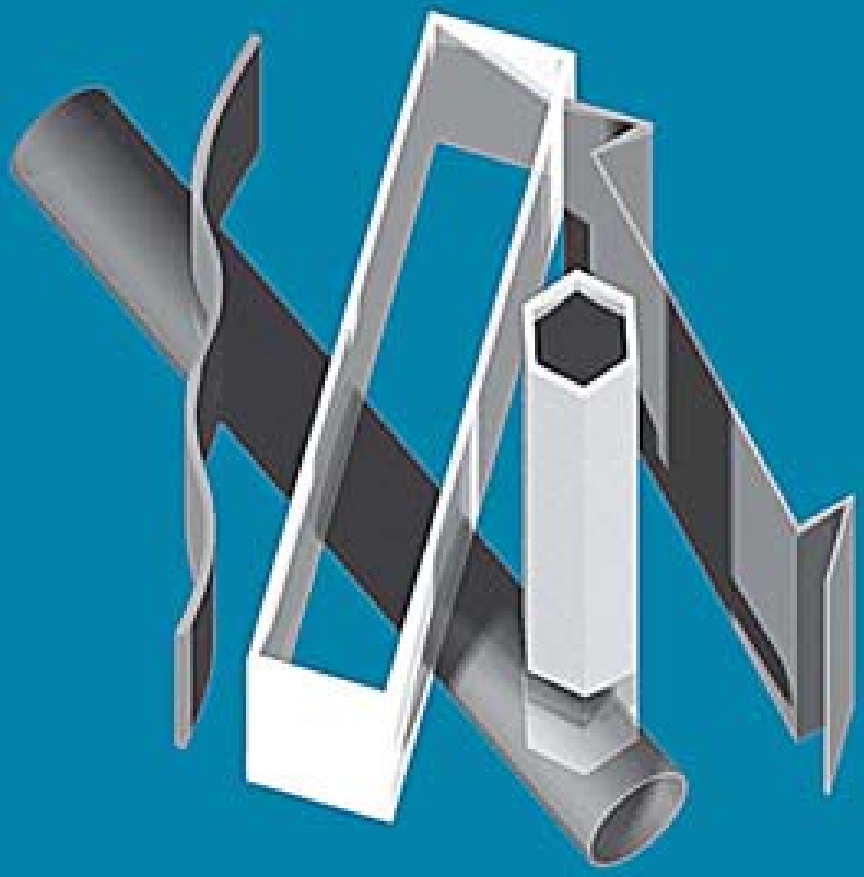
PDL HANDBOOK SERIES

Extrusion

The Definitive Processing Guide and Handbook



Harold F. Giles, Jr.
John R. Wagner, Jr.
Eldridge M. Mount, III

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Extrusion: The Definitive Processing Guide and Handbook

by

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Crescent Associates, Inc.
Rochester, New York

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EMMOUNT Technologies
Fairport, New York

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In memory of my dad, who accomplished more in his 56 years than most strive for in 100 years. To be half as good a father as he was would be a great achievement. I will never be able to teach people in my lifetime what my father taught me by example in one week.

Love,
Craig

In honor of my father. His work in the plastics industry is but a snapshot of a life dedicated to his family. May his ingenuity, creativity, determination, perseverance, and passion be an inspiration for others.

I love you Dad,
Laura

Husband, father, son, brother, friend, dedicated, hardworking, steadfast, persistent, determined, and now…author. Always willing to take on a new challenge, Harold approached each task with enthusiasm, integrity, and perseverance. His love for teaching and sharing his expertise was evident to everyone he came in contact with, from his home life to his work life. He was never too busy to stop and help.

Remembering him always with love and admiration,
Betty

Preface

Extrusion processes have long been a staple of the plastics manufacturing industry—so much so that the area has been neglected in contemporary reference literature in spite of the advances in applications and productivity that have been taking place in industry over the last decade. While a number of good books about extrusion processes exist, most are not practical for operators or for educational endeavors. The very basic books do not contain enough information to answer operators' questions, and the more advanced books are too complicated. A number of books are very theoretical and not practical for engineers or mangers.

Extrusion: the Definitive Processing Guide and Handbook is meant to fill this void in current and practical information. It consists of seven parts, in largely self-contained modular format, that comprehensively and thoroughly cover basic and advanced thermoplastics processing in the extruder. This work is a practical guide, bringing together both equipment and materials-processing considerations. It can be used as a reference for operators, engineers, and managers, or in educational courses. The chapters include information about extrusion equipment, what is happening to the material in the extruder, the extrusion process, setting up temperature profiles, starting up the extruder, extruder operation, extruder safety, auxiliary equipment, troubleshooting, and coextrusion. Each chapter ends with a list of review questions to reinforce the topics. Many chapters have references for further reading. Application examples are employed throughout, and Part 7 is devoted to extended application illustrations of contemporary applications including compounding, blown film, wire coating, and monofilament.

Certain parts of this book may best serve the information needs of particular manufacturing companies and university or other training courses. For example, companies concerned with polypropylene profiles would benefit from Parts 1, Single-Screw Extrusion, 3, Polymeric Materials, 4, Troubleshooting, and 7, Extrusion Applications.

Companies producing nylon and polyester monofilament will be especially aided by the material in Part 5, Auxiliary Equipment.

Companies engaged in compounding color concentrates on a corotating twin screw extruder will find Parts 2, Twin Screw Extrusion, 3, Polymeric Materials, 4, Troubleshooting, and 5, Auxiliary Equipment, of interest.

And any company doing coextrusion work will be interested in most of the material throughout.

Harold F. Giles, Jr., the primary author of this book, passed away before it was completed. It has been an honor and a privilege to bring Harold's work to completion. We have learned from him and we trust that others who read and study this comprehensive work will be able to profit from his understanding of extrusion processes too. Knowledge is the only thing one can give away and still retain. And in the giving, your own knowledge somehow increases.

John R. Wagner, Jr.
Eldridge M. Mount, III

for Harold F. Giles, Jr.

November 2004

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Part 1: Single Screw Extrusion

1 Extrusion Process

The extrusion of polymeric materials to produce finished products for industrial or consumer applications is an integrated process, with the extruder comprising one component of the entire line. In some applications the production lines are very long with numerous operations, requiring operators to communicate and work together to produce an acceptable finished product. If the extruder temperature profile is set incorrectly, the product ingredients are not properly formulated, the cooling on the extruder feed throat is not running properly, the melt temperature at the end of the extruder is incorrect, the cooling bath temperature is not set correctly, the puller at the end of the line is running at the wrong speed, or any other incorrect operating condition or combinations of conditions, the product may not meet customer specifications. Each step in the process adds value; consequently, the product reaches its maximum value at the end of the line. An improper setting at the beginning of the process may cause the product to be unacceptable at the end of the line after significantly more value has been added. Speeds of the different process steps must be matched to ensure product compliance.

The extrusion process is shown in Figure 1.1. Polymeric material is received, inspected, and stored. Prior to extrusion, the polymer may be blended with additives (stabilizers for heat, oxidative stability, UV stability, etc.), color pigments or concentrates, flame retardants, fillers, lubricants, reinforcements, etc., to produce the desired product property profile. Some resin systems must be dried prior to extrusion to eliminate polymer degradation due to moisture. Other resins, which do not normally require drying, may have to be dried if they are stored in a cold warehouse and brought into a warm environment, causing moisture to condense on the surface of the pellets, flake, or powder. Once the polymer or blend is properly dried and ingredients mixed, the formulation is fed to the extruder, where it is melted, mixed, and delivered to the die to shape the extrudate. After exiting the die, the product is cooled and solidified in the desired shape and pulled away from the extruder at constant velocity to attain the appropriate cross section. Secondary operations, i.e., flame treatment, printing, cutting, annealing, etc., can be done in line after the puller. Finally, the product is inspected, packaged, and shipped.

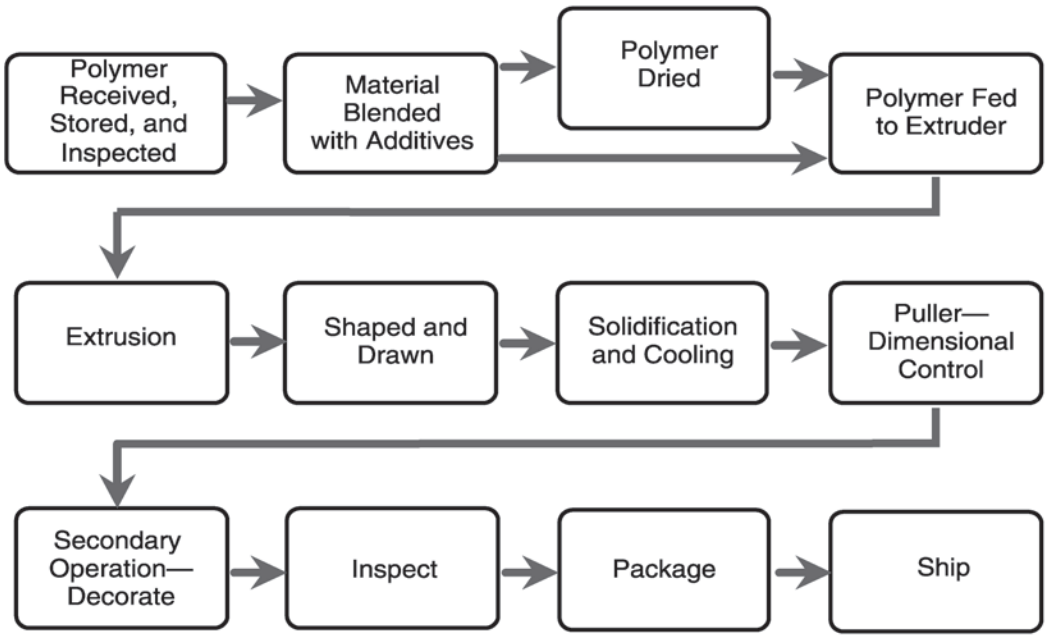


Figure 1.1. Basic extrusion process schematic.

The different parts of the process are discussed in more detail in this chapter.

1.1 Raw Material Supply

Polymer resin is shipped in different size containers depending on the quantity ordered, the processors' handling and storage capability, and the way the extruder is fed. Small lots are shipped in 50- or 55-pound bags, and large lots are shipped by tanker truck or rail. Table 1.1 shows different shipping methods. Plastic pellets can be

Table 1.1. Plastic Packaging

Package Size, Pounds	Type Package
50–55	Bags
300	Fiber Pack
1,000	Gaylord
4,000	Bulk Pack
40,000	Hopper Truck
150,000–220,000	Rail Car

air- or vacuum-conveyed around the plant to storage containers or the extruder hopper.

Pellets conveyed between storage silos, dryers, surge hoppers, and extruder hoppers must be in dedicated or properly cleaned lines to prevent product cross-contamination. All lines must be properly grounded to eliminate static electricity build-up during the resin transfer process.

Raw materials stored in warehouses without environmental controls (lack of heat or cooling) need to be brought to room temperature prior to extrusion. If the raw material temperatures vary between summer and winter, the polymer melting or softening point in the extruder will occur at a different location, leading to different melt viscosities, extrudate flow, and possible product inconsistency from season to season. Assume the raw material temperature is 50°F (10°C) in the winter and 80°F (26°C) in the summer. Additional heat must be added to the raw material during the winter months, either by a hopper dryer, allowing the polymer to come to equilibrium at room temperature, or by adding additional heat in the first zones to ensure the polymer is melting or being plasticated in the transition zone. Due to the insulative nature of polymers, a significant time period is required to heat cold pellets that sat in a cold warehouse or in a cold truck to room temperature.

Storing raw materials in a hot environment over an extended time can lead to consuming the polymer stabilization package. Most thermal stabilization packages are consumed over time as the polymer is heated. While thermal degradation happens fairly rapidly at elevated temperatures in the presence of oxygen, degradation continues at a slower rate at elevated temperatures (above room temperature but below the melting or softening temperature). Stock should be rotated to minimize long-term thermal degradation.

Many raw materials are accepted from vendors based on a "certificate of compliance." Good procedures dictate that incoming raw materials be periodically tested and a database of critical polymer properties be established. Most internal extrusion problems are not the result of raw material variations; however, in the event the wrong raw material is used, the processor should be able to identify any raw material inconsistencies immediately to minimize operating losses. Critical raw material properties for a particular application need to be identified and characterized so incoming materials are tested only for the properties that affect the final part performance. Critical properties may be viscosity, long-term heat aging, color, tensile properties, or other parameters, depending on particular end-use application.

1.2 Raw Material Blending and Mixing

Depending on the product requirements, some preblending or ingredient mixing may be required prior to extrusion. (Blending and mixing are covered in more detail in Part 5, "Auxiliary Equipment.") Unless a single polymeric material is being added to an extruder, the best way to combine different raw materials and keep them uniformly distributed prior to entering the extruder feed throat depends on different factors. Some factors to consider are

- Separation of powder and pellets
- Uniform distribution of additives introduced at low concentrations
- Separation of ingredients in flood fed hoppers
- Proper mixing
- Introduction of different levels of regrind and/or the effect of regrind particle size
- Addition of liquid additives to a single screw extruder
- Uniform distribution of powder/powder blends

The best way to meter materials and guarantee uniform component distribution is to gravimetrically feed each component with different feeders directly above the extruder feed throat. Assuming there are enough space and feeders to accommodate the various components in the formulation, gravimetric or loss-in-weight feeding ensures each component is added in the correct proportion, while addition directly above the feed throat minimizes any ingredient segregation. The downside of this approach is the cost of gravimetric feeders, the space required if there are more than four or five components, and if different size feeders are required. Assuming some components are added in very low concentrations (<1%) while other components are added in high concentration (>15%), the feeder size, feeder accuracy, and material (powder, pellets, flake, free-flowing versus compressive powder, fiber, etc.) being fed are critical to the feeder performance. If all feeders are properly sized, designed for the materials being fed (single screw feeder, twin screw feeder, vibratory, weigh belt, etc.), and there is enough

room to use a gravimetric feeder for each component, multiple feeders is the best method to ensure a repeatable, uniform formulation is being introduced to the extruder.

In many applications, a feeder is not available for each ingredient, requiring preblending. Blending depends on the ingredients being mixed and the way material is handled after blending and prior to extrusion. Assume pellets A and B are approximately the same size and are required to be premixed; proper concentrations of A and B are individually weighed and added to low intensity blending systems. Typical low intensity blending systems include tumble blenders (wide range of sizes), V-cone blender, ribbon blender, cement mixer, drum roller, or paint shaker for small lots. The same equipment can be used to mix pellets and powder. However, pellets and powder are more likely to separate when transporting the blend or loading it to a feed or extruder hopper after the blending is complete. The powder can flow between the pellets; consequently, at the beginning of an extrusion run the product may be rich in the powder component, while at the end of the extrusion run the product may be rich in the pellet component. One method to minimize this separation is to coat the pellets with a small amount of liquid such as mineral oil to provide a surface to which the powder can adhere. Of course, experimentation is required to verify the mineral oil does not affect the final product properties or performance.

Powder/powder blends can be mixed either in low intensity mixers described above or in high intensity mixers. High intensity mixers operate on the same principle as kitchen blenders. A mixing blade rotates at high speed, forming a vortex in the blender as it mixes the components. Due to the intense mixing, heat is generated and care must be taken not to melt the blend components. High intensity blenders may be jacketed to heat components or remove heat during the blend cycle. With PVC, heat softens the particle surface, allowing the heat stabilizers and plasticizers to adhere to the surface. Powder/powder blends, once properly mixed, tend not to separate during transfer, assuming the particle sizes of the different components are similar.

Uniform additive addition at low concentrations creates a mixing and blending challenge. Obviously, the best method is to feed each component directly into the feed stream with a small gravimetric feeder. However, this is not always practical or feasible. An alternative approach is to mix the additive (assume it's a powder) with some resin powder being used in the formulation and produce a masterbatch on a high intensity mixer. As an example, assume two additives, C and D, must be added at 0.5% and 0.08%, respectively, to resin B to produce a profile of material Z. A blend or masterbatch is produced by combining resin powder B with high concentrations of C and D and letting that blend down in an individual feeder. The masterbatch is added using feeder #1, and pellets of B are added via feeder #2 to produce the correct ingredients ratio in the final product. A 100-pound masterbatch is produced containing 10 pounds (10%) of component C, 1.6 pounds (1.6%) of component D, and 88.4 pounds (88.4%) of resin B. This masterbatch is let down in a 19:1 ratio, with resin B feeding at a 190 lbs/hr rate and the masterbatch feeding at a 10 lbs/hr rate to produce the correct additive ratio in the final product Z. If components C and D were fed directly to the extruder at 0.5% and 0.08%, the feed rate for each component would be 1.0 lbs/hr for component C, 0.16 lbs/hr of component D, and 198.84 lbs/hr of resin B to produce a 200-lbs/hr rate of product Z. Using a masterbatch makes feeding small concentrations of ingredients uniformly more practical.

The addition of a liquid colorant or other additive to a single screw extruder can be difficult. If the liquid is added in low concentration, it can be preblended with pellets in a tumble blender, ribbon blender, etc. Liquid can be introduced into the feed throat with a liquid feed pump, assuming the liquid does not create extruder feed problems due to pellet slippage on the barrel wall. It is critical to monitor the liquid feed rate very carefully. When using a gravimetric or loss-in-weight liquid feed pump, the pump rpm changes to keep the gravimetric feed rate constant. However, assume a volumetric liquid feed pump is being used (runs at constant rpm); the feed rate is dependent on the liquid temperature, which affects its viscosity and consequently the feed rate. Initially, a volumetric liquid feed pump must be calibrated and a graph generated showing motor rpm versus output rate in lbs/hr. On the same graph, throughput rate curves versus motor rpm curves need to be generated at different temperatures. If the liquid temperature changes during the run, the feed rate will vary and the liquid concentration will change over time. Assuming the liquid additive and feed pump are close to the extruder, it is possible the liquid temperature may increase during the run as the ambient temperature in the room increases due to the heat generated by the extruder. This results in an increase in the liquid feed rate and the wrong product formulation, unless the liquid feed pump rpm is changed to compensate for the temperature change or a gravimetric liquid feed pump is being used.

An alternative to feeding liquid into the feed throat is to pump the liquid into a two-stage extruder vent port. Modifying the vent port to accept a liquid injection nozzle connected to the liquid feed pump and a two-stage screw to accept the extra volume are the changes required. This approach eliminates the potential for feed problems associated with pellet slippage on the barrel wall in the feed zone due to the liquid. Feeding downstream does minimize mixing in the extruder and requires an appropriate screw design or static mixer (discussed later) to accomplish the mixing objectives.

1.3 Drying

Some polymers require drying prior to extrusion to prevent polymer degradation. Resins, e.g., nylon, polyester (polyethylene terephthalate [PET] and polybutylene terephthalate [PBT]), and polycarbonate, are very hygroscopic, absorbing moisture rapidly from the air. At extrusion temperatures, moisture degrades these materials to lower molecular weight polymers, resulting in poorer property performance. Proper drying to eliminate moisture is critical to obtain the optimum property performance in the final product. Other materials, e.g., acrylics, Ultem®, polysulfone, Noryl®, and acrylonitrile butadiene styrene (ABS), also absorb moisture from the air and must be dried prior to processing. Any moisture in the polymer is converted to steam in the extruder and, depending on the quantity present, can cause surface imperfections such as splay, holes in the product, or a foamy product. Some polymers, e.g., nylon, are shipped dry in moisture-proof containers. With proper handling, these resins do not normally require additional drying prior to processing. However, if the seal is broken on the container or the bag is not completely resealed after opening, the product will absorb moisture and have to be dried prior to extrusion. Polyesters are particularly sensitive to moisture and must be dried in dehumidifying dryers, transported with dry air, and blanketed with dry air or nitrogen in the extruder feed hopper. Dehumidifying dryers with –40°F (–40°C) dew points are recommended for drying most polymers. Dryers are covered in more detail in Part 5, "Auxiliary Equipment."

Formulations requiring both a dry polymer resin plus blending with other ingredients can lead to special handling requirements. Once moisture-sensitive resin is dried, it picks up moisture when exposed to the atmosphere. Additives or other components added to formulations containing hygroscopic resins need to be moisture-free. If the additives cannot be dried with the resin, special handling procedures or individual feeders are required to mix the dry resins and other additives or components at the extruder feed throat.

In some instances, resins containing moisture can be processed in a vacuum-vented extruder, with the moisture removed in the vent section. This does not work with all resins because some degradation can occur before the moisture is removed. One negative to this approach (discussed later) is the extruder is effectively shortened by approximately one-third its length, which may limit extruder throughput capacity.

Overdrying must be avoided to prevent resin degradation resulting in the loss of properties and/or the development of color bodies. Nylon 6,6 when overdried, becomes yellow and is accompanied with a loss of some properties.

1.4 Feeding Polymer to the Extruder

There are basically four ways to feed polymer to a single screw extruder:

- Flood feed
- Starve feed
- Crammer
- Melt feed

Of these, the most common is flood feeding, where resins or preblended materials are placed directly in a hopper over the feed throat, allowing gravity and the screw to feed the formulation to the extruder. A feeder with numerous hoppers to add different ingredients in the formulation simultaneously can replace the single-feed hopper. Each component is metered in the correct ratio to the feed throat. With flood feeding, the extruder throughput rate is directly proportional to the screw speed; higher screw speeds give higher throughput. Figure 1.2 shows a flood-fed extruder.

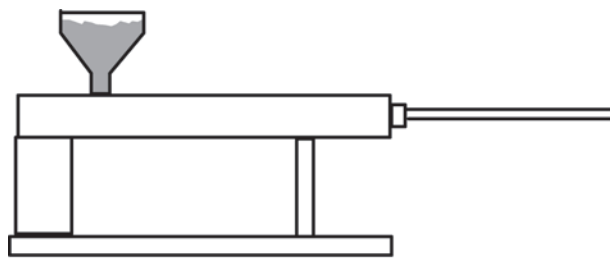


Figure 1.2. Flood-fed extruder.

Starve feeding is typically used in twin screw extrusion but can be employed with single screw extrusion. Feeders deposit the formulation directly onto the extruder screw, with the screw speed set to remove the formulation at higher rate than it is deposited on the screw. There is no material build-up in the extruder feed throat, and the throughput rate is determined by the feed rate rather than the extruder screw speed. Starve feeding has the advantage of depositing all the formulation ingredients in the proper ratio directly onto the extruder screw. Feed problems due to bridging and funneling in the feed hopper or slippage on the barrel in the extruder are

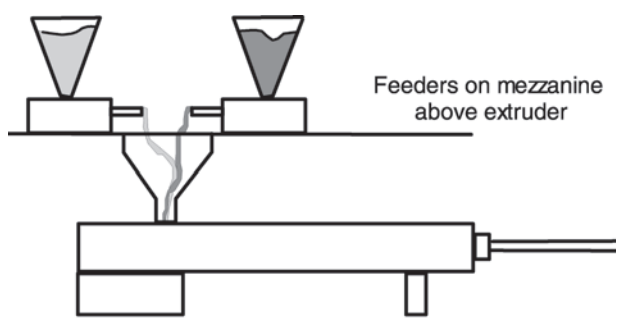


Figure 1.3. Starve-fed extruder.

typically eliminated. Feeders are normally setup directly above the feed throat or placed on a mezzanine above the feed throat to deposit materials directly on to the screw. Figure 1.3 shows a typical starve-feeding setup with two feeders on a mezzanine above the extruder.

Crammer feeding, shown in Figure 1.4, is a positive feed system that works well with low bulk density materials, materials that tend to bridge, and other hard to feed materials. A screw mechanism inside the crammer positively conveys material to the extruder. Extruder throughput rates are significantly increased with crammer feeding

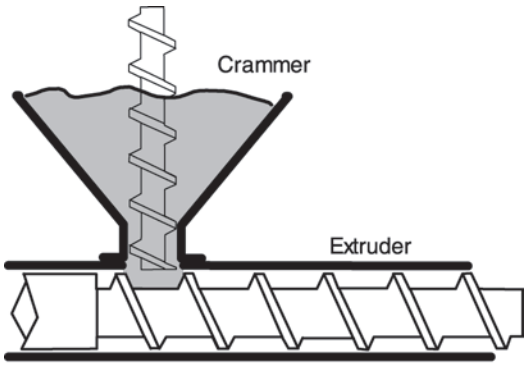


Figure 1.4. Crammer-fed extruder.

as additional material is forced into the extruder. Initially care must be taken not to overfeed the extruder to the extent the extruder is unable to melt the polymer, forcing unmelted resin into the metering section. In extreme situations overfeeding can break the extruder screw. While this positive feed system provides a good mechanism to increase throughput rates, caution must be used during start-up to minimize potential overfeeding problems.

Some extrusion operations use melt fed from another extruder, a Banbury®, roll mill, or Farrel Continuous Mixer® (FCM). A melt-fed extruder is normally shorter in length, because it is not required to melt the polymer. Basically, the extruder is a pump that generates a uniform polymer melt temperature and pressure for the die. Figure 1.5 shows a melt-fed extruder.

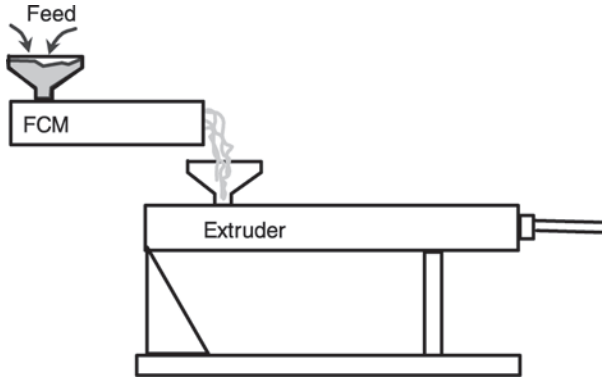


Figure 1.5. Melt-fed extruder.

1.5 Extrusion

After feeding, polymers are melted or plasticated, conveyed forward, melt mixed, and formed into a shape. These five operations within the extruder will be discussed in detail in Chapter 4, "Plasticating Behavior in the Extruder." Proper operation in each stage will produce acceptable product at high yield with proper aesthetics and the correct property balance. Figure 1.6 shows a smaller Akron Milacron single screw extruder with control cabinet.

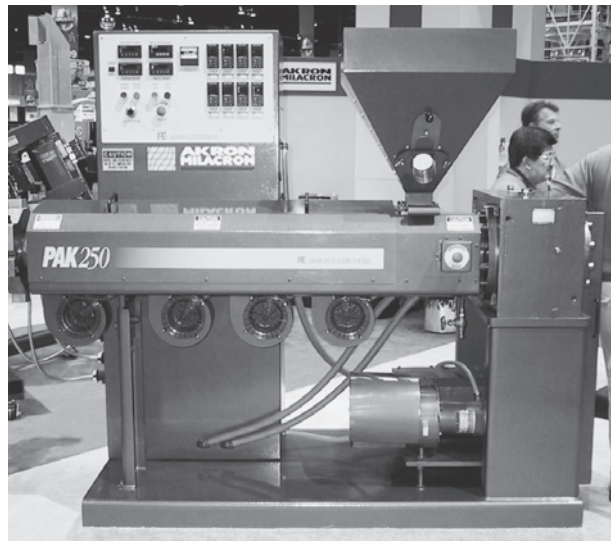


Figure 1.6. Akron Milacron single-screw extruder.

1.5.1 Shaping and Drawing

The last step in the extruder shapes the extrudate into the desired cross section. As the extrudate exits the die, the polymer molecules, which were oriented in the die land area, relax and reentangle, causing die swell. If the extrudate is allowed to droll out the die, the cross section swells, becoming larger than the die opening due to polymer relaxation. Pulling extrudate away from the extruder, with a puller farther down the line, orients the polymer molecular chains in the machine direction. Neck down or extrudate draw down is induced by this pulling action. The draw depends on the puller speed relative to the extruder output. Draw ratio is directly related to molecular orientation, resulting in higher tensile and flexural properties in the machine direction compared to the transverse position.

With a given die cross sectional area, there is one puller speed at a given extruder throughput rate that produces a product with the correct cross sectional dimensions. If the extruder throughput is increased, the puller speed must be increased proportionally to maintain the same finished product dimensions. Likewise, if the throughput is decreased, the puller speed must be decreased proportionally to maintain the same finished product cross sectional area. The draw ratio and molecular

orientation can only be increased or decreased by changing the die cross sectional area relative to the puller speed, assuming the final product dimensions are kept constant. This is easily done with sheet dies, cast film dies, or blown film dies that have adjustable die lips. Profile dies may have a fixed cross sectional opening that is not adjustable; at a given throughput rate there is only one puller speed that yields a product with the correct final dimensions. A product that tends to crack or break in the machine direction (in the plant or in field applications) may have too much molecular orientation. A new die with a different cross sectional opening is required to alter the draw ratio and change the molecular orientation to correct the problem. Higher draw ratios increase the tensile and flexural properties and the tendency to crack or split in the machine direction. Assuming most polymer molecules are aligned in one direction, it is easy to slit the product in that direction, because the molecules oriented in the perpendicular direction holding the product together are limited.

Extrudate swell, commonly known as die swell, shown in Figure 1.7, is not always visible at the die exit because the extrudate is pulled away from the extruder, causing draw down or neck down. If the extrudate is allowed to droll on the floor or is pulled from the extruder very slowly, die swell becomes very obvious. Polymer molecules in the die land area are oriented in the flow direction. The extrudate velocity profile is higher at the center of the flow front and lower near the die walls. Immediately after exiting the die, the extrudate velocity profile is identical across the entire cross section. Consequently, the velocity at the extrudate surface outside the die is identical to the velocity in the center of the extrudate. This change in the flow velocity profile gives rise to molecular relaxation outside the die and the resultant extrudate swell.

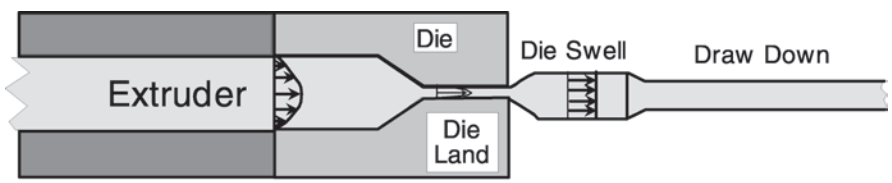


Figure 1.7. Die swell and draw down.

As the extrudate exits the die, it is quenched and possibly sized (drawn through a fixture) to maintain its final shape. Depending on the extrusion process, different methods are available to quench the final product. Cast film and sheet are quenched on rolls; blown film is quenched by air in a blown film tower; profiles, pipe, and tubing are quenched in calibration tanks filled with water and in some cases connected to a vacuum system; strands and monofilaments are quenched in water baths; wire coating is done horizontally in air or water; and large part blow molding is quenched in molds.

1.5.2 Solidification and Cooling

Extrudate cooling is normally accomplished with water, air, or contact with a cold surface. Semicrystalline polymers, i.e., polyethylene, polypropylene, nylon, polybutylene terephthalate, etc., have very sharp melting points and consequently very sharp solidification temperatures. Amorphous polymers, on the other hand, do not melt but enter a rubbery state above their T_g (glass transition temperature, discussed later). As the temperature increases, polymer chain mobility continues to increase until the polymer flows and is easy to process. When amorphous materials cool, the part temperature needs to be below the material T_g to ensure the final part dimensions. Thick cross sections form a surface skin while the center is still molten. This allows the extrusion line to be run at higher rates; however, if product dimensional tolerances are very tight, the entire product should be cooled below the melting point if it is a semicrystalline polymer and below the T_g if it is an amorphous polymer before exiting the cooling medium. Cooling from elevated to room temperature after the product is completely solidified results in additional product shrinkage and dimensional changes.

Proper part cooling is critical to produce warpage-free parts with the acceptable dimensions and performance. Part warpage is caused by differential shrinkage. To minimize differential shrinkage, the part must be cooled uniformly on all sides. If one side or area of the extrudate solidifies before another, the part will warp, bending toward the side that solidified last. If one side of the extrudate is dragged over an object in the cooling operation, molecular orientation is induced on that side, causing it to shrink differently from the other side, leading to warpage. Warpage is discussed in more detail in Part 4, "Troubleshooting the Extrusion Process."

Cooling rates with semicrystalline polymers are critical to develop the correct amount and crystal size in the final product. Rapid quenching leads to small crystal growth development and low crystallinity. Heating or annealing later (heated for a specific time and temperature above its T_g) leads to additional crystal growth in the solid state. Accompanying any increase in crystallinity is a reduction in volume, a change in the part dimensions, and possibly warpage. To maximize the crystallinity and crystal size, the extrudate should be cooled slowly. Cooling rates can be critical in maximizing product performance and reproducibility. Cooling rates are determined by throughput rates, part thickness, and cooling medium temperature (water bath, roll, or air temperatures).

Drawing products in the solid state (monofilament production, oriented film, or biaxially oriented film)

maximizes molecular orientation and directional properties. In semicrystalline polymers, drawing can lead to additional crystallinity development through molecular alignment.

In some extruded products, the cooling rate and treatment during cooling are critical to maximizing the final product properties required by the customer. In sheet or cast film, roll stack temperatures and surfaces determine product aesthetes. Highly polished rolls run at relatively high temperatures produce polished, glossy surfaces. A matte finish on the product is obtained by using rolls with a matte finish, and a matte finish on one side and a polished, glossy surface on the other are produced by using a matte finish and a highly polished roll. A vacuum sizing tank is used for hollow profiles or pipe and tubing, where the extrudate is run through sizing rings under water with a vacuum above the water. The fixturing and cooling required to maintain final dimensions depends on the application.

1.5.3 Puller

The puller controls the draw and tension on the material from the extruder exit through the cooling and solidification steps. Final product dimensions are controlled by the extruder throughput rate and the puller speed. With a fixed die opening and given throughput rate, there is only one puller speed that produces the correct product dimensions. Consequently, the puller speed must be matched to the extruder output rate. If periodic variations occur in the puller speed or extruder output (due to surging), the product dimensions continuously change. Slippage in the puller can cause thicker sections in the parts that do not meet finished product specifications. A caterpillar type puller is shown in Figure 1.8.

Pressure exerted by the puller must be sufficient to prevent product slippage in the puller, but low enough to prevent part distortion or marks on the product surface. Extreme puller pressure can crush the final part, rendering it useless. The puller may be a long distance from the extruder; however, it must be properly aligned with the extruder to prevent the part from being pulled in one direction or another, inducing molecular orientation that may lead to warpage.

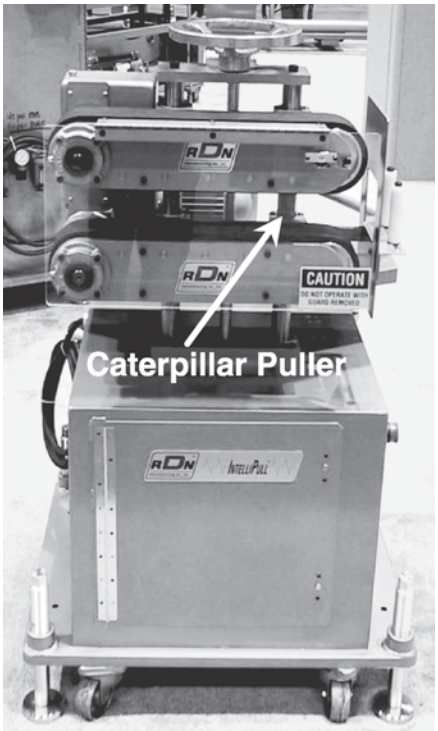


Figure 1.8. RDN puller.

Dimensional variations in the final product normally result from the extruder (surging, power input variations, slippage on the screw, poor feeding) or the puller (slippage, improper compression on the part, or power output variations).

1.6 Secondary Operations

Numerous secondary operations are performed in line to minimize product handling and improve production efficiencies. Some secondary operations done in line include cutting to length, drilling or punching holes, corona or flame treatment, decorating (painting, printing, gluing something to the surface, etc.), attaching adhesive labels, welding, etc.

1.7 Inspection, Packaging, and Shipping

Visual inspection or gauging is done at the end of the line to verify all parts meet specification. Using statistical process control (SPC) quality control guarantees products meet specifications, and visual inspection can be eliminated. The problem associated with visual inspection is some defective parts always pass through the system because of human error. Visual, subjective part inspection should be eliminated as much as possible by installing other quality control methods to assure all parts meet customer specifications. In addition to visual inspection, part weight and/or dimensions can be checked prior to packaging. Proper SPC procedures eliminate much QC work associated with product quality assurance.

Retained samples of each production lot should be stored, in the event of a customer complaint. Physical property verification or evaluation of deficiencies identified by the customer can be monitored or tested on both the customer's parts and the retained samples to assist in the identification and correction of the problem.

The final steps in the extrusion process are to package the product according to customer requirements and ship the parts.

Review Questions

1. What controls product dimensions?
2. What part of the extrusion process contributes to final part warpage?
3. Name three factors that might lead to part warpage.
4. What are the different methods of feeding polymer to a single screw extruder, and what controls the extruder feed rate in each method?
5. In the extrusion process, which steps affect physical properties and how?
6. What materials need to be dried prior to extrusion?
7. What are some methods of blending polymers? What are the best methods to use for blending powder and pellets, pellets and pellets, powder and powder?
8. What is a masterbatch?
9. What is die swell?
10. If the production rate is 300 pounds/hour and it is necessary to feed 0.07% of component X, 2% of component Y, and 1.2% of component Z with 96.73% of polymer pellets L, what is the best way to mix and add the material to an extruder?
11. At what step in the extrusion process is the product worth the most money?

2 Extruder Safety

Safety is each employee's responsibility, ranging from the janitorial staff to company president. It is the responsibility of each associate to work safely and assist other employees to operate safely, endeavoring to eliminate all unsafe acts that lead to major accidents. Of all accidents, 96 percent are caused by human error, carelessness, or the attitude, "It won't happen to me." Consequently, our personal safety plus the safety of those around us is the responsibility of each employee. It is essential to obey all work area rules and be alert for unsafe acts and conditions. Besides working safely, it is important to encourage those around you to work safely. Before a job is started, it needs to be thought through completely and determined if it can be done safely. If it can't be done safely, don't do the job until you obtain the proper equipment or develop the proper procedure to do it safely. It is important to realize the hazards associated with each job and not take any shortcuts that might put you or your associates in the way of potential danger and serious accidents.

Training new employees must include safety training plus the hazards associated with all equipment. In addition to potential equipment hazards, new employees need to know the location of all safety equipment (fire extinguishers, fire blankets, first aid, who to contact in an emergency, etc.) and understand the correct procedures to follow in the event of an accident or injured employee. Who should be contacted? What is the procedure for reporting accidents? What is my responsibility? Where is the muster point? What do different alarms mean, and what is my response?

Associates need to help each other. If you see fellow employees performing unsafe acts, help them understand proper procedure and why what they were doing is unsafe. This is an act of caring and concern for our fellow employees, not an act to belittle or make somebody look foolish.

The most important step in safety is to understand the potential hazards, realize you are not invincible, and it can happen to you. Follow procedures and think any job through thoroughly before starting to evaluate the potential for injury to yourself and others. Don't be the bull in the china shop, charging ahead without thought. **If a job can't be done safely, don't do it until procedures, methods, or equipment is available to do it safely.**

2.1 Hazards Associated with an Extruder

The three biggest potential safety hazards associated with extruders are burns, electrical shock, and falls. Without proper protective equipment, burns can be commonplace among employees working around extruders. Touching a hot die or handling extrudate without gloves normally causes burns. Long sleeves with properly approved thermal gloves should be worn when working around the die, changing the die, tightening die bolts, or other functions performed on the die. If insulation is placed around the die, make sure it is in good shape and properly installed. Hot extrudate from the extruder will stick to your skin. Since polymeric materials are great insulators, after sticking to the skin they cool very slowly, continuing to burn the skin affected.

Never stand in front of a die when a single screw extruder is starting up. Air in the extruder and possibly gas from degraded products (if the extruder has been sitting at extrusion temperature with material in the barrel for some time) is forced out of the extruder on start-up. If some polymer is left in the barrel, trapped air can be compressed, blowing the hot polymer out of the die. Standing in front of the extruder creates an unsafe condition where molten polymer can be blown out of the die, land on you, and burn you. Polymer can stick to gloves, where it holds heat for a long time, and can burn you through the gloves if the proper type of glove is not used. When removing the die and/or screw from an extruder (they are normally hot), wear the proper protective equipment (heavy duty gloves and protective thermal sleeves) to prevent burns. Dies can be heavy; therefore, a back brace or other equipment to lift and hold the die can prevent back injuries.

The potential for electrical shock exists when improperly trained employees remove the extruder covers, thus exposing bare wires and electrical connections. Extruder heater bands are normally 220 or 440 volts and can cause serious electrical shock. Check the wires to the heater bands on the die and adapters to assure there are not frayed, bare, or exposed wires that can cause electrical shock. In some extrusion processes, water-cooling baths are very close to the die, which can create additional electrical hazard. Unless properly trained, operators should never remove guards, exposing electrical terminals on heaters or open electrical cabinets, to solve electrical problems.

The third major potential safety hazard around extruders is falls. Pellets spilled on the floor are slippery and need to be cleaned up immediately. At start-up the extruder normally generates some scrap, which may be on the floor around the die. This creates tripping hazards that must be removed immediately. Occasionally processing issues arise at start-up, leading to too much material on the floor around the extruder. In these situations, the extruder should be shut down, the area cleaned, and the extruder restarted. Some extrusion processes use water for cooling. Water spills on the floor should be

removed with a wet/dry vacuum or squeegeed to a drain. Wet floors are very slippery and can cause falls.

The most dangerous area around an extruder is the exposed screw turning in the feed throat. **Never, never stick your hands or fingers into the feed throat.** If the screw is turning, there is incredible power that can quickly remove a finger. If the feed throat is hot, you may also get burned.

The most dangerous time during extruder operation is start-up. An extruder is a pressure vessel. Material is being fed into one end with a positive conveying mechanism (screw) operating at high horsepower. If the die end of the extruder is blocked with solid plastic or contaminants, incredible pressure can build up very rapidly and blow the die off. Always start the extruder screw slowly and monitor the die pressure closely until polymer is flowing continuously. Once die flow is established, the screw speed can be safely increased. As mentioned previously, never stand in front of an extruder during start-up in case molten plastic is blown out of the die under high pressure.

Extruders are equipped with rupture disks to prevent high pressure form blowing the die off the end and pressure gauges (discussed later) to monitor the pressure in and before the die. Make sure the pressure gauges are functioning properly. If the extruder does not have a rupture disk at the extruder head to relieve high pressure, it should have a pressure gauge with a feedback loop that automatically shuts the extruder down when a preset pressure is reached. Most modern extruders are equipped with both a rupture disk and high-pressure sensor that will shut the extruder down in the event of high-pressure situations. Having both the rupture disk and pressure sensor will protect in the situation where on heat-up, cold spots trap melting and expanding polymer in the head and melt pipe area. It is possible to exceed 20,000 psi in a pool of trapped polymer.

Each extruder should be equipped with a fume hood at the die or vent port to remove any fumes generated.

2.1.1 Hazards Associated with Takeoff Equipment

The safety hazards associated with takeoff equipment depend on the extrusion process and takeoff equipment. Pinch points associated with nip rolls, pullers, and roll stacks are one potential safety hazard requiring careful operation. If two operators are running equipment containing nip rolls, they must communicate to verify all operators are clear when nip rolls are closed. Loose-fitting clothing that can be caught in nip rolls or pullers must be avoided. Some lines have rolling knives or knives for edge slitting. These should be guarded, and operators must use caution when working in those areas.

High-speed rotating rolls present special hazards. Guards around all rolls and nip points must be kept in place to prevent injury. Arms, fingers, and hands can easily be pulled into high-speed rolling equipment, causing severe personal injury or dismemberment.

Scrap from start-up lying on the floor poses a tripping hazard. This should be picked up and disposed as soon as the line is running. If start-up problems continue and prevent clean up, the equipment should be shut down, the area cleaned, and the line restarted.

Noise above 80 dB requires that hearing protection be used by all people in the area. If the noise level is below 80 dB, employees may still want to wear hearing protection to prevent long-term hearing loss.

Like the extruder, identify potential safety hazards associated with the takeoff equipment. Form a plan to avoid potential hazards. Know where all emergency stop buttons are and verify that they work. Don't take the approach, **"It Won't Happen to Me."**

2.1.2 Personal Protective Equipment

Personal protective equipment exists to make your job safer. Determine what equipment you need to do your job safely and use it. Following is some of the personal protective equipment available:

- Safety glasses with or without side shields
- Safety shoes
- Ear protection
- Gloves
- Thermal insulated gloves for hot applications
- Long sleeves
- Hard hats
- Face shield
- Goggles
- Back brace
- Wrist brace
- Floor mats

2.1.3 Lock-Out, Tag, and Clear Procedure

Anyone working on equipment should have a personal lock with his or her name on it and the only key. Prior to doing maintenance or other work on the equipment, turn off the power switch and lock out the switch with your personal lock. Employees working on the line must attach their own lock. After locking out and tagging the equipment prior to doing any work, each worker attempts to start the equipment to verify it is off and cannot be turned on. Once the maintenance or other work is completed, each worker removes his own individual lock before the equipment can be restarted. This procedure prevents somebody from getting hurt while working on equipment when another person inadvertently starts the equipment.

2.2 Proper Training

Don't run equipment without proper training and understanding the potential safety hazards associated with the operation. Know where you can get hurt and understand how all the equipment and controls operate. Training includes start-up and shutdown procedures, understanding the caution or warning signs on the machines, and operating time on the equipment with an experienced operator.

2.3 Inspection and Housekeeping

Before each shift, evaluate the operating area and plant in general, looking for unsafe conditions, e.g., tripping hazards, exposed wires, water on the floor, etc. Determine what you are going to do on your shift and review the operation for safety.

Good housekeeping is directly related to safety. A cluttered, dirty area will lead to accidents and reflects on your attitude toward the job. A proper storage area for all tools and equipment makes the job easier and the plant a better place to work.

2.4 Material Safety

Understand the materials you are using by reviewing the Material Safety Data Sheets (MSDSs). Improper operating conditions or purging with the wrong materials can have serious consequences. Overheating polyvinyl chloride (PVC) generates hydrochloric acid (HCl), which attacks the lungs and rusts plant equipment. Never mix acetal with nylon, PVC, fluorinated polymers, or ionomer in an extruder, as they will react and give off formaldehyde.

PVC has limited thermal stability and should not be left in a hot extruder. PVC degrades in an autocatalytic reaction, generating HCl. Proper purge material should be available to remove PVC from the barrel if the extruder is going to be down for an extended time. Operators who have the flexibility to change extruder temperature profiles need to understand the upper processing limits when extruding PVC or other temperature-sensitive polymers.

Review Questions

1. What is the most dangerous time during extrusion and why?
2. Where are the most dangerous locations around an extruder and why?
3. What are some potential hazards associated with extrusion?
4. What is the "lock out, tag, and clear" procedure, and when should the procedure be used?
5. Why is housekeeping important?
6. What is a near miss?
7. What hazards are associated with takeoff equipment?
8. What is some personal protective equipment?
9. What materials should not be mixed with acetal in an extruder?
10. What happens if PVC is overheated?

3 Single Screw Extruder: Equipment

Introduction

What are the objectives or goals to be accomplished with the extruder in the overall extrusion process? The standard answer is to produce a quality product that meets the customer specifications 100% of the time. While this is true, the extruder has five distinct goals or objectives to achieve in the extrusion process that will result in a quality product if done correctly:

- Correct polymer melt temperature
- Uniform/constant melt temperature
- Correct melt pressure in the die
- Uniform/constant melt pressure in the die
- Homogeneous, well-mixed product

The next two chapters focus on the extruder equipment and polymer behavior in the extruder. To optimize an extrusion process, it is not sufficient to simply understand the equipment, how it operates, and how it functions; it is essential to understand the polymers and their reaction and behavior in the various process steps. To effectively troubleshoot an extrusion process, the equipment and its interaction with the material, along with the material properties, must be understood. This chapter focuses on single screw extruder equipment, and Chapter 4 focuses on the plasticating behavior in the extruder.

3.1 Equipment

The simplest extruder is a ram extruder, shown in Fig. 3.1. Pressure is applied to a piston, forcing the extrudate out the die. Heat applied to the barrel melts the material and lowers its viscosity. With the correct combination of heat and pressure, extrudate is forced out the die in the desired shape. There are many problems associated with this extruder. First, it is a discontinuous process. Second, plastic is a great insulator; consequently, it takes a relatively long time to heat the billet uniformly from the barrel surface to the center. Using high barrel temperature to increase the melting rate tends to degrade the resin at the barrel wall before the billet is molten. Lowering heater band temperatures to minimize degradation requires long heat soak times to soften the billet properly. Third, shear heating from the ram movement is minimal.

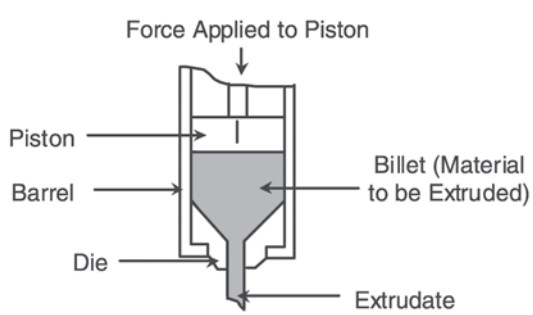


Figure 3.1. Ram extruder.

The key components in a single screw extruder are shown in Fig. 3.2. A single screw extruder has five major equipment components:

- Drive system
- Feed system
- Screw, barrel, and heaters system
- Head and die assembly
- Control system

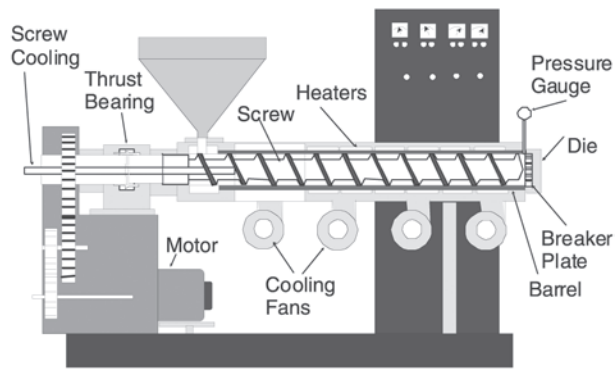


Figure 3.2. Single screw extruder.

The drive system comprises the motor, gear box, bull gear, and thrust bearing assembly. The feed system is the feed hopper, feed throat, and screw feed section. The screw, barrel, and heating systems are where solid resin is conveyed forward, melted, mixed, and pumped to the die. Extrudate is transported and shaped in the adapter and die, respectively. Finally, the control system controls the extruder electrical inputs and monitors the extruder feedback. Computer-designed extrusion controls not only run and monitor the extruder, but also can control the entire extrusion process with feedback loops that automatically change feeder settings, puller speeds, screw speeds, etc., to maintain product quality.

Table 3.1. Inch and mm Extruder Size

< 4" (100mm)		> 4" (100mm)	
inch	mm	inch	mm
0.5	15	4.5	120
0.75	20	6	150
1.0	25		180
	30	8	220
1.25	35	10	250
1.5		12	300
1.75	45	14	350
2.0	50	16	400
2.5	60	18	450
3.0	70	20	500
3.5	90	24	600

Extruders are sold based on screw or barrel diameter and the length (L) to diameter (D) ratio, called L over D or L/D, of the barrel. In the United States, extruder diameters are measured in inches; other countries use millimeters. Typical diameters are shown in Table 3.1.

Note that the match-up between inch and mm is only approximate. There are several standard sizes up to 10 inches (250 mm) in diameter. Larger sizes are usually custom built according to customer requirements.

Extruder *L/D* describes the relative length of the screw and extruder barrel. The *L/D* ratio is given by Eqs. (3.1) and (3.2):

$$L/D = \frac{\text{Flighted Length of Screw}}{\text{Outside Diameter of Screw}} \quad (3.1)$$

$$L/D = \frac{\text{Flighted Length of Screw} - \text{Axial Length of Feed Pocket}}{\text{Outside Diameter of Screw}} \quad (3.2)$$

Which *L/D* definition is used depends on the equipment manufacturer. Some manufacturers include the axial feed pocket length as part of the barrel length, while others do not.

Throughput is directly related to the extruder *L/D*. Two extruders with the same diameter but different *L/D*s have different throughput capacities. Longer extruders (higher *L/D*) have more melting and mixing capacity, allowing the extruder to be run at higher rates. Short *L/D* extruders have the following advantages:

- Less floor space required
- Lower initial investment cost
- Lower replacement part cost for screws and barrels
- Less residence time in the extruder when processing temperature-sensitive materials
- Less torque required
- Less horsepower and corresponding motor size

Longer *L/D* extruders have the following advantages:

- Higher throughput because of screw design
- More mixing capability
- Can pump at higher die pressure
- Greater melting capacity with less shear heating
- Increased conductive heating from the barrel

Some typical extruder *L/D*s are 18:1, 20:1, 24:1, 30:1, 36:1, and 40:1.

If one knows the *L/D*, one can calculate how long the barrel is. If we have a 2.5-inch extruder with a 24:1 *L/D*, the length is calculated as follows:

$$L = L/D \times D = \frac{24}{1} \times 2.5 = 60 \text{ inches}$$

Calculate the length of the barrel on a 4.5-inch diameter extruder with a 30:1 *L/D*.

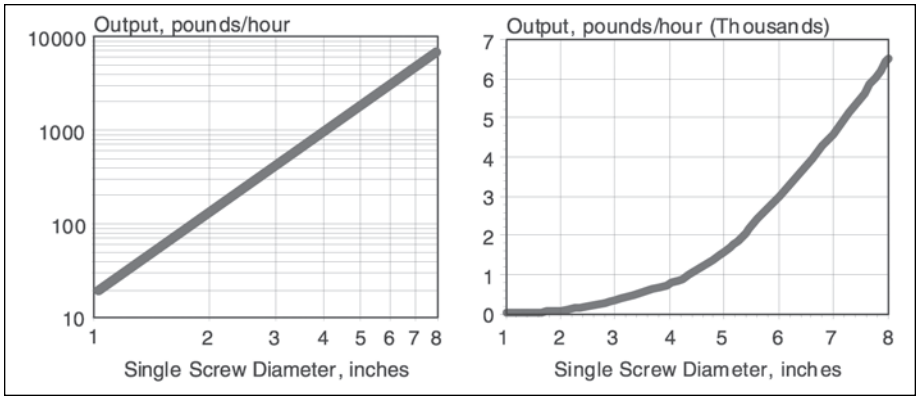


Figure 3.3. Nominal extruder throughput.

Throughout rates are directly proportional to the screw diameter. Larger diameter extruders have greater output. Figure 3.3 shows typical throughput charts for different size single screw extruders.

3.2 Drive

The drive turns the screw at a constant speed over a large speed range while supplying sufficient torque to process the polymer being used. Screw speed variations are directly proportional to throughput variations, which can cause changes in product dimensions. In practice, screw speed variations are observed either by the screw rpm or the extruder motor load readout (normally given in torque, percent load, or amps). Screw speed can be monitored with a hand-held tachometer or read from the control panel.

The most popular drive on large extruders is a DC motor connected to speed reducers that convert motor rpm to screw rpm. To generate maximum torque, DC motors are run at maximum rpm (1750 rpm). Two options for motors are standard DC motors and brushless DC motors. At one time, all extruder drives were powered with DC motors; however, AC motors are now found on smaller extruders. AC motors don't have to be run at maximum rpm to obtain maximum torque, and the new flux vector AC drives can achieve torque and speed control superior to DC brushless motors and more economically. Flux vector AC drives have solid state power switching, with a microprocessor controlling both magnetizing current and torque-producing current through vector calculations. The other option is for a variable frequency AC drive with tachometer and encoder feedback.

Direct and indirect drive systems are used to transfer motor power to the screw. A direct drive system, shown in Figure 3.4, uses a quick-change gearbox to convert the motor rpm to the desired screw rpm. Changing the gear

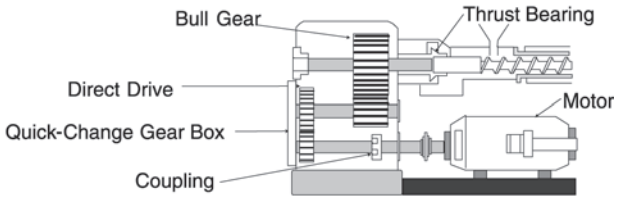


Figure 3.4. Direct drive.

ratio modifies extruder screw speed range. Before changing the gear ratio to increase the extruder screw speed, verify that the motor has sufficient horsepower to generate enough torque to process the quantity of plastic resin intended for use on the extruder at the higher screw speeds. Changing gears without determining if the motor horsepower is sufficient to process the new volume can result in high screw speed with insufficient torque or power to turn the screw when the extruder is full. Screw torque or motor power is critical in melting the polymer (discussed in Chapter 4).

An indirect drive extruder, run with pulleys, is shown in Figure 3.5. While Figure 3.5 shows the location for only two pulleys, an indirect drive extruder normally has three to five pulleys and belts. Belts are shipped in matched sets and must all be changed simultaneously.

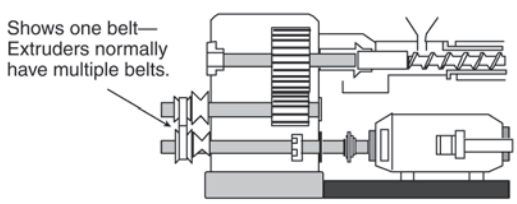


Figure 3.5. Indirect drive.

Belt slippage can result in throughput variations caused by screw speed fluctuations. To change the screw speed range, the pulley sizes are modified. When changing pulleys, make sure all pulleys are properly aligned before restarting the machine. As with the direct drive system, verify that the motor has sufficient horsepower to meet the extruder torque requirements at higher throughputs before changing any pulleys.

The thrust bearing is located between the screw shaft and the drive output shaft. As the extruder screw rotates, it is attempting to twist itself out the back of the extruder. Combining this with the die head pressure, the screw is generating high forces on the thrust bearing. Figure 3.6 shows the thrust bearing assembly. For every action there is an equal and opposite reaction; in an extruder the load on the thrust bearing is directly proportional to the head pressure and the screw diameter. Force on the thrust bearing is obtained by multiplying the screw cross sectional area times the extruder head pressure.

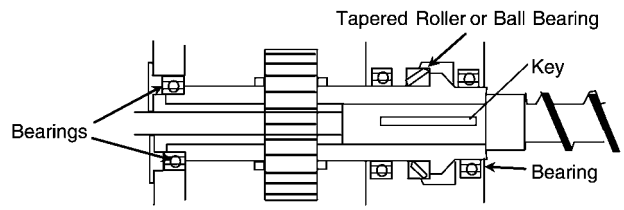


Figure 3.6. Thrust bearing assembly.

Below is a thrust bearing load calculation. Calculate the typical force on the thrust bearing of a 4.5-inch extruder running at 5000 psi die pressure.

$$\begin{aligned} \textit{Screw Cross Sectional Area} &= \pi \times r^2 \\ &= 3.1416 \times (2.25)^2 \\ &= 15.9\,\text{inch}^2 \\ \textit{Force on Thrust Bearing} &= \textit{Pressure} \times \textit{Area} \\ &= 5000\,\text{psi} \times 15.9\,\text{inch}^2 \\ &= 79{,}500\,\text{pounds force} \end{aligned}$$

Thrust bearing life is rated in B-10 life, which is measured in hours. B-10 life assumes that 90% of the thrust bearings running at 5000 psi head pressure and 100 rpm screw speed will exceed 100,000 hours or 10 years before failure. Ten percent of the thrust bearings running under these conditions fail within 10 years. If extruders are run at higher pressure or speeds, the anticipated thrust bearing life decreases. Likewise, thrust bearing life is increased at lower speeds and pressures.

Figure 3.7 shows a Davis Standard gear box with a transparent top, exposing the bull gear, other gear reduction from the motor, part of the motor, and the belt drive under the protective cover.

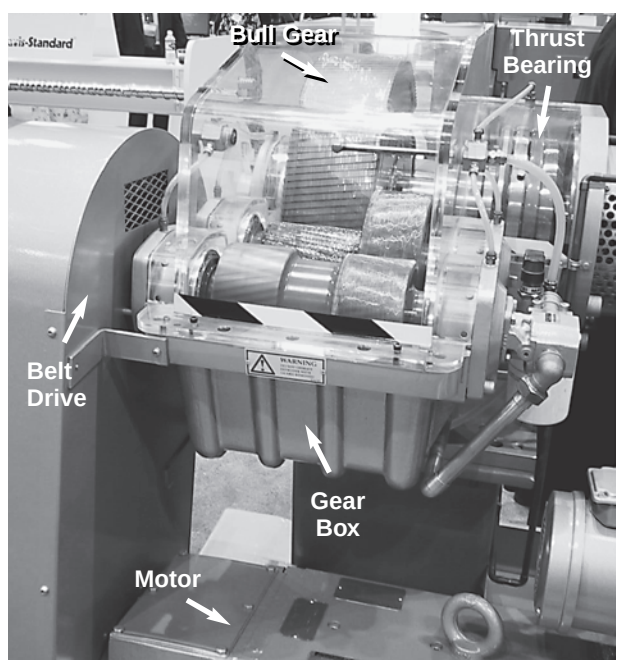


Figure 3.7. Davis Standard gear box, drive motor, and belts.

3.3 Feed

The two solid feed systems that rely on gravity are flood and starve feeding. Both feed systems have a hopper sitting directly over the extruder feed throat with the hopper opening size matched to the feed throat opening. All dead spots in the hopper and feed throat are eliminated to prevent polymer or additive build-up that can later cause cross-contamination or bridging. The feed throat section, attached directly to the extruder barrel, is water jacketed for cooling. In operation, water flow can be measured with a flow gauge on the cooling line return from the feed throat or simply by feeling the feed throat area and verifying it is not overheated. The feed throat should feel warm to the touch but not hot. Figure 3.8 shows a typical flood-fed extruder and the cooling channels in the feed throat area. The purpose of water cooling is to prevent feed materials from softening, becoming tacky, and sticking together in the feed throat, causing bridging or premature melt blockage in the feed section. A good insulative barrier is installed between the barrel and feed section to minimize heat transfer from the barrel.

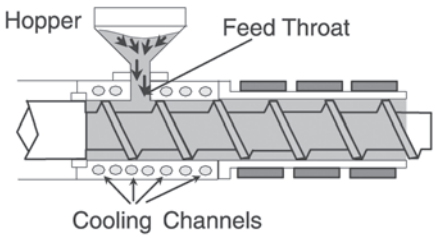


Figure 3.8. Feed throat section of flood-fed extruder.

Feed throat and hopper geometry allow material to flow freely into the extruder with minimum restriction. Standard feed throat design for pellets or powder is shown in Fig. 3.9, geometry A. Feed throat geometries B and C in Fig. 3.9 are more appropriate for melt-fed extruders. Pellets fed in configuration B can wedge between the barrel and screw, causing the screw to deflect.

Figure 3.9. Feed throat configurations for flood-fed extruders.

Grooved feed throats are popular in blown film and other applications to increase extruder output. Figure 3.10 shows a grooved feed section from a large Davis Standard feed section. Notice how the grooves are deep in the beginning of the feed section under the feed hopper and disappear just prior to going into barrel zone 1.

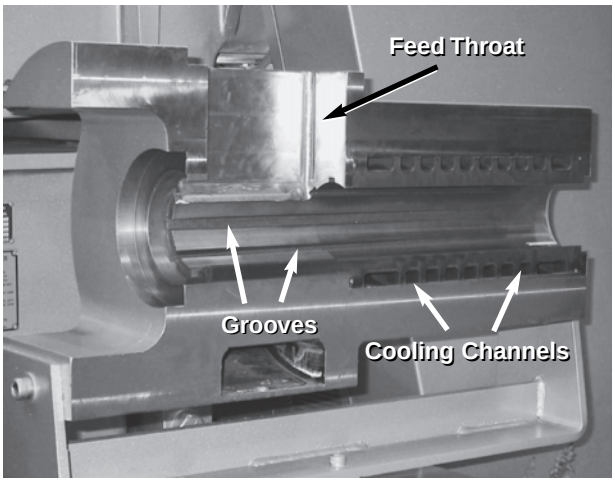


Figure 3.10. Davis Standard grooved feed throat.

The cooling channels around the feed section remove frictional heat generated by the rotating screw and pellet compression into the screw channels, preventing the pellets from premature melting. The grooves shown in Fig. 3.10 are in the axial direction but can also be helical around the feed section. The advantage of a grooved feed throat is increased friction between the pellets and the barrel wall, resulting in higher throughputs. Extruders with grooved feed sections require three alterations:

- Excellent feed throat cooling to dissipate the frictional heat generated and the capability of handling high pressures (15,000 psi plus) in the grooved feed section
- A good insulative barrier between the barrel and the feed section to minimize heat transfer
- Lower compression ratio extruder screws to handle the increased throughput rate

Normal screw compression ratios (discussed later) are approximately 1.5:1.

Other ways to feed (crammer and melt) are discussed in Chapter 1.

3.4 Screw, Barrel, and Heaters

The screw conveys material forward, contributing to the heating and melting, homogenizing and mixing the melt, and delivering the melt to the die. The barrel and heaters help heat and melt the polymer by controlling the temperature in the different zones, preventing material from overheating and degrading. The screw, in combination with the barrel, feeds polymer to the die, building pressure in the die.

Barrel hardware is shown in Fig. 3.11. Heaters are located along the barrel, with thermocouples in each zone to control the heaters and barrel temperature. The heaters

cover as much barrel surface area as practical, minimizing hot and cold spots along the barrel length. In an individual extruder temperature zone, there may be one, two, or three heater bands with one thermocouple controlling them. Assume the heater band closest to the thermocouple burns out; the other two heater bands have to supply all the external energy required, creating the possibility that the area is hotter near the two heater bands that are working. In the event the band farthest from the thermocouple burns out, the barrel area under the burnt-out heater is anticipated to be cooler than areas where the heaters are functioning properly near the controlling thermocouple. Burnt-out heater bands should be replaced as soon as possible to assure uniform heat input. Thermocouples placed in the barrel wall penetrate as close to the barrel liner as practical. Water- or air-cooling in each zone is used to control the barrel temperature. At the extruder head prior to the breaker plate, there is a pressure transducer to measure head pressure and a rupture disk for safety, in case there is a sudden and/or unexpected pressure rise. Barrels may be lined with a bimetallic liner to increase service life.

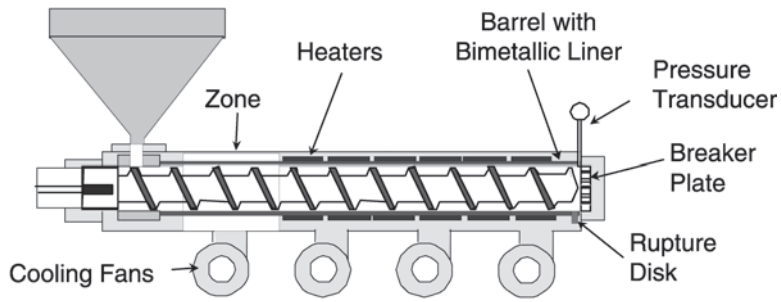


Figure 3.11. Screw, barrel, heater band configuration.

Barrels are fabricated from solid carbon steel or other material. Nitriding about 0.3 mm deep hardens the inside barrel surface. However, nitriding is not particularly effective when running abrasive fillers such as glass, mineral fillers, or other fiber reinforcements. Stainless steel barrels with hardened interiors are an option for small extruders. However, hardening stainless steel lessens the corrosion resistance, and stainless steel is not a particularly good heat transfer medium. A second approach to improve corrosion or abrasion resistance in the barrel is to use a bimetallic coating. Coating is thicker (1.5–3 mm) than nitriding, providing better wear resistance. Table 3.2 shows some coatings and their wear properties. The third approach to improve abrasion or corrosion resistance is a barrel liner, which is a thin-walled tube of stainless steel, nickel-based alloy, or hardened carbon steel, inserted in the barrel. Heat transfer may suffer slightly if there is an air gap between the liner outside diameter and barrel inside diameter. The barrel inside surface should be harder than the screw to prevent barrel wear. Screws tend to wear faster than the barrel because the barrel surface area to screw surface area is about a 10:1 ratio, meaning the screw flights come in contact with only 10% of the barrel wall during each revolution.

Table 3.2. Bimetallic Coating

Base Elements	Other Elements	Rockwell C Hardness	Comments
Fe	Ni, Si, B, Cr	50–65	Excellent wear resistance, no corrosion protection
Ni/Co	Cr, Si, B, Fe	45–60	Good wear resistance, best for corrosion protection
Ni/Cr	W, B, Fe, Si	60–65	Best for wear resistance, best for highly filled materials, very good corrosion protection

In a new barrel or extruder installation, the barrel should be bore-scoped to define the extruder centerline and verify that the thrust bearing and shaft are properly aligned with the feed throat and barrel. Leveling an extruder barrel does not necessarily mean the extruder centerline is completely level. Bore-scoping verifies that the center support and end support for the barrel are properly aligned with the feed section and thrust bearing. Proper alignment of the barrel, feed throat, and thrust bearing allows the screw to slide easily in and out when the extruder is cold. If the extruder barrel must be heated to insert or remove the screw or to turn it easily, something is misaligned or the screw is bent. Running the extruder without proper alignment can result in serious damage. Bore-scoping can be done with lasers that are attached to the extruder, or an outside contractor can be brought in to your facility to bore-scope a new extruder installation.

Barrel wear is measured with a cylinder gauge that measures the barrel inside diameter (ID) versus the barrel length. Starrett and Sunnen produce two acceptable cylinder gauges.

High pressures in extruder barrels can be very dangerous. Consequently, a rupture disk is installed at the extruder head for safety purposes. In the event there is an increase in melt pressure in the barrel, the rupture disk breaks, relieving the pressure. Barrels are normally designed to withstand 10,000 psi pressure. Rupture disks are bought with specific pressure ratings, e.g., 7,500 psi, 8,000 psi, etc., that will fail below the 10,000 psi barrel pressure rating. The rupture disk screws into a standard

pressure transducer hole in the barrel flush with the inside barrel wall to assure no dead space is present for polymer to collect and degrade. Figure 3.12 shows a Fike rupture disk that screws into an extruder barrel.

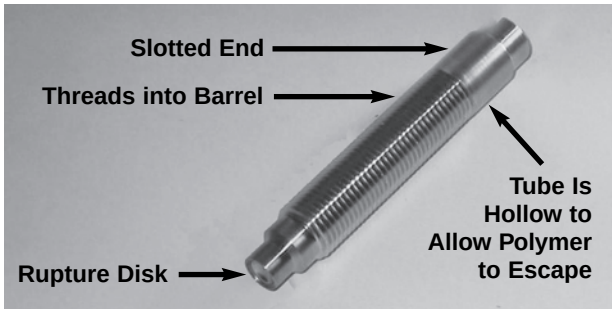


Figure 3.12. Fike rupture disk.

Three heater styles are available to heat the extruder barrel and adapters: mica, ceramic, and cast. Heaters must cover the maximum barrel area and be tightly clamped around the barrel to prevent hot spots and provide uniform heating. Large extruders normally have cast heaters, and smaller extruders use band heaters. Ceramic heaters are designed for higher temperature than mica, and both heaters have a wide operating temperature range.

Barrel cooling is accomplished with either water or air. Water is a better cooling medium with better heat transfer characteristics than air and provides better control. However, water costs more to install and requires a recirculating system or once-through water. Water lines can become dirty and clogged, solenoids must be maintained so they work properly, and a recirculating water system requires water treatment. Water has the advantage that it does not force hot air back into the room, heating the extrusion area. If a water-cooling system is properly maintained, it is very efficient and works well. Cooling systems are shown in Fig. 3.13 for both air and water. The ribbed spacers around the barrel in the air-cooled system provide additional surface area for heat removal and increased cooling efficiency. Air-cooled systems have a damper valve above the fan to adjust the airflow, providing maximum efficiency with different polymer processes.

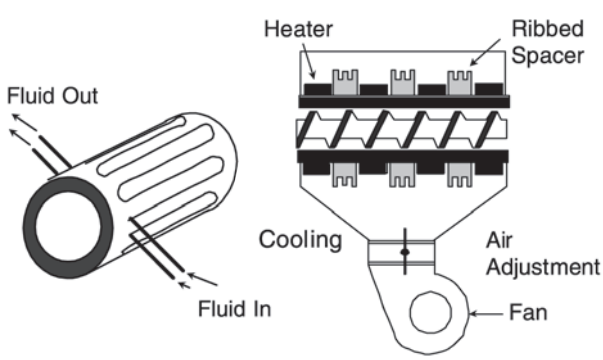


Figure 3.13. Water and air cooling on extruder barrel.

A single screw extruder screw typically has three different sections, as shown in Fig. 3.14. The feed section has deep flights to transport powder or pellets away from the feed throat. The transition section changes gradually from deep flights with unmelted pellets to shallow flights containing the melt. Resin is compressed in the transition

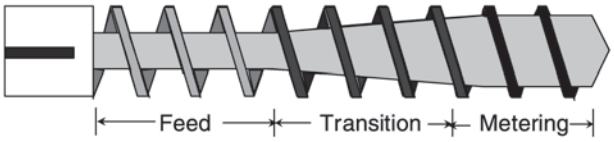


Figure 3.14. Extruder screw stages.

section during the melting process. Metering is the last screw section and has the shallowest flight depths. Screw nomenclature is defined below and shown in Fig. 3.15:

Channel depth: Distance from the top of the flight to the root

Channel: Space between flights

Trailing flight flank: Back edge of flight

Pushing flight flank: Front edge of flight

Pitch: Distance between consecutive flights

Helix angle: Angle flights make from a line perpendicular to the screw shaft

Screw diameter: Distance between furthest flights across the screw shaft

Keyway: End of screw containing the key that fits into the shaft surrounded by the thrust bearing

Root diameter: Distance from the channel bottom on one side to the channel bottom on the opposite side

Length: Distance from hopper to screw tip

L/D ratio: Screw length divided by diameter

Compression ratio: Ratio of the feed channel depth to the meter channel depth

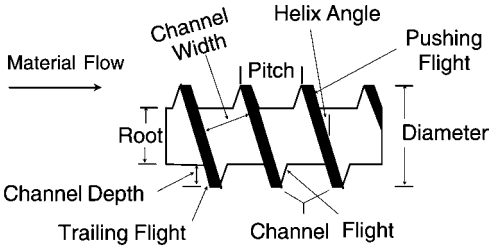
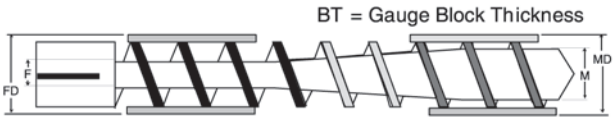


Figure 3.15. Definition of screw elements.

The screw compression ratio is critical in processing different polymeric materials. While it is desirable to have one general purpose screw that will process all materials efficiently at high rates, in practice this does not occur because different polymers have different viscoelastic properties. Some polymers run better on screws

with a 2.5:1 compression ratio, while other materials process better on screws with a 3.5:1 or 4:1 compression ratio. For this reason it is important to be able to measure the screw compression ratio and know which screw works best with different polymers. Figure 3.16 shows how to make screw measurements with gauge blocks and calculate the compression ratio. *F* is the root diameter in



Compression Ratio = Depth Feed Zone / Depth Metering Zone
Depth Feed Zone Channel = (FD – 2BT) – F
Depth Metering Zone Channel = (MD – 2BT) – M
Compression Ratio = [(FD – 2BT) – F] / [(MD – 2BT) – M]

Figure 3.16. Screw compression ratio calculation.

the feed zone, and *M* is the root diameter in the metering zone. *FD* is the screw outside diameter including the gauge blocks in the feed zone, and *MD* is the outside diameter including the gauge blocks in the metering zone. If the gauge blocks are the same size, *MD* and *FD* should be equivalent. Use the following information to calculate the compression ratio of a hypothetical screw:

- Root diameter in the feed section = 2.688"
- Gauge blocks thickness = 0.250"
- Distance between the outside of the gauge blocks in the feed section = 3.994"
- Root diameter in the metering section = 3.250"
- Distance between the outside of the gauge blocks in the metering section = 3.995"

The compression ratio of the screw is calculated as follows:

$$Feed\ Zone\ Channel\ Depth = 3.994 - 2 \times 0.250 - 2.688 = 0.806$$

$$Metering\ Zone\ Channel\ Depth = 3.994 - 2 \times 0.250 - 3.250 = 0.245$$

$$Compression\ Ratio\ (CR) = \frac{Feed\ Depth}{Meter\ Depth} = \frac{0.806}{0.245} = 3.29$$

$$Compression\ Ratio\ (CR) = 3.29:1$$

Gauge blocks are used to measure the outside screw diameter to determine screw wear. Gauge blocks are placed across the top and bottom of the flights (as shown in Figure 3.16) using calipers or a micrometer to measure the total distance. The gauge block thickness is subtracted from the measurement to obtain the outside screw diameter. Diameters of new screws should be measured and recorded prior to initial use to compare with the specifications and to use as a benchmark later when determining the screw wear.

An easier way to inspect a screw and get the data to calculate the screw profile and compression ratio is with a dial indicator mounted in a bar, as shown in Figure 3.17. The bar should be 2.2 times the length of the screw diameter. This allows zeroing the dial indicator by spanning three flights and zeroing the dial indicator on the middle flight.

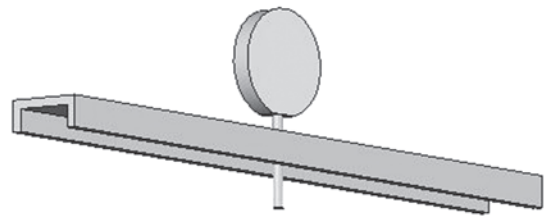


Figure 3.17. Screw channel depth measuring tool.

Clearance between the screw and barrel wall for small extruders is normally 0.001 inch (0.025 mm) times the screw diameter. For large extruders the normal clearance between screw and barrel wall is 0.004 inch (0.1 mm). These are screw clearance guidelines only; each individual extruder should be benchmarked prior to the installation of a new screw or after installation of a new barrel or liner.

A screw with a square pitch has a 17.66 degree helix angle; for each complete flight revolution around the screw, the pitch is 1*D*. Therefore, for a 4.5 inch extruder screw with a 17.66 degree helix angle, the pitch is 4.5 inches. With a square pitch, the number of flights is equal to the screw *L/D*, i.e., a 30:1 *L/D* has 30 flights or channels. (The actual number depends on whether the feed pocket is included in the *L/D*.) Flight width is normally 0.1*D*. This makes the flight strong enough to prevent chipping or breaking while still leaving sufficient room in the channel to process polymer. Unless the screw contains a barrier section, most screw designs have one parallel flight. Screws are usually between 20*D* and 30*D* long with four to eight flights in the feed section, six to 10 flights in the metering section, and the remaining flights in the transition section. Feed channel depth is normally 0.10*D*–0.30*D*, with the compression ratio between 2 and 4:1.

Early screw designs were determined empirically by trial and error. Today extruder screws are designed by computer programs based on polymer rheological data combined with the desired throughput rates, machine horsepower, mixing required for the application, and the polymer and additives being processed to generate the optimum screw design. If a current screw design is to be modified to a different design when purchasing a new screw, understand the objective or reasons for changing the screw design. What are the deficiencies in the current design? What is expected from the new screw design that can't be achieved with the existing screw? Some potential reasons for changing screw design are higher

throughput requirements; polymer supply has changed from pellets to powder; the need for more mixing to distribute or disperse additives, colorants, fillers, etc.; current screw generates too much or too little shear heat; the desire for better melt temperature control; running two different polymers that require two radically different screws and seeking a general purpose screw that runs both products reasonably well; the need to run a different polymer with substantially different rheology than current production; etc. Before making radical screw design changes, run trials with the new screw geometry in the vendor's facility to verify that it meets all processing criteria. Changing to a new screw design solely for the purpose of doing something different without a specific extrusion objective is not smart.

Screws can be cored for either a heating or cooling fluid during processing to add additional heat to the polymer or remove excess shear heat, respectively. Heat transfer fluid is pumped into or removed from the screw through a rotary union at the screw shank. Cooling is added to the feed zone to assist polymer feeding by preventing the screw or material from overheating, which may cause the polymer to stick to the screw root in the feed section (discussed in more detail in Chapter 4). Figure 3.18 shows a cored screw. Cooling required on both

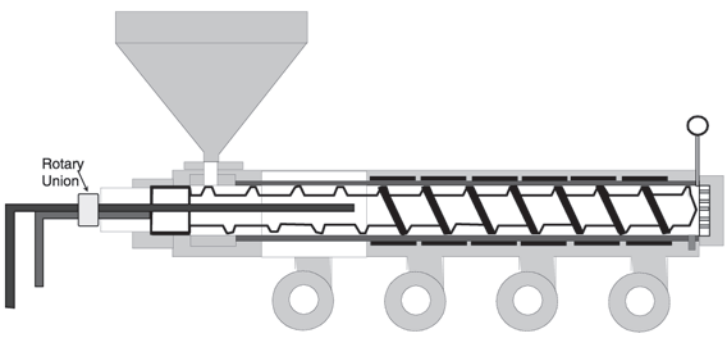


Figure 3.18. Screw cooling.

the barrel and the screw may indicate improper screw design and/or operating conditions. Heat removal from the extruder through screw cooling is not an energy-efficient operation.

Specifications required when purchasing a new screw include the following: polymer to be processed, throughput requirements, screw diameter, extruder manufacturer (needed to determine if *L/D* includes feed pocket and the keyway design), *L/D*, compression ratio, flight depth in feed or metering zone, barrier screw (discussed in Chapter 5), mixing elements (discussed in Chapter 4), number of flights in feed zone, number of flights in transition zone, number of flights in metering zone, helix angle, screw cooling, and number of screw stages (single or two-stage screw).

3.5 Die and Adapter

The extruder head assembly includes breaker plate, adapter to connect the die assembly to the extruder, and die. The breaker plate, shown in Fig. 3.19, is a round disk

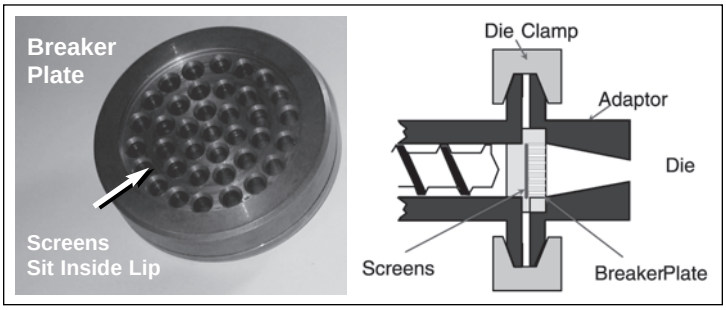


Figure 3.19. Breaker plate and seal at end of extruder.

containing many holes that is placed between the extruder and adapter. The breaker plate has the following functions:

- It stops the spiraling action of the polymer melt coming off the screw by forcing the polymer in straight lines as it passes through the breaker plate.
- It provides a seal between the extruder and the die/adapter. If the sealing surface is damaged or if the surface has been refinished a number of times to remove nicks and/or dents and the breaker plate is now too thin to provide a good seal, molten polymer will leak around the gate between the die/adapter and the extruder. High pressure is generated in this area, so the sealing surfaces must be smooth and pressure evenly applied around the seal to prevent polymer leakage.
- Screens in the breaker plate filter contamination from the polymer melt and create pressure at the extruder head. The screen and breaker plate combination assists in providing thermal homogeneity in the polymer melt. Screens clogged with contaminants cause high pressure at the extruder head and reduce the extruder throughput. If the formulation contains fillers or reinforcements, all screens must be removed.

Filtration is accomplished with wire mesh screens (square or twill weave), sintered powder, or metal fibers. Table 3.3 compares various filtration media and their characteristics. A square weave screen, Figure 3.20, has every

Table 3.3. Characteristics of Different Filtration Methods

Characteristics	Square Weave Screen	Dutch Twill Weave Screen	Sintered Powder	Metal Fiber
Gel Capture	Poor	Fair	Good	Very Good
Contaminate Capacity	Fair	Good	Fair	Very Good
Permeability	Very Good	Poor	Fair	Good

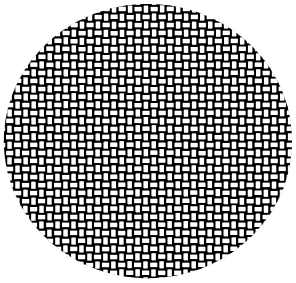


Figure 3.20. Square weave wire screen.

other wire over and under, while twill weave has every second wire over and under. Screen mesh measures the number of wires per inch; the higher the mesh, the more wires per inch, resulting in finer hole size and better contamination removal. Screen placement in a breaker plate starts with a coarse screen closest to the screw, followed by a finer and finer mesh, with the last screen in the group being a coarse mesh to act as a support for the fine mesh screens. The last coarse mesh prevents holes from being blown in the fine mesh screens in front of the breaker plate holes when high pressure is present. A typical 20/40/60/20 screen pack has the screens ordered from the extruder screw as 20 mesh, 40 mesh, 60 mesh, and finally a 20-mesh screen to support the 60 mesh.

Applications with high contamination requiring numerous screen changes may need an automatic screen changer to run economically. Many screen changer designs are available. Their sophistication depends on the application, the running time between screen contamination, and the expense associated with shutting down and restarting the process. Continuous-operation screen changers index as the screens become clogged, replacing the dirty screens with new screens, and never shutting down the process. Noncontinuous operation requires the process to be shut down, the screens changed, and the process restarted. Figure 3.21 shows a two-position

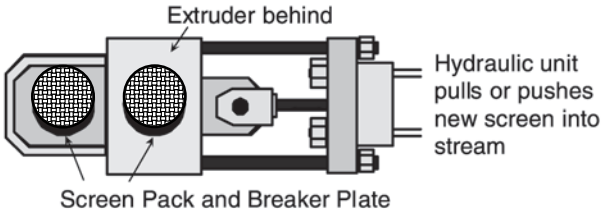


Figure 3.21. Hydraulic screen changer.

hydraulic screen changer with one screen in the polymer stream and the second clean screen waiting to be transferred into the polymer stream once the first screen becomes contaminated. Hydraulic screen changers may move slowly, requiring the process to be stopped, or very rapidly, where they can be changed on the fly with only minimal loss in product and time. Other screen changers available for continuous processes are rotary screen changers and double bolt screen changers. As screen packs become contaminated, the head pressure increases and the output decreases. If the pressure increases substantially during a run, either the extruder screw speed must be increased or the puller speed decreased to maintain product dimensions. Screen changers are discussed in detail in Part 5, "Auxiliary Equipment."

The die is attached either directly to the extruder or to a transfer pipe or adapter that is connected to the extruder. Polymer melt temperature in the adapter must be maintained. Transfer pipes, like extruder barrels, should have heater bands covering as much area as possible to minimize hot or cold spots. Remember, two extruder objectives are to provide uniform melt temperature and the proper melt temperature. Transfer pipes normally have smaller diameters than the extruder barrel, with a converging hole as material enters the transfer pipe from the extruder and a diverging hole as it exits the transfer pipe to the die. This helps maintain good polymer velocity through the transfer pipes and adapters.

Many die designs are available, depending on the extrusion process and the product being produced. Compounders use strand dies to make continuous strands that are chopped into pellets. Sheet and cast film producers use flat dies. A film or sheet die design is shown in Fig. 3.22. This produces a flat web of a specific thickness that passes through a three-roll stack or lays flat on a cast roll. Tubing and pipe dies are similar to that shown in Fig. 3.23, where extrudate exiting the die enters a vacuum sizing tank to set the product dimensions. Profile dies are all shapes and types and depend on the complexity of the product profile. Wire coating dies are similar in configuration to tubing dies, except a crossead

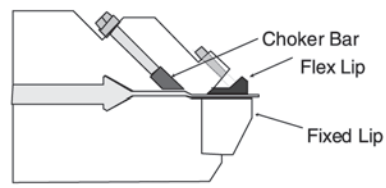


Figure 3.22. Sheet/film die.

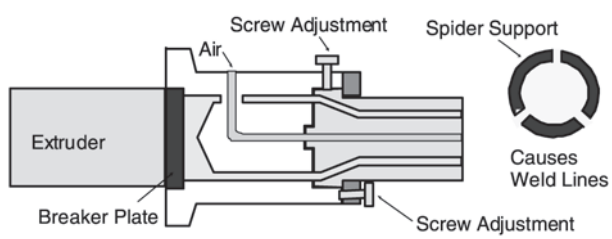


Figure 3.23. Pipe or tubing die.

die is used running perpendicular to the extruder. Dies are discussed in more detail in Part 7.

Extruder melt pressure is the pressure generated by the extruder screw pumping melt through the breaker plate, transition or adapter, and die. Small die openings combined with high throughput rates generate very high die pressure, resulting in high extruder backpressure, which reduces the throughput rates. Extruder throughput is the drag flow rate created by the screw rotation

minus the pressure flow generated by the die and/or screen pack resistance. If the die is removed from the extruder, backward pressure flow is absent and maximum output is attained. With the breaker plate, screens, and die attached to the extruder, the material forced through the die openings combined with the viscoelastic properties of the resin at a particular melt temperature determine the pressure generated. The polymer velocity profile in the metering section is depicted in Fig. 3.24. Polymer flow in the metering zone is accomplished by drag flow. Drag flow has high velocity along the barrel wall as material is scraped off the barrel with

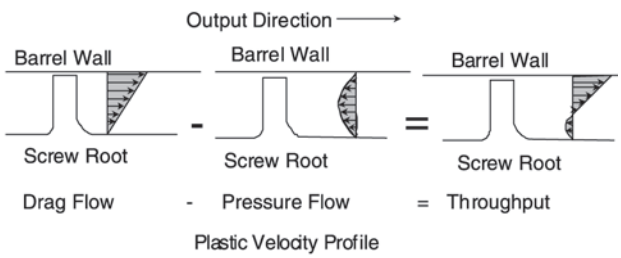


Figure 3.24. Plastic throughput profiles.

the screw flight, and zero velocity at the screw root. Pressure flow is the result of die pressure or head pressure forcing the material backward into the metering zones. The velocity profile is similar to a plug flow velocity profile with high velocity in the center and zero velocity at the barrel wall and screw interface. Combining these two velocity profiles forms the throughput velocity profile in the metering section. The throughput velocity profile shows high polymer velocity at the barrel wall, zero velocity at approximately two-thirds the depth of the channel (from the barrel wall), and a negative velocity near the screw root. Extruder throughput is given by Eq. (3.3):

$$\begin{aligned} &Plastic\ Output\ (Q) = \\ &Drag\ Flow\ (Q_D) - Pressure\ Flow\ (Q_P) - Leakage\ Flow\ (Q_{LF}) \end{aligned} \tag{3.3}$$

Leakage flow is the flow over the top of the extruder flights. Leakage flow is normally negligible and disregarded except for a worn screw, where leakage flow can be significant.

3.6 Controls

Without good process controls plus an understanding of the extrusion process, the extruder becomes nothing more than a black box where pellets are put in one end and an extruded shape exits the other. What occurs in the extruder? How do you know the system is in control? How do you know the extruder is running properly? The answer to these questions is process control, with both input and feedback loops to verify that the process is operating properly and at equilibrium.

Control is making a measurement, determining if something needs to be changed, making a decision, and taking action. If the system is operating at equilibrium, the decision is that everything is running properly and no changes are necessary. If the product is borderline acceptable or unacceptable, or a process step is outside the SPC control chart limits, the decision is that something is wrong and a change is required to get the system back into control.

With improved computer capability and durability, extruder controls have become more sophisticated, resulting in better process control. The entire process can be controlled from a central location, ranging from the individual feeder throughput to the puller speed and everything in between. Feedback loops are available to plot SPC data every second (or your chosen interval) for all temperature controls, screw speed, melt temperature, melt pressure, motor load, puller speed, on-line gauges for thickness or dimensions, feed rates, water temperature in cooling tanks, vacuum levels, windup speeds, roll pressures, and so forth. With the entire process instrumented, it can be monitored from a remote site. Engineers sitting in an office can monitor each plant line, verifying that all lines are running properly and the processes are in control based on SPC data.

In comparison with other plastic processes, an extruder has very few independent control variables that can be changed by an operator to alter the process. Assuming the correct screw is in the extruder, the proper die is installed, the screen pack is clean, and the equipment is operating properly (all heater bands and thermocouples are functioning properly, air or water cooling on each heating zone is working, and cooling on the feed throat is operating properly), the only extruder variables that can be changed are the temperature setpoints and screw rpm. Several parameters are monitored to assure the process is in control and running properly: barrel temperatures (actual and setpoint), extruder load (percent load, torque, or amps), screw speed, melt temperature, melt pressures (prior to the screen pack, in the die, and in a two-stage extruder at the first stage metering section), feed throat cooling water, temperature of raw materials entering the extruder, vacuum level if vacuum venting is being used, and cooling on the individual barrel zones. Other parameters to be monitored prior to the extruder include blend ratio, feed rates, moisture content of hygroscopic and moisture-sensitive materials, liquid feed rates, and raw material lot numbers. Downstream equipment parameters to be set and monitored include

- Roll and/or puller speeds
- Roll gaps
- Cooling bath temperature

- Vacuum level
- Windup speeds

With all this information, how do you know if the product is good? If process changes are required, how do you make changes? Even with all this information available, in some instances the information is not recorded, raising the question, how is troubleshooting done?

The first step in setting up the extruder and other process equipment to produce a different product is to obtain the standard operating procedures (SOP) for that product. This is a controlled document specifying all operating conditions and settings prior to start-up. Process parameters are normally controlled over a small range. If the final product does not meet specifications at start-up, the SOP for the process should be compared with the setup to verify that all equipment is set properly and operating within the ranges specified by the SOP. If the raw materials are the same as last time and all process conditions are the same, this raises the question, what has changed since the last run to produce an unacceptable product? Investigation to determine what has changed requires a systematic approach. Do not start trial and error experimentation, making random changes without an appropriate action plan to fix the problem.

Good extruder operation includes recording all operating conditions when the process is running optimally and also when problems exist. The tendency is to monitor the process and record operating conditions only when it is not running smoothly and there are problems. When problems occur, something has changed. If baseline data under normal product operating conditions are not available, it is difficult to determine what has changed. It is impossible to record too much data and have records that are too detailed. However, it is possible to record too little information and not have the necessary data when problems arise to do proper troubleshooting to identify the root cause.

3.6.1 Temperature Zone Control

Each extruder temperature zone has at least one heater and possibly multiple heaters controlled by a thermocouple. A signal from the thermocouple communicates with the controller, indicating whether the heater is to be turned on or off. For the controller and heaters to function properly, the thermocouple must operate properly. A faulty thermocouple with an open circuit indicates the temperature is low, resulting in the heaters staying on and causing substantial overheating. A closed thermocouple indicates the temperature is high; heaters remain off and the temperature zone cools. If a thermocouple is not responding properly, it must be replaced. (More information is presented on thermocouples later in this section.) The thermocouple well in the barrel should be at least 1.2 inches (30 mm) deep and installed away from the heaters. Never sandwich the thermocouple between the heater and the barrel wall; the thermocouple will be responsive to the heater temperature and not the barrel temperature. Figure 3.25 shows the thermocouple location relative to the heaters. Thermocouples must be plugged into the correct temperature controller; otherwise the thermocouple will control the wrong heaters. For example, if the thermocouple from zone 2 is plugged

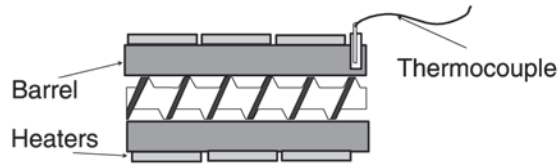


Figure 3.25. Heater and thermocouple assembly.

into the controller for zone 3 while the zone 3 thermocouple is plugged into zone 2, and the temperature for zone 2 is set at 420°F (216°C) and zone 3 is set at 450°F (232°C), the actual extruder temperatures in zone 2 will be 450°F (232°C) and in zone 3 will be 420°F (216°C). The melt temperature thermocouple works on the same principle as the zone temperature thermocouples, except the melt thermocouples protrude into the melt stream to obtain an accurate melt temperature. (This is discussed in more detail later.)

Barrel thermocouples should be spring loaded to guarantee good contact with the barrel wall and insulated to minimize heat loss along the thermocouple stem. Figure 3.26 shows the difference between a conventional and an insulated thermocouple as a function of thermo-

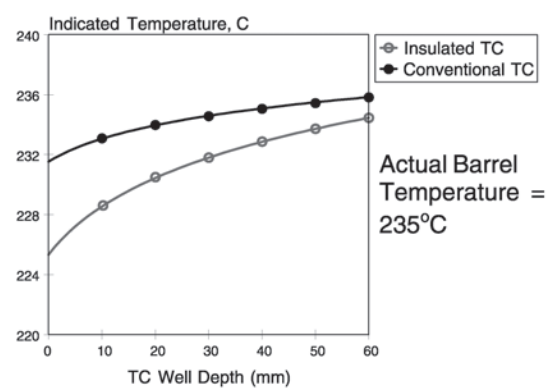


Figure 3.26. Thermocouple well depth vs. temperature.

couple well depth in the barrel wall. As the thermocouple well depth increases, the difference between actual barrel temperature and the measured temperature decreases.

Each extruder temperature zone should contain a heater ammeter readout. It may be on the control panel or in a computer-driven control system. With all heaters functioning properly, the amps drawn in a particular zone when the heaters are on 100% power is a fixed value.

Assuming a zone has more than one heater and the ammeter reading is lower than normal, at least one heater is burnt out or not working properly. Nonfunctioning heaters need to be replaced as soon as possible to prevent hot or cold spots along the extruder barrel, resulting in nonuniform polymer heating. Ammeter readings need to be checked daily to verify that all heaters are functioning properly.

If the actual and set temperatures in a particular zone are very different, this is the extruder's method of communicating with the operator to indicate the following possible scenarios:

- Temperature is not controlling in that particular zone
- Thermocouple is not operating properly and may need replacing
- Temperature setting for the material being processed is wrong
- Excess shear heat is being generated in that zone
- Extruder cooling is not functioning properly
- Thermocouple wire from the extruder is not connected to the correct controller
- Polymer is being overfed to that section of the screw

Temperature controllers have evolved over the years with improvements in electronics to provide very accurate control (within a couple of degrees) on all zones. Temperature controllers in general are either direct or reverse acting. Extruder temperature controllers all are reverse acting, meaning the output is reduced as the heat increases. With heat/cool applications, the heat output opposes the temperature output. This is shown graphically in Fig. 3.27.

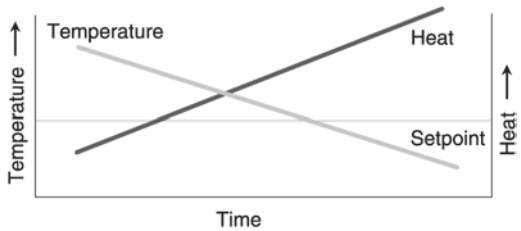


Figure 3.27. Reverse acting controller.

The first temperature controllers were on/off closed loop or feedback controllers. If the temperature was below the setpoint, the controller called for 100% power, and above the setpoint the power was off. This resulted in significant temperature over- and underride ranges attributed to the thermal lag caused by the massive barrel wall, polymer, and screw. (Thermal lag is the time delay caused by putting high temperature or heat on the barrel outside surface, and the time it takes to heat the thermocouple. The barrel temperature may be significantly hotter than the inside barrel surface. When the thermocouple communicates with the controller to shut the power off, outside surface temperature already exceeds the setpoint, resulting in the barrel temperature exceeding the setpoint. The same phenomenon occurs on cooling.) For large extruders the thermal lag could be several minutes, resulting in overheating during heat-up and overcooling when cooling down. A typical temperature/power curve versus time is shown in Fig. 3.28. Shallow-well thermocouples reduce the temperature lag, but the polymer melt temperature has less influence on the barrel temperature with shallow-well thermocouples.

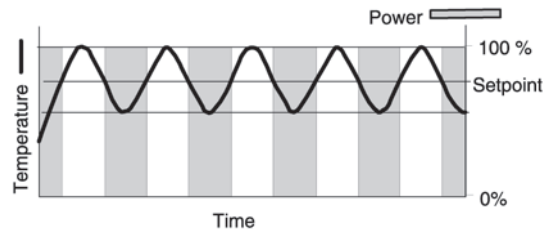


Figure 3.28. Time-temperature power curve.

To improve on/off controllers, hysteresis is added to reduce thermal lag. Hysteresis requires the temperature on cooling to exceed the setpoint by a fixed amount before the power comes back on. The hysteresis band can be set to improve controller accuracy. Figures 3.29 and 3.30 show the temperature versus power curves with hysteresis added.

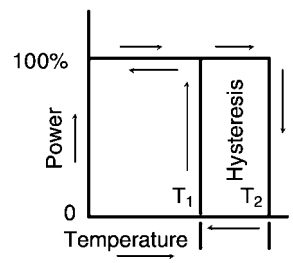


Figure 3.29. Power curve with hysteresis.

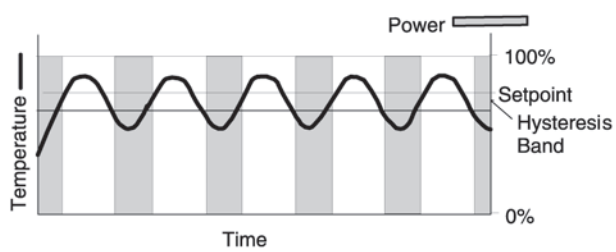


Figure 3.30. Addition of hysteresis to controller.

The next improvement in controllers came with proportional controllers, where the power input is reduced from 100% as the temperature approaches the setpoint. The proportional band, given in percentage about the setpoint, is the temperature range where the power is on some proportional amount between 0 and 100%. As an example, a 5% proportional band with a setpoint of 500°F (260°C) is ±25°F (13°C). The electrical power input curve versus temperature for a proportional con-

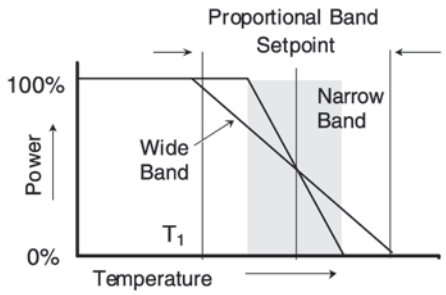


Figure 3.31. Power curve with a proportional controller.

troller is shown in Fig. 3.31. The gain is the percent of the span given by Eq. (3.4):

$$Gain = \frac{100\%}{Proportional\ Band\ in\ Percent\ Span} \tag{3.4}$$

Smaller proportional bands have higher gains. In operation, the temperature cycles through increases and decreases until an equilibrium level is reached. The difference between the setpoint and the equilibrium level is called the offset or droop. Narrowing the proportional band decreases the droop. However, the proportional band can only be decreased to a certain level before control instability becomes an issue. Figure 3.32 plots time versus temperature for a proportional controller.

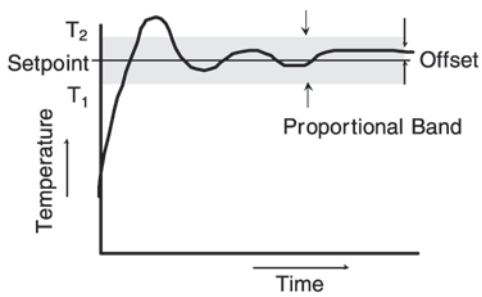


Figure 3.32. Temperature vs. time for a proportional controller.

In practice, the extruder power requirements are based on the heat needed to process the polymer plus the heat lost by the system through conduction, convection, and radiation. Insulating an extruder and die reduces the heat loss and power requirements. The heat loss or power loss curve intersects the proportional band (Fig. 3.33). If the intersection of the proportional band curve is below the setpoint, there is a negative offset, while intersection above the setpoint is a positive offset. Automatically resetting the proportional band so the power loss curve intersects the proportional band at the setpoint is done by integration, providing long-term controller accuracy with changing external and internal conditions. This is a PI controller, standing for Proportional + Integral. The integral time constant is the time required for the proportional band to move. Movement must occur relatively slowly to assure oscillations above and below the setpoint are not introduced. Figure 3.34 shows the integral action on the power versus temperature curve. The integrator will continue to adjust

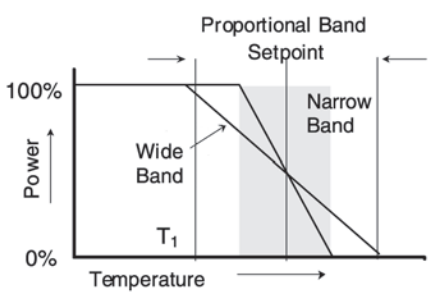


Figure 3.33. Power loss curve combined with proportional band.

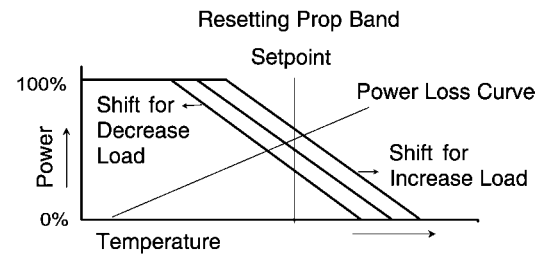


Figure 3.34. Integral action on power loss and temperature curves.

the proportional band until the deviation from setpoint to the power loss curve is zero. Figure 3.35 shows the time versus temperature curve for a PI controller. Temperature exceeds the setpoint on start-up, because the automatic reset begins at the proportional band lower limit. Integral

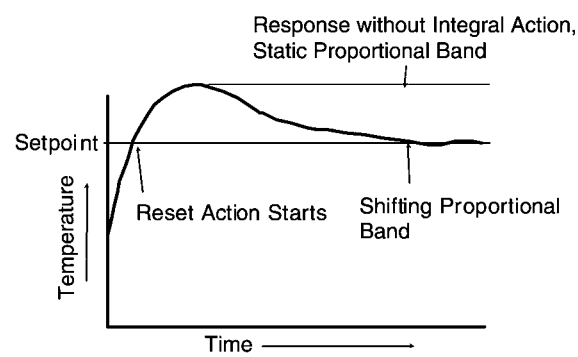


Figure 3.35. Time-temperature curve for a PI controller.

action resets the proportional band until the temperature is at the setpoint. PI controllers may respond improperly if the integrator acts on an error signal when the temperature is outside the proportional limit.

PID controllers combine Proportional + Integral + Derivative control. Derivative control adds more stability to controllers by compensating for rapidly changing conditions. As mentioned above, PI controllers are slow to respond. Adding the second corrective term to proportional controllers results in a faster response time. When the temperature exceeds the setpoint, the derivative signal changes the sign and the integrator brings the proportional

band back into position. The derivative term allows the proportional band to be narrowed while preventing oscillations. Determining PID parameters for a particular control system is called tuning. Three methods are available for tuning PID controllers. The first method is to take the PID parameters off an existing controller when replacing controllers. The second approach is to use "Self Tuning," which is a software package to automatically determine the PID and other control parameters. The third tuning method is called "adaptive tuning," which uses a program to monitor the current PID parameter performance and calculates new parameters, if necessary. More information is available on temperature controllers and tuning from controller manufacturers, e.g., Eurotherm,[1] Barber Coleman, Omega, Chromalox, and others.

3.6.2 Melt Temperature

Polymer melt temperature control is critical to the control and reproducibility of the extrusion process. Melt thermocouples must protrude into the melt stream to obtain accurate readings. Melt temperature measurement inside the extruder barrel is not practical because the turning screw would shear off a melt probe sticking down in the melt stream. Melt temperature measurements have been made in barrels with infrared probes that are either part of a pressure transducer or are sized to fit into a barrel pressure transducer hole. The accuracy of infrared measurements depends on the temperature being recorded and the source emissivity. Arguments are made that infrared radiation penetrates the polymer and measures the melt temperature at some specified depth. This depth, relative to the material present, is somewhat questionable.

Melt temperature measurements in an adapter, transition pipe, or die depend on the thermocouple penetration into the melt stream. Plug flow in the adapter or transition pipe results in polymer velocity profiles having higher flow in the middle of the melt stream than along the walls. As a result, the melt temperature is normally hotter in the center of the melt stream as the polymer temperature near the wall approaches the wall temperature. Heat conduction along the thermocouple can also lead to erroneous melt temperature readings. Another factor to consider is the energy dissipation at the probe due to polymer shear heating resulting from the polymer flowing around the probe.

The different melt temperature probes available are shown in Fig. 3.36. The flush mount thermocouple, A, does not disrupt the resin flow in the channel; however, the temperature is more apt to resemble the metal temperature than the melt temperature. When using a flush mount probe, the melt temperature is similar to the adapter, transition, or die temperature, even when these temperatures are significantly higher or lower than the extruder barrel temperature. It is likely the temperature measured is the metal, and not melt. Polymer temperature at the wall always matches the metal temperature. Thermocouples B and C in Fig. 3.36 are straight probes protruding into the melt stream. These are simple, sturdy thermocouples that provide a good melt temperature in the channel. Measurement error comes from shear heating caused by the polymer flowing around the thermocouple and conduction along the probe. Thermocouple D in Fig. 3.36 is radially adjustable in the melt stream. This adjustable thermocouple provides a way to measure the melt uniformity across the melt channel, where temperatures may vary up to 20°F (11°C). When comparing processing data from run to run with radially adjustable thermocouples, the thermocouple probe location in the melt stream must be constant. During process start-up these probes are retracted from the melt stream to prevent breakage. Currently, radially adjustable probes are more common in R&D labs and on process development lines than in production environments. Finally, thermocouple bridges, E in Fig. 3.36, allow simultaneous measurement at different locations in the melt stream, providing an instantaneous cross-channel temperature profile.

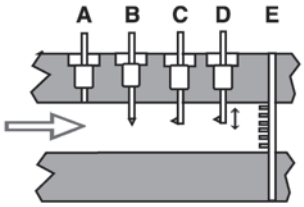


Figure 3.36. Melt thermocouples.

Melt temperature is critical to process control and is a parameter that has to be carefully monitored when setting up, evaluating, or troubleshooting a process. Polymer viscosity and flow are directly related to the melt temperature. Increases in melt temperature result in lower viscosity, leading to lower resistance to flow in the melt pipe and die as well as higher back pressure flow in the extruder. This combination usually results in lower output. Decreases in melt temperature lead to higher viscosity and higher resistance to flow in the melt pipe and die as well as lower back pressure flow in the extruder. This combination usually results in higher output. Figure 3.37 shows a K-type melt thermocouple that protrudes into the melt stream.

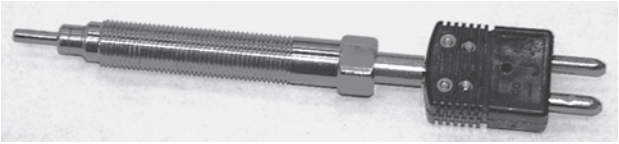


Figure 3.37. K-type melt thermocouple.

Thermocouples are made from two dissimilar metal wires connected at both ends, with one end being at a different temperature than the other end. An electromotive force (thermal EMF) is generated, resulting in an electric current flowing in the closed loop. With the current flow

being predictable and one connection or junction being kept at a known temperature, it is possible to determine the other temperature. The wire length or the wire size does not affect temperature measurement. Any two dissimilar wires can be used to make a thermocouple; however, to obtain reproducible results, the same wire combination must be used each time. A J-type thermocouple is shown in Fig. 3.38. Thermocouple technology is possible because

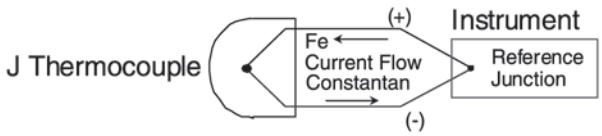


Figure 3.38. Standard thermocouple.

of the Laws of Homogeneous Circuits and Intermediate Metals. The Law of Homogeneous Circuits states that homogeneous thermocouple conductors are not affected by intermediate temperatures or temperature differentials along the wires. This allows long thermocouple leads to be used in all temperature environments. The Law of Intermediate Metals states a third metal can be introduced into the circuit without affecting the temperature, providing the junctions with the third metal are at the same temperature. This allows the instrument making the measurement to be a different metal than the thermocouple wires.

Over time, certain metals have been selected for producing standard thermocouples; some are shown in Table 3.4. The most common type of thermocouple used on extruders is the K thermocouple. Hand-held pyrometers used in plastic processing are typically either J or K thermocouples. Each thermocouple type has a unique connector, color, and thermocouple wire. When ordering thermocouple wire, verify that the correct wire and size is purchased for the thermocouple type and application. It is important to match the thermocouple to the application.

Three thermocouple junctions, shown in Fig. 3.39, are commercially available, depending on the application. Grounded junctions are used where electromagnetic induction or radio frequency interference is present and can interfere with the thermocouple signal or measurement. These

Table 3.4. Some Types of Thermocouples Available

Type/Color	Conductors	1/4" Diam. Probe 18 AWG Wire	3/16" Diam. Probe 24 AWG Wire	1/8" Diam. Probe 24 AWG Wire	1/16" Diam. Probe 30 AWF Wire
J/Black	Fe/CuNi	482° C (900° F)	482° C (900° F)	371° C (700° F)	315° C (600° F)
K/Yellow	NiCr/NiAl	982° C (1800° F)	982° C (1800° F)	871° C (1600° F)	760° C (1400° F)
T/Blue	Cu/CuNi	287° C (550° F)	260° C (500° F)	204° C (400° F)	148° C (300° F)
E/Purple	NiCr/CuNi	537° C (1000° F)	537° C (1000° F)	426° C (800° F)	371° C (700° F)

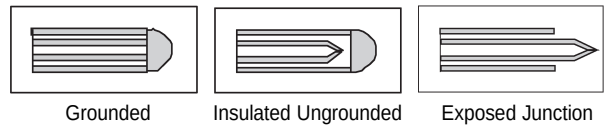


Figure 3.39. Thermocouple junctions.

have faster response times than insulated junctions. A second junction is the insulated ungrounded model with a slower response time due to the insulation. These are typically used in extruders and hand-held pyrometers. Exposed junction thermocouples have the fastest response time; however, with the exposed junction they are easily damaged and susceptible to oxidation and temperature deterioration.

Thermocouple response time depends on the sheath diameter and the junction type. Smaller diameters have a faster response. The advantages of thermocouples are their ruggedness, inexpensive nature, high responsiveness, broad temperature range, and tip sensitivity. The disadvantages are that their outputs are not linear and the accuracy depends on the purity of the metal wires used. The output nonlinearity is not a serious limitation, as the measurement equipment normally takes this into account when correcting the response to yield the proper reading.

The other temperature measurement device is called aresistance temperature detector (RTD). RTD operation is based on electrical resistance increasing or decreasing with temperature in a predictable manner. A probe, shown in Fig. 3.40, is an assembly containing an element, sheath, lead wires, and connections. The element that measures the temperature is platinum, copper, or nickel wire wound around a ceramic or glass core. To specify an RTD it is necessary to know the element length, element diameter, and resistance in ohms at 0°C. Advantages of RTDs versus thermocouples are the response is more linear and the signal stability is better. The main disadvantage is slower response time compared to a thermocouple.

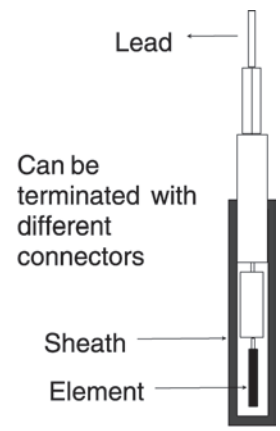


Figure 3.40. RTD probe.

3.6.3 Pressure Measurement

Melt pressure measurements at the extruder head are as critical as melt temperature. Constant polymer melt temperature and pressure at the die produce consistent output, resulting in uniform product cross sectional

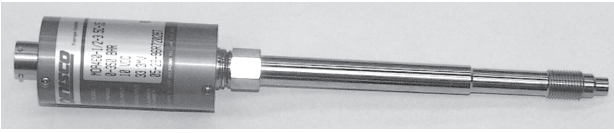


Figure 3.41. Pressure transducer.

dimension, assuming the extruder and puller speeds are properly set. Like melt temperature, melt pressure tells the operator what is going on inside the extruder and die. Die pressure fluctuations correspond with output fluctuations and dimensional changes. A Dynisco® pressure transducer is shown in Fig. 3.41.

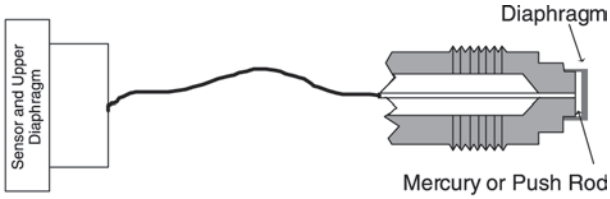


Figure 3.42. Pressure transducer schematic.

A pressure transducer works on the same principles as a Wheatstone bridge. Pressure in the melt causes the transducer diaphragm to bend. The stress or strain induced in the diaphragm is transferred to a resistor. As pressure increases or decreases, the resistance changes, resulting in different pressure readings. Figure 3.42 shows a melt pressure transducer cross section.

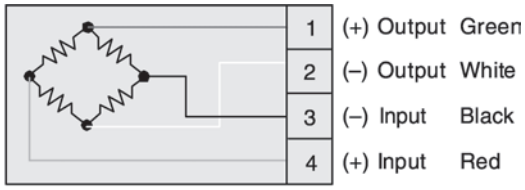


Figure 3.43. Schematic of pin configuration.

A schematic for the pin configuration in the sensor head is shown in Fig. 3.43.

Pressure measurements are taken before the breaker plate and in the die. If the die pressure is constant, the output will be constant. Increased pressure before the breaker plate due to screen pack contamination or blockage causes pressure decreases in the die, as less material is being pumped from the extruder into the die, resulting in a reduction in the product cross sectional area. To compensate for reduction in die pressure, the screw speed is increased. While this increases the pressure prior to the breaker plate, it allows a constant output until a clean screen pack can be installed. In a two-stage extruder, pressure is normally measured at the first-stage metering section in addition to the extruder head and die.

A pressure transducer located at the end of a 1-inch single screw extruder with the die removed is shown in Fig. 3.44. Note the cooling fans on the barrel for air cooling and the temperature thermocouple near the end of the extruder controlling the last temperature zone.

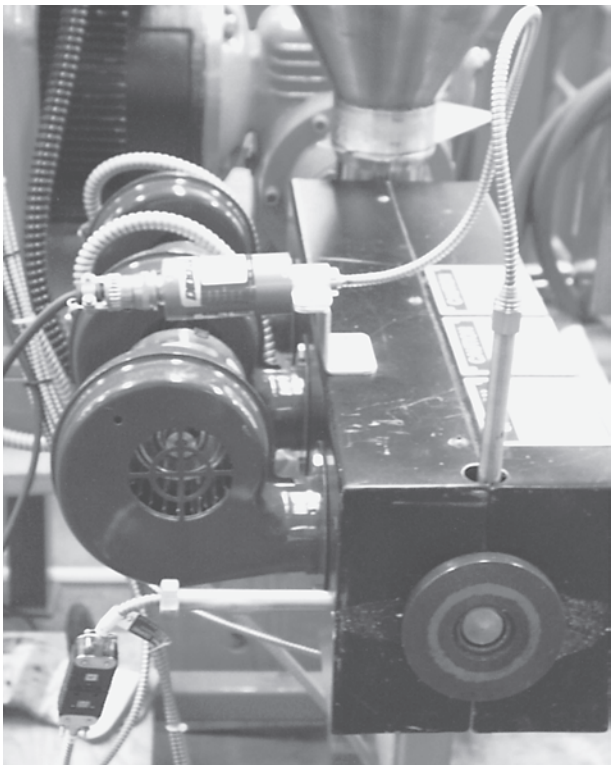


Figure 3.44. Pressure transducer at end of extruder.

Typical melt pressure curves recorded at the end of the barrel, the adapter, and the die are shown in Fig. 3.45. At the end of the screw, the pressure variation is quite significant due to the extrudate spiraling action coming off the screw. Pressure fluctuations are caused by the screw flight and the pressure gradient in the screw channel. Low pressure occurs where there is a break in the melt stream due to the screw flight. High pressure occurs from the material at the edge of the pushing flight. For this reason, head pressure measurement should not be taken where the flight will pass over the pressure transducer. A good place to measure the head pressure is between the last flight and the screen pack. The screens and breaker plate reorient the extrudate flow, reducing the differences between the peaks and

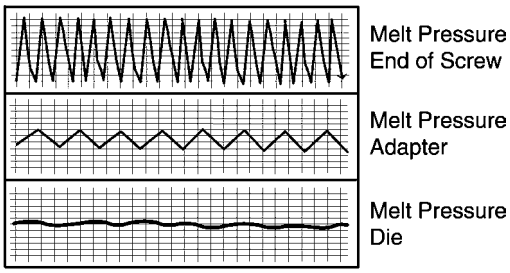


Figure 3.45. Melt pressure.

valleys observed at the end of the extruder. As the extrudate flows through the die, the flow channel capacitance further dampens the high-frequency pressure fluctuations. If the pressure variations in the die were the same as at the end of the extruder barrel, the product cross sectional dimensions would be constantly changing, and the final product tolerances would be very broad.

Output as a function of the low-frequency pressure variation measured in the melt pipe or die is given by Eq. (3.5):

$$\Delta Q(\%) = \frac{\Delta P(\%)}{n} \tag{3.5}$$

where

ΔQ = Change in volumetric output (percent)

ΔP = Change in pressure (percent)

n = Power law index

Assume the polymer being processed is polypropylene with a power law index of 0.35 (n = 0.35) and the die pressure is varying from 1,950 to 2,050 psi for a ΔP of 100 psi.

$$\Delta P(\%) = \frac{100}{2000} \times 100 = 5\%$$

$$\Delta Q(\%) = \frac{5}{0.35} = 14.3\%$$

Assuming the process is running at 200 pounds/hour, the variation in output is 29 pounds/hour. This flow variation would have a major effect on the final product dimension. Head pressure variation should be kept as low as possible, preferably 1% or less.

3.6.4 On-Line Measurements

Process instrumentation helps remove the black box around the extruder and gain insight into what is happening to the product and process in real time. In addition to making process troubleshooting much easier, on-line process instrumentation provides a tool to improve product quality and plant yields. Polymer and product properties that can be measured on-line include rheology, additive levels, polymer ratios, and product dimensions. Producing a more consistent product, verifying that the product meets customer specifications, and determining the correct additive level in each formulation improve product quality. Plant yields are improved by verifying that the product being run has the same properties from lot to lot, identifying product deficiencies in real time, and providing data to make troubleshooting easier.

3.6.4.1 Rheology

Polymer rheology (discussed more in Chapter 4) is critical in producing identical products from run to run. Remember, the extruder objectives to accomplish are constant melt temperature, uniform melt temperature, constant melt pressure, uniform melt pressure, and homogeneous product; these factors all relate to the polymer rheology. Unexpected increases in viscosity lead to lower throughput, resulting in product dimensional changes, higher shear heating causing possible resin degradation, and higher motor loads. Unexpected decreases in viscosity lead to higher throughputs causing dimensional changes, lower shear heating, and different melting and metering characteristics. Changes in viscosity lead to unacceptable finished product dimensions unless some corrective action is taken. Other than equipment problems, viscosity variations are an issue associated with running different regrinds at various levels, blends of two or more different melt flow index (MFI) resins, or running coextrusions.

Viscosity information available prior to a production run normally includes MFI data from vendor certification or in-house measurements and/or shear rate versus viscosity data supplied by the vendor or measured in-house. Real time viscosity is measured on-line with a small side stream extruder. Real time measurements allow early detection of viscosity changes during the production run. Correcting the melt viscosity can be done by increasing or

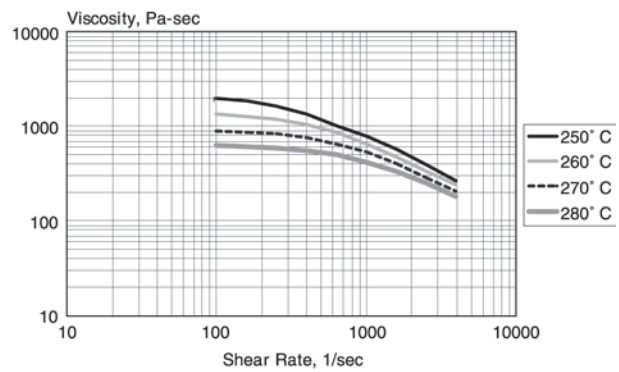


Figure 3.46. Shear rate vs. viscosity data.

decreasing the polymer melt temperature. On-line viscosity measurements allow the distinction between a real time problem and normal process variations.

Typical shear rate versus viscosity curves at four different temperatures for a resin with temperature-sensitive viscosity are given in Fig. 3.46. (A temperature-sensitive viscosity resin is one whose viscosity changes more with temperature than with shear rate, while a shear-sensitive resin is one whose viscosity changes more with shear rate than with temperature.) Extrusion shear rates are normally between 50 and 1000 sec^{-1}. An Arrhenius plot to determine how the viscosity changes with temperature is obtained by plotting the viscosity at a given shear rate versus $1/T$ (T = temperature in Kelvin). Using the rheology data in Fig. 3.46 at 400 sec^{-1}, an Arrhenius plot

yielding a straight line relationship is generated in Fig. 3.47. The Arrhenius plot allows for extrapolating the viscosity to other temperatures. (Arrhenius plots provide data to calculate the process temperature in coextrusion where the resins being coextruded will have the same viscosity.) As an example (using the data in Fig. 3.47),

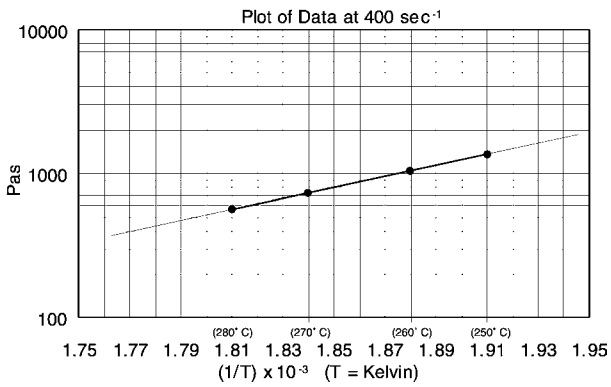


Figure 3.47. Arrhenius plot of temperature vs. viscosity.

assume under normal operating conditions the melt temperature is 500°F (260°C) and the viscosity is 1039 Pa·s at 400 sec^{-1}. Extruder output and all downstream equipment is set up based on an extruder throughput having a viscosity of 1039 Pa·s. Using the same setup with regrind added to the formulation, the output viscosity is only 800 Pa·s. What melt temperature is necessary to get the process to run as though the viscosity was 1039 Pa·s? Using a second set of viscosity versus shear rate curves based on the formulation with regrind, plot a new Arrhenius graph (Fig. 3.48) at 400 sec^{-1} and extrapolate to determine the melt temperature where the viscosity is 1039 Pa·s. The desired melt temperature to run the process under the same conditions used previously is $1/T$ = 0.00190 or 487°F (253°C). Using an in-line rheometer, the viscosity can be constantly monitored and temperatures adjusted until the correct viscosity is attained.

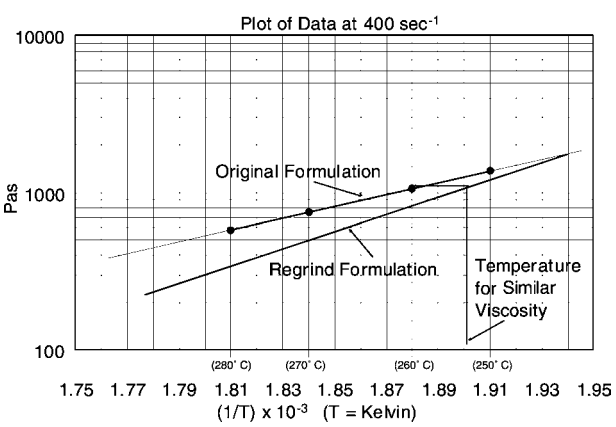


Figure 3.48. Arrhenius plot with regrind formulation and original formulation.

Most in-line rheometers operate similar to MFI testers with 1-point data. However, some in-line rheometers are available that generate shear rate versus viscosity curves. Besides providing information for process control, these rheometers give information on polymer molecular weight and molecular weight distribution. Increases or decreases in viscosity in the low-shear region correlate with molecular weight, while viscosity changes in the high-shear region relate to molecular weight distribution or branching.

3.6.4.2 Spectroscopy

Composition and additive levels in a formulation can be measured with Fourier Transform Infrared (FTIR). Molten polymer is diverted through an infrared cell, providing real time measurements. In-line measurements eliminate the problems associated with sampling methodology, sample preparation, and time delays. In addition, the infrared (IR) crystalline absorption bands are not present as the polymer is molten and amorphous. In-line systems must be able to handle high temperatures and high melt viscosities, and be able to perform in a manufacturing environment with delicate equipment. Kayeness makes a system called IROS-100 that uses a unique polymer FTIR flow cell with an adjustable path length and diamond windows; it withstands temperatures to 752°F (400°C) and pressures to 4,000 psi. IR applications in extrusion include identification of additives and their concentration, determination of the degree of poly-

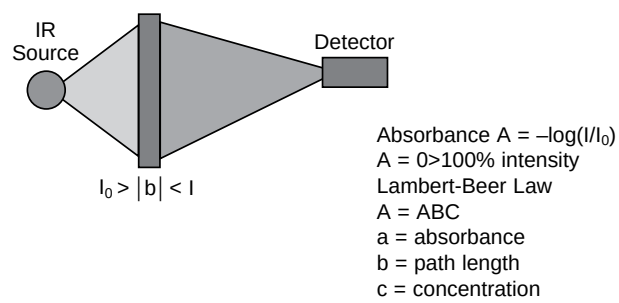


Figure 3.49. FTIR cell, IR source, and detector.

merization (end-group analysis), polymer concentrations in blends, thermal degradation, and cross-linking. Other advantages of this technology include verifying that sufficient concentration of costly additives is present in a formulation without excessive quantities, determining when a product change is complete, validating polymer ratios, etc., in real time. The theory behind the calculations is based on any typical IR technology; see Fig. 3.49.

As an example,[2] consider an extruder product conversion from high density polyethylene (HDPE) to thermoplastic elastomer (TPE). Figure 3.50 shows the IR scans over time and a plot of the absorbance at 3480 cm^{-1} versus time. The absorbance peak at 3480 cm^{-1} is attributable to TPE and is not present in the HDPE spectrum. As the TPE

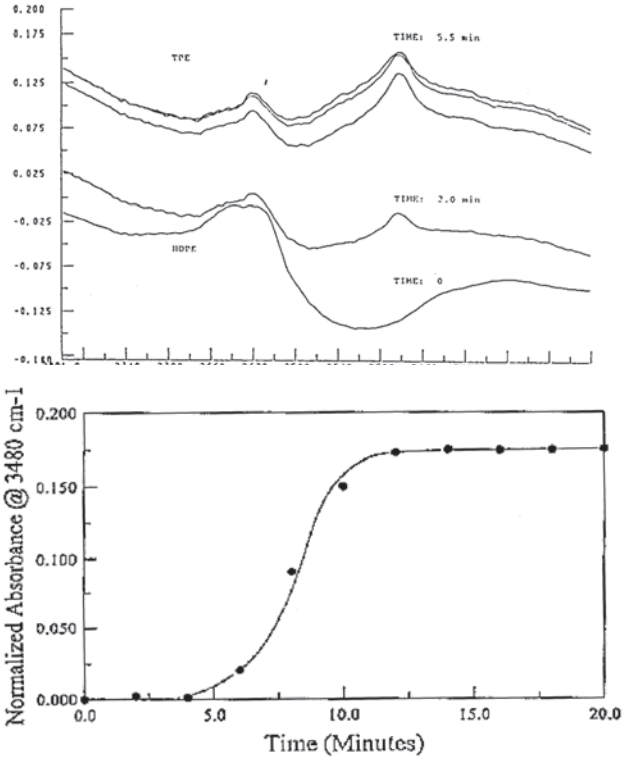


Figure 3.50. FTIR monitoring of product change from HDPE to TPE.

forces the HDPE out of the extruder and the HDPE concentration decreases, the TPE concentration increases, corresponding to increased peak intensity at 3480 cm^{-1}. The second graph in Fig. 3.50 shows the normalized absorbance at 3480 cm^{-1} versus time. Approximately 12.5 minutes after the product change is initiated, the product exiting the extruder is 100% TPE and can be collected. Other applications are defined in Dynisco literature.[2]

3.6.4.3 Laser Micrometers and Thickness Gauges

Laser micrometers are used with processes such as profiles, pipe, monofilament, tubing, etc., to make in-line dimensional measurements on the products. Feedback loops to increase or decrease the puller speed, based on the product dimensions, can be used to immediately place products back into specification. Without feedback loops, laser micrometers are used for quality assurance and real time measurements of product dimensions.

Thickness gauges are used with sheet and/or film processes to determine sheet or film thickness and product uniformity. Similar to laser micrometers, the output can be used strictly for quality assurance purposes or in feedback loops to the pullers or the extruder screw speed to change thickness. Sheet and film dies are available with automatic die lip adjustments to correct cross-machine direction sheet uniformity.

3.6.5 Control Summary

Good production practices dictate that all operating conditions and observations need to be recorded to establish a database for troubleshooting product or process problems. New computerized control systems provide process data continuously; however, any unusual occurrence or abnormality needs to be documented on the control chart to identify the cause and effect. It is impossible to record too much information during a run. Records should include setpoint and actual temperatures, pressures, screw speeds, motor load (percent load, amps, or screw torque), melt temperatures, product appearance, raw material type and lot numbers, formulations, drying time, drying temperature, puller speeds, cooling medium temperature, takeoff equipment speeds, and any other observations.

New computerized control and data collection systems keep formulation data, record input operating conditions, and generate control charts for SPC. Error messages are printed out along with process data at prescribed intervals and can document the entire run.

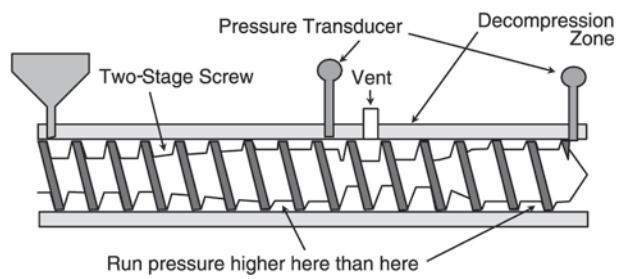


Figure 3.51. Vented extruder with two-stage screw.

3.7 Extruder Devolatilization

A two-stage screw in a vented extruder is used to remove volatiles and/or moisture from polymers. The first screw stage has a feed, transition, and metering section, followed by a decompression zone at the vent location, and finally a pumping section at the end of the screw. A cross section of a two-stage screw in a vented extruder is shown in Fig. 3.51. A pressure gauge at the end of the first stage just prior to decompression assists the operator in properly setting the temperature profile to prevent polymer from flowing out of the vent. When purchasing a vented extruder, the vent port can be specified either on the top or side of the extruder. With proper operation, molten resin does not normally flow out the vent; however, at start-up or during process development with a new product, molten polymer may flow out the vent port if operating conditions aren't properly balanced. Polymeric material flowing out the vent is easier to handle

if the vent is on the side of the extruder where material flow can be collected in a bucket under the vent. Vent ports on the top of the extruder allow polymer to flow out the vent, then flow down on the barrel and around the barrel heaters. A vacuum pump can be attached to the vent port to assist in the removal of volatiles.

The disadvantage of a vented or two-stage extruder is the effective shortening of the extruder *L/D*. Normally, the vent port is about two-thirds the distance between the feed hopper and die; therefore, resin must now be fed, melted, and metered in about two-thirds the barrel length. A 30:1 *L/D* effectively becomes a 20:1 *L/D* extruder, limiting the time required to melt the polymer at a particular rate. Having a mixing section may not be practical because of the decompression zone and the need to use the metering section in the second stage to pump molten polymer to the die.

Single screw extraction can remove approximately 15% of volatiles by weight with high *L/D* extruders and multiple vent ports. One vent port can effectively remove approximately 5% of volatiles by weight. Extruders with multiple vent ports have *L/D*s of 40 to 50:1. Screw sagging or buckling can become an issue, with additional supports required for the extruder barrel.

3.8 Vertical Extruders

Vertical extruders are normally found in smaller equipment. Their advantages are they take up less floor space, can generate higher torques without breaking the screw because the motor is on the metering end where the screw root diameter is larger, require smaller thrust bearings, and eliminate screw sag. Since the extruder screw protrudes into the feed hopper, positive extruder feed similar to a crammer feeder is obtained with the screw biting into the feed and conveying it into the extruder. The disadvantages are the need for higher overhead space and the possibility of melt leakage at the screw shank. Figure 3.52 shows a vertical extruder.

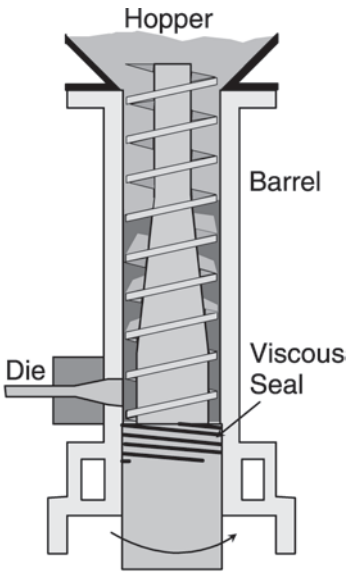


Figure 3.52. Vertical extruder.

REFERENCES

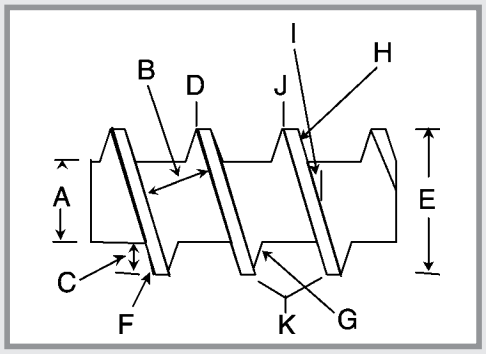
1. Eurotherm Controls, Inc., 11485 Sunset Hills Road, Reston, VA 22090-5286, Technical Information Sheet, Number 101, "Closed-Loop Cycling Method of PID Tuning," May 3, 1996.
2. Dynisco/Kayeness Polymer Test Systems, "FTIR Spectroscopy of Polymers and Additives with IROS," Reference, December 1996.

Review Questions

1. What are the objectives or goals to be accomplished with the extruder in an extrusion process?
2. What are the five major components of a single screw extruder?
3. What is the L/D of an extruder, how is it measured, and how does it differ from one extruder to another?
4. What is the compression ratio of a screw and how is it measured?

Review Questions (continued)

5. Calculate the compression ratio and L/D of a 6-inch diameter screw that is 15 feet long with a feed section channel depth of 1.2 inches and a metering section with a root diameter of 5.08 inches and screw diameter measured with two 0.250-inch gauge blocks of 6.46 inches.
6. What is the purpose of a grooved feed throat?
7. What is the difference between a direct drive and indirect drive extruder?
8. What are the typical sections of an extruder screw?
9. Identify the screw components in the drawing.



10. Why is cooling used on an extruder in both the feed throat and the barrel?
11. What is the purpose of screw cooling?
12. What is the function of the breaker plate?
13. How are screens placed in an extruder breaker plate for filtration? Place 40-, 80-, and 120-mesh screens in the proper order.
14. What parameters can be changed and controlled on an extruder to alter the process parameters when troubleshooting an extrusion problem?
15. When is a two-stage screw used?
16. What parameters need to be specified when ordering a new extruder screw?
17. What are some reasons for changing the current screw design to something new?
18. What is used at the end of an extruder barrel for safety in the event of high pressure?
19. What happens if a thermocouple fails in the closed position?
20. What are the three types of thermocouple junctions?
21. Explain how thermocouples work.

Review Questions (continued)

22. Why are melt temperature and pressures imperative in controlling and troubleshooting an extrusion process?
23. How do changes in die pressure affect the extruder output?
24. What is the difference between an on/off proportional controller and a proportional controller with integrator?
25. What is the typical range of shear rates in an extruder?
26. What differences, if any, are observed in pressure at the end of the extruder, in the adapter, and die?
27. Processing polycarbonate (n = 0.7) at 455 pounds/hour in a 4.5-inch diameter extruder, the pressure gauge before the breaker plate is oscillating between 1900 and 2700 psi, while the pressure gauge in the die is oscillating between 2300 and 2500 psi. Is this process in control and what is the variation in throughput?
28. What is the advantage of screw cooling?
29. What is B-10 life and what does it measure?
30. What is the difference between a DC and an AC motor to drive an extruder?
31. What are the requirements for an extruder motor?
32. How does a pressure transducer operate?
33. Does a flush mount melt thermocouple give accurate melt temperature readings and why or why not?
34. What types of barrel surfaces are available?

4 Plastic Behavior in the Extruder

Polymer response or behavior in the extruder combined with the extruder processing conditions, i.e., barrel temperatures, screw speed, and screw design, are what allow the extruder to extrude a homogeneous polymer melt at constant pressure and temperature. In Chapter 3, the extruder equipment was broken down into five distinct areas:

- Drive system
- Feed
- Screw, barrel, and heating
- Die and adapter
- Controls

This chapter discusses the polymer behavior in six processing zones:

- Feed
- Melt or plasticating
- Melt conveying
- Mixing
- Devolatilization
- Die forming

The name of each zone defines the activity in that zone. In the feed zone, unmelted polymer flows from the feed hopper through the feed throat and feed section to extruder zone 1. In zone 1 (often referred to as the feed zone), the formulation is being compacted as it is conveyed forward to the transition zone. Melting or plasticating is the section where polymer is converted from a solid to a melt for semicrystalline polymers or softened to an acceptable processing viscosity with amorphous polymers. After plastication, the melt is conveyed forward through the metering section and pumped to the die. In a two-stage extruder, the melt may go through a devolatilization or decompression zone where moisture, volatiles, or solvent are removed. To achieve a homogeneous polymer melt, the melt can be mixed before it is conveyed to the die forming zone, where the product is shaped. Each zone is discussed in detail in this chapter.

4.1 Feed Zone

The simplest feed system is gravity-induced solid conveying of polymer pellets, powder, additives, reinforcements, etc., from the feed hopper into the screw channel. The driving force is gravity, material weight, and the formulation fluidity in the solid state. Typical polymer densities in the solid or melt range from 0.9 to 1.7 gram/cm^3 (56–106 pounds/ft^3), depending on the polymer type, additives, and filler contents. Powder and pellet bulk densities are significantly lower than solid or melt densities due to packing effects and the air between individual particles. Bulk densities range from 0.3 to 0.7 gram/cm^3 (19–44 pounds/ft^3). Additives with material bulk density below 0.2 gram/cm^3 (13 pounds/ft^3) do not feed well on conventional equipment that relies strictly on gravity as the driving force for feeding. Fumed silica is an example of a low bulk density additive requiring a crammer feeder to provide adequate feed rates to the extruder.

A second bulk property affecting feed characteristics is material compressibility. Compressibility is the ability of loose material to occupy less volume during shipping or handling. Material compressibility is easily determined by taking a handful of product and squeezing it tightly. If the volume is reduced on squeezing, the material is compressible. A measure of the compressibility is the difference between the bulk density of loose and packed particles. Percent compressibility is defined by Eq. (4.1):

$$Compressibility\,(\%) = \frac{Loose\ Bulk\ Density - Packed\ Bulk\ Density}{Packed\ Bulk\ Density} \times 100 \tag{4.1}$$

Above 20% compressibility, the distinction between free flowing and nonfree flowing material is normally defined. Material having compressibility greater than 20% is anticipated to have feed problems in conventional gravity-fed extrusion systems.

Feed problems encountered in a flood-fed feed hopper include bridging, funneling or “rat holing,” and selective entrapment. Bridging is caused by

- Compressibility of polymer or additives in the feed hopper
- Nonfree flowing formulations
- Material softening in the feed throat and adhering to the feed hopper or the feed throat walls
- Large chunks of material (particularly regrind)
- Fibrous regrind
- Fiber reinforced pellets
- Low bulk density materials (fiber fluff, fumed silica, etc.)

that hinders or prevents free flow. Figure 4.1 shows bridging in a feed hopper near the entrance to the extruder feed throat. Material forms a bridge across the bottom of the hopper, preventing polymer from flowing into the extruder. Bridging can happen either very rapidly or slowly over time, depending on the mechanism causing the phenomenon. If it is caused by heat, the material softens or becomes tacky at the bottom of the feed hopper or in the extruder feed throat. The material builds up slowly over

Figure 4.1. Feed hopper bridging.

time on the feed throat wall, decreasing the feed rate gradually over time until the throat is completely blocked. With a slow blockage of the feed throat, the product dimensions change gradually over time, accompanied by a decrease in the extruder motor load. Since material is still feeding and product is exiting the extruder, bridging is not always readily recognizable as the problem. With fluff, low bulk density materials, or regrind, bridging normally occurs very rapidly (over a minute or two), accompanied by a loss of feed, no extruder output, and the extruder motor load decreasing to almost zero. Corrective action for bridging is to shut down the extruder and remove the feed material from the hopper and the material causing the bridge. Sometimes the bridge can be broken or eliminated without shutting down the extruder and restringing the line by simply poking a plastic rod into the feed throat area to force the bridged material into the screw. Preferably the plastic rod is the same polymeric material being processed so it can be forced into the feed throat and rotated to break up the bridging material in the feed hopper or feed throat area. Assume the plastic rod gets caught on the screw; the plastic is cut by the screw rotation and conveyed into the extruder, melted, and becomes part of the output. If the rod is the same composition as the formulation being processed, contamination does not create a product issue. Another approach is to use a wood rod to achieve the same objective. The downside of a wood rod is if it gets caught on the rotating screw and sheared, wood chips are transported through the extruder and deposited on the screens, partially blocking the screens and increasing the backpressure. **Do not stick your fingers into the feed throat with the screw running to break up a bridge; you could lose a finger or more. Do not stick a metal rod into the feed throat to break up a bridge with the extruder running; this could break a screw flight or worse if it gets caught on the screw.**

Intermittent or partial bridging in the feed hopper or feed throat results in nonuniform feeding, translating to output variations. Assume the feed is about to bridge and suddenly breaks loose. A short time later it starts to bridge again and breaks loose. The product dimensions are constantly varying. Extruder output uniformity is dependent on feed uniformity.

If bridging is a constant problem with a particular formulation or feed stream, an air vibrator or air hammer installed on the hopper, operating either intermittently or continuously, may eliminate the problem. However, if bridging is caused by materials compressing and sticking together in the feed hopper, installing an external air vibrator on the hopper may aggravate the problem by assisting compaction and creating an even more severe bridging problem. Hoppers containing rotating screws or other rotating mechanisms act as bridge breakers. Individual volumetric or gravimetric feeders are normally equipped with some rotating paddles, auger, flex-wall hoppers, or blades above the feed screw to prevent bridging. Depending on the bulk density and nature of the feedstock, bridging can still occur above the bridge-breaking mechanism in the feeder. Extruder starve feeding is another method to overcome bridging in the extruder feed hopper, provided that bridging is not an issue in either the volumetric or gravimetric feeder feeding the extruder.

Figure 4.2. Funneling.

Funneling or "rat holing," shown in Fig. 4.2, occurs when material is not free flowing and sticks to the hopper walls. Center flow in the hopper leaves material to build up on the hopper walls. Compressible material, sticky material that tends to agglomerate, nonfree flowing powder or pellets, regrind or recycled plastics, and materials that stick to hopper walls may create the phenomenon described as funneling. Potential methods to eliminate the problem include using a plastic rod of the same polymer being run in the formulation to break up the materials clumped on the feed hopper walls, tapping the hopper walls to break up material agglomerated on the side walls, or installing an air vibrator on the side of the hopper. Depending on the bulk material characteristics, an air vibrator may aggravate the problem by compressing the feed material to a greater degree. Adding talc, at low levels, or another additive to improve the formulation fluidity may eliminate funneling; however, the talc or other additives should not affect the physical property profile or product specifications.

Selective entrapment occurs in mixtures with different size particles, i.e., pellets and powder, where the particles can separate. The bottom of the hopper becomes richer in smaller size particle material, while the top is richer in large size particle material. Consequently, at the beginning of the run there may be a higher concentration of the smaller size particle in the product, while the end of the run is richer in the larger size particle material. Figure 4.3 shows selective entrapment. Selective entrapment is a concern with pellet/powder blends or large regrind

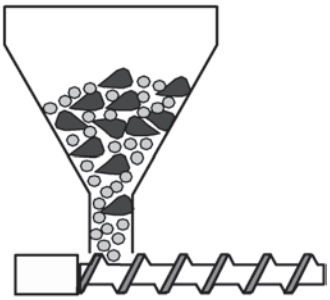


Figure 4.3. Selective entrapment.

particles combined with smaller virgin pellets or powder. Care must be taken in handling mixtures of different size particles to prevent segregation during transfer and moving. In many instances, a formulation is prepared in a blending location away from the extruder. After a formulation is properly blended, it is transferred to a box, bin, drum, etc., and transported to the extruder. The blend is then loaded to the feed hopper pneumatically, by vacuum, or by buckets. During transfer or material flow to the extruder from the feed hopper, different size components in the formulation can separate. If the blend is being compounded into pellets, postblending the pellets ensures a uniform product. However, if the product is being extruded into film, profile, wire coating, or other finished form, postblending is not feasible; therefore, care is required in all handling and transporting operations to minimize formulation separation. The best method to eliminate selective entrapment is to use separate feeders for each component and meter the product into the extruder by starve feeding or feeding just enough material to cover the screw in a flood-fed situation.

Large particle size, lubricants, or liquids can cause slippage in the feed section. Figure 4.4 shows a slippage situation caused by large particles. Grinding or pulverizing the particles to a smaller size can eliminate this problem. Slippage due to either internal or external lubricants coating the barrel walls is an issue that is extremely difficult to overcome in normal operation when lubricants or mold releases are being compounded. Two ways to eliminate or reduce slippage are a crammer feeder or grooved feed throat, which automatically increases the frictional characteristics between the feed material and the barrel walls.

Figure 4.4. Slippage.

Cutting large pellets or regrind chunks between the screw flight and feed throat is another feed problem. This problem does not create the same consequences and output variations as other feed issues discussed previously. Depending on the material being processed, cutting can lead to screw wear in the feed section over time. Figure 4.5 shows cutting.

Figure 4.5. Cutting.

The final feed problem is called ejection; it is where large rubbery particles can be ejected or forced out the extruder by screw rotation. Ejection is shown in Fig. 4.6.

Figure 4.6. Ejection.

Feeding is achieved by dragging the polymer formulation forward in the extruder barrel, away from the feed hopper. Figure 4.7 shows the material feeding and being conveyed forward in the barrel. Frictional properties between individual particles within the formulation and between the formulation and the feed hopper, feed throat, screw, and barrel wall determine the feeding characteristics and efficiency of the materials being extruded. Two types of friction are present in the feed area:

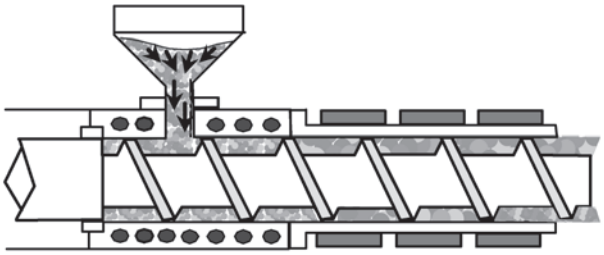


Figure 4.7. Flood feeding in feed section of extruder.

1) Internal friction between the individual polymer particles, formulation ingredients, and each other
2) External friction between the formulation components and the metal surfaces of the feed hopper, feed throat, barrel walls, and screw root

Good feed characteristics in the feed hopper require both the internal friction and external friction to be low. Low internal friction allows the formulation components to flow freely past one another into the extruder (eliminating many feed problems discussed previously), while low external friction allows the formulation to flow freely along the hopper walls into the extruder. In the feed zone, high internal friction is desirable to encourage material entering the extruder to stick together as a plug and be dragged forward in the feed section to the transition section. External friction between the formulation components and the barrel wall needs to be high, encouraging the material to stick to the barrel wall, while the external friction between the screw root and the polymer must be low, allowing the polymer to slip on the screw root. This can be simulated with a nut and bolt. Assume the nut is the polymer formulation moving as a plug, and the threaded bolt is the extruder screw. If the nut is held on the outside and the bolt turned, simulating high external friction between the polymer and barrel wall, the nut moves down the bolt similarly to the polymer being conveyed down the extruder barrel. If the nut is not held and the bolt is still rotating, simulating high external friction between the polymer and the screw root with low external friction between the polymer and the barrel wall, the nut is not conveyed down the screw but stays in the same location, rotating with the screw. The highest conveying rate is attained when all plastic moves forward down the screw as a plug.

The relationship between solids conveying rate and the coefficient of friction between the barrel wall and the formulation components is shown graphically in Fig. 4.8.[1]

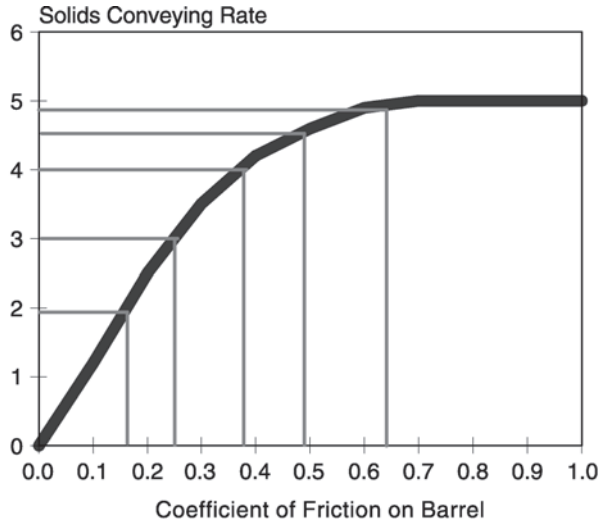


Figure 4.8. Effect of barrel/polymer coefficient of friction on solid conveying rate.

With a low coefficient of friction, the solids conveying rate changes dramatically with small changes in the coefficient of friction. Above approximately 0.65 coefficient of friction, the solids conveying rate changes very little as the coefficient of friction changes. In practice, if the processing conditions cause the solids conveying rate to be on the steep side of the curve (low coefficient of friction), small changes in process conditions in either the feed throat or extruder zone 1 can result in dramatic differences in output rate. Low coefficient of friction between the polymer and barrel wall relative to the material coefficient of friction with the screw root leads to poor feeding characteristics. Poor feeding characteristics translate to polymer flow variations, causing output variations. Maximum friction between the polymer and the screw root results in the total loss of polymer feed, creating a condition called melt-over or melt plug. A melt plug is solid plastic melted around one or more flights in the feed zone, blocking resin feed and preventing flow. As the screw turns, the polymer sits in one position—similar to the nut and bolt simulation, where the bolt is being turned and the nut is not being held, the nut stays in the same relative position on the screw.

Temperature in the first extruder zone is critical for good feed characteristics. If zone 1 temperature is too high, premature melting may produce a lubricating film on the barrel wall, reducing the external friction between the barrel wall and polymer and leading to polymer slippage. If the zone temperature is too low, the polymer external friction drops, giving poorer solids conveying. The ideal zone 1 temperature allows the pellet surface to become tacky, generating high internal friction to assist the polymer moving forward as a plug with screw rotation.

If zone 1 temperature is too hot and the extruder screw is stopped with material in the feed hopper, the polymer can melt, forming a molten plug or melt-over around the channel in the feed section. If this situation happens with the extruder running and extrudate uniformly flowing out the die, suddenly the extrudate stops flowing from the die with the screw still turning and the polymer level in the feed hopper still full. The screw, according to the rpm gauge, is still turning at the original setting, but the motor load is significantly reduced or almost zero. This indicates either a melt-over or melt plug in the feed zone or a broken screw. A melt-over is caused by zone 1 temperature being too high; the polymer feed becomes molten, sticks to the screw root, and slips on the barrel wall. Removing a melt plug can be a challenge. A number of removal strategies can be attempted; however, if they all fail it may be necessary to remove the screw from the extruder and clean it. Screw removal is a drastic, messy procedure with potential safety hazards. The screw is coated with hot molten plastic, making removal difficult and increasing the possibility of burns and fume inhalation. The first step in any corrective action procedure is to stop the extruder, remove all pellets and additives from the feed hopper, and vacuum all polymer from the feed throat, removing as much unmelted material as possible. With the feed throat clean, try some of the procedures listed below to remove the melt plug before pulling and cleaning the screw.

- Raise extruder zone 1 temperature to lower the melt plug viscosity. Force the melt plug out the feed zone with new polymer, starve fed to the extruder. If successful, lower zone 1 temperature while continuing to feed polymer. The downside of this procedure is the possible generation of degraded polymer that may be difficult to purge from the extruder once the plug is eliminated. If the melt plug is PVC, high temperatures may start an autocatalytic degradation reaction generating HCl and degraded polymer, requiring the screw to be pulled to properly clean the extruder before restarting.
- Cut a block of wood approximately the same size as the feed throat. Place a small amount of pellets of the same formulation in the feed throat and use the block of wood as a crammer to force the pellets into the extruder screw.
- Using strips of plastic of the same material being extruded, force the plastic strips into the feed throat, letting the screw pull them into the melt plug, attempting to break the melt-over and encouraging the formulation to flow freely again.
- Stop the extruder screw and force a piece of metal banding strap around the screw like a plumber's snake to break the melt plug. This is difficult to do in small extruders. **Do not restart the extruder until the banding strap has been removed.**

- The last resort is to pull and clean the screw and barrel, reinstall the screw, and restart the extruder.

Prevention of melt-over or melt plugs can be accomplished using the following procedures:

- The feed throat temperature needs to be properly controlled. There are two sources of feed throat cooling in normal operation. One is the water cooling jacket around the feed throat with adequate water circulation at the correct temperature, and the second is the raw material flow to the extruder. The feed throat jacket normally feels warm to the touch, but not hot. Normal feed throat operating temperature is 100–120°F (38–49°C).
- The zone 1 temperature is set below the polymer melting point to prevent polymer from completely melting or plasticating in the feed throat or zone 1. Verify the heater bands and thermocouples are functioning properly.
- Don't stop the extruder with material left in the feed throat, where it can absorb heat from zone 1 and melt or plasticate. Stop the material flow from the feed hopper using the slide gate at the feed hopper, and run the extruder empty before shutting down. Alternatively, run both the feed hopper and extruder empty before shutting down.

Output surging is often caused by poor or inconsistent feeding. Surging is defined as output variations where the material exiting the extruder per unit of time is inconsistent. Possible causes of surging resulting from polymer feed are

- Poor feeding characteristics from the feed hopper, resulting from a nonfree flowing formulation, size variations in feedstock that cause partial bridging in the feed hopper or feed throat, or nonuniformity in the feed stream
- Polymer or some component of the formulation sticking to the screw root and not the barrel wall as a result of the wrong temperature profile, slippery material, or size variations in the feed stream (pellet/powder mix), causing some components of the formulation to melt prematurely
- Condensation in the feed throat caused by too much feed throat cooling
- Slippage at the polymer barrel interface due to high lubricity of pellets or formulation additives, e.g., silicone oil, external or internal lubricants, or liquid feed into the feed throat.

Zone 1 temperature is a critical temperature in the barrel temperature profile. If extrusion problems arise, zone 1 temperature is probably the most overlooked temperature zone as operators and engineers change temperature settings toward the exit end of the extruder. Without consistent feed at a proper rate, all other functions in the extruder can be overshadowed.

To accomplish consistent feed in gravity-fed flood feeding, the level in the feed hopper needs to be maintained at a constant level. When the hopper is completely full, there is more weight pushing down on the material flowing into the extruder than when the hopper is almost empty. To achieve tight dimensional tolerances in the final product, the feed hopper level must be maintained constant. One way to accomplish this is with a vacuum loader that automatically reloads the hopper as the level decreases.

With an understanding of how the extruder feed mechanism works, it is easier to understand why a grooved feed section results in significantly higher feed rates. Pellets trapped in the grooved wall create higher external friction between the material and the barrel wall. As the screw rotates, the pellets in the grooves are forced forward in the feed throat by the screw flights. High external friction correlates with very high conveying rates (Fig. 4.8). To handle these high feed rates, the compression ratio of the screw is reduced from 3–4:1 to 1.5–2:1. The feed section wall strength must be increased to compensate for the higher pressures generated by the solids compaction. The increased volume of pellets in the grooves plus the pellet volume in the screw channel in the first few screw flights are combined at the end of the feed section into the same screw channel volume in zone 1 that was present in the feed section.

4.2 Polymer Melting or Plastication

Melting or plastication starts in the feed zone approximately five or six screw diameters from the feed opening. Semicrystalline polymers go through a sharp melting point, while amorphous polymers will continue to soften above T_g until the viscosity is low enough to process the polymer. Two heat sources are available to melt or plasticate the polymer: conduction from the barrel and viscous shear heating. Most heat is generated by the extruder motor turning the screw, creating polymer shear heating. Approximately 80–90% of the heat necessary to melt the polymer is generated through shear by the screw rotation, while the other 10–20% comes from the barrel heaters.

Shear heat comes from two sources; one is the scraping of the resin in contact with the barrel wall by the screw, and the other is the individual layers of polymeric materials sliding over or under each other during the laminar flow, generating viscous heat. Viscous shear heating is similar to having two pieces of paper slide by each other, generating frictional heat between the papers. With polymer flow, the molecular friction or resistance in one layer as it flows over the molecules in another

layer causes heat to be generated and the polymer to melt. *Shear* is defined as the movement of one layer in a fluid or solid relative to a parallel or adjacent layer. Shear in the extruder is caused by molten plastic moving in a direction parallel to a fixed surface such as the barrel wall or screw, in addition to movement relative to other layers within the polymer. Shearing occurs when fluids flow through channels, as in the extruder or tubes in transfer pipes and dies. Two other terms associated with shear are *shear strain,* defined as the movement of one polymer layer relative to an adjacent layer divided by the layer thickness, and *shear rate,* defined as the rate of change in shear strain over time. Increases in screw speed and throughput rates are directly proportional to increases in shear rate. As the shear rate increases, polymer viscosities decrease, increasing the polymer flow. *Shear stress* is defined as the tangential force acting on the molecular plane to move the polymer per unit area. Shear stress is the polymer viscosity multiplied by the shear rate. With viscoelastic materials, shear stress is a function of both shear strain and shear rate. *Shear flow* is the sliding of parallel layers relative to each other. Viscous heat generated during the extrusion process is due to the force necessary to make the polymer molecules flow. Due to the high polymer viscosity and relatively high shear rates in an extruder, viscous heat generation is the most significant factor contributing to polymer melting and plastication.

Shear rate versus viscosity curves were introduced in Chapter 3, Fig. 3.46, showing the effect of temperature on viscosity at different shear rates. Molten polymers are viscoelastic fluids, meaning the polymer flow has both a viscous component and an elastic component. A Newtonian fluid like water has only a viscous component. Referring back to Fig. 3.46 in Chapter 3, increasing or decreasing the shear rate on water does not change its viscosity; consequently the viscosity versus shear rate curve for water is a straight line parallel to the X axis. The elastic component in polymer flow makes polymers non-Newtonian in nature, resulting in shear thinning or polymer viscosity decreases with increasing shear rate. Unfortunately, as the shear rate increases, more heat is generated at the shearing surfaces, leading to possible polymer degradation and/or discoloration. In addition, changes in viscosity can create either sizing or cooling problems in the takeoff equipment. Shear rate calculations in different areas of the extruder are presented in Chapter 8.

The polymer melting process in the extruder is shown in Fig. 4.9 and described below:

- The plastic solid bed is pushed forward by the pushing edge of the flight in the feed section.
- As the polymer approaches the screw transition section, a thin melt film develops between the solid bed and the barrel wall.
- In the transition section, the polymer is compressed by the solids conveying pressure and by the increasing screw root diameter as the material is conveyed forward.
- Polymer melts in the thin melt film region generated between the solid bed and barrel wall from the combination of shear and barrel heating.
- Viscous heat generation occurs at the boundary between the melt film and the solid bed.
- The pushing flight scrapes the melt off the barrel wall, creating a melt pool against the pushing flight and forcing the solid bed forward against the trailing flight.
- As polymer is conveyed forward in the transition zone, the melt pool increases in size as the solid bed melts.

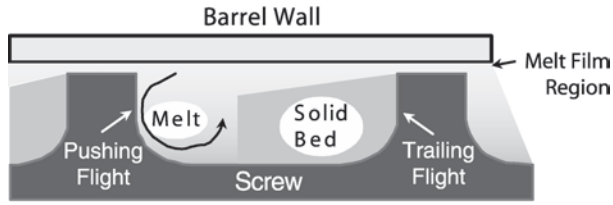


Figure 4.9. Polymer melting model in the transition section of the screw.

Figure 4.10 shows the melting mechanism in the transition zone.

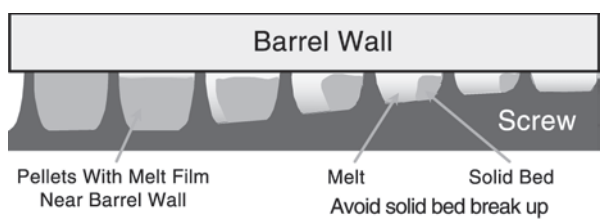


Figure 4.10. Melting mechanism in transition zone of single screw extruder.

The melt film is sheared at a high rate between the solid bed and the barrel wall, resulting in substantial heat generation. A small percentage of melt flows over the screw flight between the barrel wall unless the screw is badly worn. The melt pool at the pushing flight has a circular flow due to the screw velocity. As the melt pool size increases, the solid bed decreases. As the screw channel volume decreases, pressure increases on the solid bed, forcing it against the barrel wall. Ideally the solid bed is completely melted by the end of the transition zone, and there is no solid bed breakup into small unmelted pieces floating in the melt. Assuming the solid bed breaks up, it is very difficult to melt any remaining solid particles as

the high shear region between the solid bed and the wall is no longer present, resulting in significantly reduced viscous heat generation. With solid bed breakup and the absence of compression in the metering zone, the driving forces for polymer melting are reduced. It is similar to ice floating in cold water and trying to put pressure on the ice to melt it. The ice simply moves out of the way. The goal is to complete the melting by the end of the transition section and prevent solid bed breakup.

Variables affecting melting rate include solid bed width, melt film thickness, and barrel temperature. To maximize polymer melting, the solid bed needs to be as wide as possible to create more surface area for melt film formation and the high shear region for viscous heat generation. Figure 4.11 shows wide and narrow solid beds. The wider solid bed has a greater melting rate due to the potential for higher viscous heat generation. These concepts are important when discussing screw design and

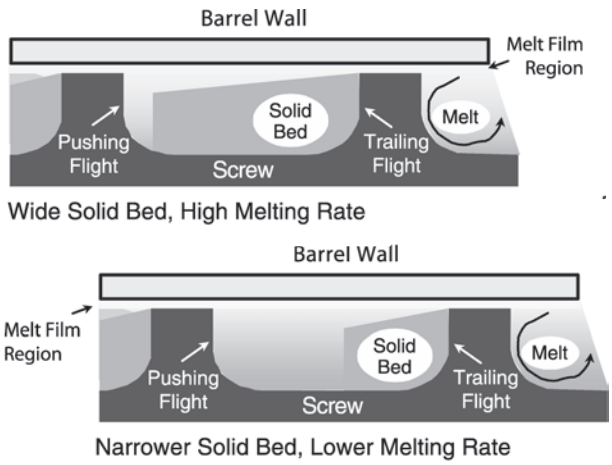


Figure 4.11. Effect of solid bed width on melting rate.

alternative designs in the transition section for improved melting capacity. The second variable affecting melting rate is melt film thickness, which determines the shear rate in the film. Thinner films create higher shear rates, generating more viscous heating and more efficient melting. Extruder barrel or screw wear in the transition zone can increase the melt film thickness and decrease the polymer melting efficiency. The third variable affecting melting rate is the conductive heat from the barrel heaters. Increasing the barrel temperatures may decrease the melting rate. Higher barrel temperatures can increase the melt film thickness, decreasing the shear rate and viscous heat generation. Figure 4.12 shows the effect of increasing the barrel temperature from the barrel heaters and the increased melt film thickness. A higher barrel temperature also lowers the polymer viscosity at the barrel wall. Lower resin viscosity acts as a lubricant to the melt, lowering the viscous heat generation and possibly reducing the overall melting capacity. Barrel temperatures

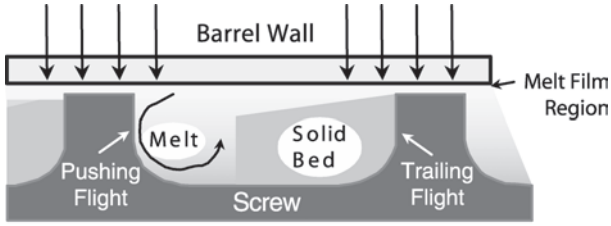


Figure 4.12. Effect of increasing the barrel temperature on the melt film thickness.

in extruder zones 2 and 3 are critical to obtaining the optimum polymer melting. If the temperature is too low, barrel zones may override their setpoint due to the high viscous or shear heating. If temperatures are too high, melting rate may be reduced because the melt film region is too thick and insufficient shear heat is generated. With reduced shear heating and high throughput rates, the possibility exists that all the solid material may not be melted by the end of the transition zone, increasing the likelihood of solid bed breakup or overfeeding the metering zone.

Compression of the material by the decreasing channel depth in the transition section is necessary because the solid bulk density is lower than the melt density. Consequently, the solid requires more volume than the melt; as the solid melts, the transition section compresses the melt, forcing the solid bed against the trailing flight and the barrel wall, continuously generating a new melt film region. Compression in the feed and transition section forces the air between the solid particles back through the feed throat, eliminating any air in the final product.

Polymeric materials most affected by increasing the barrel temperatures are resins with temperature-sensitive viscosities. Polymers that are more Newtonian at the shear rates (50–1000 sec^{-1}) experienced in extrusion exhibit larger viscosity changes with increases or decreases in polymer melt temperature than with changes in shear rate. Generally, amorphous polymer viscosity tends to be more susceptible to temperature than semicrystalline polymer viscosity.

Barrel heating alone to melt polymeric materials is slow and inefficient due to the poor thermal conductivity of polymers. Without appropriate screw designs and barrel wall clearances, viscous heat generation is limited, and continuous extrusion at throughput rates practiced today would be nonexistent.

4.3 Melt Conveying

The third polymer zone in the extruder is called the conveying or pumping zone, where melt is moved from the transition zone to the die. Similar to the melting model in the transition zone, the conveying zone has a

model to simulate melt flow behavior during conveying.[2] In modeling the conveying, assume the barrel is rotating and the screw is stationary. Figure 4.13 shows the polymer flow path in the metering channel. A particular point in the melt (point 1 in Fig. 4.13) close to the barrel wall moves in the direction of the barrel as it rotates until it comes in contact with the pushing flight.

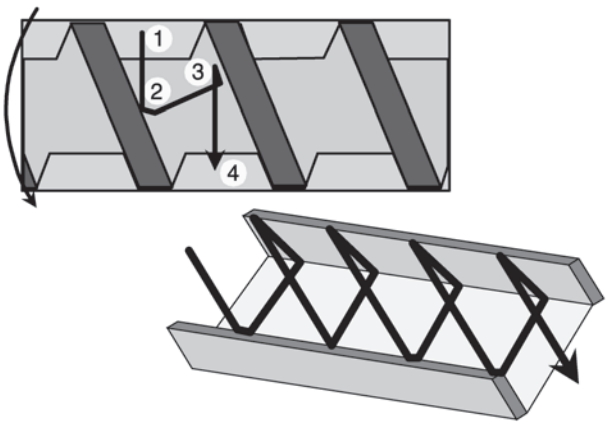


Figure 4.13. Melt conveying model. Material flows down the extruder channel in a spiral motion. Flow is shown as though the screw is stationary and the barrel is rotating. This action provides the total mixing in a single screw extruder when mixing elements are not present.

As the barrel continues to rotate, the plastic is forced down the pushing flight, moving across the channel, where it reaches the trailing flight and turns upward toward the barrel surface. Near the barrel surface, it rotates with the barrel surface again until it comes in contact with the next pushing flight. This spiraling motion shears and mixes the plastic in the metering zone. Single screw extruders with no added mixing are rather poor melt mixers. The only mixing action (assuming no mixing elements on the screw) is provided by this circular motion in the screw channel as molten plastic is pumped to the extruder die plus the backward flow created by the head pressure.

The extruder output is given by Eq. (4.2):

$$Plastic\ Output\ (Q) = Drag\ Flow\ (Q_D) - Pressure\ Flow\ (Q_p) - Leakage\ Flow\ (Q_{LF}) \tag{4.2}$$

Drag flow (Fig. 4.14) is flow created by the rotation of the screw in the barrel. The flow velocity profile near the barrel wall is high, decreasing to zero at the screw root. *Pressure flow* is backward flow of the polymer away from the die toward the feed throat, caused by the pressure prior to the screen pack, in the die, or combinations thereof. The pressure flow velocity profile is similar to plug flow in a pipe with zero velocity at the barrel wall and the screw root and maximum velocity in the center of

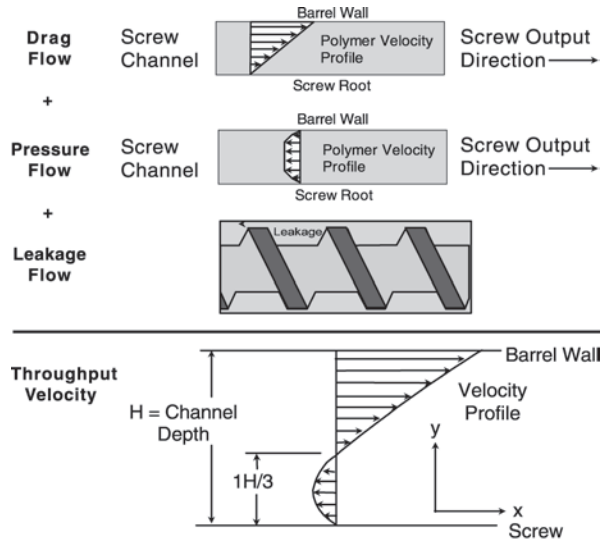


Figure 4.14. Plastic velocity profile in the metering section of the screw.

the channel. *Leakage flow* is the flow over the screw flight. This occurs with worn screws where the flight to barrel clearance is larger than normal; otherwise leakage flow is minimal and normally ignored. Figure 4.14 shows the velocity gradients for the different flows and summation of the drag and pressure flow profile to show the polymer flow in the metering section. Pressure flow assists mixing as the backward flow increases the spiraling action in the channel. Material close to the barrel wall travels at a high rate in the cross-channel direction, while material at two-thirds channel depth has zero cross-channel velocity vector. Molten polymer in the area between zero and one-third up the channel depth from the screw root is traveling toward the feed throat. While some melt mixing occurs in the screw channel, gross mixing of formulation components is done outside the extruder prior to entering the extruder feed throat.

4.4 Mixing

Mixing is another critical extruder function, even when processing only one virgin polymer stream. The extrusion goal of achieving a uniform melt temperature and homogeneous melt can be helped by adding a mixing head. Mixing is determined by the residence time and the shear rate the fluid is exposed to in the mixing section. Single screw extruders without mixing sections do a relatively poor mixing job; the spiraling flow in the metering section and the large variations in shear histories, depending where the material is in the screw channel, do not lead to extensive mixing. Which mixing section to add to the extruder screw depends on the polymer being processed and mixing required.

One method to improve the mixing is to increase the pressure at the end of the extruder or in the die to induce greater pressure flow back into the extruder. This mixing method was tried with a blue color concentrate run in polypropylene in a 30:1, 2.5-inch single screw extruder using 1000 psi and 200 psi head pressure. Core pulls from each material (screw was stopped, cooled, and pulled out of the machine with the polymer still wrapped around the screw) were removed and placed side by side. Negligible differences were observed in mixing with the two different head pressures. At the end of the screw just prior to the material passing through the breaker plate, streaks of dark blue were visible from the color concentrate in both samples. Increases in backpressure or head pressure lead to decreased throughput, higher material residence time, increases in residence time distribution, higher material temperature, and increased chance for material degradation. While backpressure assists mixing, it contributes only marginal improvements to differences in mixing. Definitions of residence time and residence time distributions are:

- *Residence time* is defined as the total time a given particle spends in the extruder from the moment it enters the extruder feed throat until it exits the die.
- *Residence time distribution* is the range in time various particles spend in the extruder. Some particles, depending on their path through the metering section, spend a longer time in the extruder compared to other particles. At a given screw speed, there is an average residence time materials spend in the extruder; some are longer and some are shorter.

General guidelines for single screw mixing element design are[2]

- Mixing section produces a minimal pressure drop with forward pumping capability
- No material hang-up or dead spots exist where polymer can agglomerate and degrade
- Mixing device completely wipes the barrel surface
- Mixing device is easy to disassemble, clean, and reassemble
- Mixing device is reasonably priced relative to the machining cost

The two types of mixing occurring in the extruder are distributive and dispersive mixing. As their names describe, *distributive mixing* evenly distributes particles throughout the melt, while *dispersive mixing* breaks up agglomerates or large particles and disperses them evenly throughout the melt. Distributive and dispersive mixing is shown schematically in Fig. 4.15. Distributive mixing is a low shear process accomplished by repeatedly changing the flow directions by breaking the molten polymer into channels and recombining the melt. Distributive mixing is

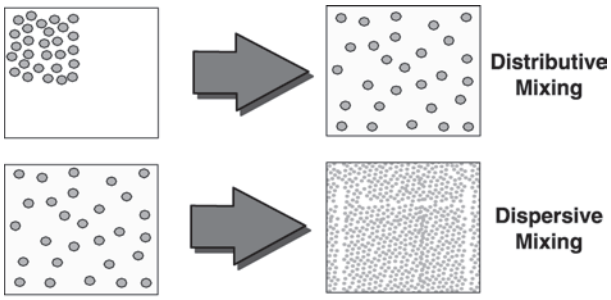


Figure 4.15. Comparison of distributive and dispersive mixing.

used with fibers, reinforcing fillers, and shear-sensitive materials to provide a uniform melt temperature. Dispersive mixing is a high-shear process where molten polymer is forced through very small openings, generating significant shear heat. Dispersive mixing is used in alloying different plastics, pigment dispersement, and mixing nonreinforcing fillers and additives, such as flame retardants, impact modifiers, lubricants, and so forth.

Flow direction changes determine the distributive mixing. Distributive mixing sections divide the flow into multiple channels, recombine the flow, break the flow, etc. Some distributive mixers are shown in Fig. 4.16. Table 4.1 provides and compares different distributive

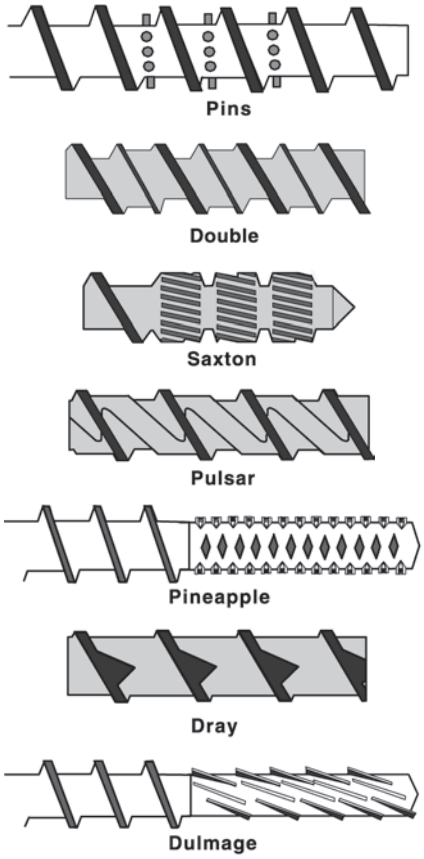


Figure 4.16. Distributive mixers.

Table 4.1. Comparison of Distributive Mixers

Type	Pressure Drop	Streamlined Flow	Barrel Wipe	Operator Friendliness	Machining Rate	Shear Strain	Splitting/ Reorientation	Overall Rating
Saxton	4	4	5	4	4	4	5	30
Axon	4	4	4	4	5	4	3	28
Dulmage	4	4	2	4	4	4	5	27
Stat-Dyn	2	4	4	4	3	4	5	26
Pineapple	2	4	3	4	4	4	5	25
Stratablend	4	3	4	4	3	3	2	25
Double Wave	4	4	4	4	2	4	2	24
Pulsar	4	4	4	4	3	3	2	24
Pins	2	2	3	4	5	2	4	22
Dray	2	3	4	4	3	3	2	21
Cavity Transfer	1	3	2	1	1	4	5	17
Blockhead	1	1	1	4	4	2	4	17

mixers.[2] Ratings are based on the following requirements for a good mixing section:

- Low pressure drop
- Streamlined flow
- Barrel wipe
- Operator friendliness (ease of cleaning)
- Machine rate (cost effective)
- Shear strain (amount one polymer layer moves relative to another)
- Splitting and recombining of different streams

Higher ratings produce better distributive mixing.

As an example, consider the Dulmage, pineapple, or Saxton mixing section shown in Fig. 4.16. Polymer exiting the metering section is pumped in a spiraling motion to the mixing section. Once the material gets into the mixing section, the flow is broken up from one large channel or flow front into numerous small channels or flow fronts. Flow fronts recombine and are subsequently redistributed into small flow channels again and then recombined. Other distributive mixers have obstacles in the screw channel to change the polymer flow direction and to achieve the required mixing.

Dispersive mixing requires high shear rates and high shear stresses that are achieved by forcing the melt over a restrictive barrier. Dispersive mixing sections require more energy than distributive, which may raise the polymer temperature and cause degradation. Three common dispersive mixing sections are shown in Fig. 4.17. A Maddock mixer (also called a Leroy mixer) contains numerous channels that open on one end and are parallel to the screw length. Alternating channel openings face toward the metering section and the die end. Molten polymer, flowing into the channel from the metering section, is forced over a restrictive barrier into the next channel before exiting toward the die. All material is forced over this restrictive barrier with low clearance between the flight and the barrel wall before it can exit the extruder. The second example of a dispersive mixing section in Fig. 4.17 is a blister ring. The clearance between the barrel wall and the blister ring is greater than the distance between a screw flight and the barrel wall. All molten polymer flows over the blister ring in a high-shear environment. The third example shown in Fig. 4.17 is a barrier screw. A secondary screw flight with a slightly smaller outside flight diameter and different pitch splits off from the primary flight. Initially the secondary channel width is very small compared to the primary screw channel. As the flights revolve around the screw, the secondary channel width increases and the primary channel width decreases. Eventually, the secondary channel becomes the principal

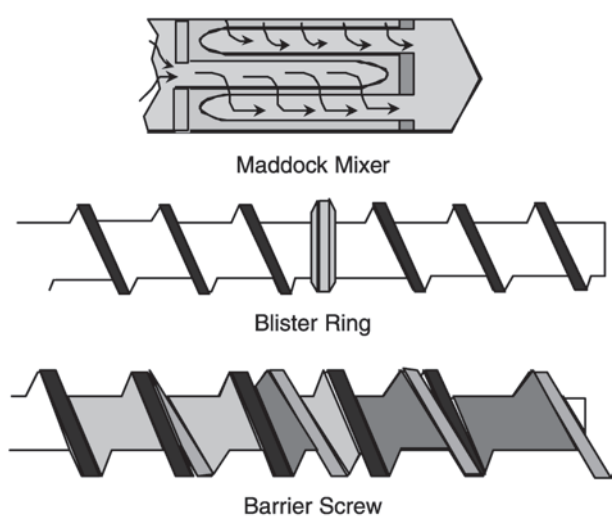


Figure 4.17. Distributive mixing sections.

channel and the primary channel disappears. Molten polymer must cross over the secondary flight into the new or secondary channel. As the primary channel and screw flights disappear, the polymer crosses the barrier or secondary flight between the flight and the barrel wall. In Fig. 4.17, the barrier screw shows only three flights, where the secondary flight starts and the primary flight stops. In actual screw designs this may occur over three to six flights. Barrier flights in the screw metering section are designed for dispersive mixing, while barrier flights in the transition zone are designed for improving the melting capacity. Dispersive mixers prevent solid polymer from entering the die. Solid particles will not pass through the tight tolerances between the barrier and barrel wall.

Another mixer is called a cavity mixer, with cavities in the screw designed for distributive mixing. These mixers are inferior to the other distributive mixers listed in Table 4.1 as they do not wipe the barrel, are not operator-friendly because they are hard to clean, have no forward pumping capabilities, and generate large pressure drops. Cavity mixers are normally added as extensions to the screw when mixing is inadequate.

Another approach to use when mixing is inadequate and no mixing elements are present on the screw is a static mixer in a transition pipe between the extruder head and the die. Many static mixer designs are available that provide good distributive mixing by constantly dividing and recombining the flow stream. Static mixers are installed in pipes where the fluid motion is essentially plug flow. Figure 4.18 shows the polymer temperature profile before and after a static mixer.[3] In addition to improved melt uniformity with a static mixer, color and additive homogenization is improved so lower concentrations may be used in the formulation to attain the same results. Static mixers are used extensively in fiber production. Figure 4.19 shows a static mixer. Static mixers are available in many different configurations and geometries, depending on the mixing requirements.

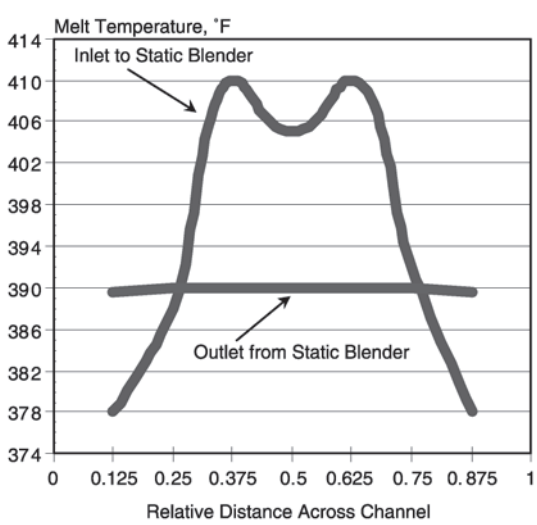


Figure 4.18. Effect of static mixer on melt temperature uniformity.

Figure 4.19. Kenics static mixer.

4.5 Extruder Throughput Calculations

The extruder output is defined by a process limiting factor. Limiting factors in a particular process include extruder capacity (melting capacity and motor power limits), calibration and downstream cooling capacity (limited by heat transfer between the polymer and cooling media), capacity for manipulating the extrudate (postextrusion shaping, changing spools on winders, etc.), die performance (stress, pressure velocity, and temperature), polymer rheological properties, and line length as defined by space limitation (can't install more cooling due to inadequate space). Ideally the throughput limiting factor is the most expensive equipment in the extrusion line. In other words, the throughput should not be limited by an undersized cooling bath, since it is relatively inexpensive compared to an extruder.

Extruder output is equal to the volumetric drag flow rate minus the pressure flow rate minus any leakage flow, Eq. (4.2). The polymer velocity profile in the screw channel for the different components used to calculate the throughput was shown in Fig. 4.14. Volumetric drag flow rate, Q_D, for a Newtonian fluid is given by Eqs. (4.3) and (4.4):

$$Q_D = \frac{W \times H \times V_Z}{2} \tag{4.3}$$

$$V_Z = \pi \times D \times N \times \text{Cos}\varphi \tag{4.4}$$

where

W = channel width
H = channel depth
V_Z = plastic velocity in the channel
D = screw diameter
N = screw speed in rpm
φ = helix angle

Figure 4.20 shows the parameters relative to the screw geometry to calculate the volumetric drag flow and pressure flow rates. Since all the parameters are known, the vol-

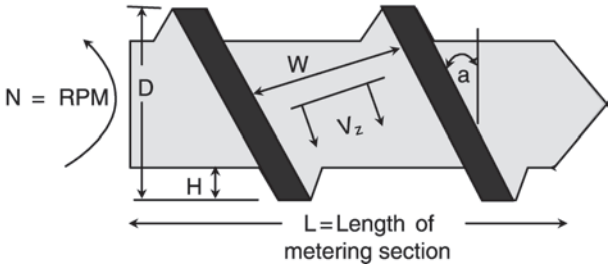


Figure 4.20. Parameters to calculate volumetric drag and pressure flow rates.

umetric drag flow rate can be calculated assuming the plastic is Newtonian. Plastics are viscoelastic, non-Newtonian materials that have viscous and elastic components, and their viscosity is a function of shear rate. A power law model is a simple model used to describe melt viscosity as a function of shear rate. Multiplying Q_D in Eq. (4.3) by the correction factor $(4 + n)/5$, where n is the power law index, yields a good approximation of the screw throughput due to drag flow. For a Newtonian fluid, $n = 1$; so the closer n is to 1, the more Newtonian the polymer flow behavior. Polymers such as polycarbonate with a power law index of 0.7 display Newtonian behavior at low shear rates and non-Newtonian behavior at high shear rates.

The volumetric pressure flow rate, Q_P, for a Newtonian fluid is given by Eq. (4.5):

$$Q_P = \frac{W \times H^3 \times \mathrm{Sin}\,\phi}{12 \times \eta} \times \frac{\Delta P}{L} \tag{4.5}$$

where

$\Delta P/L$ = pressure change in the metering section of length L

η = plastic viscocity at a given shear rate

The power law correction factor for the volumetric pressure flow for non-Newtonian fluids is $3/(1+2n)$.

Table 4.2 lists various polymer properties, including the power law index.[1] As n approaches 1, the polymer viscosity is more temperature-sensitive and less shear-sensitive, meaning the melt viscosity changes more with temperature than with shear rate in an extrusion shear rate range of 50–1000 sec^{-1}. Materials with low n values tend to have melt viscosities that are more shear sensitive and less temperature-sensitive. For example, for a polypropylene with a 0.35 power law index, the melt viscosity changes more with increasing or decreasing shear rate than by changing the melt temperature. The second column in Table 4.2 gives the melting point of various polymeric materials. The first six materials in column 2, polystyrene through polycarbonate, exhibit no melting points because these are amorphous polymers, which soften above T_g, versus semicrystalline polymers that exhibit a distinct melting point. Other polymer properties in Table 4.2 include solid density, thermal conductivity, and specific heat.

Comparing the throughput rates of polycarbonate (PC), $n = 0.7$, and polypropylene (PP), $n = 0.35$, at different head pressures shows lower throughputs for PP (Fig. 4.21). Since PP has a more shear-sensitive viscosity than PC, PP experiences more shear thinning at higher head pressures. Pressure flow increases with lower melt viscosity, decreasing the throughput. Comparison of the two polymers at 2000 psi head pressure shows PC has a throughput of 6 $inch^3$/sec, compared to 2 $inch^3$/sec for PP.

Actual extruder output may differ from that calculated due to factors not considered in the equations. Each polymer has an optimum screw channel depth, helix angle, and melt temperature that may not be optimized. Increased channel depth results in higher backflow or pressure flow, lowering product throughput. Die opening, combined with material flow through that opening, determines head pressure. Large die openings with low throughput have low die pressure, while small die openings with high throughput generate high die pressure. Screw design and rpm determine the shear rate. Deeper

Table 4.2. Polymer Properties

Material	Power Law Index, n	Melting Point, Tm (°C)	Glass Transition T_g (°C)	Density g/cc	Thermal Conductivity, k (J/msC)	Specific Heat, cp (kJ/kgC)
PS	0.3	—	101	1.06	0.12	1.20
PVC	0.3	—	80	1.40	0.21	1.10
PMMA	0.25	—	105	1.18	0.2	1.45
SAN	0.3	—	115	1.08	0.12	1.40
ABS	0.25	—	115	1.02	0.25	1.40
PC	0.7	—	150	1.20	0.19	1.40
LDPE	0.35	120	–120	0.92	0.24	2.30
LLDPE	0.6	125	–120	0.92	0.24	2.30
HDPE	0.5	130	–120	0.95	0.25	2.25
PP	0.35	175	–10	0.91	0.15	2.10
PA-6	0.7	225	50	1.13	0.25	2.15
PA-6,6	0.75	265	55	1.14	0.24	2.15
PET	0.6	275	70	1.35	0.29	1.55
PBT	0.6	250	45	1.35	0.21	1.25
PVF	0.38	275	–40	1.76	0.16	1.38
PA-6,6	0.6	275	70	2.15	0.2	1.18

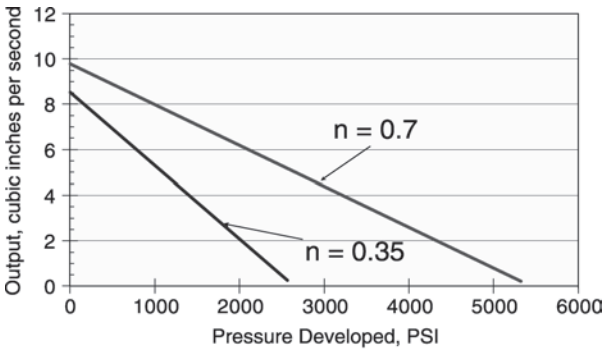


Figure 4.21. Output comparisons of two different power law index polymers.

channel depth in the metering section shears the polymer less. Higher viscosity resins generate more viscous heat, raising the melt temperature, and higher melt temperature lowers the melt viscosity. Obviously, the throughput is dependent on many factors, including the effects of shear rate and temperature on melt viscosity. There are many factors interacting with each other, and the optimum throughput is a fine balance between all the different factors controlling melt temperature, head pressure, melt viscosity, and shear rate.

4.6 Devolatilization

Figure 4.22 shows a vented extruder with a two-stage screw. The first stage is similar to a standard screw, with feed, transition, and metering section followed by a decompression zone. The second stage has a sharp transition to the final pumping section. Pressure is essentially zero and screw fill is low in the decompression zone, preventing material from flowing out the vent. Vented extruders are normally longer (30:1, 36:1, or higher *L/D*), with the vent port approximately two-thirds down the barrel. In a 30:1 extruder, the first screw stage is approximately 20:1 *L/D*, requiring the same feed, plastication, and conveying functions as a 24:1 or 30:1 extruder. With shorter feed, transition, and/or metering zones, it is impossible to melt the same volume of polymer that can be melted in a longer extruder; consequently, rates in vented extruders are normally reduced unless the extruder has a very high *L/D*.

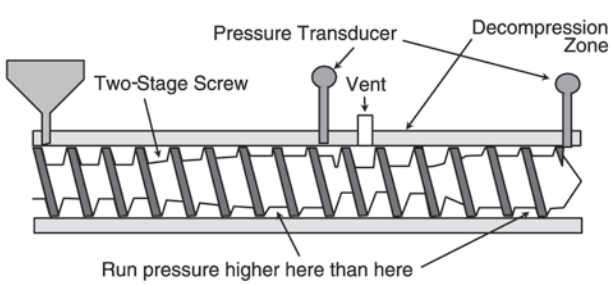


Figure 4.22. Cross section of vented extruder.

The second stage pumps molten polymer as fast as the first stage to ensure resin does not build up in the decompression zone and flow out the vent. Processing conditions (temperature setpoints) can be critical in obtaining the proper balance between the first and second stages to ensure vent flow does not occur. Screw design and particularly metering channel depth in the first and second stages are critical.

To remove volatiles or moisture, a vacuum system is connected to the vent port, lowering the vapor pressure and fostering easier volatile escape. While moisture can be removed at the vent, polymers such as PET, PBT, PC, nylons (PA), other polyesters, etc., will degrade in the presence of moisture and heat. These polymers need to be properly predried prior to extrusion to prevent polymer degradation in the first stage. Consequently, even though the vent can remove moisture, the polymer degradation will already have occurred if the polymers are not properly predried. Polymer degradation results in shorter molecular chains and lower physical properties.

The vent can be used as a liquid injection port to inject liquid colorants or other liquid additives into the extruder. Depending on the screw configuration, uniform additive mixing with the polymer matrix may create a special challenge because of a short second stage. One solution to ensure adequate mixing is to install a static mixer in the transition between the extruder and die. Liquid additive addition downstream alleviates feed problems created by adding liquid to the extruder feed section.

4.7 Die

Some functions the die serves are

- Shapes the melt pumped from the extruder to provide the desired cross sectional dimensions at a specific throughput rate
- Contributes to the physical properties by controlling molecular orientation in the product
- Controls product surface aesthetics

An efficient die provides a specific cross sectional area with tight tolerances, acceptable pressure drop in the die, good surface aesthetics, and good melt homogeneity at high throughput rates. Sophisticated dies with feedback loops sense product variations in the cross sectional dimension, automatically changing the die settings to produce the correct dimensional profile. Dies producing complex cross sectional dimensions require uniform wall thickness and flow in the various die channels to minimize the potential for warpage in the final product.

Proper die construction ensures ease of maintenance, efficient sealing between all elements to prevent polymer leakage, quick connection and disconnection to the extruder, sufficient mechanical strength to minimize deformation under pressure, correct location and sufficient electrical power to provide uniform heat to the die, ability to use the die for more than one polymeric material, and low manufacturing cost. Three zones within the die produce the final cross section. First, the entrance zone, including the screen pack and breaker plate, filters the melt while reducing the spiraling action of the polymer off the screw end and the pressure spikes. Second, the transition zone or adapter changes the extruder circular cross section to a wide assortment of shapes (depending on application), leading up to the die lips. Third, the parallel zone is the location where the melt acquires its final characteristics and shape before exiting the die. The parallel zone or die land area controls to a certain degree the die swell, back or head pressure, and flow uniformity in the extruded part cross section. In an ideal world, the die opening would be flexible enough to alter the profile dimensions and control the melt temperature without requiring a different die. Examples of die adjustments are adjustable mandrels, choker bars, die lip adjustments, and localized temperature controls in different die sections. Die adjustments can help compensate for differences in melt rheology and/or material. In practice, die adjustments are often left to operators. Since this is a critical step in obtaining proper product dimensions, experienced people with both material and equipment understanding are required to produce good parts.

Die lip dimensions are normally different from the product dimensions due to die swell, draw down, and pressure variations across the die. Differences in characteristics (amorphous or semicrystalline) and rheological properties (molecular weight and molecular weight distribution) of similar materials affect the difference in magnitude between the product cross section and the actual die dimensions. *Die swell*, or more properly called extrudate swell, is the actual material swelling as it exits the die (Fig. 4.23). Polymer molecules or chains oriented

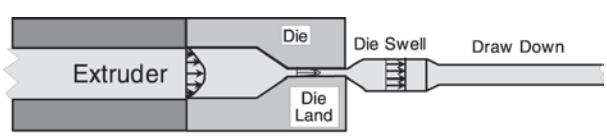


Figure 4.23. Die swell.

in the flow direction in the die land area relax and reentangle once the material exits the die due to the polymer elastic component. This relaxation and reentanglement of polymer chains causes distortion of the extrudate cross sectional area compared to the die cross section. Die swell (extrudate swell) depends on

- Shear rates in the die
- Melt temperature
- Die land length
- Reservoir length

High die shear rates and low melt temperature create more die (extrudate) swell, while longer die land lengths and lower reservoir-to-land length lead to less extrudate swell or die swell. Combinations of these factors can create different conditions, giving the same cross sectional profile. Figure 4.24 shows typical distortion in a square cross section and the die shape required to produce a square rod.

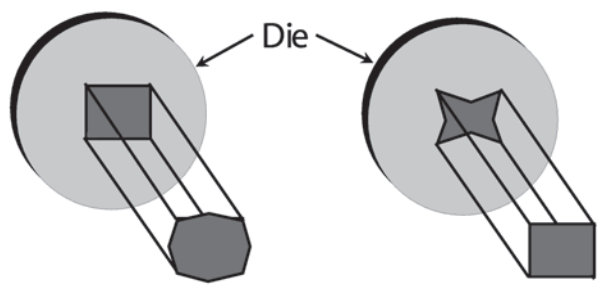


Figure 4.24. Extrudate distortion due to die swell.

Product is pulled away from the die and drawn down to its final dimensions by the puller. When polymer is drooling out the die onto the floor, extrudate swell at the die lips is obvious; however, when the extrudate is pulled away from the extruder, extrudate swell is hidden as the product is necked down by the pulling or drawing operation. The draw depends on the product size exiting the die versus the size required in the final product application. As the draw increases, molecular orientation in the machine direction increases, resulting in higher property (tensile and flexural) performance in the machine direction versus the transverse or cross-machine direction. Depending on the application, high molecular orientation can cause the material to easily split in the machine direction. Extrudate is normally oversized at the die to compensate for the draw between the die and puller. Table 4.3 provides cross sectional guidelines to be used with dif-

Table 4.3. Cross Sectional Guidelines

Polymer	Increase in Orifice Size Relative to Cross Sectional Area of Part, %
Polyethylene	15–20
Rigid PVC	12–15
Flexible PVC	Small Profile: 8–10
Flexible PVC	5–10
Flexible PVC	Large Profile: 3–5
Polystyrene	8–10
Polyamide	15–20

ferent polymers. In dies with adjustable die lips, the die opening is not critical relative to the final product dimensions because the die opening size can be adjusted. On the other hand, if the die is designed to produce a given cross section and no die lip adjustments are available, the die has to be cut correctly the first time; otherwise it is useless and has to be discarded. When cutting new dies, the cross sectional area is normally cut smaller than the required dimensions because it is easier to recut the die opening, increasing the channel size, than to weld steel back to the die and recut to decrease the channel size.

Part of determining die dimensions is understanding polymer shrinkage that occurs on cooling. All polymers shrink when cooled, with semicrystalline materials shrinking more than amorphous polymers. Product dimensions at high temperature are larger than at room temperature. In semicrystalline polymers, shrinkage due to crystallization continues in the solid state if the part temperature is above the glass transition temperature until the equilibrium crystallinity level is obtained. As semicrystalline polymers develop higher crystallinity, polymer molecules pack closer together and the part volume decreases or shrinks. While part shrinkage on cooling may be insignificant in wall thickness dimensions (depending on the tolerance requirements) because the walls are relatively thin, length shrinkage can be quite substantial if the part is not completely cooled to room temperature or completely crystallized prior to cutting. Assume production of a thick PP profile that is 8 feet long; it cools slowly and is cut to length at 120°F (49°C). PP has a coefficient of thermal expansion of 9.0×10^{-5} in/in/°F (5.0×10^{-5} mm/mm/°C). After the part cools to room temperature (assume 72°F [23°C]) from 120°F (49°C), the length has decreased by

$$48°F \times 9.0 \times 10^{-5} \frac{\text{inch}}{\text{inch} - °F} \times 96\,\text{inch} = 0.41\,\text{inch}\,(10\,\text{mm})$$

The part length reaching the customer is 95.59 inches instead of 96 inches (243.8 cm).

A prerequisite for the die is to produce extrudate with a uniform velocity profile throughout the entire cross section. The resistance to flow in the die channels is given by Eq. (4.6):

$$R = \frac{\Delta P}{Q} \tag{4.6}$$

where

R = resistance to flow

ΔP = pressure drop

Q = volumetric flow rate

Changes in melt temperature and die geometry affect the resistance. Uniform resistance to flow can be obtained by modifying the die land lengths in different sections of the die. Balanced polymer flow helps prevent distortions or warpage later in the process due to differential shrinkage. For a specific polymer, only a limited number of die geometries yield the optimized process in terms of mass flow rate, dimensions, and balance of physical properties. Polymer rheology effects altering die performance include

- Melt temperature fluctuations, which affect melt viscosities, flow, and resistance to flow. Hot and/or cold spots in the die can change the flow characteristics in that die section.
- Polymer batch-to-batch variations, particularly if regrind levels are changed or there is a larger variation in the regrind used.
- Stress developed in the polymer during convergence from the reservoir to the land area as a part of extensional flow. (Extension flow is caused by stretching a material while reducing the cross sectional area. In the die, this is associated with stretching a molten polymer.)
- Stresses developed by stretching the extrudate outside the die.

Polymer problems associated with the die are "fisheyes," plate-out, and melt instabilities. These extrusion problems are encountered in all extrusion processes, ranging from compounding, to sheet and film, to profile extrusion. "Fisheyes" are hard polymer specks protruding from the extrudate surface. These specks are high-molecular-weight particles originating from the polymer production or are created in the die by fusion of materials, or in dead spaces where material has long residence time, allowing it to branch or cross-link. Small high-molecular-weight particles from the resin supplier may not melt in the extruder, passing through the screen pack and die. Plate-out or die lip build-up is from low-molecular-weight polymer or additives that migrate to the extrudate surface and are deposited on the die lips. Die build-up needs to be chemically analyzed to determine the source. Low-molecular-weight polymer is in the low-end tail on the polymer molecular weight distribution curve. To eliminate plate-out, a different resin can be purchased with a narrower molecular weight band. If the deposits are additives that volatilize and recondense on the die surface, alternative additives can be used or processing conditions changed (lower temperatures) to minimize the deposits. Regardless of the cause, solutions are required to minimize plate-out in order to maximize product yields and process efficiencies.

Melt instabilities include "sharkskin" and melt fracture. Sharkskin, also referred to as surface matteness, is a repetitious wavy or regular-ridged surface distortion perpendicular to the extrusion direction. Less severe shark-

skin produces a matte finish where a glossy, smooth finish is anticipated. Sharkskin is formed in the die land area due to the rapid acceleration of surface layers in the extrudate as it exits the die. In severe instances, the surface area may actually fracture. Sharkskin appears to be dependent on the melt temperature and linear extrusion speed. Corrective actions to eliminate this problem are increasing the die temperature, reducing the extrusion velocity, or using an external lubricant. High viscosity polymers with narrow molecular weight ranges tend to be the most susceptible to the sharkskin effect. Factors not contributing to this phenomenon are shear rates, die dimensions, approach angle to the die reservoir, die land surface roughness, extruder *L/D* ratio, extruder temperature profiles, and construction materials.

A second melt instability, called melt fracture, is defined as highly distorted extrudate associated with die pressure fluctuations. It is not a surface phenomenon as with sharkskin; instead it goes completely through the extrudate. The surface may remain smooth but the extrudate is distorted. Some severe melt fracture examples, shown in Fig. 4.25, are called spiraling, bambooing, regular ripple, or random fracture, A–D, respectively. While melt fracture, similar to sharkskin, is caused in the die land area, a universal understanding of cause and effect

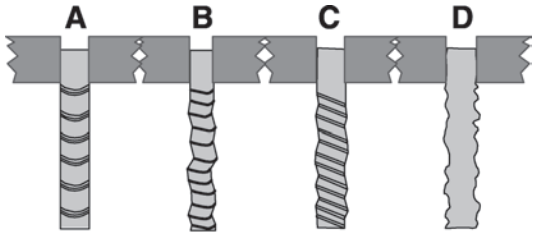


Figure 4.25. Examples of different types of severe melt fracture.

for severe melt fracture varies from polymer to polymer. If a critical wall shear stress in the die is present (15–60 psi), melt fracture is likely to occur. Corrective actions to eliminate melt fracture include

- Streamlining the die
- Slowing the extruder rate
- Reducing the melt viscosity
- Increasing the die land temperature
- Increasing the product cross sectional area
- Adding external lubricants to the formulation

Die design, particularly in the profile industry, has been a trial and error process, with a good die designer able to design and produce an acceptable die in three to four tries, while an inexperienced die designer may require six to eight attempts. Science has been added to die design in recent years with the addition of computer-aided design and finite element analysis (FEA) to design dies and to predict flow uniformity, flow analysis inside and outside of the die, melt pressure, melt temperature, residence time, particle paths, and stress analysis in parts. Computer-aided design is particularly useful in blow molding, coextrusion, and profile applications. Sheet and film dies have a very sophisticated die design for specific thickness, polymeric materials, and throughput rates. Computer-aided die design benefits are

- Reduced design time and elimination of trial and error process
- Reduced testing time
- Applicability to many polymers

Some die design schematics, shown below, include

- Flat sheet or film die (Fig. 4.26)
- Pipe or tubing die (Fig. 4.27)
- Blown film die (Fig. 4.28)

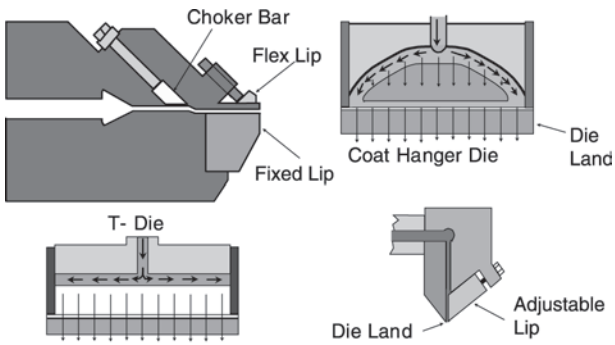


Figure 4.26. Sheet and cast film dies.

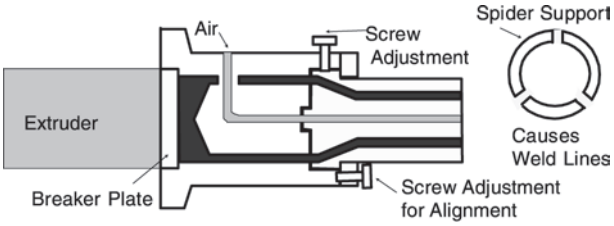


Figure 4.27. Pipe or tubing die.

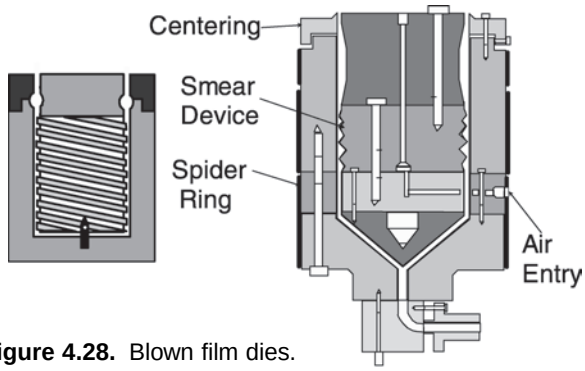


Figure 4.28. Blown film dies.

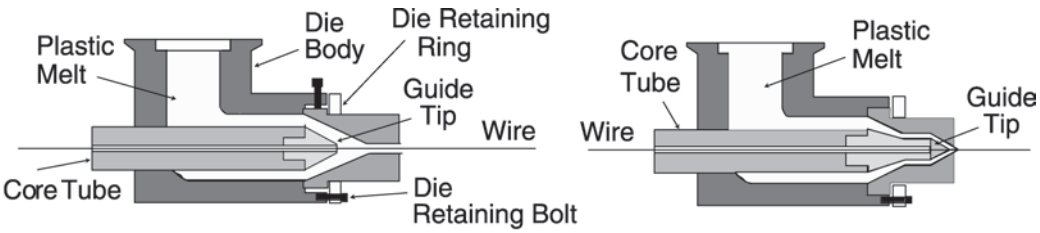


Figure 4.29. Two types of wire coating dies.

- Wire coating die (Fig. 4.29)
- Coextrusion blown film die (Fig. 4.30)
- Profile die (Fig. 4.31)

Every extrusion process matches an extruder output to a particular puller speed to produce products with the correct dimensions. There is only one throughput and puller speed ratio for a particular die opening that produces the correct product. Inherent with that ratio is the molecular orientation, which dictates the polymer properties in the machine and transverse direction. Changing the puller speed or extruder speed independent of each other without changing the die opening alters the product dimensions, orientation, and property profile.

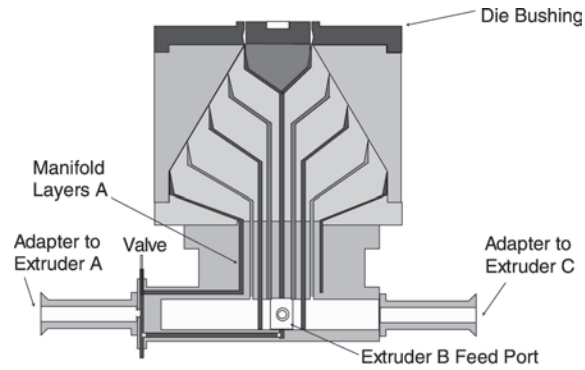


Figure 4.30. Coextrusion blown film die.

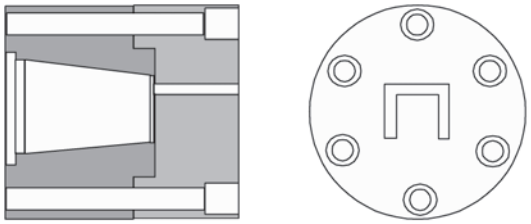


Figure 4.31. Profile die.

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2. Rauwendaal, Chris, *Plastics World*, November 1990, p. 45.
3. Langley, I. L., and Crandall, F., *Fiber Producer*, April 1982, p. 2.

Review Questions

1. How does screw cooling improve extruder efficiency?
2. What are two types of mixing?
3. What mixing elements are used for each mixing?
4. What type of mixing is required for
   - Uniform melt temperature?
   - Blending or alloying two different polymers?
   - Coloring with pigments?
   - Blending fillers?
   - Adding regrind?
   - Adding reinforcing fibers?
   - Adding stabilizers?

Review Questions (continued)

5. What determines the extruder throughput? Describe the flow velocity profile at the end of the metering section.
6. What are the six possible functional zones in an extruder?
7. Why use a two-stage extruder?
8. Describe the feed mechanism in a single screw extruder.
9. Describe the melting mechanism in the transition zone.
10. What are the heat sources to melt plastic, which is the most essential, and why?
11. Describe the melt conveying mechanism.
12. Does increasing the barrel temperature increase the melting capacity? Explain!
13. Explain die swell.
14. Describe melt fracture and its causes.
15. What is sharkskin and what causes it?
16. With a given profile, can the product attributes be changed by changing the extruder speed or puller speed? Explain!
17. Describe four feed problems that might be encountered in single screw extrusion and how to correct them.
18. What causes a melt plug or melt-over?
19. Why is coefficient of friction between the polymer and the barrel wall so critical?
20. What is a static mixer and how does it work?
21. Calculate the volumetric drag flow output rate from a 4.5-inch extruder running at 85 rpm with polystyrene, assuming a square pitch screw with a channel depth of 0.275 inch and flight widths of 0.45 inch. What are the units in your answer? What is the volumetric drag flow in pounds/hour?
22. Explain how a two-stage extruder operates.
23. What are three different areas in the die?

5 Screw Design

The optimum screw design for a particular application depends on the polymer being processed, throughput rate, mixing requirements, venting, and die design. Figure 5.1 shows an extruder cross section and screw, summarizing the extruder, melting mechanism, and polymer processing[1] presented in Chapters 3 and 4. A well-designed screw is critical to optimize productivity, extruder control, and product yields. Improper screw designs can lead to high frequency instabilities caused by melt temperature or pressure variations in the extruder. *High frequency instabilities* are defined as variations in process conditions ranging from a second or two up to 10 seconds or more. While control systems with feedback loops can monitor and make automatic corrections to process conditions for temperature and pressure variations, high frequency instabilities happen so quickly that process control schemes do not have time to respond. Assume the melt temperature varied every five seconds; temperature control feedback loops to barrel heaters cannot respond quickly enough to correct this problem. In addition, the extruder barrel, screw, and polymer system is too massive for temperature changes to occur quickly. Changes between setpoint and actual temperatures depend on longer time frames.

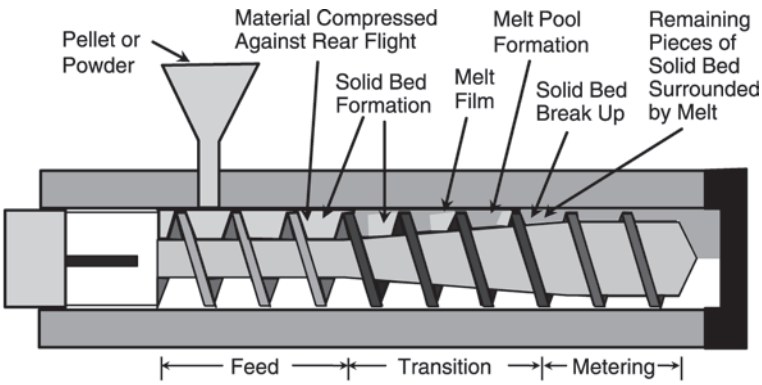

Figure 5.1. Cross section of extruder showing screw elements and polymer processing.

Companies with extrusion operations are interested in purchasing general purpose screws that process all their polymer formulations in the plant at high rates, with good yield, great temperature control, uniform mixing, and pumping capability to generate sufficient and uniform die pressure. The advantage of a general purpose screw is it can produce any product with any polymer and be run at any time on any extruder with minimal supervision, equipment setup, or product attention during production. Unfortunately, in the real world, a screw designed to run high density polyethylene or polypropylene does not work well with polycarbonate, polyethylene terephthalate, or nylon. A general purpose screw designed to process all polymers does not exist. A given screw design may produce product with most polymers; however, the process is not optimized for maximum throughput rate with uniform stable melt temperature and pressure at high throughput rates. If two extruders are available, it is better to produce materials that process well with a particular screw design on one machine and another formulation that processes better with a different screw design on the second extruder. Assume HDPE is run 80% of the time and PC 20% of the time; the screw that produces HDPE at the highest throughput rate with the most stable process and the highest yield is used. When running PC, the screw can be replaced with a screw designed for PC. If this is not practical because the PC run size is too small and/or the time to change the screw is too long, group all the PC production together at one time and make enough material for three to six months to justify the downtime required to pull the screw and replace it with a PC screw. Assuming neither of these options is practical, run PC production at a lower throughput rate, obtaining the most stable process possible with the best yield. There is no such thing as a general purpose screw that processes all polymer combinations at high throughputs with good polymer melt stability in high yields. If the screw is not optimum for the material being processed, it is normally better to run a lower throughput rate with higher first-quality yield than to run at high rates making borderline-acceptable product or scrap.

Several factors to consider when purchasing a new screw include output requirements, venting requirements, feed section geometry, metering section channel depth, formulation being processed, and polymer rheology. Total output capacity is determined more by the extruder diameter, motor torque available, and screw speed limits than by screw design. A properly designed screw allows the maximum screw speed with the motor torque capacity to produce the highest throughput possible. Feed and metering channel depths relate to how much material can be fed and melted in the extruder with the available motor torque. Figure 5.2 shows guidelines for typical pounds/hour that can be anticipated with different size extruders.

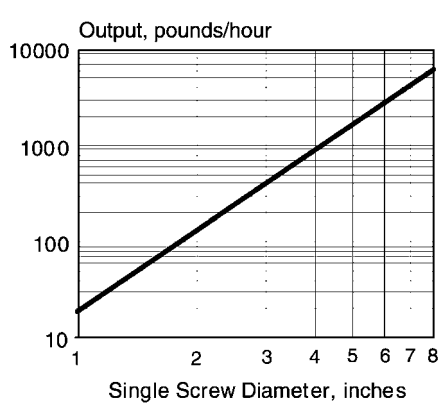

Figure 5.2. Typical extruder throughput rates based on screw diameter.

Most extruders have some differences in motor torque, screw speed, and *L/D*, leading to different throughput capabilities. In 1997, HPM introduced a 2.5-inch diameter extruder combining a high torque motor (500 Hp) with an *L/D* of 50:1 and a 1000 rpm screw speed to produce 2000 pounds/hour. If the screw is designed to melt more material than the extruder motor torque can process, motor horsepower will limit the process output, never allowing the extruder screw to operate at maximum efficiency.

Venting requirements dictate the need for a multistage screw. The vent port location determines the position on the devolatilization section on the screw. Screw design specifications for channel depth in the feed and metering sections in the first stage are based on the motor horsepower, feed type (powder versus pellet), and extruder *L/D* to properly melt a particular polymer at the desired throughput rate. The second stage metering section is designed to provide sufficient pumping capacity to maintain the pressure in the devolatilization zone at zero; otherwise polymer will flow out the vent port.

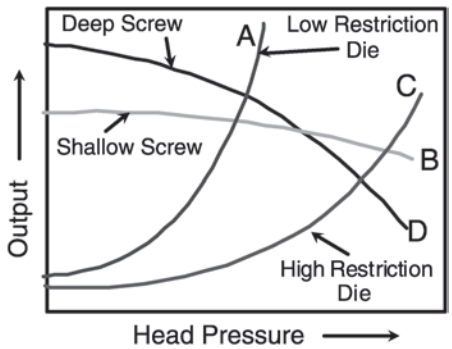

Figure 5.3. Effect of head pressure on output with different die restrictions and metering channel depths.

If a grooved feed section is present, the screw characteristics change considerably. Grooved feed extruders have significantly higher solids conveying rates, generating forces in the feed section as high as 10,000–20,000 psi. Since the additional solid (solid pellets in the grooved channels plus the original pellets in the screw channel) is forced into the screw channel in the feed zone, the screw compression in the transition areas is reduced to ensure the channel volume in the screw metering section can handle the higher melt volume. This leads to screw compression ratios closer to 1.5:1 than 3:1 or 4:1, preventing overfeeding in the metering section. Due to the high solids conveying in the feed section, it is not necessary to have as efficient pumping in the metering section to generate sufficient die pressure. High output rates provide less time for efficient melting and mixing. Therefore, very efficient mixing devices are required to provide a homogeneous melt.

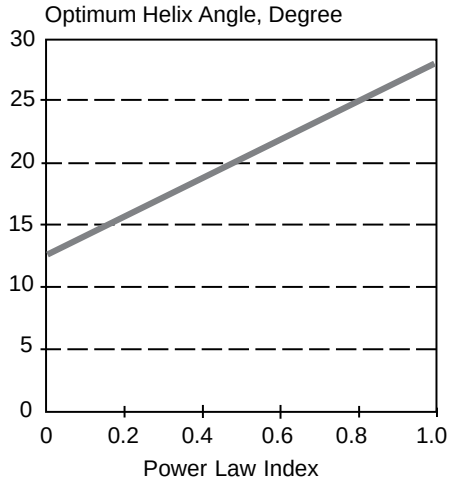

Figure 5.4. Power law index versus optimum helix angle in metering section.

With smooth bore feed sections, the feedstock bulk density is important in determining the feed section channel depth. Bulk densities at least 50% of the actual melt density are easily conveyed with feed channel depths of 0.1–0.2 times the screw diameter and compression ratios of 2:1–3:1. Feed material bulk density less than 50% of the melt or solid density requires deeper feed channels and compression ratios on the order of 3:1–5:1. Bulk densities less than 30% of the melt density may not extrude well without a crammer feeder coupled with a special screw design.

The metering section design depends on the polymer rheological properties and the pressure required in the die. The optimum metering section depth is dependent on matching the die restrictions to the metering channel depth.[2][3][4] Figure 5.3 shows head pressure versus output with different die restrictions using either a deep or shallow metering channel depth. A low restriction die, curve A, has a higher output with a deep metering channel, intersection of curves A and D, compared to a shallow channel depth, intersection of curves A and B. A high restriction die, curve C, has a higher throughput with a shallow channel depth in the screw metering section compared to a deep channel. If a gear pump is used, the extruder screw is required to generate lower pressure, with the gear pump generating the polymer flow and required die pressure. Generally, when a gear pump is used, a deeper channel in the screw metering section generates higher throughput.

Polymer rheology is an important property in determining the appropriate screw design. Melt viscosity is directly proportional to the shear rate. However, all polymer melt viscosities do not respond to shear rate in the same manner. The power law is the simplest representation of pseudoplastic melt flow over several decades of shear. Most plastics exhibit a power law index, *n*, of 0.25–0.90. (See Table 4.2 in Chapter 4.) Low viscosity polymers require a shallow channel depth in the metering section to pump enough material to the die to generate high pressure. Related to the power law index, the optimum helix angle for the screw flights in the metering section is shown in Fig. 5.4. While numerous screws simply use a square pitch with a 17.66° helix angle, an optimum helix angle in the metering section exists for each polymer based on its power law index.

Pressure variations at the extruder end depend on the screw frequency or screw beat. Each time a flight passes the feed throat, polymer flow is interrupted by the flight,

preventing polymer from flowing freely into the channel. Higher screw speeds generate a quicker screw beat. At 120 rpm there are two beats per second, while at 60 rpm only one screw beat per second occurs. Pressure variations also exist between the pushing and trailing flights on the screw. These pressure variations are obvious in a pressure versus time curve measured at the extruder end prior to the breaker plate. (See Fig. 3.44 in Chapter 3.) Melt temperature variations occur in the extruder where high viscous heat generation occurs. Since plastics are great insulators and the extruder barrel and screw mass is so large, it is difficult to remove excess viscous heat from the polymer inside the extruder. Hot spots in the extruder, adapter, or die can contribute to nonuniform melt temperature, leading to differences in viscosity and polymer flow.

When purchasing a new screw with a different design for a current extruder or application, identify the objectives to be accomplished with the new screw that can't be done with the current screw. Do not change the screw design just to change screw designs or because a supplier has approached you with the newest and greatest technology that will solve all your extrusion problems. After outlining the objectives to be accomplished with both the current and new screws, go to the vendor's lab and run tests. Determine new screw performance with the resin system and product currently being used to verify that the predicted extruder performance is improved. A machine shop may be a suitable place to duplicate or repair a screw; it is not necessarily the best place to obtain a radically new screw design. When ordering a new screw, the following variables are specified: extruder diameter, extruder manufacturer (needed to understand extruder *L/D*, shank, and keyway specifications), *L/D*, compression ratio, feed flight depth, screw flights in the feed section, screw flights in the transition section, screw flights in the metering section, single or multistage screw, mixing sections, barrier flights, helix angle, flight surface treatment, and surface coating on screw. A quick reference guide to screw design and some performance factors are given in Table 5.1.[4]

Table 5.1. General Guide to Screw Design

Parameter	Effect on Screw
Output	Calculate screw diameter from chart.
Number of vent or feed ports	0 L/D = 25 1 L/D = 30 2 L/D = 35–40 3 L/D = 45
Barrel	Smooth: Use normal compression ratio. Grooved: Use low compression ratio.
Bulk/actual density ratio	>50% of melt density: Use normal feed depth. <50% of melt density: Use large feed depth.
Melt viscosity	Helix angle from power law index. High viscosity, use a deep metering section. Low viscosity. use shallow metering section.
Melting rate	For higher melting rate, use small clearance, large helix angle, and multiple flights. Can also use internal screw heating.
Melt quality	For higher melt quality, use efficient mixing section.

5.1 Barrier Screw

A so-called "standard" screw with a single flight is a good conveyor of solid pellets in the feed section, provides a uniform melt, and mixes most plastics. Its limitations occur with difficult-to-melt polymers where the screw controls the production rate (can't obtain higher rates because the screw cannot melt enough material), better mixing is required for improved product homogenization, or higher screw speeds generate too much viscous heat. Figure 5.5 shows a standard screw design, and Fig. 5.6 shows the melting model associated with the standard screw. The feed section initiates solid conveying and resin heating. At the beginning of the transition zone,

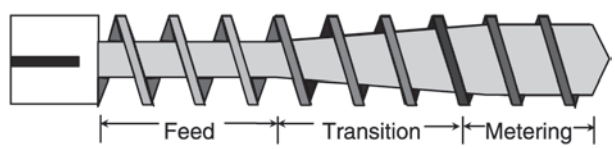

Figure 5.5. Standard single flighted screw.

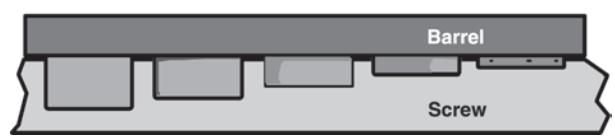

Figure 5.6. Melt model on single flighted screw.

the resin preheated by the feed section is forced against the barrel wall, resulting in sliding friction of the solid against the barrel wall. Combining this friction with the barrel, conductive heating creates a melt film at the barrel surface. In the middle of the transition zone, with additional melting and compression, the polymer is divided into three distinct areas: a compacted solid bed, melt film at the barrel wall, and polymer melt pool. Near the end of the transition zone, the channel depth is still decreasing, melting continues with the solid bed size decreasing, and the possibility of solid bed breakup looms. In the metering section, the plastic is pumped down the shallow channel to the discharge end. Potential problems occur if unmelted pellets are floating in the melt from solid bed breakup. Unmelted pellets in the metering section can cause nonuniform melt, poor mixing of color or other

additives, and inferior properties. Laminar polymer flow in the metering channel does not lead to good mixing.

Melting efficiency can be improved by using multiple flights in the transition section with more flight clearance between the barrel wall and the secondary flight. Barrier flights improve the melting efficiency while eliminating solid bed breakup. Figure 5.7 shows the transition section of a barrier screw with the associated melting model. The

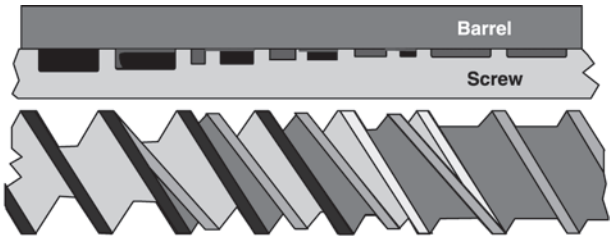

Figure 5.7. Barrier flight in transition section with melt model.

barrier or secondary flight coming off the primary flight is slightly smaller in diameter, so molten polymer flows across the flight and is collected in a separate melt pool in the secondary channel. Solid remaining in the primary channel becomes compressed against the barrel wall, where viscous and conductive heating continues to form new melt film. There is no chance of solid bed breakup, as the entire solid has to be melted and move across the barrier or secondary flight into the new channel before it can exit the transition zone. At the end of the transition zone, the primary flight and its channel completely disappear, and the secondary flight or the barrier flight that formed the new secondary channel in the transition section becomes the only channel in the metering section. All the polymer entering the metering section has to cross the barrier flight in the transition zone, preventing any solid from entering the metering section. Barrier flights in the transition section are designed to improve melting performance.

Figure 5.8 shows a melting model diagram associated with a Spirex Meltpro screw. The advantage of this barrier versus the one shown in Fig. 5.7 is the solid bed width is kept as wide as possible. Melting capacity, discussed in Chapter 4, is directly related to the solid bed width (the wider the solid bed, the higher the melting rate). The solid is compressed in the transition section as

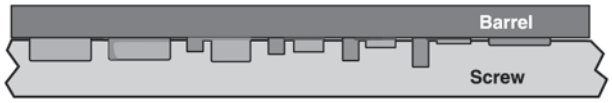

Figure 5.8. Spirex Meltpro screw melting model and transition section channel design.

the channel becomes shallower, forcing the solid against the barrel wall. Melt collects in the secondary flight, which becomes progressively larger as the channel depth in the transition section gets deeper. In the metering zone, the secondary channel formed by the barrier flight is transformed to the normal channel width and depth.

Other transition barrier flights are shown in Fig. 5.9. While numerous barriers are available in the transition section, they are all designed to improve the melting capacity and eliminate solid bed breakup.

Barriers in the metering section are designed to improve mixing, as all melt exiting the extruder must flow across the barrier to provide dispersive mixing. Screws with no barrier in the transition section but having a barrier in the metering section prevent any solid resulting from solid bed breakup from reaching the die. Barriers in

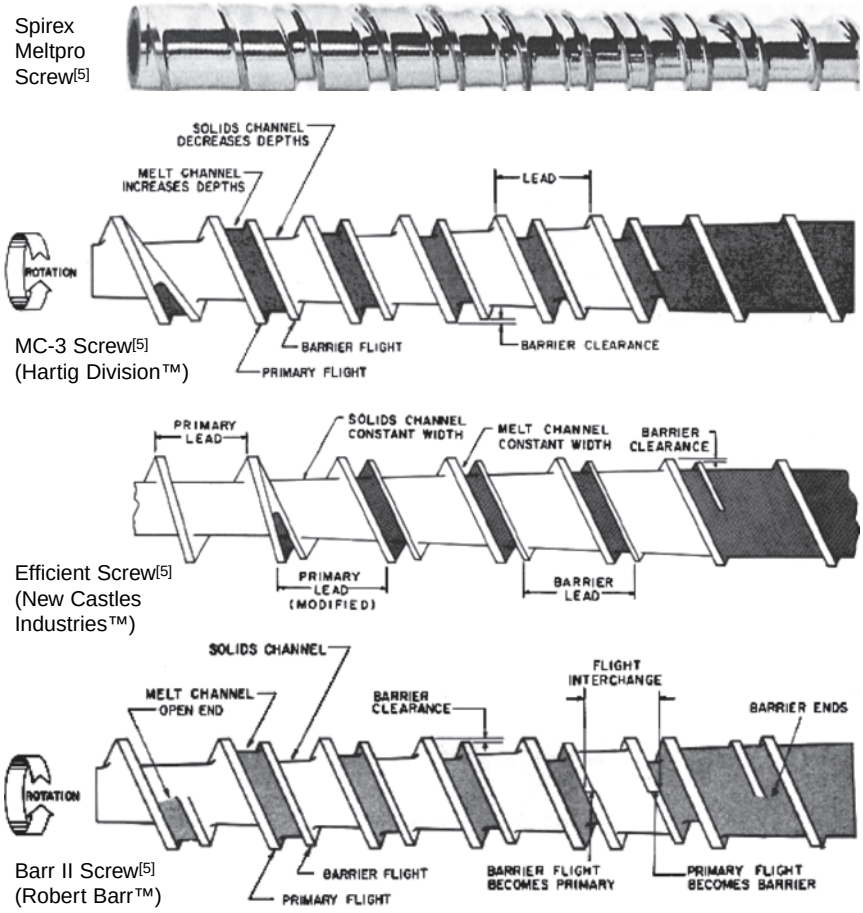

Figure 5.9. Transition section barriers.

Figure 5.10. Barrier in metering section for mixing.

both the transition and metering section are designed to provide efficient melting and good mixing, respectively. Figure 5.10 shows a barrier in the metering section.

5.2 Mixing Screw

There are many mixing sections, patented by different companies making screws to meet the demand for dispersive mixing, distributive mixing, and combinations of both dispersive and distributive mixing. Some screw elements were covered briefly in Chapter 4. The best mixing section to incorporate into a particular screw depends on the process, the rate, and the materials being mixed.

A Pulsar® mixing section from Spirex Corporation is shown in Fig. 5.11. This mixing section is touted to be

Figure 5.11. Pulsar® mixing section from Spirex Corporation.[5]

particularly effective with temperature-sensitive resins such as PVC. The metering channel is divided into alternating sections that are deeper or shallower than the average metering channel depth. Molten polymer is forced from one channel to another as it is pumped through the metering section. Constant directional changes provide excellent mixing and melt uniformity. As with all good mixing sections, there are no dead spots.

A Dulmage mixing section, named after Fred Dulmage of The Dow Chemical Company, is shown in Fig. 5.12. A Dulmage mixing section can be at the screw end or incorporated as an integral part of the screw, normally several

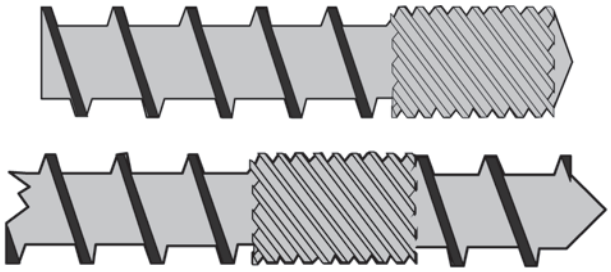

Figure 5.12. Dulmage mixing section.

flights from the discharge end. Multiple Dulmage sections provide discharge and recombination many times to improve distributive mixing. Dulmage mixers are frequently used in foam processing.

A Flex flight mixing screw patented by Spirex Corporation is useful for both distributive and dispersive mixing. A barrier flight that is not parallel to the primary flight is introduced in the metering section (Fig. 5.13). The barrier creates two channels of constantly changing width within the original channel width. Molten resin is squeezed over the barrier as the channel width decreases into the wider channel, causing a tumbling and mixing action. Optimization for different resins is obtained by

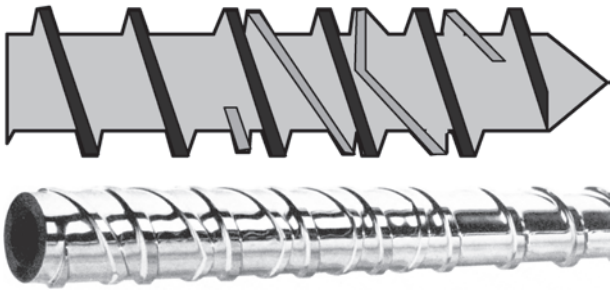

Figure 5.13. Spirex Flex flight mixing section.[5]

changing the clearance between the barrier flight and the barrel wall. The Flex flight eliminates the high shear area normally associated with a barrier screw, because the primary flight never ends in the transition to a secondary channel or barrier flight. One distinct advantage of the Flex flight is it can be added to an existing screw by welding the flight into the metering section.

Mixing pins were one of the first distributive mixers due to the ease of inserting metal dowel pins radially in the root of an existing screw in a plant environment. Screws did not have to be sent to a fabricator or machine shop. The second advantage is the processing effect on the product can be determined almost immediately. Pins interrupt the laminar flow in the channel by dividing and recombining the melt. Typical pin arrangement may have a row of pins at the entrance to the metering section, a second group halfway through the metering section length, and finally a row of pins one flight before the screw end. Figure 5.14 shows a screw with three sets of pins in the metering section.

A Maddock mixer, developed by Bruce Maddock of Union Carbide Corporation, is also called a Union Carbide

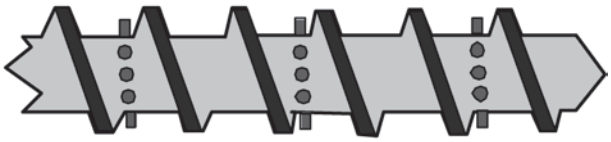

Figure 5.14. Pin mixers.

mixer or Leroy mixer. The mixer is a series of grooves parallel to the screw length that are open on alternating ends to the feed and discharge. Melt is pumped from the metering section into one groove, where it passes over a barrier and exits from the discharge groove toward the die. The land between the entrance and discharge groove is slightly lower than the screw diameter, while the land area between the discharge groove and the next entrance groove is the same diameter as the screw. The barrier lands are called mixing or discharge lands, while the land between the discharge and next entrance land is called the wiping land, as it wipes the barrel clean of polymer. Polymer going across the discharge land is subject to high shear, generating significant viscous heat as the polymer is mixed. Figure 5.15 shows a Maddock mixer at the screw end. Frequently, a Maddock mixer may be incorporated a couple of flights from the screw end to provide time to dissipate the shear heat generated by the polymer crossing the discharge land.

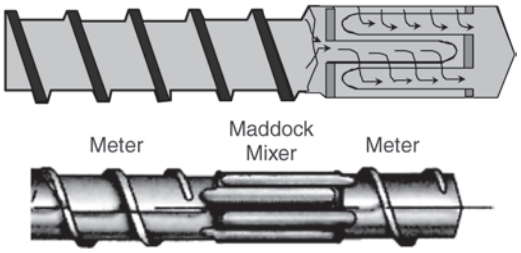

Figure 5.15. Maddock mixer at end of screw and in metering section.[6]

Maddock mixing sections also come in Spiral Maddock, where the channels rotate partway around the screw rather than being parallel to the screw axis, as shown in Fig. 5.15.

A Saxton mixer is a good distributive mixer developed by Ronald Saxton in 1961. A melt channel going in the opposite direction interrupts secondary flights with a different helix angle than the primary flight. The dividing and recombining of the melt stream make this an efficient distributive mixer. Flighted interruptions do prevent positive forward conveyance of the melt stream. The total number of minor flights can be modified to vary the mixing for different applications. Figure 5.16 shows two drawings of a Saxton mixer.

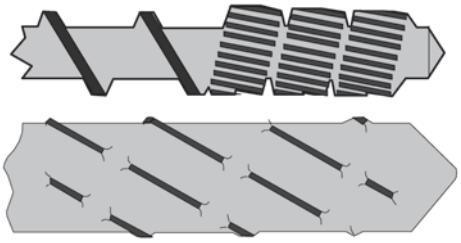

Figure 5.16. Saxton mixers.[5]

Double and triple wave screws are patented by HPM Corporation. The double and triple waves have two and three equal channel widths, respectively, within the main channel, separated by a barrier flight between the channels. Melt is forced back and forth across the barrier by the deep and shallow section in a particular channel, making up the wave. The wave can be in the transition section to improve the melting efficiency and/or in the metering section to improve both the distributive and dispersive mixing.[7] In the metering section, good melt homogenization and mixing occur, with the compressing wave forcing melt across the barrier into the neighboring channel. When the second channel wave compresses the melt, it is forced back across the barrier. This action reduces the work on the melt, resulting in better temperature control. Wave screws were originally designed for rigid PVC, but have found use in all applications. Figure 5.17 shows a double wave screw. A double or triple wave through the transition and metering section can be combined with a dispersive mixer, e.g., a Maddock, near the screw end to provide a higher mixing if required by the application.

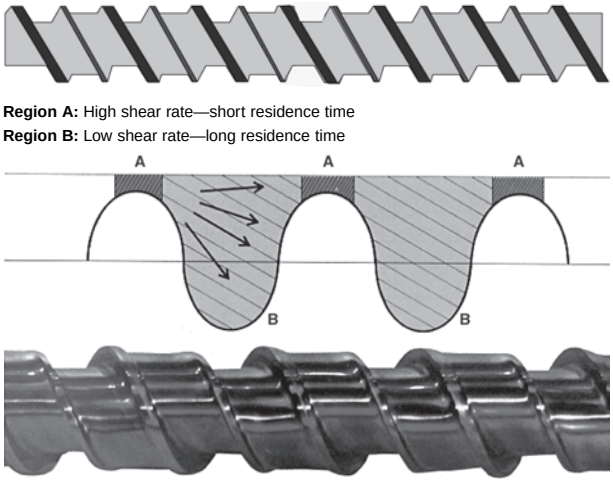

Figure 5.17. Double wave screw from HPM.[7]

A Spirex Z mixer, shown in Fig. 5.18,[8] has primary flights interrupted by shallow plateaus that force melt up and down between shallow and deep valleys in the metering section. Interruptions in the primary flights allow the polymer to flow from one primary valley to another, providing distributive mixing among the many valleys.

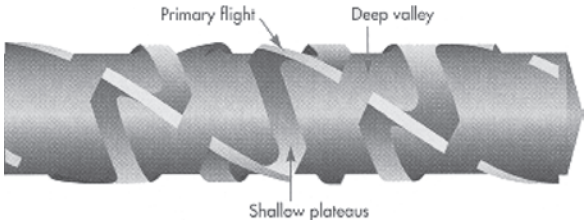

Figure 5.18. Spirex Corporation Z mixer.

A Barr E.T.® screw (Fig. 5.19) has a patented low shear mixing geometry in the metering section. Approximately 70 to 80% of the polymer is melted by high shear,

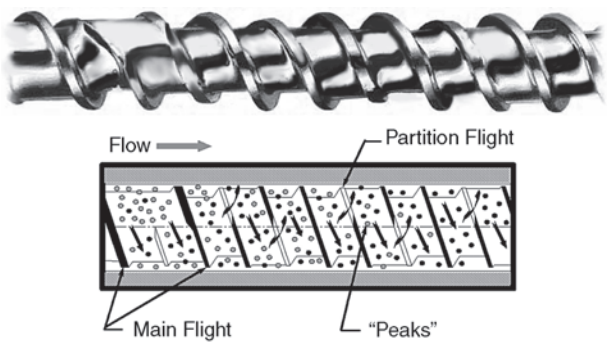

Figure 5.19. Barr E.T.® melting section.

and the remaining solid is melted in the E.T.® section. Utilizing the natural drag force of screw rotation, the remaining pellets are mixed with the hotter melt. Heat conducted from the hotter melt to the remaining pellets completes the melting process at low shear rates. This allows lower melt temperature and higher energy efficiency.

Numerous other mixing elements and screw types are available from screw and extruder suppliers. Each supplier has done internal studies and development work to define their "optimum screw design." The optimum screw for your plant depends on the polymer being processed, die design, melt temperature, throughput rate, and melt pressure. Trials with new screw designs are recommended before purchase to verify that the screw design selected meets your processing demands.

New screw designs are modeled by computer to predict the output rate, melt temperature, and die pressure. The appropriate screw design for your application depends on past experiences, experimental trials to support computer modeling, output stability tests, melt limitations, and extruder venting. It is possible to process only the polymer that can be effectively melted in the transition section. If the transition section is too short, excessive shear heat can be generated as the screw overfeeds the metering section because the entire polymer is not melted. Polymer factors affecting screw design are melt viscosity, melting rate (determined by heat capacity and thermal conductivity), solids conveying rate (friction factor), lubricants in the polymer or formulation, material flow in the feed hopper, bulk density, polymer temperature limitations, and any chemical or physical changes occurring during processing.

5.3 Screw Wear

There are many factors affecting screw and barrel wear:[5][9][10][11][12]

- Cold-starting the extruder screw
- Running the screw with no plastic present
- Processing the wrong material (corrosive polymer on a screw not designed for corrosive material)
- Nonuniform barrel heating
- Poor barrel support
- Improper screw, barrel, and drive alignment
- Screw straightness
- Abrasive additives or reinforcements in the polymer formulation
- Wrong screw or barrel materials
- Excessive die weight on the end of the barrel
- Corrosion caused by polymer and/or additive degradation

The three types of screw wear are abrasive, corrosive, and adhesive. Abrasive wear is caused by continuous contact of hard abrasive particles (e.g., glass fibers, glass-filled materials, silica, calcium carbonate, talc, flame retardants) with the screw at high temperature and pressure. Particle size, shape, hardness, and loading directly correlate to the degree and rate of abrasive wear. Screw hardness is critical to minimize abrasive wear. Glass fibers wear the screw root at the pushing flight edge in the transition and feed sections. Corrosive wear is caused by chemical attack of any component in the formulation to the screw root, resulting in a pitted surface. Chemicals attacking the screw root can come from polymer degradation, flame retardant additives, fluorinated polymers, etc. Normally chemical attack is in the metering section, where the formulation is at its highest temperature for the longest time, creating a higher possibility of degradation products for chemical attack. Chemical attack may remove coatings designed to minimize abrasive wear and eat into the screw base material. Adhesive attack results from metal-metal contact leading to galling. Screw rotation can cause momentary contact between the screw flight and the barrel wall at high pressures, resulting in a momentary weld between the two surfaces. As the weld separates, metal is removed. This metal can slide on the screw root, causing abrasive wear of the screw root. Proper extruder alignment along with compatible barrel and screw materials can virtually eliminate adhesive wear between the barrel and screw flights.

Minimizing wear depends on identifying the wearing mechanism and eliminating the root cause. Screw surface modification to minimize abrasive wear is accomplished by welding harder materials to the screw surface, surface treatment of the base material, plating with nickel or chrome, or spraying hard alloys onto the surface. Base materials used for most screws include

- Alloy steels (AISI 4140, AISI 4340, and Nitralloy 135-M)
- Tool steel (CPM-10V, CPM-T440V, CPM-9V, D-2, and H-13)
- Stainless steel (304, 316, 17-4 PH and 15-5 PH)
- Specialty high temperature metals (Duranickel 301, Inconel 718, and Hastelloy C-276)

Table 5.2. Comparison of Surface Treatments[10]

Method	Hardness (Rockwell C)	Hardness Depth (Inches)	Cost	Abrasive Resistance	Corrosion Resistance	Adhesive Resistance
Nitriding	60–70	0.010–0.030	C	8	5	7
Dynablue	65–70	0.005–0.010	C	10	6	8
Ion Nitriding	>70	0.010–0.030	1.75C	10	6	8
Flame Hardening	50–55	0.015–0.250	0.5C	5	4	5
Age Hardening	35–45	Fully Hard	4C	8	8	8

Comparison number based on 10 = best, 1 = worst; C = average cost of nitriding.

Many alloys can be welded to the flights to prevent or reduce flight wear. With very abrasive fillers, e.g., glass fibers, cost justification exists to weld hard materials to the screw root and sides of flights in the feed and transition zones. Due to the nature of adhesive wear, hard surface coated materials (e.g., Stellite-6, Stellite-12, Colmonoy 6, Colmonoy 56, Colmonoy 83, Colmonoy 88, Coltung 1, Certanium 27B, and Stody 101HC) are welded to the top of the screw flights to form a wear-resistant surface in contact with the extruder barrel.

Surface treatments coat the entire screw, not just the top of the flights. Surface treatments improve both wear and corrosion resistance, but they are not as effective in treating wear as welding hardened alloys to the surface. The driving factor is cost, with surface treatment being significantly less expensive than welding alloys to screw flights or surfaces. Surface treatments include nitriding, flame hardening, age hardening, Dynablue, and ion nitriding. Table 5.2 compares different surface treatments.

Screw surface plating is done with chrome, nickel, or Nye-Carb® (silicon carbide particles suspended in a nickel-phosphorous matrix) to reduce corrosion or improve processing. Chrome is the most common plating material and is deposited on the screw root and the flight sides from 0.001–0.0015 inch thick. Wear resistance improvement from chrome is minimal; most chrome plating is done to facilitate screw cleaning. Corrosion resistance is not significantly improved because chrome has micro-porosity. Nickel plating provides about the same results as chrome, with the exception of improved corrosion resistance to fluoropolymers. Nickel is normally applied using the electroless nickel process, providing a denser coating and better corrosion resistance than plating. Electro Coating Inc. of Houston, Texas, patented the Nye-Carb process, which works with small diameter screws.

Spraying hard surfacing alloys onto the screw root and sides of the flights works well with new screws but is not economical to do with rebuilt screws because the entire initial surfacing alloy must be removed. Many alloys can be sprayed onto the screw, including U-car, high velocity oxy-fuel (HVOF), and powder flame spray. U-car is a patented process from Union Carbide Corporation where different alloys are applied to the screw root with a D-gun that shoots the powder alloys at hypersonic speed, producing bonding on impact. The HVOF process uses fuel gas and oxygen combustion to accelerate powder at supersonic speeds prior to impinging on the screw root. Powder flame spray introduces the alloy powder into a burning oxygen-fuel gas stream, where the heated particles are carried by flame to the screw root. While this is the most common method used for spraying hard alloys onto the screw root, it is inferior to either U-car or HVOF because the bond to the screw root is not as strong.

Table 5.3 shows some common polymers with their abrasive and corrosive characteristics toward screws and

Table 5.3. Polymer Characteristics and Screw Materials

Polymer	Abrasiveness	Corrosiveness	Base Material	Flight Hardening Material
Nylon 6, LDPE	Soft	Not Corrosive	AISI 4140	Stellite 12, Flame Hardened to 50 RC
Flexible PVC	Soft	Medium	AISI 4140	Stellite 12
FEP	Soft	High	Inconel 718; Hastelloy C-276; AISI 4140	Stellite 12, Age Hardened to 39–42 RC
HDPE, PP, GP PS	Medium	Not Corrosive	AISI 4140	Colomony 56
Rigid PVC	Medium	Medium	AISI 4140; Nitralloy 135M	Colomony 56
High Loadings FR	Medium	High	Inconel 718; Hastelloy C-276	Colomony 56, Age Hardened to 39–43 RC
Glass Filled	High	Not Corrosive	CPM 9V/10V; AISI D2	Heat Treated to 39–43 RC
Clay Filled PVC	High	Medium	Nitralloy 135M	Coltung 1
PTFE/TFE Filled	High	High	AISI 4140	Coltung 1

barrels, recommended screw flight material, and hard flight facing. Screw and barrel materials need to be matched to obtain optimum wear resistance. The correct screw material combination for a given polymer may not be the same when compounding glass fibers into that same polymer matrix.

5.3.1 Worn Screws

A new screw needs to be benchmarked with a given resin formulation when it is first installed in the extruder. Generate graphs that show how screw speed with a given temperature profile affects throughput rate, melt temperatures, head pressure generated with a standard die configuration, motor load, and melt stability over time. Periodically, the screw needs to be benchmarked with the same resin formulation and extruder configuration to determine if any process variations from the original parameters are occurring due to screw wear. Just measuring screw diameter for wear is not sufficient. Minor wear may actually produce higher output and better operating conditions. While throughput reduction is the primary result of screw wear, melt and pressure stability may be as critical in producing repeatable high tolerance products. If the melt pressure and temperature vary as a result of screw wear, output stability will deviate from the current mean, resulting in higher standard deviation associated with product dimensions. A process that was in control, based on SPC data and 3 or 6 sigma capability, may become out of control, or the process deviations may produce more data near the upper or lower control limits. Over time, a worn screw or barrel may turn a stable process into a process that is just barely capable, or in some cases a process that is incapable of producing the product within SPC limits.

Figure 5.20 shows the proper way to measure screw diameter wear with a micrometer and gauge block. Screw root diameter can be measured directly with the micrometer across the root. A bent screw is sometimes difficult to determine. Place the screw across a flat granite table and measure the space beneath the screw as it is rolled across the table. If the screw is bent and raised off the granite table, the bending can be measured with a feeler gauge. Verify that the distance between the screw and table is actually the screw bending and not screw flight wear.

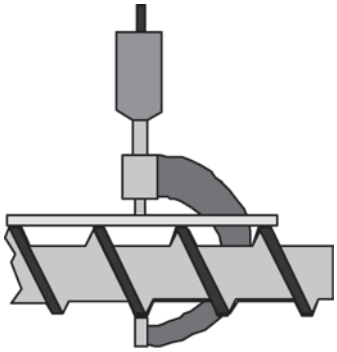

Figure 5.20. Measuring screw wear with gauge block and micrometer.

5.4 Screw Compression Ratios

Table 5.4 lists recommended screw compression ratios for numerous commercial polymers.

Table 5.4. Typical Screw Compression Ratios for Commercial Polymers

Material	Compression Rate
ABS	2.75:1
Nylon 6	3.9:1
Nylon 6,6	3.6:1
LDPE	3.5:1
LLDPE	Variable Pitch
HDPE	3.0:1
PMMA	1.8:1
PP	3.0:1
Flexible PVC	3.0:1
Rigid PVC	2.5:1
HIPS	2.5:1
PC	2.25:1
PET	3.25:1
PBT	2.5:1
Noryl®	2.1:1
Ultem®	2.1:1

5.5 Screw Performance[13]

There is no such thing as a general purpose screw that processes all materials at peak efficiency. Some general purpose screws may run one polymer or formulation at optimum rates with good melt stability, while the second polymer must be run at lower rates to produce an acceptable product. Work done by Ed Steward while at Davis-Standard[1][3] shows how different screw characteristics effect output rate and melt temperature. The second part of his study quantified the effect of five different barrier screws and two single flighted screws equipped with a Maddock mixer and mixing pins, respectively, on throughput rates and pressure stability with HDPE and high impact polystyrene (HIPS). The results in Figs. 5.21 and 5.22 show output and pressure stability of a 2.5-inch extruder run at 150 rpm. Based on the data presented, the best screw design (best balance between throughput and pressure variation) for HDPE is B1, while the best screw for HIPS is B2.

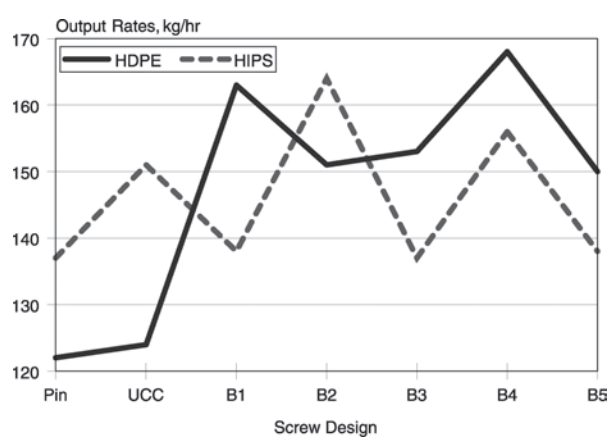

Figure 5.21. Screw design versus output.

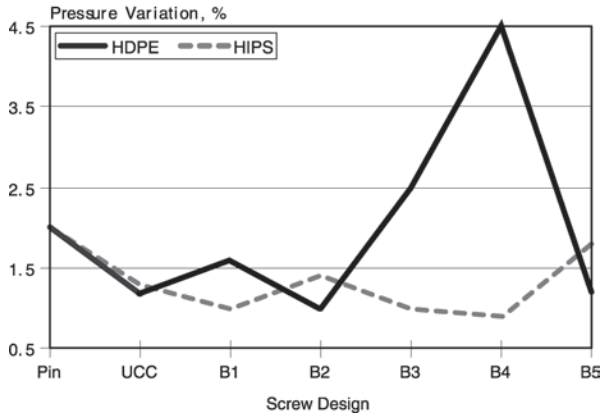

Figure 5.22. Screw design versus pressure stability.

B4 yields the highest throughput with HDPE but also the largest pressure variation. Using B4 with HIPS yields a high throughput and very low pressure variation. Selecting one screw design to process both materials, the best choice is probably B2, even though the HDPE throughput is lower. However, if the extrusion line processed 70% HDPE and 30% HIPS, the optimum screw design compromise is probably B1, due to the higher throughput and better melt stability process with HDPE.

5.6 Summary

Screw design and screw selection depend on the following criteria:

- Polymer to be processed
- Throughput rate
- Output stability (melt pressure and temperature stability) required for the application
- Pressure development in the die
- Polymer melt temperature limitations
- Mixing efficiency and type required for the resin formulation being processed
- Power usage

Screw design selection determines the overall extrusion process capability. A poor screw design or worn screw costs money in both performance and efficiency. Payback time for a new screw or properly designed screw might be significant. Since an overall general purpose screw does not exist to process all polymers, changing screws from one job to the next may make economic sense, depending on

- Time required to change the screw
- Extruder size (dictates time to change screw)
- Cleanup required between screw changes
- Production run size versus improved efficiency with a different screw

If more than one product is processed on an extruder, screw design will always be a compromise. The optimum screw design is the one that processes several polymers at high throughput, good melt stability, and maximum yield while allowing any other formulations to be processed under stable conditions with high yields.

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Review Questions

1. What parameters need to be specified when ordering a screw?
2. Where is the barrier located on a barrier screw and why?
3. What is the typical compression ratio of a screw used with a grooved feed throat and why?
4. What are the two types of mixing screws?
5. What causes screw or barrel wear?
6. Explain the relationship of throughput to screw channel depth in the metering section of the screw and the effect of die restriction.
7. Why purchase a new screw or new screw design?
8. Explain how a barrier in the transition section works, its function, and in your opinion the best barrier design and why.
9. What mixers do you use for distributive mixing?
10. What mixers work both for distributive and dispersive mixing?
11. Are there any mixers that are designed just for dispersive mixing, and if so what are they?
12. What are some screw hardening techniques and surface hardening procedures?
13. How do you measure screw and barrel wear?

6 Processing Conditions

This chapter covers extruder temperature profiles, setup, start-up, steady-state operation, and shutdown procedures.

6.1 Extruder Temperature Profiles

What is the best method to set the optimum extruder temperature profile? Options are available to assist in selecting the best temperature profile, ranging from recommendations by the raw material supplier to trial and error methods in a production environment. With products made previously, experience or documented SOPs dictate the best process and conditions to use. However, extruding new products the first time can create a dilemma. Material suppliers provide recommended temperature profiles, optimum screw design, proper drying conditions, recommended extruder *L/D,* appropriate die designs and land lengths, and suitable draw ratios. Some equipment parameters are fixed, such as the screw design based on the screws in the extruder or those available in-house; the draw ratio is set based on the die opening and the final cross sectional dimension requirements; and the extruder *L/D* is fixed. The barrel and die temperature profile and screw rpm are varied until the product quality and output rates meet acceptable standards. During process optimization, processing conditions are recorded with descriptions of the product quality to establish the best relationships between processing temperature, screw speed, and throughput rates with product performance and appearance. A design of experiments (DOE) can be employed to optimize the processing conditions. Once equilibrium processing conditions are obtained, all operating conditions, setpoints, and process observations are recorded as baseline data for future production runs.

The possible barrel temperature combinations available to operate the extruder are almost limitless. However, only some temperature profiles will produce an acceptable product in high yield under optimum process conditions. Given the melting mechanism and the extrusion objectives to melt all the polymer in the transition section, while providing a well-mixed, homogeneous melt in the die at both the correct and constant pressure and temperature required for the material, only a few temperature profiles provide the optimum extruder performance. Possible temperature profiles include

- A progressive or increasing temperature profile with the setpoints increasing continually from the feed throat to the die
- An inverted or decreasing temperature profile with the setpoints decreasing from zone 1 and to the die
- A straight temperature profile where all temperature zones are set at exactly the same temperature setpoint
- A humped profile where the temperature is lowest in zone 1, gradually increases toward the middle, and then decreases progressively toward the die

The temperature profile that works best with a given product and extruder setup depends on the resin type and viscosity, screw design, and the throughput rate. Polymer melt temperature is a critical property in controlling the extrusion process and optimizing throughput, while minimizing resin degradation, based on the screw design limitations.

An increasing or progressive temperature profile increases the temperature across the extruder, with zone 1 having the lowest temperature and the last zone or the die the highest. Low temperature in zone 1 prevents premature melting and minimizes melt plug formation around the screw or bridging in the feed throat, resulting from polymers or additives becoming sticky and adhering to the feed throat wall. Zone 1, as discussed previously, is an extremely critical temperature zone for good operation and feed. If zone 1 is too hot, in addition to premature melting or melt plug formation, a melt film produced on the barrel wall reduces the coefficient of friction between the barrel wall and polymer, possibly leading to poor feeding characteristics and lower throughput. If zone 1 is too cold, polymer will not stick together and move as a plug, polymer may not adhere to the barrel wall, or the resin may not be sufficiently preheated going into the transition section. Zone 1 temperature controls resin feed characteristics and to some degree surging in the extruder. Assuming zones 2 and 3 are the transition section, the objective is to melt all the resin in these zones. Consequently, zone 2 temperature is set higher than zone 1 and zone 3 higher than zone 2 to assist the melting mechanism with conductive heat from the barrel. At the end of zone 3, all the polymer is theoretically melted, and temperature setpoints are very close to the desired melt temperature. The setpoint versus actual zone temperatures is the extruder's method of communicating with the operator what is occurring in each extruder zone. If the actual temperature is significantly overriding the setpoint temperature, the screw in that particular zone is generating extra shear heat. Overheating may occur because the quantity of resin and additives being processed is inappropriate for the screw design; because of the polymer type; because of a high shear area in the screw; and/or because the temperature setpoints are too low, creating an overfeeding condition in a particular extruder zone. Assume the zone 3 actual temperature is significantly higher than the setpoint; signifi-

cant viscous heat is being generated by the screw rotation in this section. To compensate for the viscous heat generation, the temperature is increased in zone 3, zone 2, and possibly zone 1 to provide more conductive heating to the polymer approaching zone 3. If melting is not completed by the end of the transition section, solid bed breakup occurs, with the melting completed in the metering section. Because the metering section is designed to pump and convey polymer, not to melt it, overheating may occur in zone 4 as the screw is trying to compress and force unmelted resin into the shallow metering channel. This can lead to screw burning, resulting in the screw turning blue from overheating. Depending on the mixing section and the viscous shear heat generated, setpoint and actual temperatures can differ in the mixing area. The die zone may be set lower than the last extruder temperature zone to remove heat from the melt, to improve the polymer melt strength at the die exit, and to increase the die pressure.

An inverted temperature profile has a high setpoint temperature in zone 1 with a progressively decreasing temperature profile across the extruder toward the die. Normally this is not a desired temperature profile, as it encourages melt plugs in the first zone and does not use the extruder motor and drive power efficiently to melt the polymer. However, if the wrong screw design is in the extruder for the polymer being processed, it may be necessary to use an inverted profile to produce a quality product. One example of this is running PBT on a 3.5:1 compression ratio screw equipped with a Maddock mixing section designed for processing PP. PP is processed efficiently at high rates with a humped or increasing temperature profile, using the screw described. However, PBT runs best on a 2.5:1 compression ratio screw with no Maddock mixing section. To efficiently run PBT on this particular PP screw, an inverted temperature profile is required with low screw rpm.[1] Otherwise, PBT degradation occurs in the Maddock mixing section.

A flat temperature profile (all zones are set at the same temperature) does not take advantage of the temperature control capability on the extruder. Usually the feed end is too hot and the metering section too cold. This setup does not provide good melt temperature control, as the metering end is constantly trying to remove excess temperature from the process. A flat temperature profile does not encourage melting in the transition section.

A humped temperature profile has zone 1 temperature low, the transition section zones set higher to encourage the melting process, and the metering zones are flat or slightly lower in temperature to prevent resin degradation. Once the resin is melted, it is not necessary to add additional heat to lower the viscosity. Set the extruder so the metering zones are controlling the desired melt temperature and not just constantly cooling or removing heat. Some processes, like extrusion lamination, require very hot polymer to promote adhesion.

One way to optimize the temperature profile is discussed below[2]:

- Set zone 1 temperature 50°F (28°C) below the resin melting point with all other zones at their normal setting.
- Process resin through the extruder until the equilibrium is attained.
- At typical production screw speed, measure the extruder throughput rate.
- Increase zone 1 temperature 50°F (28°C), allow the extruder to come to equilibrium, and measure the throughput rate.
- Continue to increase the temperature in 50°F (28°C) increments until a plot of throughput rate versus zone temperature similar to Fig. 6.1 is produced.
- After zone 1 is set, follow the same procedure for the other zones.

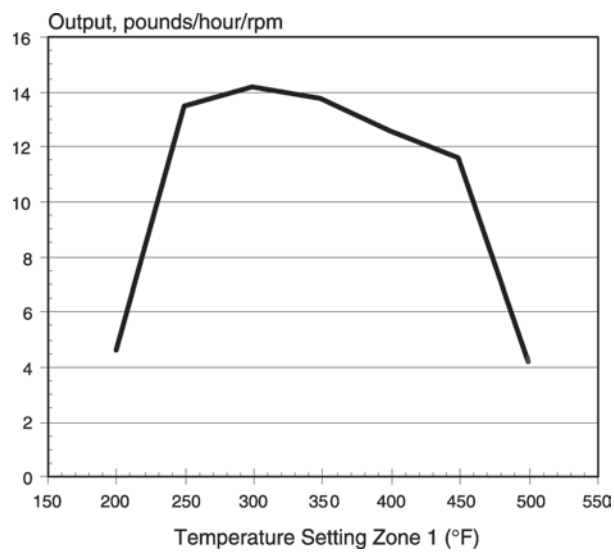

Figure 6.1. Throughput versus zone temperature.

The high point of the curve corresponds to the maximum throughput. Based on the processing conditions and resin used to generate Fig. 6.1, the optimum zone 1 temperature is approximately 300°F (144°C). The low temperature (200°F or 93°C), low throughput point is caused by resin sliding on the screw root and not sticking to the barrel wall because the barrel wall temperature is too low. At the other extreme, high temperature (500°F or 260°C), the melt film produced by the high barrel temperature in zone 1 acts as a lubricant, thereby lowering the coefficient of friction between the barrel wall and the resin, leading to lower throughput. Zone 1 temperature has to be properly set to encourage the polymer to stick to the barrel wall (higher friction) and slip on the screw root. Once the zone 1 temperature is properly established, use the same

procedure for zones 2, 3, 4, etc. After optimizing all temperature zones, go back and redo zone 1 to determine if the optimum temperature has changed as a result of downstream temperature changes. This procedure finds a useful temperature profile with a particular resin, screw design, and screw rpm to maximize the throughput rate.

Screw speeds are normally run near but not at the maximum extruder speed to optimize the throughput rate. It may be impossible to run the screw speed near maximum rpm and produce good product if the melting is not completed in the transition section or the barrier or other dispersive mixer generates too much heat. It is better to run at a slower screw speed with 100% good product than to run at a higher rate with an 80% quality yield. Excessive screw speed can result in overfeeding, leading to either resin degradation from too much shear heat or incomplete melting and mixing.

Table 6.1 gives "standard" processing conditions and screw compression ratios for various commercial resins. Depending on the resin modification (flame retardant additives, impact modifiers, fillers and reinforcements, UV stabilizers, etc.), extruder *L/D*, die design, screw design, etc., the actual processing conditions may vary from those recommended.

Table 6.1. Setpoints for Various Resin Systems

Material	Screw Compression Ratio	Feed Zone, °F (°C)	Transition Zone, °F (°C)	Metering Zone, °F (°C)	Die Zone, °F (°C)
ABS	2.75:1	400° (204°)	425° (219°)	440° (227°)	460° (238°)
Nylon 6	3.9:1	420° (216°)	460° (238°)	480° (249°)	500° (260°)
Nylon 6,6	3.6:1	530° (277°)	535° (280°)	545° (285°)	540° (282°)
LDPE	3.5:1	340° (171°)	355° (180°)	365° (185°)	375° (191°)
LLDPE		300° (149°)	325° (163°)	364° (185°)	410° (210°)
HDPE	3:1	340° (171°)	380° (193°)	400° (204°)	400° (204°)
PP	3:1	375° (190°)	410° (210°)	430° (221°)	430° (221°)
Polystyrene	3:1	350° (177°)	400° (204°)	440° (227°)	450° (232°)
HIPS	2.5:1	375° (191°)	420° (216°)	450° (232°)	450° (232°)
PMMA	1.8:1	360° (182°)	400° (204°)	430° (221°)	445° (230°)
Flexible PVC	2.5:1	265° (130°)	340° (171°)	355° (181°)	365° (181°)
Rigid PVC	2.5:1	300° (149°)	320° (160°)	340° (171°)	365° (181°)
PC	2.25:1	510° (266°)	530° (277°)	550° (288°)	560° (293°)
Noryl®	2.1:1	450° (232°)	480° (249°)	510° (266°)	510° (266°)
Ultem®	2.1:1	600° (316°)	640° (338°)	675° (357°)	675° (357°)
PET	3:1	520° (270°)	550° (290°)	510° (265°)	510° (265°)
PBT	2.5:1	470° (243°)	490° (254°)	500° (260°)	500° (260°)
Polysulfone	2.5:1	550° (288°)	600° (316°)	650° (343°)	650° (343°)
Acetal	3.6:1	400° (204°)	390° (199°)	400° (204°)	410° (210°)
Thermoplastic Polyurethane	3:1	330° (166°)	360° (182°)	380° (193°)	380° (193°)

6.2 Extruder or Production Run Setup

Setting up a new production run or changing from product 1 to product 2 requires all transfer lines, feed hoppers, and downstream equipment to be clean to prevent cross-contamination of product 2 with product 1. If the same dryer is used for both products, the dryer and vacuum transfer lines must be completely cleaned to prevent cross-contamination. The extruder is run empty and possibly purged with appropriate purge material, depending on the next product to be processed. It may be necessary to pull the screw and use a new screw depending on the next polymer's processing requirements. Changing screws to optimize throughput and product quality is advantageous if the new product is going to be run for an extended time or the products scheduled after this product change also require the new screw.

Check all raw material availability and verify that all formulation ingredients are at room temperature or in the dryer for the appropriate time and temperature to yield the correct moisture content. If raw materials are stored in either a cold or hot warehouse, the raw materials should be brought to the processing plant and allowed to come to equilibrium temperature at least 24 hours prior to the run start time. Plastics are great insulators; consequently gaylords or other large containers require long times to come to equilibrium temperature. Adding either hot or cold resin to the extruder can alter the processing parameters, moving the melting point toward either the feed or die end, resulting in the processing being different from previous production runs.

Before starting any production, compare the SOP with previous run records to determine how the product ran last time. This tells the operator what to anticipate and whether any changes may be required at start-up. If the SOP has been changed and/or reissued since the previous run, ask questions to understand what necessitated the changes. Check all setpoints versus the SOP and verify that the equipment (heater bands, thermocouples, temperature controllers, cooling on the feed throat and barrel, and so forth) is functioning properly.

Before starting a new run, check and correct any safety concerns on or around the extruder, such as bare wires or loose connections on the die, housekeeping around the extruder, proper insulation on the die, etc.

If the extruder has been down for an extended time or shut off completely, it is necessary to heat soak the extruder, bringing all temperature zones and dies up to their setpoints before starting. The time required to reach equilibrium temperature depends on the extruder size. If the die comes up to temperature much slower than the extruder barrel, turn on the power to the die heaters prior to the extruder. An hour is normally enough time to bring an extruder up to temperature; additional time heat soaking may result in degrading resin left in the extruder barrel generating black specks at start-up. Black specks generated by resin degradation need to be completely purged from the system before collecting first-grade product.

If either volumetric or gravimetric feeders are used to feed resin and/or additives, they may require calibration or resetting from the previous run. Gravimetric or loss-in-weight feeders are easily set by inserting the correct pounds/hour (kilograms/hour) throughput rate required for each feed stream in the formulation. Volumetric feeders may need to be recalibrated for the formulation ingredient being fed with a particular feeder. The rate is manually checked, verifying that each feeder is delivering the correct pounds/hour (kilograms/hour). No feeder calibration is required when flood feeding a premixed blend or virgin material.

Downstream equipment is checked for safety, proper operation, and maintenance prior to starting. Speeds are set, slitting or trim knives changed, rollers cleaned, nip roll gaps set, roll temperatures or cooling tanks adjusted, etc.

6.3 Start-up

Prior to starting the extruder, retorque the die bolts after the die and extruder are up to temperature. With feed resin in the hopper, start the extruder at 10 to 20 rpm, until molten polymer is flowing freely from the die. During this time, monitor the breaker plate pressure. If there is a solid polymer plug in the die or transfer pipe or some other die blockage, the pressure will increase very rapidly, causing the rupture disk to fail unless the screw rotation is stopped. If a pressure spike occurs, immediately turn off the screw motor and determine the cause before proceeding. **Do not walk in front of the die during start-up or until material is flowing freely from the die and no more air is being forced out of the extruder die.** Once molten polymer exits the die, the extruder speed can be safely increased.

Regardless of the extrusion process, the downstream equipment is normally strung up at low extruder rpm and takeoff speeds. After the downstream equipment is properly strung up and product is running satisfactorily, increase the takeoff and extruder screw speeds simultaneously to the desired settings. Bringing the equipment up to speed too fast or using the wrong puller speeds initially can cause the molten polymer web to be easily broken, requiring the process to be restarted. The pull or draw on the molten web depends on the polymer melt strength. Extrudate with high polymer melt strength can be easily handled, while low melt strength makes stringing up the downstream equipment a special challenge. Elastomeric materials and impact-modified products exhibit very good melt strength and are easy to handle, whereas high MFI crystalline materials and highly filled materials may have poor melt strength and break easily.

After running the extruder for a few minutes, collect product samples and determine if the product meets specifications. Depending on the extruder size, it takes a while for the entire system to come to equilibrium. Continue to collect product samples and monitor the product and process while the extruder comes to equilibrium, making any minor die adjustments required to bring the product into dimensional specification. Changes in extruder screw speed or puller speed are required if gross product dimensional changes are required.

Once equilibrium is attained, record all operating conditions, i.e., setpoint and actual temperatures, melt pressure, melt temperature, extruder load, screw rpm, feed rates, dryer conditions, takeoff speeds, takeoff equipment temperature, pressures associated with takeoff equipment, vacuum levels, product quality, and product dimensions.

6.4 Steady-state Operation

Steady state depends on the extruder size. Small extruders (2.5") may take 20–40 minutes to reach equilibrium, while large extruders (8" or greater) may require two or more hours. Equilibrium time is dependent on the metal mass being heated, the resin processed per hour (resin is cold and takes heat away from the barrel), and the shear heat generated by the motor and screw. Changes in process conditions are never done before the equipment reaches equilibrium. After changes are made, the equipment must be allowed to reach equilibrium before the need for any further change can be determined. If an extruder is running properly at start-up, producing a quality part, and 45 minutes to an hour later is running poorly and no process changes have occurred, the product does not run well at equilibrium conditions but does run well under the actual start-up temperatures.

New products being scaled up from either R&D or from a small to a large production extruder may require process changes to establish standard operating conditions that produce the most robust product. If temperature or speed changes are required, make a significant change in conditions to verify that the change had an effect. For example, changing zone 2 temperature 2°F (1°C) is not

normally large enough to have any significant effect on the product. Instead, change zone temperatures 15–25°F (8–14°C) to determine if the change has significant effect on the product variable in question. Allow the extruder time to reach equilibrium before passing judgment on the effect the change has on the variable property in question. Effects of screw speed changes are not instantaneous. The extruder barrel temperatures have to reach equilibrium conditions resulting from viscous heat generated by the screw speed change. Due to the metal mass in the barrel, heating and cooling systems require time to compensate for temperature changes resulting from screw speed changes that add or remove shear heat from the process.

Any process change that does not show an effect on the product can be returned to its original settings before another change is made. After a process change, allow time for the extruder to come to equilibrium and change only one variable at a time unless a DOE study is being performed. **Do not be a knob turner** when trying to solve extrusion problems. The extruder needs time to come to equilibrium. Rapid changes make it hard, if not impossible, to define optimum process conditions.

6.5 Shutdown and Product Changes

Product change, maintenance, or weekend work stoppage requires shutting down the extruder. Sometimes the operator is required to operate a different line. The shutdown or changeover procedures used depend on many factors:

- How long will the system be down?
- Is the product change from a natural or lighter-color resin to a darker color?
- Is the product change from a darker to a lighter-color or natural resin?
- Is the product change to a completely different resin system?
  - — Does the new formulation process at higher or lower melt temperature?
  - — Is the new resin formulation compatible with the previous resin?
  - — Is the new resin melt viscosity higher or lower than the old resin at a given melt temperature?

The first question, “How long will the extruder be down?” dictates the procedure to follow. If rigid PVC is being processed and the extruder needs to be shut down for more than a few minutes, PVC is purged from the extruder to prevent degradation. PVC goes through an autocatalytic degradation reaction; once degradation starts, it becomes progressively worse. Black specks and degraded material are generated very rapidly. With other resin systems, if the extruder is going to be down for more than 10 minutes, remove all material from the feed throat (close the slide gate between the feed hopper and extruder and run the extruder dry) to prevent melt plug formation around the screw in the feed zone and to minimize bridging in the feed throat.

If the line is going to be shut down overnight and restarted in the morning, thermally stable resins are run out of the barrel and the extruder temperatures turned off or down to a lower level, then brought back up to temperature prior to the desired start-up time. If the resin system is thermally stable but not oxidatively stable, run the extruder until the feed section is clear, while leaving material in the die and breaker plate area to prevent oxygen from coming in contact with the molten resin on cool-down or reheating. Depending on the time the extruder is down, either lower or turn off the extruder temperatures. Finally, if the resin is thermally unstable, run the resin out of the barrel and purge with compatible, thermally stable resin that does not degrade. Turn the extruder temperatures off or lower them, depending on the time the extruder will be sitting idle. Restart the extruder, bring all zones up to temperature, and run virgin resin through to remove all purge and any degraded resin before starting the process.

Product changes resulting from color change are straightforward when going from a lighter to a darker color. Run the extruder hopper empty (verify that the lighter-colored formulation is removed from all transfer lines), then add the new darker formulation to the extruder. This change is made easier by slowing down the screw speed and downstream equipment during transition from product 1 to product 2 and discarding any transition material. (Transition product can be ground up and used at a later time as regrind in a darker or black product if the resin is the same in both formulations.) After the original lighter-color product is removed from the system, collect the new product. If the product change is from a darker to a lighter-color formulation, the product change requires more time and possibly a purge step. Run the feed hopper and extruder empty, purge with a low-cost material or regrind to remove the darker resin, run the purge out of the extruder, and add the new lighter-color resin to the feed hopper and extruder. With all the purge flushed out of the system, restring the line and start collecting new lighter-colored product.

When changing products requires an entirely different resin matrix, the conversion procedure depends on the difference between the old and new resins. Does the conversion require higher melt temperatures, are the original and the new resins chemically or molecularly compatible, what is the viscosity of each resin at product changeover, and is a different screw required? Figure 6.2 shows possible resin scenarios (each situation is numbered) when changing

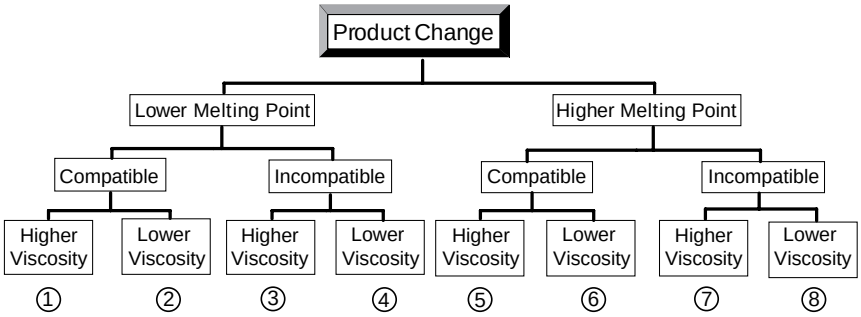

Figure 6.2. Various product changeover scenarios.

from resin A to resin B. The rest of this section describes procedures to follow where resin changes are required.

Scenario 1

Scenario 1 in Fig. 6.2 is conversion to product B with a ***lower*** melting point polymer that is ***compatible*** with and has a ***higher*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Add formulation B to the feed hopper.
- Lower the barrel and die temperatures.
- Run the extruder and takeoff equipment at slower speed until the barrel temperatures and melt temperature stabilize at their new setpoints.
- Once the system reaches equilibrium, ramp the extruder screw speed back to its operating range, followed by the takeoff equipment, and start collecting the new product.

Scenario 2

Scenario 2 in Fig. 6.2 is conversion to product B with a ***lower*** melting point polymer that is ***compatible*** with and has a ***lower*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Add new resin to the hopper.
- Lower barrel and die temperatures.
- Run the extruder and takeoff equipment at slower speed until the temperatures stabilize at their new setpoints.
- Depending on the melt strength of product B, the process may continue to run without a web break. If the extrudate web breaks, run the transition product to the floor until material A is purged from the extruder.
- Once all of formulation A is out of the extruder and the temperatures have reached their setpoints, restring the takeoff equipment.
- Increase extruder screw and takeoff speed and collect product.

Scenario 3

Scenario 3 in Fig. 6.2 is conversion to product B with a ***lower*** melting point polymer that is ***incompatible*** with and has a ***higher*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Understand the degree and reason for resin incompatibility between products A and B:
  - — Cannot be mixed because the resins react chemically, generating noxious gases.
  - — Cannot be added to the extruder because the barrel and die temperatures are too high and degradation of resin B will result.
  - — Not compatible because of molecular polarity. Polar resins (resins containing dipole moments) tend to be compatible with polar resins, and nonpolar resins tend to be compatible with nonpolar resins, while mixtures of polar and nonpolar resins are incompatible.
- Procedure to use when resins can't be mixed because of chemical reaction or thermal decomposition:
  - — Purge resin A out of the extruder with an appropriate purge material.
  - — Lower barrel and die temperatures.
  - — At the proper extrusion temperature for resin B, add the formulation to the extruder.
  - — Start up at low screw rpm until resin is flowing freely out the die.
  - — Restring the process at low speed.
  - — Increase extruder screw and takeoff speeds and start collecting product B.
- Procedure to use with incompatible resins due to the combination of polar and nonpolar polymers:
  - — Run formulation A out of the extruder.
  - — Lower barrel and die temperatures.
  - — Start feeding resin B to push resin A out of the extruder. Initially run at high rpm until all of resin A is flushed from the extruder and die. When the system is completely purged, lower the extruder speed until all barrel and die temperatures reach setpoints.
  - — At the proper extrusion temperatures, restring the process at low speed.
  - — Increase all speeds and start collecting product.

Scenario 4

Scenario 4 in Fig. 6.2 is conversion to product B with a ***lower*** melting point polymer that is ***incompatible*** with and has a ***lower*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Procedure to use for resins that can't be mixed because of chemical reaction or degradation due to temperature:
  - — Purge resin A out of the extruder with appropriate purge material.
  - — Lower barrel and die temperatures.
  - — At the proper extrusion temperature, add resin B to the feed hopper.
  - — Start extruder up at low rpm until extrudate is flowing freely from the die.
  - — Restring the process at low speed.
  - — Increase extruder screw and takeoff speeds and start collecting product.
- Procedure to use for resins that are incompatible because of molecular polarity:
  - — Run or purge resin A out of the extruder.
  - — Lower barrel temperatures.
  - — Start feeding resin B to the system to force the purge of resin A from the extruder. Due to the lower viscosity of resin B, it may be difficult to force the purge of resin A from the extruder. Two extreme alternatives to remove resin A are to pull and clean the screw, barrel, and die, or to find a third resin that is compatible with both resins A and B and use this resin as a purge.
  - — Initially run the screw at high rpm until all of resin A is removed from the extruder.
  - — After removal of resin A, lower the extruder rpm and run at slow speed until the die and barrel temperatures are stabilized.
  - — Restring the process at low speeds.
  - — Increase all speeds, and start collecting product B.

Scenario 5

Scenario 5 in Fig. 6.2 is conversion to product B with a ***higher*** melting point polymer that is ***compatible*** with and has a ***higher*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Increase barrel and die temperatures just prior to the end of the current production run, before the feed hopper is completely empty.
- With resin A present, slow the extruder and takeoff speeds.
- As the new temperature setpoints stabilize, add resin B to the extruder feed hopper and increase speeds.
- When resin A is completely purged from the extruder and all speeds are properly set, start to collect product B.

Scenario 6

Scenario 6 in Fig. 6.2 is conversion to product B with a ***higher*** melting point polymer that is ***compatible*** with and has a ***lower*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Increase barrel and die temperatures just prior to the end of the current production run.
- Run the extruder feed hopper empty.
- Procedure to use depends on the temperature increase required. With a 50°F (28°C) change, use the following procedure:
  - — Add resin B to the extruder hopper and continue to run at a high rate through the entire line.
  - — Once all material A has exited the extruder, slow the extruder screw speed and takeoff equipment.
  - — When the temperatures stabilize at the new setpoints, increase extruder screw and takeoff speeds.
  - — Start to collect product.
- If the temperature has to be raised 50°–200°F (28°–111°C), use the following procedure:
  - — Run the extruder empty.
  - — Raise the setpoint temperatures 50°–100°F (28°–56°C) and add resin B to the extruder.
  - — Try to run the extruder very slowly. If both the drive load and head pressures are low, increase the extruder speed slowly to purge resin A from the extruder.
  - — Lower the screw speed and continue to run as the temperatures stabilize at the new setpoints.
  - — Restring the line.
  - — Bring the screw and takeoff speeds up to their normal operating levels.

Scenario 7

Scenario 7 in Fig. 6.2 is conversion to product B with a ***higher*** melting point polymer that is ***incompatible*** with and has a ***higher*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Procedure to use if the resins can't be mixed because of chemical decomposition or degradation:
  - — Purge material A out of the extruder and die.

— Raise the barrel and die temperatures.

— Once the temperature reaches the setpoint, add resin B to the extruder.

— Start up at low screw speed.

— Restring the process at low speeds.

— Increase extruder screw and takeoff speeds and collect product.

- Procedure to follow if resins are incompatible because of molecular polarity:

— Run resin A out of the extruder and die.

— Raise the barrel and die temperatures.

— Start feeding the new resin very slowly to the extruder to force resin A out. Initially run the extruder at low rpm and verify that the drive load is not too high or the polymer is not freezing in the die, causing excessively high head pressures. As resin B flows freely from the die, increase screw speed to generate more shear heat, assisting the barrel temperature increase.

— Verify that resin A is out of the extruder.

— Shut the extruder down or run very slowly until all temperatures stabilize at their setpoints.

— Restring the process at low extruder and takeoff speeds.

— Increase all speeds and start to collect product.

Scenario 8

Scenario 8 in Fig. 6.2 is conversion to product B with a ***higher*** melting point polymer that is ***incompatible*** with and has a ***lower*** viscosity than the polymer in product A. A procedure to follow in changing from product A to B is—

- Run the extruder feed hopper empty.
- Procedure if the resins cannot be mixed due to chemical decomposition or degradation:

— Purge formulation A out of the extruder and die using appropriate purge material.

— Raise the barrel and die temperatures.

— Add resin B to the extruder at lower temperatures than the setpoints so the viscosity is sufficient to flush the purge material from the extruder.

— Allow the extruder and die temperatures to stabilize at the setpoint temperatures.

— Start up the system at low extruder screw speed and restring the process at low speeds.

— Increase extruder and takeoff speeds and collect product.

- Procedure to follow if resin incompatibility is due to molecular polarity:

— Run the extruder feed hopper, extruder, and die empty.

— Raise the barrel temperatures.

— Start feeding resin B to push out the original formulation. Initially run at low rpm and monitor the extruder load and die pressure very carefully. Resin B may freeze off or clog the die. Once material is flowing freely from the extruder and the drive load or head pressure is not excessive, increase the screw speed until formulation A is completely flushed from the extruder.

— Turn the screw off or run at low rpm until the extruder and die temperatures reach their setpoints.

— Restring the process at low speed.

— Increase extruder and takeoff speeds and start to collect the product.

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Review Questions

1. What is meant by "heat soaking" an extruder and how long does it take?
2. What are SOPs and where are they used?
3. What are some feasible extruder temperature profiles?
4. How do you set extruder processing conditions?
5. With the correct screw design for a particular resin, what is a good temperature profile and why?

6. Give the ideal screw and temperature profile to process polycarbonate, HDPE, acrylic, nylon, and polypropylene.
7. Describe a procedure to change extrusion products from resin A to resin B assuming resin A has a higher melting point, is lower in viscosity, and is incompatible with resin B.
8. Describe a procedure to change from processing resin C to resin D assuming resin D has a higher melting point, is compatible with resin C, and has a higher viscosity.
9. You need to schedule a production run to produce five colors (natural, black, light blue, yellow, and white) in polypropylene. What order do you schedule the resins to be processed and why?
10. While producing rigid PVC siding, a heater band fails on the extruder. What is the best course of action?
11. You have successfully produced PP sheet for the last three days on a 60-inch wide sheet die on a 6-inch extruder. Another two days are required to complete the order, it is Friday night, and the extruder needs to be shut down for the weekend. What are the recommended procedures for shutting down the extruder and restarting Monday morning?
12. You have been running a 4.5-inch extruder with PET and need to shut it down for about two hours while everyone in the plant attends a safety meeting. What action steps are to be taken with the extruder?

7 Scale Up

Scaling up from an R&D extruder or small production extruder to a larger extruder can create unexpected problems and product challenges. While it may be possible to scale up from a 0.75-inch to a 1.5-inch extruder in the laboratory, scaling up from smaller to larger extruders can be a real challenge. Any process data generated on extruders smaller than 2.5-inch diameter are not readily scalable to large production extruders. Product development is normally performed on small laboratory extruders due to either the lack of experimental resin, additives or other ingredients, or to minimize the experimental product produced under equilibrium extrusion conditions. As an example, experimental runs on a 2.5-inch extruder with a nominal throughput of 150 pounds/hour require extruding 50–75 pounds before the extruder reaches equilibrium processing conditions producing a representative sample. Developing or optimizing a new formulation with three ingredients using a mixture experiment (DOE approach) of 10 or 12 treatment combinations generates 750 to 900 pounds of experimental products, where only a few pounds are required for internal evaluation. Consequently, most product development is done on smaller extruders.

Assume either a new product formulation or a new process is being developed on a small extruder (1.5-inch diameter or less) to save material and facilitate development work, and the product or process has to be scaled up to a production extruder. This is best accomplished by developing a process on a small production extruder (2.5-inch diameter) with a process DOE to determine the critical processing parameters and interactions. The DOE defines the processing window, the critical process parameters, and any process-product interactions. A several hour validation run on a 2.5-inch extruder can establish both product and process viability and the process criteria required for scale-up to a larger extruder, and provide confidence that scale-up is feasible.

A minor processing problem on a small extruder can become a major problem on a large production extruder. As an example, vent flow on a 1.5-inch extruder where 3 to 4 pounds/hour flows out the vent stack is not difficult to control. Scaling up to a 6-inch extruder, the vent flow can become 50–100 pounds/hour, creating major housekeeping, collection, and disposal problems. As a second example, minor surging on a small extruder results in the product going out of dimension every 5 to 10 minutes. In the development stage, sufficient product is produced with correct product dimensions for evaluation. Assuming the yield because of surging is 80%, only 2 pounds out of every 10 are discarded. If all product evaluations are done with 20–40 pounds of product, surging may never be identified as a product/processing issue. Scaling the process up to a larger extruder magnifies the surging problem, possibly resulting in 25% rejects, and at throughputs of 500 pounds/hour this translates to 125 pounds/hour of rejected material and an incapable process.

As a third example of a scale-up issue, degradation that results from shear heating on a small extruder may be a minor or nonexistent problem. Assume gas is generated from polymer degradation, creating a small hole in the extrudate every 10 feet. In 100 feet of experimental production, only 10 holes occur and sufficient material is collected for evaluation. If you scale up to a larger extruder, where additional shear heat magnifies the polymer degradation, gas generation that produces a small hole every 10 feet in 4000 feet of production might necessitate the whole production run being rejected and scrapped.

In addition to product issues, extruder issues can occur during scale-up to larger diameter extruders. Extruder areas that might be affected include

- Drive motor torque sufficiency
- Thrust bearings
- Heaters
- Screw optimization
- Melt pressure uniformity
- Absolute melt pressure
- Melt temperature uniformity
- Absolute melt temperature
- Degree of mixing
- Shear heat
- Die design

During any scale-up, the process limiting factors need to be identified and their limits established. Potential limiting factors to the extrusion process include

- Motor power
- Volume capacity of the extruder
- Downstream cooling capacity
- Product properties, such as
  - — Shear heat
  - — Degradation
  - — Moisture
  - — Loss in physical properties at high temperature

Don’t run a process with low throughput just to make product; in the long term, a low throughput solution to a process issue is not economically viable. With low viscosity resins, heat transfer may be the limiting factor, due to lower conductive heat transfer in a large extruder compared to the viscous heat generated in a smaller extruder.

Table 7.1. Scale-up Parameters Proportional to

Parameter	Proportional to
Volumetric output, Q	D^2Nh
Mechanical power required, E	$(D^{n+2}N^{n+1}L)/h^n$
Specific mechanical power, E/Q	$(D^nN^nL)/h^{n+1}$
Pressure gradient, dp/dz	$(DN)/h^2$
Die pressure, P (assume Newtonian)	$(DNL)/h^2$
Wall shear rate, γ (assume Newtonian)	$(DN)/h$
Mean residence time, t_{res}	$L/(DN)$
Shear strain in mixing, γt_{res}	L/h
Specific heating surface, A/Q	DL/Q

Process scale-up factors and their proportionality are shown in Table 7.1.[1] Definitions of terms in Table 7.1 are

Q = Volumetric throughput
D = Screw diameter
L = Extruder length
h = Channel depth in the metering section
n = Power law index
N = Screw speed
E = Mechanical power

Specific mechanical power (E/Q) is for an adiabatic temperature (constant heat: no heat loss or gain from the surroundings). Heat is added by heaters or removed by cooling to move from the adiabatic temperature. Under adiabatic extrusion conditions, heat is neither added nor removed from the extruders.

One scale-up strategy is to increase all extruder dimensions in the same proportion. With a square pitch screw, the number of flights stays the same for a larger diameter extruder with the same L/D. Channel depth, h, is proportional to the diameter, so the output (Q) increases as a function of D^3N and the mechanical power required increases as function of D^3N^{n+1}. The adiabatic temperature (E/Q) is independent of D but increases as a function of the screw speed N^n. Die pressure and wall shear rate both increase as a function of N. Mean residence time in the extruder is proportional to $1/N$, while the specific heating surface area is proportional to $1/DN$. These data imply that doubling the extruder diameter yields approximately a 10-fold increase in output while not affecting the melt temperature, pressure, or mixing. Unfortunately, the 10-fold increase in output must be heated with approximately 4.5 times the barrel surface area, which may not be practical. The volumetric throughput increases by D^3, but the barrel surface area only increases by D^2, so limitations in melting capacity may exist at the same screw speed with the larger extruder. Improper melting can result in output variations and/or lower throughput.

A more common scale-up strategy is to maintain constant shear in the extruder, providing constant melt temperature through similar viscous heat generation. To accomplish this objective, the screw speed in the larger extruder is reduced by $D^{-1/2}$ and the channel depth in the metering section is increased by $(D_2/D_1)^{1/2}$, where D_2 is the diameter of the larger extruder and D_1 is the diameter of the smaller extruder being scaled up. The volumetric output of the new extruder is proportional to $(D_2/D_1)^2(Q_1)$, while the power required is proportional to $(D_2/D_1)^2(E_1)$. In practice the screw speed of the larger extruder is normally run at a higher rate than $(1/\sqrt{D})$. This results in the volumetric throughput, power requirements, and shear rates being higher than observed in the smaller extruder.

Table 7.2. Scale-up Factors at Constant Shear

Factor	Small Extruder, 1	Large Extruder, 2
Diameter, D	D_1	D_2
Channel width, W	W_1	$W_1[D_2/D_1]$
Channel depth, h	h_1	$h_1\sqrt{[D_2/D_1]}$
Screw speed, N	N_1	$N_1\sqrt{[D_2/D_1]}$
Volumetric output, Q	Q_1	$Q_1[D_2/D_1]^2$
Shear rate, γ	γ_1	$\gamma_1[D_2/D_1]^0$
Circumferential speed, V	V_1	$V_1\sqrt{[D_2/D_1]}$
Residence time, t_{res}	t_1	$t_1\sqrt{[D_2/D_1]}$
Melting capacity, M_p	M_{p1}	$M_{p1}[D_2/D_1]^{1.75}$
Solids conveying rate, M_s	M_{s1}	$M_{s1}[D_2/D_1]^2$
Screw power, Z	Z_1	$Z_1[D_2/D_1]^{2.5}$
Specific energy consumption, Z/Q_T	Z/Q_{T1}	$[Z/Q_{T1}]\sqrt{[D_2/D_1]}$

Common scale-up factors are shown in Table 7.2.[2][3] The operating parameters for extruder 1 are given by the following equations:

$$Melt\ Output\ Rate\,(Q) = Q_D - Q_P \tag{7.1}$$

$$Shear\ Rate\,(\gamma) = \frac{\pi \times D \times N}{h} \tag{7.2}$$

$$Screw\ \ Circumferential\ Speed\,(V) = \pi \times D \times N \tag{7.3}$$

$$Residence\ Time\,(\tau_{res}) = \frac{L}{V} = \frac{L}{\pi \times D \times N} \tag{7.4}$$

$$Solids\ Conveying\ Rate\,(M_S) = \rho \times H \times W \times N \times V\left(\frac{\mathrm{Sin}\theta}{\mathrm{Sin}\theta + \phi}\right) \tag{7.5}$$

where ρ = solid bulk density
θ = angle solid bed is moving
ϕ = helix angle

$$Specific\ Energy\ Consumption\,(SEC) = \frac{Z}{Q} = t_{res} \times \gamma^2 \times \eta \tag{7.6}$$

where η is the viscosity.

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1. Stevens, M. J., and Covas, J. A., Extruder Principles and Operation, Chapman and Hall, 2nd ed., 1995.
2. Rauwendaal, C., *Polymer Extrusion*, Macmillian (1986), New York, p. 434.
3. Bigio, D. I., "Scale-up in Polymer Processing," Extrusion Division RETEC, Asheville, NC, 1997.

Review Questions

1. What are some extruder limiting factors that may affect scale-up?
2. Why is scaling from one size extruder to another a challenge?
3. What is the smallest single screw extruder that can be scaled reasonably well to a large production extruder?
4. Assuming the shear rate is kept constant, a small extruder (2.5-inch diameter) has been running HDPE at 85 rpm and 175 pounds/hour. What is the anticipated throughput on a 6-inch extruder, and what screw speed is required to obtain that throughput rate?
5. In question 4, calculate the residence time in both the small and large extruders.

8 Shear Rates, Pressure Drops, and Other Extruder Calculations

Significant discussion has centered on shear rates, shear heating, and viscous heat generation in the transition and mixing sections. This chapter demonstrates how to calculate shear rates in the extruder and die, pressure drops in the die, and provides other calculations related to single screw extruders. Derivations of the equations are beyond the scope of this book and can be found elsewhere.

8.1 Shear Rates

Calculation of shear rate (γ) in an extruder screw channel is done using Eq. (8.1):

$$\gamma(Screw\ Channel) = \frac{\pi \times D \times N}{h} \tag{8.1}$$

where γ = Shear rate in the screw channel, sec^{-1}
D = Screw diameter
N = Screw speed
h = Channel depth

Shear rate is given in sec^{-1}. Shear rates between the screw flight and the barrel wall are calculated using Eq. (8.1) except h = distance between the screw flight and the barrel wall.

Calculation of shear rates in the die land area depends on the die shape. In a round channel (rod), the shear rate is calculated from Eq. (8.2):

$$\gamma(Round\ Channel) = \frac{4 \times Q}{\pi \times R^3} \tag{8.2}$$

where γ = Shear rate in round channel
Q = Volumetric flow rate calculated from Eq. (8.3)
R = Radius of the channel

The volumetric flow rate (Q) is found from Eq. (8.3):

$$Q(Round\ Channel) = k \times \frac{\Delta P}{\eta} = \frac{\pi \times R^4}{8 \times L} \times \frac{\Delta P}{\eta} \tag{8.3}$$

where k = Resistance factor ($k = \pi R^4/8L$)
R = Radius of the channel
ΔP = Pressure drop across the channel
L = Length of the channel
η = Polymer viscosity

The pressure drop in the channel is calculated from Eq. (8.4):

$$\Delta P = 2 \times \tau \times \frac{L}{R} \tag{8.4}$$

where τ = Shear stress = F/A = Force applied per unit area

An example of the resistance factor in a rod die on the flow rate is shown in Fig. 8.1. As the hole diameter decreases, the pressure increases based on k or the resistance factor.

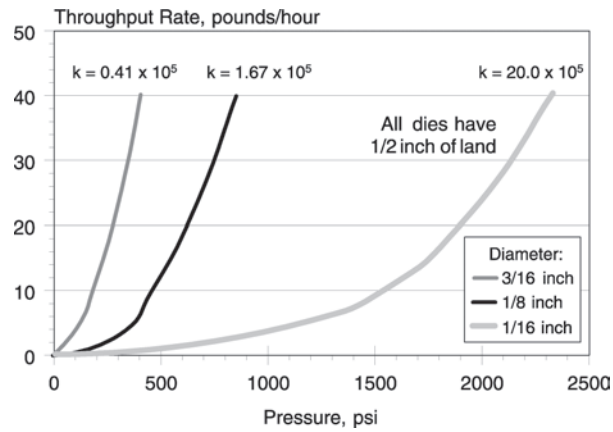


Figure 8.1. Effect of die resistance on throughput.[1]

Shear rates (γ) in a rectangular die channel (calculation for shear rates in sheet, film, and profiles dies) are given by Eq. (8.5):

$$\gamma(Flat\ Channel) = \frac{6 \times Q}{W \times h^2} \tag{8.5}$$

where γ = Shear rate in rectangular channel
Q = Volumetric flow rate
W = Channel width
h = Channel height

The volumetric flow rate (Q) is calculated from Eq. (8.6):

$$Q(Flat\ Channel) = k \times \frac{\Delta P}{\eta} = \frac{W \times h^3}{12L} \times \frac{\Delta P}{\eta} \tag{8.6}$$

where $k = (Wh^3)/12L$, a geometrical constant

Shear rates in annular dies (blown film, pipe, and tubing) are given by Eq. (8.7):

$$\gamma(Annular\ Die) = \frac{6Q}{\pi(R_1 + R_2)h^2} \tag{8.7}$$

where R_1 and R_2 = Inner and outer radius
h = Die gap

The volumetric output is calculated from Eq. (8.8):

$$Q(Annular\ Die) = k \times \frac{\Delta P}{\eta} = \frac{C_m \times h^3}{12 \times L} \times \frac{\Delta P}{\eta} \tag{8.8}$$

where $k = (C_m h^3)/12L$, a geometrical constant
$C_m = \pi(R_1 + R_2)$, the mean circumference

An example of shear rate calculations at different locations in the extrusion process follows. Polypropylene is being extruded in a 6-inch extruder equipped with a

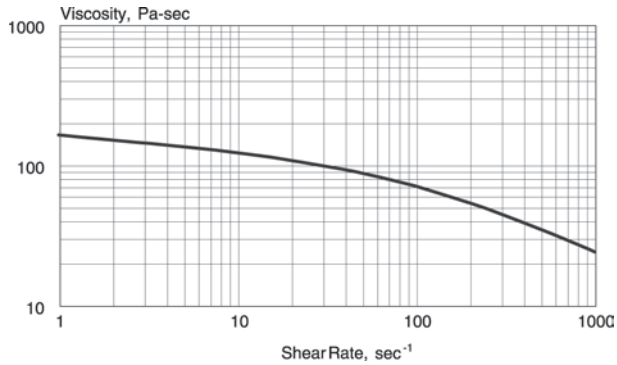


Figure 8.2. Shear rate versus viscosity for 12 MFI PP.

barrier flight in the metering section, having a barrel-to-wall clearance of 0.05 inches, a channel depth of 0.35 inches, and a screw speed of 75 rpm, producing a rate of 600 pounds/hour. Calculate the viscosity in the metering channel, between the barrier flight and barrel wall, and in the die land area of a sheet die that is 60 inches wide with a 0.125-inch opening. The density of PP is 0.91 g/cc, melt density is 0.75 g/cc, and the shear rate versus viscosity curve is given in Fig. 8.2.

$$\gamma\,(Screw\ Channel) = \frac{\pi DN}{60h} = \frac{3.1416\times 6\times 75}{60\div 0.35} = 67\,\text{sec}^{-1}$$

Extrapolation of Fig. 8.2 shows the viscosity at 67 sec^{-1} is approximately 80 Pa-sec.

Again from Fig. 8.2, the viscosity of plastic moving across the barrier flight is approximately 35 Pa-sec.

$$\gamma\,(Flat\ Channel) = \frac{6Q}{W\times h^2}$$

Using the equation for a flat channel (Eq. [8.5]), we see that we are given W and h. Q is given in pounds/hour rather than in^3/sec. To convert from pounds/hour to in^3/sec, use the 600 pounds/hour output and convert to a volumetric flow rate based on the melt density of PP being 0.75 g/cc.

$$Q = \frac{600\,\text{lb}}{\text{hr}}\times\frac{1\,\text{hr}}{3600\,\text{sec}}\times\frac{1\,\text{cc}}{0.75\,\text{g}}\times\frac{454\,\text{g}}{1\,\text{lb}}\times\frac{1\,\text{in}^3}{2.54^3\,\text{cm}^3} = 6.16\,\frac{\text{in}^3}{\text{sec}}$$

Now we can calculate the shear rate (γ) for the die land.

$$\gamma\,(Flat\ Channel) = \frac{6Q}{W\times h^2} = \frac{6\times 6.16}{60\times 0.125^2} = 39\,\text{sec}^{-1}$$

This does not yield the absolute shear rate or viscosity because the resistance to flow by the die is not included. However, the shear rates and viscosity will be close to the actual value using the volumetric output for the volumetric flow calculation, Eq. (8.6).

From Fig. 8.2, the viscosity in the die lip area is approximately 92 Pa-sec.

8.2 Extruder Screw Calculations[2]

Certain calculations can be made for energy usage, conveying rates, helix angles, and other properties associated with screw designs and dies. The power required to heat polymer from the feed throat to the molten state is given by Eqs. (8.9) and (8.10):

$$Power = \begin{pmatrix} energy\ required\ to\ heat\ polymer \\ from\ feed\ to\ melt\ temperature \end{pmatrix} - \Delta H_{\text{fusion}} \tag{8.9}$$

or

$$Power = mC_p\Delta T + m\Delta H_{\text{fusion}} \tag{8.10}$$

where m = Mass flow rate in kg/hr
C_p = Heat capacity in kJ/kg °C
ΔT = Difference in temperature between the feed and melt temperature in °C
ΔH_{fusion} = Heat of fusion for the polymer matrix in kJ/kg.

Note that the heat of fusion (ΔH_{fusion}) for amorphous polymers is zero.

Using Eq. (8.9), what is the power required to heat polypropylene running at 400 pounds/hour from 25°C to 235°C? Heat capacity of PP = 2.10 kJ/kg ºC and ΔH_{fusion} = 102 J/gm. First convert the 400 pounds/hour to kg/hr and ΔH_{fusion} to kJ/kg:

$$400\,\frac{\text{lb}}{\text{hr}}\times 0.454\,\frac{\text{kg}}{\text{lb}} = 182\,\frac{\text{kg}}{\text{hr}}$$

One must recognize that for the ΔH_{fusion} = 102 J/g is also = 102 kJ/kg. Calculating the power:

$$\begin{aligned} Power &= mC_p\Delta T + m\Delta H \\ &= 182\,\frac{\text{kg}}{\text{hr}}\times 210\,\frac{\text{kJ}}{\text{kg–°C}}\times(235\text{–}25)\text{°C} \\ &= 80{,}262\,\frac{\text{kJ}}{\text{hr}} + 18{,}564\,\frac{\text{kJ}}{\text{hr}} = 98{,}826\,\frac{\text{kJ}}{\text{hr}} \end{aligned}$$

Divide power by 3600 sec/hour to convert to kW:

$$Power = 98{,}826\,\frac{\text{kJ}}{\text{hr}}\times\frac{1\,\text{hr}}{3600\,\text{sec}} = 27.5\,\text{kW}$$

Since 1 kW = 1.36 HP or 1 kW = 860 kcal/hr, the power calculation becomes

$$Power = 27.5\,\text{kW}\times 1.36\,\frac{\text{HP}}{\text{kW}} = 37.4\,\text{HP}$$

or

$$Power = 27.5\,\text{kW}\times 980\,\frac{\text{kcal}}{\text{hr}} = 23{,}650\,\frac{\text{kcal}}{\text{hr}}\,Energy$$

From the perspective of energy costs, it takes 27.5 kW per hour to process 400 pounds/hour of PP resin at 235°C, based on the information in this example. Since extrusion equipment is not 100% efficient, it actually takes more energy than calculated to process the PP, since some of the heat is lost to the surroundings and the motors and drives are not 100% efficient.

Energy losses in an extruder come from feed throat cooling, barrel cooling, convective heat loss from the barrel and die, heat loss from DC drives, gear box losses, and losses due to operation of pumps, cooling fans, and the control panel. Approximately 61% of the energy put into the process is actually used to convert solid polymer to a molten resin for processing.

8.3 Calculations of Output in Different Sections of the Extruder

The maximum output possible in the feed section is calculated based on either the mass flow rate or the volumetric flow rates. Both calculations are based on maximum output, which is not normally attained because

- The coefficient of friction of the polymer with the barrel wall or the screw root is not ideal,
- The internal coefficient of friction forcing the resin to flow as a plug is not optimum,
- The extruder generates pressure, and/or
- Bulk density varies.

The volumetric flow rate is given by Eq. (8.11) and the mass flow rate by Eq. (8.12):

$$Volumetric\ Flow\ Rate = v_z \times W \times h \tag{8.11}$$

where v_z = Channel velocity = $\pi\ D\ N\ Cos(\varphi)$

$$Mass\ Flow\ Rate = D^2 \cdot N \cdot h \cdot \rho_{bulk} \tag{8.12}$$

The volumetric flow rate is given in inches^3/minute and the mass flow rate in pounds/minute based on

W = Feed channel width in inches
H = Feed channel depth in inches
D = Screw diameter in inches
N = Screw speed in revolutions per minute
φ = Helix angle
ρ_{bulk} = Bulk density in pounds per cubic inch

The same formulas (Eqs. [8.11] and [8.12]) are used to calculate the volumetric and mass flows in the metering zone, with the exception that the channel depth and width are based on the channel in the metering section, and the ρ_{melt} is the melt density rather than the bulk density.

The optimum channel depth in the screw metering section is calculated from Eq. (8.13):

$$Optimum\ Channel\ Depth = \sqrt{\frac{2 \cdot \eta \cdot v_z \cdot \Delta L}{\Delta P}} \tag{8.13}$$

where ΔP = Pressure change over the metering section

The optimum helix angle in the metering section is calculated from Eq. (8.14):

$$Optimum\ Helix\ Angle = \tan 2\alpha = \frac{6 \cdot \eta \cdot v_z \cdot \Delta L}{\Delta P \cdot h} \tag{8.14}$$

REFERENCES

1. Christensen, R. E., "Science of Extrusion Course," March 10, 1997.
2. Kruder, G. A., "Extruder Efficiency: How You Measure It—How You Get It," *Plastics Engineering*, June 1981. p. 20.

Review Questions

1. Calculate the shear rates in the screw channel and over the barrier flight while extruding acrylonitrile butadiene styrene (ABS) on a 4.5-inch extruder running at 133 rpm with a mixing screw that has a barrier flight in the metering section. Channel depth in the metering section is 0.38 inches, and the clearance between the barrier and barrel wall is 0.04 inches.
2. Assume the ABS in question 1 is running at 642 pounds/hour through a sheet die producing 0.20-inch thick film 48 inches wide. What is the approximate shear rate in the die lip area? Melt density of ABS is 0.88 g/cc.
3. Calculate the approximate shear rate in the die lip area of a blown film die producing HDPE at 330 pounds/hour (melt density = 0.72 g/cc) on a nominal 8-inch diameter blown film die with a gap opening of 0.05 inches.
4. Calculate the power required to heat polycarbonate (heat capacity of PC is 1.40 kJ/kg °C) from 25°C to 310°C at a rate of 650 pounds/hour throughput.
5. Calculate the power required to heat nylon 6,6 running at 617 pounds/hour from 25°C to 275°C. The heat of fusion for nylon 6,6 is 20.5 J/kg$\times 10^{-4}$, and the heat capacity is 2.15 kJ/kg °C.

Part 2: Twin Screw Extrusion

9 Twin Screw Extrusion Process

Extrusion of polymeric materials to produce finished products for industrial or consumer applications is an integrated process, with the extruder comprising one component of the entire line. In some applications the production lines are very long with numerous operators, requiring operators to communicate and work together to produce an acceptable finished product. If the extruder temperature profile is set incorrectly, the product ingredients are not properly formulated, the cooling on the extruder feed throat is not running properly, the melt temperature at the extruder discharge is incorrect, the cooling bath temperature is not correctly set, the puller is running at the wrong speed, or any other incorrect operating condition or combination of conditions, the product may not meet customer specifications. Each step in the process adds value; consequently, the product reaches its maximum value at the end of the line. An improper setting at the beginning of the process may cause the product to be unacceptable at the end of the line after significantly more value has been added. Speeds at the different process steps must be matched to ensure product compliance.

Figure 9.1 is an extrusion process block diagram. Polymeric material is received, inspected, and stored. Prior to extrusion the polymer may be blended with additives (stabilizers for heat, oxidative stability, UV stability), color pigments or concentrates, flame retardants, fillers, lubricants, reinforcements, etc., to produce the desired product property profile. Some resin systems must be dried prior to extrusion to eliminate polymer degradation due to moisture. Other resins, which do not normally require drying, may have to be dried if they are stored in a cold warehouse and brought into a warm environment, causing moisture to condense on the pellet, flake, or powder surface. Once the polymer or blend is properly dried and ingredients mixed, the formulation is fed to the extruder, where it is melted, mixed, and delivered to the die to shape the extrudate. After exiting the die, the product is cooled and solidified in the desired shape and pulled at constant velocity to attain the appropriate cross section. Secondary operations, e.g., flame treatment, printing, cutting, annealing, are done in line after the puller. Finally the product is inspected, packaged, and shipped.

The different process parts are discussed in more detail in this chapter.

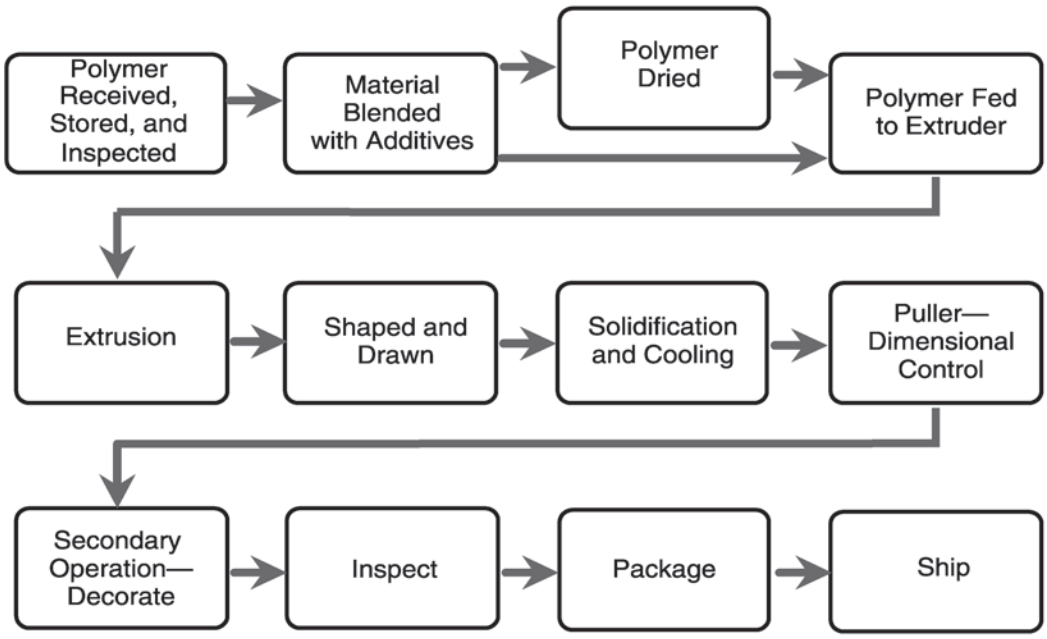


Figure 9.1. Basic extrusion process schematic.

9.1 Raw Material Supply

Polymer resin is shipped in different size containers depending on the quantity ordered, the processor's handling and storage method, and the extruder feed method. Small lots are shipped in 50- or 55-pound bags, with large lots shipped by tanker truck or rail. Table 9.1 shows a combination of packaging and shipping methods. Plastic pellets can be air or vacuum conveyed around the plant to storage containers or the extruder hopper.

Pellets conveyed between storage silos, dryers, surge hoppers, and extruder hoppers must be in dedicated lines

Table 9.1. Plastic Packaging

Package Size, Pounds	Type Package
50–55	Bags
300	Fiber Pack
1,000	Gaylord
4,000	Bulk Pack
40,000	Hopper Truck
150,000–220,000	Rail Car

or properly cleaned lines to prevent product cross-contamination. All lines must be properly grounded to eliminate static electricity during resin transfer.

Raw materials stored in uncontrolled temperature warehouses need to be brought to room temperature prior to extrusion. If the raw material temperatures vary between summer and winter, the polymer melting or softening point in the extruder will occur at a different location, leading to different melt viscosities and extrudate flow and consistency. Assume the raw material temperature is 50°F (10°C) in the winter and 80°F (26°C) in the summer; additional heat must be added to the raw material during the winter months, either by a hopper dryer, allowing the polymer to come to equilibrium at room temperature, or by adding additional heat in the first extruder zones to ensure the polymer is melting or being plasticated in the transition zone. A significant time period is required to raise the temperature of resin pellets that sat in a cold warehouse or in a cold truck to room temperature because polymers have low thermal conductivity.

Storing raw materials in a hot environment for a long time can consume the polymer stabilization package. Most thermal stabilization packages are consumed over time as the polymer is heated. While thermal degradation happens fairly rapidly at elevated temperatures in the presence of oxygen, below the melting or softening temperature degradation continues at a slower rate. Stock should be rotated to minimize the long-term thermal degradation effects.

Many raw materials are accepted from vendors based on a "certificate of compliance." Good procedures dictate that incoming raw materials are periodically tested and a critical polymer properties database established. Most internal extrusion problems are not caused by raw material variations; however, in the event the wrong raw material is used, the processor should be able to identify raw material inconsistencies immediately to minimize operating losses. Critical raw material properties for a particular application need to be identified and characterized so that incoming materials are tested only for the properties that affect the final part performance. Critical properties may be viscosity, long-term heat aging, color, tensile properties, or other parameters, depending on particular end-use application.

9.2 Raw Material Blending and Mixing

Depending on the product requirements, some preblending or ingredient mixing may be required prior to extrusion. (Blending and mixing are covered in more detail in Part 5, Chapter 33, "Feed Systems".) Unless a single polymeric material is being added to an extruder, the best way to combine different raw materials and keep them uniformly distributed prior to entering the extruder feed throat depends on many factors. Some factors to be considered include

- Powder-pellet separation
- Uniform distribution of additives introduced at low concentrations
- Proper mixing
- Different regrind levels and/or regrind particle size
- Liquid additive addition
- Powder/powder blend uniformity

The best material metering method to guarantee uniform component distribution is to gravimetrically feed each material with different feeders directly above the extruder feed throat. Assuming there are enough space and feeders for the various components in the formulation, gravimetric or loss-in-weight feeding ensures each component is added in the correct proportion, while addition directly above the feed throat minimizes any ingredient segregation. The downside of this approach is the gravimetric feeder cost, the space required for more than four or five components, and how many different size feeders are required. Assuming some components are added in very low concentrations (<1%) while other components are added in high concentration (>15%), the feeder size, feeder accuracy, and type of material (powder, pellets, flake, free flowing versus compressive powder, fiber, etc.) being fed are critical to the feeder performance. If all feeders are properly sized, designed for the materials being fed (single screw feeder, twin screw feeder, vibratory, weigh belt, etc.), and there is enough room to use a gravimetric feeder for each component, this is the best method to ensure a repeatable, uniform formulation is introduced to the extruder.

In a significant number of applications, a feeder is not available for each ingredient, requiring some component preblending. The type of blending depends on the ingredients being mixed and the material handling after blending and prior to extrusion. Assume pellets A and B are approximately the same size and must be premixed. A and B are individually weighed and added to a low intensity blending system. Typical low intensity blending systems include tumble blenders (many sizes), V-cone blender, ribbon blender, cement mixer, drum roller, or paint shaker

for small lots. The same equipment can be used to mix pellets and powder. However, pellets and powder are more likely to separate when transporting the blend or loading it to a feed hopper. The powder can flow between the pellets. Consequently, at the beginning of the extrusion run, the product may be rich in the powder component; while at the end of the extrusion run, the product may be rich in the pellet component. One method to minimize this separation is to coat the pellets with a liquid such as mineral oil, providing a surface to which the powder can adhere. Of course experimentation is required to verify an effective oil level and that the mineral oil does not affect the final product properties or performance.

Powder/powder blends can be mixed either in low intensity mixers described above or in high intensity mixers. High intensity mixers operate on the same principle as kitchen blenders. A mixing blade rotates at high speed, forming a vortex in the blender as it mixes the components. Due to the intense nature of mixing, heat is generated; care must be taken not to melt the blend components. Blenders may be jacketed to heat components during the blend cycle. With polyvinyl chloride (PVC), generated heat softens the PVC particle surface, allowing heat stabilizers to adhere to the individual PVC particles. Powder/powder blends, once properly mixed, will not tend to separate during transfer if the different component particle sizes are similar.

Uniform additive addition at low concentrations creates a significant mixing and blending challenge. Obviously, the best method is to feed the component directly into the feed stream with a small gravimetric feeder. However, this is not always practical or feasible. An alternative approach is to mix the additive (assume it's a powder) with some resin powder being used in the formulation and produce a masterbatch on a high intensity mixer. As an example, assume two additives, C and D, must be added at 0.5% and 0.08%, respectively, to resin B to produce a profile of material Z. A blend or masterbatch is produced by combining resin powder B with high concentrations of C and D and letting that blend down in an individual feeder. The masterbatch is added using feeder #1 and pellets of B are added via feeder #2 to produce the correct ingredient ratio in the final product. A 100-pound masterbatch is produced, containing 10 pounds (10%) of component C, 1.6 pounds (1.6%) of component D, and 88.4 pounds (88.4%) of resin B. This masterbatch is let down in a 19:1 ratio, with resin B feeding at a rate of 190 pounds/hour and the masterbatch feeding at a rate of 10 pounds/hour, to produce the correct ratio of additives in the profile Z. If components C and D were fed directly to the extruder at 0.5% and 0.08%, the feed rate for the different blend components would be 1.0 pound/hour for component C, 0.16 pound/hour of component D, and 198.84 pounds/hour of resin B to produce a 200 pounds/hour rate of profile Z. Using a masterbatch makes feeding small concentrations of ingredients uniformly more practical.

Liquid color or liquid additive addition to a twin screw extruder is quite easy. Liquid can be introduced into any barrel section with a liquid feed pump and injection port. A screw designed to accept the liquid and compound it into the resin is the only criterion. With a gravimetric or loss-in-weight liquid feed pump, the pump rpm is adjusted to keep the gravimetric feed rate constant. However, if a volumetric liquid feed pump is used (runs at constant rpm), the feed rate is dependent on the liquid temperature, which affects its viscosity and consequently the feed rate. Initially a volumetric liquid feed pump must be calibrated and a graph generated, showing motor rpm versus output rate in pounds/hour. On the same graph, rate curves versus motor rpm need to be generated at different temperatures. If the liquid temperature changes during the run, the feed rate will vary, and the liquid component concentration in the final product will change over time. Assuming the liquid additive and feed pump are in the vicinity of the extruder, it is possible the liquid temperature may increase during the run as the room temperature increases from the heat generated by the extruder (liquid density will decrease). This results in a decrease in the liquid feed rate and the wrong product formulation unless the liquid feed pump rpm is increased.

9.3 Drying

Some polymers require drying prior to extrusion to prevent polymer degradation. Resins, e.g., nylon, polyester (polyethylene terephthalate [PET] and polybutylene terephthalate [PBT]), and polycarbonate (PC), are hygroscopic, absorbing moisture rapidly from the air. At higher temperatures moisture degrades these materials to lower molecular weight (shorter chains), resulting in poorer property performance. Proper drying to eliminate moisture is critical in obtaining optimum property performance in the final product. Other materials, e.g., acrylics, Ultem®, polysulfone, Noryl®, and acrylonitrile butadiene styrene (ABS), also absorb moisture from the air and must be dried prior to processing. Any moisture in the polymer is converted to steam in the extruder and, depending on the quantity present, can cause surface imperfections such as splay, holes in the product, or a foamy product. Some polymers, e.g., nylon, are shipped dry in moisture-proof containers. With proper handling, these resins do not normally require additional drying prior to processing. However, if the seal is broken on the container or the bag is not completely resealed after opening, the product will absorb moisture and have to be dried prior to extrusion. Polyesters are particularly sensitive to moisture and must be dried in dehumidifying dryers, transported with dry air, and blanketed with dry air or nitrogen in the feed hopper.

Dehumidifying dryers with –40°F (–40°C) dew points are recommended for drying most polymers. Dryers are covered in more detail in Part 5, "Auxiliary Equipment."

Formulations requiring both a dry polymer plus blending with other ingredients can lead to special handling requirements. Once moisture-sensitive resin is dried, it will pick up moisture when exposed to the atmosphere. Additives or other components added to formulations containing hygroscopic resins need to be moisture-free. If the additives cannot be dried with the resin, special handling procedures or individual feeders are required to mix the dry resins and other additives or components at the extruder feed throat.

In some instances, resins containing moisture can be processed in a vacuum-vented extruder, with the moisture removed in the vent section. This does not work with all resins because some degradation can occur before the moisture is removed.

Overdried resin should be avoided to prevent resin degradation resulting in property loss and/or undesirable color development. When overdried, nylon 6,6 becomes yellow and is accompanied with poorer properties.

9.4 Feeding Polymer to the Extruder

The proper method of feeding polymer to a twin screw extruder depends on the twin screw extruder design, the material being fed, and the feed location. In parallel counter and corotating twin screw extruders, single or multiple resin feed streams and additives are starve fed to the extruder feed section. Feeders deposit the formulation directly onto the extruder screw with the screw speed set to process the formulation at a higher rate than it is deposited on the screw. There is no material build-up in the extruder feed section, and the throughput rate is determined by the total feed streams, and not the extruder screw speed. Additives, other resins, reinforcements, etc., can be added to the extruder melt at several downstream locations by either gravity feeding or with a side feed extruder. Liquid additives are introduced downstream with the aid of a liquid feed pump and liquid injection system. Each component is metered in the correct ratio to the total extrusion rate. The advantage of starve feeding is that all the formulation ingredients are fed in the proper ratio directly onto the extruder. Feed problems due to bridging or slippage in the extruder feed hopper are typically eliminated. Feeders are normally set up directly above the feed opening or on a mezzanine above the feed opening to deposit materials directly onto the screw. Figure 9.2 shows a typical starve-feeding setup with two feeders on a mezzanine.

Conical counterrotating twin screw extruders are normally starve fed, similar to other twin screw extruders.

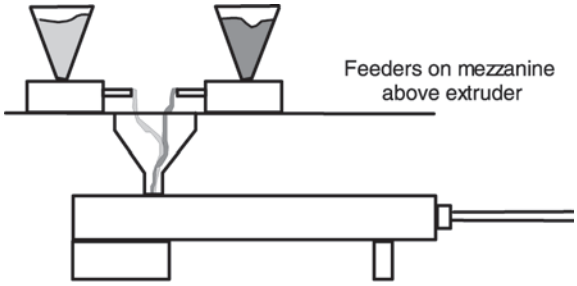


Figure 9.2. Starve-fed extruder.

Additives, reinforcements, fillers, etc., can be fed downstream into the polymer melt using a stuffer or side feed extruder. The side feed extruder rate is normally run in a starve-fed mode for additive rate control. Side feeders are either single or twin screw extruders with zero-compression screws that positively convey ingredients into the melt stream. This positive material conveying in side feed extruders provides higher downstream addition rates than by gravity feeding into a downstream barrel. When using high secondary feed rates downstream, a proper screw design is necessary to handle the high rates and accomplish the extrusion objectives of melting (if required) and mixing.

9.5 Extrusion

After feeding, polymers are melted, conveyed forward, melt mixed, and formed into a shape. These five operations within the extruder will be discussed in detail in later chapters. Proper operation in each extrusion stage is essential to produce acceptable product in high yield with proper aesthetics and the correct property balance. Figure 9.3 shows a twin screw extruder with side feed extruder, vent, and barrel cooling.

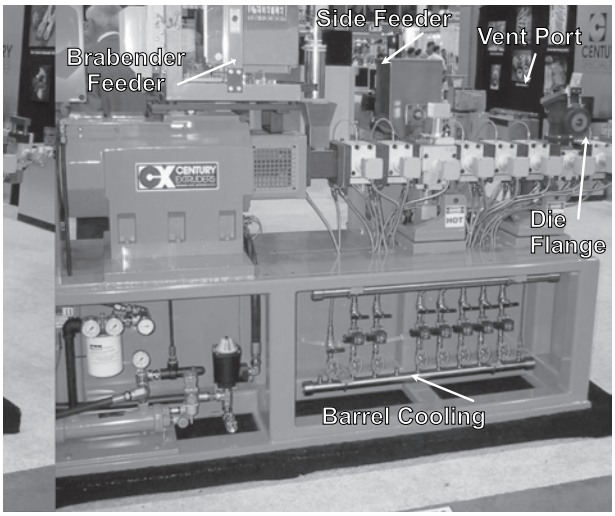


Figure 9.3. Century twin screw extruder.

9.5.1 Shaping and Drawing

The last step in the extruder shapes the extrudate into the desired cross section. As the extrudate exits the die, the polymer molecules, which were oriented in the land area, relax and reentangle, causing die swell, which is more properly called extrudate swell. If the extrudate is allowed to drool out the die, the cross section will swell and become larger than the die opening due to the molecular relaxation. Pulling extrudate away from the extruder, with a puller farther down the line, orients the polymer molecular chains in the machine direction or in the puller direction. Neck down or extrudate draw down is induced by this pulling action. The draw depends on the puller speed relative to the extruder output. Draw ratio is directly related to molecular orientation, resulting in higher tensile and flexural properties in the machine direction compared to the transverse position.

With a given die cross sectional area, there is only one puller speed to extruder throughput rate ratio that produces a product with the correct cross sectional dimensions. If the extruder throughput is increased, the puller speed must be increased proportionally to maintain the same finished product dimensions. Likewise, if the throughput is decreased, the puller speed must be decreased proportionally to maintain the same finished product cross sectional area. The draw ratio and molecular orientation can only be increased or decreased by changing the die cross sectional area relative to the puller speed, assuming the final product dimensions are kept constant. This is easily done with sheet dies, cast film dies, or blown film dies. However, profile dies may have a fixed cross section opening that is not adjustable; at a given throughput rate there is only one puller speed that yields a product with the correct final dimensions. A product that tends to crack in the machine direction (in the plant or in field applications) may have too much molecular orientation, and a new die with a different cross sectional opening is required to alter the draw ratio and correct the problem. Higher draw ratios increase the tensile and flexural properties and the tendency to crack or split in the machine direction. Assuming most polymer molecules are aligned in one direction, it is easy to slit the product in that direction, because there are fewer molecules in the perpendicular direction to hold the product together.

Die swell, shown in Fig. 9.4, is not always visible at the die exit because the extrudate is pulled away from the extruder, causing draw down or neck down. If the extrudate is allowed to drool on the floor or is pulled from the extruder very slowly, die swell is obvious. Polymer molecules in the die land are oriented in the flow direction. The extrudate velocity profile is higher in the center and lower near the die walls. Immediately after exiting the die, the extrudate velocity profile is identical across the entire cross section. Consequently, the velocity at the extrudate surface outside the die is identical to the velocity in the extrudate center. This change in the flow velocity profile gives rise to molecular relaxation outside the die and the resultant extrudate swell.

As the extrudate exits the die, it is quenched and possibly sized to maintain its final shape. Depending on the extrusion process, different methods are available to quench the final product. Cast film and sheet are quenched on rolls; blown film is quenched in air in a blown film tower; profiles, pipe, and tubing are quenched in calibration tanks filled with water and in some cases connected to a vacuum system; strands and monofilament are quenched in water baths; wire coating is done horizontally in air or water; and large part blow molding is quenched in molds.

9.5.2 Solidification and Cooling

Extrudate cooling is normally done by water, air, or contact with a cold surface. Semicrystalline polymers, e.g., polyethylene, polypropylene, nylon, and PBT, have very sharp melting points and consequently very sharp solidification temperatures. Amorphous polymers, on the other hand, do not melt but enter a rubbery state above their Tg (glass transition temperature, discussed later). As the temperature increases, polymer chain mobility continues to increase until the polymer flows and is easy to process. When cooling amorphous polymers, the temperature needs to be below the material Tg to freeze the final part dimensions. Thick cross sections can form a surface skin with the center still being molten. This allows the extrusion line to be run at higher rates. However, if product dimensional tolerances are very tight, the entire product should be cooled below the melting point if it is a semicrystalline polymer and below the Tg if it is an amorphous polymer. Cooling from elevated to room temperature after the product is completely solid results in additional product shrinkage and dimensional changes.

Proper part cooling is critical to produce warpage-free parts with acceptable dimensions and performance.

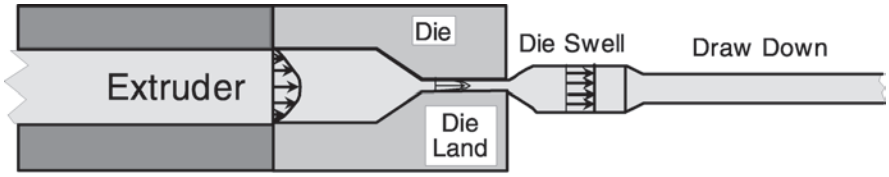


Figure 9.4. Die swell and draw down.

Part warpage is caused by differential shrinkage. To minimize differential shrinkage, the part must be cooled uniformly on all sides. If one side or area of the extrudate solidifies before another, the part will warp, bending toward the side that solidified last. If one side of the extrudate is dragged over an object in the cooling operation, molecular orientation is induced on that side, causing it to shrink differently from the other side and leading to warpage. Warpage is discussed in more detail in Part 4, "Troubleshooting the Extrusion Process."

Cooling rates with semicrystalline polymers are critical to develop the correct amount and crystal size in the final product. Rapid quenching leads to small crystal development and low crystallinity levels. Later heating or annealing (heated for a specific time and temperature above the T_g) leads to additional crystal growth in the solid state. Accompanying this increase in crystallinity is a reduction in volume, a change in the part dimensions, and possibly warpage. To maximize crystallinity, the extrudate should be cooled slowly. Cooling rates can be critical in maximizing product performance and reproducibility. Cooling rates are determined by throughput rates, part thickness, and cooling medium temperature (water bath, roll temperatures, or air temperatures).

Drawing products in their solid state (monofilament production, oriented film, or biaxially oriented film) maximizes molecular orientation and directional properties. In semicrystalline polymers, drawing can lead to additional crystallinity development through polymer molecule alignment.

With some extruded products, the cooling rate and treatment during cooling are critical to obtain the final product properties required by the customer. In sheet or cast film extrusion, roll temperatures and surfaces determine the product aesthetics. Highly polished rolls run at relatively high temperatures produce polished, glossy surfaces. A matte finish on the product is attained by using rolls with a matte finish. For a matte finish on one side and a polished, glossy surface on the other, a matte finish and a highly polished roll are used together. A vacuum sizing tank is used for hollow profiles or pipe and tubing, where the extrudate is run through sizers under water with a vacuum above the water. The fixturing type and cooling required to maintain final dimensions depends on the application.

9.5.3 Puller

The puller controls the draw and tension on the material from the extruder exit through the cooling and solidification steps. Final product dimensions are controlled by the extruder throughput rate and the puller speed. With a fixed die opening and given throughput rate, there is only one puller speed that produces the correct dimensions in the final product. Consequently, the puller speed must be matched to the extruder output rate. If puller speed or extruder output varies, the product dimensions will change continuously. Slippage in the puller can cause thicker sections or parts that do not meet finished product specifications. A caterpillar type puller is shown in Fig. 9.5.

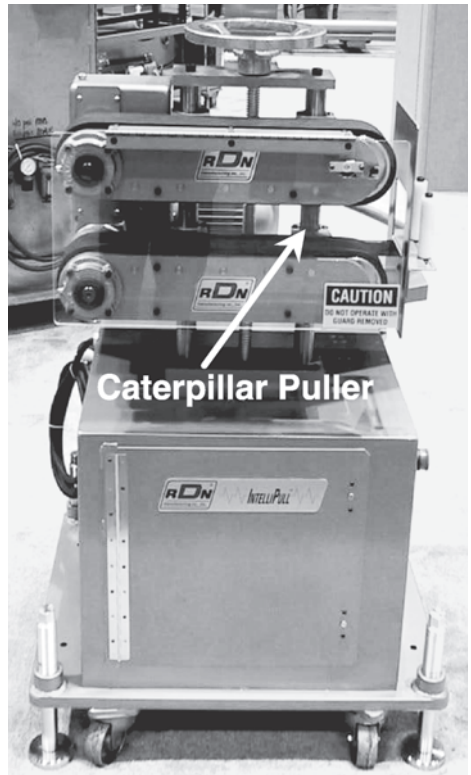


Figure 9.5. RDN puller.

Pressure exerted by the puller must be sufficient to prevent product slippage in the puller, but low enough to prevent part distortion or creating marks on the product surface. Extreme puller pressure can crush the final part, rendering it useless. The puller may be a long distance from the extruder; however, it must be properly aligned with the extruder to prevent the part from being pulled in one direction or another, inducing molecular orientation leading to warpage.

Dimensional variations in the final product normally result from the extruder (surging, power input variations, slippage on the screw, poor feeding) or the puller (slippage, improper compression of the part, or power input variations).

9.6 Secondary Operations

Numerous secondary operations are performed in-line to minimize product handling and improve production efficiencies. Some in-line secondary operations

include cutting to length, drilling or punching holes, corona or flame treatment, decorating (painting, printing, gluing something to the surface), attaching adhesive labels, and welding.

9.7 Inspection, Packaging, and Shipping

Visual part inspection or gauging is done at the end of the line to verify that all parts meet specification. Using statistical process control (SPC) quality control guarantees that all products meet specification, and visual inspection can be eliminated. The problem associated with visual inspection is some defective parts always pass through the system from human error. Visual subjective inspection needs to be eliminated as much as possible to ensure that all parts meet specifications. In addition to visual inspection, part weight and/or dimensions can be checked prior to packaging. Proper SPC procedures reduce much of the QC work associated with product quality assurance.

Samples need to be kept from each production lot for physical property verification or for evaluation if there is a customer complaint.

The final steps in the extrusion process are to package the product according to customer requirements and ship.

Review Questions

1. What controls product dimensions?
2. What part of the extrusion process contributes to final part warpage?
3. Name three factors that lead to part warpage.
4. What are the different methods of feeding polymer to a twin screw extruder and what controls the extruder feed rate in each method?
5. Explain what a side feeder is and does.
6. In the extrusion process, which steps affect physical properties and how?
7. What materials need to be dried prior to extrusion?
8. What are some methods of blending polymers? Which method should be used for blending powder and pellets, pellets and pellets, powder and powder?
9. What is a masterbatch?
10. What is die swell?
11. If the production rate is 300 pounds/hour and it is necessary to feed 0.07% of component X, 2% of component Y, 1.2% of component Z with 96.73% of polymer pellets L, what is the best way to mix and add the material to a twin screw extruder?
12. At what step in the extrusion process is the product worth the most money?

10 Extruder Safety

All employees, from the janitorial staff to company president, are responsible for safety. Each employee must work safely and assist other associates to operate safely, endeavoring to eliminate all unsafe acts that lead to accidents. Ninety-six percent of all accidents are caused by human error, carelessness, or the attitude, "It won't happen to me." Consequently, our personal safety plus the safety of those around us are each employee's responsibility. It is essential to obey all work area rules and be alert for unsafe acts and conditions. Before a job is started, it needs to be thought completely through and determined if it can be done safely. If it can't be done safely, don't do the job until you obtain the proper equipment or develop the proper procedure to do the job safely. It is important to realize the hazards associated with each job and not take any shortcuts that might put you or your associates in the way of potential danger and serious accidents.

New employees must be trained about safety and about equipment hazards. In addition to potential equipment hazards, new employees need to know

- Who to contact in an emergency
- What different alarms mean and what the proper response is
- Location of all safety equipment (fire extinguishers, fire blankets, first aid)
- Where the muster point is
- The correct paperwork procedures in the event of an accident or injured employee
- Who to contact
- The procedure for reporting accidents
- Their responsibility in the situation

Associates need to help each other. If you see fellow employees performing an unsafe act, tell them and help them understand why it is unsafe. This is an act of caring and concern for our fellow employees, not an act to belittle or make someone look foolish.

The most important step in safety is to understand the potential hazards, to realize you are not invincible and it *can* happen to you. Follow procedures and think any job through thoroughly before starting to evaluate the potential for injury to yourself and others. Don't be the bull in the china shop, charging ahead without thought. **If a job can't be done safely, don't do it until procedures, methods, or equipment are available to do it safely.**

10.1 Hazards Associated with an Extruder

The three biggest potential safety hazards associated with extruders are burns, electrical shock, and falls. Without proper protective equipment, burns can be commonplace for employees working around extruders. Burns are normally caused by touching a hot die or unprotected twin screw extruder barrel sections and handling extrudate without gloves. Long sleeves with properly approved thermal gloves should be worn when working around the die, changing the die, tightening die bolts, or other functions performed on the die. If insulation is placed around the die, make sure it is in good shape and properly installed. Barrel sections around vents and downstream feeding ports can be exposed and are hot. Proper protective equipment will prevent burns when working around an extruder. Hot extrudate from the extruder will stick to your skin. Since polymeric materials are great insulators, after sticking to the skin they cool very slowly, continuing to burn the skin. **Never stand in front of a die when a twin screw extruder is starting up. Always wear safety glasses and a face shield when working around molten polymer, especially on start-up, when something unexpected can happen.** Sometimes air in the extruder and possibly gas from degraded products (if the extruder has been sitting at temperature with material in the barrel) can spit out of the extruder on start-up. If some polymer is left in the barrel, the air at start-up becomes compressed, building up pressure that blows hot polymer out of the die. Standing in front of the extruder creates an excellent opportunity for any molten polymer blown out of the die to land on you and burn you. Polymer can stick to gloves, where it holds heat for a long time and can burn you through the gloves if the proper type is not used. When removing the die and/or screw from an extruder (they are normally hot), wear the proper protective equipment (heavy duty gloves and protective thermal sleeves) to prevent burns. Dies can be heavy; therefore, a back brace or other equipment to lift and hold the die can prevent back injuries.

The potential for electrical shock exists. Check the wires to the heaters on the barrels, die, and adapters to ensure that there are no frayed, bare, or exposed wires that can cause electrical shock. In some extrusion processes, water-cooling baths are very close to the die, which can create additional electrical hazard. Unless properly trained, operators should never remove guards, exposing electrical terminals on heaters, or open electrical cabinets to solve electrical problems.

The third major potential safety hazard around extruders is falls. Pellets spilled on the floor are slippery and need to be removed immediately. At start-up the extruder normally generates some scrap, which may be on the floor. This creates tripping hazards, and the scrap must be removed immediately. Occasionally processing issues arise at start-up, leading to a large quantity of

material on the floor around the extruder. In these situations, the extruder should be shut down, the area cleaned, and the extruder restarted. Some extrusion processes use water for cooling. Water spills on the floor should be removed with a wet/dry vacuum or squeegeed to a drain. Wet floors are very slippery and can cause falls.

The most dangerous area around an extruder is the exposed screw turning in the feed throat. **Never, never stick your hands or fingers into the extruder feed throat.** If the screw is turning, there is incredible power that can quickly remove a finger. If the feed throat is hot, you may also get burned.

The most dangerous time during extruder operation is at start-up. An extruder is a pressure vessel. Material is being fed into one end, with a positive conveying mechanism (screw) operating at high horsepower. If the die is blocked with solid plastic or contaminants, pressure can build up very rapidly, generating high pressure (>10,000 psi/68.95 MPa) at the extruder discharge. At start-up, use a low screw speed and feed polymer (starve feed) at a very low rate while monitoring the die pressure until polymer is flowing freely out the die. Once die flow is established, the screw speed and feed rate can be safely increased. As mentioned previously, never stand in front of an extruder during start-up in the event molten plastic is blown out of the die under high pressure.

Twin screw extruders are equipped with pressure gauges (discussed later) that will shut the extruder off in the event excessive die pressure is present. Verify that the pressure gauges are functioning properly. Die pressure shut-off limits can normally be set on the control panel. Once the pressure limit is reached, the extruder will automatically shut off and cannot be restarted until the pressure is removed.

Each extruder should be equipped with a fume hood at the die or vent port to remove any fumes generated by the extruder.

10.1.1 Hazards Associated with Takeoff Equipment

The safety hazards associated with takeoff equipment depend on the extrusion process and takeoff equipment. Pinch points associated with nip rolls, pullers, and roll stacks are one potential safety hazard requiring careful operation. If two operators are running equipment containing nip rolls, they must communicate to verify that all operators are clear when nip rolls are closed. Loose fitting clothing can be caught in nip rolls or pullers. Some lines have either rolling knives or straight knives to slit the edge. Knives need guarding, and operators must use caution when working in the knife area.

High-speed rotating rolls present special hazards. Guards around all rolls and nip points need to be kept in place to prevent injury. Arms, fingers, and hands can easily be pulled into high-speed rolling equipment, causing severe personal injury or dismemberment.

Scrap from start-up lying on the floor poses a tripping hazard. This needs to be picked up and disposed of as soon as the line is running. In the event scrap becomes a hazard due to start-up problems, the equipment needs to be shut down, the area cleaned, and the line restarted.

Noise above 80 decibels (dB) requires hearing protection be used by all people in the area. If noise level is below 80 dB, employees may still want to wear hearing protection to prevent long-term hearing loss.

As you would with the extruder, identify potential safety hazards associated with the takeoff equipment. Form an action plan to avoid any potential hazards. Know where all emergency stop buttons are located and verify that they work. Don't take the approach, "**It won't happen to me.**"

10.2 Personal Protective Equipment

Personal protective equipment exists to make jobs safer. Determine what equipment is required to do the job safely and use it. Following is some of the personal protective equipment available:

- Safety glasses with or without side shields
- Safety shoes
- Ear protection
- Gloves
- Thermal insulated gloves for hot applications
- Long sleeves
- Hard hats
- Face shield
- Goggles
- Back brace
- Wrist brace
- Floor mats

10.3 Lock-Out, Tag, and Clear Procedure

Anyone working on equipment needs a personal lock with his or her name on it and the only key. Prior to doing maintenance or other work on the equipment, turn off the power switch and lock out the switch with your personal lock. Employees working on the line need to attach their own lock. After locking out and tagging the equipment prior to doing any work, each worker attempts to start the equipment to verify that it is off. Once the maintenance or other work is completed, each worker removes his own

individual lock before the equipment can be restarted. This procedure prevents somebody from getting hurt while working on equipment when another person inadvertently starts the equipment, thinking all line work has been completed.

10.4 Proper Training

Don't run any equipment without proper training and an understanding of the potential safety hazards associated with the operation. Knowing where you can get hurt plus understanding new equipment and how the control panel works is essential. Part of all training includes

- Start-up and shutdown procedures
- Understanding all the caution or warning signs on the machines
- Operating time on the equipment with an experienced operator

10.5 Inspection and Housekeeping

Before each shift, evaluate the operating area and plant in general, looking for unsafe conditions, e.g., tripping hazards, exposed wires, and water on the floor. Determine what you are going to do on your shift and review the operation for safety.

Good housekeeping is directly related to safety. A cluttered, dirty area will lead to accidents and reflects your attitude toward the job. A proper storage area for all tools and equipment makes the job easier and the plant a better place to work.

10.6 Material Safety

Understand the materials you are using by reviewing the Material Safety Data Sheets (MSDSs). Improper operating conditions or purging with the wrong materials can have serious consequences. Over-heating PVC generates hydrochloric acid (HCl), which attacks the lungs as well as plant equipment, causing rust. Extruding acetal (polyoxymethylene) followed by nylon, or vice versa, generates formaldehyde when the two components are mixed in an extruder. Acetal followed by PVC, or vice versa, also generates formaldehyde when mixed in an extruder. Acetal will also react with fluorinated polymers or ionomers to give off formaldehyde.

PVC has limited thermal stability and should not be left in a hot extruder. PVC degrades in an autocatalytic reaction, generating HCl. Proper purge material should be available to remove PVC from the barrel if the extruder is going to be shut down for an extended period. Operators who have the flexibility to change extruder temperature profiles need to understand the upper processing limits when extruding PVC or other temperature-sensitive polymers.

Review Questions

1. What is the most dangerous time during extrusion and why?
2. Where are the most dangerous locations around an extruder and why?
3. What are some potential hazards associated with extrusion?
4. What is the "lock-out, tag, and clear" procedure and when should the procedure be used?
5. Why is housekeeping important?
6. What is a near miss?
7. What hazards are associated with takeoff equipment?
8. What are some types of personal protective equipment?
9. What materials should not be mixed with acetal in an extruder?
10. What happens if PVC is overheated?

11 Twin Screw Extruder Equipment

There are many twin screw extruders commercially available. The one to use depends on the end-use application. Different models have two parallel screw shafts that either rotate in the same direction (called corotating) or rotate in opposite directions (called counterrotating), with varying distances between the screw shafts. If the centerline distance between the shafts is less than the screw diameter, the screws are called intermeshing, while screws with a distance between the shafts equal to the screw diameter are nonintermeshing. Figure 11.1 shows a short segment conveying screw element with parallel corotating and counterrotating screws that are fully intermeshing. In nonintermeshing extruders, the screw lengths of the two shafts can be equal or one screw can be longer than the other to provide better pumping capability to the die. Another common twin screw extruder is a conical, where the counterrotating, intermeshing screws are tapered rather than parallel. Figure 11.2 shows the screw geometry in a conical extruder. This chapter describes the different twin screw extruders, their hardware, and how they differ from single screw extruders.

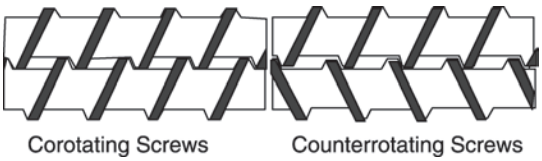


Figure 11.1. Corotating and counterrotating fully intermeshing screws.

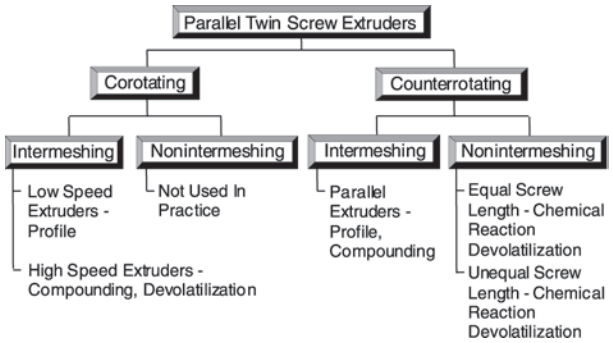


Figure 11.3. Types of parallel twin screw extruders.

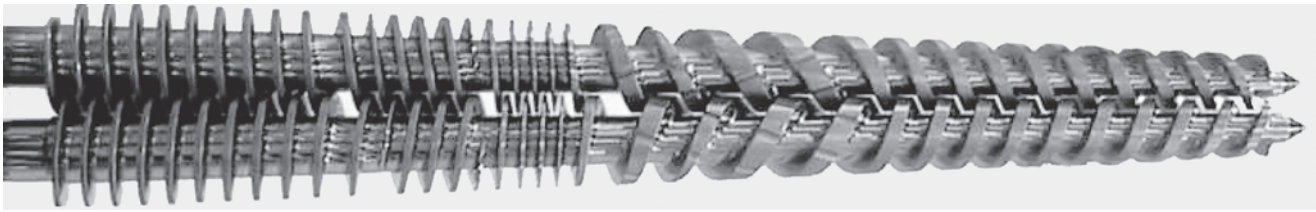


Figure 11.2. CPM conical counterrotating twin screw extruder screw shafts.[1]

The principal differences in parallel intermeshing and nonintermeshing twin screw extruders depend on whether the screws are rotating in the same direction, corotating, or in opposite directions, counter-rotating, and the distance between the screws. Figure 11.3 summarizes the different parallel twin screw extruders and applications where the different twin screw extruders are used. High-speed, corotating twin screw extruders are used for compounding resin with additives (colorants, fillers, flame retardants, reinforcements, stabilizers), devolatilization to remove solvents, and reactive extrusion (chemical reactions done in situ in the extruder). Low-speed corotating and counterrotating extruders are used to produce profiles and pipe. Counterrotating twin screws are used for compounding PVC and other resin systems. Nonintermeshing, counterrotating extruders are principally used for devolatilization and chemical reactions, i.e., grafting, polycondensation, addition, controlled cross-linking, and functionalization. Twin screw extruders are finding homes in sheet and film extrusion, where different formulation ingredients can be compounded and formed in the same extrusion. This eliminates compounding and reextruding to produce a final part.

Figure 11.4 identifies four of the five major equipment components (drive, feed, screw and barrel, and the die or head) of a parallel twin screw extruder. The fifth component is the control cabinet. The drive system is composed of a DC motor, cooling system for the motor, coupling between the motor and gear box, thrust bearing, gear box, oil lubrication and cooling for the gear box, and shaft coupling between the gear box and the extruder screws. One feed port is located at the rear of the extruder in the first barrel section. Additional feed streams can be added in numerous locations along the barrel length through gravity from a volumetric or gravimetric feeder, liquid feed using

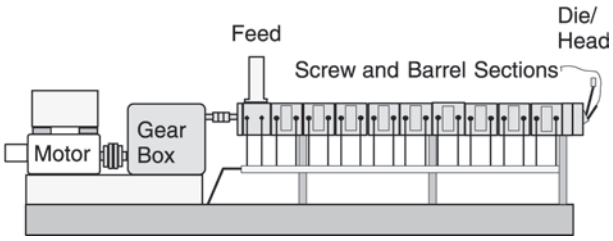


Figure 11.4. Corotating and counterrotating fully intermeshing screws.

a pump with a liquid injection nozzle, and/or a side feed extruder or stuffer to add polymer, additives, fillers, or reinforcements at locations along the barrel. Screw and barrel sections are both modular. Barrel sections can be added or removed to make the extruder barrel longer or shorter to increase or decrease compounding capabilities, depending on the product application. A set of screw shafts is required for each extruder length. Additional barrel sections are normally added to increase process flexibility for downstream feeding or venting. Each barrel section is normally cooled with water and heated with cast heaters to control barrel temperature. Screws are modular, with different elements combined in a strategic design to localize the feeding, melting, conveying, mixing, pumping, and venting at specific locations along the extruder barrel. Screw designs are easily changed or modified to optimize the processing, depending on the materials being fed and the product requirements. The adapter between the extruder and die can be equipped with a screen pack to generate additional backpressure or for melt filtration. Figure 11.5 shows an automatic screen changer and gear pump between the head and the die in a sheet extrusion line. Melt temperature and pressure transducers are located at the extruder head to monitor and/or control the process. Figure 11.6 identifies individual elements in a twin screw extruder.

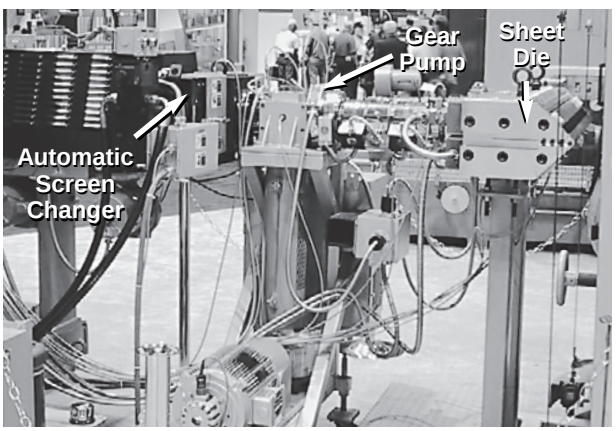


Figure 11.5. HPM setup with gear pump and automatic screen changer.

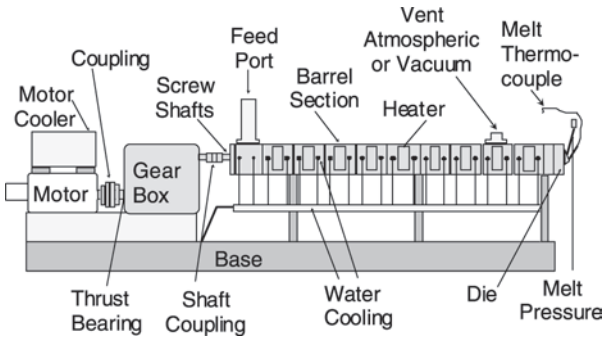


Figure 11.6. Identification of equipment elements comprising the major components in a twin screw extruder.

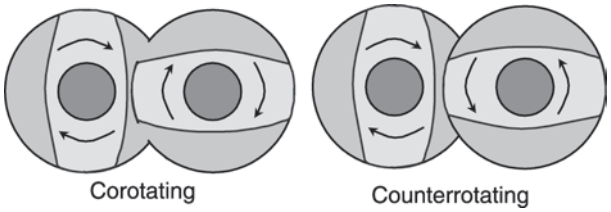


Figure 11.7. Screw rotation in corotating and counterrotating extruders.

The difference between corotating and counterrotating twin screw extruders is the screw rotation relative to each other. In corotating extruders, both screws rotate in the same direction, while in counter-rotating extruders, one screw rotates clockwise and the other screw rotates counterclockwise. With corotating screws, both screws have either right-handed or left-handed thread, depending on the screw rotation (CW vs. CCW). In counterrotating extruders, one screw has a right-handed thread and the other screw a left-handed thread. Figure 11.7 shows the corotating and counterrotating screw rotations viewed from the end of each element from the die end. Polymer flow in a fully intermeshing, corotating extruder makes a figure 8 pattern, as the material does not pass between the screws. This generates high- and low-pressure regions for the material near the extruder apex, as shown in Fig. 11.8. Polymer flow in a counterrotating extruder is forced between the two screws, resulting in a high-pressure region at the nip, where the material is being forced between the screws, and a low-pressure region at the nip exit. Figure 11.9 demonstrates the flow in a counterrotating extruder with the high- and low-pressure areas. Figure 11.1 showed the intermeshing region for corotating and counterrotating screws.

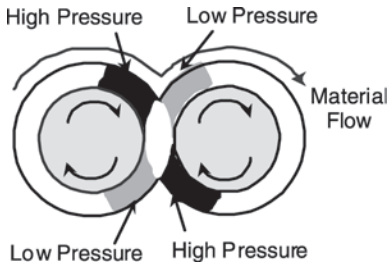


Figure 11.8. Material flow in corotating screws.

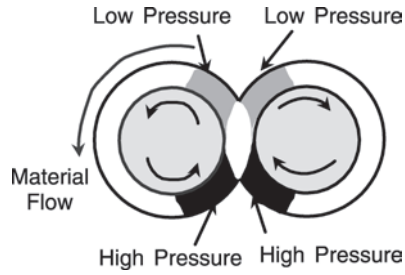
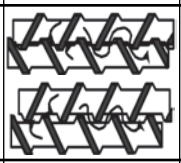
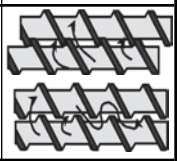
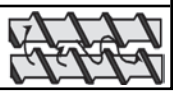


Figure 11.9. Material flow in counterrotating screws.

In addition to identifying twin screw extruders based on screw rotation, the distance between the two screws varies from fully intermeshing to nonintermeshing. Material flow and the shear generated depend on the

Table 11.1. Comparison of Intermeshing and Nonintermeshing Twin-Screw Extruders[1]

Screw Distance	Material Flow	Counterrotating	Corotating
Fully Intermeshing	Closed to Length and Cross		Impossible
	Open to Length Closed to Cross	Impossible	
	Open to Length and Cross	Possible, Not Practical	Kneading Blocks and Gear Mixers
Partially Intermeshing	Open to Length Closed to Cross		Impossible
	Open to Length and Cross		
Nonintermeshing	Open to Length and Cross		

intermeshing.[2][3] Screw designs (Table 11.1) are defined as either open or closed, based on whether material can flow in a particular direction. If material can flow in an axial or longitudinal direction from the feed throat to the die, the screw is open in the length direction. Theoretically material can move from one channel to the next channel, allowing flow in the lengthwise direction. If the screw is closed to material flow in the axial direction, the length is considered closed. Figure 11.10 shows a partially intermeshing, counterrotating screw that is open to flow in both the axial direction and cross-machine direction. In the cross-machine direction, the channel is considered open if the material can flow around a particular screw channel. Figure 11.11 shows partially intermeshing, corotating screw elements with material flow both lengthwise in the axial direction and across the extruder

Figure 11.10. Counterrotating screw open to flow in both length and across.

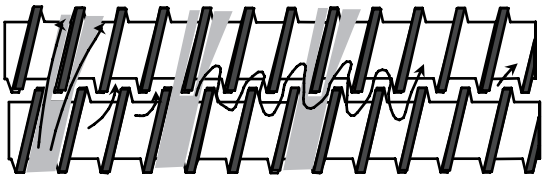


Figure 11.11. Corotating partially intermeshing screw length and cross-direction open to material flow.

as the material passes between the screws. With fully intermeshing, corotating screws, material cannot pass between the screws (refer to Fig. 11.8). One screw is rotating down and the other screw is rotating up, preventing material cross-flow. Table 11.1 shows the various twin screw extruder configurations and those screw configurations that allow material flow axially and across the channel and those that don't. Normal leakage flow caused by the requirements for mechanical clearance between the two screws is not considered in the material flow behavior. Whether the screw configurations are open to cross or lengthwise flow is directly related to the conveying, mixing, and pumping efficiency in a particular extruder.

If the axial length is closed, the pathway down the screw is divided into isolated areas with no opportunity for flow in the axial direction. Open cross-flow allows a material path around the screw and the polymer flows in the radial direction, as it is transferred back and forth between screws. If the cross-direction is closed, material cannot flow between adjacent screws, resulting in no flow in the radial direction. When both the length and cross-flow are open, good distributive mixing along with poor pressure generation result. Good distributive and dispersive mixing but poor pressure generation occur with the length open and the cross-material flow closed. If both the length and cross-flow directions are closed, good dispersive mixing results with good pressure generation. Dispersive and distributive mixing are discussed in more detail in Chapter 12. Nonintermeshing twin screw extruders are open both axially and across the barrel, regardless of whether the extruder is corotating or counterrotating.

Table 11.2[4] compares processing parameters for the three basic parallel twin screw extruders. The various

Table 11.2. Comparison of Parallel Twin-Screw Extruders

	Corotating Intermeshing	Counterrotating Intermeshing	Counterrotating Nonintermeshing
Practical Residence Time, minutes	0.35–6	0.35–6	0.35–6
Residence Time Distribution	Variable	Variable/ Tighter	Variable
Dispersion	High	High	Good
Heat Transfer	Excellent	Excellent	Excellent
Venting	Excellent	Excellent	Excellent
Pumping	Good	Excellent	Fair
Self-Wiping	Excellent	Good	Fair
Zoning	Excellent	Excellent	Good
Output Rate	High	Moderate	High
Distributive Mixing	Good	Good	Excellent

processing parameters compared are defined below. These will be discussed in more detail later.

- *Practical residence time* is the time polymer, additives, or other formulation components will spend in the extruder from feed to the die.
- *Residence time distribution* is the shortest to longest time different particles spend in the extruder.
- *Dispersion* is breaking up large particles or agglomerates and uniformly dispersing them throughout the melt.
- *Heat transfer* is the ability of the barrel heaters to transfer heat into the material being processed to create a uniform temperature profile throughout the melt.
- *Venting* is the ability of the extruder to remove volatiles or moisture through a single or multiple vent ports along the barrel length.
- *Pumping* is delivering a uniform melt pressure and material supply to the die.
- *Self-wiping* is one intermeshing screw element removing polymer from the adjacent screw element.
- *Zoning* is where specific areas or zones in the extruder accomplish specific extrusion objectives such as melting, mixing, feeding, etc.
- *Output rate* measures the throughput rate or pounds/hour that can be delivered by a specific extruder size or diameter.
- *Distributive mixing* uniformly distributes all components and melt temperature in the extrudate.

Twin screw extruders for plastics processing have evolved over the past 50 years from extruders with an over-under twin screw arrangement or geometry to a side-by-side configuration, providing better low bulk density powders and other material feeding. Early side-by-side extruders used three-lobe conveying elements (three separate flights around each element) coupled to motors with limited torque and small gear boxes. Three-lobe screw elements have been replaced with two-lobe elements (two flights and screw channels around each element; see Fig. 11.12), increasing the free volume and throughput capacity. Extruder equipment changes accompanying the two-lobe screws are larger motors that generate higher torque, higher torque screw shaft and demand arrangement to transfer the increased power to the screw, and an increased centerline distance, leading to more free volume within the extruder. Combining all the equipment changes, throughput rates have increased dramatically.

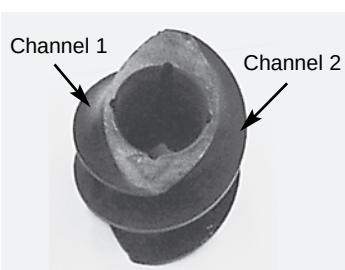


Figure 11.12. Two-lobe element showing two separate channels and flights.

With intermeshing screw elements, the shaft centerline distance and screw diameter determine the free volume, the shear rates, and the characteristic outside-to-inside screw element diameter (D_o/D_i). The conveying element's channel depth is a function of the centerline distance, the outside diameter, and the number of screw flights. The channel depth must provide enough wall thickness between the inside diameters of the screw element (defined in Fig. 11.13) and the shaft to transfer the torque from the motor and shaft to the screw elements. One or two large keyways cut in the three-lobe elements transfer the torque. In a two-lobe system, wall thickness is too thin for large keyways. Different systems have been developed to transfer the torque with the thinner screw element walls. Figure 11.14 shows the typical flight, flight clearance between the top of the flight and the barrel wall, and channel depth. The open channel depths between the screw and the barrel wall (less any channel space occupied by the intermeshing flight of the adjacent screw element) plus the space between the kneading blocks or other elements and the barrel wall make up the extruder free volume. Extruder free volume is a measure of the space available in the barrel for pellets, powder, reinforcements, fillers, and/or polymer melt with a specific screw configuration. Extruder free volume is given by Eq. (11.1):

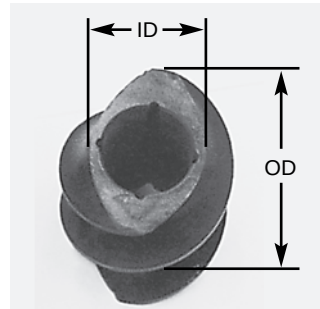


Figure 11.13. Screw element inner and outer diameters.

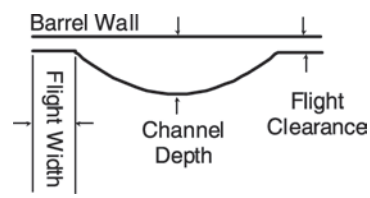


Figure 11.14. Channel and flight geometry.

$$V_0 = A_0 \times L \tag{11.1}$$

where V_o = Free volume
A_o = Screw element open area
L = Screw length

A_o can be calculated for each screw element using Eq. (11.2):

$$A_0 = A_b - A_s \tag{11.2}$$

where A_b = Barrel area
A_s = Screw area

Greater free volume gives the extruder higher throughput capacity.

Free volume is directly related to the ratio D_o/D_i and the shaft centerline distance. Figure 11.13 shows a two-lobe element with the outer and inner diameters defined. The free volume is balanced versus the extruder torque. Screw shafts must be able to transfer the available motor torque through the screws to process a specific quantity of material. As the extruder free volume increases, more open space is available in the extruder barrel, resulting from deeper screw channels. With deeper screw channels the metal between the channel bottom and the screw shaft decreases, providing less area to transfer the motor torque to each screw element. Consequently, the outer-to-inner screw element diameter ratio reaches an optimum value. At low D_o/D_i there is very little free volume in the extruder, and the distance from the screw shaft wall to the channel bottom is fairly large. In this scenario high torque can be transferred from the motor to the screw, but only limited torque is required because the material in the extruder is reduced. With high D_o/D_i ratios, higher extruder free volume is available, allowing more material to be processed. However, the screw element wall thickness between the channel bottom and the shaft wall has decreased, making it more difficult to transfer the torque required for processing from the motor to the extruder. The centerline distance between the shafts becomes important because smaller distances limit the thrust bearing size in the gear box, reducing the

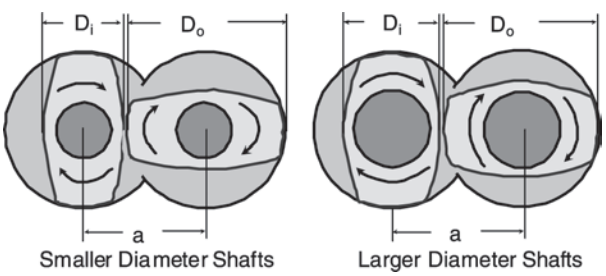


Figure 11.15. Effect of shaft diameter on centerline distance and D_o/D_i ratio.

torque that can be transferred from the motor to the screw shafts. As shown in Fig. 11.15, larger screw shafts that can transfer more torque have less free volume in the barrel and a greater centerline distance. Consequently, less material is processed when larger diameter shafts are present. This becomes a balancing act, as smaller screw shafts lead to more free volume, while the shafts, thrust bearings, centerline distance, and keyways must be able to handle increased torque. Figure 11.16 shows the relationship between torque and free volume versus D_o/D_i.

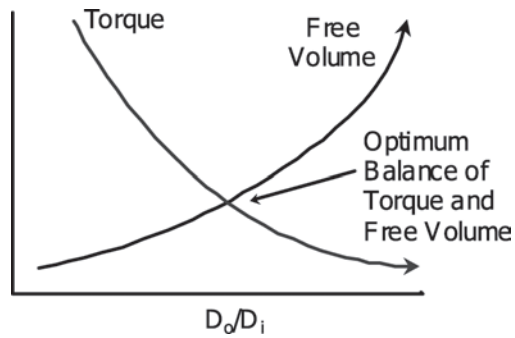


Figure 11.16. Relationship of torque and free volume relative to the outer/inner screw diameter ratio.

As extruders evolved from three-flighted, three-lobe elements to two-flighted, two-lobe elements, the free volume increased. Increasing both the D_o/D_i ratio and the centerline distance raised the free volume more. The free volume relationship relative to the number of lobes and the D_o/D_i ratio of various commercial extruders is shown in Table 11.3. Most corotating twin screw extruder manufacturers today have a centerline distance, a, and flight depth, h, given by Eqs. (11.3) and (11.4), respectively:

Table 11.3. Comparison of Free Volume for Different D_o/D_i Ratios

Type	D_o	D_i	Free Volume
3 Lobe	55	1.26	0.52
2 Lobe	57	1.44	0.86
2 Lobe	58	1.55	1.0
2 Lobe	60	1.54	1.0

$$a = \frac{D_O + D_i}{2} \tag{11.3}$$

$$h = \frac{D_O - D_i}{2} \tag{11.4}$$

As free volume increased, the extruder torque had to increase to process the additional resin capacity. With thinner screw element walls between the screw shaft surface and the channel bottom, the challenge was how to transfer the higher torque provided by larger motors and gear boxes to the screw elements. Early three-lobe machines had one large rectangular keyway on each shaft. With the two-flighted, two-lobe machines, one keyway was still used with two key channels 180 degrees apart on the inside of the screw element. As the inside screw element walls became thinner with the increased D_o/D_i ratio, the keyway was replaced by four round rods (Fig. 11.17), hexagonal or octagonal shafts, or splined shafts, depending on the extruder manufacturer. Figure 11.17 shows the various screw shaft and screw element geometries.

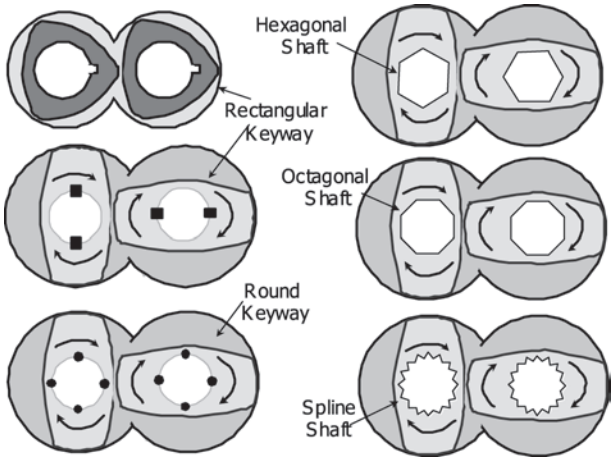


Figure 11.17. Different types of shafts to transfer the torque.

Free volume increases required higher motor torque, as most of the energy to melt plastic is supplied by the motor. As the material in the extruder increases, the energy to process the plastic must increase proportionally. Torque in rotational motion is the power divided by the angular velocity, where the angular velocity is defined as revolutions per minute multiplied by 2π to change the angular velocity to radians/minute. The torque is then calculated from Eq. (11.5):

$$Power = C \times torque \times 2\pi N \tag{11.5}$$

where C = Conversion factor (1 hp = 33,000 ft.-lb./min.)
P = Power given in horsepower (hp)
N = Screw speed given in rpm
Torque = lb.-ft.

Increasing the motor power provides more torque at a constant screw speed. High screw speed translates to more power, allowing more material to be fed per unit time.

Starve-fed twin screw extruders run at high rpm to generate enough power to convey, melt, mix, and pump the polymer to the die. In addition to increased free volume, modern twin screw extruders can run at very high rpm. The specific energy, measured in kW-hr/kg, is related to how much material can be processed based on screw speed and power input. Figure 11.18 shows a hypothetical graph relating power to screw speed. As the screw speed increases, the power goes up, allowing more material to be processed. At low screw speed, there is very little power available; consequently, if the extruder is over-fed, the motor and drive will shut down because the system torque and power requirements are exceeded. This means that when starting a starve-fed extruder, the screw is started and material is fed slowly to the extruder. Once material exits the die, the screw speed is increased before more material is fed to the extruder, while monitoring the torque. If too much material is fed, the power requirement is exceeded and the drive and motor will shut down. The amount of material that can be fed depends on the power available and the screw speed. During operation, the screw torque is constantly monitored and must be maintained below a certain value. The curve in Fig. 11.18 for a specific extruder depends on the motor and the gear box capabilities.

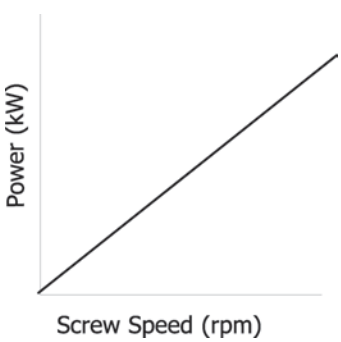


Figure 11.18. Screw speed versus power.

Related to the power and torque requirements is how much the screw flights are filled. Since the extruder is starve fed and the screw speed is running at sufficient speed to remove all the material being fed, most screw elements are not full. (Depending on the screw elements and the screw configuration, some screw sections may be completely filled. This is discussed in more detail in Chapter 13, "Screw Design".) The fill may be approximately 30%, as demonstrated in the conveying screw element channel in Fig. 11.19. If the fill is 30%, the total free volume available between the screw and the barrel wall is 70%, as the polymer occupies only 30% of the total volume. Each conveying channel may be only partially filled with solid or molten polymer, depending on the screw element pitch and the screw configuration. The percentage fill at any location along the extruder screw is a function of the feed rate, screw rpm, screw pitch, screw configuration, and pressure gradients along the screw. In calculating shear rates, the values change depending on the fill. (This is discussed in more detail in Chapter 7.)

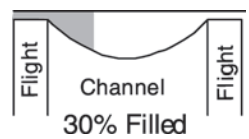


Figure 11.19. Partially filled flight.

Material in the extruder experiences a residence time distribution based on the screw design, type of twin screw extruder, length to diameter (*L/D*) ratio, screw speed, and feed rate. The average residence time is defined by t_{mean} given by Eq. (11.6):

$$t_{\text{mean}} = \frac{V}{Q} \tag{11.6}$$

where V = Free volume used
Q = Volumetric flow rate per unit time

As shown in Fig. 11.20, the residence time distribution can be narrow or broad. Counterrotating, intermeshing twin screw extruders have the narrowest residence time distribution, followed closely by corotating, intermeshing twin screw extruders. Counterrotating, noninter-

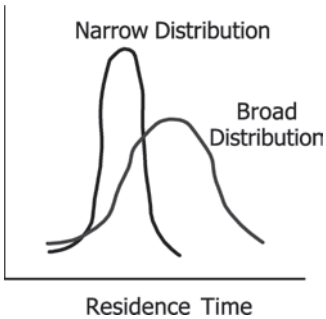


Figure 11.20. Screw speed versus power.

meshing twin screw extruders have the broadest residence time distribution.

The final area to cover before getting into the individual equipment components on the extruder is the nomenclature used to identify different screw elements. In a two-lobe screw design, each conveying element contains two flights, with the elements sometimes called two-flighted elements. Two numbers representing the pitch and element length identify conveying elements. A 45/45 conveying screw element is shown in Fig. 11.21. The first number represents the pitch or the screw element length required for a given flight to make one complete revolution around the element, while the second number represents the element length. Consequently in Fig. 11.21, the pitch is 45 mm or it takes 45 mm for a flight to complete one revolution about the element, and the element is 45 mm long. Other conveying elements with the appropriate nomenclature are shown in Fig. 11.22. The 60/60 conveying element has a 60 mm pitch or it requires 60 mm in screw length for the flight to make one complete revolution about the screw, and the element is 60 mm long. Another conveying element, not shown, is a 90/90, with a 90 mm pitch (requiring 90 mm in screw length for the flight to make one complete revolution) and 90 mm in length. What are the dimensions and pitches for a 40/20 or a 60/30 conveying element? Conveying elements can be either right-handed or left-handed pitch to convey material forward or rearward in the barrel. Figure 11.23 shows a 60/60 right-handed element followed by a 60/30 left-handed conveying element, followed by another 60/60 right-handed conveying element. The left-handed element acts as a melt seal or a place to build pressure in the screw.

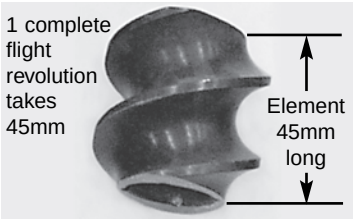


Figure 11.21. 45/45 Theysohn Corp. conveying screw element.

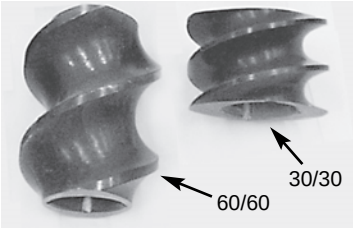


Figure 11.22. 45/45 Theysohn Corp. conveying elements and nomenclature.

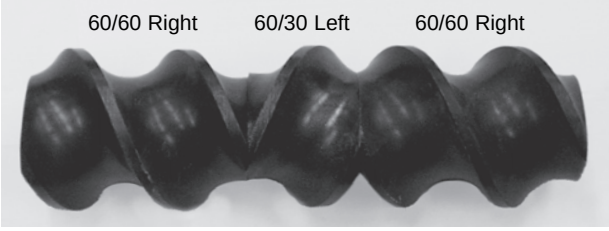


Figure 11.23. Combination of right- and left-handed Theysohn Corp. conveying elements.

Mixing elements, called kneading blocks, have disks in different spatial configurations around the element. An additional number is added to the kneading block nomenclature, indicating the number of disks. A 45/5/30 kneading block has the second disk rotated 45° from the first disk with a total of 5 disks and a length of 30 mm. A 45/5/30 right-handed kneading block is shown in Fig. 11.24. A 45/5/45 kneading block has the second disk rotated 45° with a total of 5 disks in a 45 mm length, shown in Fig. 11.25. Similar to conveying elements, there are right-handed and left-handed kneading blocks, with the disks rotating in either a right-handed or left-handed pattern. Neutral kneading blocks have the second disk at 90° rotation from the first disk. These are designated as 90/3/15, where the second disk is 90° from the first, with three disks and a total length of 15 mm. There are other special screw elements from different machine manufacturers that will be discussed in more detail later.

Figure 11.24. Theysohn Corp. 45/5/30 right-handed kneading block.

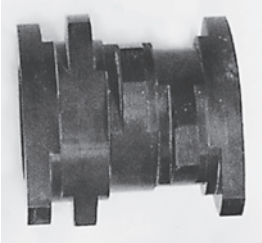


Figure 11.25. Theysohn Corp. 45/5/45 right-handed kneading block.

Some corotating twin screw extruders operate with their screws turning in a clockwise direction, while others operate in a counterclockwise direction. Discussions to this point have assumed a clockwise direction with right-handed conveying elements moving material from the feed throat to the die, and left-handed elements moving material back toward the feed throat. If the corotating screws rotate in a counterclockwise direction, left-handed elements convey material from the feed throat toward the die, while right-handed elements convey material back toward the feed throat. For counterrotating extruders, one screw rotates clockwise and the other counterclockwise. Consequently, the screw elements on the clockwise rotating screw are right-handed elements to convey material toward the die, while the screw rotating counterclockwise uses left-handed conveying elements to move the material toward the die.

Parallel, corotating, intermeshing twin screw extruders are the most common twin screw extruders available on the market today, and they are produced by several different companies. Figure 11.26 shows a picture of a 45 mm Theysohn twin screw corotating extruder with a mezzanine over the extruder to support a range of feeders used to feed product in different feed ports along the extruder. This particular extruder is a high-speed, high-torque extruder capable of running 600 to 1000 pounds/hour, depending on the type of resin and formulation. This picture shows 6 of the 10 barrel sections plus the gear box, motor, and heating and cooling on the extruder barrel. To supply the high torque and screw rpm required at such high throughputs, the motor and gear box take up almost as much space as the barrel.

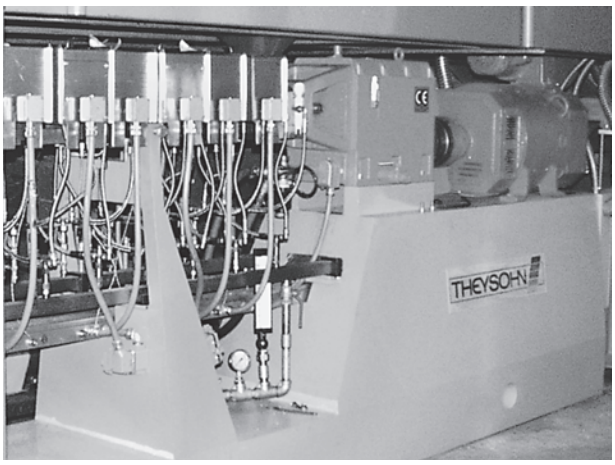


Figure 11.26. Theysohn parallel corotating twin screw extruder.

The following sections will deal with the five major equipment components required to operate any twin screw extruder:

- Drive
- Feed
- Screw and barrel heaters and coolers
- Die
- Controls

These equipment areas are identified in Fig. 11.4.

11.1 Drive

The drive is comprised of a DC motor, motor cooling, coupling between the motor and the gear box, gear box, and coupling between the gear box and the screw shaft. The DC motor provides constant torque to the screw. Figure 11.27 shows power versus screw speed at 100% torque. The gear box and torque transfer through the shaft and screw elements limits maximum power. At 100% torque, the screw speed is directly related to the power. To produce high

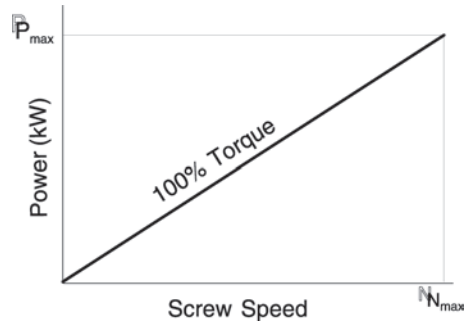


Figure 11.27. Relationship of power, screw speed, and torque.

screw speeds, torque, and throughput, motor horsepower has increased in recent years. Newer twin screw extruders have motors and gear boxes that are larger than the barrels to supply the necessary power. The DC motor is cooled with a top-mounted blower. A filter to remove any particles from the intake air keeps the cooling air contaminant-free. If the filter becomes clogged, preventing enough cooling from reaching the motor, the motor will automatically shut down to prevent damage from overheating. If the filter is dirty and insufficient air is circulating during a run, the extruder motor may shut down, and the cause or reason for shutdown is hard to detect. It is impossible to restart the motor until it cools sufficiently.

Between the motor and the gear box, Fig. 11.28, is a fully interlocked torque limiting coupler that disengages at a predefined torque limit to protect the motor from damage in the event the high torque limit is reached. If the coupler disengages, it is rotated in the reverse direction to reengage. If the torque coupler disengages, the extruder feed systems stop and all hoppers on the extruder are empty. It may be difficult to restart the extruder if it is full, as the screws will not be able to rotate before the extruder reaches maximum torque. Two possible scenarios exist: the screws can be turned a small amount before they reach maximum torque, or the screws cannot be turned at all without reaching maximum torque, causing the torque coupler to disengage. In the event the screws turn, rotate the screws very slowly, shutting down the extruder just before the torque limit is reached, disengaging the coupling mechanism. Continue this process until the screws turn continually at low rpm. As material exits the extruder, the screw speed can be gradually increased as the torque decreases. Once the screw speed reaches the desired level, gradually add polymer and other formulation ingredients to the extruder until the product is running at the desired rate. In the second scenario, where the screws cannot be turned at all before reaching the torque limit, verify that no material or other foreign object is caught in the extruder screws. If there is a foreign object, remove it and start the screws. Assuming no foreign object is found, raise the barrel temperatures until the

screws can be rotated. Once rotation is attained, gradually clear the material from the extruder by the procedure described previously. After the screw speed is back up to full rpm, lower the barrel temperatures and run the process under normal conditions. If the torque is close to the extruder torque limit at the desired throughput rate, the following steps can be taken to reduce the torque level:

- Increase the screw speed.
- Increase the resin melt temperature.
- Decrease the feed rate.
- Change the screw design to generate less shear and torque.

Gear boxes transfer the motor power to the screw shafts. At the entrance to the gear box is the torque limiting coupling, and at the exit is the gear box connection to the screw shaft. Gear boxes have increased in size with higher throughput rates, requiring larger motors and more torque. Oil is pumped to the gear box to lubricate the gears and shafts during rotation in the gear box. The oil must be cooled to prevent thermal degradation, which leads to gear box wear. Some newer extruder models have gear boxes with the ability to run both corotating and counterrotating screws, depending on the application. To change from corotating to counterrotating, another screw shaft with enough elements to construct a left-handed screw is required to intermesh with the right-handed elements on the corotating screw. A second alternative is to have two sets of shafts and screw elements to quickly make the conversion from corotating to counterrotating or vice versa.

Figure 11.28 pictures a corotating twin screw extruder drive with different components identified.

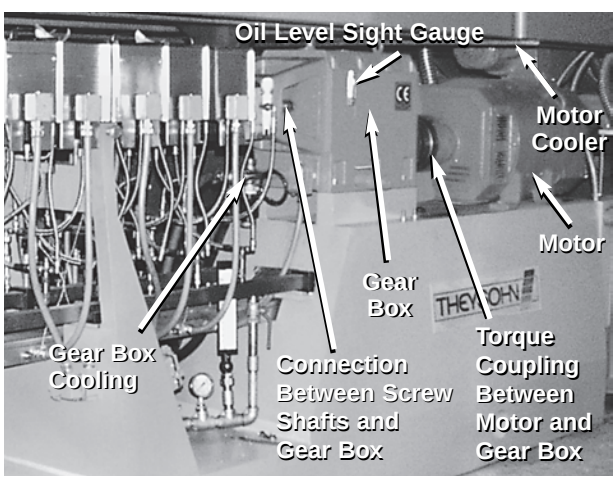


Figure 11.28. Drive system components.

11.2 Feed

Material is starve fed to parallel, intermeshing, corotating and counterrotating twin screw extruders. Some formulation components are fed above the feed opening in zone 1 through a feed pipe or feed hopper. The feeder is supported above the extruder feed throat by a mezzanine or hung from or placed on some other support system above the extruder. Extrusion throughput rates are determined by the feed rate in a starve-fed machine, and not by the screw speed. Screw speed determines the residence time in the extruder and the screw fill. Screw speed and feed rate are balanced to prevent the extruder from overtorquing and shutting down. Feeders are normally either volumetric (run at constant screw speed, vibratory speed, or belt speed) or gravimetric (sometimes called loss-in-weight). Gravimetric feeders deliver a constant weight per unit time as the screw speed, vibrations, or belt move faster or slower to produce the specified throughput rate. Each feeder can deliver a resin, premix, additives, fillers, reinforcements, colorants, or stabilizers to different locations along the extruder. The feeders used depend on the available feed ports, the feeders available, the feeder size and accuracy, and the ingredients. The ideal situation has gravimetric feeders properly sized for each ingredient. At start-up, the individual feed rate for each material stream is gradually increased with the extruder speed until all feeders are set at the desired rate and the screw speed is high enough to provide sufficient torque to run the product.

Figure 11.29 shows a feeder layout on a mezzanine above a twin screw extruder for supplying different ingredients to the extruder below. This configuration shows six individual feeders. Some are large, feeding ingredients at

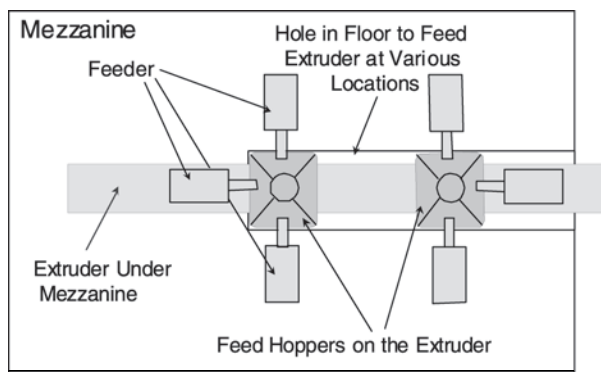


Figure 11.29. Feeder arrangement on a mezzanine.

high rates, while others are small, feeding additives at less than 1%. In many operations this many feeders are not available, and some components are remotely premixed in batches and transported to a particular feeder. Figure 11.30 shows a view from above the mezzanine looking down on the feed hopper in barrel zone 1 and the extruder below. In barrel zones 2, 4, and 6, additional materials can be fed to the extruder by removing the barrel plugs. The actual feed ports used in a particular operation depend on the ingredients, the extrusion objectives, the ingredient temperature stability, how much shear or work the additives require or

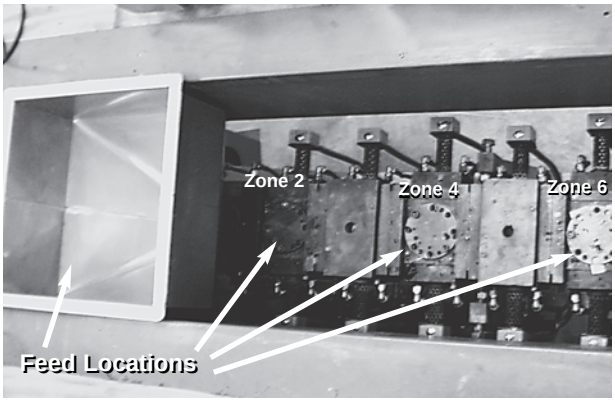


Figure 11.30. View of feed ports looking down on the extruder from the mezzanine.

can absorb, and the screw design. Figure 11.31 shows a volumetric feeder set over the feed hopper in barrel zone 1, ready to feed material to the extruder.

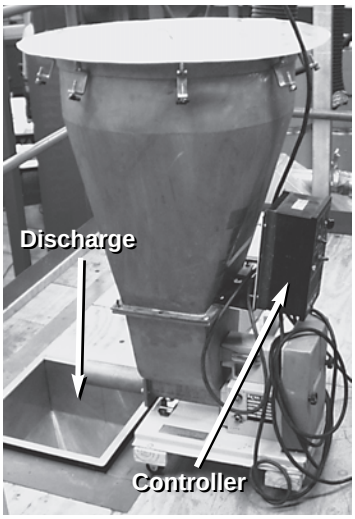


Figure 11.31. Volumetric feed to the extruder feed hopper.

Ideally, solid material fed to a corotating screw is fed over the outside screw that is turning down into the extruder, so the material is conveyed into the barrel. If it is fed on top of the screw that is coming up, the material has to be transferred to the other screw before it can be fed into the extruder. In the transfer from one screw to the other at high screw rpm, resin pellets can bounce out of the extruder throat. In counterrotating extruders, the material is fed over the outside of each screw turning down into the extruder to capture the material and convey it forward into the barrel. Figure 11.32 shows the proper feed locations in both corotating and counterrotating, intermeshing extruders. Material fed downstream by gravity also needs to be fed on the proper side of the screw turning down into the barrel to obtain a good bite on the material. A downstream feed port, where material is gravity fed to screws partially filled with molten polymer, may have an opening on only one screw side, where rotation is turning down into the barrel, to assist feed material addition in corotating and counterrotating extruder screws.

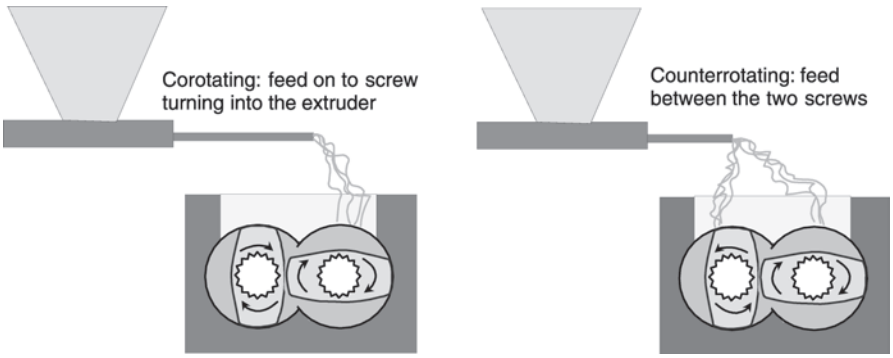


Figure 11.32. Optimum feed locations for corotating and counterrotating extruders.

Liquid can be added via an injection port and a liquid feed pump. Figure 11.33 shows a barrel cap or vent cap with a liquid injection port that was removed from either zone 2, 4, or 6 in Fig. 11.30. The vent cap is drilled for a pressure transducer; the liquid injection port is fabricated to fit into the pressure transducer type hole. Injection ports can be fabricated with different size holes to feed different quantities. The liquid feed pump is attached to the injection port on the vent cap with metal piping. For safety the liquid feed pump is equipped with a pressure gauge to measure the injection line pressure and a relief valve in the event pressure becomes excessive. Liquid feed is introduced as soon as the extruder is started, preventing molten resin from backing up into the injection port, clogging the line, and/or freezing off the liquid injector tip, preventing liquid addition. Figure 11.34 shows constant speed liquid feed pumps and a controller.

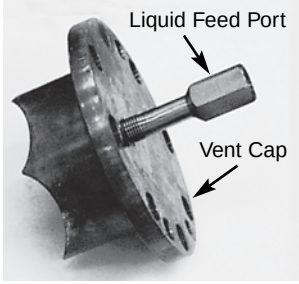


Figure 11.33. Liquid feed port.

Some materials are best fed downstream into the melt using a side feed extruder or stuffer. High aspect ratio polymer reinforcements such as fiberglass are fed downstream into the melt to minimize fiber attrition. Molten plastic acts as a buffer, minimizing fiber breakage and lubricating the screws where the fiber is added. If one feeds high aspect ratio reinforcements into the first extruder zone with pellets, the solid pellets plus the screw rotation will chop the fibers into very short lengths. Figure 11.35 shows a twin screw side feed extruder used to starve feed fillers, reinforcements, or other additives from a feeder on the mezzanine into the polymer melt downstream. Fillers and reinforcements are fed into a low-pressure zone after the resin is completely melted. (Other materials, such as mineral fillers, temperature-sensitive additives, resins, etc., can also be fed downstream with the side feed extruder.) Additive addition down-

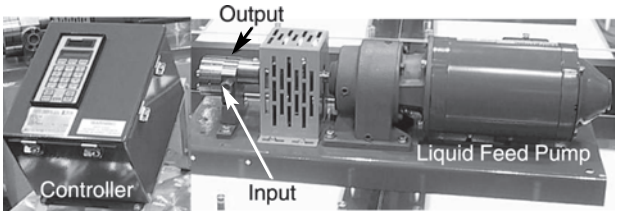


Figure 11.34. Zenith liquid feed pump.

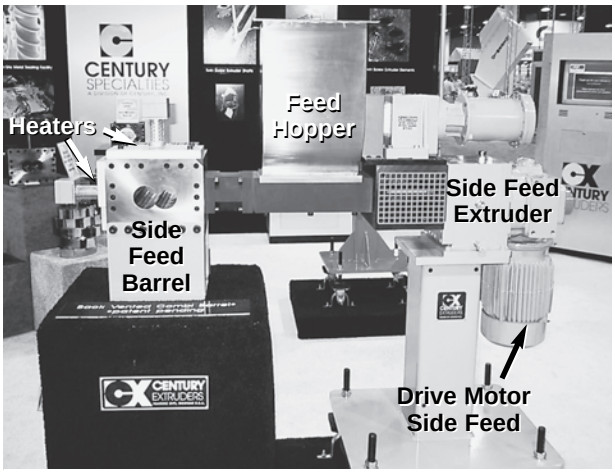


Figure 11.35. Century Speciaties side feed extruder connected to side feed extruder.

stream reduces its residence time and the shear experienced. Side feed extruders are sometimes called stuffers. These extruders have very short barrels with various screw geometries that convey but don't compress the material while feeding different resins, fibers, and/or additives. The side feed extruder barrel clamps to the main extruder barrel, with the side feed screws extending through the main extruder barrel wall almost to the main extruder screws. Material from the side feed extruder is fed at high rates directly into the polymer melt.

The first barrel section or feed zone on the extruder is normally water cooled with no heating present. Cooling is provided to prevent resin or additives from prematurely melting and sticking to the feed throat. Over time, sticky or premelted material can agglomerate around the feed throat opening, causing a bridge to block the feed opening. Feed screw elements are deep flighted with large pitch to provide the maximum open area to transport resin away from the feed opening and into the extruder; see Fig. 11.36. At start-up the screw speed is normally run at low rpm with a

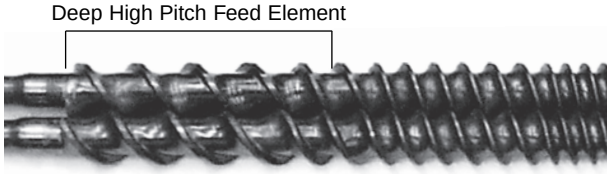


Figure 11.36. Feed elements.

significantly reduced feed rate until material exits the die. Once polymer flow through the extruder and die is established, the extruder screw speed is increased, followed by the feed rate, until the desired throughput is attained. During this time screw torque is carefully monitored to assure that the extruder does not exceed the drive torque limit, causing the machine to shut down. The feed rates into the extruder are carefully monitored to verify that the screw speed is sufficient to remove all the material being fed to the extruder and no build-up is occurring in the feed hopper. If material builds up in the extruder feed hopper, either the extruder screw speed must be increased or the feed rate decreased. The individual feeders can be slaved to the extruder; in the event the extruder shuts down, the feeders automatically stop feeding. This prevents resin build-up in the various feed locations along the barrel when the extruder stops unexpectedly. Assuming the extruder shut down because it exceeded the torque limit, any material in the feed throat (or other feed locations) needs to be vacuumed out prior to restarting the extruder. If the extruder stops due to exceeding the torque limit, the screw speed is brought back up to high rpm again before the feed stream is restarted at a low rate and gradually increased.

Feed is introduced into nonintermeshing, counterrotating extruders between the screws, as shown in Fig. 11.37. With the screws turning down into the barrel, the material is picked up and conveyed forward into the barrel.

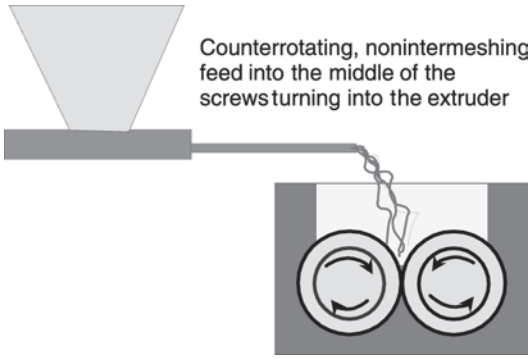


Figure 11.37. Feed location for counterrotating, nonintermeshing extruder.

11.3 Screw and Barrel Heating and Cooling

The screw and barrel is where polymer is fed, melted, conveyed, mixed, devolatilized, and pumped to the die. Both the screws and barrel sections are modular and can be arranged in any configuration necessary to accomplish a particular extrusion objective. Barrel sections normally have either rectangular or circular outside dimensions, depending on the manufacturer. They are assembled with either a rod through all the sections or with bolts holding the sections together, as shown in

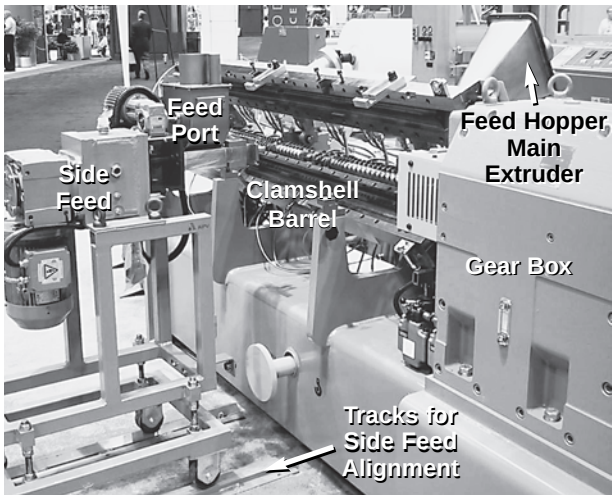


Figure 11.38. APV Clamshell twin extruder from HPM.

Fig. 11.35 and Fig. 11.43. The barrel is supported at different locations along its length to prevent it from sagging. Each barrel contains a thermocouple to control the heating and cooling input. Figures 11.39 and 11.40 are a top and side view of a twin screw extruder barrel. The heating and cooling elements, different barrel sections, vent caps, insulated barrel covers, and a barrel support

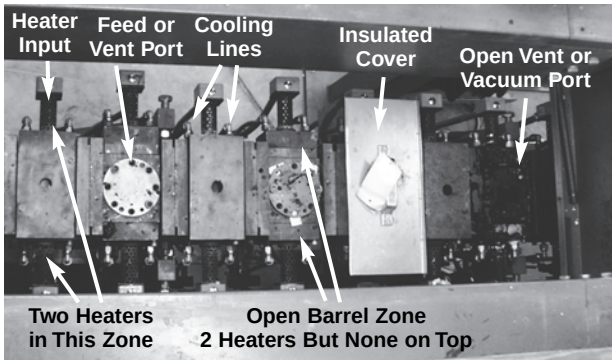


Figure 11.39. Top view of different types of barrels plus heating and coating.

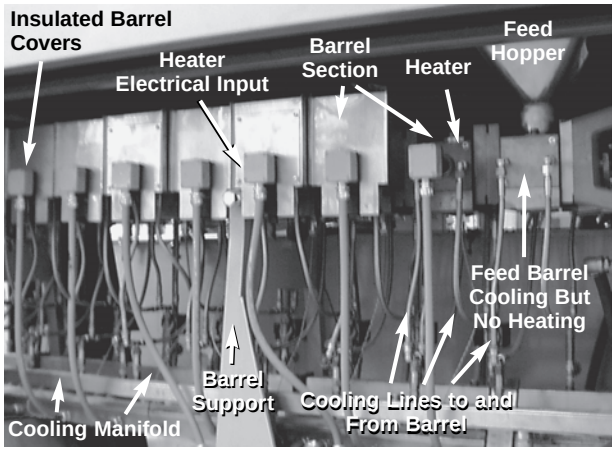


Figure 11.40. Side view of barrel sections, heaters, cooling, etc.

are identified. Barrel sections have flanges on each end for alignment and connection to the next barrel. While it is a sizable task, barrel sections can be disassembled, reconfigured, and reassembled to move feed and vacuum sections. Some machines (particularly smaller ones) may have a clamshell barrel design where the barrel separates in the middle and the entire screw length is exposed. See the APV extruder from HPM in Figs. 11.38 and 11.40. Clamshell barrels are one-piece construction rather than modular sections that bolt together. Individual heating and cooling zones along the barrel provide temperature control similar to a water-cooled single screw extruder.

Barrel sections, like extruders, come in different length to diameter ratios (*L/D*s). Typical lengths depend on the screw diameters and the manufacturer. Some common *L/D*s are 2.5, 3, 4, 5, 6, 8, 10, and 12. There are many different barrel sections:

- one used for feeding with an opening on top (vent barrel)
- feeding into the side with a vent on top (combi barrel)
- a solid barrel (closed barrel)

The vent barrel has an opening on top that may be either circular or rectangular and is used to vent volatiles from the barrel or to feed different formulation components. The first barrel section in the extruder is opened on top for feeding all or part of the formulation into the extruder. It is cooled and normally has no side heaters. The normal figure 8 pattern bore connects to an end plate that prevents the formulation from traveling backward toward the drive system, with the other end connecting to the next barrel section. Figure 11.40 shows the feed section with a hopper, water cooling, and the absence of heaters. Other vent or feed barrel sections have an opening on top with cast heaters contacting the barrel on the other three sides. Barrel section 2 (the one after the feed throat) contains a vent plug and is heated on three sides with cooling (these barrel sections have the capability to be either heated or cooled as the process requires); refer to Figs. 11.39 and 11.40. This same barrel section can be used to feed liquid through an injection nozzle, to gravity feed polymer further downstream, or to vent volatiles with a vent stack,

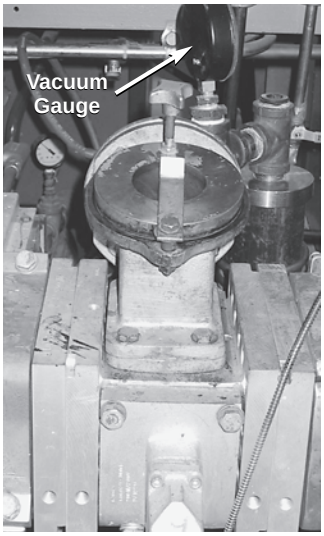


Figure 11.41. Vented barrel section.

Fig. 11.41. The vent stack can be open for atmospheric venting or it can be connected to a vacuum port to remove higher volatiles through vacuum venting. The combi barrel has an opening on top for gravity feeding and a side opening for feeding with a side feed extruder. The barrel section in Fig. 11.35 is a back-vented combi barrel (supplied by Century Extruders and is patent pending) with the side feed extruder attached. Combi barrel sections are cooled similarly to other barrels, but only two of the four sides are heated when the side feed extruder is connected. If the side feed extruder is not in use, the port can be plugged and heated similarly to the vent barrel. The last barrel section is a closed barrel with no openings on the top or side, shown in Figs. 11.42 and 11.43. This barrel is heated with cast heaters on all four sides and cooled similarly to the other sections. It is seen in Figs. 11.39 and 40 and is the barrel section with the heater on top. Each heater is L-shaped, so one heater covers the top and one side, while the other covers the bottom and opposite side. Figure 11.43 is a schematic of the different barrel sections.

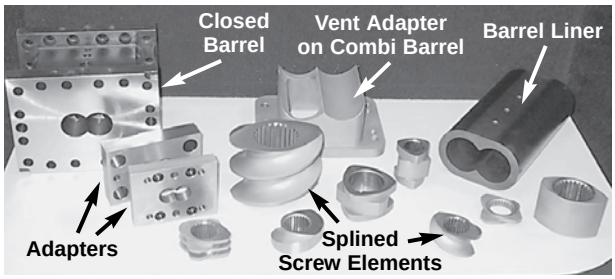


Figure 11.42. Century Extrusion spare corotating twin screw extruder parts.

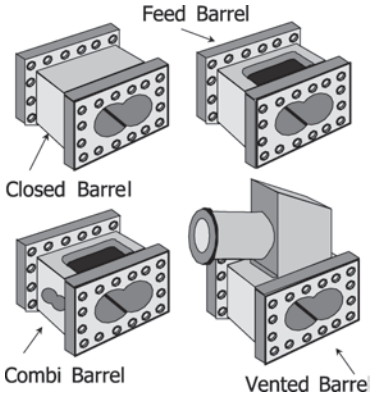


Figure 11.43. Different types of barrel sections availble.

Barrels and screw are normally nitrided steel for long life. Through-hardened barrels are available and some barrels are lined with wear-resistant liners to increase life. Liners provide resistance to corrosion and abrasion. Other metal treatments are available to increase service life. Figure 11.44 shows the end of a 45-mm Theysohn twin screw extruder barrel section with a wear resistant liner. A barrel liner is shown in Fig. 11.42.

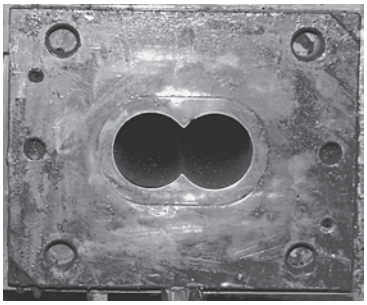


Figure 11.44. Barrel liner.

If an atmospheric or vacuum vent is used, the vacuum port covers half the screw (see Figs. 11.42 and 11.43) to prevent material from flowing out the vent. Some typical vent caps and configurations are shown in Fig. 11.45 for corotating, intermeshing twin screws.

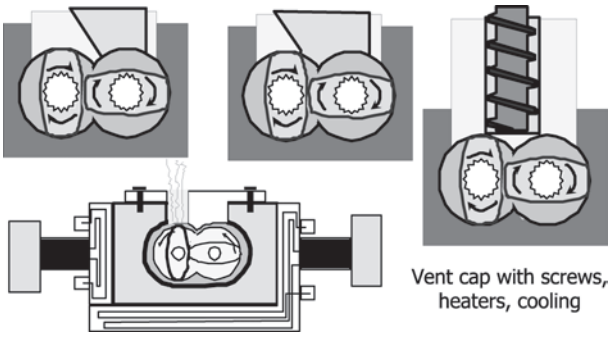


Figure 11.45. Vent cap configurations.[4]

Different cast barrel heaters are available, depending on the application and temperature requirements. Aluminum heaters have a temperature range to 660°F (350°C), while aluminum/bronze alloy heaters go up to 840°F (450°C).

Two cooling designs are available. One uses cooling bores with holes bored around the barrel next to the barrel liner (Fig. 11.46). The second uses water cooling in the heater with a loop running around the heating element (Fig. 11.47). Combining both types of cooling provides the maximum temperature control for the system. Figure 11.46 shows a barrel cross section with heater and internal cooling bores and the corresponding time–temperature response curve. This cooling is the least responsive cooling mechanism available, giving the poorest temperature control. Figure 11.47 shows a single cooling system in the

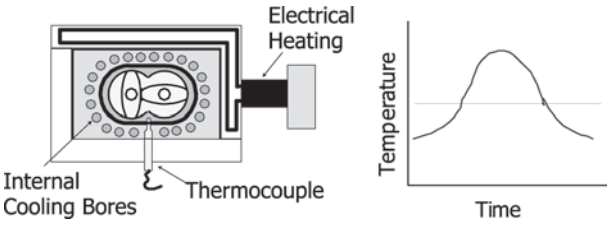


Figure 11.46. Internal bore cooling and time-temperature response.

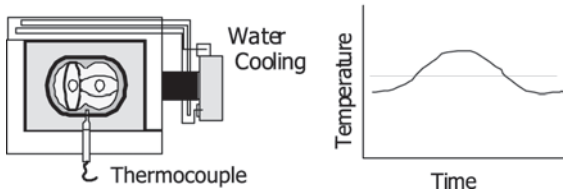


Figure 11.47. Single cooling system.

heaters. While this system performs better than the cooling bores, it still is not as good as combining internal cooling bores with water cooled heaters to control the temperature. Figure 11.48 shows a system combining cooling methods and the time–temperature response curve.

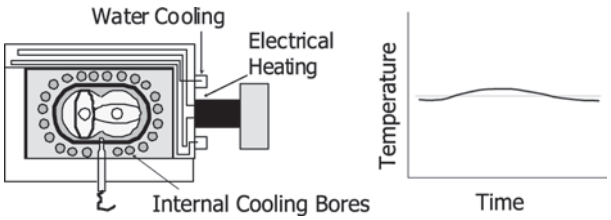


Figure 11.48. Combination of water cooling and cooling bores.

In some circumstances the screw shafts can be cooled for better temperature control if required. Water or oil is pumped in a tube down the center of the shaft and returned on the outside of the tube. This provides good heat transfer to the outside surface of the screw shaft and inside surface of the screw elements. Maximum cooling occurs at the extruder discharge end. Reversing the flow provides maximum cooling at the extruder feed end.

A Welding Engineers counterrotating, nonintermeshing twin screw extruder drive and barrel section is shown in Fig. 11.49. While this is a nonintermeshing twin screw, the barrel section does not appear radically different from the intermeshing barrels shown previously, and the drive system components are similar. The only slight difference is the barrel apex area, where the metal on top and bottom extend farther toward each other.

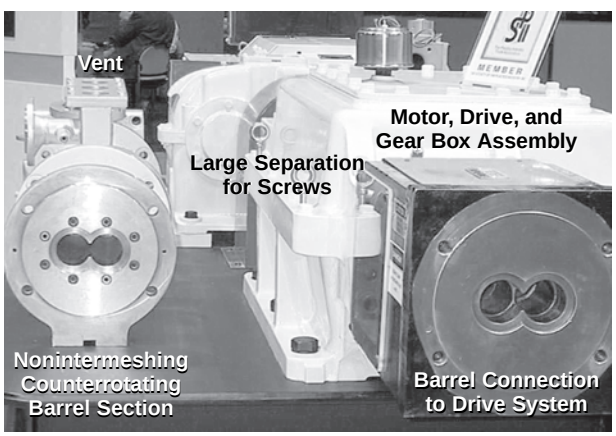


Figure 11.49. Welding Engineers counterrotating nonintermeshing twin screw barrel and drive train.

Screw nomenclature was discussed previously in this chapter. Similar to the barrel sections, screw elements are modular and are inserted onto the screw shafts to provide the proper conveying, melting, mixing, downstream component addition, venting, and pumping to produce a commercially acceptable product. Specific screw design to accomplish various objectives is discussed in Chapter 13. It is important to understand the extrusion requirements before assembling the screw.

- Where along the extruder barrel do you want the material to melt?
- What location do you want to add ingredients downstream?
- How much and what kind of mixing is required to produce a homogeneous melt?
- What rate is required?
- Does the product need atmospheric or vacuum venting and at what location?

Before assembling a specific screw configuration,

- All elements must be cleaned on both the outside and inside.
- The ends must be lapped to provide a good sealing surface between the elements.
- The shafts, keyways, or splines must be cleaned so elements slide on the shafts easily.
- Antisieze must be applied in a thin coat to all surfaces during assembly.

A specific screw design is developed for the process and documented. The design is used to assemble the elements in the correct order. Figure 11.50 shows a typical setup to use when constructing or assembling a new screw design. During screw assembly, antisieze application is essential to ensure that the screw can be disassembled later. However, excessive antisieze must be avoided. Antisieze can build up between the elements, creating a gap where polymer will flow under pressure between the elements down to the shaft. Over time at high temperature, polymer trapped between the screw elements and the shaft degrades and chars, making screw element removal very difficult.

Figure 11.50. Setup to assemble new screws.

Screw assembly is critical; the elements must be placed on the shafts in pairs (element on shaft 1 must be the same as element on shaft 2). With multiple keyways, splined shafts, or polygon shafts, there is only one correct position to install each element. Each element flight must

match the preceding and following element flight to prevent dead spaces along the shaft and create a smooth polymer flow. If an element is misaligned or a left-handed element is used on one shaft and right-handed one on the other, the screw shafts will not rotate, and the screw elements will have to be removed back to the mistake and then reinstalled. Identical kneading blocks, like conveying elements, have to be installed at the same place on the shafts or the shafts will not rotate.

Screws and barrels for conical counterrotating extruders are significantly different from the parallel intermeshing co- and counterrotating extruders. Similar to single screw extruders, the barrel is not segmented and modular, but one piece. Likewise the screws are one piece. Conical twin screws are used primarily for rigid or unplasticized PVC extrusion into profiles (window, siding, gutters, etc.) and pipe.[5] Conical screws and barrels have a large diameter at the feed throat and get progressively smaller toward the die. Conical twin screw extrusion shafts from CPM GmbH were shown in Fig. 11.2.[4] Figure 11.51 pictures a CPM GmbH barrel without any heaters or cooling elements present. A conical extruder looks similar to a single screw extruder with heater bands, cooling, and barrel covers installed.

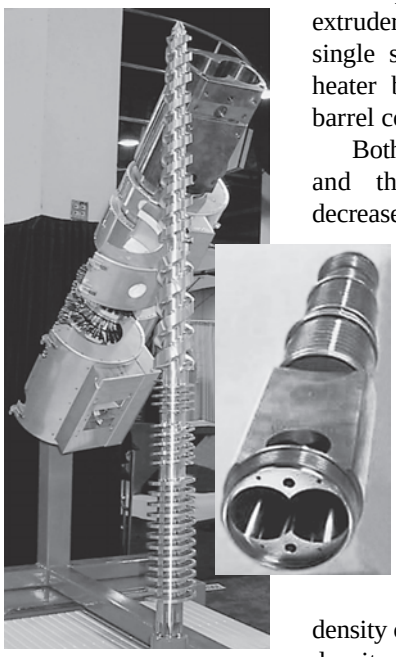


Figure 11.51. Conical barrel by CPM.[1]

Both the root diameter and the flight diameter decrease from the feed section to the die. Material is compressed from the feed end to the die through a decrease in channel volume. With conical twin screws, the feed is often PVC powder, which has a low bulk density compared to the melt density. Consequently, the channel volume change through the extruder is quite dramatic to properly compress and mix the formulation. Maximum channel volume occurs in the feed zone, assuring a uniform material feed and conveying. Figure 11.52 shows a conical extruder screw and the process zones in the extruder. The feed and plastication area is where the formulation is fed and converted to a molten polymer. After the plastication flights, a restrictive area with smaller flight volume retards the resin movement, forcing the resin to spend more time in the plastication zone while preparing the melt for devolatilization. Similar to parallel twin screw extruders, this screw section acts as a melt seal for the devolatilization area. The devolatilization zone has a larger pitch to provide maximum polymer surface area to remove volatiles. This zone is only partially filled to assist with the removal of trapped air and volatiles from the melt. Finally, the melt is recompressed and pumped to the die in the metering zone.

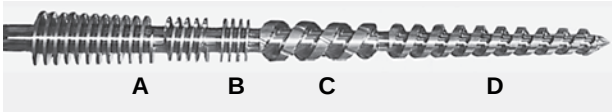


Figure 11.52. Cross section of conical twin screw extruder showing zones for (A) feed and plastication, (B) restrictive, (C) devolitilization and (D) metering.

Parallel twin screw extruders have more surface area than conical twin screws. However, the flight flanks in a conical have more surface area than in a parallel extruder and can transfer more heat to the material in the channel. The screw surface area is 40% larger compared to a parallel extruder. Due to the larger channel volumes, conical extruders generate less shear heat and more conductive heat compared to parallel twin screw extruders, making them better for processing shear-sensitive materials such as PVC.

Conical twin screws with their one-piece design have a distinct advantage in being able to transfer torque through the screws. Since most energy to melt the material is supplied by the motor and screw rotation, conical extruders have more strength where it is required. Most screw wear will occur in the feed, plasticating, and restrictive areas.

11.4 Die and Adapter

Dies for any extrusion process can be installed on a twin screw extruder to produce the desired product. Breaker plates with screen packs for filtration or to restrict the polymer flow and build backpressure before the die are used when the process requires them. The largest application for corotating twin screw extruders is compounding different resin formulations or pelletizing polymer by resin suppliers that require either a strand die (Fig. 11.53), die face pelletizer, or underwater pelletizer. While corotating twin screw extruders are excellent melt mixers, producing homogeneous products, they do not generate high die pressures. Gear pumps are often incorporated between the extruder and die (discussed in Part 5, “Auxiliary Equipment”) to generate sufficient die pressure for sheet, profile, and tubing applications. Gear pumps generate very uniform and high die pressure to produce uniform cross section dimensions in the extrudate.

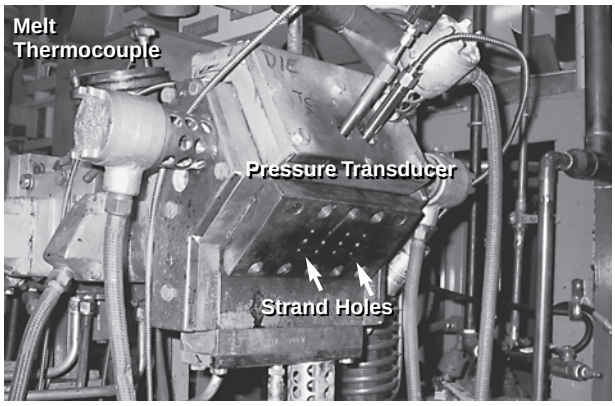


Figure 11.53. Strand die.

Counterrotating twin screw extruders, both parallel and conical, are excellent die pressure generators and are normally run at slower screw speeds with fuller flights than corotating extruders. Consequently, less shear heat is generated. Profile extrusion, particularly with rigid PVC, uses counterrotating extruders to provide

- Good ingredient fluxing during plastication
- Good mixing
- A low shear process that minimizes degradation
- Sufficient die pressure at a constant rate to produce uniform extrudate cross section in the final product

The die or die adapter connects to the barrel figure 8 cross section and converts the melt to a round cross section before entering any other adapter or die. This is normally called an 8-O transition and is required for all twin screw extruders. Once the 8-O transition is made, dies are similar to those used in single screw extrusion to obtain the desired product profile. Similar to single screw extrusion, all dies and adapters require adequate and uniform heating with no dead spaces in the flow channels to prevent hot or cold spots in the polymer flow that might alter the melt viscosity or lead to resin degradation.

Some die designs for different products were shown in Part 1, "Single Screw Extrusion," Chapters 3 and 4, and will be reviewed in more detail in Part 7, "Extrusion Processes." Different screens and filtration were discussed in Part 1, "Single Screw Extrusion," Chapter 3, and will be covered in more detail in Part 5, "Auxiliary Equipment," under automatic screen changers.

11.5 Controls

The extruder is more than a black box where pellets are put in one end and an extruded shape exits the other. It is a complicated process and needs good controls based on understanding what is occurring in the extruder. What happens in different areas of a twin screw extruder? How do you know the extruder is in control and producing a quality product? How do you know the extruder is running at peak efficiency? The answer to most of these questions is, there are both input and feedback controls to verify that the process is operating properly and at equilibrium.

The control process makes a measurement, determines if something needs to be changed, makes a decision, and takes appropriate action. If the system is operating at equilibrium with good product exiting the die at a high rate, the decision might be everything is running properly and no changes are necessary. However, if the product is borderline acceptable or unacceptable, or one process step is outside the SPC control chart limits, the decision is that some function is operating improperly and a change is required to return the system back into control.

With today's computer capability and durability, extruder controls are more sophisticated, resulting in better overall process control. The process can be controlled and monitored with feedback loops from the individual feeder throughputs to the puller speed and all operations in between. Feedback loops plot SPC data at specified time intervals for all temperature controllers, speed controls, puller speeds, in-line gauges for thickness or dimensions, feeder rates, water temperatures in cooling tanks, vacuum levels, windup speeds, roll pressures, and so forth. With the entire process instrumented, a line or multiple lines can be monitored from a remote site to evaluate process reproducibility from run to run and repeatability within a run. Engineers sitting in an office can monitor each plant line, verifying that all lines are running properly and the processes are in control based on SPC-generated data.

In comparison to other plastic processes, an extruder has very few independent control variables that can be changed by an operator to alter the process. Assuming the correct screw design is in the extruder, the proper die is installed, the screen pack is clean, and the equipment is operating properly (all heater bands and thermocouples are functioning properly, air or water cooling on the heating zones is working, and cooling on the feed throat is operating properly), the only extruder variables that can be changed are the temperature setpoints, screw speed rpm, and throughput rates. Parameters that can be monitored to ensure that the process is in control and running properly are

- Barrel temperatures (actual and setpoint)
- Extruder load (percent load, torque, or amps)
- Screw speed
- Melt temperature
- Melt pressure
- Feed throat cooling water
- Raw material temperature entering the extruder

- Vacuum level, if vacuum venting is being used
- Cooling on the individual barrel zones

The other variable with modular twin screw extruders is screw design. If the process is not optimized, the screws can be removed from the extruder and then redesigned to optimize the process and throughput rates. Other parameters in the extrusion line to monitor include

- Blend ratio
- Feed rates
- Moisture content of hygroscopic and moisture-sensitive raw materials
- Liquid feed rates
- Raw material lot numbers

Downstream equipment parameters to be set and monitored include

- Roll and/or puller speeds
- Roll gaps
- Cooling bath temperature
- Vacuum level
- Windup speeds

With all this information available, how do you know if the product is good? If process changes are required, how do you make changes? In some instances, not enough information is recorded, raising the question, how do you troubleshoot the process?

When setting up the extruder and any other auxiliary equipment in the extrusion process to produce a particular product, the first step is to obtain the standard operating procedures (SOP) for the product. This is a controlled document specifying all processing conditions and settings prior to start-up. Process parameters are normally controlled over a very limited range. If the final product does not meet specifications at start-up, the SOP for the process should be compared with the setup to verify that all equipment is properly set and within the ranges specified by the SOP. If the raw materials are the same as last time and all process conditions are the same, this raises the question, what has changed since the last run to make the product unacceptable? Investigation to determine what has changed requires a systematic approach. Do not start trial and error experimentation, making random changes without an appropriate plan to fix the problem.

Good procedure is to record all operating conditions, both when the process is running properly and when problems exist. The tendency is not to monitor the process when it is running smoothly, only when there are problems. Consequently, when problems occur because something has changed, if baseline data under good processing conditions are not available it is difficult to determine what has changed. It is impossible to record too much data and have too-detailed records. However, it is possible to record too little information and not have the necessary data when problems arise to properly troubleshoot the process.

11.5.1 Temperature Zone Control

Each barrel except the barrel in zone 1, the feed barrel, has at least one heater, and closed barrels have two heaters controlled by a thermocouple. A signal from the thermocouple communicates with the controller, turning the heater on or off. For the controller and heaters to function properly, the thermocouple must operate properly. A faulty thermocouple with an alternate closed junction at a cold temperature will indicate that the temperature is low, causing the heater to stay on, resulting in substantial overheating. An open thermocouple circuit is usually detected by the controller as a high temperature and the heaters will stay off and the temperature zone cools. If a thermocouple is not responding properly, it needs to be replaced. Thermocouples have to be plugged into the correct temperature controller; otherwise the thermocouple controls the wrong heaters. For example, if the thermocouple from zone 2 is plugged into the zone 3 controller while the zone 3 thermocouple is plugged into the zone 2 controller, and the setpoint temperature for zone 2 is 420°F (216°C) and zone 3 is 450°F (232°C), the actual extruder temperature in zone 2 and in zone 3 will be a complex relationship involving the heat load, the thermal lag, and the controller tuning parameters. Needless to say, this is an undesirable condition. Barrel temperature thermocouples are spring loaded to guarantee good contact with the barrel wall and insulated to minimize heat loss along the thermocouple stem. Melt temperature thermocouples work on the same principles as the zone temperature thermocouples, except the melt thermocouples must protrude into the melt stream to obtain accurate melt temperature. Melt temperature measurements and thermocouples are discussed in detail in Part 1, “Single Screw Extrusion,” Chapter 3.

Each heater zone should have an amp meter readout, whether it is on the control panel or an option in a menu-driven computer control system. With all heaters functioning properly, the amps drawn in a particular zone when the heaters are 100% on is a fixed value. Assuming a zone has more than one heater and the ammeter reading is lower than normal, at least one heater is burnt out or not working properly. Nonfunctioning heaters need to be replaced as soon as possible to prevent hot or cold spots along the extruder barrel, resulting in nonuniform polymer heating. Ammeter readings should be checked daily to verify that all heaters are functioning properly.

If the actual zone temperature differs from the setpoint, the extruder is saying something is out of control.

Possible scenarios causing the variation of actual versus set temperature are

- Temperature is not controlling that particular zone.
- Thermocouple is not operating properly and may need replacing.
- Temperature setting for the material being processed is incorrect.
- Excess shear heat is being generated in that zone.
- Extruder cooling is not functioning properly.
- Thermocouple wire to the controller is not connected to the correct controller and/or zone.
- Screw design is incorrect.

11.5.2 Pressure Measurement

Melt pressure measurements at the extruder head are as critical as melt temperature. A melt temperature thermocouple and die pressure transducer are also shown in Fig. 11.53. To reproduce the same product from run to run, the melt temperature needs to be the same to ensure that the melt viscosity and consequently the throughput is constant. Likewise, constant pressure produces a uniform cross sectional dimension from run to run, assuming the melt viscosity is the same. As the pressure in the die increases due to unstable extrusion rate, the output increases and the product cross sectional area becomes larger. The second function of the pressure gauge is to act as a safety device to prevent excessive die pressure that may blow the die off the machine or cause other equipment issues. Normally the pressure gauge is equipped with a high-pressure alarm, which shuts the extruder off in the event a preset pressure is exceeded.

Equation (11.7) shows the effect of pressure variation on volumetric output.[6]

$$\Delta Q(\%) = \frac{\Delta P(\%)}{n} \tag{11.7}$$

where ΔQ = Change in volumetric output given in percent
ΔP = Change in pressure as a percentage
n = Power law index of the polymer being processed

Assume the polymer being processed is polypropylene with a power law index of 0.35 (n = 0.35) and the die pressure is varying from 860 to 900 psi for a ΔP of 40 psi. The ΔP in percentage is (40/880) $\times$ 100 = 4.5%. ΔQ = 4.5/0.35 or 13%. Assuming the process is running at 200 pounds/hour, the variation in output is 26 pounds/hour, which significantly impacts the product dimension.

11.5.3 In-Line Measurements

In-line measurements of viscosity and infrared spectroscopy for additives in the formula were covered in Part 1, "Single Screw Extrusion," Chapter 3. The techniques used to monitor polymer viscosity, polymer compositions, and finished product dimension are the same for both single and twin screw extrusion.

11.5.4 Control Summary

Good production practices dictate that all operating conditions and observations need to be recorded to establish a database for troubleshooting product or process problems. New computerized control systems provide continuous process data collection. Any unusual occurrence or abnormality needs to be documented on the control chart to identify cause and effect. It is impossible to record too much information during a developmental trial or a production run. Documentation includes

- Setpoint and actual temperatures
- Pressures
- Screw speeds
- Motor load (percent load, amps, or screw torque)
- Melt temperatures
- Product appearance
- Raw material type and lot numbers
- Formulations
- Drying time and temperatures
- Puller speeds
- Cooling medium temperatures
- Takeoff equipment settings

New computerized control and data collection systems contain formulation data and input operating conditions and generate control charts for SPC. Error messages are printed out along with process data at prescribed intervals and can document an entire run.

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Review Questions

1. What is the difference between parallel and conical twin screw extruders?
2. What are the different types of twin screw extruders and where are they used?
3. Name the five major equipment components of a twin screw extruder.
4. Where are the high-pressure areas in material flow in corotating and counterrotating, intermeshing twin screw extruders?
5. What is the difference between three-lobe and two-lobe screw elements?
6. In screw element nomenclature, describe the following screw elements: 60/60, 45/5/30, 60/30 left, 25/25, 40/20, 90/3/30, 45/5/45 left.
7. What is meant by the diameter ratio D_o/D_i and why is it important?
8. What are the keys to obtaining a high throughput in a corotating, intermeshing twin screw extruder?
9. What does free volume in a twin screw extruder mean?
10. What is meant by the extruder is 40% full?
11. Why is it necessary to have splined, polygonal, or other keyway designs in newer twin screw extruders versus the old single rectangular keyway in three-lobe extruders?
12. What is the residence time distribution and how does it apply to the different twin screw extruders?
13. Describe the start-up procedure for an intermeshing, corotating twin screw extruder and tell why the procedure used is important.
14. What will happen if the filter on the motor cooler is dirty?
15. What feed mechanism is used with corotating, counterrotating conical, and counterrotating slow-speed parallel twin screw extruders?
16. What are the different barrel sections on modular twin screw extruders?
17. What and where are the different feed ports in a twin screw extruder and how is each fed?
18. What is the best barrel cooling method and why?
19. What are the various vent or vacuum ports?
20. What process parameters can be controlled on a twin screw extruder, and what can't be controlled but need to be monitored to ensure a quality product?
21. How do twin screw extruders differ from single screw extruders?
22. What is the main application for corotating twin screw extruders?
23. Where are conical twin screw extruders used and why?

12 Plastic Behavior in Twin Screw Extruders

Similar to single screw extrusion, polymer exposure in the extruder can be broken down into eight distinct areas:

- Feed
- Plastication
- Conveying
- Mixing
- Downstream feeding
- Devolatilization
- Pumping
- Die

These areas may overlap and not be as distinct as in a single screw extruder because a twin screw extruder is more flexible and versatile than a single screw extruder. However, the extrusion objectives to be accomplished with the twin screw extruder remain constant:

- Create a homogenous product to feed the die.
- Obtain a constant melt temperature.
- Process the polymer at a specific melt temperature.
- Obtain a constant melt pressure.
- Process the polymer at a specific melt pressure.

With these objectives in mind, the screw design and operating conditions are chosen to meet the extrusion goals while creating a quality product.

12.1 Feed

As discussed in Chapter 11, either liquid or solid feeding can occur at different locations in a parallel twin screw extruder. Initially, feeding in barrel zone 1 is discussed, followed by downstream addition of either solid or liquid material. Corotating and counterrotating, intermeshing, parallel resin extruders are starve fed. The extruder throughput rate is determined by the feed rate to the extruder, and not the extruder screw speed. The minimum acceptable extruder screw speed conveys all the feed material away from the feed throat while running at acceptable torque levels to prevent the extruder from exceeding its torque limit. The extruder screw conveying capacity is greater than the total feeder throughput rate. Consequently, the feeder rates have to be both very accurate and reproducible to produce uniform products.

The first screw element is normally a low-pitch, short element. This element prevents feed from flowing backward toward the screw seals. Immediately following the low-pitch element is a high-pitch element, a single flighted element, or a high-pitch element with an undercut that provides open space for pellets, powder, flake, etc., to enter the extruder (Fig. 12.1). The SK element has an undercut on the pushing flight (flight is almost vertical) with a normal radius on the trailing flight. This configuration helps feeding low bulk density material as well as normal resin formulations by providing more free volume in the feed section. The SKN transition element, following the SK, changes the undercut to the normal conveying element pitch. If the SKN is not used, there is a dead spot where an SK element mates to a normal conveying element. After the transition element, normal conveying elements are used to convey the feed material to the rest of the extruder. Figure 12.2 pictures two 90/90 SK feed elements,

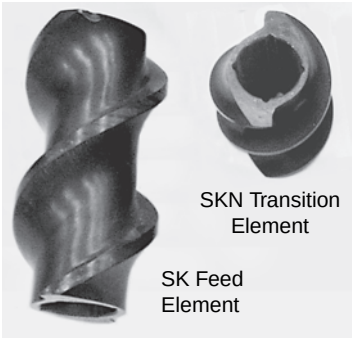

Figure 12.1. Enlarged view of SK and SKN screw elements.

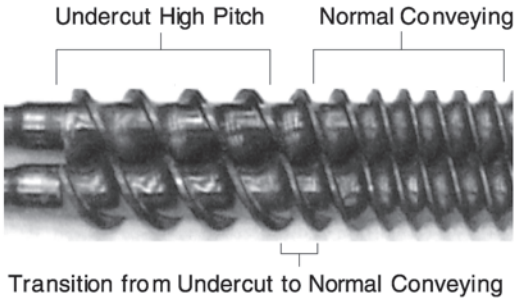

Figure 12.2. Screw elements in the feed zone.

followed by 60/60 SKNs (transition from an undercut to a standard conveying element) and normal 45/45 conveying elements. Normal large-pitch conveying elements can be used in place of the 90/90 SK screw elements in most situations to feed material. The screw elements shown and discussed go in a 45 mm corotating twin screw. Consequently, the pitch in the feed zone using 90/90 conveying or SK elements is two times the diameter. To have a similar screw free volume in a 90 mm extruder, the element would have a pitch of 180 mm with a length of 180 mm and be designated as a 180/180 SK or a 180/180 normal conveying element. In any parallel twin screw intermeshing extruder, the feed screw element pitch of either an SK or a normal conveying element is normally 1.5 to 2 times the screw diameter. Figure 12.2 shows the screw element configuration in barrel zone 1 and part of barrel zone 2 for a 45 mm corotating twin screw extruder.

Either volumetric or gravimetric feeders can be used to supply resin to the feed throat. A volumetric feeder

runs at constant speed, whether it is a screw feeder (single or twin screw), a belt feeder, or a vibratory feeder. Providing the formulation is free flowing, volumetric feeders produce a uniform flow rate. If individual feeders are used for different additives, e.g., resins, filler, volumetric feeders require calibration. Calibration requires generating a speed versus output curve for each feeder with the different resins and additives. This can be a very time-consuming task. After setting the desired feed rate using the calibration curve, the feed rate stability needs to be checked by sampling the feed rate versus time. In the event more than one feeder is used, calibration curves and throughput rate checks at the desired speed are required prior to starting the extrusion run to determine measurement precision and feeder accuracy. Another way to check the feed rate is to set one feeder, start feeding material, and check the extruder throughput rate. After the first component rate is established, the second feeder can be started and the total extruder throughput for the two feeders determined. The difference in the two throughput rates is the second feeder rate. If the rates are not in the proper ratio, either feeder speed can be adjusted to establish the correct ratio. Throughput rate checks during production can be used to establish if the total feed is correct. This alternate method is not effective when checking low additive feed rates.

Gravimetric feeders control the mass flow per unit time. Each ingredient rate is dialed into the control panel for its feeder, and the sum is the extruder throughput rate. Setup time with gravimetric feeders is very rapid, as calibration curves to determine the setpoint are not required. In the event both volumetric and gravimetric feeders are being used to feed different components, calibrate the volumetric feeder to establish the feeder setpoint for the total extruder throughput rate. Then use the gravimetric feeder(s) to establish the correct ingredient ratio.

In Chapter 11, a picture of a volumetric feeder, feeding material to a twin screw extruder on a mezzanine, was shown in Fig. 11.31, with different feeder configurations shown in Fig. 11.29 and extruder feed locations in Fig. 11.30. Figure 12.3 shows a picture of a gravimetric weigh belt feeder. Feeders are discussed in more detail in Part 5, "Auxiliary Equipment."

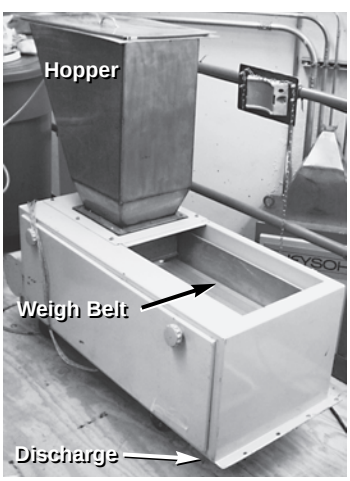

Figure 12.3. A weigh belt feeder.

If feeders are limited and all components in a given formulation cannot be fed into the extruder feed throat, ingredients can be premixed at a remote site, transported to the extruder, and added to one feeder. This method uses less expensive feeders in the extrusion process. A potential problem is that particle size or density differences can cause materials to separate during transport to the extruder and/or transfer to the feeder hopper. A second concern is this process requires more labor and is more prone to human error. Extrudate uniformity is directly related to the preblend uniformity and the separation during handling prior to extrusion. Separate individual component feed streams at the extruder increase the process flexibility.

How material is fed into a twin screw extruder depends on the screw rotation. In corotating, intermeshing extruders, it is the outer screw that turns down into the extruder and draws the material into the barrel. A feed pocket can enhance the feeding characteristics. Figure 11.32 in Chapter 11 shows the preferred feeding for corotating and counterrotating, intermeshing extruders. The feed configuration depicted in Fig. 12.4 for a corotating, intermeshing extruder is particularly effective for downstream feeding without the aid of a side feed extruder or stuffer. It is important to add new ingredients where the screw turns down into the barrel to prevent build-up in the barrel opening. Counterrotating, intermeshing extruder screws rotate toward the outside barrel wall, pulling material into the barrel as the screws turn down into the barrel. Nonintermeshing, parallel twin screw extruders draw material into the apex area as material is drawn in between the two screws. The free volume in nonintermeshing extruders can be very large, which makes ingredient addition easier. The solids conveying zone in nonintermeshing twin screws is very similar to a single screw, and grooved feed zones have been used to increase the extruder throughput capacity. Figure 11.37 in Chapter 11 showed the feeding location in a nonintermeshing, parallel twin screw extruder.

Figure 12.4. Downstream feed port corotating extruder.

Downstream solid feeding is normally done with a side feed extruder (shown in Fig. 11.35, Chapter 11) or a stuffer. While gravity feeding downstream is possible, the screw flights must be deep enough to accept the additional product plus the melt that is already present. To decrease the fill, a melt seal or restricted area in the screw design is incorporated prior to large-pitch elements where the downstream ingredients are being fed. This creates elements with low fill, while providing a high free volume for adding other ingredients. A side feed extruder or stuffer is a positive feed system that forces material into the screw channel down-

stream. Most side feed extruders are corotating twin screw extruders with no flight compression. Different screws are available, depending on the materials to be fed. Powders are fed with powder screws that have low pitch, low channel depth, and high intermeshing. Pellets or reinforcements are fed with large-pitch screws with minimum intermeshing. Pellets have to flow freely without binding between the screws and the barrel wall, while reinforcements (glass, carbon, aramid, etc.) are transferred to minimize fiber attrition. The additive concentration in the final composition depends on the feeder accuracy. Gravimetric downstream feed ports are normally round or rectangular and expose as much of the screw flights as possible to increase the fill. If a side feed extruder is used, a figure 8 transition is required to feed material to the extruder barrel.

Ingredients added downstream include

- Temperature-sensitive material that may degrade if subjected to the same heat as the other ingredients in the formulation.
- Reinforcements that require maintaining their aspect ratio. Addition to the melt is less stressful to fibers or other high aspect ratio materials than addition to the feed throat with hard pellets present. Conveying fiber reinforcements in the feed throat with plastics pellets generates significant fiber attrition and screw wear.
- Shear-sensitive materials that may degrade if run through the entire extruder.
- Liquids that can be mixed into the melt easily. Feeding liquids into the extruder feed throat can produce a lubricating film on the barrel wall, causing the resin to slip and hindering feeding efficiency. Pellets and liquids do not mix very well, as pellets cannot absorb the liquid. Once a melt pool is present, liquid is readily dispersed in the polymer melt to provide a uniform mixture.

Liquids are fed downstream through a liquid injection nozzle (Fig. 11.33, Chapter 11) using a positive displacement liquid additive pump. The pump has to generate enough pressure to overcome the pressure in the barrel. A gravimetric liquid feed pump that pumps mass per unit time versus running at a set rate is more accurate. Feeding volumetrically (fixed pump speed) works very well for most low- to high-viscosity fluids. However, if the liquid temperature increases during the production run or trial, the liquid density normally decreases, causing less liquid by weight to be pumped at the set pump rate. Temperature variations in the liquid can create dramatic changes in the liquid feed rate using a positive displacement pump. At start-up the liquid injection pump is turned on at a slow rate, simultaneous to the resin formulation addition to the extruder. This prevents any molten resin from flowing into the injection nozzle and freezing off or fouling the injection nozzle so it won't feed properly during the run. Unlike solid feed, liquid can be fed into any screw element, not just conveying elements. Feeding liquid into kneading blocks or gear mixers may actually improve the mixing and product performance.

Counterrotating, intermeshing, conical extruders normally meter the entire formulation into the extruder feed throat. A large feed opening with deep, narrow screw channels is used to ensure free powder flow into the screw flights. (Most conical twin screw extruders are used to process rigid PVC in powder form. The PVC is preblended with stabilizers and other ingredients, then fluxed and heated in a high intensity mixer prior to the extruder. The preblended ingredients may still be warm when added to the extruder, so less heat is required to be supplied by the extruder screw and barrel.) The feed zone determines the polymer feed uniformity and percent fill over the extruder length. Double flighted conical screws feed material better and compress material more efficiently than single flighted conical screws that have the same channel depth over the entire screw length. Figure 12.5 shows the difference between single and double flighted conical feed sections.[1] With their deep narrow channels, conical twin screws have more surface area and better heat conduction than parallel extruders. Conical extruders have 45% more surface area for preheating material, compared to parallel extruders.[2]

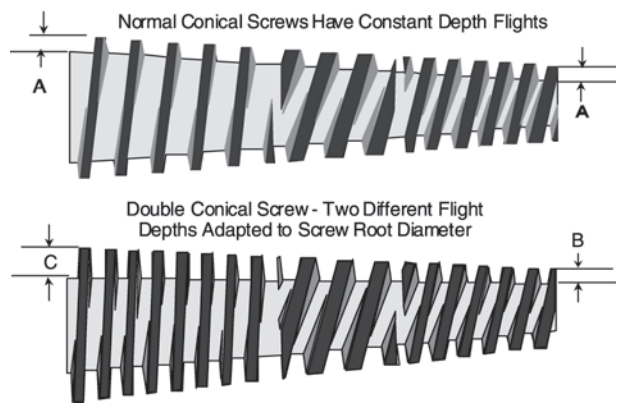

Figure 12.5. Types of conical screws.[1]

12.2 Plasticating and Melting

There are four mechanisms to melt polymer in an extruder:

- Conduction melting without melt removal
- Conduction melting with drag melt removal
- Melt initiation by plastic energy dissipation (PED) and frictional energy dissipation (FED)
- Dissipative mix melting[3]

Under normal operating conditions, the extruder motor, through shear heating, supplies 80–90% of the

energy to plasticate polymer. As polymer is conveyed down the extruder barrel, conductive heat is transferred to the plastic from the barrel walls. Resin and additives are compressed by reducing the conveying element pitch as the material moves forward. Reducing the element pitch decreases the free volume, compressing the material while supplying some shear heat. As the formulation is conveyed forward in the first two or three barrel sections, it is preheated and compressed in the conveying elements. The melting point location is defined by the screw configuration and the objective to be accomplished with the material once melting is complete. Using a high L/D extruder with many barrel sections, the melting point location depends on what other functions are to be accomplished downstream, such as mixing, venting, liquid feed addition, or solid feed addition. Melting can be accomplished in a relatively short L/D after the material is added to the extruder, depending on the screw elements used in the screw configuration. Adding the formulation into barrel zone 1 (no barrel heat present), material is preheated and compressed in barrel sections 2 and 3, and melted in barrel zones 3 and 4. Depending on the extruder length, it can be melted in barrel section 3. Figure 12.6 shows a screw design to complete melting in barrel section 4 on a corotating extruder. Feed is fed into zone 1,

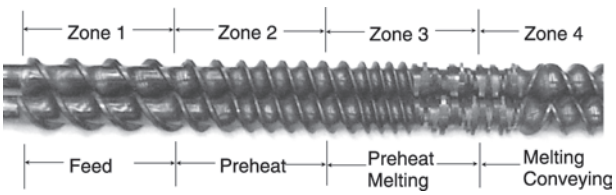

Figure 12.6. Typical screw design to melt polymer.

where deep-flight, high free volume SK elements are present to accept the polymer formulation. Zone 2 has an SKN transition element to go from the undercut SK flights to the normal conveying elements. Material is preheated in zone 2 and compressed as the screw element flight pitch decreases from 90 mm to 45 mm to 30 mm in zone 3. A set of right-handed kneading blocks starts to work the polymer toward the end of barrel zone 3. It is followed by a second, smaller set of right-handed kneading blocks, a small neutral kneading block, a small left-handed kneading block, and a left-handed conveying element. The left-handed kneading block and conveying element act as a melt seal, with positive conveying back toward the feed throat. Pressure is generated on the polymer melt by the forward pumping capacity of the conveying elements in zones 2 and 3. The left-handed conveying element and left-handed kneading block in zone 4 force the material back toward the feed throat. When the pressure is sufficiently high from the polymer being pumped forward to force the material past the left-handed conveying element and left-handed kneading block, the molten polymer flows over the kneading blocks and left-handed conveying element as it moves downstream. This action melts the formulation before it can pass to the next conveying elements at the end of zone 4. Work supplied by the screw and motor completely melt the formulation in this section.

The melt seal developed by the left-handed conveying elements prevents any air in the partially filled screws prior to the kneading blocks from passing downstream. The left-handed conveying elements force the flights just prior to it to be completely filled with polymer. Therefore, any air entering the extruder with the formulation is forced back through the feed throat. This can have a negative effect on polymer feeding, particularly if it is powder. Since the powder bulk density is normally low, air blowing back out the extruder countercurrent to the powder being added can lower the bulk density more, creating feed issues. One alternative is to put an atmospheric vent in zone 2 to remove any air being forced back toward the feed throat by the restrictive elements at the end of the melt section. The second problem associated with air coming back out the feed throat is dusting and housekeeping from airborne powder or fines being blown out the feed throat.

In the event the polymer added to the extruder is powder, it may be advantageous to change the melting section to contain only kneading blocks that convey material toward the die, removing all reverse elements. Forward conveying kneading blocks do not create as tight a melt seal, allowing some air entering the extruder with the feed to pass through the kneading blocks. The air can be removed downstream by either a vacuum vent or atmospheric vent. A possible screw design, shown in Fig. 12.7, melts the powder feed while allowing some air to pass

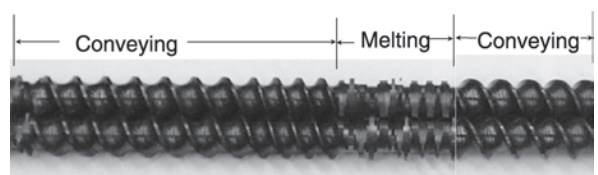

Figure 12.7. Melting section with no rearward conveying elements.

through the melting section to vent downstream. How melting is accomplished depends on the resin's oxidative thermal stability. In some situations, oxygen presence in the melt state causes resin degradation.

General screw design guidelines suggest conveying elements between the feed throat and the melting zone have reduced pitch to decrease the free volume and increase the percent fill. Restrictive elements are used in the melting zone or at the end of the melting zone to work the material, creating higher shear heat to melt the formulation. If the melt temperature rise in the melting section is too high due to the shear heat generated by the restrictive elements used in the melting section, convey-

ing elements can be placed between kneading blocks to assist in reducing and controlling temperature.

The screw design and elements used to melt the formulation depend on the formulation particle size (pellets versus powder), the resin softening temperature or melting point (amorphous versus crystalline), and whether the polymer viscosity is heat or shear sensitive. Other factors to be considered include liquid addition, formulation lubricity, and other additives present that may affect the screw's ability to feed and convey forward.

Melting in a counterrotating, parallel extruder is similar to that in a corotating twin screw. The melting section is designed to shear and work the formulation in a given location in the extruder by using reduced pitch conveying elements, kneading blocks, and restrictive elements to localize melting.

In a conical twin screw, the material is compressed as the screw flight volume decreases. As the screw root diameter and flight diameter steadily decrease, the channel volume decreases, compressing the polymer. The reduced volume coupled with the intermeshing, counterrotating screws work the material as it travels between the gap in the two screws, supplying the shear energy required to flux and soften the PVC formulations. Figure 12.8 shows a screw set used in conical extruders and

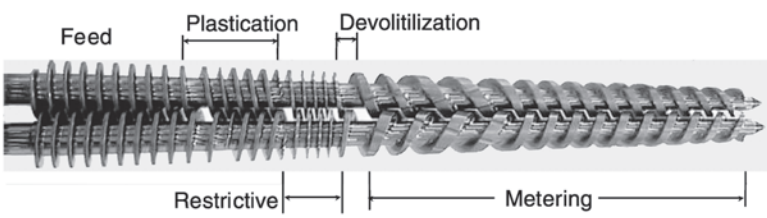

Figure 12.8. Conical twin screw feeding and plasticating.[4]

locates the melt or plastication pool. The narrow screw channels provide conductive heat, which assists plastic melting. The large flights conduct heat from the barrel wall to the screw root, heating the polymer in the narrow channels and reducing the shear heating requirements.

12.3 Melt Conveying

Once the material is melted, it is conveyed forward in the barrel by conveying elements. Figure 12.9 shows a possible screw element configuration after the melting section to convey the melt forward to a mixing or downstream feeding section. This particular conveying section uses large-pitch elements to minimize the work and reduce the pressure on the melt. Conveying in a two-lobe, intermeshing, parallel extruder contains three distinct melt pools around the screw elements that are moving down the screw in three distinct channels. Figure 12.10 shows an end view of the screw elements with an outline of a simulated barrel and the location of the three distinct lobal pools of molten polymer. As the polymer is conveyed down the extruder barrel, each lobal pool is in a different screw channel. There are (2*n*-1) lobal pools, where *n* is the number of lobes on the screw element. For a three-lobe screw element, there are five lobal pools.

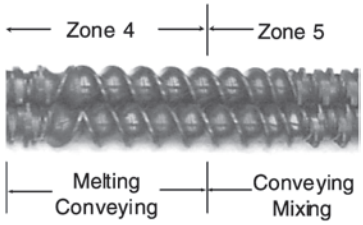

Figure 12.9. Conveying after mixing.

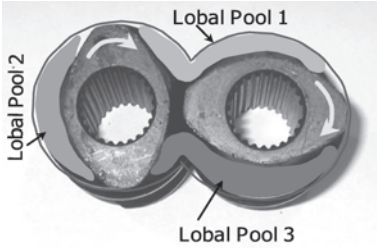

Figure 12.10. Three distinct lobal pools.

12.4 Mixing

There are two types of mixing that occur in twin screw extruders, regardless of whether they are corotating or counterrotating, parallel or conical, or intermeshing or nonintermeshing. Distributive mixing, as its name implies, distributes particles uniformly throughout the melt. Distributive mixing is a low-shear process and is accomplished by breaking and recombining the melt stream. Normal applications for distributive mixing include distributing:

- Fibrous materials (carbon fiber, fiberglass, aramid fiber, etc.)
- Fillers with high aspect ratios (mica) or
- Mixing shear-sensitive polymers and additives

The second type of mixing is dispersive, which breaks up large particles and disperses them as smaller particles throughout the melt. Dispersive mixing is a high-shear process used to

- Disperse pigments, liquid additives, and nonreinforcing fillers
- Alloy or blend two or more polymer resin systems

Figure 12.11 shows a diagram of both distributive and dispersive mixing. Different mixing elements and screw geometries are used to accomplish each type of mixing. Mixing effectiveness or level is dependent on

- Screw speed
- Percent fill in the screw channels
- Temperature
- Screw geometry

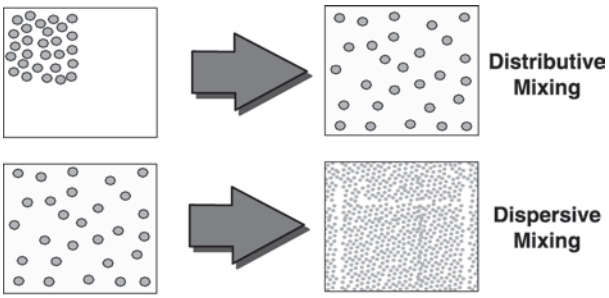

Figure 12.11. Comparison of distributive and dispersive mixing.

Screw geometry affects the shear rate, which influences the resin viscosity.

Narrow kneading blocks provide good distributive mixing as material flows in and around the blocks, but not between the block and the barrel wall. Different kneading blocks are available, ranging from wide blocks to narrow blocks, forward or rearward conveying, and nonconveying (neutral block) configurations. Figure 12.12 shows forward and reverse narrow disk kneading blocks produced by Theysohn Corporation. Each block is cut away on both edges to make the blocks narrower, which is better for distributive mixing. Kneading blocks were also shown in Fig. 11.43 in Chapter 11. Figure 12.13 shows wide block kneading elements without the shoulder on each side of the block. Gear mixers, shown in Fig. 12.14, are supplied by Theysohn America Incorporated,[5] Davis Standard®,[7] and other twin screw manufacturers. Gear mixers are used for distributive and dispersive mixing. The number of teeth, the conveying angle, and the spacer width may vary slightly, depending on the supplier, but these mixing elements all split the melt stream in numerous places and then recombine it on the other side of the gear in the spacer area. Sometimes three or four gear mixers will be used in succession to break and recombine the melt stream. Gear mixers are particularly useful when mixing liquid additives into the polymer melt stream.

Figure 12.12. Theysohn narrow kneading blocks.[5]

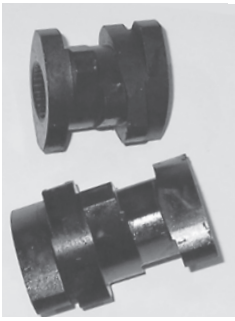

Figure 12.13. Wide forward conveying kneading blocks.

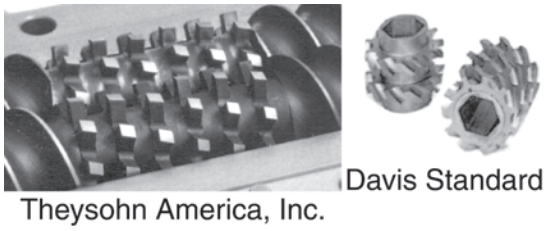

Figure 12.14. Gear mixers.[5][7]

Wider kneading blocks are used for dispersive mixing. With the narrow disks, molten polymer and additives flow around the disks; but with the wider disks shown in Fig. 12.13, material goes over the top of the disks as well as around them, providing dispersive mixing. William Thiele[8] shows how material flows over wide kneading blocks and around narrow blocks. A lobal pool formed in front of a wide block goes through the small gap between the flight top and the barrel wall. As it passes over the block, dispersive mixing occurs. This is shown graphically in Fig. 12.15. Figure 11.42 in Chapter 11 has an example of a wide kneading block, shown in the lower right-hand corner.

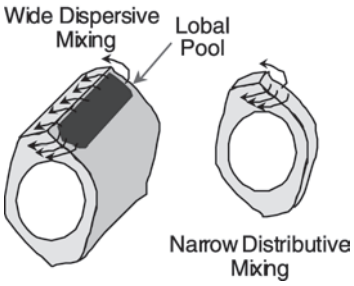

Figure 12.15. Material flowing around and over kneading blocks.

Neutral kneading blocks can be used for either distributive or dispersive mixing, depending on each kneading block width. Neutral kneading blocks have neither forward nor rearward conveying attributes in the extruder. They come in different lengths with either three or five blocks. Figure 12.16 shows a 90/5/45 neutral kneading block. The kneading blocks are positioned 90° from each other. If the blocks are wide, they provide more dispersive mixing, while narrow blocks provide more distributive mixing. As an example, in a 40 mm twin screw extruder, a 90/3/15 is 15 mm long with three blocks or lobes that are 5 mm wide; this element provides good distributive mixing. A 90/5/60 in a 40 mm twin screw is 60 mm long with five blocks that are 12 mm wide; it will provide more dispersive mixing. The wider individual disk provides the opportunity for dispersive mixing. Whether it provides the appropriate dispersive mixing for a given system requires testing.

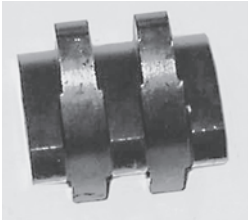

Figure 12.16. Neutral kneading block.

Figure 12.17 shows two mixing elements supplied by Farrel Corporation.[9] The patented Polygon mixing element is used for distributive mixing. It squeezes material,

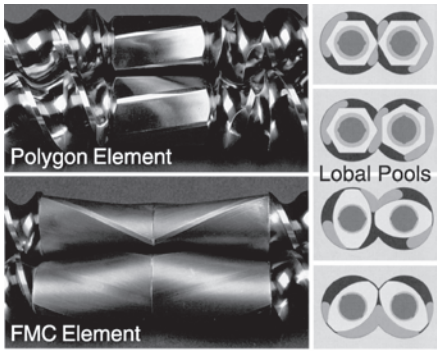

Figure 12.17. Dispersive Mixing Elements.[9]

conveys melt forward, and continuously splits the melt stream to provide distributive mixing. The second element is a continuous mixing element similar to a wide kneading block. However, the shape induces cross-channel flow, providing aggressive mixing. Many special elements are available from different extruder manufacturers to provide specialized mixing capabilities with their equipment.

Similar to melting, various elements are combined to provide specific mixing capabilities for the formulations being processed. To minimize heat build-up in multiple kneading blocks due to shear, the blocks can be separated with conveying elements. The mixing section design depends on

- Mixing type required by the formulation
- Screw mixing length available
- Polymer and additives temperature sensitivity
- Polymer and additive shear sensitivity in the formulation

Some mixing is accomplished strictly with conveying elements by going from a large-pitch element to a small-pitch element back to a large-pitch element, and so forth. This repeated material compression and expansion does provide some mixing; however, kneading blocks and other mixing elements are much more efficient mixers.

In conical twin screw extruders, a restrictive zone, shown in Fig. 12.8, follows the plastication or melting zone. The restrictive zone seals off the plastication zone from the devolatilization zone. Mixing will occur in the restrictive zone, compressing the polymer in the narrower channels. The restrictive zone compressive capacity is determined by its length and the channel volume. Mixing also occurs downstream in the metering section as the plasticated polymer is homogenized by the reduction in channel size.

12.5 Downstream Feeding

Unlike single screw extrusion, numerous materials, both solid and liquid, are fed downstream in parallel twin screw extrusion. This provides maximum extrusion flexibility by allowing materials to be introduced into the melt at different stages or locations along the extruder barrel. Examples of downstream feeding advantages are

- Adding fibrous material (glass, carbon fiber, stainless steel, aramid) to minimize fiber attrition
- Adding high aspect ratio fillers (mica) to minimize attrition in aspect ratio
- Adding shear- or temperature-sensitive materials that may degrade if put through the entire extruder
- Adding liquid feed to the melt as a plasticizer, liquid colorant, stabilizer, or lubricant

When added to the feed throat with solid polymer, high aspect ratio fillers and/or reinforcements can experience severe size attrition. This aspect ratio reduction lowers properties. Fibers conveyed in the feed section and compressed with solid pellets break and cause extensive screw wear. Downstream feeding allows the introduction of the high aspect ratio fillers or reinforcements into a melt, which assists feeding by acting more as a lubricant, conveying the material forward in the barrel. Mixing elements used after reinforcement addition downstream are relatively mild to provide sufficient mixing while minimizing fiber attrition. When processing high-temperature polymers that require low-temperature additive addition, downstream feeding decreases additive volatilization or degradation. As an example, assume a concentrate is being made with polycarbonate that uses an additive with limited thermal stability. The polycarbonate is added to the extruder feed throat. After it is melted, the additive is introduced downstream to minimize the thermal degradation that is a function of time, temperature, and shear history.

Liquid addition in the feed throat can lead to feed problems, due to the liquid lubricity on the barrel wall in the feed zone. In addition, any restrictive elements used to melt or plasticate the polymer in barrel zone 3 or 4 may prevent the liquid feed from passing through this section. In that event, the liquid feed will back up and come out of the extruder feed throat. Liquid is normally added more efficiently downstream by a liquid feed pump.

Downstream feeding introduces room temperature material into a polymer melt that is substantially higher in temperature. If the downstream additive volume is relatively high compared to the polymer volume, this can substantially alter the polymer melt temperature. Care must be taken to ensure that crystalline materials with sharp melting points are not cooled below their melting points, where localized solidification can happen. In this situation, additional heat is required to remelt the crystalline material, because the solid polymer may damage high aspect ratio fillers or reinforcements, causing higher attrition as the lubrication from the polymer melt is reduced.

Fillers, fibers, and polymers for alloying or blending are fed downstream. Additives such as flame retardants,

stabilizers, lubricants, and colorants are done by gravity feeding a stuffer box or a side feed extruder. Selecting large-pitch screw elements (Fig. 12.1) that aren't full of polymer provides space for the added ingredient. A vacuum or atmospheric vent is normally used to remove the air introduced with the downstream addition. Otherwise the air has to be removed through the barrel section where the downstream feeding takes place. Air removal through the same port as feed addition occurs may lead to poor additive feed characteristics, particularly with low bulk density powders. Figure 12.18 shows a Davis Standard® D-Tex twin screw extruder with two downstream feed locations using side feed extruders.

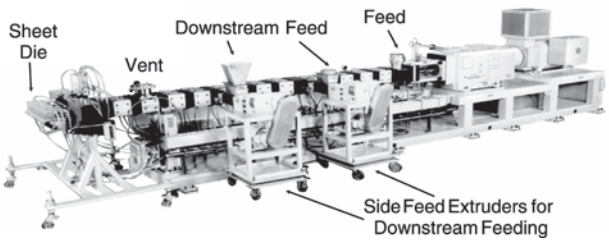

Figure 12.18. Davis Standard® D-Tex system showing downstream feeding in two locations.

In downstream feeding of solids, the following factors (discussed generally above) need to be considered for successful feeding:

- Lower the screw channel percent fill by using larger-pitch elements to accommodate the new feed stream entering the extruder.
- Feed stream temperature differential compared to the extruder melt may cool the melt, raising the resin viscosity.
- Downstream feed addition location must provide sufficient extruder length after the addition for melt mixing and homogenization with the new ingredients.

The screw design selected must accommodate these criteria to produce an acceptable product.

Figure 12.19 shows a screw design with

- Mixing prior to the side feed extruder
- Large-pitch elements in the area of the side feed
- Mixing after the side feed extruder to mix and homogenize the resin plus the new additive

Just prior to introducing the downstream feed stream via the side feed extruder, there are a number of kneading blocks that inhibit the molten polymer flow, forming a partial melt seal, so the large-pitch elements at the side feed are only partially filled with molten polymer. After the side feed there are some high-pitch conveying elements that allow the downstream addition to be partially preheated through conductive heating from the barrel before introducing shear heat through compression using lower-pitch elements. Once the material is at least partially heated, it is mixed with kneading blocks. At the end of the mixing section, a rearward conveying element acts as a melt seal prior to the vent. Large-pitch elements under the vent provide the melt with a high surface area with low pressure, where volatiles and air are easily removed. At the extruder end, small-pitch elements are used to foster melt pumping to the die.

Figure 12.19. Suggested screw geometry for downstream feeding.

Liquid feed downstream is pumped in through a liquid injection nozzle connected to a positive displacement pump running at a fixed speed. Assuming the temperature of the fluid being fed is constant, the liquid feed rate remains constant. If the temperature changes, the fluid viscosity and density increases or decreases as the temperature goes down or up. In a constant-speed pump, the throughput will decrease or increase depending on the change in viscosity or density. Consequently, it is important to maintain the liquid at uniform temperature. If a gravimetric liquid feed pump is available, the temperature is not critical, as the pump rate increases or decreases to supply the same liquid mass per unit time. A liquid feed port in a vent plug is shown in Chapter 11, Fig. 11.33. Going back to Fig. 11.30, remove any extruder vent plugs and replace them with a liquid injection port vent plug. The injection port is connected through metal tubing to a liquid feed pump output, shown in Fig. 12.20. For safety reasons,

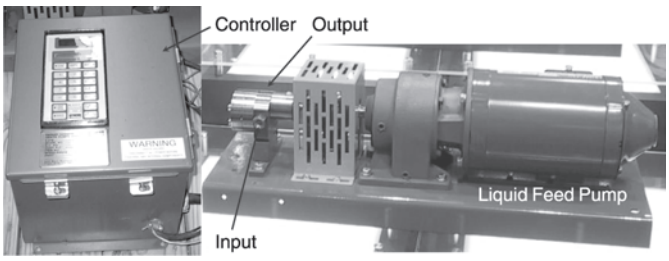

Figure 12.20. Zenith liquid feed pump and controller.[10]

either a pressure relief valve or pressure gauge is installed in the line going to the extruder to monitor the pressure and prevent line rupture in the event high pressure occurs due to feed port blockage. A volumetric liquid feed pump must be calibrated. This is done by plotting various throughput rates versus pump speed. Based on the curve, a given pump speed is selected to supply a specific mass per unit time. Once the pump speed is selected, two or three rate checks are run at the specified speed to ensure a uniform feed rate.

The screw design under the liquid injection port is not as critical as where the large-pitch elements are used at either the main feed throat or side feed entrance. Provided the liquid feed pump pressure is higher than the melt pressure in the extruder where the liquid is injected, the pump will deliver a uniform feed to the extruder. Some studies have suggested that liquid feed over kneading blocks or gear mixers (Fig. 12.14) gives better mixing than feeding into large free volume conveying elements and mixing farther downstream. When a large liquid volume is added, it may be added with two or more pumps in different locations along the barrel to obtain a uniform distribution in the melt. This is not a problem, provided there are enough liquid feed pumps and vent barrels to add the liquid and enough compounding capacity to mix the liquid into the polymer stream.

When starting the extruder, small amounts of liquid are added through the various liquid injection ports to prevent pressurized molten polymer from flowing into the nozzle, freezing off and plugging the injection nozzle in that location. If the injection nozzle gets plugged with polymer, then the extruder is stopped and the liquid injection nozzle cleaned or replaced with a clean nozzle. At lower throughput rates, smaller-diameter nozzles are required, while larger nozzles are needed for higher throughput rates.

Downstream feeding must be planned when setting up the extruder to provide a suitable mechanism to feed resin, additives, fillers, reinforcements, or liquids downstream. Some methods available to support feeders are

- Hang them from a frame above the extruder.
- Place them on a mezzanine.
- Place them on adjustable carts beside the extruder.

Important criteria to consider when locating feeders are

- How is the hopper feed level going to be maintained during the run?
- Is the feeder easy to remove and clean between runs?
- Can different size feeders be installed at different locations to maximize the compounding line flexibility?
- Is the feeder isolated from vibrations during operation to provide uniform feed rates?
- What is the material flow path to the feeder?
- Is drying critical? If so, how is the dried material transferred to the feeder and kept dry during the operation?
- Is dusting a housekeeping problem during operation?
- Is cross-contamination possible, and is this an issue from one run to the next?

12.6 Devolatilization

Venting, whether opening the barrel vent port to the atmosphere or drawing a vacuum, is used to remove air, moisture, and/or volatiles prior to the die. The vent port is normally at least one barrel section from the extruder end to prevent pressurized polymer from backing up the barrel and flowing out the vent. If the die pressure is too high, the screw flights will fill with polymer back to the vent port. Or if the screw speed is too slow to pump the material out the die, polymer can back up in the extruder and exit through the vacuum port or vent.

In intermeshing extruders, a melt seal is designed into the screw configuration prior to the vent, followed by large-pitch elements under the vent port to provide the surface area and the low melt pressure. The melt seal provides a high pressure drop prior to the vent to assist devolatilization. Melt seals are most efficiently created by rearward conveying elements (both conveying and kneading blocks). Blister rings also function well as melt seals. Neutral kneading blocks and other elements that compress or hold up the melt can be used, but are not as efficient. Normally melt seals are used before a vacuum vent or downstream feeding. Low percent fill, Fig. 12.21, combined with a high pressure drop provides better devolatilization on venting. In counterrotating, intermeshing, parallel extruders, multiflighted forward conveying elements are used. Figure 12.18 shows the vent section of parallel, corotating screws. Figure 12.19 shows the screw design on a 45 mm corotating screw with large-pitch elements under the vacuum port with kneading blocks and a melt gear directly ahead of the vacuum area. Just prior to the large-pitch element is a rearward conveying element that acts as a melt seal.

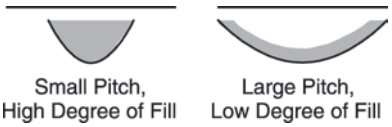

Figure 12.21. Comparison of degree of fill.

Different vent port configurations were shown schematically in Fig. 11.45, Chapter 11. Figure 11.41, Chapter 11, shows an actual vent port with cover glass. In operation, the cover glass becomes dirty from the volatiles being removed from the polymer melt and condensing on the sight glass. To view the vent operation under vacuum, the vent glass must be kept clean. A trap is placed between the vacuum pump and the vent port to trap volatiles and prevent melt from flowing out the vent into the vacuum pump. Different vacuum pumps can be used. Ideally, the pump is large enough to draw a vacuum to approximately 28 inches of mercury.

As mentioned previously, a vacuum or atmospheric vent may be required near a downstream feed port to remove any air entrapped in the feed stream that is enter-

ing the extruder. Melt seals at different locations along the extruder (before a vent and most feed ports) prevent air from traveling along the extruder length to the closest atmospheric or vacuum vent. Consequently, any air entering with a particular feed stream must be vented from the extruder or flow back out the feed port it entered. Air does not pass through a good polymer melt seal in the extruder.

Devolatilization in conical twin screw extruders occurs right after the restrictive zone and is shown in Fig. 12.8. Devolatilization removes air from the plasticated PVC and removes volatiles that can cause holes, pores, or a foamy structure in the final product. Similar to other twin screws, the devolatilization zone has a coarse pitch screw flight with partial filling, providing both an appropriate pressure drop and high polymer surface area.

12.7 Pumping

The last area in the extruder pumps or meters molten polymer to the die. Figure 12.19 shows the pumping area at the screw discharge. Polymer in large-pitch screw elements under the vent is gradually compressed by the lower-pitch elements to fill the melt channel. Small-pitch elements in the metering section provide good pumping capability. Smaller-pitch conveying elements compress the melt into smaller and smaller channel areas and may increase the melt temperature slightly. Singleflighted, forwardconveying elements can be used in the metering section to reduce backward pressure flow resulting from die pressure and leakage flow backward over the flights. Singleflighted screws can minimize melt temperature increases that occur when smaller- and smaller-pitched conveying elements are used.

One problem with parallel, corotating, intermeshing twin screw extruders is their poor pressure generation. For years, corotating, intermeshing twin screws were used primarily for compounding, while single screw extruders were used for extrusion processes such as film, sheet, and other processes requiring higher die pressures. Today corotating twin screws find a home in the sheet and film industry, as processors integrate compounding customized formulations into the sheeting process. To produce uniform thickness sheet, gear pumps or short single screw extruders are installed after the twin screw extruder to provide uniform polymer melt flow, melt pressure, and melt temperature to the die. Figure 12.22 shows a coextrusion sheet line at the end of the corotating twin screw extruder with an automatic screen changer and a gear pump between the extruder end and the die. Gear pumps are added to numerous extrusion lines to provide positive and constant polymer melt pressure to dies to ensure constant product dimensions. Gear pumps are covered in more detail in Part 5, "Auxiliary Equipment."

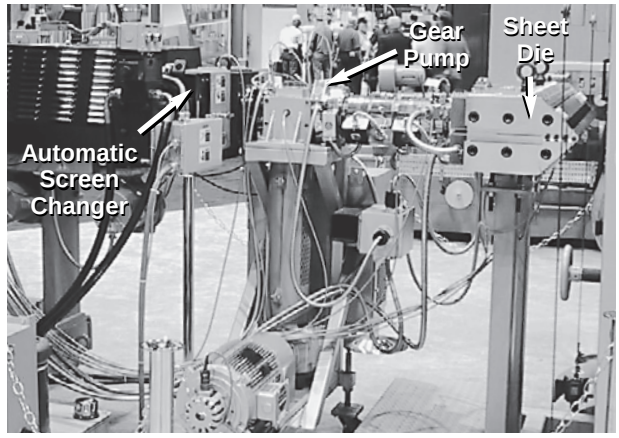

Figure 12.22. HPM twin screw extruder with automatic screen changer, gear pump, and sheet die.[11]

Counterrotating, intermeshing twin screw extruders generate higher and more uniform die pressure than corotating extruders. Counterrotating extruders are used extensively with rigid PVC and other resin formulations in profile and pipe extrusion. The extruders normally run at lower speed with high screw fill and high melt pressure.

Nonintermeshing twin screw extruders may have one screw longer than the other, using the longer screw to act as a pump to provide sufficient and uniform die pressure. This twin screw extruder is used for melting, mixing, and devolatilization, and the single screw extruder is used for pumping. Figure 12.23 shows a twin screw extruder schematic with one screw longer than the other; the single screw feeds the die.

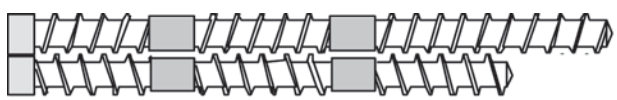

Figure 12.23. Nonintermeshing twin screw extruder with one screw longer than the other.

12.8 Die

The die is attached to the extruder end either directly or with a transfer pipe or adapter, depending on the die size and the process. In Fig. 12.22 an automatic screen changer is attached to the extruder end, followed by a transition piece to the gear pump, followed by a second transfer pipe connecting the die to the gear pump.

Some functions the die provides are

- Shape the extrudate into the desired cross sectional dimensions at a specific throughput rate
- Contribute to the physical properties by controlling molecular orientation
- Control product surface aesthetics

An efficient die provides a specific cross sectional area with tight tolerances, acceptable die pressure drop,

good surface aesthetics, and good melt homogeneity at high throughput rates. Sophisticated dies with feedback loops sense product variations in the cross sectional dimension, automatically changing the die settings to produce the correct profile. Dies producing complex cross sectional dimensions require uniform wall thickness and flow in the various die channels to eliminate final product warpage.

Good die construction ensures

- Ease of maintenance
- Efficient sealing between all elements to prevent polymer leakage
- Quick connection and disconnection to the extruder
- Sufficient mechanical strength to minimize deformation under pressure
- Correct location and sufficient electrical power to provide uniform heat to the die
- Ability to use the die for more than one polymeric material
- Low manufacturing cost

Four zones between the extruder exit and the die exit produce the final cross section. First, the transition zone or adapter transitions from the figure 8 to a round cross section. Second, the screen pack and breaker plate filters the melt. Third, different shapes (depending on application) lead to the die lips. Fourth, the melt acquires its final characteristics and shape in the parallel zone before exiting the die. The parallel zone or die land area controls to a degree the die swell, back or head pressure, and flow uniformity in the extruded part cross section. In an ideal world, the die opening has the flexibility to be modified by altering the profile dimensions and controlling the melt temperature, which controls the polymer and product characteristics without requiring a different die. Things that can be adjusted in a die are

- Mandrels
- Choker bars
- Die lip
- Localized temperature controls

Die adjustments can help compensate for differences in melt rheology and/or material. In practice, die adjustments are often left to operators. Since this is a critical step in obtaining proper product dimensions, experienced operators with both material and equipment understanding are required to produce high-quality parts.

Many corotating twin screw extruders are used for high-speed compounding, where the die is either a strand die to make pellets or a die face pelletizer that pelletizes extrudate immediately upon exiting the die. Figure 12.24 shows a small strand die for pelletizing. These relatively unsophisticated dies have an 8–O transition (going from

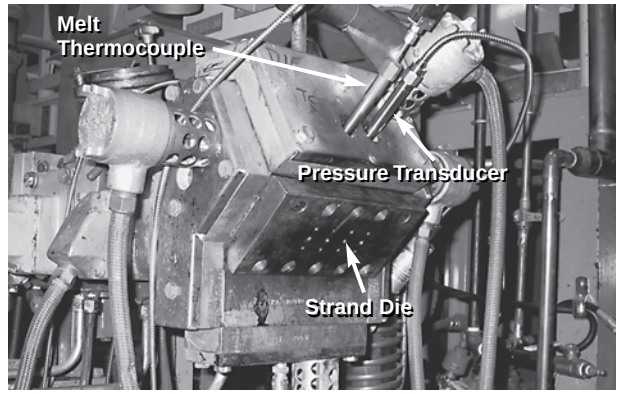

Figure 12.24. Four hole strand die.

the figure 8 barrel configuration to a round die configuration) at the die entrance. The die fans out from the round entrance to a flat strip in front of the strand holes. There are no dead spots for polymer to accumulate. A melt thermocouple and melt transducer measure the polymer melt pressure, temperature uniformity, and optimum processing values. Depending on the extruder output, these dies can be very wide. Strand cooling is critical in determining the optimum water bath length required. As the throughput increases (normally measured in pounds/hour/hole), the time spent in the water bath decreases, resulting in high-temperature strands during cutting. Strands with insufficient cooling prior to packing may stick together and form clumps in packages.

Corotating twin screw extrusion applications are expanding from strictly compounding to producing finished parts. Twin screw extrusion provides increased application flexibility by allowing end users to custom formulate their own materials without paying a compounder to produce a given formulation and then remelt it and reprocess it into an extruded shape. In addition to increased flexibility and productivity, the resin experiences one less heat history. There is better control, less regrind and start-up material, and it is easier to control the formulation ingredients. In some applications a gear pump is required between the twin screw extruder and die to ensure adequate die pressure.

While counterrotating twin screw extrusion has historically been used for profile extrusion, it is currently used in some compounding applications run at high screw speeds and rates. For compounding, the industry standard remains corotating, parallel, intermeshing extruders.

In counterrotating twin screw profile extrusion, the die lip dimensions are normally different from the product dimensions due to die swell, draw down, and pressure variations across the die. Differences in materials (amorphous or semicrystalline) and rheological properties (molecular weight and molecular weight distribution) affect the size difference between the product cross section and the actual die dimensions. Die swell is due to

extrudate swelling as it exits the die, Fig. 12.25. (Author's note: "Dies don't swell; extrudate swells," quoting Bryce Maxwell.) Polymer molecules or chains oriented in the flow direction in the die land area relax and reentangle once the material exits the die due to the elastic component. This relaxation and reentanglement process distorts the extrudate cross sectional area compared to the die cross sectional shape. Die swell depends on

- Shear rates in the die
- Melt temperature
- Die land length
- Reservoir length

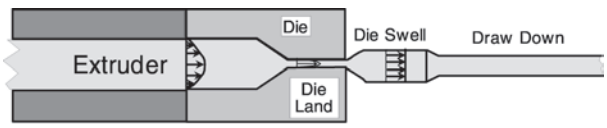

Figure 12.25. Die swell.

High die shear rates and melt temperatures create more die swell, while longer die land lengths and lower reservoir-to-land lengths lead to less die swell. Combinations of these factors can create a number of conditions giving the same cross sectional profile. Figure 12.26 shows typical distortion in a square cross section and the die shape required to produce a square shape.

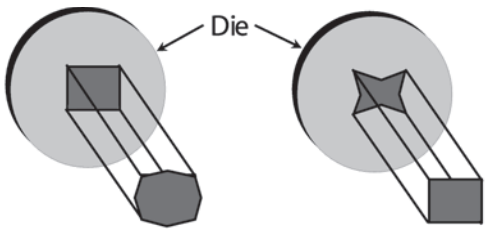

Figure 12.26. Extrudate distortion.

Product is pulled away from the die and drawn down to its final dimensions by the puller. When polymer drools from the die onto the floor, die swell at the die lips is obvious; however, when extrudate is pulled away from the extruder, die swell is normally hidden because the product necks down by the pulling or drawing operation. The draw ratio depends on the product size exiting the die versus the required product size. As draw increases, machine-direction molecular orientation increases, resulting in higher machine-direction product property (tensile and flexural) performance versus the transverse or cross-machine direction. Depending on the application, high molecular orientation can result in the material easily splitting in that direction. Extrudate at the die is oversized to compensate for the draw between the die and the puller. Table 12.1 provides cross sectional guidelines to be used with different polymers. The die opening size is not critical on dies with adjustable lips or openings relative to the final product dimensions because the die opening is adjustable. On the other hand, if the die is designed to produce a given cross section and no die lip adjustments are available, the die has to be cut correctly the first time; otherwise it will have to be discarded, recut, or metal welded back on to the opening and remachined. When cutting new dies, the cross sectional area is normally cut smaller than the required dimensions, because it is easier to recut the die opening, increasing the channel size, than to weld steel back to the die and recut to decrease the channel size.

Table 12.1. Cross Sectional Dimensions versus Channel Size

Polymer	Increase in Orifice Size Relative to Cross Sectional Area of Part, %
Polyethylene	15–20
Rigid PVC	12–15
Flexible PVC	Small Profile: 8–10
Flexible PVC	5–10
Flexible PVC	Large Profile: 3–5
Polystyrene	8–10
Polyamide	15–20

Calculating die dimensions requires understanding polymer shrinkage due to cooling. All polymers shrink on cooling, with semicrystalline resins shrinking more than amorphous polymers. Product dimensions at high temperature are larger than at room temperature. With semicrystalline polymers, shrinkage can continue in the solid state if the part is heated above the glass transition temperature until the equilibrium crystallinity level is obtained. As semicrystalline polymers develop higher crystallinity, polymer molecules pack closer together and the part volume decreases, or shrinks. While shrinkage on part cooling may be insignificant in wall thickness dimensions (depending on the tolerance requirements) because the walls are relatively thin, length shrinkage can be quite substantial if the part is not completely cooled to room temperature prior to cutting. Assume you are producing a thick polypropylene (PP) profile that is eight feet long. It cools slowly and is cut to length at 120°F (49°C). PP has a linear coefficient of thermal expansion of 9.0×10^{-5} in/in/°F (5.0×10^{-5}mm/mm/°C). After the part cools to room temperature (assume 72°F [22°C]) from 120°F [49°C], the length has decreased by

$$\Delta L = L_0 \times \alpha \times \Delta T = 96 \text{ inch} \times 9.0 \times 10^{-5} \times (120 - 72) = 0.41 \text{inch}$$

or 10.4 mm. This means the part length reaching the customer is 95.59 inches (242.7 cm) instead of 96 inches (243.8 cm). This does not factor in any additional shrinkage that may result from increased crystallinity in the part. PP glass transition temperature is below room temperature; consequently, during cooling additional crystallinity development can still be occurring.

A die prerequisite is to produce extrudate with a uniform velocity profile throughout the entire cross section. The resistance to flow is given by Eq. (12.1):

$$R = \frac{\Delta P}{Q} \tag{12.1}$$

where R = Resistance to flow
ΔP = Pressure drop
Q = Volumetric flow rate

Changes in melt temperature and die geometry affect the resistance. Uniform resistance to flow is obtained by modifying the die land lengths in different sections of the die. Balanced polymer flow can help prevent distortion or warpage resulting from differential shrinkage later in the process. For a specific polymer, only a few die geometries yield the optimized process in terms of mass flow rate, dimensions, and physical property balance. Polymer rheology effects of altering die performance include the following:

- Melt temperature fluctuations affect melt viscosities, flow, and resistance to flow. Hot and/or cold spots in the die can change flow characteristics in that section of the die.
- Polymer batch-to-batch variations occur, particularly if regrind levels are changed or there is a larger variation in the regrind type used.
- Stress develops in the polymer during convergence from the reservoir to the land area as a part of extensional flow. (Extension flow is caused by stretching a material while reducing the cross sectional area. In the die, this is associated with stretching a molten polymer.)
- Stresses develop from stretching the extrudate outside the die.

Polymer problems associated with the die are: "fisheyes," plate-out, and melt instabilities. These extrusion problems are encountered in all extrusion processes, ranging from compounding to sheet and film to profile extrusion. Fisheyes are hard polymer specks protruding from the extrudate surface. These specks are high-molecular-weight particles made during the polymer production or are generated in the die dead spaces, where material has long residence time, allowing it to branch and/or cross-link. Small high-molecular-weight particles from the resin supplier may not melt in the extruder, passing through the screen pack and die. (This assumes a screen pack is used; in many twin screw extrusion applications there is no screen pack.) Plate-out or die lip build-up is from low-molecular-weight polymer or additive that migrates to the extrudate surface and is deposited or condenses on the die lips. Die build-up has to be chemically analyzed to determine the source. Low-molecular-weight polymer is from the low end tail on the polymer molecular weight distribution curve. To eliminate plate-out, a different resin can be purchased with a narrower molecular weight band or less low-molecular tail. If the deposits are additives that volatilize and recondense on the die surface, alternative additives can be used or processing conditions changed (lower temperatures) to minimize deposits. Regardless of the cause, solutions are necessary to minimize plate-out in order to maximize product yields and process efficiencies.

Melt instabilities include "sharkskin" and melt fracture. Sharkskin, also referred to as surface matteness, is a repetitious wavy or regular-ridged surface distortion perpendicular to the extrusion direction. Less severe sharkskin produces a matte finish, where a glossy, smooth finish is anticipated. Sharkskin is formed in the die land area due to rapid surface layer acceleration as the extrudate exits the die. In severe instances, the surface area can actually fracture. Sharkskin appears to be dependent on melt temperature and linear extrusion speed. Corrective actions to eliminate this problem are

- Increase the die temperature.
- Reduce the extrusion velocity.
- Use an external lubricant.

High viscosity polymers with narrow molecular weight ranges tend to be the most susceptible. Factors not contributing to this phenomenon are

- Shear rate
- Die dimensions
- Reservoir approach angle
- Die land surface roughness
- Extruder L/D ratio
- Extruder temperature profiles
- Materials used in die construction

A second melt instability, called melt fracture, is a highly distorted extrudate associated with die pressure fluctuations. It is not a surface phenomenon, as with sharkskin; instead, it goes completely through the extrudate. The surface may remain smooth, but the extrudate is distorted. Figure 12.27 shows four types of severe melt fractures with extruded strands:

A. Spiraling
B. Bambooing
C. Regular ripple
D. Random fracture

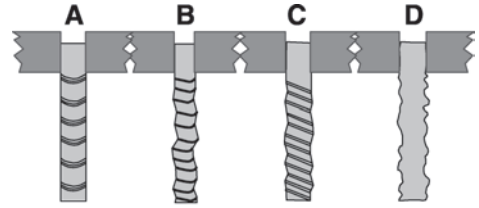

Figure 12.27. Examples of different types of severe melt fracture.

While melt fracture is similar to sharkskin, it is caused in the die land area; understanding of the cause and effect for severe melt fracture varies from polymer to polymer. If a critical wall shear stress in the die is present (between 15–60 psi [0.1–0.4 mPa]), melt fracture is likely to occur. Corrective actions to eliminate melt fracture include streamlining the die, slowing the extruder rate, reducing the melt viscosity, increasing the die land temperature, increasing the product cross sectional area, or adding external lubricants to the formulation.

Die design, particularly in the profile industry, has been a trial and error process. A good die designer may be able to design and produce an acceptable die in three to four attempts, while an inexperienced die designer may require six to eight attempts. Science has improved die design in recent years with computer-aided design and finite element analysis (FEA) to design dies and predict

- Flow uniformity
- Flow analysis inside and outside of the die
- Melt pressure
- Melt temperature
- Residence time
- Particle paths
- Stress analysis in the parts

Computer-aided design is particularly useful in blow molding, coextrusion, and profile applications.[12] Sheet, film, and coextrusion dies and blocks have sophisticated designs for specific cross sections, polymeric materials, and throughput rates. The benefits of computer-aided die design are

- It cuts design time and eliminates the trial and error process.
- It reduces testing time.
- It is applicable to many polymers.

Some die designs include a pipe or tubing die—Fig. 12.28; wire coating die—Fig. 12.29; flat sheet or film die—Fig. 12.30; blown film die—Fig. 12.31; flat plate profile die—Fig. 12.32; and coextrusion blown film die—Fig. 12.33.

Every extrusion process matches the extruder output to a particular puller speed to produce products with the

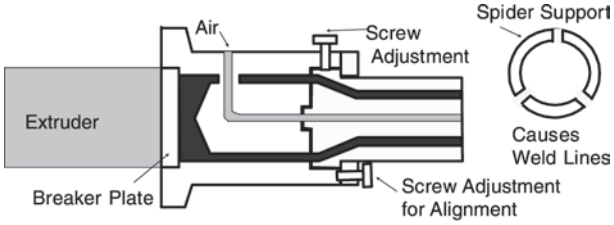

Figure 12.28. Pipe or tubing die.

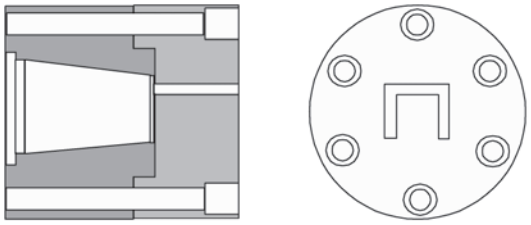

Figure 12.31. Profile die.

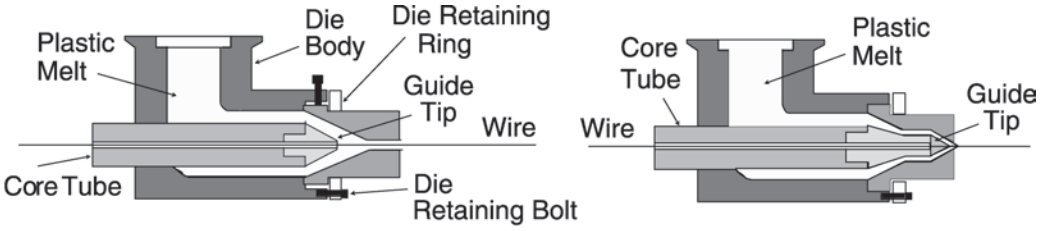

Figure 12.29. Two types of wire coating dies.

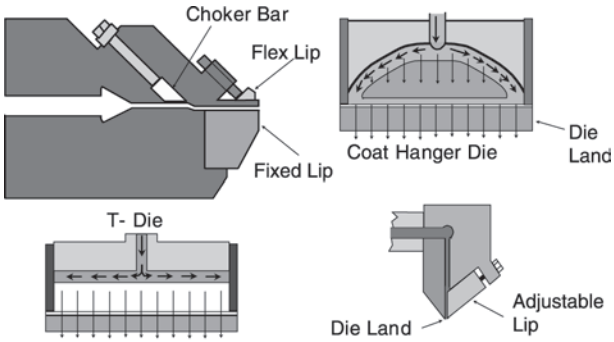

Figure 12.30. Sheet and cast film dies.

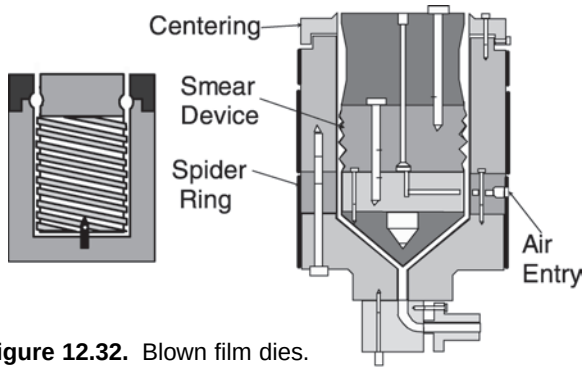

Figure 12.32. Blown film dies.

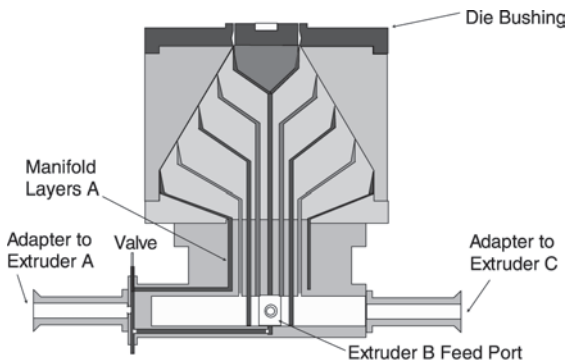

Figure 12.33. Coextrusion blown film die.

correct dimensions. There is only one throughput and puller speed ratio for a particular die opening that produces the correct product. Inherent with the ratio is a fixed molecular orientation that dictates the machine and transverse direction properties. Changing the puller speed or extruder speed independently, without changing the die opening, alters the product dimensions, orientation, and property profile.

12.9 Melt Pressure Profile

The polymer melt within the extruder goes through high-pressure and low-pressure regions along the barrel between the feed throat and die, depending on the screw configuration. Kneading blocks and melt seals generate high-melt-pressure regions along the screw length, while large-pitch conveying elements lead to low-pressure areas. In an area where only conveying elements are present, changing the pitch from a large pitch to a small pitch increases the melt pressure. Figure 12.34 shows a screw configuration in the first few barrel zones and a curve representing the corresponding pressure generation in the barrel. Melt pressure attains its maximum value at the rearward conveying element. To force molten polymer through the rearward conveying element, pressure is generated by the forward conveying elements pumping material forward as the rearward element is pumping it backward. When the pressure generated by the forward conveying elements overcomes the pressure generated by the rearward conveying elements, molten polymer passes through the melt seal. Resin is pumped into the kneading blocks and melting section, the force required to push material past the rearward conveying element is exceeded, and material moves through to the low-pressure area or to the large-pitch conveying element.

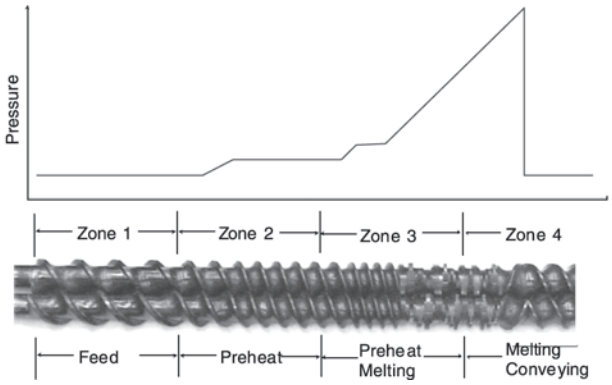

Figure 12.34. Melt pressure curve associated with a given screw design.

Figure 12.35 shows the melt pressure profile near the extruder end, where mixing, venting, and pumping to the die occur. Pressure is high in the mixing area and low in the devolatilization area. The small-pitch conveying elements at the screw end increase the melt pressure to the die. A similar chart for melt temperature along the screw has a similar profile to the melt pressure. In each high-pressure region along the screw length, the shear heating increases, resulting in higher melt temperature. In Fig. 12.34 at the end of the melting section (rearward conveying element), the polymer melt temperature is higher than the resin melting or plastication point. Kneading blocks and rearward conveying elements, if not properly designed, can cause resin degradation from thermal overheating.

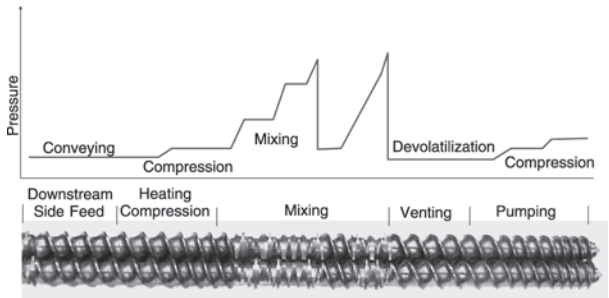

Figure 12.35. Relative pressure profile at different locations along the screw.

12.10 Residence Time

The residence time is the time each particle spends in the extruder, from when it enters the first barrel section until it exits the die. Each screw configuration is associated with a residence time distribution depending on

- Throughput rate
- Number of conveying elements
- Kneading blocks
- Rearward conveying elements
- Any other restrictive elements
- Screw speed

There is a residence time distribution curve for resin fed into the feed throat; a second distribution curve for

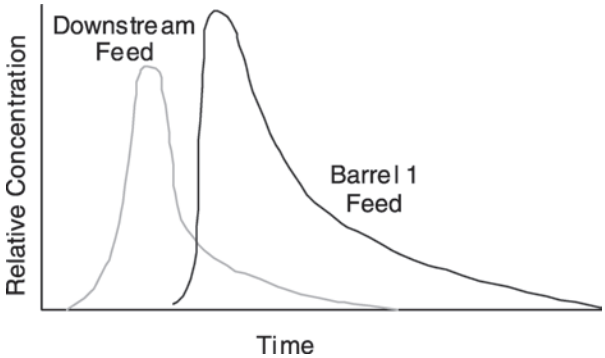

Figure 12.36. Typical residence time distribution curve.

fillers, reinforcements, additives, or polymers added downstream; and another distribution curve for liquid fed into the extruder at another port. Each ingredient fed into a different port in the extruder has its own residence time distribution curve. Figure 12.36 shows a typical residence time distribution curve for two materials fed into the extruder in different locations. Residence time in the extruder is important because it relates to the resin time temperature curve and resin degradation. A polymer can be heated for a long time at low temperature or a short time at high temperature with no degradation. However, long residence time at high temperature may cause resin degradation. Over long time periods, polymers degrade in the solid state at elevated temperatures once the thermal stabilizers are consumed. Long resident times will consume more thermal stabilizer, detracting from the product long-term thermal stability.

With a specific screw configuration and feed rate, and comparing two different screw speeds, the higher screw speed generates

- Shorter residence time
- Higher melt temperature
- Lower torque
- Potentially greater mixing
- With reinforcements, higher aspect ratio attrition

The residence time may be short or long, depending on the path through the extruder. The initial or shortest residence time (τ_i) in seconds is shown in Table 12.2 for various screw speeds for a 40 mm corotating twin screw extruder running at 100 pounds/hour.

Table 12.2. Shortest Residence Time versus RPM

Screw RPM	τ_i (sec)
75	105
125	90
175	80
225	70

This is the time it took for the first material to exit the extruder. These residence times were measured by adding black concentrate to the feed throat and monitoring the time it took to show up at the die exit. At 225 rpm the black color was first visible at 70 seconds. At 80–100 seconds it was a deep gray, and then the color started to disappear. Accurate time measurements provide a residence time distribution at a specific throughput rate, screw speed, and screw configuration. Change any one of the three variables and the material residence time in the extruder is altered.

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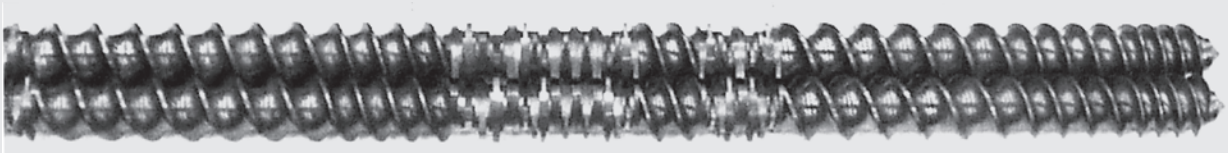
Review Questions

1. What are the eight possible processing areas in a twin screw extruder?
2. What is the difference between an undercut SKN type conveying element and a normal conveying element; what is an SKN element used for?
3. What is the difference between a gravimetric or loss-in-weight feeder and a volumetric feeder and how is each feeder set up to run 250 pounds/hour?
4. What elements are used in the feed section of a parallel, corotating, intermeshing twin screw extruder?
5. What is the difference in feed geometry between a parallel and a conical twin screw extruder?
6. What materials are added downstream in a twin screw extruder and why?
7. How is the melting zone located in a twin screw extruder?
8. What is the difference between the melting sections in parallel and conical twin screw extruders?
9. What elements are used to convey and pump molten polymer toward the die?
10. How is mixing done in a parallel, intermeshing twin screw extruder?
11. Where does mixing normally occur in a given screw configuration?
12. In a three-lobe twin screw extruder, how many lobal pools of molten polymer exist?
13. What is the difference between dispersive and distributive mixing and what types of materials are mixed with each type of mixing?
14. What screw elements are used for dispersive mixing and how do they work?
15. What screw elements are used for distributive mixing and how do they work?
16. Where are gear mixers used and why?
17. What are the different ways of feeding materials to the extruder downstream?
18. How does a side feed extruder work?
19. What screw geometry and elements are used for downstream feeding?
20. What screw geometry and elements are used for atmospheric or vacuum venting?
21. Why might a gear pump be needed on a parallel, intermeshing, corotating twin screw extruder?
22. What is the main application for corotating twin screw extruders?
23. Where are conical twin screw extruders used and why?
24. What die geometry is required to make a square rod and why?
25. Knowing the linear coefficient of expansion of PVC is 75×10^{-6} in/in/°C, what is the shrinkage of a 10-foot section of pipe cooling from 140°F to room temperature (75°F)?

Review Questions (continued)

26. What are some polymer rheology effects that alter the performance of dies?
27. What is melt fracture and what causes it?
28. What is the melt pressure and melt temperature profile in the following group of elements?



29. What is meant by mean residence time?

13 Screw Design

Screw design was touched upon briefly in Chapter 12 while discussing polymer behavior in the extruder and the different segments in screw configurations required for feeding, conveying, melting, mixing, venting or devolatilization, downstream feeding, and pumping molten polymer to the die. This chapter will discuss the particular screw elements in more detail to provide a better insight into which elements to use in particular situations.

The available barrel sections determine the screw length. Assume a particular extruder has 10 barrel sections, each barrel section L/D (length to diameter ratio) is 4:1, and the screw diameter is 60 mm. When designing the screw configuration, what is the screw element length in mm required to completely fill the screw shafts? Each barrel section with a barrel L/D of 4:1 is 240 mm long ($60 \times 4 = 240$ mm); with 10 barrel sections present, the screw length is 2400 mm. The actual screw shafts are slightly shorter than 2400 mm; with 2400 mm of screw element length, the screw tips force the elements tightly together, preventing polymer from flowing between the screw elements on the shaft. The tip is threaded into the end of the screw shaft; it is tightened, squeezing the screw elements together since the last element extends past the end of the shaft. When constructing the screw configuration (placing the elements on the shafts), verify that each element mating surface is flat with no scratches or nicks, and the shaft and inside surface of the screw elements are clean. Screw elements slide onto the screw shafts easily. Shafts and keyways have a thin coat of antisieze applied to them during screw construction to assist element removal at a later date when changing the screw design. It is advisable to lap the end of each element on a flat surface with fine emery cloth prior to assembly. Excess antisieze is avoided; otherwise it builds up between the elements, preventing a good contact seal that allows polymer to flow between the elements. If polymer flows between the elements in high-pressure regions of the screw during processing, over time the polymer degrades and chars, making element removal very difficult. Figure 13.1 shows a new screw design that is ready for assembly. The screw shafts are clean, a screw configuration or design is available to use as a guide in assembling the screws, and the appropriate screw elements are arranged in identical pairs, ready to be inserted on to the screw shafts.

Figure 13.1. Setup for assembly of new screw design.

Screw elements are available in discrete lengths to be combined in any of a number of ways to provide flexibility in locating the melting, mixing, conveying, feeding, etc., zones in unique configurations for different product formulations. Screw design configurations are the key to operating efficiency. By designing the operating steps—i.e., melting, mixing, addition of other ingredients, venting, and so forth—to optimize the melt temperature, die pressure, and throughput rates, each extruder is optimized for the processes and products being produced on that extruder.

13.1 Conveying

Available conveying elements have a distinct pitch, a defined length, and a specific conveying direction. The screw element nomenclature is pitch/length, so that a 60/60 element has a 60 mm pitch (flight makes one complete revolution around the screw element every 60 mm in length) and its length is 60 mm. A 60/30 element has a 60 mm pitch and is 30 mm long. Forward conveying elements move material down the barrel from the feed throat to the die, while rear conveying elements move material back toward the feed throat. Figure 13.2 shows forward and rear conveying elements.

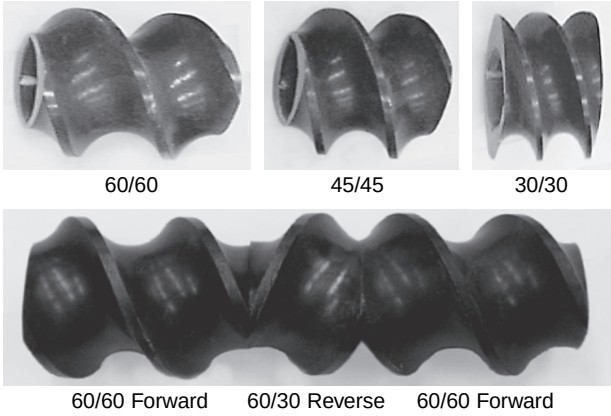


Figure 13.2. Conveying screw elements.

Pitch influences the percent fill in intermeshing extruders. A large-pitch element after a melt seal or restrictive element has a low percent fill. As the screw pitch is reduced, the percent fill increases. A schematic for this is shown in Fig. 13.3. As the pitch decreases, the mean residence time in the extruder increases, since the material must make more revolutions around the screw shaft to be conveyed

Figure 13.3. Degree of fill with different pitch conveying elements.

forward the same distance in the barrel. An example is a 60/60 element; the material makes one complete revolution while moving downstream 60 mm. Material in a 30/30 element must make two revolutions to move forward 60 mm.

Pitch is normally related to the element diameter. Conveying elements shown in Fig. 13.2 are from a 40 mm corotating, intermeshing twin screw extruder. A 60-pitch element is 1.5 times the diameter or 1.5D. Large-pitch elements normally have a 1.5D or 2D pitch. In a 90 mm twin screw extruder, large-pitch elements have either a 135 mm or 180 mm pitch. A typical large-pitch element on a 90 mm twin screw might be 180/180 or 135/90. Large-pitch elements have the lowest percent fill and the highest conveying rate. They are normally used in either the feed or venting section. Medium-pitch conveying elements are approximately 1D or 45/45 on a 45 mm extruder, 60/60 on a 60 mm extruder, or 90/90 on a 90 mm extruder. (The element length does not have to equal the pitch, as suggested in the examples above. A 90 mm extruder may be a 90/45 element, where the flights complete 1/2 revolution and the element is 45 mm long.) Medium-pitched elements have medium conveying speed with a higher percent fill compared to large-pitch conveying elements. They are used for feed or melt compression after large-pitch conveying in either a feed zone or venting area. Small-pitch conveying elements have the slowest forward conveying speed and the highest percent fill. Their pitch is normally 0.25D to 0.75D. These elements have the maximum percent fill without generating significant melt pressure and are used for pumping and heat transfer.

Nonintermeshing twin screw extruder conveying elements usually employ a square pitch, where the pitch equals the extruder diameter. A 60 mm nonintermeshing extruder square pitch screw has a 60 mm pitch. While alternative pitches are available, square pitch is normally standard.

The screw flights on each conveying element can vary from one to three. Single flighted elements are standard for nonintermeshing twin screw extruders, while most intermeshing, parallel twin screw extruders are bilobal, having two lobes or two channel flights on each element. The early intermeshing, parallel, corotating twin screw extruders had three lobes or three flights per element. Most triple flighted elements in more recent twin screw extruders are specialized kneading blocks. Single flighted elements used in two-lobed extruders are principally employed in the feed section. As mentioned in Chapter 12, the melt is divided into channels depending on how many lobes or flights are in the element design. The number of polymer channels is given by Eq. (13.1):

$$Polymer\ Channels = 2n - 1 \tag{13.1}$$

where n = number of lobes or flights on each element.

Single flighted elements have large flight widths with very good pumping efficiency and feed capacity. The shear generated in single flighted elements is low in comparison to multiflighted elements. Figure 13.4 compares intermeshing screws with single flighted elements in both corotating and counterrotating extruders.

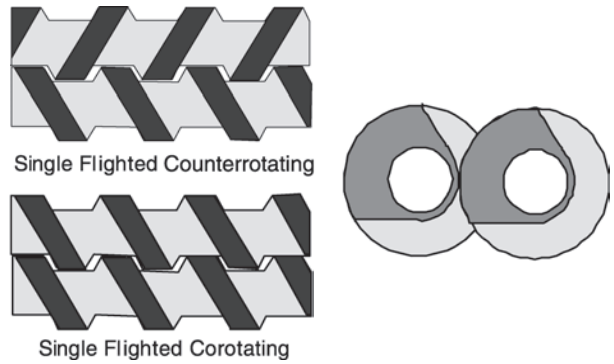


Figure 13.4. Single flighted screw elements.

Double flighted or bilobal elements have been discussed in detail in Chapters 11 and 12. They are the standard in the industry today for high torque, high throughput extruders. Double flighted elements are normally located downstream from any single flighted elements, dividing the polymer stream into two channels. Conveying elements are used principally for

- Feeding
- Conveying
- Downstream feeding
- Devolatilization
- Pumping

Triple flighted elements are still present in older twin screw extruders. They generate higher shear than either single or double flighted elements because the channel depth is shallower. These elements are used for the same applications as double flighted elements in newer machines.

13.2 Mixing Elements

Numerous mixing element designs are available with different functions based on their geometry. Kneading blocks are the most common mixing element. They are characterized by their

- Length
- Stagger angle
- Number and width of disks
- Number of flights
- Conveying direction

Kneading block nomenclature is stagger or offset angle/number of disks/kneading block length. A common 40 mm twin screw kneading block is a 45/5/60 with the first disk in the vertical direction, the second rotated 45 degrees, the third rotated another 45 degrees from the second, and so forth. There are five disks on the kneading block with an overall length of 60 mm. See Figs. 12.12, 12.13, and 12.16 in Chapter 12 for pictures of standard kneading blocks. The width of the disk can vary, depending on the mixing to be accomplished. Wide disks are used for dispersive mixing, while narrow disks are used in distributive mixing. Figure 13.5 shows two approximately equal length kneading blocks with different disk widths. The narrower disks are more open, allowing polymer to flow around the blocks both in the forward and reverse directions, resulting in the melt stream splitting and recombining numerous times. In the narrow kneading blocks, each disk has a shoulder on each side of the disk at the screw element root, resulting in a smaller width disk to work the polymer.

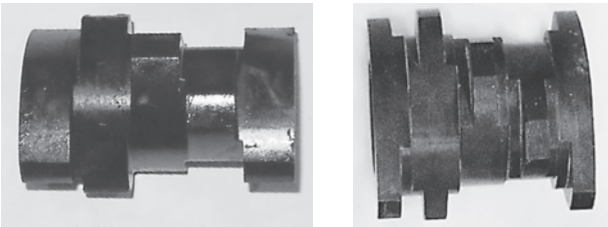


Figure 13.5. Comparison of wide (left) and narrow kneading blocks.

Kneading blocks' ability to convey melt forward in the barrel depends on the individual disk rotation on each block. Forward conveying kneading blocks convey polymer toward the die, while reverse kneading blocks convey melt back toward the feed throat. With reverse kneading disks, all polymer flows between the disk gaps as it is conveyed forward in the extruder. Using a combination of reverse kneading blocks and/or conveying elements in succession leads to significant increase in both melt pressure and melt temperature. The smaller the stagger angle, the greater the reverse flow and work put into the polymer, yielding better mixing accompanied by higher melt temperature and pressure. Reverse kneading blocks, like reverse conveying elements, are used to develop melt seals prior to downstream feeding or venting. Neutral kneading blocks convey material neither forward nor reverse. Each disk is rotated 90° from the previous disk (Fig. 12.16, Chapter 12). Long, wide, neutral kneading blocks work the polymer to a greater degree than forward conveying kneading blocks, and to a lesser degree than rearward conveying kneading blocks. Wide neutral disks tend to capture the melt pool, leading to dispersive mixing, while narrow neutral disks divide and recombine the melt, resulting in distributive mixing.

The stagger angle determines the polymer backflow that passes through the gaps between the disks. The higher the stagger angle, the greater the open area allowing backflow, with decreased forward conveying capacity. A 30/7/45 kneading block has more forward conveying capacity and less polymer flowing backward through the gaps toward the feed throat than a 60/4/45 kneading block. Neutral kneading blocks have more open area for backflow compared to forward conveying kneading blocks, since they exhibit no forward conveying. If there is no melt pressure forcing material toward the die, molten polymer in a neutral kneading disk is as likely to go forward as backward.

Three-lobe kneading blocks are used in two-lobe machines to improve mixing in a shorter length. Transition elements are required to go from two-lobe to three-lobe elements and then back to two-lobe elements after the kneading blocks.

Table 13.1 shows the relative characteristics of the different kneading blocks for mixing and conveying performance.

Table 13.1. Effect of Kneading Block Size and Type on Processing

	Narrow Forward Conveying	Narrow Reverse Conveying	Narrow Neutral Kneading Block	Wide Forward Conveying	Wide Reverse Conveying	Wide Neutral Kneading Block
Conveying Efficiency	Good	Negative	None	Poor	Negative	None
Distributive Mixing	Good	Excellent	Excellent	Poor	Okay	Okay
Dispersive Mixing	Poor	Okay	Good	Excellent	Excellent	Excellent
Shear Heating	Low	Average	Average	High	Very High	High
Pressure Generation	Average	High	High	Very High	Exceptionally High	Very High

Other specialized mixing elements are available from different twin screw manufacturers. Farrel Corporation has its patented Polygon element for distributive mixing and the continuous mixing element for dispersive mixing, shown in Chapter 12, Fig. 12.17. The Polygon is a six-flighted kneading block with a reverse pitch that has multiple screw-to-screw clearances as the screw rotates. The continuous mixing element has both forward and reverse pitch angles. This particular geometry evolved from Farrel's continuous mixer technology. Numerous companies have gear or turbine-type mixing elements, shown in Chapter 12, Fig. 12.14. While each manufacturer has slight variations on its gear mixer geometry, they are all designed to convey material forward in the barrel while dividing and recombining the melt stream to provide distributive mixing. While gear mixers are typically used for mixing liquid additives with molten polymer, they have been tried with glass fibers and other applications. Gear teeth are arranged to provide a forward, reverse, or neutral conveying. The number of gear teeth is varied to change the mixing intensity. Orifice disks or blister rings are used for dispersive mixing. These elements form a flow barrier that creates 100% fill prior to the disk. Single kneading elements, shown in Fig. 12.15, Chapter 12, and Fig. 11.42, Chapter 11, are used for dispersive mixing and reactive extrusion.[1] Leistritz lobal elements are used in both corotating and counterrotating extruders and can be forward, reverse, or neutral conveying. Lobal elements shown in Fig. 13.6 are designed to run at 100% fill.

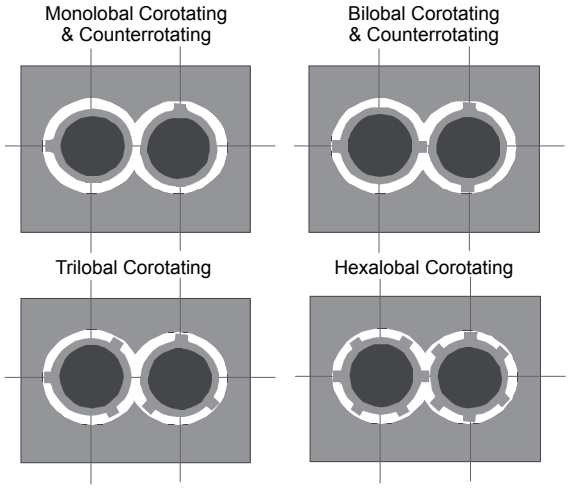


Figure 13.6. Lobal elements.[1]

In intermeshing twin screw extruders, most elements are self-wiping and self-cleaning, meaning the rotating screws continuously wipe molten resin from each other and the barrel walls, respectively, during processing to prevent polymer build-up and degradation over time. Most intermeshing twin screw extruders are built to be self-wiping and self-cleaning. Some mixing elements are nonself-wiping, such as gear mixers and blister rings. Care must be taken when using these elements if the polymer being processed is subject to thermal degradation over time.

Distrubitive mixing in nonintermeshing twin screw extruders is provided by staggered screw flights in counterrotating geometry. Optimum mixing is obtained with 50% adjacent screw flight stagger. This geometry promotes molten polymer transfer from one screw to the other screw. During transfer, the polymer melt is split by the adjacent screw flights. It recombines with the melt when transferred back to the original screw. Figure 13.7 shows the nonintermeshing staggered flights used for mixing. Dispersive mixing is obtained using a series of elements including a blister-type cylinder followed by forward and then reverse conveying elements. The reverse conveying element creates 100% fill upstream, forcing the molten polymer to be worked. Figure 13.8 shows a dispersive mixing section in a counterrotating, nonintermeshing twin screw extruder.

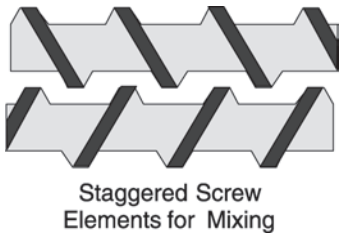


Figure 13.7. Distributive mixing nonintermeshing twin screw.

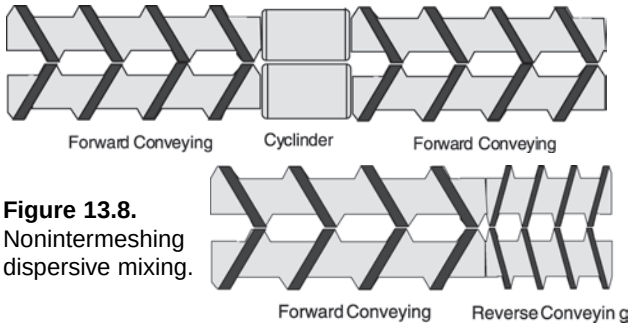


Figure 13.8. Nonintermeshing dispersive mixing.

13.3 Screw Design Applications

To construct the proper screw design, it is important to know the objectives to be accomplished with the extruder. A sheet containing the barrel sections as they are configured is combined with the screw element arrangement to design a screw geometry that melts, mixes, pumps, conveys, and so forth. As an example, assume a new screw configuration is being designed for an intermeshing, corotating extruder with eight barrel sections. A typical configuration, shown in Fig. 13.9, has

- Polymer feed in barrel section 1
- Atmospheric or vacuum vent in barrel section 7
- Side feed extruder in barrel section 5
- Venting, gravity feed, or liquid injection ports in barrel sections 2 and 4

A 10-barrel configuration is shown in Fig. 13.10 with

- Feed in barrel section 1
- Downstream feed via a side feed extruder in barrel section 6
- Atmospheric or vacuum vent in barrel section 9
- Barrel sections 2, 4, and 8 available for other feed streams or venting
- Closed barrel sections 3, 5, 7, and 10, where no material feeding or venting is possible

These two barrel configurations will be used in the following compounding examples to develop different screw designs to process specific formulations.

In example 1, two resins, A and B, are to be compounded together to produce a new resin blend, C. Both resins have similar thermal stability. Resin A is powder, and resin B is pelletized. The objective is to develop an appropriate screw design to mix and compound these ingredients. While either the 8- or 10-section barrel configuration will work, Fig. 13.11 with eight barrel sections shows a screw configuration that meets this example's objective. Both resins A and B are fed into the feed throat with two different gravimetric resin feeders to maintain the proper resin A to B ratio. The formulation is preheated and compressed in barrel sections 2 and 3 and melted in barrel section 4. The melt is conveyed downstream with additional mixing in barrel sections 5 and 6 and vented to remove any volatiles in barrel section 7. The vent can be either an atmospheric or vacuum vent, depending on the volatile concentration and how hard it is to remove. The melt is pumped to the die in barrel section 8. The left-handed kneading block and left-handed conveying elements in barrel section 4 ensure a restrictive barrier and complete melting in barrel section 4. The left-handed elements at the end of barrel section 6 provide a melt seal to improve vacuum venting at the vent in barrel section 7. Mixing elements used in sections 5 and 6 ensure both good distributive and good dispersive mixing. The screw elements used in these examples are based on a 40 mm corotating, intermeshing twin screw extruder. To scale up from a 40 mm extruder to a larger extruder, ratio the element lengths and the pitch-to-diameter ratios to the larger machine diameter. As an example, assume this is to be scaled to a 60 mm twin screw. The 60/60 conveying element in the 40 mm extruder feed section is 1.5 times the pitch and 1.5 times the length. A similar element in a 60 mm extruder is a 90/90 SK screw element. The 40/40 in the 40 mm becomes a 60/60 in the 60 mm, and the 25/25 conveying element in the 40 mm becomes approximately a 40/40 in the 60 mm. This 60 mm screw design is another design that will accomplish the objective. Any particular screw design used in a specific production or R&D environment needs to be optimized for the resins and formulations being produced.

For example 2, let us use the same resin formulation in the example above with the same barrel configuration, but with resin B more thermally stable than resin A.

Figure 13.9. Typical barrel configuration with eight barrel sections.

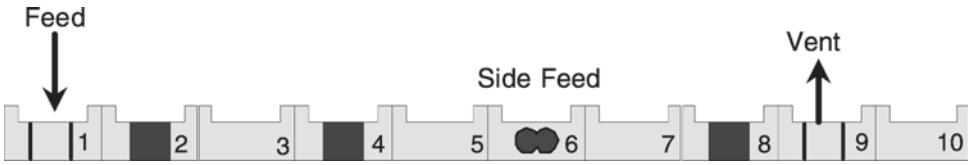


Figure 13.10. Typical barrel configuration with ten barrel sections.

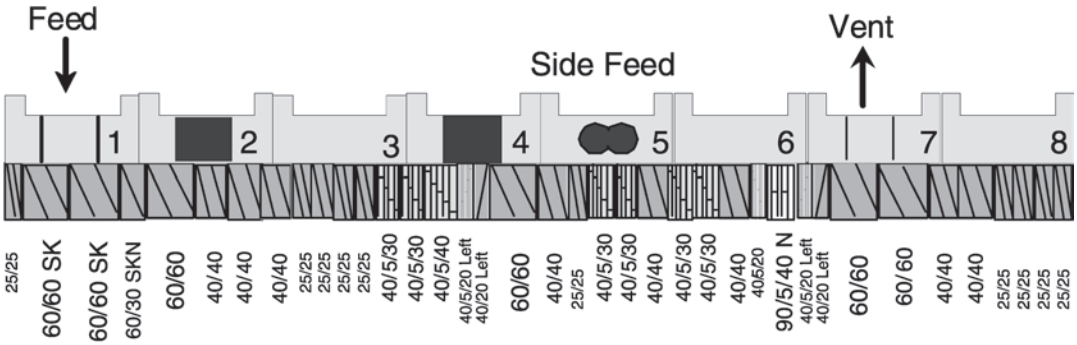


Figure 13.11. Potential screw design to mix resins A and B.

Consequently, resin A is added to the resin B melt downstream. What is an appropriate screw design to accomplish the objective in this second example? Figure 13.12 shows a screw modification to feed resin A downstream, with the side feed extruder, and provides acceptable melt mixing with resin B prior to the vent.

The third example has resins A and B added to the extruder in barrel section 1 and filler added downstream at 30% by weight. As it is filler being added and not a reinforcement that requires good dispersion, there is no concern for particle attrition. The 10-barrel extruder configuration is used for this example. Figure 13.13 shows a screw design that adds filler downstream with the side feed extruder. Melting of resins A and B, added at the feed throat in barrel section 1, is complete by barrel section 4. If the two resins require additional mixing prior to filler addition, it can be done in barrel section 5 by replacing the conveying elements with kneading blocks. Filler is added in barrel section 6 by a side feed extruder, heated and compressed in barrel 7, and mixed in barrel 8. The vent in barrel 9 removes any volatiles or gases entrapped in the melt.

In a fourth example, similar to example 3, the filler is replaced with a fiber reinforcement fed downstream through a side feed extruder. Figure 13.14 shows a screw design to process the reinforcing fiber, where the major concern is fiber attrition. The melt section is changed slightly from example three to example four by separating the kneading blocks with conveying elements. This helps to minimize the melt temperature rise by decreasing the work in the kneading blocks. Kneading blocks are not used after the side feed extruder to minimize fiber attrition. The left-handed conveying element prior to the vent was removed, decreasing the shear and fiber attrition.

In a last screw design example, use the situation in example 4 with fiber reinforcement, but add a liquid colorant to the formulation so the resulting product is a colored reinforced thermoplastic composite. The colorant is pumped into the resin melt in barrel section 4. Figure 13.15 shows an appropriate screw design to add liquid with the proper mixing. Gear mixers can be used in place of the kneading blocks in barrel section 5 to mix the liquid colorant. In all examples presented here, specialized mixing elements are intentionally avoided to gain a better understanding of what can be accomplished with conveying elements and standard kneading blocks. Melting is complete in barrel section 3, and the liquid is injected

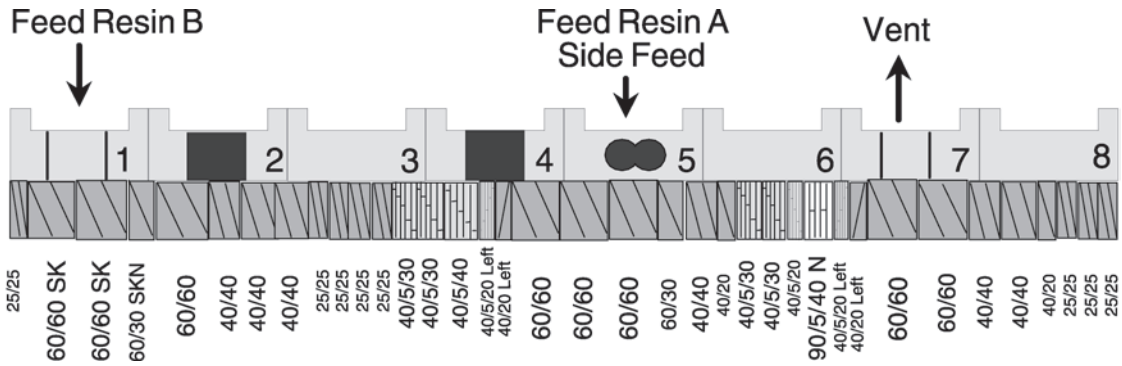


Figure 13.12. Screw configuration for Example 2—Resins A and B fed in different barrel locations.

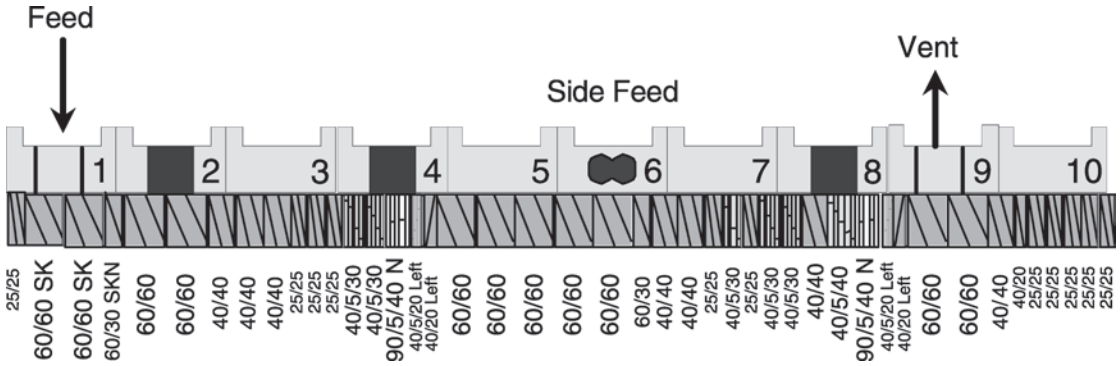


Figure 13.13. Resins A and B are fed in barrel 1 and filler in the side feed extruder.

Figure 13.14. Screw design to process fiber reinforcements with downstream feeding.

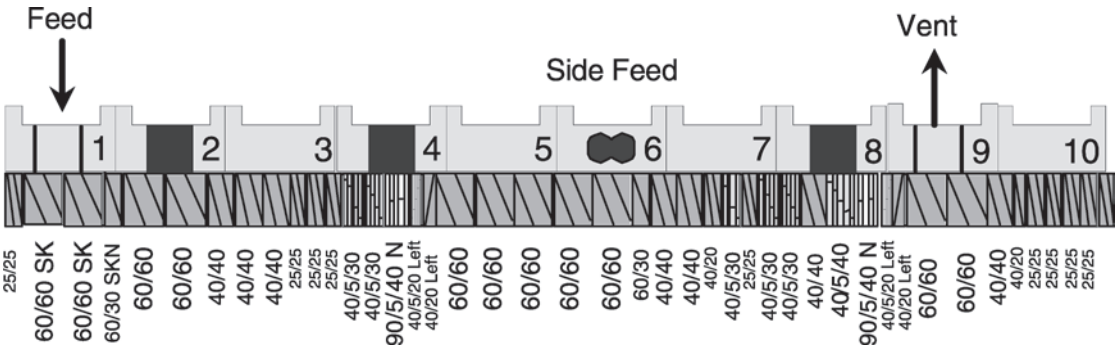


Figure 13.15. Compounding of two resins with liquid colorant and reinforcement.

into barrel section 4 and mixed with kneading blocks in barrel sections 4 and 5. While conveying elements are used where the liquid is injected, it is possible to inject liquid directly over kneading blocks or gear mixers, assuming the pressure supplied by the liquid feed pump is greater than the melt pressure generated in the extruder. Barrel sections 6 through 10, where the reinforcing fiber is fed and the extruder is vented, are essentially unchanged from example 4.

These examples demonstrate appropriate screw designs and an understanding of how to arrange the various screw elements to accomplish specific objectives. In actual practice, screw designs may have to be modified slightly for different formulation components to obtain

- Optimum throughput
- Dispersive and distributive mixing
- Correct melt temperature profile
- Optimum feed locations

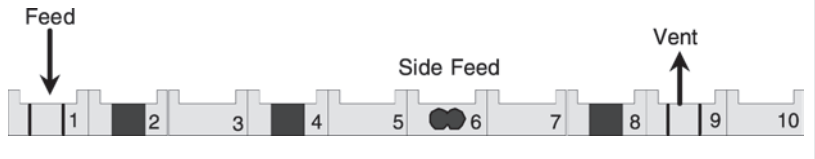
REFERENCE

[1] Thiele, William, "Reactive Compounding with Your Extruder," *Formulating & Compounding, 2(6)*, 1996.

Review Questions

1. What is the difference between conveying elements and kneading blocks?
2. What preparation is required for constructing a new screw design?
3. What are some typical large-pitch conveying screw elements for 135 mm, 57 mm, and 90 mm twin screw extruders?
4. What are some typical small-pitch conveying elements for 30 mm, 45 mm, and 60 mm twin screw extruders?
5. What are the differences between forward and reverse conveying elements and forward, reverse, and neutral kneading blocks?
6. What are the differences between single flighted, double flighted, and triple flighted conveying elements?
7. What are some mixing elements other than kneading blocks, and do they work better as distributive or dispersive mixers?
8. Design a screw to make a color concentrate (liquid colorant) on the eight-barrel extruder shown at right. The extruder is 60 mm in diameter, and each barrel section has a 4:1 L/D. Determine what elements to use based on the extruder size, and arrange them in the proper screw configuration to produce an acceptable product for the customer.

9. Develop a screw design to plasticize resin C by adding a liquid plasticizer to the extruder. The formulation calls for 75% resin C and 25% plasticizer. Use a 45 mm twin screw corotating, intermeshing extruder with the same barrel geometry shown in question 8 with a 4:1 L/D barrel ratio.
10. Develop a screw design and define the process to compound resin D with a two-component flame retardant system (both flame retardant ingredients are powders—3% S and 7% H in the final formulation), a reinforcing fiber at 22% of the final formulation, and a liquid colorant at 4.5% of the total formulation. Use the 10-barrel, 40 mm twin screw extruder shown above, assuming a 4:1 L/D barrel ratio.



14 Processing Conditions

This chapter covers extruder temperature profiles, setup, start-up, steady-state operation, and shutdown procedures.

14.1 Extruder Temperature Profile

What is the best method to set the optimum extruder temperature profile? Several options are available to select the correct temperature profile, ranging from recommendations by the raw material supplier to trial and error methods in a production environment. With products made previously, experience or documented standard operating procedures (SOPs) dictate the best processing conditions. However, the first time a new product is extruded, a dilemma can be created. Material suppliers provide recommended temperature profiles, proper drying conditions, appropriate die design and land length, and suitable draw down ratios. In addition, material suppliers provide optimum screw design for single screw extruders but do not recommend screw designs for twin screw extruders. Therefore, it is the twin screw extruder user's responsibility to develop the proper screw design for the particular application. Using the guidelines for screw development shown in Chapter 13, a suitable starting point can be developed and optimized. The product draw ratio is set based on a fixed die opening and the final dimensional requirements. With new products, the barrel and die temperature profiles and screw rpm are varied until the product quality and output rates meet acceptable standards. During process optimization, processing conditions are recorded with product quality descriptions to establish the best relationships between processing temperature profile, screw speed, and throughput rate with product quality. A design of experiments (DOE) can be employed to optimize the processing conditions. Once equilibrium processing conditions are obtained, all operating conditions, setpoints, and process observations are recorded as baseline data for future production runs.

The possible barrel temperature combinations available for operating the extruder are almost limitless. However, only a few temperature profiles will produce an acceptable product while providing optimum process conditions. When one knows where the polymer melts or softens, the mixing required after melting, and any downstream feeding and venting, then one can select temperature setpoints based on the processing requirements for a particular resin formulation. Possible temperature profiles include

- Progressive or increasing temperature, with the setpoints increasing continually from the feed throat to the die
- An inverted or decreasing temperature profile, with the setpoints decreasing from the first heated barrel to the die
- A straight temperature profile, where all barrels are set at exactly the same temperature setpoint
- A humped profile, where the temperature is lowest in the first heated barrel, gradually increases toward the middle of the extruder, and then decreases progressively toward the die
- Some combination of these profiles

The temperature profile that works best with a given product and screw design depends on

- Resin type
- Resin viscosity
- Percent mixing
- Downstream feeding
- Venting
- Throughput rate

Polymer melt temperature is a critical property in controlling the extrusion process and optimizing the throughput while minimizing resin degradation.

Barrel section 1 is normally water cooled with no heating available. This provides uniform resin flow, preventing premature melting or polymer sticking to the feed hopper or the feed throat opening. It also prevents premature melting in barrel section 1, where a lubricating melt film might lead to feed problems later in the run. Barrel temperature in zone 2 is normally set 20–30°F (11–17°C) below the melting point of semicrystalline polymers or 125°F (70°C) above the T_g of amorphous polymers. In barrel section 2, the polymer is compressed and preheated as it moves forward to barrel section 3. Temperatures are raised progressively in barrel sections 3 and 4 to the desired polymer melt temperature. Temperature guidelines for barrel zone 3 are 20°F (11°C) below the desired melt temperature for either semicrystalline or amorphous resins. Set barrel 4 temperature at the desired melt temperature. Depending on the screw design, a resin normally is melted in barrel 4. The remaining barrels are used to maintain the melt temperature or cool, depending on the functions occurring in each barrel section. If excessive work is being done in a particular barrel through either dispersive and/or distributive mixing, heat is generated by shear, and the resin may have to be cooled by removing heat in that barrel section plus the downstream barrel sections. This helps prevent resin degradation. Heat in barrel sections just prior to, during, and after downstream feed addition may be raised to bring the melt temperature back to the desired level. Adding cold material (original resin plus additional resins, additives, fillers, or reinforcements fed downstream) lowers the average extru-

sion temperature, requiring heat to be added to maintain the desired melt temperature. The melt temperature is held constant during venting to maintain the volatility of gases being removed. The final barrel sections are used to optimize the melt temperature for good flow in the die. The last few barrel zones can also be used to provide the desired melt strength, and not just cool the resin. Depending on the downstream process, melt strength can be a critical parameter. Melt strength typically increases with lower melt viscosity, which decreases with lower melt temperatures.

Table 14.1 provides "standard" processing conditions for various commercial resins. The actual processing conditions may vary from those recommended, depending on

- Resin modification
  - — Flame retardant additives
  - — Impact modifiers
  - — Fillers and reinforcements
  - — UV stabilizers, and so forth
- Extruder L/D
- Die design
- Screw design, and so forth

Table 14.1. Recommended Processing Temperatures

Material	Feed Zone, °F (°C)	Transition Zone, °F (°C)	Metering Zone, °F (°C)	Die Zone, °F (°C)
ABS	400° (204°)	425° (219°)	440° (227°)	460° (238°)
Nylon 6	420° (216°)	460° (238°)	480° (249°)	500° (260°)
Nylon 6,6	530° (277°)	535° (280°)	545° (285°)	540° (282°)
LDPE	340° (171°)	355° (180°)	365° (185°)	375° (191°)
LLDPE	300° (149°)	325° (163°)	364° (185°)	410° (210°)
HDPE	340° (171°)	380° (193°)	400° (204°)	400° (204°)
PP	375° (190°)	410° (210°)	430° (221°)	430° (221°)
Polystyrene	350° (177°)	400° (204°)	440° (227°)	450° (232°)
HIPS	375° (191°)	420° (216°)	450° (232°)	450° (232°)
PMMA	360° (182°)	400° (204°)	430° (221°)	445° (230°)
Flexible PVC	265° (130°)	340° (171°)	355° (181°)	365° (181°)
Rigid PVC	300° (149°)	320° (160°)	340° (171°)	365° (181°)
PC	510° (266°)	530° (277°)	550° (288°)	560° (293°)
Noryl®	450° (232°)	480° (249°)	510° (266°)	510° (266°)
Ultem®	600° (316°)	640° (338°)	675° (357°)	675° (357°)
PET	520° (270°)	550° (290°)	510° (265°)	510° (265°)
PBT	470° (243°)	490° (121°)	500° (260°)	500° (260°)
Polysulfone	550° (288°)	600° (316°)	650° (343°)	650° (343°)
Acetal	400° (204°)	390° (199°)	400° (204°)	410° (210°)
Thermoplastic Polyurethane	330° (166°)	360° (182°)	380° (193°)	380° (193°)

14.2 Screw Speed

Screw speed in a starve-fed extruder is a critical process parameter in controlling

- Mixing
- Melting
- Pressure generation
- Melt temperature

Screw speed determines

- The percent fill of the various screw flights
- The residence time in the extruder
- The torque level

The screw is run at sufficient rpm to generate enough power to prevent the extruder from exceeding its torque limit and shutting down. During start-up, the screw is run at high rpm until the desired throughput is reached. This prevents the extruder from generating too much torque and shutting down prematurely, while feed rates are being increased. At a given throughput, the screw motor is normally run at approximately 90% torque. The effect of higher screw speed is to

- Lower the torque
- Put more work into the polymer matrix
- Raise the melt temperature
- Provide greater mixing
- Increase resin degradation due to thermal energy

In addition to lower torque, higher rpm reduces the residence time and the percent screw fill. Lower rpm at the same throughput leads to

- Higher torque
- Longer residence time in the extruder
- Higher screw fill

In a steady-state process with uniform feeding and no torque spikes attributable to raw material feed, torque can be run at 95% or higher. However, if a feed spike or other abnormality occurs, the motor may exceed its torque limit and shut down. Monitoring the torque limit at equilibrium for 30 minutes to an hour will indicate if the screw speed can be safely run at a lower speed and higher torque or at the same speed and higher rates. One reason for decreasing the screw speed and increasing the processing torque is to minimize the melt temperature resulting from the higher shear at higher speeds.

Processing is always a balance of processing conditions, where

- Throughput
- Screw speed
- Melt temperature
- Residence time
- Mixing
- Screw design

are optimized to produce the highest quality product. Higher screw speed may generate so much shear heat that the melt temperature cannot be controlled. In this event the throughput is reduced, allowing the screw speed to be reduced, lowering the melt temperature. Ultimately, a different screw design may be required to optimize the process for screw speed, throughput, torque, and melt temperature. Remember, in starve-fed extrusion the throughput is determined by all the feed rates and not the screw speed. Screw speed determines the residence time in the extruder and the percent screw fill. If feed is being added downstream, do not turn the screw speed down to optimize the torque until all the feed streams are running at their proper rate. Assuming the screw is not running at its maximum speed and mixing is borderline, increasing the screw speed will improve the mixing until the screw design can be optimized for better mixing. If reinforcements are being added downstream, higher rpm increases the fiber attrition, possibly resulting in a lower property profile.

14.3 Process Variables

Twin screw extruders are characterized by their response to various process variables, such as

- Throughput
- Screw speed
- Barrel temperatures
- Head pressures
- Screw design
- Polymer viscosity

This section looks at the effect of the independent variables on some response variable, such as

- Torque
- Specific energy
- Melt pressure
- Melt temperature
- Viscosity changes
- Throughput rates

Increasing the throughput rate results in a higher percent screw fill, a decrease in residence time, and an increase in molten polymer backup length at melt seal locations along the screw. Higher throughput requires more torque, while generating higher head pressure; normally the melt temperature and specific energy (kilowatt-hour/pound or kilogram) decrease.

Increasing the screw speed reduces the percent screw fill, the residence time in the extruder, and the molten polymer backed up prior to a melt seal. Higher screw speed results in higher specific energy and melt temperature, with lower torque and head pressure.

Higher barrel temperatures decrease the resin melt viscosity. Higher barrel temperatures lead to lower torque, specific energy, and head pressure, accompanied by higher melt temperature.

As head pressure goes up, the molten polymer backup length in the extruder just prior to the die increases. At very high head pressures with atmospheric or vacuum vent in either the last or next to last barrel section, polymer can be forced out the vent port. As head pressure increases, the torque, specific energy, and melt temperature all increase, while the throughput rate decreases.

Screw design changes that increase the shear in the extruder result in higher specific energy, a greater percent screw fill, and a higher residence time. Other process responses affected by increased shear include higher torque and melt temperature, with lower throughput and head pressure.

To optimize the process, all independent process variables must be considered in conjunction with the polymer and what is occurring in the different extruder sections. Balancing screw speed, throughput, melt temperature, screw design, and barrel temperatures to obtain both the correct and uniform melt temperature and pressure is critical in developing an optimum, cost-effective process producing a quality product.

14.4 Extruder or Production Run Setup

Setting up a new production run or changing from product 1 to product 2 requires all transfer lines, feed hoppers, and downstream equipment to be clean to prevent product 2 cross-contamination with product 1. If the same dryer is used for more than one product, the dryer and vacuum transfer lines must be completely cleaned to prevent cross-contamination. The extruder is run empty and possibly purged with appropriate purge material, depending on the next product to be processed. It may be necessary to pull the screw and change the screw design if product 2 is totally different from product 1. Changing the screw design to optimize throughput and product quality is advantageous if product 2 is going to be run for an extended time, or the products scheduled after product 2 will process better with a different screw design.

Check raw material availability and verify all formulation ingredients are at room temperature or in the dryer for the appropriate time and temperature to yield the correct moisture content. If raw materials are stored in either a cold or hot warehouse, the raw materials should be brought into the processing plant and allowed to come to equilibrium temperature at least 24 hours prior to starting the run. Plastics are great insulators; consequently, gaylords or other large containers require a long time to come to temperature equilibrium. Adding either hot or cold resin or additives to the extruder downstream can alter the processing parameters so the product is not the same as last time.

Before starting any production, compare the SOP with previous run records to determine how the product ran last time. This helps the operator to anticipate whether any changes may be required at start-up. If the SOP has been changed and/or reissued since the previous run, ask questions to understand what necessitated the changes. Check all setpoints versus the SOP, and verify that all equipment (heaters, thermocouples, temperature controllers, cooling on the feed throat and barrel, vacuum, etc.) is functioning properly.

Before starting a new run, check and correct any safety concerns on or around the extruder, e.g., bare wires or loose connections on the die, housekeeping around the extruder, and proper insulation on the die.

If the extruder has been down for an extended period or shut off completely, it is necessary to heat soak the extruder to bring all temperature zones and dies up to their setpoints before starting. The time required to reach equilibrium temperature depends on the extruder size. If the die comes up to temperature much more slowly than the extruder barrel temperatures, turn the die heater power on prior to the rest of the extruder. An hour is normally enough time to bring the extruder barrel up to temperature; additional heat soak time may result in any resin left in the extruder barrel generating black specks at start-up due to degradation. Black specks generated by resin degradation need to be completely purged from the system before collecting first-grade product.

If either volumetric or gravimetric feeders are used to feed resin and/or additives, they may require calibration or resetting prior to the next run. Gravimetric or loss-in-weight feeders are easily set by inserting the correct pounds/hour (kilograms/hour) required for each feed stream in the formulation. Volumetric feeders may need to be recalibrated for the formulation ingredient being fed by that particular feeder. The rate is manually checked to verify that each feeder is delivering the correct pounds/hour (kilograms/hour).

Downstream equipment is checked for safety, proper operation, and maintenance prior to starting. Speeds are set, slitting or trim knives changed, rollers cleaned, nip roll gaps set, roll or cooling tank temperatures adjusted, etc.

14.5 Start-Up

Prior to starting the extruder, retorque the die bolts after the system is up to temperature. Start the extruder at 20 to 50 rpm and gradually add resin to the feed throat until the torque reaches 50 to 60% and molten resin is flowing freely from the die. During this time, monitor the torque and die pressure. If there is a solid polymer plug in the die or transfer pipe or some other die blockage, the pressure will increase very rapidly, causing a dangerous situation. In the event a pressure spike occurs, immediately turn off the screw and determine the cause before proceeding. If the pressure exceeds the preset cut-off limit, the extruder will automatically shut down. ***Do not walk in front of the die during start-up or until material is flowing freely from the die and no more air is being forced out of the extruder die.*** Once molten polymer exits the die, the extruder screw speed is increased, followed by increased feed rates. Feed is added slowly to prevent the extruder from exceeding the high-torque limit, which will automatically shut down the extruder.

Regardless of the extrusion process, the downstream equipment is normally strung up at low rates and takeoff speeds. After the downstream equipment is properly strung up and product is running satisfactorily, increase the feed rate (while monitoring the extruder torque), then the takeoff speed, until the desired settings are reached. Bringing the equipment up to speed too fast or using the wrong puller speeds can cause the molten polymer web to be easily broken, requiring the process to be restarted. The amount one can pull or draw on the molten web depends on the polymer melt strength. Extrudate with high polymer melt strength can be easily handled, while low melt strength makes stringing up the downstream equipment a special challenge. Elastomeric materials and impact-modified products exhibit very good melt strength and are easy to handle, whereas crystalline materials with a high melt flow index (MFI) and highly filled materials may have poor melt strength and break easily.

After running a few minutes, collect product samples and determine if the product meets specifications. Depending on the extruder size, it takes a while for the entire system to equilibrate. Continue to collect product samples and monitor the product and process while the extruder comes to equilibrium, making any minor die adjustments required to bring the product into dimensional specification. Changes in extruder feed rate or puller speed are required if gross product dimensional changes are required.

Once equilibrium is attained, record all operating conditions, such as set and actual temperatures, melt pressure, melt temperature, extruder load, screw rpm, feed rates, dryer conditions, takeoff speeds, takeoff equipment temperature, pressures associated with takeoff equipment, vacuum levels, product quality, and product dimensions.

14.6 Steady-State Operation

Steady state, or equilibrium condition, depends on the extruder size. Small extruders (40 mm) may take 20 to 40 minutes to reach equilibrium, while large extruders (135 mm) may require one or two hours. Equilibrium time is dependent on

- The metal mass being heated (barrels and screws)
- The resin processed (resin is cold and takes heat away from the extruder)
- The shear heat generated by the motor and screw

Changes in process conditions should never be done before the equipment reaches equilibrium. After changes are made, the equipment must be allowed to reach equilibrium again before the effect can be determined. If an extruder is running properly at start-up, producing a quality product, and 45 minutes to an hour into the run the product is running poorly and no process changes have occurred, the product does not run well under equilibrium conditions but does run well at the actual start-up temperatures.

New products being scaled up from R&D or from a small to large production extruder may require process changes to establish the standard operating conditions that produce the most robust product. If temperature, screw speed, or throughput changes are required, make a significant change to verify that the change did or did not have an effect. For example, changing zone 2 temperature by 2°F (1°C) is not large enough to demonstrate any effect on the product. Changing zone temperatures 15–25°F (8–14°C) is necessary to determine if the change has a significant effect on the product variable in question. Allow the extruder time to reach equilibrium before passing judgment on the effect the change has on the variable property. Effects of screw speed changes are not instantaneous. The extruder barrel temperatures have to reach equilibrium conditions resulting from viscous heat generated or removed by the screw speed change. Due to the metal mass in the barrel, heating and cooling systems require time to compensate for temperature changes resulting from screw speed changes.

Any process change that does not show an effect on the product can be returned to its original settings before another change is made. After a process change, allow time for the extruder to come to equilibrium. Change only one variable at a time unless a DOE study is being run. **Do not be known as a knob turner** when trying to solve extrusion problems. With quick changes, the extruder never has an opportunity to come to equilibrium, and optimum process conditions are hard if not impossible to define.

14.7 Shutdown and Product Changes

Product change, maintenance, or weekend work stoppage requires shutting down the extruder. Sometimes the operator is required to operate a different line. The shutdown or changeover procedure used depends on several factors:

- How long will the system be down?
- Is the product change from a natural or lighter color resin to a darker color?
- Is the product change from a darker to a lighter color or natural resin?
- Is the product change to a completely different resin system?
  - — Does the new formulation process at a higher or lower melt temperature?
  - — Is the new resin formulation compatible with the previous resin?
  - — Is the new resin melt viscosity higher or lower than the old resin at a given melt temperature?

The first question, "How long will the extruder be down?" dictates the course of action and procedures to follow. If rigid PVC is being processed on a counterrotating extruder (parallel or conical) and the extruder needs to be shut down for more than a few minutes, PVC is purged out of the extruder to prevent degradation. PVC undergoes an autocatalytic degradation reaction; once degradation starts, it becomes progressively worse. Black specks and degraded material are generated very rapidly. With other resin systems, if the extruder is going to be down for more than 10 minutes, turn the feed off, run the extruder dry, and turn off the screw.

If the line is going to be shut down overnight and restarted in the morning, thermally stable resins are run out of the barrel and the extruder temperatures turned off or set at a lower level, then brought back up to temperature prior to the desired start-up time. If the resin is thermally unstable, run the resin out the barrel and die and follow with a thermally stable purge. After purging for a short time, run the barrel and die dry. Turn the extruder temperatures off or lower them, depending on the time the extruder will be sitting idle. Restart the extruder, bring all zones up to temperature, and run virgin resin through to remove all purge and any degraded resin before starting the process.

Product changes requiring color change are straightforward when going from a lighter to darker color. Run the feeders empty, add new material to the feeder, and continue running. Discard any transition material until the new color is obtained. If this is done properly, the screw speed may never have to be slowed down. If the change can't be made at high speed, the change is easily accomplished by slowing the feed rate during the transition from product 1 to product 2, slowing the downstream equipment, and discarding any transition material. Depending on the transfer system delivering the new, darker color resin to the feed hopper, cleaning may be required to ensure there is no

cross-contamination of the darker product. Once all the lighter color formulation is removed from the system, collect the new product. If the product change is going from a darker to lighter color formulation, the product change requires more time and possibly a purge step. Run the feeders and extruder empty, purge with a low-cost material or regrind to remove the darker resin, run the purge out, and add the new lighter color resin to the feeders and extruder. With the entire purge flushed out, restring the line and start collecting new lighter color product.

When changing to a product requiring an entirely different resin matrix, the conversion procedure depends on the old and new resin characteristics.

- Does the conversion require higher melt temperatures?
- Are the original and the new resins chemically or molecularly compatible?
- What is the viscosity of each resin at the time of product changeover?
- Is a different screw design required?

Figure 14.1 shows the possible resin scenarios (each situation is numbered) when changing from resin A to resin B. The rest of this section describes procedures to follow when changes to different resin systems are required.

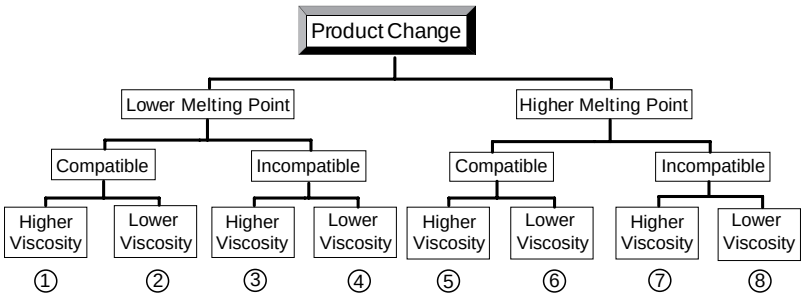


Figure 14.1. Various product changeover scenarios.

14.7.1 Disassembling Screw Elements

Product changes requiring a different screw design require special attention. To remove the extruder screws, stop all feed streams and purge with an acceptable resin to remove polymer from the barrel. During the purge, vary the screw speed over a wide range to clean the various screw elements and barrel as much as possible before removing the screws. Stop the screws and remove the die. With the die heat still on, clean the die channels of all resin and purge using a brass wire brush and scrapers. A small air-driven drill equipped with a brass wire brush wrapped in copper gauze does an excellent job of cleaning the die and transition area. Be careful not to damage any melt temperature transducers sticking down into the melt stream. Once the die is clean and removed, purge the extruder without the die. Run the extruder dry and use the following procedure to clean the screws and barrel:

1. Stop the extruder.
2. Leave all barrel temperatures at their setpoints until the screws are removed and the barrel is clean.
3. Disconnect and remove the vacuum port.
4. Remove the feeders from over the feed hoppers and clean both the feed hopper and feeders.
5. Remove the screw tips with the special tool provided for tip removal.
6. Disconnect the screw coupling between the screws and the gear box.
7. Slowly pull the screws from the extruder using the screw puller.
8. As the screws are being removed, use a brass wire brush or brass wire wheel connected to a drill motor to remove any excess resin.
9. After 5 to 10 screw elements are outside the barrel, stop pulling the screw and slide the elements off the shaft. Place the elements in a location where they can be cleaned later. The best procedure is to remove as much resin as possible when the screw elements are hot, using a wire brush. The elements can be placed in an oven, pyrolysis unit, or fluidized bed later to remove all resin and totally clean the elements.
10. Once the first 5 to 10 elements on each shaft are removed, pull out another 5 to 10 screw elements. Follow the same procedure to wire brush the elements and remove them from the shaft.
11. While removing the elements from each shaft, it is important to keep all molten resin off the screw shaft.
12. Continue the procedure for removing elements a few at a time until all screw elements have been disassembled from the shaft.
13. Using a barrel brush wrapped in copper gauze and attached to a drill motor, ream the barrels out numerous times.
14. Blow out any copper left in the barrel or feed hoppers.
15. Turn the barrel heaters off.
16. Clean up the area and put the tools away.

In the event elements are hard to remove from the shaft, it may be necessary to pull the shafts out of the extruder without first removing the screw elements. For this scenario, follow the procedure below:

1. While pulling the screw shafts out of the extruder with the screw elements in place, wire brush the elements to remove as much resin as possible.
2. Once the shafts are removed, place each screw in a wooden or brass cradle and continue to wire

brush the elements to remove as much resin as possible while the resin is still molten.

3. After the excess resin is removed, insert one screw shaft back into the extruder barrel to keep it hot while working on the other one.
4. Using a block of oak or maple, place the wood block against one end of the screw element and knock the screw element off the shaft with a hammer. Be careful not to hit the element with the hammer or damage the element; screw elements are expensive to replace. Continue to remove one screw element at a time until they become too hard to remove.
5. When no more screw elements can be removed from the shaft, take the heated screw shaft out of the extruder barrel and replace it with the shaft that was being disassembled.
6. Using the same procedure, remove the elements from the second shaft until they become hard to remove, and then reheat the shaft.
7. Alternate between the two shafts, storing them in the barrel to maintain temperature, until all screw elements have been removed.

In preparation for future production runs, the disassembled screw elements are cleaned and prepared for future use. Follow the procedures below to clean the screw elements once they are removed from the screw shaft:

1. One option is to wire brush all elements until they are clean; this is very time consuming and tedious.
2. The inside surface of the screw element is as critical to clean as the outside surface.
3. The best option is to heat all elements in a furnace to about 930°F (500°C) for one to three hours. Other ways to remove resin are with pyrolysis units or fluidized beds.
4. After the elements are cooled to room temperature, wire brush them to remove any char. (Remember the inside surface must also be clean).
5. If the elements are to be stored, spray a light coat of mold protectant on the surface to prevent rust.

Prior to reassembling the screw, each element matting surface is lapped on a smooth surface using fine emery cloth. The outside surface of the shaft is cleaned with emery cloth to remove any resin degradation. If a splined shaft is being used, it is critical to remove all resin from between the splines on both the shaft and the inside diameter of the screw bushing. Once everything is clean and the elements are lapped, the screw is ready for reassembly in another configuration.

14.7.2 Scenario #1

The first scenario in Fig. 14.1 is a product conversion to product B, which has a ***lower*** melting point than product A, is ***compatible*** with A, and has a ***higher*** viscosity than product A. A procedure to follow in changing the product is—

- Run the feeder empty.
- Add formulation B to the feeder.
- Lower the barrel and die temperatures.
- Run the feeder, screw speed, and takeoff equipment at slower speed until the barrel temperatures and melt temperature stabilize at their new setpoints.
- Once the system reaches equilibrium, ramp the extruder screw speed back to its operating range, gradually increase the feed, then takeoff equipment speeds, and start collecting the new product.

14.7.3 Scenario #2

The second scenario in Fig. 14.1 is product conversion to product B, which has a ***lower*** melting point than product A, is ***compatible*** with A, and has ***lower*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Run the feeder empty.
- Add new resin to the feeder.
- Lower barrel and die temperatures.
- Run the feeder, extruder screw speed, and takeoff equipment at slower speed until the temperatures stabilize at their new setpoints.
- Depending on the melt strength of product B, the process may continue to run without a web break. If the extrudate web breaks, run the transition product to the floor until material A is purged from the extruder.
- Once all of formulation A is purged and the temperatures have reached their setpoints, restring the takeoff equipment.
- Increase the extruder screw speed, feed rate, and takeoff speeds, and collect the product.

14.7.4 Scenario #3

The third scenario in Fig. 14.1 is product conversion to product B, which has a ***lower*** melting point than product A, is ***incompatible*** with A, and has a ***higher*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Run the feeder empty.
- Understand the degree and reason for resin incompatibility between products A and B:
  - — Cannot be mixed because the resins react chemically, generating noxious gases.
  - — Cannot be added to the extruder because the barrel and die temperatures are too high and resin B will degrade.

— Not compatible because of molecular polarity. Polar resins (resins containing dipole moment) tend to be compatible with polar resins, and nonpolar resins tend to be compatible with nonpolar resins, while polar and nonpolar resin mixtures are incompatible.

- When resins can't be mixed because of chemical reaction or thermal decomposition—
  — Purge resin A with an appropriate purge material.
  — Lower barrel and die temperatures.
  — At the proper extrusion temperature for resin B, add the formulation to the feeder.
  — Start up at low screw rpm and low feed rates until the resin is flowing freely out the die.
  — Restring the process at low speed.
  — Increase the screw speed, followed by the feeder rate and takeoff equipment; finally, start collecting product B.
- With incompatible resins due to the combination of polar and nonpolar polymers—
  — Run formulation A out of the feeder and extruder.
  — Lower barrel and die temperatures.
  — Start feeding resin B to push resin A out of the extruder. Initially run at high screw rpm but relatively low B feed rate until resin A is flushed from the extruder and die. When the system is completely purged, lower the extruder speed until all barrel and die temperatures reach their setpoints.
  — At the proper extrusion temperatures, restring the process at low speed.
  — Increase all speeds and feed rates and start collecting the product.

14.7.5 Scenario #4

The fourth scenario in Fig. 14.1 is product conversion to product B, which has a ***lower*** melting point than product A, is ***incompatible*** with A, and has a ***higher*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Run the feed hopper empty.
- For resins that can't be mixed because of chemical reaction or degradation due to temperature—
  — Purge resin A with an appropriate purge material.
  — Lower barrel and die temperatures.
  — At the proper extrusion temperature, add resin B to the feed hopper.
  — Start extruder up at low rpm and low B feed rate until extrudate is flowing freely from the die.
  — Restring the process at low speed.
  — Increase extruder screw speed, feed rates, and takeoff speeds, and start collecting the product.
- For resins that are incompatible because of molecular polarity—
  — Run out or purge resin A.
  — Lower barrel temperatures.
  — Start feeding resin B to the system to force the purge of resin A from the extruder. Due to the lower viscosity of resin B, it may be difficult to force the purge of resin A from the extruder. Two extreme alternatives to remove resin A are to pull and clean the screw, barrel, and die, or to find a third resin that is compatible with both resins A and B and use this resin as a purge.
  — Initially run the screw speed at high rpm until all of resin A is removed from the extruder.
  — After resin A removal, lower the screw rpm and feed product B at a low rate while running the process at slow speed, until the die and barrel temperatures are stabilized.
  — Restring the process at low speeds.
  — Increase screw speed, feed rate, and takeoff speeds, and start collecting product B.

14.7.6 Scenario #5

The fifth scenario in Fig. 14.1 is product conversion to product B, which has a ***higher*** melting point than product A, is ***compatible*** with A, and has a ***higher*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Increase barrel and die temperatures just prior to the end of the current production run, before the feeder is completely empty.
- With resin A present, slow the feed rate, extruder screw speed, and takeoff speeds.
- As the new temperature setpoints stabilize, add resin B to the feeder and increase speeds.
- When resin A is completely purged from the extruder and all speeds are properly set, start to collect product B.

14.7.7 Scenario #6

The sixth scenario in Fig. 14.1 is product conversion to product B, which has a ***higher*** melting point than product A, is ***compatible*** with A, and has a ***lower*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Increase barrel and die temperatures just prior to the end of the current production run.
- Run the feeder empty.

- The procedure to use depends on the magnitude of temperature increase required. With a 50°F (28°C) increase, use the following procedure:
  - — Add resin B to the feeder and continue to run at a high rate through the entire line.
  - — Once all of material A has exited the extruder, slow the feed rate, extruder screw speed, and takeoff equipment.
  - — When the temperatures stabilize at the new setpoints, increase extruder screw speed, feed rate, and takeoff speeds.
  - — Start to collect the product.
- If the temperature has to be raised between 50° and 200°F (28°–111°C), use the following procedure:
  - — Run the feeder empty.
  - — Raise the setpoint temperatures 50°–100°F (28°–56°C) and start feeding resin B to the extruder.
  - — Try to run the feed rate very slowly with the extruder screw speed relatively high. If both the drive load and head pressures are low, increase the feed rate slowly to purge resin A from the extruder.
  - — Lower the feed rate and continue to run as the temperatures stabilize at the new setpoints.
  - — Restring the line.
  - — Bring the screw speed, feed rate, and takeoff speeds up to their normal operating rates.

14.7.8 Scenario #7

The seventh scenario in Fig. 14.1 is product conversion to product B, which has a ***higher*** melting point than product A, is ***incompatible*** with A, and has a ***higher*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Run feeder empty.
- If the resins cannot be mixed because of chemical decomposition or degradation—
  - — Purge material A from the extruder and die.
  - — Raise the barrel and die temperatures.
  - — Once the temperature reaches the setpoint, add resin B to the extruder.
  - — Start up at low screw speed, feeding B very slowly to the extruder.
  - — Restring the process at low speeds.
  - — Increase extruder screw speed, followed by feed rate and takeoff speeds, and collect the product.
- If resins are incompatible because of molecular polarity—
  - — Run resin A from the extruder and die.
  - — Raise the barrel and die temperatures.
  - — Start feeding the new resin very slowly to the extruder at high screw speed to force resin A out. Initially run the extruder at high rpm and verify that the drive load is not too high and polymer is not freezing in the die, causing excessively high head pressures.
  - — Verify that resin A is completely out of the system.
  - — Shut the extruder down or run at low feed rates with high screw rpm until all temperatures have stabilized at their new setpoints.
  - — Restring the process if necessary at low extruder and takeoff speeds.
  - — Increase screw speed, followed by the feed rate, and start to collect the product.

14.7.9 Scenario #8

The eighth scenario in Fig. 14.1 is product conversion to product B, which has a ***higher*** melting point than product A, is ***incompatible*** with A, and has a ***lower*** viscosity than resin A. A procedure to follow in changing from product A to B is—

- Run the feeder empty.
- If the resins cannot be mixed due to chemical decomposition or degradation—
  - — Purge formulation A from the extruder and die using appropriate purge material.
  - — Raise the barrel and die temperatures.
  - — Add resin B to the extruder at lower temperatures than the setpoints so the viscosity is sufficient to flush the purge material from the extruder.
  - — Allow the extruder and die temperatures to stabilize at the setpoint temperatures.
  - — Start up the system at low extruder screw speed and feed rates and restring the process at low speeds.
  - — Increase extruder screw speed, feed rates, and takeoff speeds, and collect the product.
- If resin is incompatible due to molecular polarity—
  - — Run the feeder, extruder, and die empty.
  - — Raise the barrel temperatures.
  - — Start feeding resin B to push out the original formulation. Initially, run at high rpm and low feed rates, while monitoring the extruder load and die pressure very carefully. Resin B may freeze off or clog the die. Once material is flow-

ing freely from the extruder and the drive load or head pressure is not excessive, increase the screw speed and feed rate until formulation A is completely flushed from the extruder.

— Turn the screw off or run at low rpm and low feed rate until the extruder and die temperatures reach their setpoints.

— Restring the process at low speed.

— Increase screw speed, feed rate, and takeoff speeds, and start to collect the product.

Review Questions

1. How is the extruder barrel temperature profile set? What are some options?
2. What is the relationship of screw speed to torque and throughput rates?
3. In setting up an extruder with a 1000 rpm maximum screw speed, where should the speed be set?
4. What determines the throughput rate in a starve-fed extruder and why?
5. What is the relationship of increased screw speed to torque, melt temperature, head pressure, residence time, and specific energy? Explain in each case.
6. How do the different process variables (barrel temperature, screw speed, throughput rate, head pressure, and screw design) affect melt temperature?
7. How do the different process variables (barrel temperature, screw speed, throughput rate, head pressure, and screw design) affect torque?
8. What should be done before starting a new production run?
9. How long does it take for an extruder to come to equilibrium and why? Should product be collected or process changes be made prior to reaching equilibrium and why?
10. How do you change a screw configuration?
11. What is critical in setting up a different screw configuration?
12. What is the best way to disassemble an extruder screw?
13. What is meant by heat soaking an extruder and how long does it take?
14. What is an SOP?
15. Define a possible temperature profile to process resin A being fed into barrel zone 1 and filler being fed into barrel zone 6 of a 10-barrel machine.
16. Describe a procedure to change from resin F to resin M, where F has a higher melting point than M, is incompatible chemically, and has a higher viscosity than M.
17. What is the effect of screw speed on throughput rates, percent screw fill by the polymer, and residence time?
18. How does residence time relate to polymer degradation?

15 Applications

Historically, twin screw extruders have been used for compounding, profile extrusion (particularly PVC in conical or counterrotating parallel), devolatilization, and reactive extrusion. Recently, twin screws are finding use in other extruder applications, as processors are realizing the cost advantage of formulating their own compounds and extruding them directly into finished products such as sheet or film, foam, recycle, and other extruded shapes. The chart in Table 15.1 compares twin screw extrusion applications for compounding and profile extrusion with the equipment used and the processing conditions. Some comparisons are very obvious, while others require explanation. In compounding, the percent screw fill is low because of the high speed and the short residence time in the extruder. Twin screw use is compared to single screw extrusion, where a significant amount of compounding and profile extrusion is done. Tangential screws are nonintermeshing, parallel, counterrotating twin screw extruders. These are mainly used for compounding, devolatilization, and reactive extrusion; they are not used in profile extrusion.

Table 15.1 Comparison of Twin Screw Applications

	Compounding	Profile
Screw Speed	High	Low
Degree of Screw Fill	Low	High
Die Pressure Uniformity	Fair	Good
Screw Speed	100–2000	5–50
Melt Pressure	Low	High
Twin Screw Use	45%	55%
Twin Screw Type	Corotating—75% Counterrotating—20% Tangential—5%	Corotating—95% Counterrotating—5% Tangential—0%
Downstream Feeding	Common	Not Common
Dispersive Mixing	Common	Not Common
Distributive Mixing	Common	Sometimes

15.1 Compounding

The principal uses for twin screw extrusion remain either compounding resins and formulations (mainly corotating) or profile extrusion with rigid PVC for house siding, gutters, downspouts, pipe, flooring, and window profiles (which use counterrotating, conical, and parallel extruders). In compounding, each user develops the screw configuration for particular applications, and that screw configuration is treated as proprietary information. The flexibility and versatility of twin screw compounding extruders are demonstrated by their

- Ability to feed liquids easily
- Downstream feeding
- Positive downstream feeding with a side feed extruder
- Easy ingredient addition at different feed ports with gravimetric or volumetric feeders
- Vacuum venting for volatiles removal
- Ability to develop dispersive and/or distributive mixing
- Ease of modifying screw designs to optimize a proprietary screw design to meet a specific objective

High additive levels (color concentrates, fillers,[1] flame retardants, etc.) can be divided into multiple side feed streams to uniformly disperse the additives at high concentrations. Reinforcing fiber[2][3][4] and fillers can be added downstream to the melt to minimize fiber or particle aspect ratio attrition that normally occurs when fiber or filler is added with cold pellets in the extruder feed throat. In addition, the reinforcements are not exposed to the high shear region associated with resin melting. Downstream addition allows distributive mixing to be incorporated in the screw design after the fiber addition to properly distribute the fibers and break up the fiber bundles, without destroying the fiber length.

The same downstream addition approach is used for shear-sensitive additives. Blowing agents, microspheres, or other additives can be destroyed in the extruder feed throat or damaged in the melting section due to high shear. Screw designs can be optimized to gently mix the ingredients into the melt, while providing the best additive distribution and dispersion in the final product.

Liquid additives such as color concentrates, plasticizers, lubricants, mineral oil, etc., are easily added downstream with twin screw extrusion. Depending on how much liquid is added and its temperature stability, the liquid can be added in more than one location along the barrel to provide the proper concentration in the final product and the proper mixing. Gear mixers placed either directly below the liquid injection port or farther downstream effectively mix the liquid into the molten resin. If the liquid concentration is high, the addition may be split in half or thirds, with one-half or two-thirds added into the first liquid injection port and the last one-half or one-third added into a second liquid injection port farther downstream. Proper mixing screw elements provide optimum distribution.

Different feeder configurations are used to add components at different entry points along the extruder. Some installations have a mezzanine or second floor over the

extruder, supporting the feed equipment. Feed streams flow freely by gravity through pipes to the extruder feed zones. Feeders are easily moved from one feed location to another to deliver material to different extruder locations. Acrison Inc. makes a continuous mixer,[5] shown in Fig. 15.1, which is a trough with a screw conveyor. Different feeders are aligned over the trough in various locations, delivering a constant feed rate to the trough, where it is continually removed by the screw conveyor and fed directly to the extruder. This allows feeders to easily add multiple ingredients to one feed stream that goes directly to the extruder feed throat in the first barrel section or to a side feed extruder downstream. There are various feed systems available to feed multiple resin and additive streams into one extruder feed opening. Feeders have been suspended from framework over the extruder to supply resin at the various feed locations. Different approaches work, but all approaches are guided by the following requirements:

- Available space
- Ability to clean equipment between runs
- System to deliver resin or additives to the feed hopper
- Flexibility to change feeder locations
- Need to calibrate and change from one setting to another

Other applications in compounding include alloying or blending two or more different resin systems. The blend components can be a physical mix where one or more resins are dispersed in a carrier resin.[6] The carrier is the continuous phase. Under photo microscopy, one can observe the noncontinuous phase as small islands dispersed in the large continuous matrix resin. Addition and dispersion of a second or third resin in the continuous phase determines the compound property performance, based on the disperse phase size and adhesion to the continuous phase. With a twin screw extruder, the mixing sections can be modified to provide the proper dispersion of the noncontinuous or disperse phase in the continuous phase. Alloying is mixing two different resin systems with two distinct glass transition temperatures, e.g., polystyrene and polyphenylene oxide, in an extruder to produce a third component with one glass transition temperature. An alloying example is Noryl®, which is produced by mixing polystyrene and polyphenylene oxide in a twin screw extruder. As with blending, mixing during alloying is critical to obtain the proper mixing and component particle size. To alloy or blend resins, all resins may be added to barrel section 1 or the second resin may be added downstream, depending on its thermal stability. If three resins are involved, two can be added to the feed throat in barrel section 1, and the third component downstream.

Another application where dispersion and particle size are important is blending impact modifiers into a resin matrix to modify its impact properties. Particle size, particle size distribution, and particle adhesion to the matrix is critical in obtaining the maximum property performance. Where the impact modifier is added and the proper mixing to disperse the impact modifier in the continuous phase are critical.

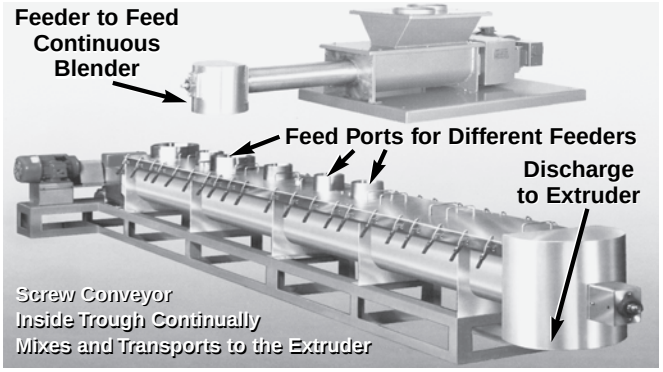


Figure 15.1. Continuous blender produced by Aricson Inc.[5]

Color concentrate production is accomplished by adding colors in the feed throat or downstream to the melted resin. Colorants are either liquids or solids, depending on the colorant, the resin it is mixed with, and its thermal stability.

15.2 Reactive Extrusion and Devolatilization

Reactive extrusion[7][8][9] and devolatilization are accomplished on corotating, parallel, intermeshing twin screw extruders, and counterrotating, nonintermeshing twin screw extruders. Reactive extrusion occurs when two or more components are added to an extruder and a chemical reaction takes place in the extruder. Typical reactions performed in an extruder include

- Cross-linking
- Grafting
- Polycondensation
- Bulk polymerization
- Functionalizing resin backbones
- Compatibilizing
- Controlled depolymerization or degradation

Reactive extrusion is currently done with several commercially available resin systems to produce new polymers, or by combining two or more noncompatible polymers using a third material to compatibilize the first two and produce a product with unique properties. Some commercially available materials made by reactive extrusion are

- Polypropylene grafted maleic anhydride
- Noryl GTX®

- Nylon grafted maleic anhydride
- Polyurethanes
- Polypropylene depolymerization

The advantages of using an extruder as a reactor are

- Ease of running at high temperature to accelerate the reaction
- Sequential ingredient addition to accomplish the desired addition sequence
- Efficient dispersive and distributive mixing at the proper location
- Intimate contact of the various ingredients resulting from the low volume in the extruder reaction area
- Controlled reaction time accomplished with the narrow residence time distribution
- Increased reaction speed by controlling reaction temperature
- Ability to achieve reaction in a high-viscosity medium and still have good mixing

The most significant disadvantage is that the reaction time and reaction kinetics have to be short due to the extruder length. To obtain a reasonable throughput rate, short reaction times are required to match the relatively short residence time in the extruder. Additional barrel sections can be added to increase the extruder length and reaction time; however, at high throughput, only incremental increases in reaction time are available. And they are at a high expense.

Obviously, the most important criterion is to control the chemical reaction in the extruder. This requires the correct stoichiometry for addition reactions, like grafting or polyurethane production, that are controlled by the ingredient feed stream rates. Polycondensation reactions are controlled by water removal; consequently, devolatilization and venting are critical to the reaction sequence. Controlled polymer rheology uses peroxide addition to polypropylene to lower the polymer molecular weight, which is dependent on the peroxide-to-polypropylene ratio. It may also be necessary to provide an oxygen-free environment. The reaction requirements must be understood to design the extruder screw and barrel configuration to meet the objectives.

In practice, reactive extrusion is viewed as several small processes, each dependent on the previous process to produce the correct final product. Reactive components are combined in one extruder section; if one component needs preheating and melting prior to adding the second component, this requires a separate extruder section; volatiles are removed in another section; mixing to ensure complete reaction occurs elsewhere; and the final product must be homogenized and extruded through the die. Catalysts or compatibilizers may be critical in the reaction sequence; consequently, addition in the correct order, in particular extruder barrels, is essential to the reaction kinetics.

Corotating, intermeshing and counterrotating, nonintermeshing twin screw extruders are used for devolatilization, to remove solvents and water from polymerization reactions, or to drive polymerization reactions to completion. Twin screw extruders are significantly more effective in removing high volatiles concentration, compared to single screw extruders. Devolatilization is driven by good melt seals and high vacuum to isolate the different stages within the extruder. To remove high solvent levels to parts per million (ppm) concentrations, high vacuums are required in multiple stages, so the material is subject to vacuum three or four times along the extruder barrel length. If organic solvents are being removed, they are condensed and reused. In some situations, a stripping agent will be injected into the polymer melt to assist the solvent removal.

15.3 Profile and Other Twin Screw Extrusion Applications

As mentioned earlier, counterrotating twin screw extruders, particularly conicals, are used in rigid PVC profile extrusion to produce pipe, window profiles, siding, and so forth. Counterrotating extruders are particularly effective for this application, for the following reasons (refer to Table 15.1):

- They run at low screw speed, generating low shear heat.
- Their positive pumping mechanism develops uniform, high die pressure.
- They provide superior mixing for PVC formulation.
- They have a high screw fill.

For high-volume rigid PVC production, the PVC resin is purchased in PVC powder, which is premixed with the necessary stabilizers, plasticizers, colorants, and fillers in batch mixers. The PVC is subsequently fluxed for a prescribed time and temperature to adhere the thermal stabilizers and plasticizers to the PVC particles. The colorants and fillers are added sequentially to the batch to produce a thermally stable formulation that is extruded into a custom profile. Since PVC is an amorphous resin, it does not have a defined melting point, and it continuously softens as the temperature is increased until the viscosity is low enough to flow and be processed by the extruder. The softening and flowing is called plastication. Conical twin screws take the formulation and gently heat it and massage it until it is hot enough to flow in the extruder. In some instances, the melting process is called fluxing, because PVC does not go through a well-defined melting point; instead, it continuously softens at higher temperatures until all the ingredi-

ents are fluxed and mixed together. As a result, downstream feeding is not normally done, as all the ingredients are preblended prior to addition to the extruder.

Once the formulation is properly fluxed, it is devolatilized to remove any air or volatile gases that can cause problems in the final products. The self-wiping, positive-displacement, counterrotating screws convey the formulation forward and provide very uniform pumping to the die, providing the constant high pressure required in profile extrusion. PVC is very temperature-sensitive and easily degrades at high temperatures in an autocatalytic reaction, generating hydrochloric acid. Once degradation starts, it cannot be stopped, and it becomes progressively worse until the extruder must be purged and cleaned. Consequently, the shear heating generated during processing is critical. The low screw speed associated with the counterrotating twin screws is advantageous to control PVC melt temperature. PVC is not typically processed on high speed corotating twin screw extruders due to the shear heat generation and resin degradation.

Counterrotating twin screw extruders work very well in profile extrusion because of the high, uniform head pressure that is generated. Corotating twin screws are finding uses in profile, sheet, and film applications with a gear pump added to the extruder to generate both sufficient and constant pressure. The advantage of using twin screw extruders for these applications is that the custom or captive extrusion operation can purchase virgin resins and additives and do its own compounding prior to sheeting or filming. Previously, a film or sheet converter purchased a preformulated material from a compounder for his operation. It is more cost-effective to produce the compound internally and go directly into sheet or film in one operation. In addition, this procedure has the added advantages of one less heat history plus the ability to modify and tweak the formulation to optimize product quality and performance. Numerous corotating twin screw extruders are currently used in film and sheet applications.[10] Figure 15.2 shows a coextrusion setup by American Leistritz using a twin screw extruder in combination with a Merritt Davis single screw extruder to produce a coextruded sheet.

Twin screw extruders are used for processing continuous foam articles in either sheets or profiles.[11] Gas is injected into the polymer after the melting zone, where the melt seal prevents the gas from blowing back upstream and exiting the extruder. Material is introduced as a gas, liquid, or chemical blowing agent to generate the gas for foaming. Chemical blowing agents generate gas at a specific decomposition temperature. This allows the solid chemical blowing agent to be introduced downstream or in the feed throat at the appropriate melt temperature that gas is generated. Carbon dioxide and nitrogen are used as gas and are directly injected into the extruder under pressure. Direct gas injection can produce lower density products than chemical blowing agents. The gas needs to be soluble in the resin matrix to achieve the desired foam density. Pentane is a liquid that vaporizes at very low temperatures; it is injected as a liquid and, upon vaporization, forms a foamy product. Three mixing zones are required to properly disperse the blowing agent in the resin. First, the dynamic seal or melt seal that prevents gas from escaping back through the feed throat must be sufficient to prevent the gas from flowing backward through the feed port. The melt seal is normally accomplished with dispersive mixing elements and a restrictive barrier. In the second mixing zone, distributive mixing combines the gas with the molten resin to uniformly distribute it throughout the melt. The third mixing zone is for final melt homogenization prior to exiting the extruder.

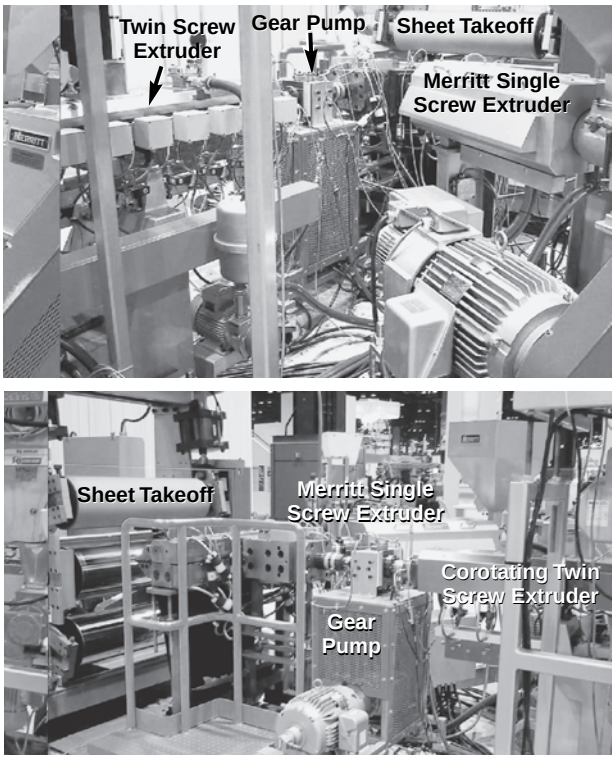


Figure 15.2. American Leistritz corotating twin screw coextrusion sheet line.

Twin screw extrusion is a more flexible and versatile process than single screw extrusion. Since twin screw extruders cost significantly more than single screw extruders, twin screw extrusion is justified by its increased capability and flexibility to accomplish projects and objectives that can't be done with single screw extrusion.

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Review Questions

1. Why is a corotating twin screw extruder commonly used in compounding applications and a counterrotating twin screw extruder found in profile extrusion?
2. What is the proper twin screw extruder setup to compound fiberglass reinforcement?
3. What types of reactive extrusion are done in twin screw extruders and what requirements exist for reactive extrusion?
4. To ensure uniform film or sheet production with a corotating twin screw extruder, what is required?
5. Why is PVC profile extrusion commonly found on a counterrotating extruder?
6. What twin screw extruder setup is anticipated for alloying and blending?

16 Scale-Up

Scaling up from one size twin screw extruder to another can always create processing issues and problems to be solved. If the preliminary product work is done on a 20 to 30 mm twin screw extruder and the product attributes are very good, the next step is to scale it up to a production machine to develop the process. Small problems that seemed insignificant while running on a 20 to 30 mm twin screw extruder at 10–40 pounds/hour can become very large problems on bigger machines. Typically, small laboratory extruders used for product development purposes are not directly scalable to a large size machine. Process data generated on small extruders are not readily scalable to a larger extruder. If the product development is done on a 40–50 mm extruder and the process is optimized to run smoothly on the 40–50 mm extruder, it normally is easily scalable to a larger extruder. However, if a new product and process are developed on an older, three-lobe, 40–53 mm extruder and are being scaled to a 90 mm, two-lobe machine, some process changes might be anticipated.

In scaling up a process from one size extruder to another, any limiting factor, whether it is product or process related, needs to be identified and monitored. The product limiting factor is property deterioration due to higher processing rates. As an example, the property profile at low throughput rates may exhibit high impact and flexural properties. However, at high throughput rates, more shear heat is generated. This increases the melt temperature, causing a reduction in molecular weight. This lower molecular weight leads to lower impact properties. With the larger extruder, impact properties are equivalent to the smaller extruder when the throughput rate and screw speeds are decreased, because the polymer melt temperature is lower. It may be possible to change the screw design to lower the melt temperature; however, this may not provide enough mixing, resulting in some other product deficiency. Product limiting factors can be difficult to overcome and need to be identified early in the process development and scale-up work so they can be addressed.

The other limiting factors involve process limitations. There are many process limiting factors in scale-up. These include heat input, motor power, downstream equipment, and feed capacity. The first step in process scale-up is a volumetric scale from one extruder to another. Volumetric scale-up assumes the volume that can be fed to the extruder by each feeder limits the throughput. We have the following information about a smaller extruder operating at a high extrusion rate:

- Screw configuration
- Maximum screw speed
- Torque limitations
- Maximum volume that can be introduced (the material fed into barrel zone 1 plus any downstream feed)
- Vacuum or atmospheric vent that can remove all the volatiles

We can then assume that the smaller extruder is operating at its volumetric limit. In order to scale to a larger extruder, we need to have the following conditions:

- Similar screw design and volumetric ratio in the larger extruder
- Same speed as defined by the outer flight speed or element tip speed in each extruder
- Same percent fill and average residence time in the extruder

For extruders with the same outer-to-inner diameter ratio for the screw elements (D_o/D_i ratio) operating at the same screw tip speed, we will have the same average shear rate, translating to similar shear heating. Then scale-up to the anticipated volumetric throughput is given by Eq. (16.1).

$$Q_2 = Q_1 \times \left(\frac{D_2}{D_1}\right)^3 \tag{16.1}$$

where Q_2 = Anticipated throughput of the larger extruder with diameter D_2

Q_1 = Measured throughput of the smaller extruder with diameter D_1

The equation changes slightly if the D_o/D_i ratio is not the same for both extruders. Use Eq. (16.2) to calculate the anticipated volumetric throughput rate if the D_o/D_i ratio is not the same, but similar screw geometries are used in both extruders with similar screw tip flight speed.

$$Q_2 = Q_1 \times \left(\frac{Free\ Volume_2}{Free\ Volume_1}\right) \tag{16.2}$$

where $Free\ Volume_1$ = Screw free volume in the original extruder

$Free\ Volume_2$ = Screw free volume in the larger extruder

Assuming the volumetric scale-up is possible, a second process limitation may be available power. With a given screw design and screw speed, is there enough power to generate sufficient torque? There may not be enough power to process the desired throughput rate without exceeding the motor torque limitations. In the event this happens, lower throughput rates can be run with the same screw design or the screw design may have

to be changed to generate less torque and shear heat. Restrictive elements (rearward conveying elements) generate significantly higher torque requirements. Reducing the number or type of restrictive elements in the melting area or before downstream feeding or venting lowers the torque and power requirements, while reducing mixing and work put into the material. This must be done carefully to ensure the process and/or product is not compromised in any way.

Power is scaled up based on constant specific energy. Specific energy is defined as the energy divided by the output rate and is measured in kilowatts/kilogram; see Eq. (16.3).

$$SE = \frac{P}{Q} \tag{16.1}$$

where SE = Specific energy (kW-hr/kg per hr or HP-hr/lb per hr)
P = Drive motor power (kW or HP)
Q = Throughput (kg/hr or lb/hr)

Power scale-up to a large machine is based on

- Similar screw geometry
- Same percent fill
- Equivalent screw speed as measured by the flight tip velocity

Equation (16.3) gives the scaled-up power required based on each extruder having the same D_o/D_i ratio. Consequently, if the specific energy of the small extruder is known, using a constant specific energy and the larger extruder throughput, the power required can be calculated. It may be necessary to estimate the larger extruder throughput in this calculation to determine the power required.

With different D_o/D_i ratios for the two extruders and different percent fills, the maximum theoretical throughput is calculated using Eq. (16.4):

$$Q_2 = \frac{P_{\text{at a given RPM}}}{SE} \tag{16.4}$$

When one knows the specific energy of the smaller extruder and the power available in the larger extruder, one can calculate the throughput. The specific energy value for the smaller extruder in this equation should be based on 90% torque to provide a safety factor.

A third potential process limiting factor is the ability to transfer heat into or from the resin to obtain the correct melt temperature. If the limitation is caused by insufficient heat, this is normally easy to overcome by changing the screw design and/or raising the barrel temperatures. In large extruders with high throughput rates and high screw speeds, the process limitation is more likely to be heat removal to prevent the polymer from overheating and degrading. Barrel cooling alone may not remove sufficient heat at high screw speed to prevent resin degradation. One way to lower the polymer melt temperature is to lower the screw speed; however, this may create high motor torque, shutting the extruder down. A second alternative is to change the screw design to generate less shear heat by moving the melting point toward the die. This is done by adding conveying elements between the kneading blocks in the melting section to dissipate heat build-up, or decreasing the mixing downstream. Altering the screw design may change the processing characteristics by moving the polymer melting point to another location in the extruder or may provide insufficient mixing to generate the same product. If the polymer is shear-sensitive (resin viscosity thins more with shear rate than temperature) and has limited thermal stability, the screw melting and mixing section design is very critical. Due to the short residence time at high throughput rates, barrel temperatures alone are not effective in supplying or removing heat from the melt. Increasing the barrel screw flight clearance will decrease the shear rate, reducing the melt temperature. Undersized mixing elements or large gaps between dispersive mixing disks may reduce the melt temperature.

Heat transferred from the barrel heaters to the resin is controlled by several factors. Heaters located on the outside of the barrels are dependent on several heat transfer coefficients to transfer heat from the barrel to the resin. Heat transfer is dependent on the coolant, barrel thickness, the air gap between the barrel and the liner, and heat transfer from the barrel liner to the melt film in contact with the barrel wall and finally to the bulk resin. The polymer melt film and the air gap between the barrel and barrel liner are poor heat conductors. Neither of these factors aids heating or cooling the bulk polymer.

Direct scale-up for heat transfer is based on heat transfer coefficients being constant for both extruders, which is based on heat conduction from the barrel heater to the bulk melt. As with other scale-up assumptions, the percent fill must be the same from the smaller to larger extruder. Scale-up based on heat transfer for the same D_o/D_i ratio is given by Eq. (16.5):

$$Q_2 = Q_1 \times \left(\frac{D_2}{D_1}\right)^2 \tag{16.5}$$

where Q_2 = Throughput of the larger extruder with the same melt temperature and heat transfer as the smaller extruder
Q_1 = Throughput of the smaller extruder
D_1 = Diameter of extruder 1
D_2 = Diameter of extruder 2
with the same melt temperature and heat transfer.

For extruders with different D_o/D_i ratios, the throughput with the same heat transfer is given by Eq. (16.6):

$$Q_2 = Q_1 \times \frac{Barrel\ Area_2}{Barrel\ Area_1} \tag{16.6}$$

Scale-up of nonintermeshing twin screw extruders is similar to single screw extruders, which was covered in Part 1, Chapter 7. The throughput anticipated in a larger nonintermeshing twin screw extruder (extruder number 2), based on a smaller extruder (extruder number 1), is given by Eq. (16.7):

$$Q_2 = Q_1 \times \left(\frac{D_2}{D_1}\right)^{2.5} \tag{16.7}$$

Equation (16.7) is based on the die open area being scaled up the same, so the backpressure and resin backflow into the extruder remain approximately equivalent.

Review Questions

1. What is meant by limiting factors in scale-up?
2. What are some product and process limiting factors that may interfere with scale-up?
3. Assuming the same D_o/D_i ratio, the same free volume in the extruder, approximately the same screw configuration, and similar screw flight tip speed, what is the throughput of a 90 mm twin screw extruder based on a 57 mm twin screw extruder running at 800 pounds/hour? Assume a volumetric scale-up.
4. Using the same assumptions in problem 3, calculate the throughput based on a similar scale-up in the heat transfer between the two extruders.
5. During actual scale-up, if excess heat is built up in a large extruder that affects the thermal stability of the polymer, what options does an operator have to correct the problem?

17 Shear Rate, Pressure Drop, and Other Extruder Calculations

This chapter shows how to calculate shear rates in the extruder and die, pressure drops in the die, and other calculations in twin screw extruders.

17.1 Shear Rates

Shear rate in the extruder screw channel is given by Eq. (17.1):

$$\dot{\gamma} = \frac{\pi \times D \times N}{60 \times h} \tag{17.1}$$

where $\dot{\gamma}$ = Shear rate in screw channel
D = Screw diameter in mm
N = Screw speed in revolutions/minute
h = Channel depth in mm, and shear rate is given in sec^{-1}

Shear rates between the screw flight tip and the barrel wall are calculated using Eq. (17.1), except h = distance between the screw flight and the barrel wall, given in mm. The shear rates between the screw flight tip and the barrel wall are high because h is small, and low between the bottom of the screw channel and the barrel wall because h is large there.

Shear rate calculation in the die lip area depends on the die shape. In a round die channel (rod or strand for compounding), the shear rate is calculated from Eq. (17.2):

$$\dot{\gamma} = \frac{4 \times Q}{\pi \times R^3} \tag{17.2}$$

where $\dot{\gamma}$ = Shear rate in round channel
R = Radius of the hole
Q = Volumetric flow rate calculated from Eq. (17.3)

$$Q = k \times \frac{\Delta P}{\eta} = \frac{\pi \times R^4}{8 \times L} \times \frac{\Delta P}{\eta} \tag{17.3}$$

where $k = \pi R^4/8L$ (resistance factor)
L = Channel length
ΔP = Pressure drop across the channel
η = Polymer viscosity

The pressure drop in the channel is calculated from Eq. (17.4):

$$\Delta P = \frac{2 \times \tau \times L}{R} \tag{17.4}$$

where τ = Shear stress = F/A = Force applied per unit area

An example of the effect of the resistance factor in a rod or strand die on the flow rate is shown in Fig. 17.1. As the hole diameter decreases, the pressure increases based on k or the resistance factor.

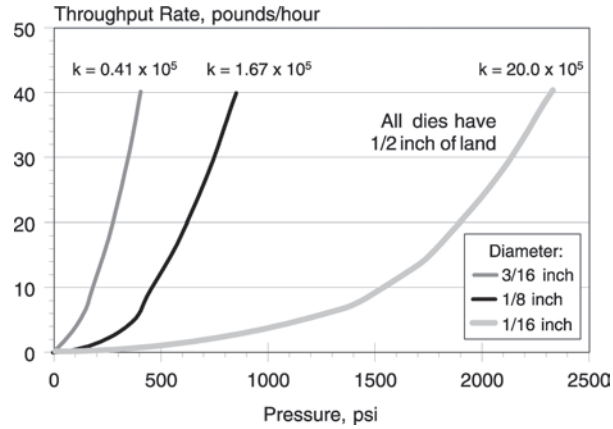

Figure 17.1. Effect of die resistance on throughput.[1]

Shear rates in a rectangular die channel (in sheet, film, and profiles dies) are given by Eq. (17.5):

$$\dot{\gamma} = \frac{6 \times Q}{W \times h^2} \tag{17.5}$$

where W = Channel width
h = Channel height
Q = Volumetric flow rate calculated from Eq. (17.6)

$$Q = k \times \frac{\Delta P}{\eta} = \frac{W \times h^3}{12 \times L} \times \frac{\Delta P}{\eta} \tag{17.6}$$

where $k = Wh^3/12L$ (resistance factor)
L = Channel length
ΔP = Pressure drop across the channel
η = Polymer viscosity

Shear rates in annular dies (blown film, pipe, and tubing) are calculated using Eq. (17.7):

$$\dot{\gamma} = \frac{6 \times Q}{\pi \times (R_i + R_o) \times h^2} \tag{17.7}$$

where R_i and R_o = Inner and outer radius
h = Die gap
Q = Volumetric output calculated from Eq. (17.8)

$$Q = k \times \frac{\Delta P}{\eta} = \frac{C_m \times h^3}{12 \times L} \times \frac{\Delta P}{\eta} \tag{17.8}$$

where C_m = Mean circumference $[2\pi (R_i + R_o)]$

An example of shear rate calculations at different locations in the extrusion process follows. Base your calculations on the following system:

- 57 mm intermeshing, corotating twin screw extruder
- equipped with a 57/57 conveying element
- having a barrel-to-wall clearance of 0.5 mm

- a channel depth of 9.5 mm
- and running at a screw speed of 400 rpm
- to produce a throughput rate of 227 kg/hour.

Calculate viscosity

- at the bottom of the channel
- between the flight and barrel wall
- and in the die land area of a sheet die that is
  — 1524 mm wide
  — with 3.2 mm opening.
- The density of PP is 0.91 g/cc
- the melt density = 0.75 g/cc
- and shear rate versus viscosity curve is given in Fig. 17.2.

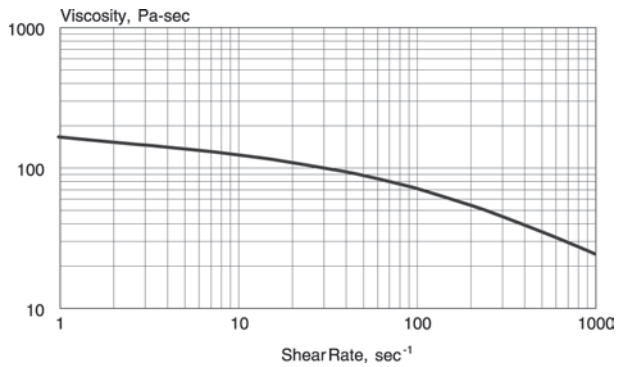

Figure 17.2. Shear rate versus viscosity for 12 MFI PP.

Using Eq. (17.1),

$$\dot{\gamma} = \frac{\pi \times D \times N}{60 \times h} = \frac{3.1416 \times 57 \times 400}{60 \times 57} = 126\,\text{sec}^{-1}$$

extrapolation in Fig. 17.2 shows the viscosity at 126 sec^{-1} is approximately 68 Pa-sec.

For the shear rate over the flights, again use Eq. (17.1):

$$\dot{\gamma} = \frac{\pi \times D \times N}{60 \times h} = \frac{3.1416 \times 57 \times 400}{60 \times 0.5} = 2387\,\text{sec}^{-1}$$

Again from Fig. 17.2, the viscosity moving across the barrier flight is approximately 16 Pa-sec. (This number is extrapolated from the shear rate at 2387 sec^{-1}).

For shear rate in the die land, use Eq. (17.5):

$$\dot{\gamma} = \frac{6 \times Q}{W \times h^2} = \frac{6 \times Q}{1524 \times 3.2^2}$$

To calculate Q, use Eq. (17.6):

$$Q = k \times \frac{\Delta P}{\eta} = \frac{W \times h^3}{12 \times L} \times \frac{\Delta P}{\eta}$$

Since the pressure drop and viscosity are not known in this example, go back and use the 227 kg/hour output and convert to a volumetric flow rate based on PP melt density being 0.75 g/cc. This does not give the absolute shear rate or viscosity due to the fact that the resistance to flow in the die is not included. The shear rate and viscosity will be close to the actual value using the volumetric output for the volumetric flow calculation:

$$Q = \frac{227\ \text{Kg}}{\text{hr}} \times \frac{1\ \text{hr}}{3600\ \text{sec}} \times \frac{1\ \text{cc}}{0.75\ \text{g}} \times \frac{1000\ \text{g}}{\text{Kg}} \times \frac{1000\ \text{mm}^3}{\text{cm}^3} = 84{,}100\ \frac{\text{mm}^3}{\text{sec}}$$

Now we can use Eq. (17.5) to calculate the shear rate in die lip area:

$$\dot{\gamma} = \frac{6 \times Q}{W \times h^2} = \frac{6 \times Q}{1524 \times 3.2^2} = \frac{6 \times 84{,}100}{1524 \times 3.2^2} = 32.2\ \text{sec}^{-1}$$

And from Fig. 17.2 we see that the viscosity in the die lip area is approximately 98 Pa-sec.

17.2 Extruder Calculations

The power required to heat the polymer from room temperature to the molten state is given by Eqs. (17.9) and (17.10):

$$P = \textit{Heat to raise polymer from room to melt temperature} + \textit{Heat of Fusion} \tag{17.9}$$

$$P = (m \times C_p \times \Delta T) + \Delta H_{\text{fusion}} \tag{17.10}$$

where P = Power
m = Mass flow rate in kg/hr
C_p = Heat capacity in kJ/kg °C
ΔT = Difference in temperature between room temperature and melt temperature in °C
ΔH_{fusion} = Heat of fusion for the polymer matrix

Remember that for amorphous polymers ΔH_{fusion} is zero. Using Eq. (17.9), what is the power required to heat polypropylene (PP) running at 400 pounds/hour from 25°C to 235°C?

- Heat capacity of PP = 2.10 kJ/kg °C
- ΔH_{fusion} = 102 J/g

First convert the 400 pounds/hour to kg/hr and ΔH_{fusion} to kJ/kg:

$$m = \frac{400\,\text{lbs}}{\text{hr}} \times \frac{0.454\,\text{kg}}{\text{lb}} = 182\,\frac{\text{kg}}{\text{hr}}$$

$$\Delta H_{\text{fusion}} = \frac{102\,\text{J}}{\text{g}} \times \frac{1000\,\text{g}}{\text{kg}} = \frac{102{,}000\,\text{J}}{\text{kg}} \times \frac{1\,\text{kJ}}{1000\,\text{J}} = 102\,\frac{\text{kJ}}{\text{kg}}$$

Using Eq. (17.10),

$$\begin{aligned} P &= (m \times C_p \times \Delta T) + \Delta H_{\text{fusion}} \\ &= \frac{182\ \text{kg}}{\text{hr}} \times \frac{2.10\ \text{kJ}}{\text{kg °C}} \times \frac{235°\ \text{C} - 25°\ \text{C}}{} + \frac{182\ \text{kg}}{\text{hr}} \times \frac{102\ \text{kJ}}{\text{kg}} \\ &= \frac{80{,}262\ \text{kJ}}{\text{hr}} + \frac{18{,}564\ \text{kJ}}{\text{hr}} = 98{,}826\ \frac{\text{kJ}}{\text{hr}} \end{aligned}$$

As 1 Joule per second = one watt, we divide the power by 3600 sec/hour to convert to kW:

$$P = \frac{98{,}826\text{kJ}}{\text{hr}} \times \frac{1\text{hr}}{3600\,\text{sec}} = 27.5\text{kW}$$

And as 1 kW = 1.36HP or 1 kW = 860 kcal/hr, the power equation becomes:

$$P = \frac{27.5\text{kW}}{} \times \frac{1.36\text{HP}}{\text{kW}} = 37.4\text{HP}$$

or

$$P = \frac{27.5\text{kW}}{} \times \frac{860\text{kcal}}{\text{hr-kW}} = 23{,}650\frac{\text{kcal}}{\text{hr}}\,of\ Energy$$

From the perspective of energy costs, it takes approximately 27.5 kW per hour to process 400 pounds/hour of PP resin at 235°C in this example. Since extrusion equipment is not 100% efficient, it actually requires more energy to process than calculated, because some heat is lost to the surroundings and the motors and drives are not 100% efficient. This considers only the energy to melt the resin and not the energy to mix and work the resin downstream in the extruder.

Energy losses in an extruder come from feed throat cooling, barrel cooling, convective heat loss from the barrel and die, heat loss from DC drives, gear box losses, energy to run pumps, energy for cooling fans, and power to energize the panel.[2] Approximately 61% of the energy put into the process is actually used to convert solid polymer to a molten resin for processing.

17.3 Other Calculations

Numerous other theoretical calculations are possible on twin screw extruders, but they are beyond the scope of this book. Burkhardt, Herrmann, and Jakopin showed the differences between corotating and counterrotating extruders, comparing velocity and stress distribution down the channel on both types of extruders.[3]

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1. Christensen, R. E., "Science of Extrusion Course," March 10, 1997.
2. Kruder, G. A., "Extruder Efficiency: How You Measure It—How You Get It," *Plastics Engineering*, p. 20, June 1981.
3. Burkhardt, K., Hermann, H., and Jakopin, S., "Plasticizing and Mixing Principles of Intermeshing Corotating and Counterrotating Twin Screw Extruders," *ANTEC, XXIV*, p. 498, 1978.

Review Questions

1. Calculate the shear rates in the conveying channel of a 30/60 element in a 60 mm intermeshing twin screw extruder and at the flight tip while extruding acrylonitrile butadiene styrene (ABS) running at 700 rpm. Channel depth in the conveying element is 10.0 mm, and the clearance between the flight tip and the barrel wall is 1 mm.
2. Assume the ABS in question 1 is running at 620 pounds/hour through a sheet die, producing 5 mm thick film 1220 mm wide. What is the approximate shear rate in the die lip area? Melt density of ABS is 0.88 g/cc.
3. Calculate the shear rate in a 60 mm two-lobe twin screw extruder with a D_o/D_i ratio of 1:54 in two adjoining kneading blocks with thin disks and wide disks. The clearance between the top of the disks and the barrel wall is 0.75 mm, and the bottom of the disks is 8.2 mm from the barrel wall. What is the range of shear rates in the wide and narrow kneading blocks?
4. Calculate the power required to heat polystyrene (heat capacity is 1.20 kJ/kg °C) from 25°C to 210°C at a throughput rate of 650 pounds/hour.
5. Calculate the power required to heat nylon 6,6 running at 617 pounds/hour from 25°C to 295°C. The heat of fusion for nylon 6,6 is 20.5 (J/kg) $\times 10^4$ and the heat capacity is 2.15 kJ/kg °C.

Part 3: Polymeric Materials

18 Polymer Overview and Definitions

To troubleshoot and solve extrusion process problems, it is important to understand both the equipment operation and the polymer being processed. Understanding either the equipment or the polymers alone may not provide enough information to solve a particular extrusion problem. Polymer behavior in the molten state is different from normal liquids such as water, because polymer melts have both elastic and viscous components, whereas liquids such as water have only a viscous component. For this reason polymers are called non-Newtonian fluids and water is a Newtonian fluid. The shear rate versus viscosity behavior, discussed briefly in Part 1, Chapter 3, is that polymer viscosity decreases with increasing shear rate. A Newtonian fluid shows no viscosity change as the shear rate changes. This drop in polymer viscosity is referred to as shear thinning. The elastic component makes molten polymer behave differently from nonelastic liquids.

This chapter covers the definition of polymers, how they are formed, different types of polymers, and their structures.

18.1 Overview

Polymers are comprised of many atoms connected together to form long chains, referred to as the polymer backbone. The atoms in the backbone are usually carbon (C), oxygen (O), nitrogen (N), or sulfur (S), combined in a configuration that is unique for each polymer. Other atoms attached to the long polymer backbone are normally some combination of hydrogen (H), carbon, oxygen, chlorine (Cl), fluorine (F), and/or bromine (Br), and these differentiate one polymer from another. Each polymer has a unique spatial arrangement of atoms combined in a regular, repeating pattern with specific physical properties and rheological (flow) properties, resulting in different property performance in end-use applications and processing characteristics. No two polymer types have the exact same processing characteristics in extrusion or the same property performance profile relative to impact, tensile and flexural strength, tensile and flexural modulus, color, heat resistance, transparency, electrical properties, and so forth.

Carbon is the major atom in a polymer backbone. Each carbon atom has four chemical bonds spatially arranged around it in a tetrahedron. The simplest carbon molecule is methane. Table 18.1 lists the name, chemical formula, approximate melting points, and boiling points of the first five alkanes, from methane to pentane. Note that methane has the lowest boiling point at –259˚F (–162˚C). Adding more carbon atoms to the molecule increases the boiling point. Methane through butane are gases at room temperature, and pentane is a liquid at room temperature.

Table 18.1. Formulas, Melting Points, and Boiling Points of Five Alkanes

Hydrocarbon	Formula	Melting Point °F	Melting Point °C	Boiling Point °F	Boiling Point °C
Methane	CH_4	–297	–183	–259	–162
Ethane	C_2H_6	–298	–183	–127	–89
Propane	C_3H_8	–306	–188	–44	–42
Butane	C_4H_{10}	–217	–134	31	–0.5
Pentane	C_5H_{12}	–202	–130	97	36

Below 5 carbon atoms in the molecular backbone, the materials are flammable gases at room temperature. Between 5 and 20 carbon atoms, the molecules become flammable liquids at room temperature. Above 20 carbon atoms, molecules become solids at room temperature. Above 36 carbon atoms in the molecular backbone, the material becomes grease. Above 140 carbon atoms, the product is a polyethylene wax. Increasing the carbon atoms to greater than 500 results in a polyethylene polymer. Commercially available polyethylene polymers have on average approximately 7000–15,000 carbon atoms bound together to form the polymer backbone.

The carbon bond angle when four other atoms are bound to it is 109˚. This results in a tetrahedral structure about the central carbon atom. Figure 18.1 models methane (CH_4). The four hydrogen atoms are uniformly distributed around the central carbon atom in a tetrahedral arrangement. Due to the 109˚ bond angle, adjacent carbon atoms are not linearly aligned, as suggested by two-dimensional chemical formulas drawn on paper. Figure 18.2 models a butane molecule with four carbon atoms. Note how the carbon atoms are not aligned in a straight line or in a flat plane. Instead, the

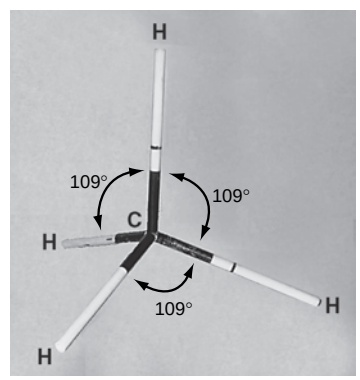


Figure 18.1. Simulated methane molecule with bond angles.

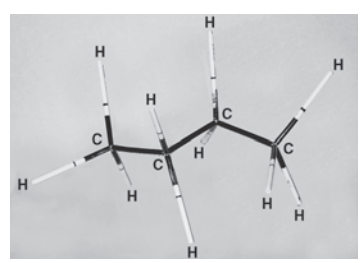


Figure 18.2. Model butane molecule.

carbon backbone in Fig. 18.2 forms a zigzag-type repeating pattern. Combining the zigzag structure with the ability of atoms to rotate 360˚ about their bond axes, entire sections can move from one location to another by a simple rotation about a C–C bond in the chemical backbone. Rotations do not occur in the solid state. However, as heat energy is added to the system, the distance between individual molecules increases. With sufficient heating, a state change to a liquid or gas occurs, and rotation about the C–C bonds constantly occurs. It is easy to envision long polymer chains with thousands of carbon atoms becoming entangled with each other and with themselves as these rotations occur. In the polymer melt, rotations cause entanglements that remain after the polymer solidifies. Polymer chain entanglement in the solid state is responsible for the high strengths and toughness of polymeric materials. It is important to understand the difference between solid, liquid, and gaseous states and the molecular chains in different polymeric materials, as this directly impacts the polymer property performance and processing characteristics.

Long chain polymer molecules are similar to a pot of boiling spaghetti where the individual molecules, like spaghetti strands, become entangled with other polymer molecules. In the molten state the polymer chains resemble the boiling spaghetti. For amorphous polymers in the solid form, the polymer chains resemble the spaghetti after the water has been removed and the spaghetti remains in relatively discrete locations. Figure 18.3 models a short linear polyethylene chain section that is oriented in one direction. Imagine a 10,000 carbon atom chain pulled out in one direction. Assume there is a second chain next to this that is also pulled out or stretched in the same manner to produce limited entanglement between the adjacent chains. Now assume there are many molecular chains lying parallel to each other. Figure 18.4 represents this, with oriented strings lying side by side. If you pick up the ends of the various strings or polymer molecules and pull on them, they are very strong and it takes a large force to break the strings or, in the case of polymers, the molecular chains. However, if the strings or polymer molecules are pulled perpendicular to the lengthwise direction of the strings or polymer molecules, it is very easy to pull the strings apart.

This string analogy can be applied to extruded polymer molecules. As the molten polymer exits the die, the puller draws it in the machine direction, tending to stretch and align the molecules. Consequently, the extruded product has higher tensile and flexural properties in the machine direction than in the cross-machine or transverse direction. In oriented film or monofilaments, the polymer is drawn with supplemental heat outside of the extruder to orient the molecules in the machine direction, resulting in high properties in one direction. Biaxial-oriented film is pulled in the solid state, in both the machine and transverse directions, under controlled temperatures outside the extruder to yield improved properties in both directions. Figure 18.3 shows a short molecular section that is oriented in one direction. Figure 18.5 shows that same short segment with some rotation about the C–C bonds. This allows the molecule to fold back on itself or change direction. Figure 18.5 is a better representation of the polymer as it exits the die. Drawing by the puller aligns some of the molecules in the machine direction; however, unless the draw ratio is very large, a large percentage of the molecules tend to relax and entangle with each other as they leave the die land area. In the die land area, the molecules tend to be oriented in the machine direction due to the polymer flow in the die land area. Upon exiting the die, the polymer relaxes and folds back on itself, entangling with itself and neighboring molecules. This generates the extrudate swell, sometimes referred to as "die swell," associated with extrusion.

Going back to the string analogy, consider the strings going through the die. It is difficult to push a string through a die hole and the associated die land area. As it is pushed, one end tends to buckle and be crunched up into a small volume. Instead, the string is pulled through the die

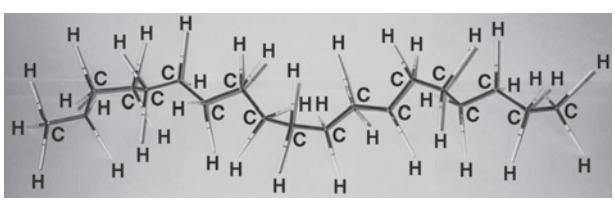


Figure 18.3. Model of a short linear section of polyethylene molecules.

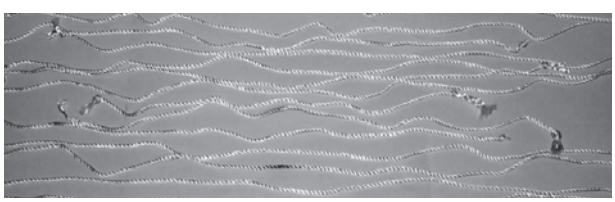


Figure 18.4. Oriented strings simulating oriented polymer molecules.

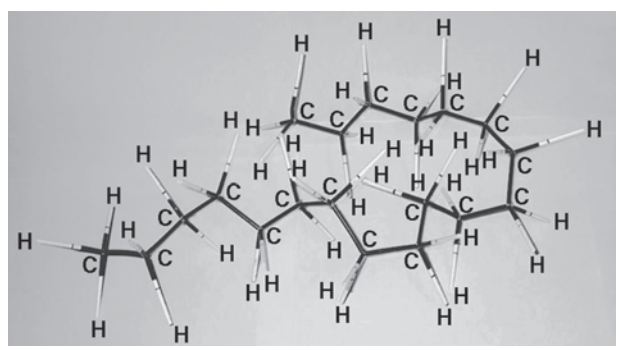


Figure 18.5. Model of short section of folded chain of polyethylene.

hole, elongating in the machine direction. Immediately upon leaving the die, the string relaxes and reentangles as it folds back on itself and neighboring polymers.

Envision long polymer chains made up of a huge number of carbon atoms connected together in some particular spatial arrangement. Figure 18.6 shows a polyethylene polymer backbone with a large number of carbon atoms connected. (The number of carbon atoms depicted is significantly less than is normally seen in a polyethylene molecule). Removing the carbon atoms from the drawing in Fig. 18.6, polymer chains are represented by the curved line shown in Fig. 18.7. As the polymer has

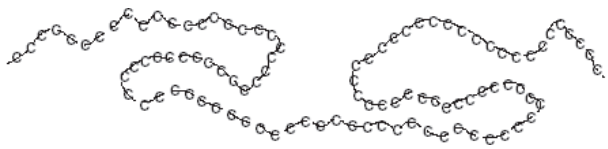
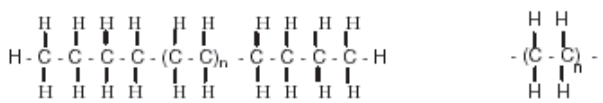


Figure 18.6. Simulated polymer chain of polyethylene.

Figure 18.7. Typical representation of a polymer chain.

many chemical entities that repeat numerous times to form the long polymer chain, the chemical formula is written as



where *n* is the number of repeating units. The number of repeating units, *n*, will vary, depending on the polymer and how many repeating units are required to produce the optimum property performance profile with acceptable processing characteristics. For polyethylene, *n* is approximately 4000. Figure 18.8 shows polymer chains entangled together to form a section of a plastic part. (Different shades are used to identify the individual chains in the drawing.) In the solid state, polymer chains are entangled with limited space between the chains. As the polymer is heated, the space between the chains becomes larger, and in the molten state the space is very large, allowing the chains to actually flow past one another. As the space between chains becomes larger, rotations about the C–C bonds in the polymer backbone become more numerous. Polymer degradation occurs when the C–C bonds or

Figure 18.8. Schematic of a group of individual polymer chains.

other bonds along the polymer backbone rupture and the chains become shorter. Going back to Fig. 18.8, one can start to understand why shorter chains with less entanglement will have lower properties. As the entanglement and chain length decrease, the polymer matrix can support less stress. Going back to the spaghetti analogy, longer spaghetti lengths in the solid state, compared to shorter lengths, are more difficult to pull out and remove from the bulk spaghetti mass.

18.2 Thermoplastic versus Thermoset

Polymers fall in two distinct categories: thermoplastic or thermoset. Thermoplastics can be repeatedly softened and resolidified by adding or removing heat, allowing materials to be processed in extrusion, injection molding, and other melt processing equipment numerous times. Thermosets are heated to accelerate a chemical reaction called curing, where a new chemical species is formed into a solid. After cooling, reheating a thermoset does not result in softening and polymer flow. Thermoplastics are easily recycled into other shapes or forms by heating to the softening or melting point and reprocessing. At one time, thermosets were not considered recyclable; however, different processes are in use today, providing mechanisms to recycle the material as filler or to reclaim the raw materials. However, unlike thermoplastics, thermosets cannot simply be heated above their melting or softening point and reprocessed into another article.

Thermoplastics include polymers such as

- polyethylene (PE)
- polypropylene (PP)
- polystyrene (PS)
- polyvinyl chloride (PVC)
- polymethylmethacrylate (PMMA or acrylic)
- acrylonitrile butadiene styrene (ABS)
- polycarbonate (PC)
- polyethylene terephthalate (PET)
- polybutylene terephthalate (PBT)
- polyamide (PA or nylon)
- polyphenylene sulfide (PPS)
- polyphenylene oxide/polystyrene blend (Noryl®)
- polyetherimide (PEI)
- polysulfone (PSO)

The polymers listed above are the more common thermoplastic polymers used in extrusion applications.

Thermoset materials are made by a chemical reaction between different components to produce a new cross-linked matrix that cannot be remelted. A typical thermoset

reaction is epoxy, which is composed of a resin and hardener that are mixed together and cured, forming a strong new material that is impervious to heat and most chemicals. Polyurethanes are formed by reacting an isocyanate with a polyol. Some thermoset materials are produced by cross-linking a resin matrix to produce a solid part that cannot be remelted during processing. EPDM rubber is a thermoset material that can contain a cross-linking agent. The material containing a cross-linking agent is extruded at low temperature into a desired shape, e.g., a gasket around a car door. After extrusion, the product is heated to promote the cross-linking reaction in the solid state, producing a final product that is impervious to heat and will not melt because of the cross-linking step.

Elastomeric polymers are materials that can be repeatedly deformed when a force is applied and return to their original shape after the force is removed. Elastomers can be either thermoplastic or thermoset polymers, depending on their chemistry and the processing mechanism. The EPDM cross-linked rubber mentioned in the previous paragraph is a thermoset elastomer. Thermoplastic elastomers include styrene butadiene copolymers, ethylene propylene polymers, and polyurethanes.

Cross-linking to produce thermoset products is accomplished when sites on two adjacent polymer chains react to form a bond joining the two individual chains together. Once the two chains are bonded together, heat does not allow the chains to separate and flow individually. High cross-linking levels join the individual polymer backbones, producing one large structure. This prevents cross-linked materials from flowing by either melting or softening. Cross-linking is a chemical reaction that changes the polymer matrix. The two general methods used to form cross-linked material are

- Adding a chemical additive that reacts with each chain to form a common site, bonding two polymer chains together.
- Breaking bonds between carbon atoms by heating them or by keeping them at temperature for too long a time. Broken bonds form free radicals, which react to form cross-linked sites between two polymer molecules.

Figure 18.9 shows polymer chains that are cross-linked and their cross-link sites. Polyethylene is known to cross-link during extrusion if polymer gets held up in a dead spot within the extruder or die and sits for a long time at high temperature. Small particles that do not melt and appear in the final product as small hard particles, described as “fisheyes” or gels, may be cross-linked and thermoset in nature. It is important that all flow paths between the extruder and die and within the dies be streamlined to prevent stagnation, as stagnation allows cross-linking or degradation to occur over time.

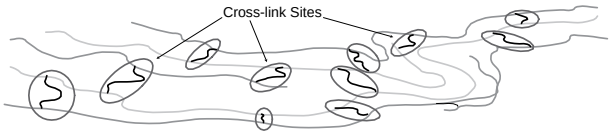


Figure 18.9. Example of cross-linked polymer chains.

18.3 Polymer Formation

Small molecules are combined in polymerization reactions using specific processing conditions, procedures, raw materials, end groups, and catalysts to generate long chain polymers of approximately the same length. The starting molecules are called monomers. Polyethylene is produced from ethylene gas, which is shown in the model in Fig. 18.10. Each carbon atom has four bonds to other atoms, with two electrons (small negative particles) shared between carbon and the adjoining atom to which it is bonded. In ethylene, the four bonds around each carbon atom come from bonds to two hydrogen atoms and bonds to the two adjacent carbon atoms. If a carbon atom bond is broken and a single electron stays with each carbon atom, a free radical is formed. The free radical now has three bonds around each atom (two to hydrogen and one to carbon) plus a free electron. Free radicals are not a stable chemical species, as the carbon atoms want to attract another electron to form a fourth bond. If a free radical comes together with another molecule, a bond is formed with the free radical moving to the end of the added molecule. This can be the addition of a second ethylene group to the initial ethylene group. By repeating this process numerous times, polymers are formed by a process called an addition reaction, as monomers are added to the polymer chain. When the chain becomes the right length to provide the optimum property profile and processing characteristics, the chain length is terminated with an end group that has a free radical or free electron on the molecule.

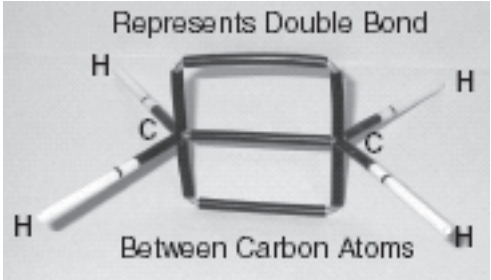
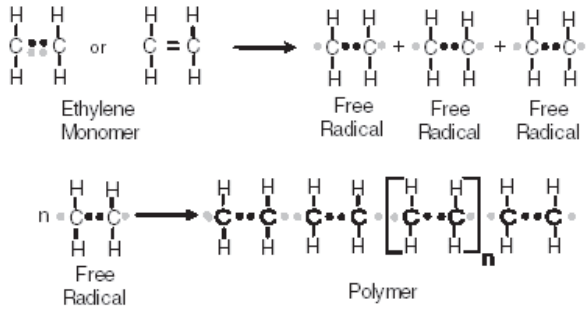


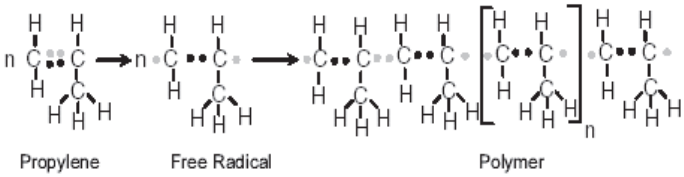
Figure 18.10. Model of ethylene molecule with double bond between carbon atoms.

The reaction sequence for the formation of polyethylene is as follows:



The single electrons on the free radicals combine to form a bond. Free radicals continue to combine until a high-molecular-weight polymer is formed.

A similar reaction sequence is used to produce polypropylene. A free radical addition reaction that polymerizes propylene molecules is similar to polyethylene polymerization. Styrene free radical formation and the subsequent polymerization steps are shown below.



Numerous polymers are made by this free radical addition, as an ethylene double bond is broken to form different polymers with unique properties. When the hydrogen atoms in polyethylene are replaced with other atoms or groups of atoms, a different polymer is produced. Table 18.2 shows a listing of the different groups that replace one or more hydrogen atoms in Structure I to form various polymers. The structures of some common polymers referred to in Table 18.2 are given in Fig. 18.11.

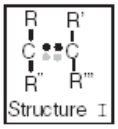
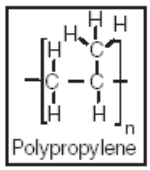
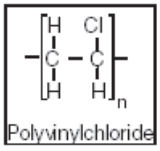


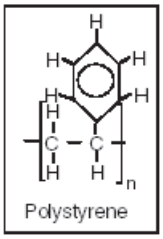
Table 18.2. Polymer Structures Based on C–C

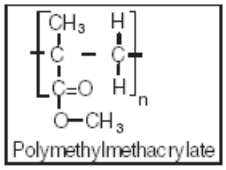
Polymer	R	R"	R'	R'"
Polyethylene	H	H	H	H
Polypropylene	CH_3	H	H	H
Polystyrene	C_6H_5	H	H	H
Polyvinylchloride	Cl	H	H	H
Polymethymethacrylate	CH_3	$COOCH_3$	H	H
Polyvinylidenechloride	Cl	Cl	H	H
Polytetrafluoroethylene (Teflon)	F	F	F	F
Polyvinyldiene fluoride	F	F	H	H
Polychlorotrifluoroethylene	F	F	F	Cl
Polyhexafluoropropylene	F	F	F	CF_3
Polyacrylonitrile	CN	H	H	H
Polybutylene	CH_2CH_3	H	H	H
Polybutylene	CH_3	CH_3	H	H
Ethylene vinyl alcohol	OH	H	H	H
Polyvinylacetate	$OOCCH_3$	H	H	H





Polypropylene, polystyrene, and other similar polymers exist in three distinct structural configurations, depending on the polymerization conditions. For polypropylene, when the pendant methyl (CH_3) groups are all on one side of the polymer chain, this is the isotactic form. When the methyl groups are on alternating sides of the polymer chain, it is the syndiotactic form. When the methyl groups are randomly oriented along the polymer chain, it is the atactic form. These forms are shown for polypropylene in Fig. 18.12 with all the hydrogen atoms eliminated from the structure for simplicity. Polypropylene sold for commercial applications is almost 100% the isotactic grade. Isotactic polypropylene crystallizes (discussed later) easily to provide its property profile. The pendant methyl group orientation in syndiotactic and atactic polypropylene hinders crystallization, resulting in lower property profiles. Commercial polystyrene has the atactic configuration. The pendant benzyl groups off the ethylene backbone create steric hindrance that prevents crystallization. Polystyrene is amorphous with an atactic structure and has outstanding transparency.





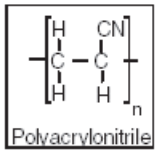


Figure 18.11. Structures of some common polymers.

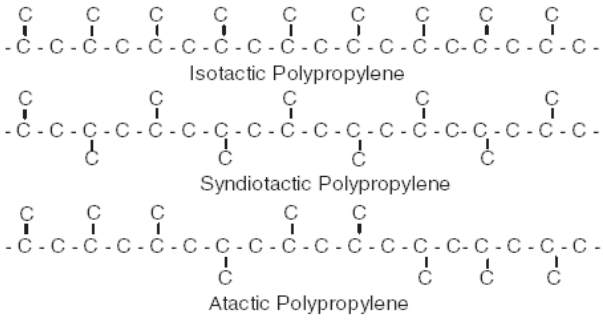


Figure 18.12. Various structural configurations of polypropylene and other polymers.

18.4 Molecular Weight

Each chemical element has a specific weight called the atomic mass unit. The atomic mass unit, or a.m.u., is an arbitrarily defined unit of mass using carbon as a reference and was created by scientists for measuring the masses of atoms and molecules. Adding the atomic weights of the atoms in an individual polymer chain gives a value called the molecular weight. Generally, the greater the polymer molecular weight, the higher the tensile, flexural, and impact properties become, while the resistance to flow or viscosity increases, making the polymers more difficult to process.

Each element has a specific atomic weight that is given in the chemical Periodic Table of Elements. Typical atomic weights found in polymers are

- Hydrogen – 1
- Carbon – 12
- Nitrogen – 14
- Oxygen – 16
- Fluorine – 19
- Sulfur – 32
- Chlorine – 35.5
- Bromine – 80

Molecular weight calculations for the small molecules discussed earlier are

- Methane, CH_4, = 16. Carbon a.m.u. is 12 and hydrogen 1. so the total = 12 + 4(1) = 16.
- Ethane, C_2H_6, = 30. Two carbon atoms plus 6 hydrogen atoms is 2(12) + 6(1) = 30.
- Butane, C_4H_{10}, = 58. 4(12) + 10(1) = 58.

As the molecular weight increases, the molecules at room temperature go from gas to liquid and finally solid. Each small molecule discussed above has a distinctive molecular weight that is unique to that molecule. Pentane, C_5H_{12}, will always have a molecular weight of 72. This is true as long as special isotopes are not present. Isotopes are atoms of the same element that vary in the number of neutrons in their nuclei.

Polymer molecular weights, unlike small molecules, do not have a unique value; instead, polymers have a molecular weight distribution that relates to the different polymer chain lengths. Polymerization is a random process, so each polymer chain does not end up with the same number of atoms. The molecular weight is a range that is presented as a molecular weight distribution. Figure 18.13 shows a typical molecular weight distribution curve. There are both short and long polymer lengths, with an average molecular weight near the center of the distribution curve. Plotted on the y-axis is the number of polymer chains with a given molecular weight, while the x-axis is the individual molecular weight. Molecular weight distributions can have different shapes, depending on the chain lengths within the polymer. A normal distribution is the bell-shaped curve shown as curve A in Fig. 18.13. In a normal distribution there are as many long high-molecular-weight chains as there are short low-molecular-weight chains. And the majority has a common molecular weight range in the center of the graph. Within a normal molecular weight distribution, it is possible to have a narrow molecular weight distribution, curve B, or a broad molecular weight distribution, curve C in Fig. 18.13. Other molecular weight distribution curves, shown in Fig. 18.14, can be skewed toward low or high molecular weight, curves D and E, or bimodal, with two molecular ranges, curve F. Polymers associated with the curves in Figs. 18.13 and 18.14 process differently because the molecular weight distribution and average molecular weights differ. If all the materials had the same average molecular weight, they would process differently because the molecular weight distribution curves aren't the same.

The polymer molecular weight and the molecular weight distribution directly affect the extrusion process. Comparing two polymers with different molecular weights, the polymer with the higher molecular weight

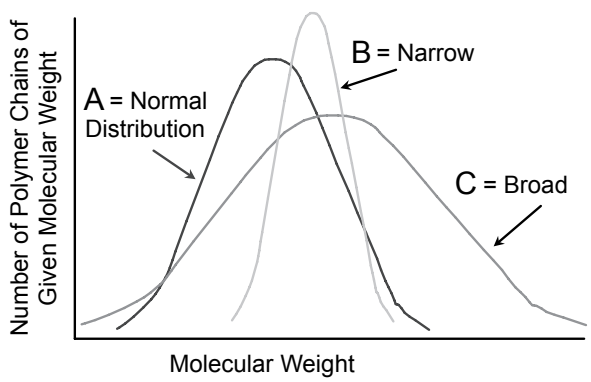


Figure 18.13. Normal molecular weight distribution curves.

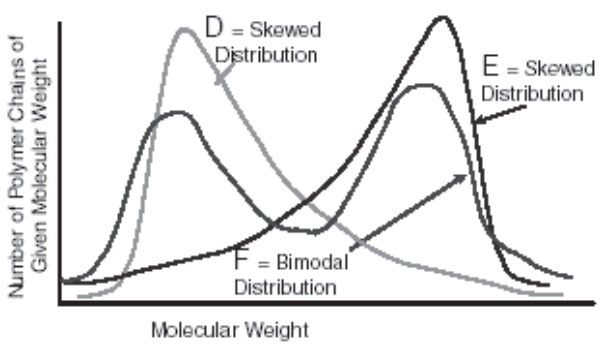


Figure 18.14. Skewed and bimodal molecular weight distribution curves.

will require more energy to process, i.e., higher drive torque or amps due to higher melt temperature, have better melt strength, be stiffer, and possess better ductility. Conversely, lower molecular weight has lower viscosity at a given temperature (flows more easily), poorer melt strength, potential handling difficulties as low viscosity polymer sticks more to gloves and other handling tools in the string up procedure, and requires less energy for the extrusion process. Extrusion processes normally use higher-molecular-weight polymers for better melt strength and melt handling characteristics, while injection molding tends to use lower-molecular-weight resins that flow more easily into thin sections of parts.

One molecular weight measure is the melt flow index (MFI), which measures how much polymer flows through a specified die orifice in 10 minutes. Figure 18.15 shows the test equipment for measuring melt flow index. The piston is removed and resin is added to the heated barrel, where it is compacted and allowed to come to equilibrium temperature. After a few minutes the weight is added to the piston, and resin flows out the orifice at the barrel bottom. The grams extruded out the barrel in 10 minutes with an orifice, barrel temperature, and weight specified by ASTM (American Society for Testing and Materials) test method D-1238 for a particular polymer resin are the resin MFI. The higher the MFI, the lower the resin viscosity and molecular weight. Low MFI correlates with high viscosity (less resin extruded through the orifice during the 10 minute test) and high-molecular-weight resins. Actual molecular weight and molecular weight curves are generated by gel permeation chromatography, also known as GPC.

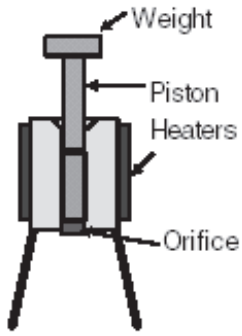


Figure 18.15. Measuring melt flow index.

Referring back to the molecular weight distribution curves shown in Figs. 18.13 and 18.14, answer the following processing questions (assume the molecular weight values are equivalent at the origin of each graph and the x-axis scale is similar):

- Identify the polymer(s) with the highest MFI value.
- Identify the polymer(s) with the lowest MFI value.
- Identify the polymer that is easiest to extrude and explain why.
- Identify the polymer with the broadest processing window.
- Identify the polymer with the smallest processing window.
- Identify the polymer that is most likely to contain gels. (Gels are defined as high-molecular-weight particles that pass through the extruder without melting.)
- Identify the polymer(s) you want purchased for processing in your plant and explain why.

The polymer with the highest MFI value is the material with the lowest average molecular weight, which is probably curve A in Fig. 18.13. A case could be made for curve D in Fig. 18.14 due to the high level of low-molecular-weight polymer; however, curve D has a long high-molecular-weight tail that will retard the flow through the MFI barrel in 10 minutes. The polymer with the lowest MFI is sample E in Figure 18.14, which appears to have the highest average molecular weight plus the highest concentration of high-molecular-weight polymer.

The easiest resins to extrude are polymers A and B, as both have relatively small and uniform molecular weight distribution. Both polymers will melt in a relatively small temperature range because the molecular weights of all the polymer chains in each sample are similar. While both resins A and B are easy to process, polymer B, with the narrowest molecular weight distribution, is probably the material of choice. All polymer molecules will melt at approximately the same melt temperature. There is not a significant amount of low-molecular-weight resin that is apt to melt early in the feed zone of the extruder and coat the barrel walls, causing succeeding feed material to slip on the barrel walls, leading to poor feed characteristics. Likewise, there is not a large amount of high-molecular-weight particles that are hard to melt and may be unmelted at the end of the transition zone in a single screw extruder. High-molecular-weight particles require more energy to be put into the melting section in a twin screw extruder to ensure all melting is complete.

The polymers with the largest processing windows are polymers A and B, assuming the largest processing window is defined as the widest temperature range over which all the polymer molecules can be uniformly melted. While A and B have the narrowest molecular weight distributions, both materials can be processed in a small range of melt temperature due to the narrow weight distribution. However, if both materials have good thermal stability, they can also be processed at a higher melt temperature range with assurance that all the material will melt and process uniformly because of the small molecular weight distribution. Therefore, these two polymers have a large temperature processing window where all the polymer is anticipated to either melt or soften uniformly.

The smallest processing window is difficult to determine from polymer curves D, E, and F. All three materials will have a narrow processing window for different reasons. Consider the skewed molecular weight distribu-

tion curve for D first. A lower molecular weight requires less energy and heat in the extruder to melt and process the material. Selecting an appropriate temperature to process the low-molecular-weight material may not supply enough energy to melt the polymer in the high-molecular-weight tail. If a low temperature is set to prevent the low-molecular-weight material from melting prematurely in the feed section, it may be impossible to supply enough energy at an acceptable throughput rate to melt the high-molecular-weight polymer. In a single screw extruder, there may still be solid particles at the end of the transition zone, causing solid bed breakup and solid floating in the melt in the metering zone. Assuming the solid in the melt in the metering zone is high-molecular-weight material, it may be impossible to melt by the end of the extruder, clogging the screen pack. If a barrier screw is used for mixing, excessive heat generated in this section to melt any solid high-molecular-weight particles may cause lower-molecular-weight particles to degrade.

Consider the polymer E molecular weight distribution curve next. Extruder temperatures are set higher than in the case of sample D to ensure all the high molecular weight material is melted in the transition section of a single screw extruder or the melting section of a twin screw extruder. However, premature melting due to the higher barrel temperatures in the feed area may lead to feed problems caused by surging, or in the worse scenario, a melt bridge. A low-viscosity melt film acts as a lubricant in the feed section, allowing the feed material to slip on the barrel walls. A second potential problem is the low-molecular-weight material acts as a lubricant at the barrel wall interface in the transition section, reducing the shear heating in the melting film region. This can interfere with the melting mechanism of the high-molecular-weight material, preventing all the resin from being melted in the transition section, leading to solid bed breakup and high-molecular-weight particles in the melt in the metering section. Finally, and maybe the most important consideration, does the heat required to melt the high-molecular-weight fraction cause the low-molecular-weight material fraction to degrade into even shorter polymer segments that can affect physical properties, i.e., tensile, flexural and impact?

Curve F, with the bimodal molecular weight distribution, is probably the worst scenario. At low temperatures the high-molecular-weight fraction does not melt in the transition section and is carried along as a filler particle floating in a molten resin matrix. At higher processing temperatures, premature low-molecular-weight melting in the feed section can cause feed problems due to slippage on the barrel wall. In addition, a low-viscosity material in the melt region may not allow as much shear heat to be generated, reducing the melt ing rate of the high-molecular-weight material. The barrel temperatures and screw speeds have to be set correctly in a small processing range to balance all effects and ensure there is sufficient heat to melt all materials, while not supplying so much heat that feeding and other issues are created.

The polymers most likely to have gels are those represented by molecular weight distribution curves D and F. Gels are high-molecular-weight particles that do not melt during processing, due to their high molecular weight or cross-linked particles. Molecular weight distribution curve D, with the majority of the polymer being low molecular weight and requiring lower processing temperatures, is likely to have some high-molecular-weight particles in the long high-molecular-weight tail that do not melt. Curve F, a bimodal distribution, is a likely candidate for gels from the small high-molecular-weight hump and tail in the higher molecular weights.

Of the six materials represented in curves A through F in Figs. 18.13 and 18.14, the most desirable resins for extrusion processing are molecular weight distribution curves A and B. They have the smallest molecular weight distribution, so all of the polymer chains will melt or soften over a small temperature range. If the polymers are thermally stable, it is relatively easy to have a very broad temperature range over which these materials can be extruded.

Not enough attention is paid to the molecular weight and molecular weight distribution of polymers being processed during extrusion. If a material with a molecular weight distribution curve similar to E in Fig. 18.14 is being processed, it is easy to envision supplying too much heat to the low-molecular-weight chains, causing them to degrade and become even shorter, thus affecting the physical properties.

A common practice in extrusion operations is to regrind out of specification material and add it back to the extrusion process with the virgin resin to reduce the raw material cost and reduce issues associated with scrap disposal. Does the previous heat history change the molecular weight distribution curve? Assuming some thermal degradation occurs during the first or second heat history experienced by the resin, the molecular weight distribution curve is now shifted toward lower molecular weight. Adding regrind that is partially degraded may create a bimodal molecular weight distribution for processing. The blend has the potential problems discussed with processing material represented by curve F in Fig. 18.14. In addition to a smaller processing window, the regrind or lower-molecular-weight material melts first, coating the barrel wall and preventing the high-molecular-weight material from processing correctly. Add to this the effect of feeding and melting different size regrind particles, and serious processing issues can result.

Blending two very different melt flow resins results in a bimodal molecular weight distribution. Processing conditions must be properly balanced to melt both the low- and high-molecular-weight portions efficiently, while preventing any feed problems associated with premature melting of the low-molecular-weight fraction or not melting the higher-molecular-weight resin in the transition section of a single screw extruder. The same problems are not as serious in twin screw extruders, where a melt seal can be formed to maximize melting in one section followed by efficient dispersive mixing to uniformly mix the high- and low-molecular-weight entities.

A problem is what data to use to compare the molecular weights of two different resins. The MFI does not describe the molecular weight distribution; it supplies average resin flow at very low shear rate. (Shear rate is at least 10-fold below the shear rates experienced in extrusion.) End-group analysis, boiling point elevation, freezing point depression, vapor pressure, and osmotic pressure are techniques used to measure the number average molecular weight. Number average molecular weight is a direct average of the molecular weights, where the molecular weight of each chain is added together to obtain a total molecular weight, which is divided by the number of polymer chains. Using a normal molecular weight distribution curve, the number average is the center of the bell shaped curve, as shown in Fig. 18.16.

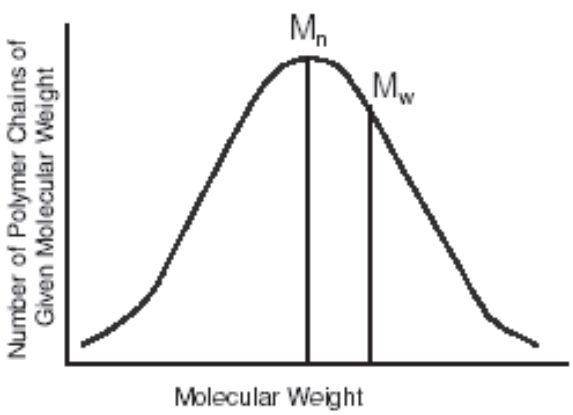


Figure 18.16. Normal molecular weight distribution curves.

The number average molecular weight is calculated from Eq. (18.1):

$$M_n = \frac{\sum n_i (MW_i)}{\sum n_i} \qquad \text{Eq. (18.1)}$$

where M_n = Number average molecular weight
MW_i = Molecular weight of chain i
n_i = Number of polymer chains of molecular weight MW_i

The weight average molecular weight is weighted toward higher-molecular-weight polymer chains. The weight average molecular weight is always higher than the number average molecular weight and is measured by gel permeation chromatography or light scattering techniques. The weight average molecular weight is given by Eq. (18.2):

$$M_w = \frac{\sum w_i (MW_i)}{\sum w_i} \qquad \text{Eq. (18.2)}$$

where $w_i = n_i(MW)_i$.

The degree of polymerization, given by Eq. (18.3), is the average number of monomer units per polymer molecule:

$$DP = \frac{M_n}{MW_{mer}} \qquad \text{Eq. (18.3)}$$

where

DP = Degree of polymerization
MW_{mer} = Molecular weight of the monomer

Generally polymers with acceptable physical properties have a DP greater than 500, with most commercial resins having a DP in the range of 1000 to 10,000.

Polydispersity index (PI) is the weight average-to-number average molecular weight ratio and is given by Eq. (18.4):

$$PI = \frac{M_w}{M_n} \qquad \text{Eq. (18.4)}$$

Usually polydispersity is between 1.5 and 2. However, higher polydispersity, between 3 and 8, is quite common. If the polydispersity is not between 1.5 and 8, the molecular weight distribution is unusual.

Another viscosity measure is intrinsic viscosity (IV), which is commonly used with PET, PBT, and nylon. Intrinsic viscosity is directly proportional to the polymer viscosity and correlates with higher molecular weight and typically higher physical properties. Intrinsic viscosities are solution-based viscosities and are normally supplied by resin manufacturers.

To ultimately understand resin viscosity and how a particular resin performs in either extrusion or injection molding, it is necessary to compare shear rate versus viscosities curves. These will be discussed in more detail in Chapter 20.

The physical properties to molecular weight and processability relationship is shown in Fig. 18.17. As the molecular weight increases, the processability decreases and the property performance increases. Polymers of low molecular weight have low viscosity and are easy

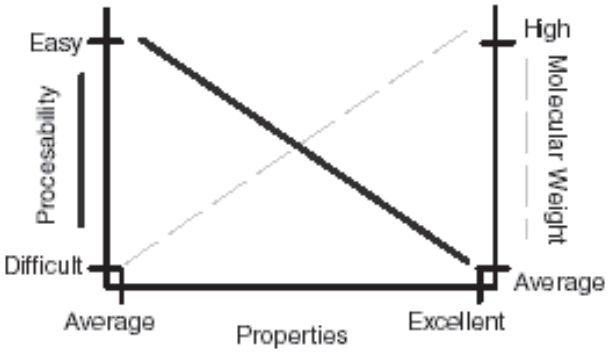


Figure 18.17. Comparison of processing, properties, and molecular weight.

to shape, but may have inferior properties and end-use performance.

18.5 Copolymers

Copolymers are formed when two monomers are polymerized together to form a polymer. Terpolymers are formed when three monomers are used. Copolymers are classified as random or block copolymers, depending on whether the repeating groups in the backbone are randomly arranged or are in segments. Assume copolymer is produced from monomers A and B. Figure 18.18 shows random and block copolymer backbones. Alternating copolymers have every other monomer changing identity, as –A–B–A–B–, etc. Block copolymers are categorized as diblocks, shown in Fig. 18.18, where alternating monomer A and B blocks occur, or triblocks, where there is a block of A followed by a block of B and then another block of A.

Block copolymers are formed by polymerizing oligomers, which are short chains. Oligomer polymer chains by themselves are too short and have insufficient molecular weight to have acceptable properties. These short chains are later polymerized to form high-molecular-weight block copolymers from small blocks of A and B oligomers.

Copolymers are produced to combine the optimum resin properties of at least two different monomers, generating a new resin with superior properties (in some specific area) relative to the individual components. Styrene and butadiene are combined as a hard and soft segment, respectively, imparting elastomeric-type properties to a copolymer that contains hard segments to yield higher modulus and stiffness. Ethylene

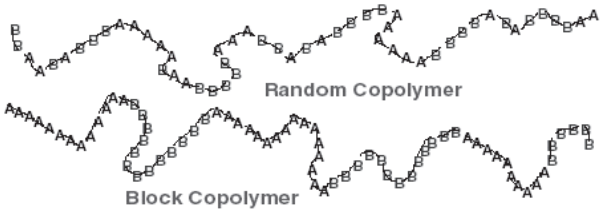


Figure 18.18. Schematic of random and block polymers.

vinyl acetate (EVA) combines ethylene monomers with ethylene vinyl acetate monomers. The copolymer has good adhesion to both polar (resin systems containing atoms other than carbon or hydrogen) and nonpolar resin systems (resins containing only carbon or hydrogen). EVA is used in extrusion as a tie layer or adhesive between different substrates in coextrusion, film coatings, or lamination. Copolymers are used as compatibilizers when blending two dissimilar resin matrices. The copolymer provides an interface to enhance adhesion and mechanical compatibility of the two different phases.

Some commercial copolymers are shown in Fig. 18.19.

18.6 Polymer Structures

Long chain polymer molecules distributed in a polymer matrix were discussed previously and shown in Fig. 18.8. The long chains are like wet spaghetti strands

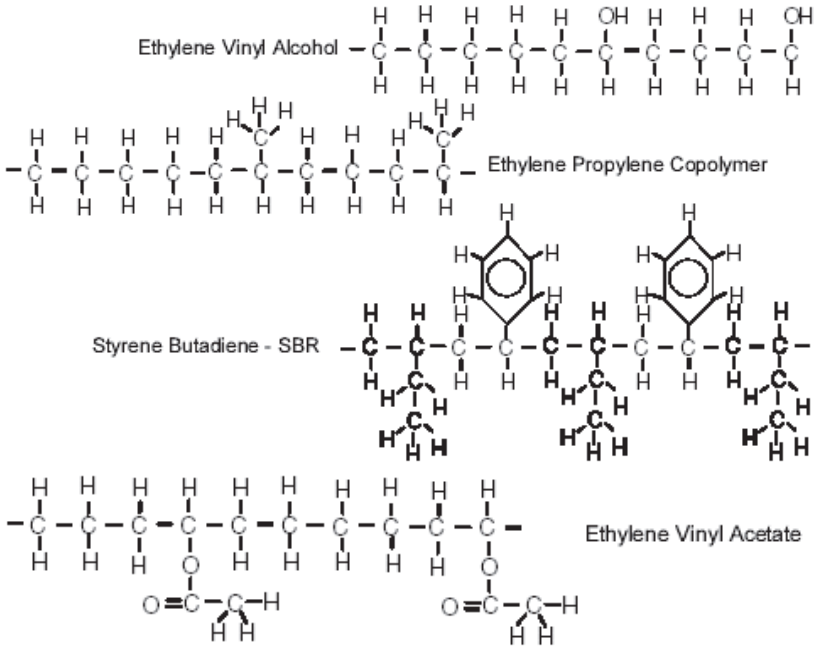


Figure 18.19. Schematic of random and block polymers.

entangled with each other to impart significant properties to the polymer matrix. As the polymer backbone is modified with atoms that provide steric hindrance, the backbone becomes stiffer, with less molecular motion (rotation) about the carbon bonds. Stiffer polymer backbones generally have higher property performance, as measured by tensile, flexural, impact, and heat properties. Polymers discussed to this point have been based on ethylene chemistry, with the C–C double bond being broken to form free radicals followed by polymerization through free radical addition. Other chemical reactions (i.e., condensation and grafting) are used to generate polymers with different chemical backbones. Some polymers with stiffer backbones and higher properties are shown in Fig. 18.20.

Cross-linked structures occurring in thermoset materials were discussed previously and shown in Fig. 18.9. A chemical bond occurs between different polymer chains, locking in the polymer structure. These chemical bonds prevent a cross-linked polymer from melting or softening. Typically, two or more components are mixed together at room temperature and placed in a mold at high temperature where the reaction is accelerated to produce a final part, which is removed from the tool or mold at the elevated temperature. Cycle times are determined by the time required for the chemical reaction to be completed.

Some thermoplastic polymers have side chain branches. These branches alter the resin processing and physical properties. Figure 18.21 shows side chain branching in a polyethylene molecule. The major difference between low density polyethylene (LDPE), high density polyethylene (HDPE), and linear low density polyethylene (LLDPE) is the type and degree of side chain branching. Generally polyethylene has 1 to 10 branches for each 100 carbon atoms along the polymer backbone. Low density has very frequent, irregular-length chain branching. HDPE has longer chain branches that occur less frequently compared to LDPE. The branching in LLDPE is controlled by the type and con-

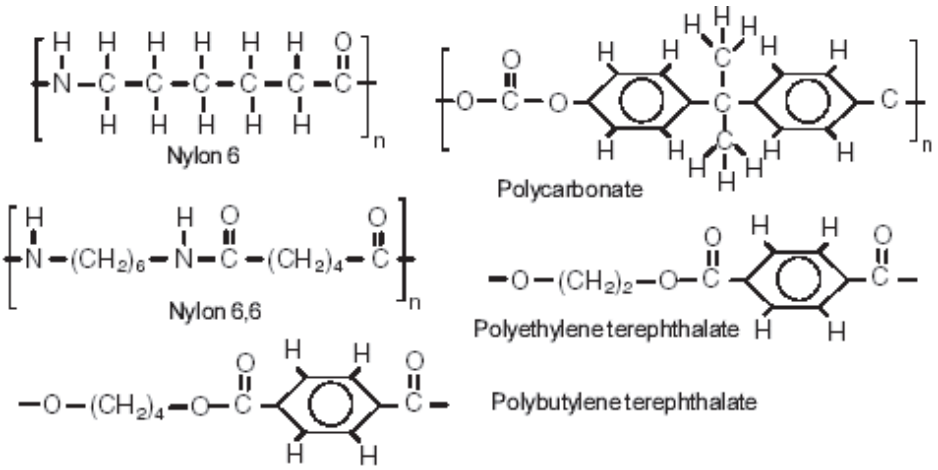


Figure 18.20. Chemical composition of some engineering themoplastics.

Figure 18.21. Branching off a polyethylene.

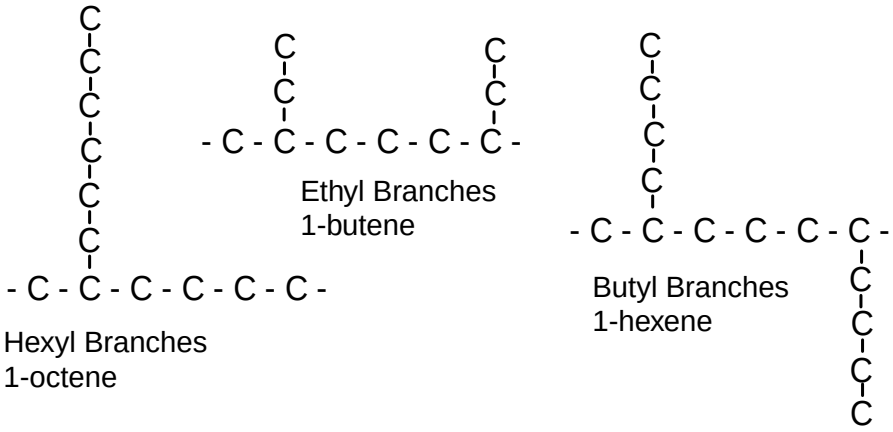


Figure 18.22. Different types of branching in LLDPE depending on the monomers used in the polymerization.

centration of comonomers added during the polymerization reaction. Using 1-octene as a comonomer generates hexyl branches along the polymer backbone, 1-hexene results in butyl branches, and 1-butene generates ethyl branches. The various comonomers are shown in Fig. 18.22.

Comparing branched and linear polymers in extrusion processing, the branched polymers require higher energy to process, are stiffer, have higher viscosity and lower MFI, better melt strength, and better handling characteristics. Branched resins are particularly useful in extrusion blow molding, where high polymer melt strength is critical to support a large parison. The melt strength is directly proportional to the degree of branching.

The last type of polymer structure is a network, where one polymer is interspersed with another and the cross-links between the polymer chains actually cross one another. Figure 18.23 shows a network polymer matrix.

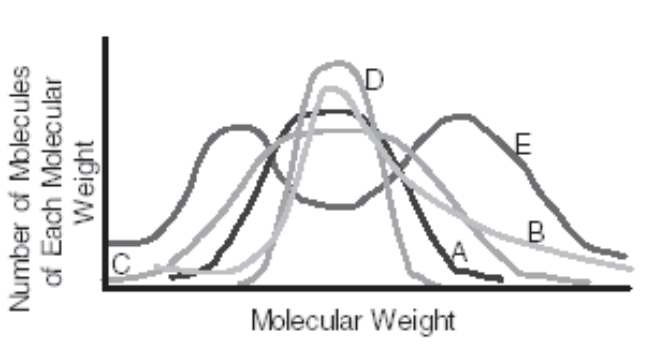
Figure 18.23. Network polymer structure.

Review Questions

1. What is a polymer?
2. What is the importance of polymer molecular weight?
3. What causes molecular orientation and how does this affect physical properties?
4. Why is rotation about carbon atoms important for physical properties?
5. How are typical molecular chains in a polymer matrix represented?
6. What is the difference between thermoplastic and thermoset polymers?
7. Are elastomers thermoplastic or thermoset polymers? Explain.
8. Can thermoset plastics be extruded, and if so, what cautions must be taken during extrusion?
9. What is cross-linking?
10. What gives polymers their strength and properties?
11. What is a copolymer?
12. How are polyethylene and polystyrene formed by free radical reactions?
13. What is the importance of polymer molecular weight and how does it affect processing?
14. What is MFI, how is it measured, what does it measure, and what is the difference between a 12 MFI resin and a 35 MFI resin?
15. How does viscosity relate to molecular weight? In extrusion, is a high or low viscosity better and why?

Review Questions (continued)

16. Using the molecular weight distribution curves given below, answer the following questions with explanations:
    a. Lowest molecular weight?
    b. Highest viscosity?
    c. Smallest processing window?
    d. Largest processing window?
    e. Most likely to contain gels?
    f. Most difficult to process?
    g. Lowest melt flow index?
    h. Highest melt flow index?



17. Why might the addition of regrind interfere with the extrusion of virgin resin?
18. What is the difference between weight average and number average molecular weight?
19. What is the degree of polymerization?
20. What is polydispersity index?
21. What is the difference between a block and a random copolymer?
22. What is the difference in branching on HDPE, LDPE, and LLDPE?
23. Why is it important to understand polymer behavior and polymers in general to troubleshoot extrusion problems?

Polymers in the solid state have either a semicrystalline or amorphous structure. The resin structure affects both the physical properties and extrusion processing. Amorphous polymers are completely random polymer chains in the solid state, while semicrystalline polymers have both random and ordered polymer chains in the solid state.

19.1 Amorphous Polymers

Figure 18.8 in Chapter 18 represents amorphous polymer molecules. Polymer chains are randomly arranged with no order or common association either within or between adjacent polymer chains. The chains are randomly distributed through the polymer matrix with no relationship to adjacent polymer chains. Amorphous polymers do not have a distinct melting point; instead, the polymers move from a glassy state to a rubbery state as heat is applied until they soften enough to flow when a force is applied.

The reversible transition from a glassy, hard, and brittle state to a rubbery, elastomeric state occurs at a unique temperature for each polymer, defined as the glass transition temperature, T_g. Typical molecular motion includes rotation and vibration about the different bonds in a polymer molecule. In the solid, glassy state, the polymer chains are fixed in a given spatial configuration with no rotation about the polymer backbone. (Rotation about a bond in the polymer backbone requires pendant groups to move within the solid matrix.) Molecular chains in the glassy state are coiled, entangled, and motionless within the polymer matrix. Minor vibrations about the carbon atoms and pendant hydrogen atom rotations about their axes do occur, and this does not disrupt the glassy polymer matrix. Below T_g all the atoms are reasonably fixed in their geometric spatial arrangement relative to other atoms, with no large-scale polymer chain motion. Figure 19.1 shows a propane molecule with some potential molecular motions. Assume this picture represents a part of a much larger polymer chain. Rotation about the C–C bond between atoms C_1 and C_2 rotates molecule C_3 either into or out of the page. (Page is that which has the two-dimensional representation of the three-dimensional molecule.) Attaching hundreds or thousands of other carbon atoms onto either C_1 or C_2 has a dramatic effect on the molecular structure of the polymer chain in the event a rotation about the C_1–C_2 bond occurs. Increasing the polymer temperature leads to more rotations about C–C bonds and substantial polymer backbone movement within the rubbery state. Properties reflect this increased molecular motion, as the polymers are tougher and more flexible. Rotations about the carbon–hydrogen bonds are possible with only minimal effect on the polymer chain. Some rotation about the carbon–hydrogen bonds occurs below T_g.

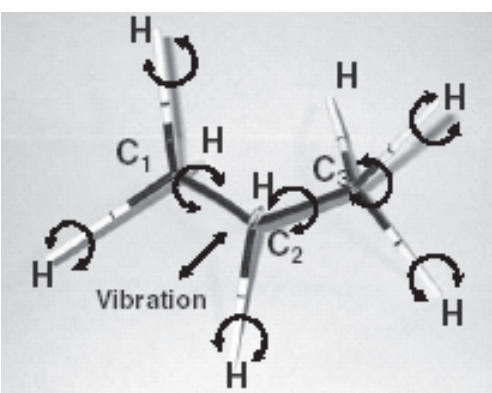


Figure 19.1. Typical polymer rotations and vibrations shown on a small molecule.

The other molecular motion shown in Fig. 19.1 is vibrational motion, demonstrated by the straight arrow. Carbon and hydrogen atoms can vibrate back and forth in any direction about the axis. This is the principal carbon and hydrogen atomic motion below the glass transition temperature. Above the glass transition temperature, entire segments of carbon atoms start to vibrate. As more heat energy is applied to the system, vibrations become more and more rapid and the structure expands. As heat is added and the matrix expands, the polymer molecules move farther apart, allowing more molecular motion and the onset of polymer flow as a force is applied to the rubbery molecules.

In the glassy state, the polymer is brittle and behaves more like a glass. Above T_g the polymer behaves more like a rubbery or elastomeric material. Adding heat to the matrix increases the molecular vibrations and finally rotation about the C–C bond. As the matrix expands from the thermal energy, the viscosity decreases until the resin flows. Figure 19.2 shows a cooling curve for an amorphous material and the associated volume change in the resin matrix. In the polymer melt, the amorphous polymer has softened enough to flow when force is applied.

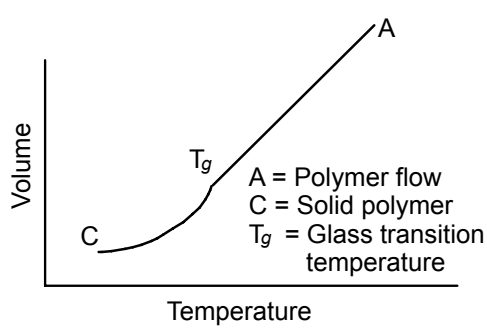


Figure 19.2. Cooling curve.

By adding heat, the polymer matrix expands to the volume at point A in Fig. 19.2. Removing heat contracts the entire matrix. This contraction is associated with a reduction in atomic molecular motion until T_g is reached. At T_g the volume continues to shrink, as there are fewer vibration movements and no rotation. At temperature C, most vibrations along the backbone have ceased and there is mainly only a little rotational and vibration motion associated with the hydrogen atoms.

Above T_g, an amorphous polymer, in theory, can be processed by applying sufficient force to make the resin flow. In practice, the force necessary to move the polymer is too high for any effective processing without adding more heat. T_g is critical in cooling an amorphous polymer outside the extruder. Once the polymer temperature is below T_g and the glass state is attained, the polymer dimensions are fixed and the extrudate can be removed from any sizing equipment. In amorphous polymer processing, cooling the extrudate below its T_g assures reproducible product dimensions. Depending on the cooling rate and the part thickness, a solid, glassy skin can be formed with a molten polymer (polymer above T_g) in the center. While the outside dimensions are fixed because the part surface is below T_g, the inside of a thick part may still be above T_g, with high rates of molecular motion and shrinkage still taking place. If the skin gets too thick without sufficient center cooling in thick parts, the center cools later and shrinks away from itself, creating voids or holes. These are not caused by trapped air or gases; they are caused by shrinkage. Once the outside is solidified and the center continues to shrink, it cannot draw the outside surface tighter. Instead, as the inside material shrinks, a void is formed. Figure 19.3 shows a shrinkage (vacuum) void inside a polycarbonate strand. As polycarbonate is transparent, the voids are readily visible. However, in an opaque part, the voids are not visible and part failure can occur due to the voids. Voids are normally formed in the thickest section. During extrusion, the rate and temperatures used to cool the entire part are critical to obtaining optimum performance. Parts made with amorphous resins need to be cooled below their T_g. The cooling rate is critical in eliminating vacuum voids. Slow cooling that allows the center to cool and solidify at the same rate as the surface eliminates or minimizes void formation; however, in actual operation, cooling this slowly is not always practical.

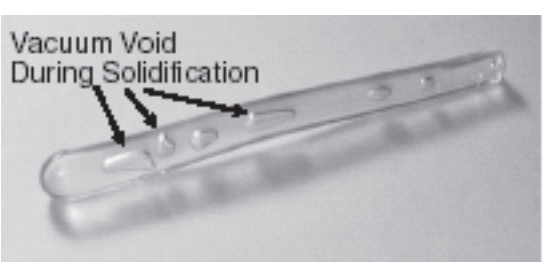


Figure 19.3. Vacuum voids.

Orientating amorphous polymers by drawing the extrudate out the die, drawing in a sizing fixture, or downstream drawing with heat has a similar effect on the polymer molecules. Figure 19.4 shows amorphous polymer molecules before and after molecular orientation.

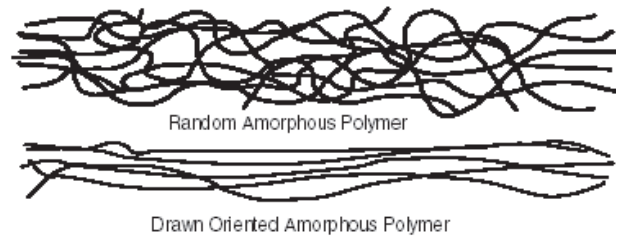


Figure 19.4. Comparison of oriented and random unoriented amorphous polymer.

Chain alignment in oriented structures leads to high tensile and flexural properties in the orientation direction and lower properties in the transverse or perpendicular direction. Stretching the chains in a tensile- or flexural-type test requires chemical bonds to be broken to rupture the chains. In the transverse or cross-machine direction, very few bonds have to be broken, as the polymer chains are simply pulled apart. Going back to the string analogy in Chapter 18, Fig. 18.4, consider strings laid parallel in one direction and the energy required to break them when pulling on the ends. Lay the strings back down and pull on them in the transverse or perpendicular direction. The energy required to separate the strings is very low compared to the energy required in the perpendicular direction to break them.

Molecular orientation allows oriented samples to be more easily slit in the orientation direction, while slitting across the polymer chains is more difficult. Extruded part failure can occur because molecular orientation is too high. When doing failure analysis on extruded parts, determine the crack propagation direction. If cracks propagate easily in the machine direction, failure may be caused by too much molecular orientation generated during the drawing operation. Similar strength in all directions comes from completely random polymer molecules, similar to cooked spaghetti strands in a bowl.

Amorphous polymers include the following resins: polyvinyl chloride (PVC), polystyrene (PS), acrylic, cellulosics, acrylonitrile butadiene styrene (ABS), polycarbonate (PC), polyetherimide (PEI), polysulfone (PS), polymethyl methacrylate (PMMA), polyphenylene oxide and polystyrene blends, styrene acrylonitrile (SAN), polyether sulfone (PES), and polyether ether ketone (PEEK).

Properties and processing amorphous polymers versus semicrystalline polymers will be discussed after the semicrystalline polymers section.

19.2 Semicrystalline Polymers

Unlike amorphous polymers, semicrystalline polymers have some regions where the polymer chains are arranged in specific spatial patterns relative to other polymer chains within their polymer matrix. Segments of polymer backbones, from either adjacent polymer molecules or within the same polymer molecule (polymer chain folded back on itself) are aligned in an orderly manner with the exact same spatial arrangement and inter- or intramolecular distance from one set or group of atoms to the next. Semicrystalline polymers contain both amorphous and crystalline regions within the same polymer matrix. The crystalline phase has a very distinct melting point or temperature to which the polymer has to be heated before it can be processed in an extruder. The amorphous phase is similar to that discussed previously, with a completely random structure and a T_g where the onset of molecular motion occurs. Molecular motion does not occur to the same degree in the crystalline section due to the inter- and/or intramolecular forces holding the ordered groups of atoms and molecules together.

Figure 19.5 shows a semicrystalline polymer matrix. The ordered crystalline areas are circled for identification; other areas with the random chains are amorphous.

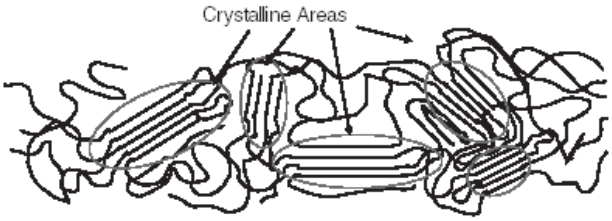


Figure 19.5. Semicrystalline polymer matrix.

The crystalline areas melt at the melting point and the polymer, in the molten state, is the same as an amorphous polymer. Once the polymer is cooled below its melting point, the polymer again crystallizes to form a semicrystalline matrix. Semicrystalline is used interchangeably in plastics jargon with crystalline polymer. In all cases, crystalline polymers contain both crystalline and amorphous regions, with all crystalline polymers being semicrystalline in the solid state. The crystallinity and the crystalline region sizes vary, depending on the polymer and the processing conditions. Some polymers crystallize very rapidly below the melting point and others crystallize very slowly, requiring long times and possibly a post-crystallization step to reach their equilibrium crystallization levels.

Different crystalline polymers crystallize at different rates and different degrees. Consider the two crystalline polymers, PBT and PET. PBT crystallizes very readily after exiting the extruder. PET, with no additives other than stabilizers, is transparent and clear as it exits the extruder and is cooled. PET directly from the extruder is in its amorphous state, and there is virtually no crystallinity present, as witnessed by its transparent nature. To crystallize PET, it is heated above its glass transition temperature. Initially, PET can be heated to 212˚F (100˚C) for two hours, followed by progressively increasing the temperature until the polymer is crystallized. The crystallization rate depends on the product thickness and additives (nucleation ingredients) in the polymer. If the resin is heated too hot initially, while it is in the amorphous state, it will become tacky (above T_g) and stick together. The product is transformed from a colorless transparent material to an opaque crystalline product through this procedure. If the product is directly heated to 266–302˚F (130–150˚C) in the amorphous state, the pellets stick together and form a solid mass. At 150˚C, PET is above its T_g and the amorphous region softens and becomes sticky, encouraging the pellets to adhere to each other and form a large solid mass. PET must be crystallized before it is sold due to its hygroscopic nature and propensity to absorb moisture. To process PET by either extrusion or injection molding, the moisture content has to be below 0.02% to prevent hydrolysis during processing. Drying at 150˚C in a dehumidifying desiccant dryer for four hours is required to obtain 0.02% moisture. If the product is not precrystallized, a solid mass is formed in the dryer.

The crystal size and quantity formed in rapidly crystallizing polymers such as PP, PBT, PA, PE, etc., depend on both the quench rate and quench temperature. Slow cooling promotes large crystal growth with maximum crystallinity. Rapid cooling promotes small crystal growth combined with minimum crystallinity. Above the glass transition temperature, where the amorphous region has molecular motion, additional crystallization occurs until the polymer reaches its equilibrium crystal content. Polymers such as PE and PP are above their glass transition temperatures at room temperature; consequently, even at room temperature these materials will continue to crystallize slowly until they reach equilibrium. Rapid cooling PE and PP generates smaller crystals compared to slow cooling. The equilibrium crystal content over time is generally equivalent. Polymer or part annealing, heating to allow the molecules to relax and rearrange, removes internal stresses. Annealing crystalline polymers at temperatures above their T_g promotes crystal growth and additional crystallization.

Drawing fibers to align the polymer chains promotes additional crystal growth. Since crystallization aligns polymer chains, drawing the polymer also aligns the molecular chains, encouraging crystal growth. A solid state drawing process is done below the melting point and aligns the amorphous chains with some rotation and reordering of the crystalline regions.

Since the crystalline regions have sections of molecular chains aligned and uniformly spaced and packed, the atoms in a crystalline domain take up less space than the atoms in the amorphous domain. Consequently, crystallization leads to higher shrinkage on cooling compared to cooling an amorphous matrix. A simple analogy is a pile of split logs randomly thrown into a pile. The random pile is larger and takes up more space than the same logs stacked neatly side by side. Neatly stacked logs simulate the crystalline region, while the randomly piled logs simulate the amorphous state. As a result, crystalline polymers exhibit higher shrinkage than amorphous polymers. A profile extrusion or finished part made from crystalline resin that is not completely crystallized in the process and is later heated, annealed, or used in a hot environment can undergo additional crystallization, resulting in the part dimensions decreasing. Thick parts taken off a continuous line in a relatively hot state and packed may be slightly smaller when they reach the customer due to additional crystallization occurring in the package. Slow cooling, combined with the fact that polymers are great insulators, holds the higher temperature for a long time, allowing additional crystallization to take place.

Film, sheet, or other extruded polypropylene products are transparent as they exit the extruder. On cooling the transparency disappears and the material turns opaque due to crystallization. Polyethylene blown film is a great example of crystallization. Exiting the die, the film is clear and transparent until it reaches the frost line, where it has cooled sufficiently for crystallization to take place. The frost line, or beginning of opacity, is caused by crystallization.

With amorphous products, it is recommended that the product be cooled below its T_g in any tooling fixture or quenching operation to hold dimension. Crystalline polymers solidify as soon as they are cooled below their melting point. Cooling generates a solid skin around the part with the center cooling after the outside surface; the time required for the center to cool and crystallize depends on the part thickness and the cooling medium. Once the skin is sufficiently thick, parts can be removed from the fixture or sizing device. It is important to remember that dimensions are constantly changing as the hot resin core continues to crystallize. Part dimensions continue to change until the amorphous regions are cooled below their T_g and the crystalline domains reach their equilibrium crystallinity. Figure 19.6 shows a cooling curve similar to Fig. 19.2 with the volume change for a semicrystalline material.

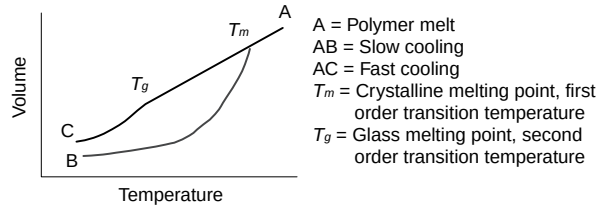


Figure 19.6. Volume change associated with cooling semicrystalline matrix.

Note that using a slow cooling approach ends up with a smaller volume part (curve AB) than rapid cooling (curve AC) due to additional crystallization induced by slow cooling. Annealing the part from point C by heating above T_g for an hour or two and cooling down to point C temperature again, the part volume is anticipated to be very similar to point B. Annealing the sample, it is conceivable that part C dimensions may be less than part B.

Examples of crystalline polymers are polyethylene, polypropylene, PBT, PET, most nylons, acetal, and liquid crystalline polymers.

19.3 Comparison of Semicrystalline and Amorphous Polymers

There are some unique differences between semicrystalline and amorphous polymers. Table 19.1 compares polymer characteristics for semicrystalline and amorphous polymers. The uniformly packed molecules present in semicrystalline polymers are not found in amorphous polymers; this is quite obvious from previous discussions.

Table 19.1. Comparison of Semicrystalline and Amorphous Polymers

Characteristics	Amorphous	Semicrystalline
Uniformly packed molecules	No	Yes
Randomly packed molecules	Yes	Yes
Sharp melting point	No	Yes
Clarity	Transparent	Opaque/ Translucent
Shrinkage	Low	High
Chemical Resistance	Poor	Good
Hardness	Soft	Hard
Energy to melt	Low	High

In extrusion or injection molding, semicrystalline polymers have a very sharp melting point, while amorphous polymers soften slowly above their T_g until the polymers start to flow. Starting up an extruder with a semicrystalline polymer in the barrel or die, it is necessary to heat soak the barrel to ensure the temperature is above the polymer melting point and the polymer is molten. If a crystalline polymer is left in a large die with poor heat transfer characteristics, it may require a long time to heat the die sufficiently to raise the resin temperature above its melting point. With crystalline resins extra precautions are required at start-up to ensure the die is not blocked with solid polymer. If solid polymer is in the die, the extruder head pressure builds up very rapidly and, in extreme cases, can blow the die off the machine or blow the rupture disk or cause polymer to flow back out the vent, if one exists. In some situations it is necessary to preheat the die prior to the extruder barrels because the die needs extra time to come to equilibrium. Amorphous materials do not have a sharp melting point; instead, the equipment can be started with the screw generating the shear heat required to soften the material sufficiently for processing. Generally, it is easier to start an extruder at low temperature with an amorphous material than with a semicrystalline resin.

Above the melting point, semicrystalline polymers flow very easily with low viscosity. Amorphous resin viscosity in general tends to be more temperature-sensitive, with the viscosity decreasing more rapidly as the resin temperature is increased. Close to their T_g, amorphous polymers in general are more apt to be very thick and require higher motor load to process.

The last category in Table 19.1 shows that crystalline resins require more energy to melt than amorphous materials. Assuming two polymers process at the same temperature and one is semicrystalline while the other is amorphous, the crystalline resin will require more energy to provide the additional heat required for the heat of fusion (ΔH_{fusion}) to melt the crystal phase. As semicrystalline resins are heated, energy is added until the melting point is reached. At the melting point additional heat must be added to melt the crystalline phase; this is called the heat of fusion.

Another characteristic shown in Table 19.1 is that amorphous polymers are transparent. This assumes there are no additives or other components that would destroy the resin transparency. Consider rigid or unplasticized PVC, which is opaque due to filler, colorants, and other components added to the formulation. Plasticized or flexible PVC is transparent in many applications. Another good example is crystal polystyrene, which is called crystal because of its clarity and similarity to glass crystal. Polystyrene clarity is the result of its amorphous structure. High impact polystyrene is also amorphous, and it is opaque because of the rubber particles added for impact modification. Light scattering from density differences between the crystal and amorphous regions and crystals that are larger than the wavelength of light makes semicrystalline polymers opaque or translucent. Light passing through the matrix is scattered by the crystalline regions, causing the resin to be opaque or translucent. As a crystalline material, such as PP, exits the extruder, it is transparent because the melt is amorphous. On cooling, the extrudate turns hazy due to crystallization.

The effect of polymer structure on shrinkage is related to the volume required to pack the polymer molecules. Amorphous molecules packed in a random configuration take up more volume than orderly, closely packed molecular chains. Consequently, shrinkage is greater in semicrystalline matrices. As amorphous regions in semicrystalline polymers continue to crystallize, the plastic part volume continues to decrease until the equilibrium crystallinity level is attained.

Amorphous polymers tend to be softer than semicrystalline polymers due to the closely packed molecules in the semicrystalline polymers. It is more difficult to move closely packed, ordered atoms than a loosely packed configuration where the polymer chains can spread or move when a force is applied, such as an indentation force like a Rockwell tester.

In general, semicrystalline polymers tend to be more resistant to chemical attack than amorphous polymers. Each polymer system has unique chemical resistance characteristics to various organic and inorganic solvents. While a certain amorphous polymer may be susceptible to chemical attack by solvent XYZ, that solvent may not attack a crystalline polymer. The orderly structures in semicrystalline resins in general are more thermodynamically stable and tighter packed, making chemical attack more difficult.

Density is weight per unit volume. Crystalline regions, due to their uniformly closely packed atoms, have higher densities than the amorphous regions in semicrystalline polymers. The final semicrystalline polymer density is a weighted average of the amorphous and crystalline regions.

Table 19.2 lists crystalline and amorphous polymers.

Table 19.2. Comparison of Test Results Measuring Polyethylene Crystallinity

Sample	Density	IR	NMR	X-Ray	Calorimetry
Sample 1	67	67	62		
Sample 2	74	80	82		
Sample 3		53		57	
Sample 4		72		74	
Sample 5			69		93
Sample 6			81		94

19.4 Crystallinity Measurement

Numerous methods are used to measure percent crystallinity, with different methods yielding different values. The key to comparing percent crystallinity is to ensure that all measurements being evaluated are made using the same measurement techniques. Table 19.2 compares polyethylene percent crystallinity as measured by different methods. Measurements were made using density gradient tube, infrared (IR) spectroscopy, nuclear magnetic resonance (NMR), X-ray crystallography, and calorimetry. Note that the values vary based on the measurement technique.

19.5 Polyethylene Crystallinity

The difference between high density polyethylene (HDPE) and low density polyethylene (LDPE) is their different crystallinity. In the last chapter, branching along the polyethylene backbone was discussed, with LDPE having more side chain branching and HDPE having fewer long chain branches. The higher side chain branching in LDPE prevents the polymer from crystallizing to the same level as HDPE. With more branching it is harder to pack the molecular segments as closely and uniformly together, resulting in lower crystallinity. The more molecules that can be packed into the same unit volume, the higher the density. With fewer longer chain branches, HDPE molecules and atoms pack closer together than LDPE, resulting in higher density. It is similar to taking a box and trying to fill it with straws. If the straws are randomly thrown in the box, simulating an amorphous structure, the box weighs less and the density is lower because there is less mass filling the same volume. Crystalline and amorphous regions are both required for the performance of polyethylene. Totally amorphous polyethylene is too soft and greaselike, making it useless as a polymer for extrusion. At the other extreme is 100% crystallinity, which is too hard and brittle. Highly crystalline HDPE with low levels of branching can have up to 70% crystallinity. Highly amorphous LDPE has approximately 40% crystallinity. The higher the crystallinity, the tighter the structure becomes and the less permeable the structure is to gases and moisture. Small crystalline regions obtained by rapid quenching have better clarity, ductility, properties, and stress crack resistance.

Linear low density polyethylene (LLDPE) is made with monomers that control the degree of side chain branching. The side chains are very short but very frequent, preventing the polymer chains from packing closely together in a crystalline structure.

Due to the level of crystallinity, the processing temperatures for HDPE and LDPE are slightly different. HDPE melts at 257–266˚F (125–130˚C), while the LDPE melting point is 230–240˚F (110–130˚C).

19.6 Polypropylene Crystallinity

In Chapter 18 commercial polypropylene was described as isotactic, where all the pendant methyl groups were on the same side of the polymer backbone. This allows PP to easily crystallize. In either the atactic or syndiotactic form, the pendant methyl groups disrupt the close uniform packing arrangement required for crystallization.

19.7 Polystyrene Crystallinity

Since commercial polystyrene is in the atactic form, the random benzyl pendant groups prevent polystyrene from crystallizing. However, recent catalyst developments have produced syndiotactic polystyrene, which is semicrystalline with a 518–543˚F (270–284˚C) melting point.

Review Questions

1. What is the difference in chemical structure between amorphous and semicrystalline polymers?
2. What is the glass transition temperature?
3. What happens to amorphous polymers above T_g?
4. What is the difference in starting up two extruders, where one is filled with nylon and the other filled with ABS, assuming they have both been off all night?
5. What is the difference in cooling an extruded profile of PBT and polystyrene?
6. What are possible methods of measuring density?
7. Explain the differences in semicrystalline and amorphous polymers based on the following parameters: shrinkage, chemical resistance, melting temperature, clarity, molecular packing, and energy required to extrude each type of resin.
8. Explain the differences between HDPE, LDPE, and LLDPE.
9. What causes vacuum voids in extruded parts?
10. Identify six amorphous polymers.
11. Name six semicrystalline polymers.
12. Explain how drawing a product out of the extruder or orientating a product downstream can assist or retard crystallinity development.
13. What are the effects of rapid versus slow cooling on crystallinity?

20 Polymer Rheology

Rheology is the science dealing with the deformation and flow of materials. For polymers, understanding the deformation and flow, both in the extruder and die, is critical to optimum operation of the extrusion process. In coextrusion, it is critical to match resin layer viscosities at processing temperature to eliminate interfacial instabilities that would make the product useless. This chapter covers different aspects of polymer rheology and its importance to extrusion processes.

Polymers, unlike water, oil, organic solvents, and most liquids encountered every day, are non-Newtonian fluids. Fluids by definition deform when a force is applied and continue to deform until the force is removed. In a Newtonian fluid, the rate of deformation is directly proportional to the force applied. Rheology deals with the relationships between stress (applied force), strain (deformation resulting from an applied force—elongation), and time.[1] As a force is applied to a Newtonian fluid, elongation occurs; when the force is removed, the fluid stays in that position until another force is applied. This is shown in Fig. 20.1, where there is no strain or elongation until a constant stress is applied. When the constant stress is removed, the elongation remains constant until the stress is reapplied, at which time the fluid again moves at a constant rate that is directly proportional to the stress applied. Consider, as an example, a drop of water that is pushed with your finger across a surface. The pushing is the stress or force applied, and the elongation or strain is the movement of the water drop from one location to another. As the stress is applied, the drop moves; when the stress is removed, the drop remains in its new location. If the force is applied a second time, the drop moves to a new location. Movement or elongation is dependent on the force applied and the time the force is acting on the drop. For elastic materials, when a force is applied, deformation occurs until the force is removed, whereupon the elastic material returns to its original configuration, assuming the material has not ruptured.

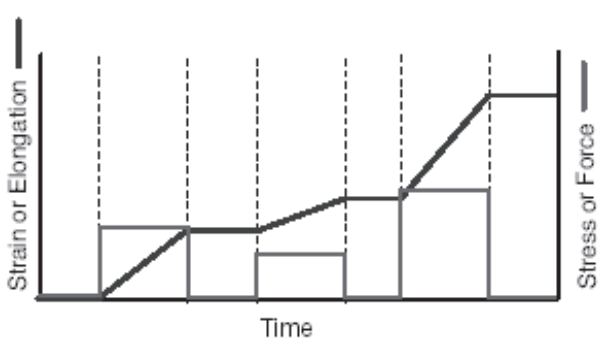


Figure 20.1. Relationship of stress and strain in Newtonian fluid.

Polymers in their molten state do not exhibit a direct relationship between the rate of deformation and the force or stress applied to the melt, producing a non-Newtonian response. Molten polymers have both a viscous and an elastic component. When the force is applied to a polymer melt, deformation occurs; the viscous component stays deformed when the force is removed, while the elastic component springs back. As stress is applied to molten polymer, three things can happen:

- Viscous flow—Material deforms as long as a stress is applied, shown in Fig. 20.1.
- Elastic deformation—Material deforms as soon as stress is applied, but when the stress is removed, the material returns to its original form, shown in Fig. 20.2.
- Rupture—Material deforms in the elastic mode to a specific elongation where it ruptures, preventing it from returning to its original form after the stress is removed.

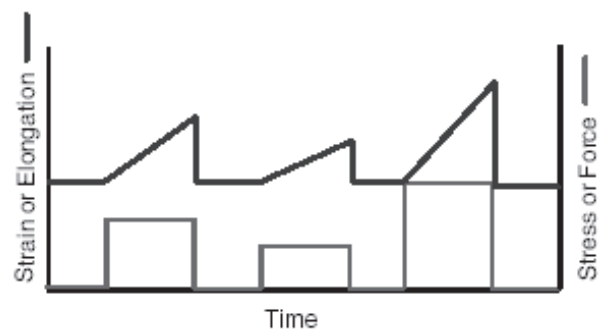


Figure 20.2. Elastic behavior to stress.

In viscous flow, the viscosity is defined as the ratio of applied stress divided by the rate of strain. Common viscosity units are: Pa•s, Poise, lb_f s/in^2 or Newton s/m^2. In elastic deformation, the modulus is defined as applied stress divided by recoverable deformation, measured in Newton s/m^2 or lb_f/in^2. Recoverable deformation is used because some deformation may be so extreme that rupture occurs.

Molecular weight, discussed in Chapter 18, is the single most important property in determining polymer viscosity. The relationship in a narrow molecular weight distribution is given by Eq. (20.1):

$$\eta = k \times MW^{3.5} \tag{20.1}$$

where η = Viscosity
k = A constant
MW = Molecular weight

Equation (20.1) teaches us that molecular weight has strong effect on the resin viscosity. Doubling the molecular weight gives approximately a 10-fold increase in viscosity.

20.1 Definitions

Common terms in discussing rheology are shear, shear rate, shear stress, shear modulus, shear flow, and extensional flow. It is important to understand these terms in order to grasp basic rheological concepts.

- ***Shear*** is the movement in either a solid or fluid of parallel layers within the sample. Consider two pieces of paper sliding past one another and generating frictional heat during the sliding operation. This is called *shear heat* due to friction caused by the sliding layers.
- ***Shear rate*** is the velocity gradient across a channel in which the fluids are sliding past each other in laminar flow. It is a measure of the deformation of a polymer melt, calculated from the flow rate and the geometry through which the melt is passing. Shear rate is the rate of change of velocity at which one layer passes over another. Normal shear rate units are reciprocal seconds.
- ***Strain*** is the ratio of the change in length or volume to the initial length or volume.
- ***Shear strain,*** like strain, is the ratio of deformation to original dimensions. For shear strain it is the deformation perpendicular to a line, rather than parallel to it. The ratio equals tan α, where α is the angle the sheared line makes with its original orientation.
- ***Shear stress*** is the force per unit area required to sustain a constant rate of movement.
- ***Shear modulus*** is the ratio of the shear stress to the shear strain. The elastic shear modulus (G') is a measure of the recoverable portion of the elastic deformation; it relates to extrudate swell. The higher G', the greater the melt elasticity, which is associated with greater extrudate swell.
- ***Shear flow*** is molten polymer flow caused by relatively parallel or concentric motion of surfaces, such as the screw in an extruder barrel. Shear flow can be caused by a pressure drop in the flow direction, as occurs in the die.
- ***Extensional flow*** is flow created by pulling on a molten polymer, forcing layers to move past one another. This occurs in the extrudate exiting the die as it is being drawn by the puller or in a converging flow channel.
- ***Viscous modulus*** (G") is a measure of the viscous component of flow. A higher ratio of G" to G' is associated with lower melt elasticity and less extrudate swell.
- ***Complex viscosity,*** designated as η^*, is equal to the shear stress divided by the shear rate; it is a measure of the polymer's resistance to flow.
- ***Shear thinning*** is the decrease in the polymer viscosity with increased shear rate, resulting from alignment of polymer molecules during processing.

20.2 Measurement

Polymer viscosity is measured differently, depending on the polymer state. The relative viscosity of one polymer compared to another within the same class is normally measured by MFI or solution viscosity, also called intrinsic viscosity. These measurements do not describe the polymer viscosity characteristics at the shear rates used in polymer processing. Measuring viscosity versus shear rate is done by oscillating plate rheometry at low shear rates (< 300 sec^{-1}) and capillary rheometry (100 to 30,000 sec^{-1}) at high shear rates. MFI was described in Chapter 18. Figure 20.3 shows a shear rate versus viscosity graph and the general shear rate areas where different polymer processes occur. Low shear rate processes include compression molding and the molding cycle in large-part extrusion blow molding. Extruder operations are typically between low and high shear rate processing, with shear rates generally ranging from about 50 sec^{-1} to several hundred reciprocal seconds. Injection molding is a high shear rate process, as polymer is forced through a nozzle, small gates, and runners at high speeds. Shear

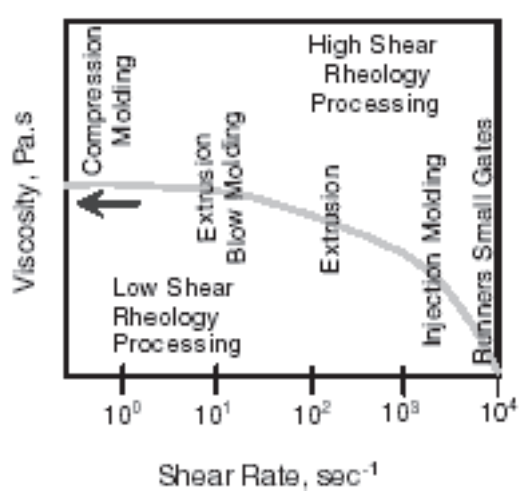


Figure 20.3. Processing shear rates.

rates in injection molding normally range in the thousands of reciprocal seconds.

An oscillating plate rheometer, or cone and plate rheometer, has two parallel plates: one oscillates and the other is fixed as the rheometer measures torque. The top plate oscillates at predetermined rates. Both plates are heated to be able to measure the viscosity at a particular temperature. In addition to measuring viscosity, these instruments also determine G' (storage modulus) and G'' (loss modulus). Typical data at low shear rates are shown in Fig. 20.4 for polypropylene (PP) at 235˚C. The viscosity decreases gradually as the shear rate increases. G' and G'' are shown to increase with increasing shear rate; above a shear rate of 100 sec^{-1}, G' and G'' converge. To use these data, additional graphs at other temperatures and different melt flow polypropylene are needed for comparison. If a lower MFI PP is measured and the G' value is higher at a particular shear rate based on the die land calculations, the extrudate swell exiting the die is anticipated to be larger.

Information on resin thermal stability can be obtained using oscillating plate rheometry by determining viscosity at a given temperature and shear rate versus time. Measuring the time when a specific resin degrades at several temperatures provides data to use in determining the time-temperature thermal stability for a given resin system in an extruder at specific shear rates. Thermal stability determination can be done at a shear rate that is higher than any experienced by the resin in the extruder to provide a safety factor in the actual process. Figure 20.5 shows viscosity versus time at a specific shear rate and temperature. The time when the resin starts to degrade under these conditions provides a guideline for the time resin degradation is anticipated to occur in an extruder at a specific melt temperature and shear rate. With several plots similar to Fig. 20.5, one can create a time-temperature degradation diagram similar to Fig. 20.6. Figure 20.6 is based on a given shear rate and shows the time it takes for resin degradation to occur at a particular temperature. Resins are stable for a long time at low temperature; as the temperature is raised, thermal degradation happens more rapidly. Each polymer system, with its own stabilization package, has a specific time-temperature curve for resin degradation. During processing, knowledge of the time-temperature curve can assist in setting the extruder temperature profile and to understanding how many times a resin can be reprocessed before it degrades and loses property performance.

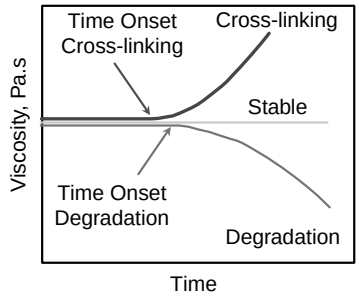


Figure 20.5. Time vs. viscosity.

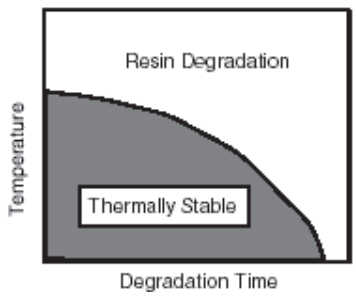


Figure 20.6. Time-temperature curve.

High shear rate versus viscosity data are generated using a capillary rheometer, shown in Fig. 20.7. Depending on the shear rate, the viscosity data cover flow in both extrusion and injection molding applications. A piston attached to a load cell forces molten resin through a capillary die at different rates. The test sequence is to place polymer in the barrel, allow it to come to an equilibrium temperature, and force it through a specific size orifice or capillary. The force required to push the resin through the orifice at increasing rates is measured. From this data a viscosity versus shear rate curve is calculated. A typical viscosity versus shear rate curve at four different temperatures for polycarbonate (Lexan 121) is shown in Fig. 20.8. A Newtonian fluid viscosity curve, which is independent of shear, is a straight line parallel to the x-axis. Lexan 121 viscosity at low shear rate is relatively Newtonian; however, at higher shear rates the polymer becomes

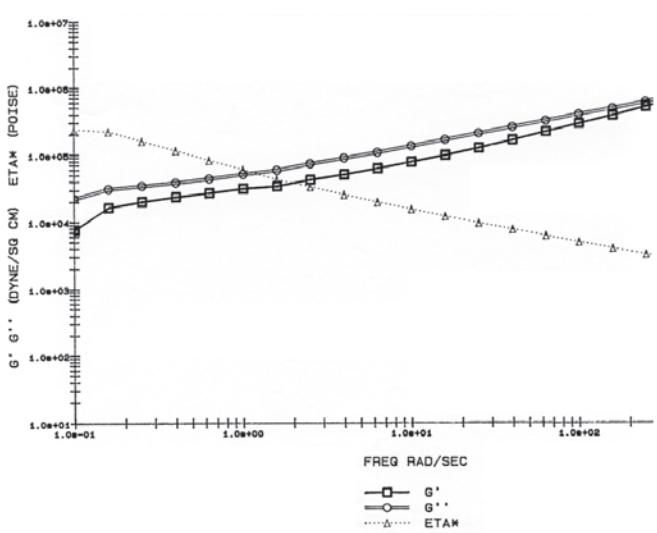


Figure 20.4. Oscillating plate rheometry data for polypropylene at 235˚C.

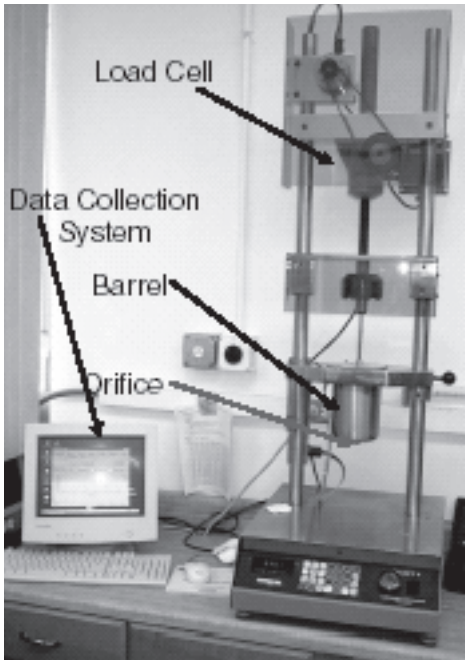


Figure 20.7. Kayeness capillary rheometer.

more non-Newtonian. These data can show how sensitive a resin is to temperature and shear rate. Changing the temperature from 250˚C to 280˚C at 250 sec^{-1} decreases the viscosity from 1627 to 589 Pa•sec or a factor of 3 for a 30˚C change. Changing the shear rate from 200 to 1000 sec^{-1} at 280˚C changes the viscosity from 589 to 481 Pa•sec. Linear polycarbonate viscosity is much more temperature-dependent than shear-dependent, meaning the viscosity changes more with temperature changes than with shear rate changes.

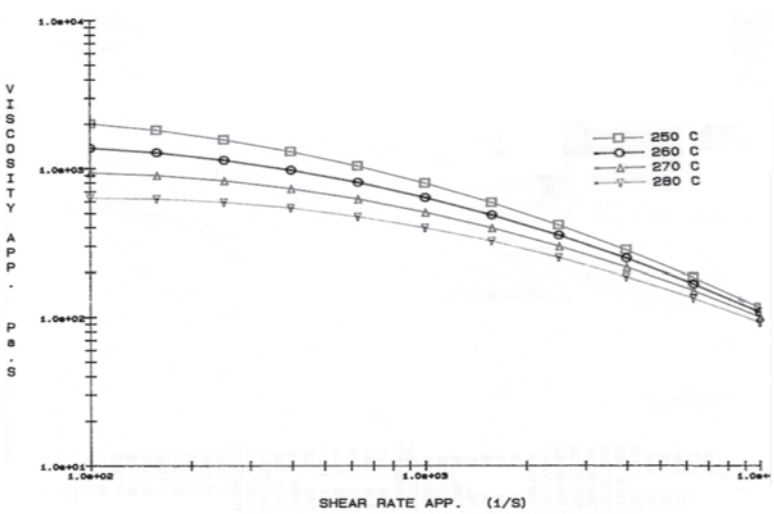


Figure 20.8. Capillary rheometry shear rate vs. viscosity data for Lexan 121.

Measurement errors can occur when using high rates in capillary flow rheometers. However, unless the data from capillary rheometry are being used for research purposes, the data are comparative and can be used without corrections. Typical corrections include the Bagley end correction for pressure drop and the Rabinowitsch correction for a nonparabolic flow velocity profile through the capillary. Newtonian fluids have a parabolic velocity profile, dilatants an extended parabolic profile, and pseudoplastics a flattened parabolic velocity flow profile.

The most important polymer flow in an extruder and die is shear flow, where one molten polymer layer slides next to another layer, applying a shearing force. The metering section velocity profile in a single screw extruder is due to screw drag flow and the backpressure flow from the breaker plate and screen pack or die resistance (discussed in Part 1, "Single Screw Extrusion"). The apparent viscosity, η, is given by Eq. (20.2):

$$\eta = \frac{shear\ stress}{shear\ rate} = \frac{\tau}{\gamma} \tag{20.2}$$

where shear stress, τ, is given by Eq. (20.3), and shear rate, γ, is given by Eq. (20.4).

$$Shear\ Stress = \tau = \frac{\Delta P \times R}{2L} \tag{20.3}$$

where

ΔP = Pressure drop
R = Capillary radius
L = Capillary length

$$Shear\ Rate = \gamma = \frac{4Q}{\pi R^3} = \frac{4\pi r_b^2 \times S}{\pi r_c^3} = \frac{4 \times r_b^2 \times S}{\pi \times r_c^3} \tag{20.4}$$

where

Q = Volumetric flow rate
R = Capillary radius
r_b = Barrel radius
S = Piston or ram speed
r_c = Die radius

Figure 20.9 shows shear stress versus shear rate for pseudoplastics, Newtonian fluids, and dilatants. With a Newtonian fluid, the slope of the line is constant as the shear stress and shear rate change. With either a pseudoplastic or a dilatant fluid, the viscosity changes as a function of shear rate. In Fig. 20.9, η_1 does not equal η_2 and η_3 does not equal η_4, as the viscosity changes with shear rate and shear stress.

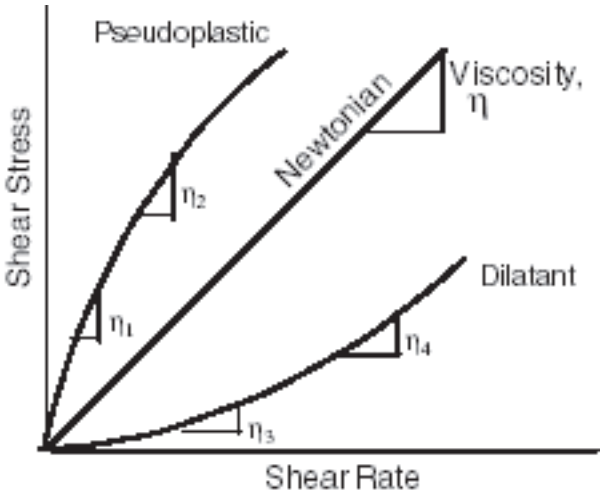


Figure 20.9. Relationship of shear stress to shear rate.

20.3 Viscosity in Extrusion

Polymer viscosity is important in extrusion to understand the processing window, the role temperature plays in viscosity, and the importance of shear rate during processing. Figure 20.8 is the viscosity versus shear rate curve for polycarbonate (Lexan 121). Typical extrusion conditions experience 50–1000 sec^{-1} shear rates, and for Lexan 121 the viscosity versus shear rate curve shows large differences with temperature changes and only small differences with shear rate changes. To lower Lexan 121 viscosity during extrusion, it is more effective to decrease the melt temperature. Going to a higher shear screw in either single or twin screw extrusion does not dramatically alter the resin viscosity. However, higher shear rate does induce shear heating, which lowers the polymer viscosity and can lead to resin degradation.

There are other resins, like PP, that are shear-sensitive but not temperature-sensitive. Figure 20.10 shows viscosity versus shear rate curves for PP at three different temperatures. Comparing these curves with Fig. 20.8, the slope in the shear rate range for extrusion is much steeper, indicating that a change in shear rate affects viscosity more than temperature. Changing the temperature from 190˚C to 230˚C at 200 sec^{-1} decreases the viscosity from 280 to 190 Pa•sec, while changing the shear rate from 200 to 1000 sec^{-1} at 210˚C changes the viscosity from 230 to 80 Pa•sec. Unlike polycarbonate, PP is a shear-sensitive rather than temperature-sensitive polymer. Common practice is to raise PP melt temperature during both extrusion and injection molding to lower the viscosity. Unfortunately, temperature has only a minor effect on PP melt viscosity, so other than using more energy and consuming more thermal stabilizer, there is not a lot accomplished at higher processing temperatures. For PP, higher shear will lower the viscosity and provide higher flow.

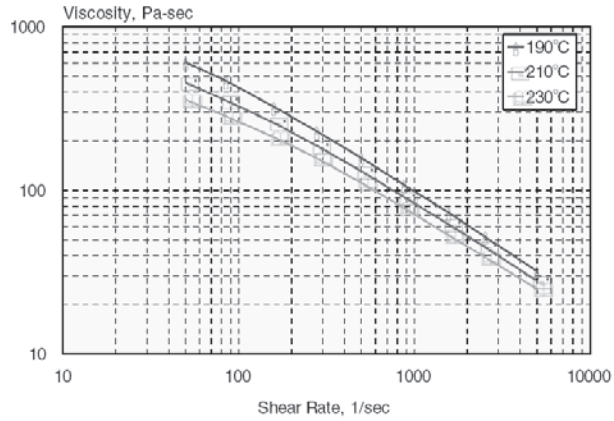


Figure 20.10. Shear rate vs. viscosity for 12 MFI PP.

Some resin systems exhibit both strong temperature and strong shear dependence. In these systems, while both temperature and shear have significant effects on viscosity, changes in shear rate affect viscosity more than changes in temperature. Screw design in both the mixing and melting stages of either a single or twin screw extruder is important in obtaining an appropriate shear rate for the polymer being processed and getting optimum extruder performance. Table 20.1 shows the temperature and/or shear dependence for several resin systems.

Table 20.1. Temperature/Shear Sensitivity

Shear Sensitive	Temperature Sensitive	Temperature and Shear Sensitive*
PP	PC	ABS
LDPE	PBT	PA 6
LLDPE	PET	PA 6,6
HDPE	Rigid PVC	Polystyrene
		Flexible PVC

*Shear has stronger effect than temperature

The viscosity versus shear rate curve visually shows the sensitivity to shear thinning. Numerous polymer processors buy resin based on MFI, which is a viscosity measurement at lower shear rates. Without having a viscosity versus shear rate curve, the processor has no idea what the resin viscosity is doing at extrusion shear rates and temperatures used during processing. Approximate shear rates are easy to calculate in both single and twin screw extruders, with equations given in Parts 1 and 2. Consider the effect of shear thinning, shown in Fig. 20.11. Polymer A has a higher viscosity at low shear rate and a lower viscosity at high shear rate, compared to polymer B. When viscosity versus shear rate curves

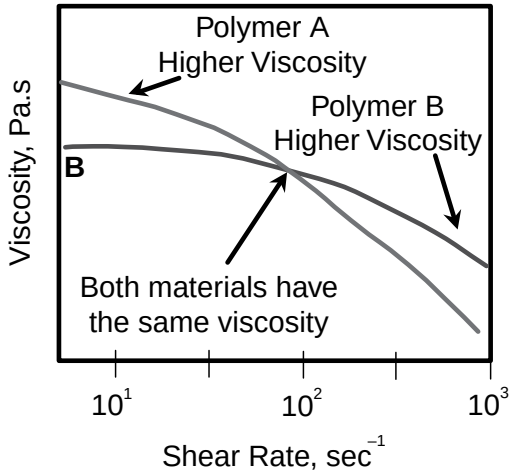


Figure 20.11. Two polymers at the same temperature with different shear rate vs. viscosity curves.

cross, the effect of viscosity will change, depending on the processing being done and where on the viscosity curve the process is located.

The other area where viscosity versus shear rate data are critical is in coextrusion.[2] To prevent interfacial instability between adjoining polymer layers in a two-layer coextrusion, it is essential for the resins to have the same viscosity in the die or feed block, where the different resin systems are brought together in melt form. Taking Fig. 20.11 as an example, if the shear rate where the two resins are brought together is approximately 100 sec^{-1}, polymers A and B have the same viscosity and the extrusion is anticipated to run very smoothly with no interfacial instabilities. However, if the shear rate where polymers A and B are brought together is 50 sec^{-1} or 500 sec^{-1} instead of 100 sec^{-1}, the resin viscosities are quite different and interfacial instabilities might create a problem. To determine the proper temperature to run the die and/or feed blocks, it is important to estimate the shear rate first and then determine the temperature where the resins have the same viscosities.

Once shear rate versus viscosity data are generated at three temperatures for each resin used in a coextrusion operation, and the proper shear rates are calculated where the resins come together, an Arrhenius plot can be generated to predict the proper operating temperature. An Arrhenius plot graphs log viscosity versus $1/T$ in Kelvin. Plotting viscosity at three temperatures gives a straight line. The lines for the different coextrusion resins can then be extrapolated to a temperature where the resins have the same viscosity. If the temperature is an appropriate melt processing temperature for each resin system, the two polymers can be expected to coextrude without any interfacial instability. However, if the selected temperature is not a suitable processing temperature for either resin, another polymer has to be substituted for one of the resins using either a higher or lower viscosity, depending on the intercept obtained in the first set of experiments.

For example, assume polycarbonate is going to be coextruded with polypropylene. What melt flow polypropylene is required and what melt temperature is optimum to run the coextrusion? Assume the resins are coming together in the die or feed block at very low shear rates, 10 sec^{-1}. Table 20.2 contains some rheology data collected at 10 sec^{-1} on polycarbonate (Lexan 121) and polypropylene with 0.5, 5, 12, and 35 MFI. Figure 20.12 shows the Arrhenius plot for the viscosity versus inverse temperature. If the melt temperature of PP and Lexan 121 is 236˚C, the correct PP to use in coextrusion is 0.5 MFI. However, 236˚C is too low a processing temperature for Lexan 121, so 0.5 MFI is an inappropriate choice for

Table 20.2. Viscosity Data at 10 sec^{-1} for Different MFI PP and Lexan 121

Temperature	0.5 MFI PP	5 MFI PP	12 MFI PP	35 MFI PP	Lexan 121
235° C	2976	1326	705	323	3444
250° C	2576	1219	547	180	1683
265° C	2386	911	409	140	907

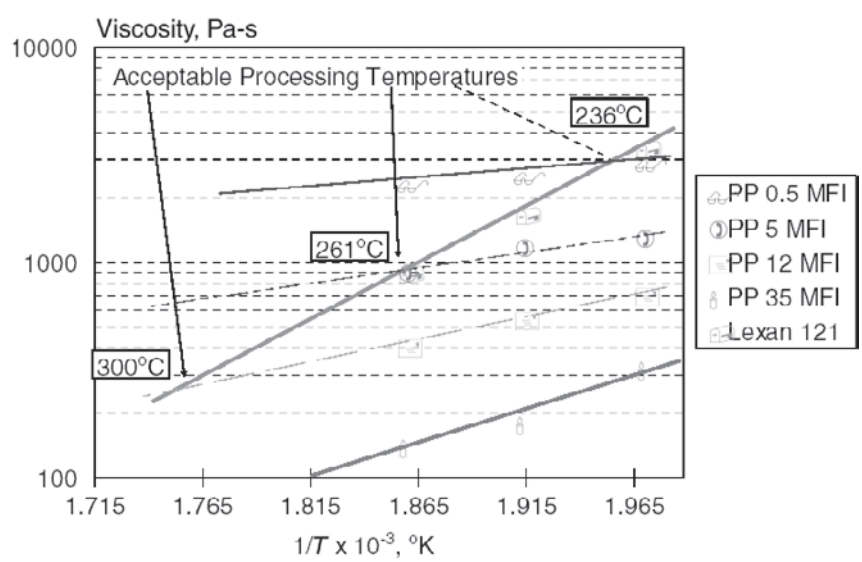


Figure 20.12. Viscosity vs temperature data.

PP resin. Five MFI PP resin at 261˚C has the same viscosity as Lexan 121 at a shear rate of 10 sec^{-1}, and 261˚C is an appropriate processing temperature for both PP and Lexan 121. Twelve MFI matches the viscosity of Lexan 121 at 300˚C melt temperature at a shear rate of 10 sec^{-1}; this is on the high end of the processing range for PP. Consequently, the resins of choice in this coextrusion application are

- 5 MFI PP with Lexan 121 at 261˚C melt temperature
- 12 MFI PP with Lexan 121 at 300˚C melt temperature
- An 8 or 10 MFI PP with Lexan 121 at an intermediate melt temperature

For multiple-layer coextrusion, more detailed analysis of the relative layer viscosities is required and is beyond the scope of this book. In general, the layer viscosities should decrease, moving from the inner polymer layer toward the outermost layers to avoid interfacial instabilities.

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2. Butler, Thomas I., Veazey, Earl W., Co-Ed, *Film Extrusion Manual: Process, Materials, Properties*, TAPPI Press, Chapter 4, 1992.

Review Questions

1. What is meant by a viscoelastic material?
2. What is the difference between a Newtonian and non-Newtonian fluid as stress is applied to the fluid over a specific time period?
3. What is the definition of viscosity?
4. Explain how changes in molecular weight alter the polymer viscosity.
5. What are two methods of measuring polymer viscosity and how do they differ?
6. Explain how viscosity can be used to measure resin thermal stability and generate a time-temperature curve.
7. What is the purpose of a time-temperature curve?
8. At what range of shear rates do the following chemical processes occur: injection molding, compression molding, extrusion, and coextrusion?
9. Name three polymers whose viscosities are both shear- and temperature-sensitive, three that are temperature-sensitive, and three that are shear-sensitive.
10. Why is viscosity important to understand in extrusion?
11. What is meant by a shear-sensitive polymer?
12. Can two polymers have the exact same melt flow index but have different resin viscosities in the extruder? Explain your answer.
13. What effect does raising the melt temperature 30˚C have on HDPE viscosity?
14. What effect does raising the melt temperature 30˚C have on PBT viscosity?
15. What effect in extrusion is expected by changing the mixing head on a single screw extruder that is extruding PP? Assume the change is from pin mixing to a Maddock mixer.

21 Testing Properties

Testing physical properties on standard test specimens obtained from injection molding or extruded parts is done in accordance with the procedures of either the American Society for Testing Materials (ASTM) or the International Organization for Standardization (ISO). Testing protocol has been changing from ASTM to ISO as resin suppliers and other custom compounders compete in global markets. At the same time, ASTM has been changing its procedures to be compatible with ISO standards. Plastics are marketed and sold based on their property performance profile and response to specific test procedures. With more than 30,000 commercially available materials, selecting the best resin formulation for a particular application is based on two criteria: the property profile/performance in a given environment and cost. Cost is generally directly proportional to temperature performance, heat resistance, and continuous use temperature. (Continuous use temperature is defined as the temperature the part can withstand in an application either over a long time or for repeated short-term durations at higher temperature.)

Thermoplastic materials are generally categorized as either commodity resins or engineering resins. Commodity resins are lower price with lower property performance and higher sales volume. Thermoplastics falling into this group include

- Polyethylene (PE)
- Polypropylene (PP)
- Polyvinyl chloride (PVC)
- Polystyrene (PS)

Engineering resins generally have higher tensile, flexural, and impact properties combined with higher temperature performance. Thermoplastics falling into this group include

- Polycarbonate (PC)
- Nylon (PA)
- Polysulfone (PSO or PSF)
- Polyetherimide (PEI)
- Polyetherether ketone (PEEK)
- Polyphenylene sulfide (PPS)
- Acrylonitrile butadiene styrene/polycarbonate blend (ABS/PC)
- Polyoxymethylene (POM)
- Thermoplastic polyesters (PET and PBT)
- Polyphthalamide (PTA)

In between these two classes are resins that have very good properties and slightly lower temperature resistance. They are not as cheap as the commodity resins and not as expensive as the engineering thermoplastics. The question has been raised, Should these resins be classified as commodity resins or engineering resins? Thermoplastics in this group include

- Acrylonitrile butadiene styrene (ABS)
- Polyphenylene oxide/polystyrene blend (Noryl® or PPO/PS)
- Polymethylmethracrylate (Plexiglass® or PMMA)
- Styrene butadiene copolymer (SB)
- Styrene acrylonitrile copolymer (SAN)

In addition to the polymers mentioned above, numerous backbone modifications to produce copolymers, branching, modified monomers, and additives within each resin are used to alter their property performance. Commodity resin producers are always looking for modifications to enhance the resin performance by adding additives, fillers, or reinforcements and becoming players in the engineering thermoplastic arena. Including fiber reinforcement or filler in crystalline resins significantly raises their heat performance. Polypropylene has a heat deflection temperature (HDT) of approximately 200˚F (93˚C) at 66 psi; adding 30% glass fiber raises the HDT to 300˚F (149˚C) at 264 psi. Simultaneously, the tensile and flexural properties increase substantially and are accompanied by a decrease in impact strength.

Numerous specialty or modified resins are available with unique product advantages for specific applications. Product performance attributes of some specialized resins systems are not always obvious from the physical property databases. Unique product attributes include

- Outstanding UV resistance and weatherability for outdoor applications
- Barrier to gases such as oxygen or moisture for food packaging
- High impact performance for bulletproof plastic sheet
- Cookware or food packaging that can be directly inserted into microwave ovens
- Resistance to sterilization techniques for medical applications
- Meeting Food and Drug Administration (FDA) guidelines for food and medical contact
- Usability for insertion into the body either as pins or replacement joints
- Chemical resistance to solvents from acids and bases to organic solvents
- Usability as a living hinge
- Flame retardance

The applications and modifications are limited only by one's imagination.

This chapter covers the basic test methods used in collecting data for developing data sheet properties. Test methods, whether specified by ASTM or ISO, must be followed to provide test data that can be compared. By following specific procedures for sample specimen preparation, sample conditioning prior to testing, and test methods (specific equipment, speeds, and conditions), data generated are comparable. Consequently, data presented in the various plastic databases can be used to predict part performance based on computer-aided design (CAD) and finite element analysis (FEA). Assume that while developing a new application, the resin system fails. Depending on the failure mode, a new resin can be selected with slightly higher property performance in the particular failure area, producing a part that meets all the end-use requirements. One example is a custom extruder that is making a plastic part for the side of a treadmill. As the part is being stapled into place during assembly, cracks occasionally appear in the longitudinal direction. The base resin is changed to a material with slightly higher impact, and the assembly problem disappears.

The second purpose for specimen testing and understanding test methods and procedures is to perform quality control procedures. Suppliers producing either raw materials or compounded resins test their product prior to sending it to their customers. Extruding finished products, whether it is sheet, film, monofilament, profiles, pipe, tubing, and so forth, requires some standardized testing to ensure quality prior to shipment. The second need for in-house testing is raw material inspection. Resin suppliers may send a "certification of compliance" with raw materials, provided the customer requests this. Under normal operations, this contains information jointly agreed upon by both the vendor and the customer. It certifies that the product shipped meets specific standards based on standardized plastic testing. It is a good practice to test incoming raw materials periodically and build an internal database in the event a problem arises later with a specific shipment. Without an internal database and understanding the material properties, it is impossible to ascertain whether the vendor sent some material out of specifications or there exists a problem in your facility, preventing the resin from processing properly.

Additives, fillers, and reinforcements that are added to plastics to modify their property performance will be discussed later in this chapter.

Only general plastics tests are discussed here. They help one understand the properties in property data sheets and databases, what they mean, and how they are derived. Many databases supply general information on commercially available materials. *Modern Plastics Encyclopedia* and *Plastics Technology Yearbook* both publish material databases in their yearbooks. Physical properties are available through the Internet at www.matweb.com and www.idesinc.com.

21.1 Density and Specific Gravity

Density, a fundamental property, is defined as the weight per unit volume. It is normally measured in grams per cubic centimeter; however, bulk density and plastic density may be given as pounds per cubic foot. In extrusion three different densities are critical:

- Raw material bulk density
- Melt density in the extruder
- Solid polymer density

The bulk density is important in determining whether potential feed problems may occur. A bulk density below 20 pounds/cubic foot (defined as the material poured into a box 1' × 1' × 1' weighing less than 20 pounds) is very fluffy and may not flow well from the feed hopper into the extruder. If it is free flowing, the feed volume per unit time may greatly reduce the anticipated throughput rate. Normal resin or powder has a bulk density of about 30–60 pounds/cubic foot and is extruded easily on either a single or twin screw extruder at high rate if it is free flowing.

The melt density is higher than the bulk density, as the air and space between the particles in the solid state are removed. In the melt state, the density is less than the final plastic part, as polymers contract as temperature decreases. In the final part, the molecular chains are tightly packed together and the air is removed that was originally present in the bulk density. Comparing high density and low density polyethylene in the same part, the high density polyethylene parts weigh more as the polymer atoms and molecules are more closely packed together in the higher crystallinity HDPE. Consequently, there is more mass per unit volume.

Density is critical when buying resins for extrusion. As an example, assume a pipe is being made that is 10 feet long with a 2.625-inch outside diameter and a 2.500-inch inside diameter. The part volume is obtained by taking the circular area and multiplying it by the length:

$$\begin{array}{ccccccc} \text{(Area of} & & \text{Area of} & & & & \text{Volume} \\ \text{Outside} & - & \text{Inside} & \times & \text{pipe} & = & \text{of} \\ \text{Diameter} & & \text{diameter)} & & \text{length} & & \text{plastic} \end{array}$$

$$(2\pi r_1 - 2\pi r_2) \times 120 \text{ inches} = \text{Volume of plastic in the pipe}$$

where $r_1 = OD/2$ 2.625/2 = 1.3125
$r_2 = ID/2$ 2.500/2 = 1.250
$L = 10 \text{ ft} \times 12 \text{ inch per ft}$ 120 inches

The calculation becomes

$$[(2)(3.1416)(1.3125) - (2)(3.1416)(1.25)] \times 120 = 47.1 \text{ inches}^3 \text{ of plastic}$$

Assume that two different raw materials work equally well in this application: material A with a 1.10 g/cc density costs $1.10/pound, and material B with a 0.98 g/cc

density costs \$1.20/pound. Which is the most cost-effective material to use to produce the part?

Part weight material A = (47.1 in.^{3}) × (1.10 g/cc) × [(2.54 cm)^3/in^3] × (1 pound/454 grams) = 1.870 pounds

Part weight material B = (47.1 in.^{3}) × (0.98 g/cc) × [(2.54 cm)^3/in^3] × (1 pound/454 grams)

Part weight material C = 1.666 pounds

The part cost with material A is: \$ = (1.870 pounds) × (\$1.10/pound) = \$2.057. The part cost with material B is: \$ = (1.666 pounds) × (\$1.20/pound) = \$1.992. The savings on a 50,000-piece production run by using the more expensive material B is: (\$2.057 – \$1.992) × 50,000 = \$3,250.

Since resin is purchased on a cost-per-pound basis, it is important to understand the density-part cost relationship. A resin with a lower density uses fewer pounds to produce the same volume part. If all other properties are equivalent, it is important to calculate the cost per part based on the resin cost and the density.

Specific gravity is a unitless number and is derived from a material's density divided by the density of water; thus all units cancel. And, since water's density is 1 gram per cubic centimeter (at specific conditions), then a polymer's specific gravity would also correspond to a polymer's density as expressed in grams per cubic centimeter.

Density or specific gravity is used when mixing two different density or specific gravity polyethylenes and calculating the anticipated resultant density or specific gravity of the final product. As an example, assume 150 pounds of resin C with a density of 0.924 g/cc is mixed with 110 pounds of resin D with a density of 0.916 g/cc. What is the anticipated density of the mixture?

Volume of resin C = (150) × (1/0.924) = 162 volume

Volume of resin D = (110) × (1/0.916) = 120 volume

Total volume of mixture = 162 + 120 = 282 volume

Total mass = 150 + 110 = 260 mass

Density of mixture of C plus D = 260/282 = 0.922

In this calculation the conversion factors were ignored going from g/cc to pounds because they are constants in the calculation and cancel each other out. Consequently, the final density of the mixture is approximately 0.922 g/cc.

21.2 Melt Flow Index

Melt flow index (MFI), melt flow rate (MFR), or melt index (MI) are the same test at different test conditions and are used to measure melt viscosity under a constant load and low shear rates. Figure 21.1 shows a Tinius Olsen melt flow index testing machine, which is essentially a ram extruder. Figure 18.15 is a melt flow index testing machine schematic, and that section discusses its operation. The test uses a specific size orifice or die in a heated barrel, with a piston or plunger and weight on top inserted into the barrel to force the polymer through the orifice. Starting with a clean barrel, polymer is added and packed into the barrel with the plunger or piston. After a specified heating period to melt or soften the resin, a weight is placed on the piston, and the molten polymer is forced through the die. The melt flow index is the amount measured in grams that exits the die in 10 minutes. Each resin system has a specific orifice length and diameter, barrel temperature, and piston load specified in ASTM D1238-86 or ISO R1133 test methods. At the end of the test, the die and barrel are cleaned.

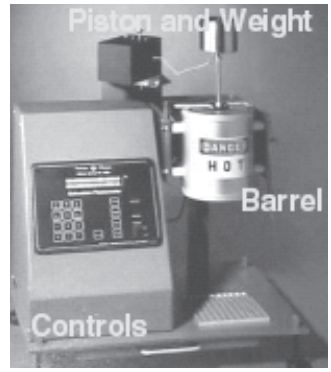


Figure 21.1. Tinius Olsen melt flow index tester.

Higher melt flow index correlates with a lower viscosity resin. As the resin viscosity decreases, the flow per unit time increases. Lower melt flow index or fewer grams passing through the die in 10 minutes relates to higher viscosity.

21.3 Tensile

Tensile testing measures the deformation resistance to stretching or pulling. Figure 21.2 shows a universal testing machine used to test tensile, flexural, and compressive properties in different modes of operation. A "dog-bone" shaped specimen, shown in Fig. 21.3, is clamped on each end, a load is applied, and the sample is pulled until it ruptures. Depending on the material, the sample may deform in the thin section of the bar by "necking down" or elongating as the load is increased. This pro-

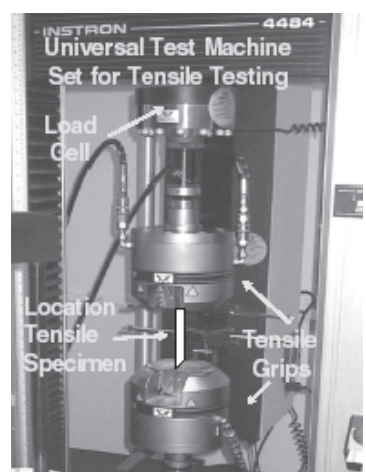


Figure 21.2. Instron universal testing machine.

vides the tensile strength at both yield and failure, the tensile modulus, and the elongation. Figure 21.4 shows a typical tensile stress-strain curve and the important measurement points. The x-axis represents the strain or sample elongation as the crossbar in Figs. 21.2 and 21.3 is raised. The y-axis shows the stress or force required to stretch the specimen a particular distance. The stress required to elongate the sample increases almost linearly until the yield point is reached. At the yield point the tensile specimen actually starts to neck down and continues necking until the sample ruptures. The initial portion of the stress-strain curve (start to the yield point) is the elastic region. If the load is removed during this section, the bar nominally returns to its starting dimension and can in theory be repeatedly stretched and released and returned to its starting point numerous times. In this region it shows elastic character similar to an elastomeric material. At the yield point, the part deforms as the specimen cross section changes and dimensions become noticeably smaller. Once the specimen has exceeded the tensile yield point, the deformation process is irreversible, and the sample bar can never be returned to its original dimensions. After the yield point the sample will continue to deform under load until it ruptures.

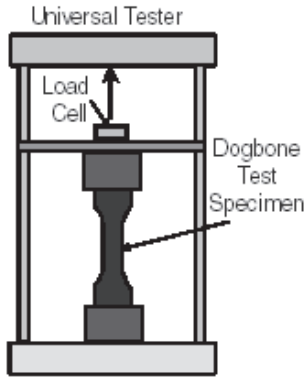


Figure 21.3. Tensile test setup.

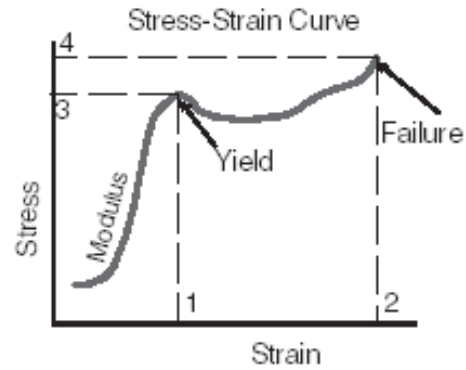


Figure 21.4. Typical stress-strain curve for unmodified plastic.

Measurements obtained from a tensile test are the tensile modulus, elongation at yield (point 1), elongation at failure (point 2), strength at yield (point 3), and strength at failure (point 4) in Fig. 21.4. The initial slope of the curve is defined as the tensile modulus by Eq. (21.1):

$$\text{Modulus} = \text{Stress/Strain} \tag{21.1}$$

The modulus is also referred to as the elastic modulus, modulus of elasticity, or Young's modulus. The elastic limit in reality is slightly before the yield point, extending only through the linear portion of the stress-strain curve.

Dogbone-shaped specimens are used to isolate the elongation to a particular section of the bar. Elongation normally occurs in the smaller section of the bar due to the reduced cross sectional area. Occasionally, specimens break in the grips, rendering the values useless. As soon as the tensile specimen is put under tension (crossbar in Fig. 21.3 starts to go up), the stress-strain curve shows the increase in force necessary to elongate the sample. In general, tensile curves for unfilled amorphous resins show a gradual increase in the tensile strength after the yield point until the sample ruptures. With semicrystalline materials, the strength required to elongate the sample after the yield point may decrease while the sample is necking and the molecular chains are being elongated. Once the amorphous domains of the semicrystalline material are elongated and necking is completed, the strength increases until failure. Reinforced or filled resins have higher strengths to break and higher modulus values. Their elongations are relatively low, as the samples normally break at the yield points.

Plastic toughness is a resin's ability to withstand fracture. Generally, polymer toughness is related to the area under the stress-strain curve. Glass-reinforced resins have very low elongation but very high stress to break, resulting in a large area under the curve. Unfilled resins, on the other hand, have lower tensile strengths with high elongations. Consequently, these materials are very tough, as defined by the area under the stress-strain curve. The relative toughness compares the area under the stress-strain curve plus the yield strength and the failure strength. Figure 21.5 compares tough and brittle materials based on the stress-strain curves (assumes the axes for all samples are identical).

ASTM D-638-99 and ISO 527-2 are the tensile test protocols.

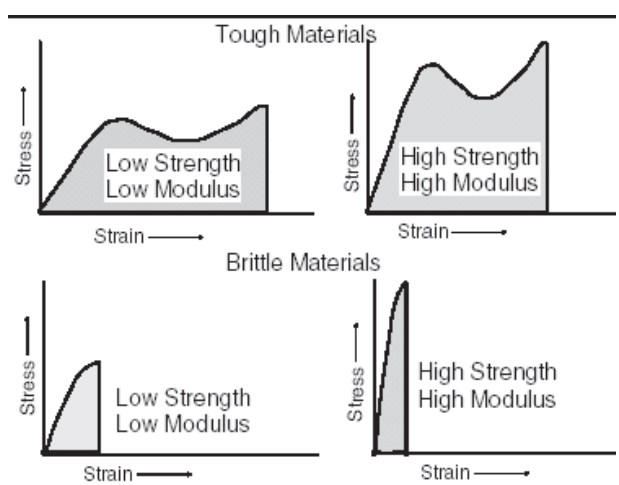


Figure 21.5. Comparisons of tough and brittle materials.

Applications where plastic parts are being used to pull something or hold something in place under a force require specific tensile properties. Some applications requiring a specific tensile strength are

- Monofilament line used in fishing, where a specific tensile strength is required to reel in a 10-pound fish versus a 30- or 40-pound fish.
- Plastic garbage bags filled to capacity that are being held by the top and carried some place for disposal.
- Glass-reinforced plastic bumper beams on cars. The beam bends inward on impact, putting the back side of the beam in tension.

Numerous other examples exist in plastic applications where tensile properties are important performance criteria.

21.4 Flexural Test

Flexural tests are performed on the same universal tester used to measure tensile tests. Flexural tests measure bending resistance. Normally, the test is run as a three-point bend, shown in a schematic in Fig. 21.6. Figure 21.7 shows a universal tester with the flexural test fixture mounted. Figure 21.8 shows a typical flexural test stress-strain curve.

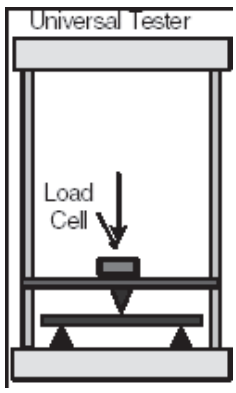


Figure 21.6. Flexural test configuration.

The flexural modulus, which is the initial slope of the stress-strain curve, represented by M in Fig. 21.8, measures the polymer stiffness and its bending resistance. The strength at 5% strain, B in Fig. 21.8, is reported as the flexural strength at 5% strain. The ultimate flexural strength is given by A in Fig. 21.8, with the strain at maximum flexural strength at point D in Fig. 21.8. Flexural strength is an important polymer property measuring the stiffness and strength in a bending mode. Flexural strength is important in structural plastic parts required to support or absorb a load, such as a bumper beam, plastic rod, plastic extruded profiles, and so forth.

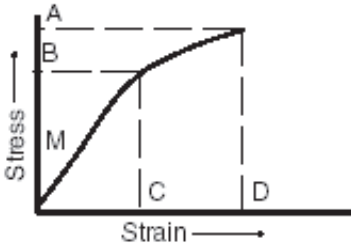


Figure 21.8. Flexural stress-strain curve.

ASTM D-790 and ISO 178:1993 are the flexural test methods.

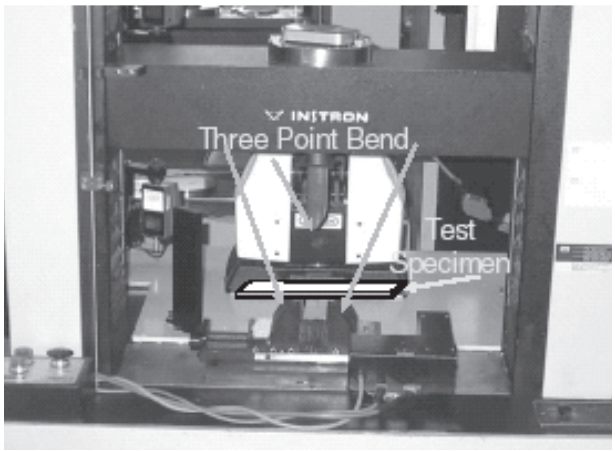


Figure 21.7. Instron universal tester with flexural fixture.

21.5 Compressive Strength

Compressive strength, like tensile and flexural strength, is measured on a universal test machine. The test specimen is a $0.5 \times 0.5 \times 2.5$ inch specimen standing on end between two flat platens. The crossbar is lowered at constant rate, and the force to compress or crush the specimen is measured. The stress-strain curve is similar to the tensile stress-strain curve, except failure occurs when the specimen is crushed. Test specimens with different cross sectional areas than that specified above can be used for testing. However, these specimens may require a support fixture to prevent the specimen from bending or flexing. Failure mode for the test specimen is crushing without any specimen bending allowed.

ASTM D-695 and ISO 604 are the compression strength test methods.

21.6 Heat Deflection Temperature

Heat deflection temperature (HDT) measures heat resistance and the temperature where the specimen deforms. The HDT test specimen is a $0.125 \times 0.50 \times 5.0$ inch ($3.2 \times 12.8 \times 128$ mm) bar. The test specimen is placed on its narrow side in a three-point bend fixture (Fig. 21.9) with either a 66 or 264 psi force applied. The bar and test fixture is placed in an oil bath, where the temperature is raised

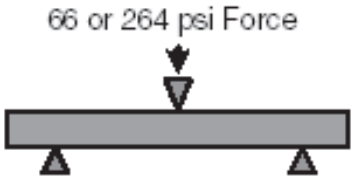


Figure 21.9. HDT test configuration.

2˚C/minute until the bar deflects 0.10 inch (2.54 mm). For amorphous resins, deflection occurs about 10˚F (5.5˚C) below its glass transition temperature, where the molecular motion makes the bar more pliable, allowing deformation. Adding fillers or reinforcing fibers to crystalline matrices greatly improves their HDT and heat performance. In amorphous materials, fillers and reinforcement have a small positive effect on the HDT values. However, the matrix still softens and becomes rubbery with reinforcements present and does deform under load.

Figure 21.10 shows a six-position HDT test machine produced by Ceast. There are individual stations for each sample submersed in the oil bath. The oil temperature increases at 2˚C/minute, with a specific weight applied to produce the correct sample loading. The temperature is automatically recorded when the specimen distorts 0.10 inch.

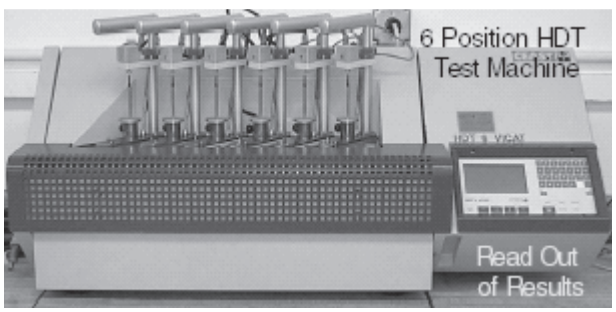


Figure 21.10. HDT/Vicat test machine made by Ceast.

Vicat is similar to HDT, with the exception that the 66 or 264 psi force that is applied across the narrow width is replaced by a point loading. Vicat measures the point indentation into the sample. Similar to HDT, this occurs when the bar softens as molecular motion increases.

ASTM D-648 and ISO 75-1:1993 are the HDT and Vicat test methods.

21.7 Long-Term Heat Aging

In some applications, heat resistance is required over extended time. Typical applications are automotive under-hood parts, internal dishwasher or dryer assemblies, internal washing machine and dryer parts, and plastic parts used in high-temperature industrial applications. At high temperature, plastic resins degrade over time. Glass-filled PP, for example, can withstand 300˚F (150˚C) for 1000 hours or more; however, if the part is exposed over its lifetime to 300˚F (150˚C) for 5000 hours, glass-reinforced PP is not the correct plastic matrix. After approximately 1000–1500 hours, the thermal stabilizer in PP is consumed and the part loses its structural integrity.

There are many long-term heat aging tests. Tensile or impact specimens can be placed in an oven at elevated temperatures (significantly higher than the end-use application) and periodically removed and tested for physical properties. A second method is to age the samples in ovens at elevated temperatures, periodically remove the samples, and weigh them. As the thermal stabilizers are consumed, the part gradually loses weight. Eventually, the thermal stabilizer is consumed and the final part loses weight very rapidly. Plotting weight loss versus time at elevated temperature defines the part life expectancy.

21.8 Thermal Properties

Other important thermal properties are the coefficient of thermal expansion, mold shrinkage, T_g, and thermal conductivity. The test methods used to determine these properties are beyond the scope of this book. The coefficient of thermal expansion describes the change in part dimensions with changes in temperature with units of inch/inch/˚F or mm/mm/˚C. This property is important because it indicates how much a part expands if the temperature is increased, or how much it shrinks if the temperature is decreased. As an example, a customer approaches your company and asks if you can make a polycarbonate profile that is 36 inches (914.4 mm) long while holding length tolerance at ± 0.02 inch (0.5 mm). This raises the question, "Is this a business your extrusion company wants to pursue?" For polycarbonate the coefficient of thermal expansion is 3.75×10^{-5} in/in/˚F (6.75×10^{-5} mm/mm/˚C). Yes, the part can be made within the tolerance requested; however, the temperature must be specified where the part is to be 36 ± 0.02 inches (914.4 ± 0.5 mm). If the customer says it is to be 36 ± 0.02 inches (914.4 ± 0.5 mm) at 80˚F (26.6˚C), the part must be allowed to cool sufficiently and stabilize at 80˚F before cutting, and the customer must measure the part under the same conditions after stabilization. Assume a truckload of parts are made that meet specification, they are shipped in the wintertime in the north, and the part temperature goes down to 0˚F in the truck. When they are delivered, the parts are measured at 50˚F (10˚C) and are rejected because they are too short. This is 30˚F (16.6˚C) below the specification temperature. Assuming the part is 36 inches at 80˚F, the length at 50˚F is

$$36 - (3.75 \times 10^{-5})(30)(36) = 36 - 0.0405 = 35.96''$$

The same situation occurs if the profiles are cut before they come to both temperature and stress equilibrium (takes about 24 hours). Assume the profiles are cut hot (120˚F [49˚C]); after they cool to 80˚F (26.6˚C) they will be shorter. It may take 24 hours for all the residual stresses from the extrusion process to be relieved and the

part to reach equilibrium. To cut the parts in-line, the shrinkage after cutting has to be determined and the appropriate adjustment made when cut.

Mold shrinkage, usually given in percent, indicates how much the material shrinks from the time it solidifies in the mold until it reaches room temperature. This is related to the coefficient of thermal expansion. Understanding and compensating for mold shrinkage is necessary when cutting dies and setting up calibration tanks and cooling fixtures. Hot parts continue to shrink after they solidify until the entire part reaches equilibrium at room temperature. Understanding shrinkage is helpful in calculating how big the part needs to be in the die, calibration tank, or fixture to have the proper dimension after shrinkage is complete.

The glass transition temperature, or T_g, is useful for understanding the temperature resistance of amorphous polymers and as a guideline to processing conditions. Guideline softening temperatures for processing amorphous polymers are approximately 90˚F (50˚C) above the glass transition temperature. Some polymers require more heat and some less. Table 21.1 gives the T_g for various resin systems.

Table 21.1 T_g of Various Resins

Material	T_g, °C
PS	101
PP	–10
PE	–120
PC	150
PVC	80
PMMA	105
ABS	115
PET	70
PBT	45
PA 6	50
PA 6,6	55
FEP	70

Thermal conductivity is the heat transfer rate through a material at steady state. It is useful for input to computer-aided engineering (CAE) programs and in determining the time required for heating and cooling different resin systems.

21.9 Time-Temperature Relationship

A specific time-temperature relationship exists for all thermoplastics resins. While the relationship varies from one resin system to another, and even slightly from one product to another within a given resin system, all thermoplastic materials degrade and lose properties if they are heated too hot or held too long a time at high temperature. Previously in this chapter, long-term heat aging was covered and demonstrated how PP loses properties in the solid state if it is held at a high temperature for too long a time. The same situation occurs in the melt state, with the time-temperature relationship being critical during processing. At high polymer melt temperatures, the polymer has a limited time before degradation begins, when the polymer is said to "unzip" and lose properties. Bonds in the polymer backbone rupture, and the polymer breaks down to lower molecular weight entities with lower property performance. Resins also degrade over time when held at above room temperature for an extended time. At these lower temperatures, the time to degradation is very long. Higher-temperature engineering resins withstand temperatures better than the lower-melting commodity resins.

Figure 21.11 shows a typical time-temperature relationship. The area under the curve is where the polymer is stable. At a certain temperature and time combination, any polymer degrades to lower molecular weight. Each polymer has its own characteristic curve.

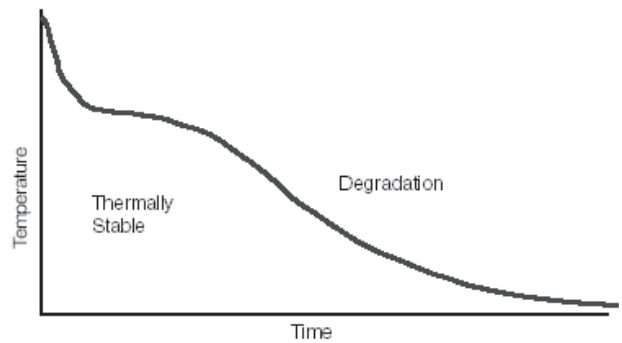


Figure 21.11. Typical time-temperature relationship.

21.10 Izod Impact

Izod impact measures energy required to break a specimen by striking a specific size bar with a pendulum. Izod normally refers to a notched specimen impact. However, in some circumstances unnotched specimens are tested. The data sheet will note that it is an unnotched bar or unnotched Izod. The notch (needs to be machined and not molded into the bar) acts as a stress concentrator, forcing the bar to break at a specific location. A specific size notch (specified in the ASTM D-256 and ISO 180:1993 test methods) is machined into one side of a test bar at a specific distance from the end. As shown in Fig. 21.12, the notch is placed in the clamp and a pendulum is released that impacts the bar, measuring the energy required to break the sample. How tightly the specimen is clamped in the holder plus any flash or defects in the injection-molded specimen can affect the results. Izod samples are normally 0.125 × 0.50 × 2.5 inches (3.2 × 12.8 × 63.5 mm) in length, with the notch machined across the 0.125 inch (3.2 mm) face and into the 0.50 inch (12.8 mm) width. The notch has specific dimensions and radius, as defined in the ASTM and ISO procedures.

Figure 21.12. Izod test configuration.

If the Izod bar is cut from a 5 inch (128 mm) long molded bar, either the gate end or dead end are compared and reported as a group. Different ends of the bars are not mixed in a sample group. If both ends are tested in the same group and compared, the standard deviation and the measurement precision suffer due to potentially different injection-molding and packing conditions during specimen formation. Izod is designated as an impact measurement, since the energy to break a bar is determined. In reality, Izod measures the notch sensitivity rather than being a true impact measurement. The notch acts as a stress concentrator and simulates what occurs if a part is scratched or cut in a specific application, making it easier to break the product. Figure 21.13 shows an Izod test machine and Fig. 21.14 shows the notcher.

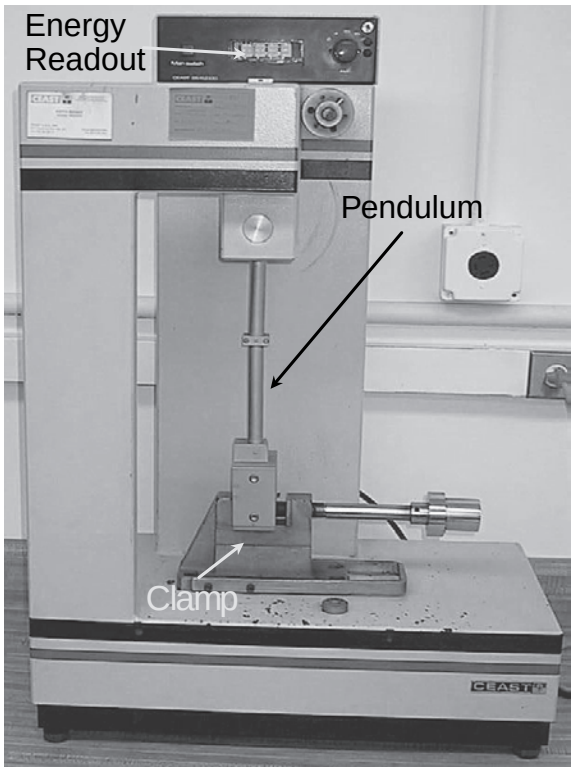


Figure 21.13. Ceast Izod impact testing machine.

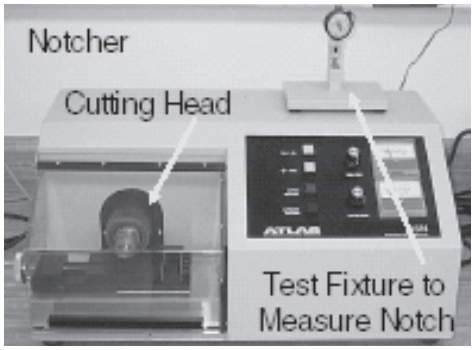


Figure 21.14. Atlas notching machine.

21.11 Charpy

Charpy is measured on the same test equipment used for Izod impact. Like Izod, Charpy measures impact. In the Izod test the specimen is clamped vertically, and in the Charpy test the bar is laid horizontally across two end supports. Then the pendulum swings down to break the bar in the middle. The bar is notched, the same as in the Izod test, and the notch is placed opposite the pendulum impact point. The pendulum is equipped with a sharp point where it impacts the specimen. The energy required to break the specimen is obtained by comparing how far the pendulum swings when no bar is present to how far the pendulum swings after it has impacted the bar. The difference in pendulum swing is the energy absorbed by the specimen. The testing equipment and specimen holder to measure Charpy are shown in Fig. 21.15. The test specimen has the same dimensions used in the Izod test with the exception that the notch is opposite the pendulum impact.

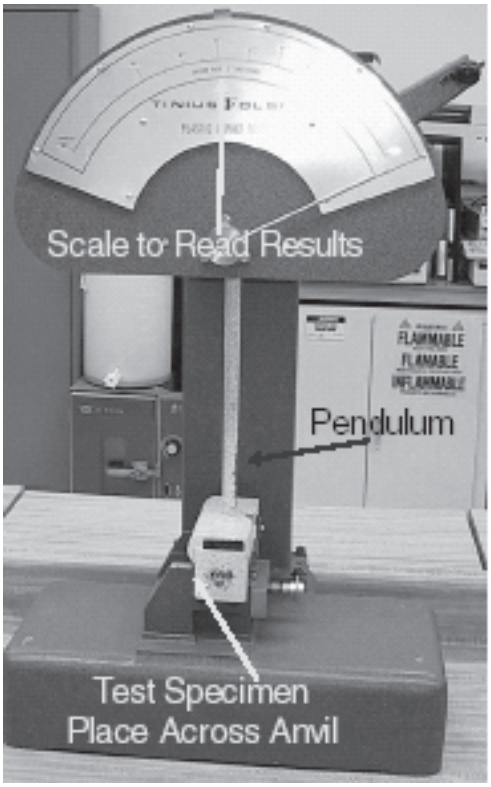


Figure 21.15. Charpy testing setup.

There is also a flatwise Charpy, where an unnotched bar is laid across the sample holder sitting on the thin edge (0.125 inch [3.2 mm]) instead of lying on the wide dimension (0.50 inch [12.8 mm]). The swinging pendulum strikes the 0.50 inch (12.8 mm) face of the bar.

ASTM D-6110-97 and ISO 179-2:1997 are the Charpy test methods.

21.12 Comparative Thermoplastic Properties

Table 21.2 compares physical properties for common unmodified thermoplastic resins. There are numerous polymer and additive modifications available. To identify a material with a specific property profile for a particular application, a comprehensive database review is required.

21.13 Polymer Additives

All resin systems contain additives or stabilizers to improve the processability and the processing window. Processing stabilizers include

- Thermal or heat stabilizers for improved processing
- Antioxidants to improve the oxidative stability during processing at high temperature and/or for high temperature end-use applications
- Lubricants, both internal and external, to improve flow and minimize adhesion to metal parts
- Plasticizers

PVC uses different plasticizers and concentrations to change the resin characteristics from rigid to flexible.

Other additives are used to modify the polymer system and add value to increase the utility and applications for each polymer system. Some additives that increase polymer value and application viability include

- UV (ultraviolet) stabilizers to improve outdoor exposure and weatherability
- Pigments and dyes for coloring the base resin
- Antiblocking agents to prevent films from sticking together
- Flame retardants to improve flammability ratings
- Impact modifiers to improve resin toughness
- Fillers and extenders to reduce cost and increase performance
- Fiber reinforcements to improve stiffness and toughness

Processing issues introduced with compounded additives are

- Reduced thermal stability
- Plate-out on the die and downstream equipment
- Blooming

Reduced thermal stability is caused by introducing an additive package (flame retardants, impact modifiers, etc.) that does not have the same processing temperature range as the virgin resin. Particularly susceptible additives are flame retardants, which may exhibit lower thermal stability than the polymer.

Plate-out is caused by additives volatilizing during processing and condensing on the take-up (takeoff) equipment or dies. In sheet and film processes, the roll stack may have to be cleaned repeatedly to eliminate deposit build-up. Deposits can build up on the die lips, where the plastic exits the extruder. Any build-up has to be removed to prevent it from contaminating the product.

Blooming is related to additives with a high volatility and a low compatibility with the resin system. These additives migrate to the surface and form a chalky layer. The chalky layer may appear between 12 and 72 hours after extrusion and is easily wiped off.

21.14 Drying

Some polymer systems are hygroscopic, meaning they absorb moisture from the air, and must be dried prior to processing. Drying is covered in detail in Part 5, "Auxiliary Equipment." In some polymer systems, moisture causes a hydrolysis reaction to occur in the

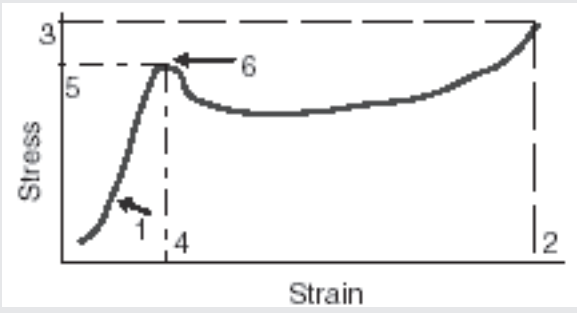
Table 21.2. Comparison of Physical Properties of Various Resins

Property	ABS	LDPE	HDPE	PS	PVC	PP	Nylon	PET	PC
Tensile strength, psi X 10^3	4.0–6.0	1.4–2.8	1.4–2.0	2.7–6.0	2.4–4.4	4.5–5.5	6.5–8.0	8.0	9.0
Flexural modulus, psi X 10^5	1.3–4.2	0.4–0.75	1.45–2.25	3.8–4.9	3.0–5.0	1.7–2.5	1.85–4.2	2.90	3.40
Notched Izod, ftlbs/in	2.5–12	1.0–9.0	0.4–4.0	0.35–0.45	1.0–8.0	0.4–1.0	0.8–2.1	1.7	12–16
Useful temperature range, to °F	190	100	130	150	120	170	140	130	220
Resistance to chemicals	Poor	Excellent	Excellent	Poor	Good	Good	Excellent	Excellent	Medium–Poor
Flammability	Burns	Burns	Burns	Burns	Flame Resistant	Burns	Flame Resistant	Flame Resistant	Flame Resistant

extruder that lowers the polymer molecular weight by reacting with the polymer backbone. This bond scission leads to lower property performance. Moisture-sensitive polymers that undergo hydrolysis reactions in the extruder when moisture is present include nylon, PET, PBT, PC, and acrylics. Other polymers, such as ABS, polyetherimide, polysulfone, polyphenylene oxide/polystyrene blends, and PPS, require drying to remove absorbed moisture. Extruding high moisture content polymers in unvented extruders causes steam to exit the die and produce a foamy structure with microvoids, surface defects, and lower property performance.

Review Questions

1. What are the two classes of thermoplastic resin, how do they differ, and what resins are in each class?
2. What is meant by time-temperature relationship? Why is it important to extrusion?
3. Why is it important to understand plastic physical property testing and how is it related to extrusion?
4. What is the difference between specific gravity and density?
5. How are tensile properties measured and what do they indicate?
6. Your company is extruding a profile that is 28 in^3 per piece after the pieces are cut. You expect to sell 575,000 profile pieces this year. Currently you are using a resin that costs $2.32 per pound and has a specific gravity of 1.21. Today a salesman came into your facility with a new resin called "Polysolvesall" that has the same properties as the resin you are currently running, except the salesman is going to give you a fabulous deal and sell it to you for $2.12 per pound. Polysolvesall has a specific gravity of 1.41. Do you excitedly run out, tell your boss you have just made a deal and signed a long-term contract for "Polysolvesall" that saves you $.20/pound, and that you are going to save your company umpteen thousands of dollars this year and next? Explain your reasoning.
7. Your company uses two PP resins with an MFI of 12 and an MFI of 35. What is the difference in the resins and which one do the operators prefer to run and why?
8. Identify the points on the following tensile stress-strain curve:



Review Questions (continued)

9. Describe a flexural test and what it measures.
10. What is HDT and how is it measured?
11. What is the difference between Izod and Charpy impact tests?
12. If you have 250 pounds of 0.949 density PE and it is blended with 736 pounds of 0.912 density, what is the anticipated density of the blend after it is extruded?
13. List three key thermal properties and tell why each is important.
14. What are some polymer additives and what do they contribute to the polymer system? Name at least five.
15. What are some possible negative effects that can occur with different stabilization additives?

22 Processing Recommendations for Various Resin Systems

Extrusion processing condition guidelines for different resin systems are summarized in this chapter. These conditions are good starting points for the various resin systems and may require optimization for a specific extrusion process to obtain the maximum throughput rate while meeting or exceeding all the quality criteria. Optimizing a particular extrusion operation depends on

- Extruder type—single or twin screw
- Extruder size—small or large
- Particular screw design being used
- Die design
- Resin melt flow
- Throughput rate
- Downstream process
- Resin type and formulation
- Additives in the resin
- Environmental factors around the extruder

22.1 Acrylonitrile Butadiene Styrene

Acrylonitrile butadiene styrene (ABS) is an amorphous resin. Its properties depend on the acrylonitrile, butadiene, and styrene monomer ratio and how these components are polymerized and formulated. For good chemical resistance, heat resistance, and long-term thermal stability, the formulation needs to be rich in acrylonitrile content. Formulations high in styrene content have good gloss, excellent moldability, and good strength and rigidity, while butadiene contributes low temperature impact and general property retention. ABS is manufactured two ways:

- In the first method, copolymers of styrene/acrylonitrile (SAN) are blended with butadiene/acrylonitrile rubber (NBR). Normally the rubber phase is broken up and uniformly dispersed in small particles as the discontinuous phase in the SAN continuous phase.
- The second manufacturing approach, used in most operations today, provides more versatility and ability to custom design specific property profiles. Acrylonitrile is grafted (chemically bonded) onto a butadiene or butadiene styrene copolymer backbone.

Polybutadiene–SAN blends are incompatible. If the two materials are blended, the polybutadiene has to be uniformly dispersed in the SAN continuous phase as very small particles. Even with the proper concentration and polybutadiene distribution, the toughness of the SAN matrix is only marginally improved due to resin incompatibility. However, if the proper styrene-acrylonitrile ratio is grafted onto a polybutadiene backbone, the polymer becomes compatible with the SAN copolymer; the two resin systems can be blended, with the resulting polymer alloy having improved toughness.

The properties anticipated for ABS depend on the ratio of the three components in a particular formulation, shown in Table 22.1. As with most polymer resin systems, all properties are not maximized in one formulation; it is always a give and take situation. If the formulation has a high butadiene concentration, the impact is good. High styrene concentrations improve processability, while high acrylonitrile concentrations improve chemical resistance and hardness. Tensile and flexural properties tend to decrease as impact increases.[1] Other monomers can be incorporated in the polymer molecule to improve the performance profile, such as substituting alpha methyl styrene for styrene to raise the heat deflection temperature.

Table 22.1. Property Ranges for ABS

Property	Maximum	Minimum
Specific Gravity	1.07	0.99
Tensile Strength, psi	11,000	2,500
Flexuran Modulus, psi	500,000	100,000
Flexural Strength, psi	15,000	2,000
Izod, ftlbs/in	13.0	0.8
Hardness, Rockwell R	123	
Hardness, Shore D		62
HDT @ 264 psi, ˚C	110	63

ABS is mildly hygroscopic. The moisture absorbed is directly related to the relative humidity in the surrounding air. Moisture absorbs on the pellet surface and migrates into the center. As ABS is not shipped in moisture-proof containers, the pellets must be dried prior to processing. Pellets that have been dried will reabsorb moisture if they are exposed to the atmosphere. Consequently, once pellets are dried, they need to be used or stored in a moisture-proof container for later use. The absorption rate depends on the relative humidity, pellet or particle size, temperature, time, and the ABS grade. Smaller particles and higher temperatures increase the absorption rate. Higher acrylonitrile grades absorb more

moisture, yielding higher equilibrium moisture. Typical equilibrium is 0.25–0.65% at 50% relative humidity, and 0.55–1.4% at 80% relative humidity. Improper drying leads to splay on the extruded part surface or, in severe cases, bubbles and foamy products (products with small internal voids). In sheet or film extrusion, moisture may cause holes to be blown in the product as steam exits the extruder. All pellets and powder need to be dried uniformly. A few pellets containing moisture in an otherwise dry feedstock can cause processing problems.

Typical drying conditions for ABS extrusion grades are 190–200˚F (88–93˚C) for four hours after the pellets have reached the drying temperature. A dehumidifying desiccant dryer with a dew point below –30˚F (–34˚C) is required for most drying. Moisture guidelines for extrusion are <0.02%. Resin should be tested for moisture content before running any extrusion. Care must be taken not to exceed the 200˚F (93˚C) drying temperature, or the resin surface can soften and stick together to form a solid mass in the dryer that is difficult to remove. Lower drying temperatures require too long a time to dry the resin properly. For example, if it takes four hours to dry the resin properly at 190˚F (88˚C), it takes eight hours to dry to the same equilibrium moisture content at 170˚F (77˚C). Drying time takes longer with pellets and regrind than with powder because moisture absorbed into the particle has to migrate through longer diffusion paths in large size particles. Drying rates are determined by the diffusion process. Larger particles take longer to dry because the diffusion path is longer. The second factor affecting drying time is the diffusion rate, which slows down as the pellets become dryer.

22.1.1 Extrusion Temperature Profile

The temperature profile given in Table 22.2 for ABS depends on the monomer ratio in the formulation. Higher acrylonitrile content requires a higher processing temperature range. The desired ABS melt temperature is 430–465˚F (221–240˚C). Flame retardant ABS is not as thermally stable as standard ABS grades, and the melt temperature should not exceed 475˚F (246˚C).

Table 22.2. ABS Extrusion Temperature Guidelines

Condition	Setting, ˚F (˚C)
Zone 1	350–380 (177–193)
Zone 2	390–420 (199–216)
Zone 3	410–440 (210–227)
Zone 4	420–465 (216–240)
Zone 5	420–465 (216–240)
Adapter	420–465 (226–240)
Die	410–465 (210–240)

ABS can be extruded on most single screw extruders with 24:1 *L/D* or greater and a screw compression ratio between 2.25:1 and 2.7:1. Twin screw extrusion using corotating twin screw extruders is straightforward, provided the screw is designed to melt the material without excessive shear heating. As long as melt temperatures are 430–465˚F (221–240˚C), no unusual extrusion problems are anticipated. Moisture removal via a vacuum vent is possible without causing a hydrolysis reaction that degrades the polymer backbone. The mixing section chosen in single screw extrusion depends on the mixing required. Tight tolerances associated with Maddock mixing heads are not recommended for ABS.

In sheet and film extrusion, die land length recommendations depend on product thickness. Table 22.3 recommends land lengths for coat hanger dies versus the sheet or film thickness.

ABS melt viscosity exhibits both shear and temperature dependence. At extrusion shear rates, increasing the melt temperature from 420 to 480˚F (216–249˚C) reduces the melt viscosity approximately 1.8 times. Increasing the shear rate from 100 to 400 sec^{-1} reduces the melt viscosity 2.0 times. Over this processing range, the melt viscosity is almost equally affected by changes in shear rate or temperature.

Table 22.3. Land Lengths for ABS Sheet and Film Dies

Sheet Thickness, inch (mm)	Land Length, inch (mm)
0.010–0.060 (0.254–1.524)	0.75–1.0 (19.05–25.40)
0.060–0.120 (1.524–3.048)	1.0–2.0 (25.40–50.80)
0.120–0.250 (3.048–6.350)	2.0–3.0 (50.80–76.20)

22.2 Polyamide

Polyamides (PA), or nylons, cover a wide range of products. There are both crystalline and amorphous forms. All polyamide molecules have a characteristic amide group (–CO–NH–) alternating with a $(CH_2)_n$ group in the polymer backbone. The common crystalline polyamides are

- Nylon 6
- Nylon 6,6
- Nylon 6,10
- Nylon 6,12
- Nylon 11
- Nylon 12

Their chemical structures are shown in Fig. 22.1. Other crystalline polyamide materials include nylon 4,6 and nylon 6,9. Amorphous nylons such as Trogamid T, sold by Creanova Inc., and Amidel, sold by Amoco Polymers, are discussed later in this section under processing.

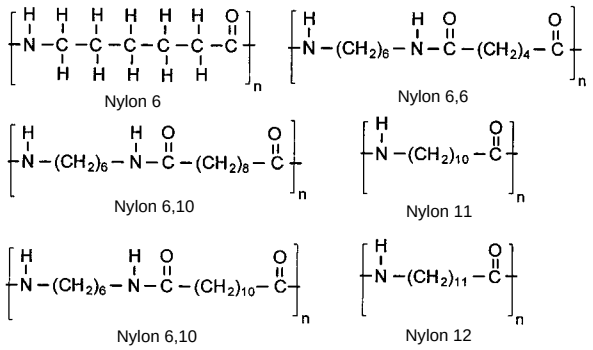


Figure 22.1. Structures of some common crystalline polyamide resins.

Table 22.4 compares nylon 6 and nylon 6,6 properties.[2] Nylon 6,6 absorbs less moisture (2.8%) than nylon 6 (3%) at 73˚F and 50% relative humidity. As nylon 6,6 melts at 504˚F (262˚C) and nylon 6 melts at 428˚F (220˚C), nylon 6,6 has a higher HDT and continuous use temperature. Nylon 6 is more dimensionally stable than nylon 6,6 due to its lower crystallinity. Nylon 12 has the highest crystallinity of the nylon resins, which gives it better chemical resistance. Nylon 12 melt temperature is lower than the other nylons, resulting in lower temperature performance as measured by HDT and continuous use temperature.

Table 22.4. Property Comparison of Nylons 6 and 6,6

Property	Comparison
Specific Gravity	6,6 > 6
Heat Resistance	6,6 > 6
Moisture Absorption	6 > 6,6
Toughness	6 has higher impact
Stiffness	6,6 has higher tensile strength
Dimensional Stability	6 > 6,6
Melt temperature	6,6 > 6

Nylon 6,10 has lower water absorption at equilibrium than nylon 6 or nylon 6,12 and is tougher than nylon 11 or nylon 12, with better low temperature toughness than nylon 6 or nylon 6,6. Nylon 11 has similar mechanical properties to nylon 6,6, poorer property retention at slightly elevated temperatures, and lower water absorption characteristics.

22.2.1 Drying

As polyamides are hygroscopic and absorb high levels of moisture quickly, it is critical to properly dry polyamides and prevent the hydrolysis reaction that occurs in molten PA in the presence of water. The hydrolysis reaction between water and the amide group degrades the polymer, breaking bonds and lowering the molecular weight. In addition, overdrying polyamide may have a negative effect on extrusion processing because moisture acts as a plasticizer for nylon. Drying nylon 6,6 below 0.08% moisture reduces the plasticizing effect and hinders its flow. Table 22.5 compares the equilibrium moisture content at saturation and at 50% relative humidity for common nylons. As carbon atoms between the amide groups increase, the equilibrium moisture content decreases. Equilibrium moisture content for nylon 6 is seven times nylon 12.

Table 22.5. Nylon Moisture Content

Nylon	Equilibrium Moisture, 50% Relative Humidity	Equilibrium Moisture, % Saturation
6	3.0	9.5
6,6	2.8	8.0
6,10	1.5	4.0
6,12	1.3	3.0
11	0.8	1.9
12	0.7	1.4

Polyamide moisture absorption produces a dramatic effect on both dimensions and physical properties. The amide group (–CO–NH–) hydrogen bonds with water. The bond is between oxygen in water and hydrogen on the nitrogen, and between hydrogen in water and oxygen on the carbon. As moisture is absorbed and bonds to the polymer backbone, the plastic swells. Dimensional changes in nylon 6 up to 0.7% occur in molded parts as the moisture increases from <0.2% to 2.8% at 50% relative humidity. This change may occur over 150 days for a part that is 0.060 inch (1.5 mm) thick. At saturation, growth is significantly larger. The dimensional changes anticipated with nylon 6,6 and nylon 11 at 50% relative humidity are 0.6% and 0.12%. Moisture absorption and matrix swelling are reversible changes. Placing a wet sample in an oven drives the moisture out, returning it to its original dimensions. The moisture absorption rate and removal depend on the nylon and the relative conditions surrounding the part. Consequently, drying, moisture absorption, moisture level, and the nylon used in extrusion are critical.

The moisture level in nylon 6,6 affects its physical properties. Since moisture is a plasticizer, higher moisture content raises impact properties and lowers the tensile and flexural properties. Consequently, the moisture content during testing must be controlled in order to provide reproducible data and be able to compare data from one lab to another.

A dehumidifying desiccant dryer is the recommended equipment for drying nylon 6 prior to extrusion. To maintain the plasticizing effect, nylon 6 should not be

dried below 0.02% moisture. For extrusion, the normal moisture content is 0.08%. For less than 20% reinforcement, maintain 0.08–0.15% moisture, and 0.08–0.13% when there is more than 20% reinforcement. The recommended drying temperature is 180˚F (82˚C). Higher drying temperatures remove moisture at a faster rate and lead to oxidative thermal degradation and discoloration. If resin is supplied in moisture-proof containers, there may be no need to dry the resin prior to extrusion. When using a moisture-sensitive resin, one should do a moisture analysis prior to extrusion to determine if drying is necessary. Once a container or bag is opened, it must be dried if it is not used immediately.

Recommendations for drying nylon 6,6 prior to extrusion are to use a dehumidifying desiccant dryer with a 0˚F (–18˚C) or lower dew point and 175˚F (80˚C) air temperature for four hours. Air temperatures higher than 200˚F (95˚C) will discolor nylon 6,6, and it should not be held at 200˚F for more than three hours. Nylon 6,6 may be shipped from the manufacturer in moisture-proof containers and need no drying prior to extrusion if the bag is sealed. In nylon 6,6 extrusion, the resin should contain less than 0.2% moisture. As with drying other resins, the drying system must be airtight to prevent room air that contains moisture from coming in contact with the dried pellets. If nylon moisture is very high, it has to be dried at low temperatures for long times to prevent oxidative thermal degradation and discoloration. Table 22.6 gives nylon 6,6 drying times for different exposure conditions.

Table 22.6. Drying Times for Nylon 6,6

Condition	Drying Time, hours*
Freshly Opened Bag with Moisture-Proof Liner	3
Bag Opened for Several Days	20
Regrind Not Stored in Moisture-Proof Container	60

22.2.2 Extrusion Conditions

Single screw nylon 6 extruders should have at least 20:1 and preferably 24:1 *L/D*, with a screw compression ratio between 3.5:1 and 4.0:1. The metering section length represents 25 to 40% of the screw, with the transition section having between three and six flights. The feed section occupies the remaining length. Temperature guidelines for the extruder barrel profile and melt temperature are given in Table 22.7. Recommended temperatures depend on the extrusion process and throughput rate. Draw ratios for wire and coating operations are between 8 and 10:1. For tubing, the vacuum sizer is between 3 and 9% larger than the final pipe or tube diameter, with the die size relative to the finished tube diameter being 1.5 to 2.0:1.

Nylon 6 viscosity versus shear rate curves show that shear and temperature play significant roles in changing the melt viscosity. Increasing the shear rate from 100 to 400 sec^{-1} decreases the viscosity 1.9 times. Increasing the melt temperature from 446˚F (230˚C) to 482˚F (250˚C) lowers the viscosity 1.6 times.

Recommended extrusion conditions for nylon 6,6 are similar to nylon 6. A single screw nylon 6,6 extruder should have at least 20:1 and preferably 24:1 *L/D*, with a 3.5:1 compression ratio. Temperature guidelines for the extruder barrel profile and melt temperature are given in Table 22.7. Nylon 6,6 viscosity versus shear rate data are similar to nylon 6 in that viscosity is shear and temperature-dependent, with shear having a slightly higher influence. Increasing the shear rate from 100 to 400 sec^{-1} decreases the viscosity 2.2 times. Increasing the melt temperature from 500˚F (260˚C) to 550˚F (288˚C) in the same shear rate range reduces the melt viscosity 1.7 times.

Single screw nylon 12 extruders should have from 20:1 to 25:1 *L/D*, with a screw compression ratio between 2.5:1 and 3.5:1. The metering section flight depth should be less than 0.06*D* to ensure a homo-

Table 22.7. Processing Guidelines for Some Crystalline Polyamide Resins

Resin	Process	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Adapter, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp.,˚F (˚C)
Nylon 6	Wire Jacket	430–480 (221–249)	480–520 (249–271)	510–540 (266–282)	510–540 (266–282)	510–550 (266–288)	510–550 (266–288)	510–550 (266–288)
Nylon 6	Monofilament, Tubing Cable	470–500 (243–260)	460–500 (238–260)	460–500 (238–260)	460–500 (238–260)	440–500 (227–260)	440–500 (227–260)	440–500 (227–260)
Nylon 6,6	Tubing, Rod, Pipe, Profile	480–500 (248–260)	520–560 (271–293)	520–560 (271–293)	520–560 (271–293)	520–560 (271–293)	440–500 (271–293)	440–500 (271–293)
Nylon 11	Tubing, Cable, Profile	330–360 (166–182)	370–400 (188–204)	380–420 (193–216)	380–420 (193–216)	380–420 (193–216)	380–420 (193–216)	380–420 (193–216)
Nylon 12	Tubing, Film, Jacketing, Shapes	310–340 (154–171)	350–380 (177–193)	370–410 (188–210)	370–410 (188–210)	370–410 (188–210)	370–410 (188–210)	370–410 (188–210)

geneous melt. Higher extrusion rate is accomplished with a screw having metering and transition sections with only a 15*D* transition and a 10*D* metering on a 25:1 *L/D* machine. Temperature guidelines for the extruder barrel profile and melt temperature are given in Table 22.7.

22.2.3 Amorphous Polyamide

Unlike crystalline polyamides, which have sharp, well-defined melting points with sharp decreases in viscosity above the melting point, amorphous polyamides like Trogamid T behave more like polycarbonate or ABS than nylon 6 or 6,6. Amorphous nylons are transparent with good impact, high stiffness, and good dimensional stability. As with other polyamides, amorphous nylon readily absorbs moisture and must be dried prior to extrusion. Amorphous nylon resins are shipped dry, in moisture-proof containers. If opened and allowed to absorb moisture, the resins can take up to 10 hours to dry in a dehumidifying desiccant drying oven at 230–248˚F (110–120˚C).

Single screw extrusion guidelines are to have a 20:1 to 30:1 *L/D* extruder with a 1.15–1.25:1 compression ratio screw. Temperature profile guidelines from the feed throat to the adapter are 482/500/509/518/518˚F with the die at 536˚F (250/260/265/265/270/270˚C with the die at 280˚C).

22.3 Polyethylene

Many polyethylene (PE) grades are available at different densities (crystallinities) and molecular weights. Table 22.8 shows polyethylene classifications based on density. Table 22.8 does not include linear low density PE (LLDPE). PE properties and performance are directly related to the density (crystallinity) and molecular weight. Table 22.9 shows the effect of density and melt index on PE property performance. Linear low density is not shown in Table 22.9 because it has different branching and structure, which affect the property performance.

Table 22.8. ABS Extrusion Temperature Guidelines

Type	Density, g/cm^3
Low	0.910–0.925
Medium	0.926–0.940
High	0.941–0.959
Very High	0.959 and Higher

Polyethylene is not hygroscopic and does not pick up moisture. Typically, no drying is required prior to extrusion unless the pellets or powder have been outside and become wet, or the resin has been stored in a cold warehouse and brought into a hot environment, where moisture can condense on the cold surface.

22.3.1 Extrusion Conditions

Many extrusion conditions are used to process PE. Processes using PE are

- Compounding
- Sheet
- Cast film
- Blown film
- Coextrusion
- Extrusion blow molding
- Wire and cable
- Monofilament
- Profile
- Pipe

With excellent thermal stability, PE can be processed over a wide temperature range without thermal degradation. PE is produced in a wide range of melt flows. This allows viscosity matching with other materials in coextrusion. Single and twin screw extruders can process PE with many different screw designs. As PE has great commercial significance, much effort has been spent optimizing extrusion processing conditions.

LDPE and HDPE process very well on 3.1–3.5:1 compression ratio screws, fitted with a high shear mixing head (Maddock mixer) near the end of the extruder or

Table 22.9. Correlation of Properties with Density and Melt Index

Maximum Physical Property	Density*	Melt Index**
Softening Point	Higher	Lower
Melting Point	Higher	Lower
Tensile Strength Yield	Higher	Lower
Tensile Strength Failure	Higher	Lower
Tensile Elongation	Higher	Lower
Resistance to Creep	Higher	Lower
Flexural Modulus	Higher	Lower
Film Impact Strength	Higher	Lower
Resistance to Warpage	Higher	Lower
Resistance to Shrinkage	Higher	Lower
Transparency	Higher	Lower
Resistance to Environmental Stress Cracking	Higher	Lower
Resistance to Grease and Oils	Higher	Lower
Film Draw Down	Higher	Lower

* Higher density correlates with higher crystallinity within the range of 0.910 to 0.959.

** Melt index correlates with molecular weight.

with barrier flights in the metering zone. LLDPE runs best with a variable-pitch screw. PE extruder temperature profile guidelines, given in Table 22.10, are very general, as polyethylene has such a wide temperature processing range. Polyethylene can be processed at high temperatures because it has good thermal stability, or low temperatures because it melts at low temperatures. Table 22.11 provides melt temperature guidelines for different melt flow PE resins used in different processes. However, similar to Table 22.10, these are initial recommendations, and each process needs to be optimized for the particular resin, screw design, and extruder. Depending on the process, the resin requirements change. Extrusion blow molding requires good melt strength, while profile requires good melt strength, good flow into thin section, plus surface quality. Melt temperature used in pipe extrusion depends on the pipe diameter, which affects resin distribution in the die. Thick sheet requires good melt strength, dictating lower melt temperature for better processability. Depending on the process, the temperature guidelines given in Table 22.10 may have to be modified to optimize the extrusion process being used.

Longer extruders, 30:1 and 32:1 *L/D*, have been used recently to improve throughput and provide a homogeneous melt to the die in pipe extrusion. There is no specific screw design recommended for PE. Both single and two-stage screws with and without mixing devices perform satisfactorily, provided the compression ratio is correct.

PE viscosity is generally more shear- than temperature-dependent. Table 22.12 shows the effect of shear rate (100–400 sec^{-1}) and temperature (range specified in the table) on viscosity. Table 22.12 shows that for the shear rate range, the viscosity decrease is 2.1–2.4 times, and for the specified temperature range, the viscosity decrease is 1.3–1.5 times.

22.4 Polymethyl Methacrylate (Acrylic)

Acrylics are amorphous polymers with excellent optical properties, UV resistance, and transparency that are extruded into sheet, profile, and stock shapes for numerous applications. Rolls on sheet takeoff equipment are embossed to produce special finishes or patterns on the final sheet. Acrylics are easily processed on single screw extruders equipped with either a single stage 2:1 compression ratio screw or a two-stage screw for venting. Acrylics have poor chemical resistance, are brittle, have high notch sensitivity, and burn easily.

Acrylics absorb moisture and must be predried prior to extrusion if exposed to the atmosphere or if the resin is not shipped in moisture-proof containers. Equilibrium moisture content for acrylics exceed 1.0% at 95˚F and 80% relative humidity. Without proper drying, extruded shapes exhibit either splay on the surface, a foamy structure and internal bubbles, or holes in the extrudate, depending on the moisture level and part thickness. A dehumidifying desiccant dryer with dew point less than –20˚F (–29˚C) is required. To produce transparent sheet with good optical properties, gloss, and surface aesthetics, the resin has to be less than 0.05% moisture. Using outside air will not provide sufficient drying. A moisture analyzer is required to provide moisture contents prior to extrusion processing.

Drying at 175–185˚F (79–85˚C) for two to three hours removes approximately 0.1% moisture. Consequently, the drying time required

Table 22.10. Extrusion Temperature Guidelines for Polyethylene

Material	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)
LLDPE	280–325 (138–163)	325–365 (163–185)	360–390 (182–199)	380–410 (193–210)	400–420 (204–216)
LDPE	300–340 (149–171)	340–365 (171–185)	360–390 (182–199)	360–390 (182–199)	360–390 (182–199)
HDPE	300–340 (149–171)	350–390 (177–199)	390–420 (199–216)	390–420 (199–216)	390–420 (199–216)

Table 22.11. Melt Temperature Guidelines for Different MFI and Processes

Process	Resin MFI	Melt Temp., ˚F (˚C)
Pipe and Small Conduit	0.3–0.7	380–420 (193–216)
Large Diameter Pipe	to 12	450–500 (232-260)
Blown Film Bags	to 12	410–450 (210–232)
Sheet	4 to 7	390–425 (199–219)

Table 22.12. Effect of Shear and Temperature on Polyethylene Melt Viscosity

Material	Viscosity Decrease Between 100 and 400 sec^{-1}	Temperature Range, ˚F(˚C)	Viscosity Decrease over Temperature Range
LLDPE	2.4X	374–482 (190–250)	1.5X
LDPE	2.3X	374–482 (171–210)	1.5X
HDPE	2.1X	356–428 (180–220)	1.3X

depends on the initial moisture content, the dehumidifying dryer dew point, and the final moisture content required to produce quality parts. Higher drying temperatures may be used to decrease the drying time; however, excessive temperature may soften the amorphous resin, allowing the pellets to stick together and become a solid mass in the dryer. The drying temperature to use depends on the resin molecular weight. High-molecular-weight, low-melt-flow resins can be dried at higher temperatures. A 2 MFI acrylic resin can be dried at 200˚F (93˚C), a 6 MFI resin at 185˚F (85˚C), a 15 MFI at 175˚F (79˚C), and a 30 MFI at 165˚F (74˚C).

22.4.1 Extrusion Conditions

A 2.0:1 compression ratio, constant-pitch screw in a 20:1 to 30:1 *L/D* single screw extruder with no screen pack is used to extrude acrylic shapes. Typical processing conditions based on resin melt flow are given in Table 22.13. The maximum draw ratio is approximately 1.15:1. For good gloss and surface quality, roll temperatures in a three-roll stack should be as hot as possible. Determine the temperature where the sheet just starts to stick to the roll and back off approximately 10˚F (6˚C). A good starting temperature is 175˚F (79˚C).

22.5 Polypropylene

Polypropylene (PP) is a versatile thermoplastic resin. It is a low-cost, semicrystalline resin with good strength and stiffness that allows it to outperform PE in numerous applications. PP is easily compounded with fillers and fiber reinforcements to produce properties similar to thermoplastic engineering resins. PP is used in many diverse applications, such as

- Automotive interiors
- Electrical applications
- Fibers
- Glass reinforced sheet
- Packaging
- Medical applications
- Appliances
- Lawn and garden applications

This rapidly crystallizing resin, which complies with FDA regulations, combines

- Outstanding heat resistance
- Chemical resistance
- Impact resistance
- Stiffness

Polypropylene does not absorb moisture and does not require drying prior to extrusion unless the resin has been moved from a cold to a hot environment and moisture condenses on the pellets. A vacuum vented extruder can remove this moisture under normal processing with no detrimental effects to the resin.

22.5.1 Extrusion Conditions

Polypropylene extrusion is relatively straightforward. It has good melt strength, low processing temperature, and a large processing window. PP viscosity is more shear- than temperature-dependent, meaning the viscosity changes more with shear than with temperature. Increasing the shear rate from 100 to 400 sec^{-1} reduces the melt viscosity 2.6 times. Over the same shear rate range, increasing the melt temperature from 400˚F (204˚C) to 500˚F (260˚C) decreases the viscosity by 1.5 times. PP viscosity shows significantly more shear than temperature sensitivity. Consequently, PP has a similar viscosity over a wide temperature range, and the best way to significantly alter the viscosity is by changing the screw configuration or screw speed. In single screw extrusion, changing the screw speed changes the throughput (assuming the extruder is flood fed); therefore, in most situations the viscosity is relatively constant because the screw speed and throughput rates are fixed. In twin screw extrusion, the screw speed can be changed to alter the viscosity.

Polypropylene processes best on a 24:1 *L/D* or longer single screw extruder with a 3.5–4:1 compression ratio, using a long feed section (50% of the screw flights) followed by a short transition section (17% of the flights) and 33% of the flights in the metering zone. Polypropylene screws are equipped with a tight barrier in the metering section or a mixing head (Maddock mixer) at the end of the metering section. Melt uniformity is improved by increasing the backpressure through screens or valving at the end of a single screw extruder.

Table 22.13. Processing Guidelines for Various MFI Acrylic Resins

MFI	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp ˚F (˚C)
2	320–360 (160–182)	360–400 (182–204)	400–440 (204–227)	400–440 (204–227)	400–440 (204–227)	400–440 (204–227)
6	320–360 (160–182)	340–380 (171–193)	380–420 (193–216)	380–420 (193–216)	380–420 (193–199)	380–420 (193–216)
8	310–350 (154–177)	340–360 (171–182)	360–380 (182–193)	360–400 (182–204)	360–400 (182–204)	360–400 (182–204)

The temperature profile depends on the MFI. The melting point is approximately 335˚F (168˚C). High-molecular-weight or low-melt-flow resins (fractional–10) are run at high melt temperatures in some applications. PP resins with 75–300 MFI are run at much lower melt temperatures. Handling characteristics of the melt and the melt strength depend on the resin viscosity. Table 22.14 gives temperature guidelines for processing different MFI PP.

Table 22.14. Extrusion Temperature Guidelines for Various Melt Flow Index Polypropylene

MFI	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp ˚F (˚C)
0.2–10	340–380 (171–193)	410–450 (210–232)	440–470 (227–243)	440–500 (227–260)	440–500 (227–260)	450–550 (232–288)
12–35	320–360 (160–182)	360–400 (182–204)	380–420 (193–216)	380–420 (193–216)	400–450 (204–232)	400–450 (204–232)
50–300	300–335 (149–168)	340–360 (171–182)	350–380 (177–193)	350–380 (177–193)	350–380 (177–193)	350–380 (177–193)

22.6 Polycarbonate

Polycarbonate (PC) is a high temperature, amorphous resin with

- Good transparency
- Outstanding optical properties
- High impact strength
- Outstanding flame retardance
- Superior dimensional stability
- Good thermal stability
- High stiffness and toughness

It is used in

- All types of glazing and is particularly effective in high vandalism areas
- Automotive products
- Lawn and garden products
- Appliances
- Office machines
- Medical products
- Lighting products
- Alloys such as ABS/PC and PBT/PC

Polycarbonate is known for its high impact strength and clarity and is used to make bullet-proof sheet.

Polycarbonate is a polyester and, like other polyesters, absorbs moisture. It must be dry prior to extrusion, as moisture leads to a hydrolysis reaction in the extruder. This reaction breaks bonds along the polymer backbone, lowering the molecular weight and reducing impact properties. A dehumidifying desiccant dryer with a –20˚F (–29˚C) dew point operating at 250˚F (121˚C) oven air temperature is required for proper drying. Polycarbonate equilibrium moisture content is approximately 0.35%. Acceptable moisture content for extrusion is 0.02%. Depending on the pellet size, drying normally requires three to four hours at the recommended temperature. However, larger pellets, regrind, or glass-filled resins may require four to six hours. Always check moisture content prior to starting extrusion to verify that moisture is less than 0.02%. Dryer size depends on the extrusion rate and the pellet size. At 300 pounds/hour, with a larger pellet requiring five hours to dry, the hopper is filled and resins dried for five hours minimum, after the dryer comes up to temperature. As resin is drawn off the bottom, new resin is added to the dryer top. To maintain the five-hour drying time, the dryer capacity must be at least 1500 pounds.

22.6.1 Extrusion Conditions

For unvented single screw extruders, a 24:1–30:1 *L/D* ratio is recommended to prevent surging associated with shorter barrels. Longer barrels may generate too much frictional heat. In vented extruders, a 30:1–34:1 *L/D* barrel is recommended. A 2.5:1 compression ratio screw with a constant, square pitch and a flight depth in the feed zone that is approximately 0.17*D* is recommended. The feed section length is 30–40% of the screw length, the transition 20–40% of the length, and the metering section is the rest. Low shear distributive mixing sections 2–3*D* long are used at the screw end.

Guidelines for an acceptable temperature profile are given in Table 22.15. Higher viscosity resin (lower MFI) requires higher processing and melt temperatures. The extruder feed throat is water cooled to below 250˚F (121˚C) to ensure uniform resin feed.

Polycarbonate melt viscosity is more temperature-sensitive than shear-sensitive. Increasing the melt temperature from 455 to 509˚F (235–265˚C) decreases the viscosity by 2.6 times. Increasing the shear rate from 100 to 400 sec^{-1} over the same temperature range reduces the

Table 22.15. Processing Temperature Guidelines for Polycarbonate

Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp ˚F (˚C)
340–380 (171–193)	340–380 (171–193)	410–450 (210–232)	440–470 (227–243)	440–500 (227–260)	440–500 (227–260)

viscosity 1.5 times. Considering that this viscosity change is over a small temperature range, temperature is very significant in affecting the melt rheology.

In sheet extrusion, the die gap is normally set 5–20% greater than the desired final dimension. A coat hanger die with a choker bar for thickness control is recommended if the sheet thickness is greater than 0.20 inch (0.5 mm). The restrictor bar controls the polymer flow distribution to the die lips. The die lips need to be tapered to extend into the roll stack. The minimum desirable nip roll pressure is 450 pounds per linear inch (pli).

In profile extrusion, polycarbonate can be hard to handle because of its low melt strength and tendency to stick to metal. To overcome these obstacles, process at the lowest melt temperature that is practical. Some general guidelines for the die geometry are

- Parallel lip length needs to be 20 times the part wall thickness.
- Draw down for tubular profiles needs to be no more than 15–20%.
- Draw down on wall thickness needs to be less than 20%.
- Optimum melt temperature at the die is 480–510˚F (249–266˚C).

22.7 Polystyrene

Polystyrene (PS) comes in two main forms: general purpose, which is transparent and sometimes called crystal polystyrene, and high impact polystyrene (HIPS), which is opaque due to rubber modification. Polystyrene is a low-cost, clear, commodity resin. General purpose PS has excellent flow characteristics and transparency, while HIPS has excellent physical properties with easy processing and better melt strength. PS is used in many applications, including

- Sheet for thermoforming
- Film
- Foam
- Coextrusion to produce food packaging
- Lighting
- Appliances
- Disposable cups
- Thermoformed dairy containers
- Vacuum formed packages

Polystyrene is not hygroscopic and does not need to be dried unless the resin is moved from a cool to hot environment and water condenses on the pellets. If the resin does require drying because it has been stored under adverse conditions, two hours at 160–180˚F (71–82˚C) will remove any surface moisture.

22.7.1 Extrusion Conditions

Most polystyrene applications use 24–30:1 *L/D* extruders with a 3:1 compression ratio screw. Screws with a 5:1 compression ratio have been used in some PS extrusions. Feed zone flight depth is 11–15% of the screw diameter. The feed zone is 15–20% of the screw length, 25–33% of the flights are in the transition zone, with the remaining flights in the metering zone. A two-stage screw can be used, with 60% of the flights in the first stage. Guidelines for extruder temperature profiles and melt temperatures are given in Table 22.16 for PS and HIPS. Typically HIPS processes at a slightly higher temperature because the rubber increases the melt viscosity.

High impact polystyrene viscosity is both shear- and temperature-dependent, with shear playing a more dominant role in determining melt viscosity. Increasing the shear rate from 100 to 400 sec^{-1} decreases the melt viscosity by 2.5 times. Increasing the melt temperature from 355 to 425˚F (180–220˚C) reduces the melt viscosity 1.9 times.

Sheet extrusion is done with a coat hanger die equipped with a flexible lip that is tapered to fit tightly into the roll stack. The sheet die lips are set 10–12% larger than the sheet thickness. Maintaining the die lip opening close to the final sheet dimensions minimizes the draw. This improves the impact strength, the consistency, and thermoforming qualities. Melt temperatures are minimized to improve the melt strength, limiting sag between the die and the rolls. Sheet gloss is controlled by the roll temperatures. Roll temperatures can range from 100 to 225˚F (38–107˚C). Hotter temperatures provide better surface smoothness and gloss. Set the roll temperature 10˚F (6˚C) below the temperature where the resin starts to stick to the rolls. As PS T_g is 101˚C, the sheet needs to be cooled below 101˚C before it is wound on a roll.

PS foam extrusion can use either a tandem extrusion system or a twin screw extruder with a chemical blowing agent, nitrogen, or pentane injection into the melt. The blowing agent foams the product as it exits the die. Roll stacks cool the foam structures without crushing the cell structure.

Table 22.16. Processing Temperature Guidelines for Polystyrene

Type	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp ˚F (˚C)
General Purpose	340–360 (171–182)	390–410 (199–210)	430–450 (221–232)	430–450 (221–232)	440–460 (227–238)	440–450 (227–232)
HIPS	350–380 (177–193)	400–425 (204–219)	430–450 (221–232)	430–450 (221–232)	450–460 (232–238)	460–480 (238–249)

22.8 Polyvinyl Chloride

PVC is a low-cost, amorphous resin system with unique physical and weathering properties that make it suitable for many applications. Depending on plasticizer concentration, PVC becomes two radically different materials. Unplasticized PVC (UPVC) or rigid PVC is used extensively in the construction industry as

- Vinyl siding
- Window profiles
- Pipe
- Blinds
- Flooring
- Wall covering

Plasticized PVC (PPVC) or flexible PVC is used in

- Tubing
- Medical applications
- Packaging
- Food wrap

A major PVC weakness is its thermal stability. As long as there is no dead space within the extruder, die, or adapters where resin can stagnate, and the processing temperatures are kept low, PVC does not degrade. However, over long runs degradation can occur, resulting in a dark streak initially that leads to total degradation and a dark, useless material. If the extrusion process is stopped, PVC in the barrel and die needs to be purged from the system.

PVC is not hygroscopic and does not require drying prior to extrusion.

Compounded PVC has thermal stabilizers, plasticizers, internal lubricants, impact modifiers, and possibly fillers, and is ready for extrusion. PVC powder, which is available at a significantly lower price, must be compounded with similar ingredients in-house as part of the extrusion process. Large PVC users purchase unstabilized powder and preblend it with the necessary plasticizers, stabilizers, lubricants, and other additives. PVC blending and stabilization is a science unto itself. The stabilizers are mixed with the PVC resin powder using high intensity mixers that generate sufficient heat to adhere the stabilizer to the resin powder. Batches are made continuously and dropped to a holding tank prior to extrusion. Under no situation should unstabilized PVC powder be run in an extruder. Unstabilized PVC degrades very rapidly, giving off hydrogen chloride (HCl) gas, which corrodes the extruder barrel and other metals in the production facility. Inhaling hydrogen chloride gas can cause severe irritation and injury to the upper respiratory tract and lungs, and exposure to high concentrations may cause death.[3]

22.8.1 Extrusion Conditions

Rigid vinyl extrusion into profiles and sheet is done in a 24:1 *L/D* extruder with a 2.4:1 compression ratio screw. Four rows of mixing pins are sometimes added to aid distributive mixing. Chrome plating is recommended on the screw to prevent chemical attack in the event polymer degradation occurs. A breaker plate with a 20/40/20 screen pack is used to increase the backpressure and to eliminate the spiraling action of the melt coming off the screw.

Rigid vinyl can be extruded on twin screw extruders with low-shear screws. Typically, single screw extruders are used to process PVC pellets, and counterrotating twin screw extruders are used when compounding and processing powder that is premixed with the necessary stabilizers, lubricants, impact modifiers, and so forth, in the converter's own facility. Most PVC is compounded or prepared for profile extrusion using counterrotating, parallel or conical twin screw extruders with low screw speeds and high screw fill.

Profile dies require smooth transitions with a 30 degree or less approach angle. Flat plate dies can be used for short runs but must be avoided for long runs due to the probability of resin degradation. Time at temperature is minimized to prevent rigid PVC from degrading. Guidelines for dies are

- Balance die flow to produce uniform exit velocity.
- Normal die pressures are between 3000 and 5000 psi.
- Balance die flow by reducing the die land length, rather than opening up the channel thickness.
- Land length is typically 10 to 15 times the die opening.
- Radius the exit and all corners in the die.

Extrusion and melt temperature guidelines for processing rigid PVC are given in Table 22.17. Unplasticized PVC viscosity is significantly more shear-sensitive than

Table 22.17. Temperature Guidelines for Extrusion of PVC

Type	Zone 1, °F (°C)	Zone 2, °F (°C)	Zone 3, °F (°C)	Zone 4, °F (°C)	Die, °F (°C)	Melt Temp °F (°C)
Rigid PVC	290–300 (143–149)	310–330 (154–166)	330–350 (166–177)	330–350 (166–177)	350–365 (177–185)	350–365 (177–185)
Flexible PVC	250–280 (121–138)	270–300 (132–149)	300–320 (132–160)	300–320 (132–160)	320–340 (160–171)	320–340 (160–171)

temperature-sensitive. Increasing the shear rate from 100 to 400 sec^{-1} decreases the resin viscosity by 2.8 times. Increasing the melt temperature from 350 to 400˚F (177–204˚C) over the same shear rate range reduces the resin viscosity 1.7 times. As rigid PVC is very temperature-sensitive, care must be taken when increasing the shear rate to avoid thermal degradation from higher melt temperatures.

The same extruder used to produce rigid profiles can be used to make rigid PVC sheet. Screw cooling is critical when extruding sheet. Oil is pumped to the screw tip, flowing back toward the feed. This keeps the tip cool, preventing resin degradation at the screw tip. Recommended oil temperature is 200–250˚F (93–121˚C). A coat hanger type die is used to distribute the resin in the die. Roll temperatures are normally run between 120 and 200˚F (49–93˚C).

Flexible vinyl extrusion into profiles and sheet is done in a 24:1 *L/D* extruder with a 2.5–3.0:1 compression ratio screw. A typical square-pitch screw has 5 feed zone flights 0.35 inch (8.9 mm) deep, 12 transition flights, and 7 metering flights 0.117–0.140 inch (2.97–3.6 mm) deep. Mixing pins and other distributive mixers can be incorporated into the screws or, if dispersion is a concern, a dispersive mixer is added. Breaker plates stop the melt spiraling flow coming off the screw end. Dies are similar to those used with rigid PVC.

Extruder and melt temperature guidelines for processing flexible PVC are given in Table 22.17. plasticized or flexible PVC has lower melt viscosity than rigid PVC at the same temperatures and shear rates. This results from the plasticizer in flexible PVC. Plasticized PVC viscosity shows both strong temperature- and shear-dependence, with the shear being slightly more dominant. Increasing the melt temperature from 350 to 400˚F (177–204˚C) reduces the melt viscosity 2.0 times. Increasing the shear rate from 100 to 400 sec^{-1} lowers the melt viscosity 2.6 times. Shear rate is more significant than temperature in affecting melt viscosity.

22.9 Polyester

There are many polyester polymers, and they are available as copolymers and homopolymers. Two polyesters presented here are polyethylene terephthalate (PET) and polybutylene terephthalate (PBT). PET is used extensively in the packaging industry (soft drink bottles) and is coextruded with other resin systems to produce products with outstanding oxygen and moisture barrier properties. PET and PBT are often reinforced with fibers and high aspect ratio fillers to provide superior engineering properties and heat resistance. Other than chemical structure, the main differences between PET and PBT are

- PBT crystallizes much faster.
- PET has higher temperature resistance.

Some PBT and PET applications include

- Underhood automotive products
- Engine heat shields
- Tool housings
- Industrial applications
- Electrical components

Both PBT and PET are hygroscopic, and drying prior to processing is critical to maintain property performance in the final product. Trace moisture in the melt state readily degrades the intrinsic viscosity (IV), which is a measure of molecular weight, resulting in products with lower properties than the base resin. The presence of moisture in the extruder causes a hydrolysis reaction between the polyester and the water, lowering the molecular weight, strength, and part toughness, while increasing the melt flow and difficulties in handling the melt.[4] Drying is accomplished with a dehumidifying desiccant dryer with a –4˚F (–20˚C) dew point. After drying, the pellets are transferred by heated dry air to the feed hopper on a single screw extruder or to the feeder on a twin screw extruder. Dry resin absorbs moisture very rapidly when exposed to air moisture. PET is processed below 0.02% moisture. When dried PET at 0.01% moisture is exposed to 50% relative humidity room temperature air for 15 minutes, it will have 0.02% moisture, which is the processing limit. Feed hoppers are either blanketed with dry nitrogen or heated with dry air, circulated from a desiccant dryer, to maintain the moisture content prior to processing. Vented barrels are not recommended for removing moisture from either PBT or PET, because the molten resin hydrolyses before the moisture can be removed. PET moisture absorption rate after it has been dried to 0.01% is shown in Table 22.18. PET is probably the most moisture-sensitive commodity resin, as it readily absorbs moisture and small amounts affect physical

Table 22.18. Rate of Moisture Pickup of PET Resin Dried to 0.01% Moisture and Exposed to Various Relative Humidities

Relative Humidity, %	15 Minute Exposure	1 Hour Exposure	24 Hour Exposure
15	0.015	0.017	0.032
50	0.02	0.03	0.082
100	0.035	.055	0.3

properties. PBT is dried to below 0.04% moisture before processing.

Factors affecting the drying rates in dehumidifying desiccant dryers are

- Air temperature
- Air flow
- Dew point

All three drying variables need to be optimized and monitored to properly dry PET and PBT. Table 22.19 provides drying times for PET and PBT. In all cases, moisture content needs to be measured before extrusion, to ensure that moisture is below the acceptable processing limit. Sample collection and analytical method are critical in getting accurate data. At no time during measurement or sample collection is PET or PBT exposed to the atmosphere, where it can absorb moisture. Sample bottles for transferring dried resin to testing are stored in a 225˚F (107˚C) oven and capped immediately upon removal from the oven. The bottle is filled with the hot resin and recapped immediately and not reopened until the sample is ready for analysis.

Table 22.19. Drying Time in Hours for PET and PBT

Drying Conditions	225˚F (107˚C)	250˚F (121˚C)	275˚F (135˚C)
Virgin PET	8	4	3
Recycled PET	8	5	4
Wet PET	8	6	4
Saturated PET	16	9	6
PBT		4	

22.9.1 Extrusion Conditions

PET or PBT extrusion is done in either single or twin screw extruders. While barrels can be vented to remove volatiles, barrels are not vented to remove moisture; that has to be done prior to extrusion. Polyesters can be processed on a 24–30:1 *L/D* extruder with a 2.5:1 compression ratio screw. Since both PET and PBT are temperature-sensitive resins (viscosity changes more with temperature than with shear rate), high-shear dispersive mixing devices do not work well. Distributive mixing with Saxton or other mixing heads that split and recombine the melt stream work better. PBT has a relatively small processing window, with the recommended melt temperature between 470 and 510˚F (243–266˚C). PET recommended melt temperature is 520–570˚F (270–300˚C). Extrusion and melt temperature guidelines are given in Table 22.20.

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Table 22.20. Temperature Guidelines for Processing Polyesters

Resin	Zone 1, ˚F (˚C)	Zone 2, ˚F (˚C)	Zone 3, ˚F (˚C)	Zone 4, ˚F (˚C)	Die, ˚F (˚C)	Melt Temp ˚F (˚C)
PET	480–520 (249–271)	520–550 (271–288)	550–570 (288–300)	550–570 (288–300)	550–570 (288–300)	540–570 (282–300)
PBT	450–470 (232–243)	480–500 (249–232)	490–510 (254–266)	490–510 (254–266)	490–510 (254–266)	500–520 (232–271)

Review Questions

1. What resins are moisture-sensitive and require drying prior to extrusion processing?
2. What happens if PET, PBT, PC, and nylon are not properly dried prior to extrusion?
3. What happens if nylon 6,6 is dried too long?
4. What determines the property profile of ABS?
5. What is unique about the different nylons?
6. What happens to nylons 6,6 and 6 at equilibrium moisture content?
7. What is the difference between LDPE, HDPE, and LLDPE?
8. How does polyethylene density affect transparency, tensile properties, melting point, warpage, and environmental stress crack resistance?
9. With PP and PMMA, what effect does MFI have on processing conditions?
10. What is the difference between PS and HIPS and how does it affect processing?
11. What is the difference between rigid and flexible PVC? What is used to change PVC from flexible to rigid or vice versa?
12. What important processing factors are considered when extruding PVC?
13. What is unique about extruding polyesters?

Part 4: Troubleshooting the Extrusion Process

23 Problem Solving

How is an extrusion problem solved? Regardless of whether it is in a production, pilot plant, or research and development environment, there are numerous approaches to take, ranging from the trial and error method to implementing systematic investigative procedures. Trial and error is just turning knobs until something happens to correct a specific problem or make it disappear. For some it is a "shot in the dark," where something is tried without a valid reason, as "all" other ideas have been tried and failed and "something" needs to be done. Another approach is "gut feel": "I have a lot of experience and I think this is the correct approach to implement to make the process run more efficiently at higher yield." The "gut feel" approach is usually based on experience and is something that has been done or seen previously. Although the operator or engineer may not know why it works, it does seem to correct the problem. Another approach is to ask a fellow employee if he or she has seen a similar problem and what approach was used to eliminate the problem. In production, if the line is not producing quality product for shipment to a customer, the company is losing money. While it may be ego-building to solve a problem on your own, if the line does not run for an entire shift, the business loses money, which is not good for the company or its employees. Always seek information from other people, probing for methods they have implemented to solve a particular problem, in an effort to get the line operational as quickly as possible. Businesses are run as teams, and cooperation between fellow employees makes the process run more efficiently with higher yields.

The next question to ask is how many variables to change at one time while trying to solve a problem. In a production environment, with time restraints to get the line up and running, the answer is probably one. However, when a problem arises, you often see people turning knobs and changing multiple variables at one time, trying to get the process to run more efficiently or to solve a problem. Once a variable has been changed, how long should the process run under those conditions, allowing it to stabilize or equilibrate before collecting samples to determine the effect of the process changes? The answer is, it depends on the extruder factor or variable changed. For example, if the screw speed is increased or decreased, the thought process is that this happens quickly (30 seconds or less) and the process is in equilibrium immediately. However, if a change is made to the barrel or die temperature profile, equilibrium may take 15 or 20 minutes, depending on extruder size, the delta *T*, and how long it takes for the actual polymer melt temperature to reach equilibrium. In reality, the time to reach equilibrium and collect a sample after a process change is made depends on the extruder size, whether it was a screw speed change, barrel temperature change, die temperature change, or some other controllable variable (screen pack, raw material, cooling, and so forth). The extruder must be allowed to reach equilibrium before the effect can be determined. As an example, increasing the screw speed from 100 to 120 rpm is almost instantaneous; however, the total effect may not be completely visible for a half hour or more, depending on the extruder size. Increasing the screw speed generates additional frictional shear heat that must be dissipated by the barrel cooling. Initially the large metal mass comprising the barrel and the screw absorbs the heat. Over time, the barrel and screw temperatures increase as equilibrium is reached. If the barrel cooling is insufficient to remove this additional heat, polymer degradation may occur 45 minutes to an hour after the screw speed change. Barrel temperature changes require as long a time to come to equilibrium because the metal mass involved is still the same. The heaters transmit the heat through the barrel and into the screw through the polymer melt. It takes time to establish equilibrium temperature conditions between the barrel, polymer, and screw. On a small extruder (3.0-inch diameter or smaller), the time to reach equilibrium after a process change is 30 to 45 minutes. For larger extruders with more barrel and screw mass, the equilibration time increases. Process changes on an 8-inch extruder may take two hours to reach steady state or equilibrium. If a process change is made and the system is not allowed to reach equilibrium before a second or third change is made, the effect of the first process change is never realized. If the extruder is not allowed to reach equilibrium before another change is made, the operator is running around in circles and may never solve the problem. Consequently, the answer to the question, "How long do you wait after making a process change to determine the effect of the change?" depends on the extruder size and the change made. In any event, the time is always longer than one to five minutes and in general between 30 minutes and two hours, depending on the extruder size.

The next question is, "How large a process change should you make when changing extrusion variables?" Changing zone 2 on either a single or twin screw extruder 4˚F (2.2˚C) is not a significant enough change to observe any effect. If you change zone 2 4˚F (2.2˚C) and wait 20 minutes for the system to come to equilibrium, then change zone 3 5˚F (2.8˚C) and allow it to come to equilibrium, it will take too long to make significant changes that affect the processing. Using the same argument for screw speed, increasing screw speed 2 rpm is a change, but it is probably not significant enough to see

any effect unless the process setup is on the very edge of the processing envelope and that small process change pushes it over the edge. An example is extruding polybutylene terephthalate (PBT) into sheet using a 3.5:1 compression ratio screw with a Maddock mixer designed for polypropylene. Running polypropylene sheet with the same setup, the extruder is run at 125 rpm, producing quality sheet at a high rate. However, running PBT at 54 rpm degrades the resin, generating gas that blows holes in the PBT sheet. At 52 rpm, defect-free sheet is produced. This 2 rpm change converted a good process and product into the unacceptable range. In most cases, 2 rpm does not greatly alter the process or product characteristics. In this example, a 2 rpm change caused a dramatic effect because the extruder setup was not designed to process PBT. With increased screw speed, the process was pushed over the edge of the processing envelope, resulting in unacceptable product.

When making process changes, the new setting needs to be significantly different from the original setting to determine if the change had an effect. Table 23.1 provides some suggestions for process changes. The actual change depends on the extruder, screw design, and the polymer being processed. Assume the current process is running rigid PVC with a barrel temperature profile of 300/340/370/370˚F and die temperature of 380˚F (149/171/188/188˚C and die temperature of 193˚C), and the suggestion is made to increase the temperature profile 30˚F (16.7˚C). The immediate concern is whether rigid PVC can be processed at that temperature without degradation. It is important to use both material and process knowledge to determine the increase and/or decrease when changing processing conditions. Lowering the barrel setpoint temperatures severely can cause the extruder to operate at excessively high torque or, in extreme cases, stop. With crystalline polymers, significantly lower barrel temperatures can cause the polymer to solidify or, with amorphous polymers, become so stiff the extruder torque limit is exceeded. The process change implemented in a specific situation has to be large enough to determine if the change had an effect on the process while remaining within the operating limits of the polymer being processed. If the operating change did not have an effect on the process, go back to the original process setting and change a different variable. In all situations, document the effect the variable change had on the process so the information can be used in the future in making process modifications. If the change had no effect on the process, it is still noteworthy. This knowledge can prevent similar changes if the problem arises again, thus saving time and valuable resources.

Table 23.1. Size of Process Changes

Size Change	Screw Speed, rpm	Temperature °F (°C)
Small	1–2	2–4 (1.1–2.2)
Medium	10–15	10–15 (5.6–8.3)
Large	25 or more	30 (17)

After a process variable change (screw speed, barrel temperature, cooling tank temperature, roll temperature, and so forth) is completed and the extrusion system is allowed to come to equilibrium, what is the next step if the product or process deficiency has not improved? Is the process variable changed back to its original setting, or are other process variables changed without converting the process back to its original setpoint? Since the first process change had no effect on the process, it doesn't really matter if the process is changed back to its original settings or the current process configuration is used as the baseline for additional process changes. In either situation a baseline is established to determine if future process modifications have an effect. Assuming the process is run using the new baseline data resulting from the process change, a second change is made that shows a very dramatic effect on the product or process. The question becomes, "Is the product or process improvement or deficiency caused by the second process change or the interaction of the first and second process changes working together?" Regardless, experimental data have to be generated to understand the effect of any extruder changes on the product and/or process.

When troubleshooting product problems, always save samples made under different processing conditions. Compare the different samples versus the process changes to determine the best direction to take for future process changes. If the product property being optimized continues to show improvement and almost meets the quality requirements but not quite, what is the next step? To a process engineer or operator, small changes in process variables may seem like the correct approach to optimize the production process. However, if after making numerous small changes the product still does not meet all the specifications, it is time to step back and evaluate the process changes versus product improvements and ask the question, "Can I get there from here?" At some point it becomes obvious that small changes in the process variables will not lead to the desired endpoint. It may be necessary to rethink the process and/or product and make a dramatic change in another direction.

If a product or process problem is identified immediately after a production run or trial is started, go back and compare the complete process setup versus the standard operating procedures (SOP). Verify that all operating conditions are properly set, including the auxiliary equipment setup, resin formulation, drying, type of die, barrel and die temperature profile, screw speed, takeoff equipment setup with the proper temperatures, and so forth.

Assume the product or process is running properly with a quality product being produced at a high rate. All of a sudden the product or process is out of control. What happened or what changed? If the processing conditions are well documented when everything is running well, what changed may be easier to identify. In troubleshooting a process, it is critical to identify what has changed in the product formulation or processing conditions to cause the product or process to be out of control. At start-up the product and process might be running very smoothly, and 45 minutes later, with no process or formulation changes, the product or process is running poorly. The operating conditions at start-up yield good product; however, after the extrusion process has come to equilibrium, the operating conditions do not produce a good product or process. The question is, What has changed between the start-up conditions and the equilibrium conditions (probably temperature difference), and what changes are required to get the extruder and process to perform similarly to the start-up conditions?

One last potential scenario is the product or process produced a quality product at a high rate last time it was run, and it is borderline acceptable or unacceptable this time. What has changed? Review the operating conditions and procedures used last time to verify that they are the same this time. Have any equipment changes been made between the two runs? Review the run sheets from the previous production campaign to verify that the operating conditions are the same as the SOP. If the operating conditions from the previous run are different from the previous SOP conditions, the SOP and current processing conditions need to be changed to the operating conditions used in the previous production run.

Problem solving and troubleshooting require good documentation comparing what is currently happening in the process to what occurred previously, and to the SOP. A systematic approach and both materials and process knowledge are necessary to solve processing and product problems.

Problem solving starts with an understanding of how equipment in the extrusion line functions and how the material interfaces and behaves in the equipment as it is being transformed through the various stages into an acceptable product. (Parts 1 and 2 covered single and twin screw extrusion equipment, while Part 3 covered the polymeric materials and their behavior in the extruder.) To understand how each piece of equipment operates, the following questions require answers:

- What is the function of a particular piece of equipment in the overall extrusion process?
- How does each piece of equipment operate?
- What function or role does each piece of equipment perform relative to other pieces of equipment in the overall extrusion line?
- What do the various gauges and output displays communicate to the operator?
- How do the controls function and what do they control?

It is necessary to understand how the extruder and the extrusion process operate when they are working together properly. What roles do extruder heating and cooling, downstream equipment heating and/or cooling, puller roll speed to extruder throughput ratio, die operation and adjustments, extruder feed throat temperature, extruder screw design, and so forth, play? Learn to identify any unusual noises or odors, as these may be the first indication some change has occurred in the process or equipment. Die defects may prevent an acceptable product from being produced; be able to identify die damage both internally (assuming die is apart) and externally.

In combination with processing conditions and equipment knowledge, an understanding of the materials and their transformation in each step is necessary.

- What is the formulation rheology and how does it change with temperature and shear?
- What are the proper drying conditions?
- How does moisture affect properties?
- What is the proper melt temperature?
- What is the material shrinkage?
- Is the polymer crystalline or amorphous?
- What are the cooling requirements (material T_g or T_m)?

Complete and well-documented setup sheets or SOPs are required for each product and process. The sheets need to be routinely reviewed and updated as process improvements and modifications are made. Specific setpoints and operating ranges for each process input (independent variable) for all equipment in the extrusion process, from drying to puller speed, are defined. SOPs are dated with revision numbers and contain the process changes made over time as different revisions are generated.

Each extrusion line needs its own log book documenting any process and equipment changes made by the operator or maintenance. Operators log in at the beginning of each shift and document any changes made to the equipment or process during their shift. Information contained in the process log includes

- Operator name
- Product changes
- Start-up and shutdown times
- Documentation of any downtime and reason
- Equipment malfunctions or changes

- Product or process deficiencies and corrective action taken to remedy product and/or process issues
- Any unusual observation or occurrence

It is impossible to record too much information, but very possible to record too little. If there is a customer complaint, product traceability and any unusual occurrence that happened during the production run greatly assist in determining if anything out of the ordinary or unusual can account for the complaint.

Processing conditions are recorded periodically throughout a particular production run. Statistical process control (SPC) data can be collected continuously throughout a run and stored on disk or CD for later evaluation or if there is a customer complaint. Newer extruder control systems may have built-in SPC documentation that can be used to record process data automatically at specific time frequencies. Process control provides data showing conditions remained constant or within specific upper and lower control limits during the run or when a process upset occurred; control charts identify the time frame during which the process was operating in control and out of control. Extruder load (torque, amps, or percent load), screw speed, and barrel temperatures define the extruder process. Include feed rates and downstream equipment speeds and temperatures in the data collection, because a process snapshot at any particular time is useful to reconstruct. Setpoints versus actual operating conditions provide information on what is changing and how the changes may have affected the product or process. Setpoint information verifies that the temperature controllers are set correctly and, more important, operating properly. These records are critical for product traceability.

Any equipment defects or malfunctions identified during a given production run need to be repaired either immediately or after the production is completed and before the next material is started. Maintenance and/or equipment issues should only be put off until the current production run is completed if safety or the product is not affected by the equipment malfunction. An example may be a heater band that burns out during a run. Since most of the energy is from shear heat, the temperature zone may still control within the specified process window at steady state with one heater band out of two or three not functioning properly. On shutdown, the heater band is replaced before the next production run is started. Always make sure any equipment malfunction, the specific time the malfunction is first observed, and when it is corrected are documented in a log book.

Processing conditions and setpoints to be recorded include

- Actual and setpoint barrel and die temperatures
- Melt temperature
- Feed throat cooling
- Screw speed
- Motor load (torque, amps, or percent load)
- Melt pressures in both the die and barrel
- Throughput rates
- Polymer pretreatment conditions, such as
  - — drying temperature
  - — moisture content
  - — lot numbers
  - — blend ratios
  - — feeder speeds prior to the extruder
  - — moisture level
- Takeoff equipment speed, such as rolls and pullers
- Takeoff equipment temperature, such as water bath, rolls, and air
- Air pressure used in cooling, such as blown film

New equipment (extruder, screw, takeoff equipment) should be run at two or three different speeds with a particular product to obtain baseline operating data. Later in the equipment life cycle, the same processing and product conditions can show if there is a throughput rate or product quality shift. Baseline data comparisons six months, one year, and two years later assist in determining the equipment life expectancy and defining when modifications might be anticipated. Any time a new barrel or screw is installed, a baseline plot of throughput rate (pounds/hour) versus screw speed (rpm) should be generated with a specific polymer for comparison later in the screw and barrel life. These data assist in predicting capital expenses and screw replacement.

Good troubleshooting and problem solving techniques are based on determining what has changed in the process. Without proper documentation, during both good and bad operation, it may be impossible to define what changes have occurred, making problem solving difficult or impossible. At the same time, collecting reams of process and product data that are never analyzed is an exercise in futility, and questions the importance of the data.

Review Questions

1. Why is process documentation necessary for extrusion troubleshooting?
2. What is the best method of solving extrusion problems in production?
3. How long do you have to wait after making a process change on an extruder to see the effect? Assume the change is either a barrel temperature or screw speed change.
4. Define equilibrium as it refers to an extrusion process.
5. How large should the extruder process change be when evaluating the effect of either barrel temperature or screw speed on process or product improvements?
6. Why is it important to understand both the process and the material in troubleshooting extrusion processes?
7. What is the purpose of a log book in the extrusion process?
8. When troubleshooting a product problem, why is it important to keep the samples from different processing conditions?
9. If a new production run is started and the product is not running well, what is the first thing to do?
10. What should be recorded in a log book?
11. What processing conditions and observations should be monitored and recorded?

24 Five Step Problem Solving Process

The "Five Step Process" is part of a total Quality Education System developed by Philip Crosby and Associates, Inc.[1] The "Five Step Process" is a procedure or process to follow to define, correct, and eliminate problems. It is used as part of a total quality program in many industries to improve product yield and quality performance. The five steps in the process are

- Define the problem.
- Fix the problem.
- Identify the root cause.
- Take corrective action to eliminate the problem.
- Follow up and monitor the corrective action to verify that it eliminated the problem.

24.1 Problem Definition

Defining the problem can be one of the most difficult steps in this process. When several people discuss the potential causes of a problem or try to define the problem, many different thoughts and ideas will be generated due to the varied backgrounds and experiences of the people present. Different techniques are used to define the problem and develop procedures to identify the root cause. As an example, assume a customer has identified that his or her extruded pipe is failing prematurely. There were no obvious defects when the pipes were made, and no deficiencies were detected in the normal quality control (QC) checks made prior to shipment. However, failures in the field have created a need to understand the source, correct the problem, and devise methods or processes to eliminate the defect in the future. The following facts are known about the defective parts and the manufacturing process used to produce them. The pipe is made from ABS on a 3.5-inch, 30:1 single screw extruder equipped with a single-stage 2.5:1 compression ratio screw. The extruder has a downstream vacuum port that is not used during the production of this particular part. A 2.7:1 two-stage screw is available but has not been used with this particular product. The ABS is predried with a 1000-pound desiccant dehumidifying dryer set at 200˚F (93.3˚C) prior to extrusion. Moisture content of the resin was checked prior to starting the run and found to be within specifications at <0.02%. The throughput rate is 300 pounds/hour, and the normal production run is six to eight hours. Downstream sizing equipment is a two-stage vacuum tank with vacuum in the first stage and 100˚F (37.8˚C) water temperature. There is no vacuum in the second stage and water temperature is 60˚F (15.6˚C). After the vacuum tank, an air knife removes excess surface moisture from the pipe; a caterpillar-type puller, printing station, cut-off saw, and a packaging station are used to prepare the product for shipping.

Six people associated with the production are assembled to define the problem, determine a corrective action to fix the problem, and identify the root cause. The six people come from customer liaison, manufacturing, quality control, R&D, and a process assistance group that interfaces between R&D and manufacturing; this group brings different backgrounds and perspectives to address the problem. The person from customer liaison is acting as the meeting facilitator to define the problem, identify a fix, determine the magnitude of the problem, and estimate how long it will take to identify the root cause and to implement a corrective action procedure.

The first step in the process is to define the problem. From the perspective of the customer liaison representative, the problem is defective parts in the field. Manufacturing and quality control both checked the parts prior to shipment and are convinced the product met all customer specifications. The person from the process assistance group took part in the production run under question and verifies that all machine specifications and manufacturing processes were properly set and followed, while the R&D representative is listening to discern the facts. The magnitude of the problem is not known. How many bad parts were shipped and where are the parts now? Of course the customer is trying to identify the magnitude of the problem and understand what the company is going to do to resolve the problem. A decision is made to define a program that can be conveyed to the customer, identifying a course of action to identify the cause and implement corrective action.

Everyone agrees the problem is premature part failure in the field. There is not agreement on whether it is caused by the company supplying the pipe, incorrect installation in the field, or excessive pressure in the application exceeding the pipe burst strength. Customer liaison is requested to obtain samples of the failed pipe for failure analysis to determine if the pipe they produced was defective. In the meantime, potential causes from the company perspective are outlined. What could cause the pipe failure?

- Wrong feedstock
- Wrong extruder temperature profile and melt temperature
- Improper drying
- Volatiles in the feedstock
- Contamination
- Wrong puller speed
- Improper cooling
- Wrong screen pack

The decision is made to obtain parts with premature failure and compare them to good parts, to evaluate the pipe failure mechanism, and in the meantime to check the processing conditions and other things on the list above to verify that everything was produced under proper operating conditions.

Two days later, with a failed part in hand and a manufacturing parameter review, the same group meets again. All manufacturing conditions (extruder temperature and die profile, melt temperature, raw materials, drying, puller speed, and so forth) were verified to be correct. Inspection shows microvoids in the internal cross section of the failed part, with no microvoids present in good parts. The problem statement is now clear. The problem is defined as pipe failure due to internal microvoids. The next step in problem definition is to list the potential causes for microvoid creation in the extrusion process. After listing the potential causes of void formation, a fishbone diagram, shown in Fig. 24.1, is generated. The potential causes are listed:

- Resin degradation
- Resin contamination
- High regrind level
- Moisture in feedstock
- Air entrapment in final product
- Volatiles in feedstock
- Cooling too rapidly

Other parts from the same production run are requisitioned to determine if all parts exhibit the same problem.

Evaluating samples throughout the production run shows only materials from the end of the run had microvoids. Production for the first three to four hours shows no microvoids present in the pipe; it is seen only in material run over the last two to three hours. After numerous meetings and investigative work, it is determined that the premature pipe failure is attributable to microvoids in the part that occur after three to four hours of production. While the root cause is not yet known, the problem has been defined and corrective action can be initiated.

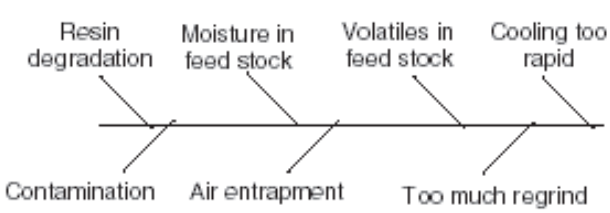

Figure 24.1. Fishbone diagram—cause of voids.

24.2 Fix the Problem

The second step in the five step process is to fix the problem. All action is designed to supply the customer with quality parts that keep the process running or the supply flowing to customers while developing a long-term solution. The temporary solution may include reworking previous materials, running at lower throughput rates, patching the process, using more expensive raw materials, or making the product on different equipment. The fix is to supply the customer with a quality product in sufficient quantity to keep the process running or the supply to customers flowing regardless of cost. While this step is not meant to be a permanent solution, it is an intermediate action until the original problem is solved permanently and corrective actions are properly implemented to eliminate the problem permanently. The fix is not the final step in the five step process, since the root cause has not been identified and the problem has not been permanently eliminated.

In the pipe example, the short-term solution is to make shorter production runs and monitor pipe for microvoids, since the product made for the first three to four hours meets all customer specifications. This is not a long-term solution because it is an inefficient operation. Start-up and shutdown is time consuming and productivity is zero. In any extrusion process, money to pay employees and produce a profit is only made when the extruder is running, producing quality products that are shipped to the customer. Making shorter production runs does ensure a quality product is sent to the customer, but it lowers efficiency.

24.3 Identify the Root Cause

This is the most crucial step in the five step process. If the root cause of the problem is never identified, it can never be corrected and will keep reappearing from time to time. There may be more than one root cause, and they must all be identified to eliminate the problem permanently. Ways to identify the root cause are "brainstorming," "fishbone diagrams," and "Pareto analysis." Brainstorming is a technique where people familiar with the problem or process get together and suggest potential causes for the problem. A person is selected as a facilitator, who states the problem. The facilitator asks individuals one at a time what they think might be the cause. The potential causes are listed on a blackboard or paper as they are stated. No discussion occurs on any ideas until they are all documented. No ideas are thrown out as unrealistic or ridiculous, and all potential solutions are initially given equal weight. After all the ideas are listed, they are weighted and listed in priority by those that should be

evaluated first. The ideas can be listed on a fishbone diagram similar to that shown in Fig. 24.1.

A fishbone diagram may be used in a brainstorming session. In certain situations, a fishbone diagram is based on the five M's: man, measurement, material, method, or milieu. All potential solutions are attributed to one of the five M's. In the case of the defective pipe, the response to the five M's might fall in the following categories:

- **Man:** Has the operator done something wrong in extrusion, blending, handling the pipe at the end of the line, setting up the line improperly, and so forth?
- **Measurement:** Are the current test methods to measure the product response acceptable to identify problems? In the pipe case, if specific gravity on the pipe or the individual pipe weights were measured, the microvoids may have been identified. Measurement normally refers to whether QC procedures are done properly, if the gauges on the extruder are working properly, if the vacuum tank vacuum gauge is functioning properly, whether the temperature and dew point measurements on the dryer are operating properly, and so forth.
- **Material:** Is the correct material being used with the correct regrind content, the proper stabilizers and additives, and the proper blending procedures?
- **Methods:** Are the correct processing procedures being followed for extruder screw speed, barrel and die temperature profiles, throughput rates, cooling tank temperatures, distance between cooling tank and extruder, take-up speeds, and so forth?
- **Milieu:** Are there any environmental conditions that may affect the process? Is the extruder near an overhead door that is opened during the wintertime to expose the extruder to cold temperatures, lowering the barrel and melt temperatures periodically?

A fishbone diagram can be constructed with the five M's and all the factors or thoughts attributed to a particular M in the list.

A Pareto diagram is used to determine which factors listed in a fishbone diagram or brainstorming session are to be addressed first. The potential causes are listed and a weight attached to each one based on all factors adding up to 100%. The most significant factors or the factors expected to have the greatest effect are tackled first.

In the pipe example, the brainstorming session and fishbone diagram are used to analyze the problem. A Pareto diagram, based on the fishbone diagram, is shown in Fig. 24.2. From Fig. 24.2, the most likely causes for microvoids are cooling too rapidly and feedstock moisture. Cooling thick extrudate too rapidly can cause shrinkage voids in the pipe walls. However, if the process data show the melt temperature was constant over the production run

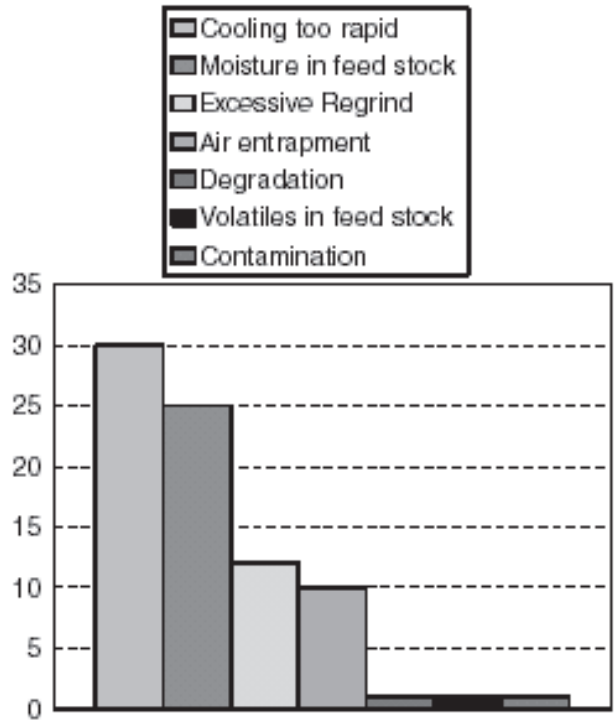

Figure 24.2. Pareto diagram for pipe problem.

and the water temperature in the sizing or vacuum tank did not change after three to four hours, it is unlikely this is the cause. If rapid cooling was the problem, microvoids should have been present in the beginning of production.

The second most likely cause in the pipe problem is feedstock moisture. A moisture analysis was done prior to the production run with moisture content measured at <0.02%. Consequently, at the start of the production run, no microvoids should have been present due to moisture. With a 1000-pound dryer running at 300 pounds/hour, the dryer is emptied of the originally dried resin after 3 hours and 20 minutes. The new material added to the dryer after the production run is started may not have had sufficient drying time to attain the proper moisture level prior to extrusion. Moisture present in a nonvented extruder can be trapped and can cause small voids. If voids start showing up sometime after three hours and increase as the production run continues, the problem may be due to diminished drying time.

The next step is to verify this hypothesis. Another production run is scheduled under the same operating conditions used when the defective pipe was produced. The moisture content is measured at start-up and after each hour until the production is complete. Assuming this is the problem cause, the moisture content will remain constant for the first three hours and gradually increase until the production run is complete. A corrective action can be instituted for both short- and long-term solutions.

An immediate corrective action is to lower the throughput to 250 pounds/hour so all ABS is properly dried for four hours. In the long term, this affects profitability due to the lower throughput rate. Long-term solutions to the problem are

- Buy a larger dryer.
- Use the two-stage screw and remove moisture through the vacuum vent.
- Move to a different line with more drying capacity.
- Buy a second dryer to supplement the first one or replace the dryer with a larger-capacity dryer.

Once the root cause is known, corrective actions can be implemented to permanently and cost-effectively solve the problem.

24.4 Take Corrective Action

The fourth step is to take corrective action to eliminate the problem. This involves changing the process, equipment, procedures, and so forth, to eliminate the problem. It is best to generate more than one possible corrective action scenario so a more cost-effective method can be used. In the pipe example, four corrective action plans are identified, and each one needs to be evaluated on its own merits. Factors affecting the decision on which corrective action is best include cost, time to implement, and reliability. Any process changes are communicated to the customer so he or she knows you have identified the cause and corrective action is in progress to eliminate the problem. Proactive steps solving customer problems are positive for your business.

The corrective action step employed has to ensure that no other critical part property is affected by the changes in process or product. Other properties that might be affected by process changes include color, physical properties, dimensions, molecular orientation, and so forth.

24.5 Process Monitoring

The fifth step in the five step process is to periodically monitor the corrective action taken to verify that the problem is truly corrected and does not recur. People need to be advised of the solution and any change in the process.

An effective "Five Step Process" can be employed to solve problems in plant environments, find short-term solutions to fix problems while the root cause is being identified, and implement changes to correct the root cause. Other methods are available to solve extrusion problems. The "Five Step Process" is only one of many approaches that will lead to a successful conclusion. Whatever methods or procedures work best in your facility should be used. As with all problem solving, a systematic approach is required that deals with facts and data, and not supposition, to provide an effective solution.

Too often somebody will say, "We have already done that and it doesn't work." The best response is, "Can we review the data to verify that it will not work in this particular situation?" The data may show something else was done other than what is now being proposed. It is essential to attack all extrusion problems with an open mind, review the data, not jump to conclusions, and do experimentation to prove or verify any hypothesis concerning the potential solutions.

REFERENCE

1. Crosby, Philip, *Quality Education System*, Winter Park, FL 1988.

Review Questions

1. What is the "Five Step Process" to problem solving?
2. What are the two most critical steps in the five step process and why?
3. What is brainstorming?
4. What is a fishbone diagram?
5. What is meant by the five Ms?
6. What is a Pareto diagram?

25 Design of Experiments

Design of experiments, referred to as DOE, is a systematic approach to understanding how process and product parameters affect response variables such as processability, physical properties, or product performance. It is a tool similar to any other tool, device, or procedure that makes the job easier. Unlike quality, mechanical, or process tools, DOE is a mathematical tool used to define the importance of specific processing and/or product variables, and how to control them to optimize the system performance while maximizing properties. DOE uses statistical methodology to analyze data and predict product property performance under all possible conditions within the limits selected for the experimental design. In addition to understanding how a particular variable affects product performance, interactions between different process and product variables are identified. Design of experiments is a technique or procedure to generate the required information with the minimum amount of experimentation, using

- Experimental limits
- Specific experimental conditions
- Mathematical analysis to predict the response at any point within the experimental limits

DOE is used to determine which factors or variables and interactions are significant in contributing to the effect being measured, and those variables and interactions that are insignificant and don't contribute to either a particular product property or processing condition. Using DOE saves both time and money by providing a useable understanding of the properties and process. The best time to use a DOE is during new product or process development, existing product or process optimization, and while solving technical problems when more than one variable is present. Where is DOE used? For solving any technical problem when you want to fully understand the response to different process or product variables that can be changed or controlled during the experimentation. Problems in industry typically get recycled from time to time, because an idea that was originally a good idea couldn't be accomplished because the technology was not available or the project was aborted because of lack of time and resources. Since the original work, new technology has been developed or resources are available that now make a solution possible. The advantage of using a DOE approach is systematic data are generated, summarized, and evaluated to definitively determine whether a project should be carried forward or if it is fundamentally impossible to resolve and needs to be dropped. Regardless of whether the DOE results are positive (experiment showed desired response) or negative (experiment showed undesired response), it is important to complete the project and document the results so the project will not be recycled at a later time. The DOE results provide an understanding of the processing and/or product parameters and their interactions over the experimental space studied.

Design of experiments is not the procedure to use to solve production problems on the floor during the second shift when the product being produced must be shipped first thing the following morning in order to keep your customer's production running. To solve problems under these circumstances, change one variable at a time, slow the production line down, do whatever is necessary to produce the product to meet the customer's requirements. After production is complete, go back and evaluate the situation to determine if a designed experiment can be used to define the process limits and capabilities so that the next time the same product is run it can be produced more cost-effectively at higher yields with fewer problems.

25.1 DOE Process

The process involved in conducting a successful design of experiments can be broken down into five steps:

- Define the problem.
- Plan the experiment.
- Run the experiment.
- Analyze the data by statistical methods.
- Report the results.

25.1.1 Defining the Problem

As with the discussion in Chapter 24, this may seem obvious, but in practice it is sometimes difficult to do. Input from different people may conflict, and deciding on the specific variables to control may be controversial. Some input may stress the need for a DOE, while other suggestions may imply a DOE is a waste of time. Verify that the problem is properly defined and clearly understood before starting any work. As an example, assume an extrusion profile at a customer is reported to have unacceptable dimensional stability and you have been requested to solve the problem. While a DOE to understand the effect of processing conditions on part dimensions may be advisable, the first step is to define what is meant by "unacceptable dimensional stability." Questions to ask are

- Do dimensional changes occur in the X, Y, or Z direction over time?

- Is one dimension outside the specification limit, and is this temperature related?
- Is the problem part warpage that is inconsistent from piece to piece?
- Do the dimensions change in a controlled manner over time?
- Have the dimensions changed because of a change in raw materials?
- Are the dimensional issues due to shrinkage caused by the raw materials?
- Are the dimensions different from run to run because of process changes?

Before deciding on the proper course of action and whether a DOE is necessary, make sure the problem is properly defined.

It is critical to fully understand the problem before you try to solve it. If the problem is only viewed as a problem in some people's eyes, and after evaluating the issues you realize that solving the problem will not generate more business, improve customer satisfaction with the product, or make the product better, you are probably wasting the company's resources by addressing or spending time on the problem. An example is a compounded product produced in an oval shape rather than the typical cylindrical pellet. Customer Service states it cannot sell an oval pellet because all customers normally receive cylindrical pellets. Since the oval pellets are free flowing and process the same in both injection molding and extrusion equipment, working on the problem to make the pellets cylindrical wastes valuable company resources that can be better spent on other developmental programs. Fortunately, the decision is made not to spend time or effort trying to solve the oval pellet problem, and the product is accepted in the field without comment.

25.1.2 Plan the Experiment

After the problem is properly defined, the second step in the DOE process is to select the independent variables, with their limits for evaluation, and the dependent variables (responses to measure for each experiment). Independent variables or factors are parameters of either processing or product that are set at specific values (levels) and controlled in the experimental design. For extrusion, some potential independent factors are barrel and die temperatures, screw speed, raw material formulations, drying time, drying temperature, puller speed, and downstream conditions such as roll temperatures, roll gaps, draw ratio, water temperature, oven temperature, vacuum level, and so forth. Dependent variables or factors are the responses being measured for each experiment to determine whether the independent variables have an effect on properties or processing conditions being evaluated. Dependent factors might include physical properties (impact, tensile, and flexural properties in the machine and transverse directions, heat resistance, and so forth), surface or appearance properties (surface roughness, gloss, distinctness of image, color, and transparency), formulation cost in a product composition experiment, and dimensions (shrinkage, warpage, thickness, length, and so forth). Many processing variables in extrusion are dependent variables related to the selected independent process variable; these include melt temperature, melt pressure (die and barrel pressure), and extruder torque. Melt temperature may be either an independent or dependent response, depending on how it is used and/or controlled in the experimental design. In some processes the melt temperature can be controlled and is treated as an independent variable; in other processes the barrel temperatures are controlled, and the melt temperature is measured as a dependent response to the selected barrel temperatures and screw speed. Melt pressure is normally a dependent variable or response unless there is a valve to control and set the melt pressure, at which time it becomes an independent variable. In deciding whether a factor or variable is independent or dependent, ask the question, "Is there a knob to turn to set that variable to a predetermined value?" If the answer is that no controller exists to set a particular variable in the process to a fixed value, it is a dependent variable and is measured as a response to factors that are preset and controlled.

Independent variables are normally quantitative in nature, meaning they are set at a specific numerical value. However, in some experimental designs, the independent variables are qualitative, meaning a particular switch is on or off, a high or low setting is used, or a control variable is yes or no. It is better to use quantitative factors if possible. With quantitative variables, the high and low levels for each factor are defined. As an example, the high/low screw speed levels for a twin screw extruder experiment may be set at 250 and 150 rpm. This sets the experimental space for that factor. After the experimentation stage is complete, mathematical models predict the response within the 150 to 250 rpm range. As this is a linear model, it does not predict well outside the experimental limits, such as 100 or 300 rpm, if the response is nonlinear. Therefore, the experimental limits need to be carefully selected in the planning stage and set as wide as experimentally practical. After the experimentation is complete, you don't want to come back and wish the limits were wider. The points selected to test over the range of the experimental limits are called levels. Going back to the screw rpm, if experimental points are run at 150, 175, 200, 225, and 250 rpm, there are five levels. In determining the experimental limits, do a few preliminary experiments to verify that the experiment can physically be run when all the independent factors are at

either their high or low values. If a particular experiment cannot be run, for example, at low screw speed and low barrel temperature because the torque exceeds the high limit on the extruder, this set of conditions is impractical and another set of low independent variables needs to be selected where the process can be run practically.

Dependent variables or responses are easy to measure and evaluate when using quantitative tests such as tensile, flexural, and impact, which provide very specific values for each experiment. In some situations, responses are more qualitative than quantitative. Normal qualitative responses such as pass/fail, good/bad, or yes/no do not provide a good response to model. A procedure has to be developed to quantitatively differentiate between the different samples for that particular response. Instead of measuring the sample as pass or fail, develop a grading system that allows differentiation on a basis of 1 to 10, even if 1 through 3 are considered failures, 4 through 7 are mediocre, and 8 through 10 are good. The best results are obtained under processing conditions that produce a 10. The number of dependent variables or factors that can be evaluated is almost limitless for all experiments as the independent variables are changed. If a particular property response is almost the same at all sets of experimental conditions, no significant effect exists for this property over the range of processing or product conditions tested. If each response value used in the analysis for a particular set of processing conditions is the average of 10 measurements, the standard deviation obtained from the measurements at each set of experimental conditions can be evaluated to determine which set of conditions generates the most reproducible results.

The number of data points evaluated for each experimental response depends on the number of tests required for statistical significance, based on the precision and accuracy of the test. Precision is how close the measurements are to each other. If the measurement precision is high, the number of samples needed to obtain an accurate mean is smaller than if the precision is low. Accuracy is how close the average value for a particular test is to the true value. If the accuracy is high, the value for the response in the test is very close to the true value; if accuracy is low, the response is significantly different from the true value. It is possible to have high precision and low accuracy.

How large does the difference in response variables have to be to indicate a significant effect is present in the data? This is determined by comparing the magnitude of the differences between the dependent responses at the various experimental points in the matrix. Duplicate or replicate experiments are run either at the center point or some other selected point within the experimental matrix. The difference in dependent responses between the replicated points determines the precision or variation within the data due to unassigned experimental error. The smaller the differences between the response values at the replicated points, the more significant larger differences are between responses at different experimental points within the design matrix. This concept will be discussed in more detail later.

The experiments comprising the DOE are run in random order to minimize systematic error that can lead to the wrong conclusions. As an example, assume a processing experiment is being run with two different raw materials. The first day all the experiments are processed using resin A. The second day the same experiments are run using resin B. If something happened to the dryer, heater bands are burnt out on the extruder, the extruder or feed throat cooling is not operating properly, or the take-off equipment temperatures are higher the second day and the malfunction or discrepancy is not noticed, the results might indicate that there is a difference between resin A and resin B. If the experiments had been run in random order with some of resins A and B processed both days, the conclusions from the DOE might be quite different.

This previous example raises a control question. Before doing any experimentation, verify that all components in the extrusion system are operating properly. The temperature controllers, heaters, thermocouples, and cooling in the various extruder zones must be operating properly. Temperature controllers, heaters, and cooling associated with the downstream equipment must be operating correctly. After all experimentation is complete, you don't want to have to apologize for the data because something mechanical malfunctioned. DOEs require significant time and effort to do properly, but the results are very rewarding, assuming everything is operating properly during the experimental phase.

Independent variables or factors are either fixed or manipulated during the experimentation. Fixed variables are not changed in the experimental design; these can be the extruder, downstream equipment, raw materials, environmental temperature, feeders used for additives, screw design, screen pack, die design, cooling temperature, draw ratio, and so forth. The other approach is to manipulate the independent variables by changing them over a specified range that defines the experiment and the experimental space.

25.1.3 Data Collection

The third step in the process is data collection. Once the experiments are defined, it is time to go into the lab or the production facility to execute the experiments and collect the data. Keep in mind, it is easier to run the experiments in a nonrandom order. However, this may introduce inherent errors that can lead to the wrong conclusion. Randomize the experimental order as much as

possible and do not run all the replicates or duplicate experiments one after the other. Collect all available processing data during the experimental phase on both the fixed and manipulated independent variables. Measure the response variables for each experimental test condition or treatment combination.

25.1.4 Data Analysis

The fourth step in the DOE process is to analyze the data. There are good computer programs available to do this. Computer programs predict which independent factors and interactions are significant (have an important effect) for a particular response, generate models to predict the dependent response at any experimental point or composition within the experimental matrix, plot the model equations to provide a visual comparison of the data, predict the experimental process condition or composition where the response is maximum, and predict or define an experimental operating range or composition where the properties meet the specifications or desired values. If the experimental design is for optimizing five different properties, one may define an experimental operating range where four of the criteria meet the goal but not all five. Possibly only two out of the five response variables can be optimized at a given set of processing conditions. The five properties can be weighted to define a processing region that maximizes the properties in order of importance. In other words, an operating range is defined where properties 1 and 4 are maximum, 2 and 5 are slightly below their maximum value, and property 3 is average. The operating conditions to obtain this were selected on the basis that properties 1 and 4 are the most important, followed by 2 and 5, and then 3.

Conclusions are based on statistical analysis and confidence levels. Statistics can't prove a factor has an effect, but they do verify reliability and test validity. The confidence limit determines a particular factor has an effect based on a specific level of confidence. Most computer programs use a 95% confidence limit as a means of saying a particular factor has a significant effect. The confidence level can be manually varied if desired. However, 95% is a good confidence limit when determining which factors or interactions have an effect.

Some computer programs to set up DOE designs and analyze the data are Design Expert by Stat-Ease, Inc.,[1] Statgraphics Plus,[2] S-Matrix,[3] JMP,[4] and MiniTab.[5]

25.1.5 Report Conclusions

The fifth and final step in a good DOE is to report the conclusions. Results generated but never reported or summarized did not happen. Without a summary report, the work will probably be repeated in the future because people will not remember the work was done or the results generated. With turnover in personnel, not only is the work forgotten, but the people who did the work also may have moved on to different jobs. If the work is important enough to do, it is important enough to write a report and take credit for a job well done.

25.2 Experimental Design

There are three experimental designs: mixture experiments, factorial design, and response surface. Each design is used in specific situations to gather information from a particular set of independent variables. Mixture designs are used to optimize formulations where the independent variables are ingredients of a mixture or recipe. Formulation optimization is based on measuring property performance (dependent variables or responses) over different compositions. Typical plastic or polymer formulation optimization includes

- Selecting the correct additive level to provide the optimum performance profile for heat, UV stability, or oxidative stability
- Evaluating the best filler concentration in one or more polymers
- Identifying the optimum polymer ratio in blends and alloys
- Determining optimum reinforcement and/or filler levels
- Optimizing an existing formulation for physical properties and cost effectiveness

Factorial design is used to screen process and/or product variables to determine which factors are significant in controlling the process. Normally, factorial designs are two-level designs, meaning a high and low value for each factor is used to determine whether an effect is present. Fractional factorial designs reduce the experimentation required and also decrease the information that can be obtained about potential interactions between independent variables in the experimental space. As the independent variables increase, the experiments required to understand which factors and interactions are significant increase.

Response surface designs are used after several independent variables have been identified and one wants a better description of their curvature and interactions in the experimental space. Response surface experiments, unlike factorial design, show curvature. A response surface is a geometric representation obtained when a response variable is plotted as a function of one or more quantitative independent factors. The results are graphed in two or three dimensions, showing the response as a function of the independent factors of interest. Response surfaces typically replace three-level factorial designs

(three-level factorials contain a low, high, and center value for each independent variable), because fewer experiments are required to obtain the same information. Mixture experiments are a special response surface experiment.

Some terms used in design of experiments are defined below:

- **Statistical model** is a model based on the dependent responses or variables. The model is based on observations and measurement error.
- **Observation** is the data obtained from the dependent responses for each set of independent variables evaluated.
- **Data** are characterized by mathematical terms; five terms are
  - — mean or average
  - — range
  - — standard deviation
  - — median
  - — coefficient of variance

The mean is the average value and is defined by Eq. (25.1):

$$\bar{X} = \frac{\sum_{i=1}^{n} x_i}{n} \tag{25.1}$$

where x_i = Response value
n = The number of x values

The range is the difference between the maximum and minimum response values.

The sample standard deviation (s), given by Eq. (25.2), measures variability within a given data set.

$$s = \sqrt{\frac{\sum_{i=1}^{n} (x_i - \bar{X})^2}{n-1}} \tag{25.2}$$

The median is the middle value in a group of samples.

The coefficient of variance is a percentage, given by Eq. (25.3):

$$COV \text{ or } CV = \frac{s}{\bar{X}} \times 100 \tag{25.3}$$

- **Experimental design** (layout) lists the experimental trials or treatment combinations. The design shows the randomization, replication, and independent factor levels within the design.
- **Effect** is the difference in the average response between two different levels of a given factor or between two experimental runs. A positive effect, shown in Fig. 25.1, increases as the independent factor changes from its low to high value. A negative effect decreases as the independent factor increases from its minimum to maximum value.

Figure 25.1. Effect definition.

- **Factor effect** is the average response for an independent factor at its high value compared to the average response for the same factor at its low value.
- **Experimental space** (factor space) is the region defined by the high and low levels for the independent factors in an experiment.
- **Interactions** occur when the response is different depending on the settings of two factors. Graphs of response Y versus independent variable A at the high and low levels of B will appear with two nonparallel lines, indicating that the effect of one factor depends on the level of the other.
- **Duplicate tests or replications** are when an entire experimental trial or treatment combination with the same independent factor levels is repeated one or more times to determine the inherent error in the various tests.
- **Degrees of freedom** is the number of independent comparisons available to estimate a parameter. It is usually the number of model parameters minus 1 or the number of free choices to estimate a parameter.
- **Error** comes from either assignable causes or random causes. Assignable causes are variations due to changes in the independent factors that are known and can be identified during the experimentation phase. Random errors are uncontrolled variations that are inherent in the measurement techniques and the test methods. These are unexplained errors that prevent responses from a similar set of experiments from producing identical results each time the experiment is run.
- **Analysis of variance (ANOVA)** is a mathematical procedure to compare the response data with the error data to determine whether an independent variable or interactions are significant. When analyzing a DOE, ANOVA uses the sum of squares to compare effects with the error variance to determine statistical significance.

- **F test** is used to test for significance of factors and interactions at a given probability level. The F test is a sampling distribution that predicts whether the results are significant at a given probability level, based on a sample variance. At a predefined confidence level (75%, 90%, 95%, 99%, and so forth), the F test predicts the probability that the F statistic is greater than the F value in an F value table. If the F value is greater than the value in the table at a given probability level, the independent factor and/or interaction from which the F test is calculated is deemed to have a significant effect on the response.
- **Model** is an empirical mathematical expression that describes the response variable in terms of the independent variables and their interactions within the experimental matrix. Typically, models are linear, quadratic, or cubic. The number of experimental trials in the design must exceed the number of terms in the model for the model to predict the response at any other point within the experimental matrix. A linear response surface model is a combination of independent factors given by Eq. (25.4):

$$Y_i = C + C_a A + C_b B + C_c C + \ldots \tag{25.4}$$

where Y_i = Dependent response variable
C = Constant
C_a = Factor for independent variable A
C_b = Factor for independent variable B
C_c = Factor for independent variable C

If more than three independent factors are evaluated, the final model can have as many terms plus the constant as there are independent variables being studied.

A quadratic model used to define the response surface is given by Eq. (25.5):

$$\begin{aligned} Y_i &= C + Linear\ Terms + C_{ab}AB + C_{ac}AC + C_{ac}BC + \ldots \\ &\quad + C_{a2}A^2 + C_{b2}B^2 + C_{c2}C^2 + \ldots \end{aligned} \tag{25.5}$$

where the linear terms are those in the linear model; AC, AB, BC, ... represent the potential two factor interactions; and A^2, B^2, C^2, ... represent the potential square terms.

A cubic model for a response surface series of experiments is given by Eq. (25.6):

$$\begin{aligned} Y_i &= C + Linear \text{ and } Quadratic\ Terms + C_{abc}ABC + \ldots \\ &\quad + C_{a2b}A^2B + C_{a2c}A^2C + C_{b2c}B^2C + \ldots \\ &\quad + C_{a3}A^3 + C_{b3}B^3 + C_{c3}C^3 + \ldots \end{aligned} \tag{25.6}$$

In response surface experiments, the quadratic model will fit approximately 95% of all problems.

In mixture experiments, the degree of the model indicates the presence of certain terms. Table 25.1 shows the degree and model description for linear, quadratic, special cubic, and cubic models.

Table 25.1. Degree and Description for Mixture Models

	Degree	Description
1	Linear	A B C ...
2	Quadratic	AB, AC, BC ...
3	Special Cubic	ABC ...
3	Cubic	AB(A – B) + AC(A – C) ...

Mixture experiments are unique, as the sum of all components equals 100%. This imparts special restrictions on the experimental points and formulations. Taking the high and low values for the experimental design points does not work, because the combination of these values must add to 100%. In most mixture experiments, a degree 2–quadratic or degree 3–special cubic model fits most of the data. In data analysis, always use the highest order model (model with the most terms) that is statistically significant. Never use a model with more terms than the number of experiments in the design matrix or evaluated during the experimental phase of the DOE. Typical mixture designs are Simplex-Lattice, Simplex-Centroid, and Simplex Augmentation.

Terms appearing in the computer analysis are

- **Residuals** refers to experimental error.
- **ANOVA** is the analysis of variance.
- **Mean square** is the sum of the squares for the model divided by the degrees of freedom.
- **Root MSE or root mean square** is the square root of the mean square, which is the standard deviation.
- **R2** is the correlation coefficient. The closer the R2 is to 1, the better the model predicts the response behavior.
- **Pure error** is the error expected in a response value if the experiment is completely rerun as a replicate.

To determine if a difference in measured responses is a true measure of a difference in the process, the data are analyzed using an ANOVA. Analysis of variance is a mathematical method that uses statistical tools to take into account the data and the errors. It gives you an understanding of what is a relevant difference versus a difference in data caused by errors through the use of an F test. The F test allows you to determine the significance of an independent variable on some dependent response

at a specific confidence interval, such as 75%, 90%, 95%, or 99% confidence limits. If the F value calculated in the ANOVA is greater than the value in the F tables, the independent factor or interaction is deemed to be significant at that probability or confidence limit.[6]

The math for calculating an ANOVA is beyond the scope of this book. An effective way to calculate an ANOVA is to use one of the computerized statistical analysis programs.

As an example, assume four materials are being compared for flow through a specific die under a fixed set of processing conditions. The four resins are designated as A, B, C, and D. Flow is critical to obtain the proper molecular orientation in the profile and surface aesthetics in the final part. Management is going to purchase one of these four materials, based on the die flow evaluation. The costs of the four resins vary slightly, with A being $1.63/pound, B is $1.78/pound, C is $1.68/pound, and D costs $1.72/pound. Four flow rates are measured for each resin, using the same processing conditions, and are given in Table 25.2 in units of pounds/hour. Based on the results, what material do you recommend management purchase and why?

Table 25.2. Experimental Data on Resin Flow

Resin	Run 1	Run 2	Run 3	Run 4
A	23.3	19.7	25.9	24.4
B	25.9	25.8	34.3	30.5
C	18.3	28.4	26.4	25.8
D	23.3	32.8	26.7	26.9

The following output was generated with MiniTab™ v. 13.3. From this output, one sees that the F value is 1.86, which gives a *p* value of 0.191. This means that there is an 80.9% chance that there is a significant difference in the data. The F value for 95% confidence can be looked up in an F distribution table for resin with 3 degrees of freedom, error at 12 degrees of freedom, and 95% confidence limit. The F value is 3.49 ($F_{\mu 1, \mu 2, \alpha} = F_{3, 12, 0.05} = 3.49$). If the F value from the MiniTab™ output is greater than 3.49, then there would be a 95% chance that there is a difference in the data. Along with the calculations, MiniTab™ provides a graphical output showing the four resins, their averages, standard deviations, and a bar graph with the averages and 95% confidence interval. From the numbers and observing the bar chart, one can see that the confidence intervals overlap, which means there is no difference in flow. In this case, one would buy the cheapest resin.

However, let us assume that you see the large variability in the flow data for each resin and decide to look into the matter. On careful examination you find some transcribing errors and some calculation errors. The revised data are presented in Table 25.3 with the corrected data highlighted in bold.

Table 25.3. Revised Resin Flow Data with Corrected Data in Bold

Resin	Run 1	Run 2	Run 3	Run 4
A	23.3	**22.5**	25.9	24.4
B	**35.9**	**32.8**	34.3	30.5
C	**28.3**	28.4	26.4	25.8
D	23.3	**25.8**	26.7	26.9

The following revised output (see next page) was generated with MiniTab™ v. 13.3. From this output, one sees that the F value is now 22.53 with a *p* value of 0.000. This means that there is almost a 100% chance that there is a significant difference in the data. Looking at the averages and the bar chart, now we see that resin B has a statistically significant higher flow than the other three resins. Even though resin B is 9.2% more expensive than the lowest cost resin, it has a 41.6% higher flow. Thus one has a case for recommending resin B because of its higher productivity.

One-way ANOVA: Flow versus Resin

```
Analysis of Variance for Flow
Source     DF        SS        MS         F         p
Resin       3      81.9      27.3      1.86     0.191
Error      12     176.5      14.7
Total      15     258.5
                                   Individual 95% CIs For Mean
                                   Based on Pooled StDev
Level       N      Mean     StDev  -----+---------+---------+---------+-
A           4    23.325     2.641  (---------*---------)
B           4    29.125     4.088                 (---------*---------)
C           4    24.725     4.425        (---------*---------)
D           4    27.425     3.946              (---------*---------)
                                   -----+---------+---------+---------+-
Pooled StDev = 3.835                 20.0      24.0      28.0      32.0
```

One-way ANOVA: Flow 2 versus Resin

```
Analysis of Variance for Flow 2
Source     DF        SS        MS          F          p
Resin       3    198.74     66.25      22.53      0.000
Error      12     35.29      2.94
Total      15    234.03
                                   Individual 95% CIs For Mean
                                   Based on Pooled StDev
Level       N      Mean     StDev  -----+---------+---------+---------+-
A           4    24.025     1.473  (-----*-----)
B           4    33.350     2.261                        (-----*-----)
C           4    27.225     1.323           (-----*-----)
D           4    25.675     1.654       (-----*-----)
                                   -----+---------+---------+---------+-
Pooled StDev = 1.715                 24.0      28.0      32.0      36.0
```

The above example shows an ANOVA analysis and how to interpret the results. When doing DOE experiments, the ANOVA is performed by the computer program. The same computer programs set up the experimental design and do the math necessary to generate the equations and models. To summarize the procedure, to do DOE experiments in extrusion, define the problem, establish the experimental limits, run the experiments, input the dependent responses into the program to analyze the data, and interpret the results.

25.3 Experiments

The last section in this chapter sets up some experimental design examples for different DOE experiments.

25.3.1 Factorial Design

Factorial experiments are used to screen many independent variables to determine what independent factors are significant in controlling the performance of various response variables being evaluated. As an example, consider a plastic sheet 0.125 inch (3.2 mm) thick, extruded on a single screw extruder using a three-roll stack and a flexible lip die. An understanding of how processing conditions affect product properties is desired. The processing conditions (independent variables) being evaluated are throughput rate, screw speed, barrel temperature, die lip opening, and roll temperatures. Physical properties and process parameters (response variables) being measured are color, gloss, impact, tensile properties in both the machine and transverse directions, melt temperature, torque, and die pressure. What is the best procedure to optimize the process? A factorial design provides a systematic approach to determine which experimental factors and interactions are significant in controlling the different responses being evaluated.

The first step is to define the experimental range for each dependent variable. Let us choose the following processing limits:

- Throughput rate: 200–300 pounds/hour
- Screw speed: 250–400 rpm
- Barrel temperatures: High and low temperature profile
- Die lip opening: 0.125–0.250 inch (3.2–6.4 mm)
- Roll temperatures: 125˚–175˚F (52˚–79˚C)

Using a full factorial design to identify all the significant effects and interactions, there are 2^k experiments, where k is the number of independent variables. For five independent variables, there are 2^5 or 32 experiments. Fractional factorial designs require less experimentation but do not define all the interactions that can be present. A 1/2 factorial with five independent variables requires 16 experiments, while a 1/4 factorial requires 8 experiments. In laying out full factorial experiments, the high and low value for each independent factor is combined with the high and low value of all other independent variables. To set up the experiments using a full factorial design, generate a table with a column for each factor. In the first column, insert the high independent factor value in the first $n/2$ (where $n = 2^k$) rows and the value for the low independent factor in the second set of $n/2$ rows. For the second independent factor, set the first $n/4$ rows at the high value, the second $n/4$ rows at the low value, the third $n/4$ rows at the high value, and the final $n/4$ rows at the low value. The same procedure is followed with the third column, except the first $n/8$ rows contain the high value and the next $n/8$ rows the low value, and so forth, until the column is full. The fourth factor is treated the same way, except the number of rows alternated is $n/16$. The fifth factor rows alternate every other row, with one being the high value and the other low, or $n/32$ rows with the high value followed by $n/32$ rows with the low value.

This procedure produces all possible combinations of maximum and minimum values.

Table 25.4 contains the 32 experiments for the full factorial design to screen the five independent processing factors. Response variables are collected for all 32 experiments or treatment combinations. The response model contains the main effects plus all the interactions. Factorials or screening experiments are used to screen many variables to determine which factors and interactions are significant. Fractional factorials are used to determine which main effects are important and the effect of some interactions, while minimizing the number of experiments. Fractional factorials are normally one of the first steps in an evaluation procedure. The important variables can be identified and then evaluated in more detail. After screening experiments are completed, the next step is to take the significant factors (hopefully a smaller set of independent variables) and perform a

Table 25.4. Full Factorial Experiments for 5 Factor Examples

Experiment	Throughput, lbs.hr	Screw Speed, rpm	Barrel Temp., F	Die Lip, in.	Roll Temp., F
1	300	400	HI	0.250	175
2	300	400	Hi	0.250	125
3	300	400	Hi	0.125	175
4	300	400	HI	0.125	125
5	300	400	Lo	0.250	175
6	300	400	Lo	0.250	125
7	300	400	Lo	0.125	175
8	300	400	Lo	0.125	125
9	300	250	HI	0.250	175
10	300	250	Hi	0.250	125
11	300	250	Hi	0.125	175
12	300	250	HI	0.125	125
13	300	250	Lo	0.250	175
14	300	250	Lo	0.250	125
15	300	250	Lo	0.125	175
16	300	250	Lo	0.125	125
17	200	400	HI	0.250	175
18	200	400	Hi	0.250	125
19	200	400	Hi	0.125	175
20	200	400	HI	0.125	125
21	200	400	Lo	0.250	175
22	200	400	Lo	0.250	125
23	200	400	Lo	0.125	175
24	200	400	Lo	0.125	125
25	200	250	HI	0.250	175
26	200	250	Hi	0.250	125
27	200	250	Hi	0.125	175
28	200	250	HI	0.125	125
29	200	250	Lo	0.250	175
30	200	250	Lo	0.250	125
31	200	250	Lo	0.125	175
32	200	250	Lo	0.125	125

response surface DOE to define the optimum processing conditions, or complete a factorial design with a center point and replicate experiments. Qualitative variables—such as the barrel temperature in the example, are high or low, on or off, and yes or no—can be used in factorial design but not in response surface designs.

25.3.2 Response Surface

A response surface is a geometrical representation of a response variable plotted as a function of the independent variables. This experiment provides more information about a dependent or response variable than a two-level factorial or fractional factorial design. A three-level factorial design has a center point included for each independent variable along with the high and low points, requiring three experiments for each independent variable. This is called a three-level factorial design because of the third factor level. Inclusion of the third factor greatly increases the number of experiments. In the previous factorial design with five variables, there are 2^k or $2^5 = 32$ experiments. The same independent factors using a three-level factorial design has 3^k or $3^5 = 243$ experiments. Qualitative variables with only two levels (high/low, on/off, yes/no) cannot be used in three-level factorial design.

A three-level, two-factor design with 3^2 or 9 experiments is shown in Table 25.5, with the high and low values for factors A and B being represented as 1 and -1, and the middle value as 0.

As three-level factorial designs generate large treatment combinations or experimental points, alternate response surface designs were developed to reduce the number of experiments required while providing similar information. Response surface analysis generates models that are used to plot contours in one or two dimensions, depending on the factors used to describe the response variable throughout the experimental space. The experiments in response surface designs are special cases of factorial design that include center points in the experimental space plus edge center points or face center points. The multi-level experimental points allow fitting the responses to quadratic or cubic equations, which provides a better model of the response variable in the experimental space. Designs that are more practical than the three-level factorial designs are Central Composite and Box Behnken designs.

Table 25.5. Three Level, Two Factor Factorial Design

Experiment	Factor A	Factor B
1	–1	1–
2	–1	0
3	–1	1
4	0	–1
5	0	0
6	0	1
7	1	–1
8	1	0
9	1	1

The Central Composite design is based on a two-level factorial design with the addition of $2k$ (k is the number of independent variables) points (star points) between the axes plus repeat points at the centroid. As with all good experimental designs, the experiments are randomized. Consider an example similar to that used for the factorial design, with the exception that the variables are

- Screw speed (250–400 rpm)
- Throughput (200–300 pounds/hour)
- Die lip opening (0.125–0.250 inch [3.2–6.4 mm])

What are the experiments and what is the appearance of the experimental space? Figure 25.2 shows the experimental space for a three-factor Central Composite design. Depending on the design, the star points or the center points in each face may extend out past the face, as shown in Fig. 25.2. In most plastic or extrusion designs, the star points have to be located in the plane with the other coordinate points (shown in the left-hand drawing of Fig. 25.2), because star points outside the plane may not be practical processing conditions. In the current example, high values associated with the star points on the throughput may exceed the torque or screw speed limitations of the equipment; the very high screw speed would be about 450 rpm. In any design, all experimental points must be physically possible to process.

The other common response surface design is a Box Behnken design, where the experimental points are located at the center points of the edges, Fig. 25.3. The centroid or center point of the experimental space is replicated. A full three-level factorial design with three independent variables requires 3^k experiments or $3^3 = 27$ experiments, shown in Fig. 25.4. In addition to the high and low exper-

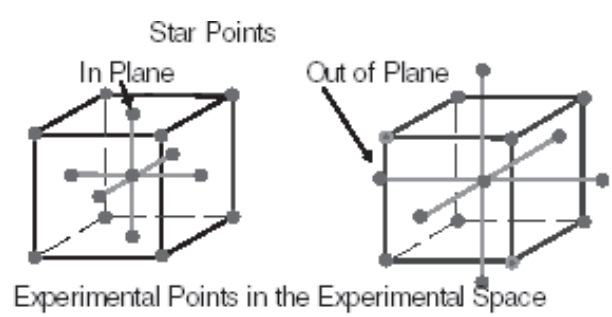

Figure 25.2. Central composite design with star points.

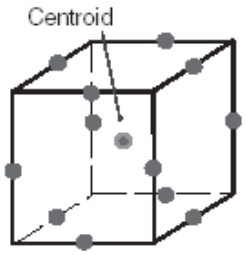

Figure 25.3.
Box Behnken design.

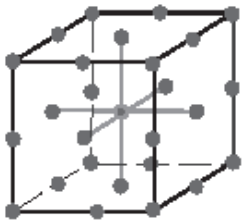

Figure 25.4.
Three-Level factorial.

imental values, the three-level factorial design includes center points on each edge, the center point of each face, and the center point plus replications to determine experimental error. A three-level factorial design to obtain similar information as a Central Composite or Box Behnken design contains 32 experiments.

Using the example to determine the effect of screw speed, throughput, and die lip opening on physical properties, the randomized experimental layout for the three designs is given in Tables 25.6–25.9, with Tables 25.6 and 25.7 showing Central Composite design with the star points in and out of the plane, corresponding to the left- and right-hand schematics in Fig. 25.2. Table 25.8 provides a Box Behnken design, and Table 25.9 shows the three-level factorial design. In Table 25.7, the star points not within the plane are experiments 3, 4, 6, 9, 13, and 16; experiment 4, making a 0.125 inch thick sheet product with a 0.08 inch die opening, is probably impossible to achieve. Experiment 9, requiring 250 pounds/hour throughput rate with a 199 rpm screw speed, may be impossible to run without torquing out the extruder. Experiment 6, with a 451 rpm screw speed, may exceed the extruder screw speed capabilities. The die opening in experiment 13 may be too large to draw down the sheet to 0.125 inch thickness without excessive molecular orientation or necking in the sheet. Central Composite experiments laid out in Table 25.6 are more practical to use for the experimental design. The center points repeated in Table 25.6 are experiments 6, 7, 10, 13, and 16. Table 25.8 shows the experimental points using a Box Behnken design. There are three fewer experiments required with this experimental design. As with the other designs, the experimental order is randomized and the center point is repeated, experiments 2, 5, 10, 13, and 17.

Table 25.6. Central Composite Design with Star Points in the Plane

Experiment	Screw Speed, rpm	Throughput, pounds/hour	Die Opening, Inch
1	325	250	0.25
2	250	300	0.12
3	400	300	0.25
4	250	250	0.19
5	400	200	0.25
6	325	250	0.19
7	325	250	0.19
8	325	200	0.19
9	250	300	0.25
10	325	250	0.19
11	250	200	0.25
12	325	300	0.19
13	325	250	0.19
14	400	250	0.19
15	325	250	0.19
16	325	250	0.19
17	400	200	0.12
18	400	300	0.12
19	250	200	0.12
20	325	250	0.12

Table 25.7. Central Composite Design with Star Points Outside the Plane

Experiment	Screw Speed, rpm	Throughput, pounds/hour	Die Opening, Inch
1	325	250	0.19
2	325	250	0.19
3	325	334	0.19
4	325	250	0.08
5	325	250	0.19
6	451	250	0.19
7	400	200	0.25
8	325	250	0.19
9	199	250	0.19
10	250	300	0.25
11	325	250	0.19
12	250	200	0.12
13	325	250	0.29
14	400	200	0.12
15	250	300	0.12
16	325	166	0.19
17	325	250	0.19
18	400	300	0.12
19	250	200	0.25
20	400	300	0.25

Table 25.8. Box Behnken Design

Experiment	Screw Speed, rpm	Throughput, pounds/hour	Die Opening, Inch
1	400	300	0.19
2	325	250	0.19
3	400	250	0.12
4	400	200	0.19
5	325	250	0.19
6	250	250	0.25
7	400	250	0.25
8	325	200	0.12
9	250	200	0.19
10	325	250	0.19
11	325	200	0.25
12	325	300	0.12
13	325	250	0.19
14	325	250	0.19
15	250	250	0.12
16	325	300	0.25
17	325	250	0.19

Table 25.9 shows the experimental points with a three-level factorial design with the center point replicated six times in experiments 4, 8, 13, 19, 21, and 31. The advantage of using either the Central Composite or Box Behnken designs to generate response surfaces is that fewer experiments are required.

Using either the Central Composite or Box Behnken design to fit a cubic model requires more experimentation so that there is at least one experiment for each term in the model. Since quadratic models fit 95% of all experimental designs, starting with a quadratic model makes sense. Additional experiments can be added later if a higher order model is required to predict the responses.

25.3.3 Mixture Experiments

Mixture experiments are used when experimenting with composition ingredients. Typical uses include

- Formulation optimization and optimizing property profiles with different additive levels
- Combinations of different resins in alloys or blends
- Determination of impact modifier concentrations
- Flame retardant formulation optimization
- Effect of filler and/or reinforcement concentrations
- Effect of regrind levels

Table 25.9. Three Level Factorial Design

Experiment	Screw Speed, rpm	Throughput, pounds/hour	Die Opening, Inch
1	250	200	0.12
2	250	250	0.25
3	400	250	0.19
4	325	250	0.19
5	400	200	0.12
6	400	300	0.12
7	400	250	0.25
8	325	250	0.19
9	325	250	0.25
10	400	200	0.19
11	325	200	0.12
12	250	300	0.25
13	325	250	0.19
14	250	300	0.12
15	325	200	0.25
16	250	250	0.19
17	250	200	0.19
18	325	300	0.25
19	325	250	0.19
20	325	300	0.19
21	325	250	0.19
22	250	300	0.19
23	400	300	0.19
24	250	250	0.12
25	400	200	0.25
26	325	200	0.19
27	250	200	0.25
28	325	250	0.12
29	325	300	0.12
30	400	250	0.12
31	325	250	0.19
32	400	300	0.25

Experimental responses include

- Physical properties
- Cost
- Flame retardance
- Chemical resistance
- UV stability
- Thermal stability
- Dimensional stability

As an example of a mixture experiment, assume resin D, resin E, and talc are being mixed, and you want to understand the proper proportions of each ingredient to optimize tensile strength, flexural modulus, and impact strength. The experimental limits are defined as

- D is present from 38 to 72%.
- E = 5 to 85%.
- Talc = 10 to 55%.

What experiments are required to establish the property profiles at any composition within the design matrix, and what is the appearance of the experimental space?

Figure 25.5 shows the experimental space for a mixture experiment with three variables. Triangular coordinates are used to satisfy the mixture constraint, which requires all components to add to 100% for each experimental condition. Any point within the experimental matrix in Fig. 25.5 totals to 100%. The horizontal axis starting at the bottom of the triangle is 0% talc, with each horizontal line representing a 10% change until the apex is reached at 100% talc. The other axes are interpreted in the same manner. Consequently, any point in the experimental space is a combination of the three components. Using the three-component coordinate space in the mixture example, the experimental space is defined by the limits shown in Fig. 25.6. Combinations of independent factors to define the response variables within the experimental space are the vertices, the midpoints of long edges, the centroid, and a checkpoint. With the duplicate experiments included to define the experimental error, the coordinates of the experimental compositions are given in Table 25.10. Thirteen experiments provide sufficient data to generate models and response curves to predict the dependent response at any composition within the experimental limits. Experiments 1 and 2, 4 and 8, 5 and 13, and 8 and 12 are repeated experiments used to determine experimental error.

Once the dependent responses are determined and models generated, a processing or compositional window can be defined where the dependent responses are better. In most situations, it is a compromise, because all response variables or, in this case, physical properties are not normally maximized at the same compositional or processing point. Optimization of one dependent response may be at a completely different location within the experimental space compared to the maximum for the next dependent response. A decision is required to determine where the best property balance is obtained within the

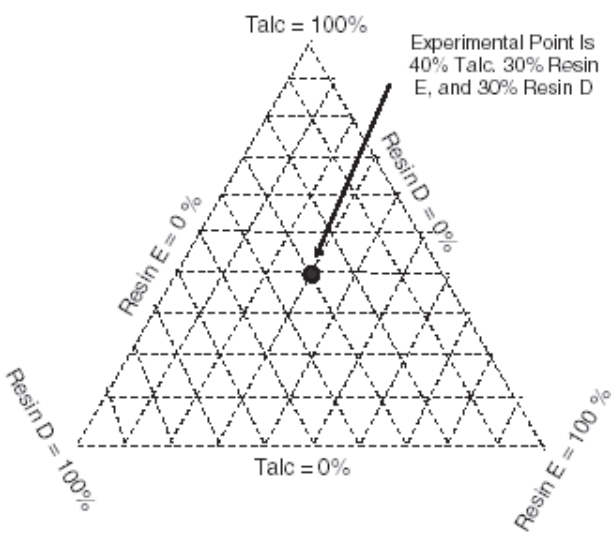

Figure 25.5. Experimental coordinates for a three-component mixture experiment.

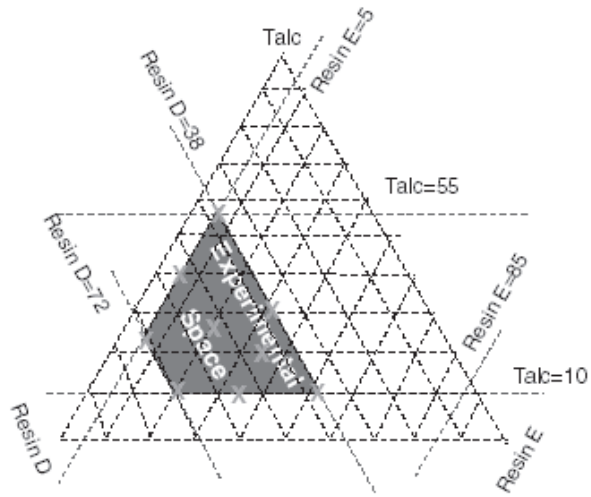

Figure 25.6. Mixture experiment example—composition points and definition of experimental space.

Table 25.10. Mixture Experiment Example

Experiment	Talc, %	Resin E, %	Resin D, %
1	32.5	29.5	38.0
2	32.5	29.5	38.0
3	10.0	35.0	55.0
4	10.0	52.0	38.0
5	10.0	18.0	72.0
6	30.6	17.4	52.0
7	39.0	5.0	56.0
8	55.0	7.0	38.0
9	10.0	52.0	38.0
10	20.3	34.7	45.0
11	23.0	5.0	72.0
12	55.0	7.0	38.0
13	10.0	18.0	72.0

experimental matrix. If a composition is selected where all the responses are not at their maximum value, does the compromise meet the customer requirements?

In any factorial, response surface, or mixture experiment, once the optimum combination of processing conditions or compositional ingredients is chosen based on a predicted response, additional experimentation is needed to verify the predictions and verify the response profile predicted is routinely attainable in practice.

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Review Questions

1. What is the definition of experimental space?
2. What are three experimental designs and where is each used?
3. Why use a DOE?
4. Where should DOE experiments be used and where are they inappropriate?
5. What are the five steps in doing a DOE? What is the objective of each step?
6. What is the difference between independent and dependent variables? Give some examples in extrusion applications.
7. What is the difference between an independent variable and a factor?
8. Why is it critical to run experiments in random order?
9. Why is it necessary to run duplicate experiments or repetitions?
10. What is a statistical model and how is it used?
11. What are four mathematical terms describing variability, and how do they differ from one another?
12. What is the definition of an effect? What is the difference between positive, negative, and no effects?
13. Why is an F test important in DOE?
14. How is analysis of variance used in DOE?
15. In response surface experiments, what terms are included in quadratic, linear, and cubic models?
16. In mixture experiments, what is the difference between quadratic, special cubic, and cubic models?

Review Questions (continued)

17. What type of experimental design can use qualitative variables?
18. How many experiments are required in a full two-level factorial with six variables, in a 1/2 factorial, and in a 1/4 factorial?
19. What is lost by doing fractional factorials?
20. What is a three-level factorial design and how many experiments are required with four independent variables?
21. What is the difference between a Central Composite and a Box Behnken design?
22. What constraint is present in mixture experiments, and how does this affect the experimental space?
23. How are data analyzed in DOEs?

Quality assurance certifies that products meet all the finished product specifications. The Quality Assurance Department strives for continuous improvement in product quality while working jointly with manufacturing, which seeks continuous improvements in plant yields and efficiency. Other Quality Assurance Department objectives related to the product are to develop and establish procedures to monitor

- Process variables
- Finished product quality and provide certification of compliance for the finished product
- Incoming specifications, certification acceptance, and raw material testing

To complete these responsibilities, the Quality Assurance Department develops and implements written procedures to

- Test and evaluate both incoming raw materials and finished product specifications
- Monitor the different steps in the extrusion process
- Develop sampling methods and frequency
- Release products and classify defective products for rework or other disposition
- Develop sampling procedures to test out-of-specification products
- Accept or reject finished product

An additional Quality Assurance Department function is to follow and monitor all procedures associated with certification by the International Organization for Standardization (ISO). Part of ISO certification is the specification and documentation of methods and procedures to use in performing certain tasks. Paper trail documentation is required for each procedure, verifying it is done the same way each time, repeated for every product, and records must be maintained substantiating all the measurements.

Raw material databases are required for each ingredient used in the various extrusion formulations. Vendor "certification of compliance" is used in accepting numerous raw materials; however, an internal database generated through periodic raw material testing is extremely valuable in case extrusion problems arise that are attributable to raw materials. Incoming specifications can be compared to historic data generated in-house to determine if something has changed that may affect the extrusion operation. Raw material requirements are based on process requirements and finished product specifications sought by customers for a particular extruded application. Test raw materials for properties that are directly related either to internal process capabilities or finished product specifications. As an example, assume you are purchasing a filled resin system; incoming raw material specification tests need to monitor filler content and uniformity, not impact properties. If thermal stability is critical for a high temperature application, a test to measure the color change at elevated temperature over time or weight loss at elevated temperature can be used to determine the resin system stability. In defining tests and specifications for raw materials, it is critical to test those properties that directly relate to your processing capabilities and customer product requirements, and not run or specify tests for "certificates of compliance" that don't correlate with either processing capabilities or customers' needs. Typical raw material specifications may include melt flow index, color, thermal stability, UV stability, flame retardance, reinforcement content, filler content, gel level, impact, and so forth.

At the other end are finished product specifications. Is your manufacturing process capable of producing a product that meets the customer specifications 100% of the time? Prior to accepting any customer order or long-term contract to produce a product for a given application, quality assurance needs to verify the product can be produced within the specification and tolerance limits required by the customer at high yield. As an example, assume the customer has specified a fiber reinforced product with $35 \pm 0.3\%$ reinforcement at ± 6 sigma; is your process capable of producing that product? If you make a 35% filled material with a standard deviation of 0.1%, the product will not always meet the specification of 0.3% at ± 6 sigma.

It is essential to understand the customer requirements and what is acceptable and what is not. Assume production of a nonvisual part that contains some surface imperfections that do not affect the product performance. What basis is used to determine the level of visual defects that can be shipped to the customer? Discussions with the customer can provide what imperfections are unacceptable prior to starting production. A display board demonstrating what is acceptable and what has to be rejected needs to be established. In a situation where a product is being visually inspected by operators, some unacceptable product will be shipped while acceptable product is occasionally discarded. Quality cannot be inspected into a product; where qualitative decisions are required by employees, some out-of-specification product will pass through the system as acceptable product. People are human and do not visualize each situation or part exactly the same. A display board with acceptable and unacceptable product assists in training and understanding what is acceptable to the customer, but it does not eliminate some human error.

In the previous example, where a visual defect does not affect product performance, establishing product acceptability with the customer is critical. While it may be possible to produce products that contain no surface defects, the manufacturing costs may be higher because the throughput rate required to run the process has to be reduced, the scrap rate increases, different raw materials that cost more are required, and so forth. Negotiations are required with the customer to establish the level of acceptable defect versus the cost of doing business at that level of quality.

The best method of assuring quality parts being shipped every time to the customer is through the use of statistical process control (SPC).

26.1 Statistical Process Control

Statistical process control is a method to ensure the product meets the customer specifications. It assumes a process run under the same operating conditions with all equipment functioning properly and using the same quality raw materials entering the process will produce the same product within the random unassignable system error every time. Instead of monitoring product performance criteria such as product properties, the process parameters are monitored and plotted using process limits to determine if the process is in control. Process control is a methodology used to produce predictable product quality through process understanding and monitoring. It provides information to be used in decision making and detects positive and negative changes in the process that affect the product.

Statistical process control uses statistical techniques and thinking to understand and solve problems within the extrusion process. The goal is continuous process improvement, focusing on error prevention and process reproducibility. Chris Rauwendaal has written a book on the use of SPC to control extrusion processes.[1]

The principal vehicle to accomplish process control objectives is control charts that are used to monitor, record, and compare process data throughout the process. Control charts are used to understand the process stability and process variation, and to provide uniform product output. Control charts contain a midline centering the process and upper and lower control limits. Control limits are calculated from the process data; they are not engineering specifications. Process data, centered around the center or midline of the chart, are a random distribution of measured values. In situations where the data are not random, or points are outside the upper or lower control limits, the process is deemed out of control and the product is susceptible to not meeting specifications.

The general theory of control charts was first proposed by Dr. Walter S. Shewart, and control charts developed according to these principles are often called Shewart control charts.[2]

A control chart, also called an X-bar chart, shown in Fig. 26.1, represents a typical process that is in control. The process data are randomly centered on the chart midline.

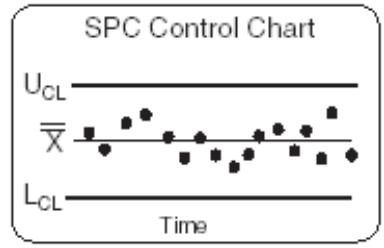

Figure 26.1. Process control chart

U_{CL} is the upper control limit defined by Eq. (26.1), where A_2 is a factor for determining the control limits based on subgroup size and R is the range; L_{CL} is the lower control limit defined by Eq. (26.2). The X-bar is the grand average and is given by Eq. (26.3), where k is the number of averages combined:

$$U_{CL} = \bar{\bar{X}} + A_2 \times R \tag{26.1}$$

$$L_{CL} = \bar{\bar{X}} - A_2 \times R \tag{26.2}$$

$$\bar{\bar{X}} = \frac{\left(\bar{X}_1 + \bar{X}_2 + \bar{X}_3 + \bar{X}_4 + \ldots\right)}{k} \tag{26.3}$$

Out of control situations are shown in Fig. 26.2. Variations in the data occur because of either common cause or special cause. Common cause is inherent random variation in the process that is always present. It accounts for the difference between two typical measurements. Special cause is variation in the process that is assignable to some specific cause or effect; for example, a heater band burned out, the cooling is not functioning properly, an outside door opened in the wintertime causing a blast of cold air on the extruder, water bath temperature is too hot, the wrong raw material or formulation ratio was used, and so forth. When a special cause is identified, it is written on the control chart, identifying the reason an out of control situation occurred or large deviations in the data were observed.

Different groups use different criteria to define when a process is out of control. One way to define out of control situations identified in Fig. 26.2 is when

- A data point or points are above or below the upper or lower control limits, respectively.
- Two out of three consecutive data points are near either the upper or lower control limits.
- A run occurs with seven or more consecutive data points either above or below the midline of the graph.
- A trend occurs where six or more data points are in either an increasing or decreasing order.

- A cycle occurs when the data points are present in a repetitive pattern.
- Zone rules are enforced in either of two situations. First, two out of three consecutive points fall more than two standard deviations from the mean. Second, four out of five consecutive points fall more than one standard deviation from the mean.
- Shift in level occurs when a number of points are below the midline followed by a group of data points above the midline.
- Stratification is defined as a situation where 15 or more points are located near the midline.
- Clusters are groups of data points occurring in one area of the chart.
- Mixture is a situation where there is an absence of points near the midline.

If any of the above situations exist, the process is deemed out of control. Determining why the process is out of control and what corrective actions are required to bring the process back into control is needed.

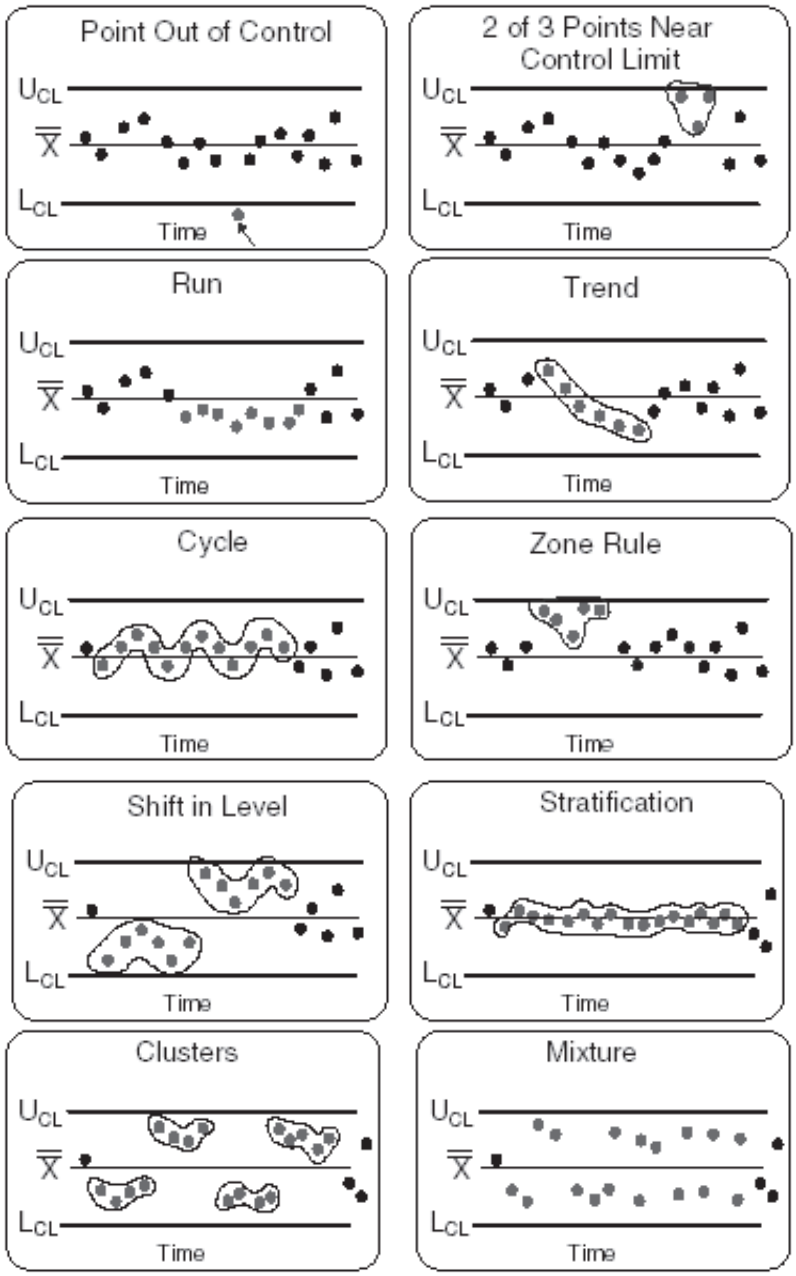

Figure 26.2. Out of control situations depicted on different control charts.

26.2 Process Capability

Process capability is a measure of the inherent process performance. It is defined by sigma (σ), the standard deviation. Different σ levels are used to determine process capability, depending on the customer's needs and specifications. The data included in different standard deviation ranges are

- ±1 σ includes 68.27% of the total area under a normal distribution curve. If the process is run at ±1 σ capability, 317,300 parts out of every million fall outside the specification limits.
- ±2 σ includes 95.45% of the total area under the normal distribution curve, with 45,500 parts out of a million falling outside the control limits.
- ±3 σ includes 99.73% of the total area under the normal distribution curve or virtually the entire area. At ±3 σ, there are still 2,700 defective parts out of each million produced.
- ±6 σ includes 99.9999998% of the area under the normal distribution curve, and 0.002 parts per million are expected to be defective.

In a 6 σ process, statisticians allow for a 1.5 σ shift. This adjustment results in 3.4 defects per million parts produced.[3]

As an example, assume a sheet product is being shipped to customer RSQ, who requests the impact strength to be 13 ±3 ft-lbs at a ±3 σ level. To supply samples to RSQ, some sheet is produced and impact properties measured. Thirty-seven data points are gathered and plotted to give a normal distribution, as shown in Fig. 26.3. Based on the data, can your company supply product to

RSQ that meets the customer requirements 100% of the time? The average impact value is 13 with a standard deviation of 1.25. At 3 σ the data are anticipated to range from 13 ±3(1.25), giving a range of 9.25 to 16.75. Based on the data without process improvements to lower the standard deviation, it is impossible to satisfy RSQ's impact requirements. The process is incapable of producing product with 13 ±3 ft-lbs at 3 σ that meets the customer's requirements 100% of the time. If the order is accepted based on the current operation, product will be produced and sent to the customer that is outside the specification limit.

Process capability measures the process repeatability relative to the customer specifications. Figure 26.4 shows two normal distribution curves defining product property profiles with specification limits. Process A is capable of producing a product that meets the customer specifications 100% of the time, while process B is an incapable process.

Process capability is measured through a capability index, C_{pk}, defined by Eqs. (26.4) and (26.5), where U_{SL} is the upper specification limit and L_{SL} is the lower specification limit. A C_{pk} value less than 1 indicates the process is not in control.

$$C_{pk} = \frac{Allowable\ Spread\ in\ Specifications}{Actual\ Spread\ in\ Specifications\ (Measurements)} \tag{26.4}$$

$$C_{pk} = \frac{U_{SL} - Mean}{3\sigma} \text{ or } \frac{L_{SL} - Mean}{3\sigma} \tag{26.5}$$

C_{pk} values of 1.33 and 2 indicate 6 parts out of every 100,000 and 2 parts out of 1,000,000,000 are defective or outside the allowable spread in specifications, respectively. Parts within the tolerance limits with a C_{pk} value of 1.33 are 99.994% in specification.

The final concept in process capability is the drive for continuous process improvements and zero defects. Zero defects is a quality system goal to remove all defects from the product.

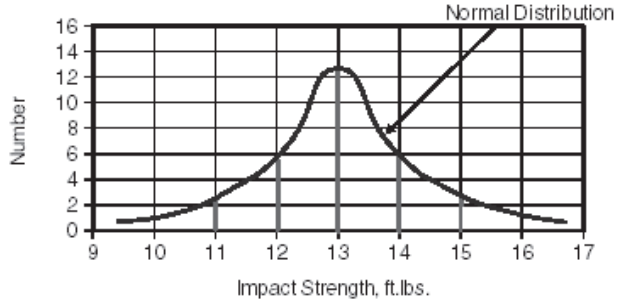

Figure 26.3. Normal distribution—example with impact data.

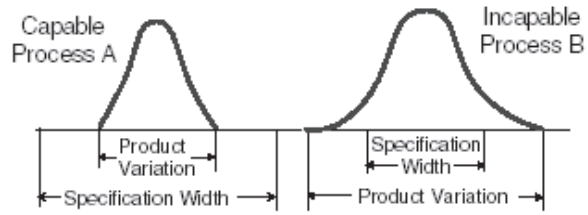

Figure 26.4. Comparison of capable and incapable processes.

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1. Rauwendaal, C., *SPC Statistical Process Control in Extrusion*, Hanser, New York, 1993.
2. Montgomery, D.C., *Introduction to Statistical Quality Control*, 3rd Ed., Whiley, p. 135, 1997.
3. Harry, M.J., *6 Sigma Producibility Analysis and Produce*, Motorola University Press, 1998.
4. Crosby, Philip, *Quality Management and Applied Statistical Process Control*, The Creative Factory, Inc., 1990.

Review Questions

1. What are some of the functions of the Quality Assurance Department?
2. What is C_{pk}? What does it measure? How is it used?
3. Using control charts, define five situations where a process is out of control and how it is recognizable on a control chart.
4. What are some possible procedures for checking incoming raw materials?
5. What are the purposes of using control charts, and how can they improve quality and productivity?

27 Troubleshooting Mechanical Extrusion Problems

Solutions and corrective actions to implement when troubleshooting various extrusion problems are grouped together depending on the extrusion process and the problem. This chapter addresses mechanical problems associated with the extruder; Chapter 28 addresses issues with materials or products produced during extrusion, while Chapters 29–32 address problems unique to specific extrusion processes. Coextrusion troubleshooting is presented with the coextrusion information in Part 6. Some problems where material and equipment interact are addressed in Chapter 28.

To be a good problem solver, it is necessary to understand both the materials and the equipment and how they interact with each other. An engineer or operator who knows the equipment is able to solve some equipment-related problems, but other problems require understanding both equipment and material behavior in the extruder. To correctly identify the problem root causes and institute the proper corrective action in the most expedient manner, one needs to understand the equipment, process, product, and their interactions.

Potential mechanical extrusion problems associated with the equipment are identified below with suggested causes and corrective actions to eliminate the problem. While all suggested corrective actions do not apply in all situations, going through the causes and suggested responses will help identify the proper corrective steps to employ in different situations.

27.1 Problem 1—Extruder Screw Doesn't Turn

When the screw speed potentiometer is turned to start the screw rotation, nothing happens; the extruder screw does not rotate. Possible causes with corrective actions for this problem are

- Voltage to the motor is not turned on. Check all circuit breakers and power to the extruder.
- Barrel temperatures are too cold. Resin was left in the extruder. Barrel temperatures are not sufficiently high to melt the resin (crystalline) or soften the resin (amorphous) to allow the screw to rotate. More time or temperature is required to heat soak the barrel to reach the correct processing temperatures.
- Gear box oil pump is not on. Oil is not being pumped to the gear box to lubricate the bearings. Either the power to the pump is off, resulting in no power, or the oil in the gear box is overheated, and the pump turned itself off automatically through the high temperature alarm sensor in the oil.
- Water to the gear box is not turned on. Oil in the gear box overheated because the water cooling to the gear box is not turned on, causing the oil to overheat and shut down automatically.
- Foreign object in the feed throat prevents the screw from turning. Metal or other foreign object is wedged in the extruder feed throat, preventing the screw from rotating.
- Torque converter in twin screw extruder is disengaged. When the maximum torque limit is exceeded, the torque converter between the screw and the gear box will disengage. Torque converter has to be reengaged and the extruder screw started back up slowly.
- Torque limit exceeded. The torque limit or motor load is exceeded. Torque converter might have disengaged in a twin screw extruder. Torque overload in either a single or twin screw extruder will shut the power off to the drive. Reset the extruder and, if necessary, reengage the torque converter on a twin screw extruder, and start the screw speed up slowly without feeding any resin until extruder screw speed is normal and the torque is reduced.

Figure 27.1 summarizes the potential causes and corrective actions to take when an extruder screw won't turn.

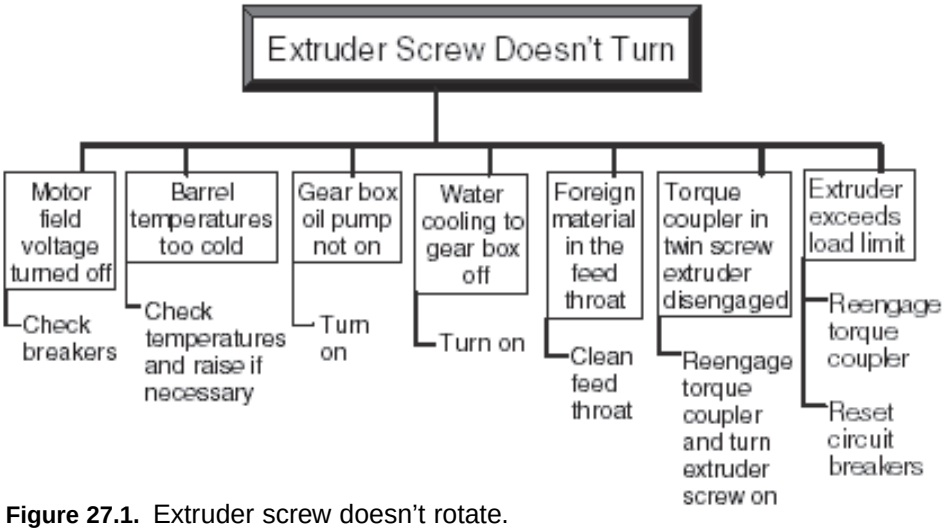

Figure 27.1. Extruder screw doesn't rotate.

27.2 Problem 2—DC Motor Will Not Start[1]

This problem occurs when the power is energized to the extruder and drive, but the motor still does not start. Potential causes and corrective actions to implement in this situation are

- Low armature voltage. Check the nameplate on the motor to verify that the line voltage meets the motor voltage requirements.
- Weak field. Check the armature circuit for the correct resistance; the motor may be burnt out and need replacing.
- Open circuit in the field or armature. Check the motor for open circuits. If an open circuit is detected, it may require motor replacement.
- Worn brushes. With older DC motors using brushes, verify that the brushes are in good condition.

Figure 27.2 summarizes the potential causes and corrective actions to use when the DC motor doesn't start.

27.3 Problem 3—Drive Train Problems[1]

Potential problems and corrective actions associated with the drive system include the following:

- Belt slippage in belt-driven extruder. This can be caused by factors such as the belts are worn; the screw is not turning because the barrel is insufficiently heated to melt the material being processed; a foreign object is lodged in the feed throat; cooling is not operating properly; and a barrel or barrel sections cooled too much, solidifying a crystalline or stiffening an amorphous polymer so that it can't be processed.
- Belt squeal. Squeal is caused by belts slipping on the drive pulley of a single screw extruder. Potential causes for squeal are motor and gear box shaft misalignment, improperly tensioned belts, missing belts, or belts that were replaced in a mismatched set.
- Thrust bearing failure. Over time the thrust bearing can fail, leading to the potential for a catastrophic problem such as a cracked gear box. Thrust bearings are a critical element. If an extruder is being overhauled or rebuilt, replace the thrust bearing.
- Bearing overheating. This problem is caused by improper belt tension, shaft misalignment, or insufficient or improper lubrication.

Figure 27.3 summarizes potential drive train problems with corrective actions to eliminate the problem.

27.4 Problem 4—Rupture Disk Failure in Barrel

Potential causes and corrective actions to implement for rupture disk failure are

- Sudden major contamination clogging screen pack. Foreign material in the feed stream passes through the extruder, does not melt, and becomes trapped on the screen pack, preventing molten polymer from flowing through the screens to the die. The foreign material can be a higher-temperature resin in the regrind that did not melt, reinforcing fibers or fillers that can't pass through a fine mesh filter, the wrong resin, paper, aluminum foil, and other debris.
- Die or adapter is too cold. The die or adapter contains unmelted resin at start-up that blocks the die exit. Molten resin being pumped by the extruder screw has no place to go, causing the pressure to build very rapidly.

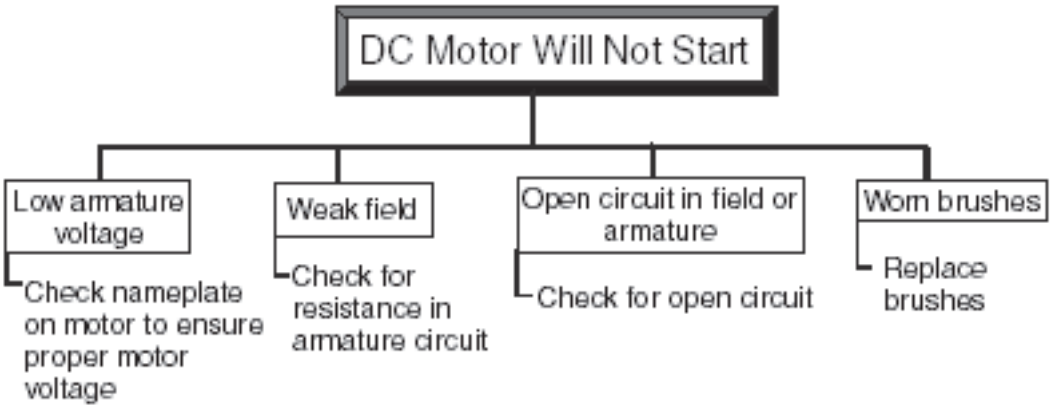

Figure 27.2. EDC motor doesn't start.

- Wrong rated rupture disk. Rupture disks are fabricated with different psi ratings; if a disk is rated below the actual processing pressure, the rupture disk will fail. An example is a rupture disk rated for 7500 psi being used in a process that normally runs between 7000 and 8500 psi. At pressures above 7500 psi, the rupture disk fails. A corrective action is to replace the rupture disk with one having a 9500–10,000 psi failure rating.
- Gear pump is not running. A gear pump, installed between the extruder and die, is not turned on simultaneously with the extruder screw. Alternatively the gear pump is turned on but is running at too low a speed to remove the polymer being pumped to it by the extruder, causing excessive pressure to quickly build between the extruder head and the gear pump.

Figure 27.4 summarizes potential causes and corrective actions to implement to eliminate rupture disk failure.

27.5 Problem 5—Barrel/Screw Wear[1]

Some potential causes and corrective actions for excessive wear on either the extruder barrel or screw are summarized below:

- Extruder alignment. The barrel and thrust bearing are not properly aligned. Larger extruders (typically greater than 2.5 inch diameter) require borescoping to ensure the barrel, feed throat, and thrust bearing are properly aligned. A straight screw should slide in and out of the barrel easily when the extruder and screw are cold. Alignment guidelines

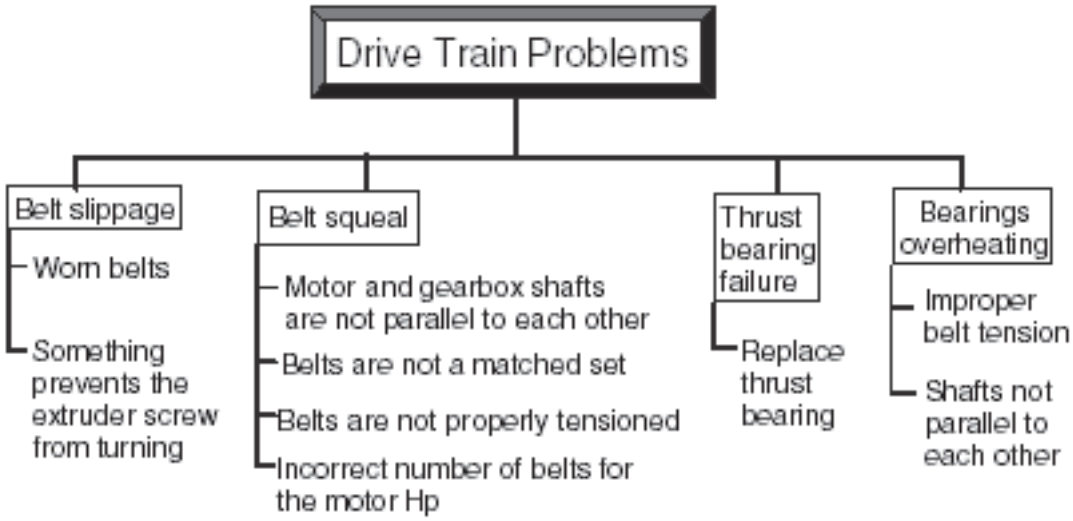

Figure 27.3. Drive train problems cause and effect.

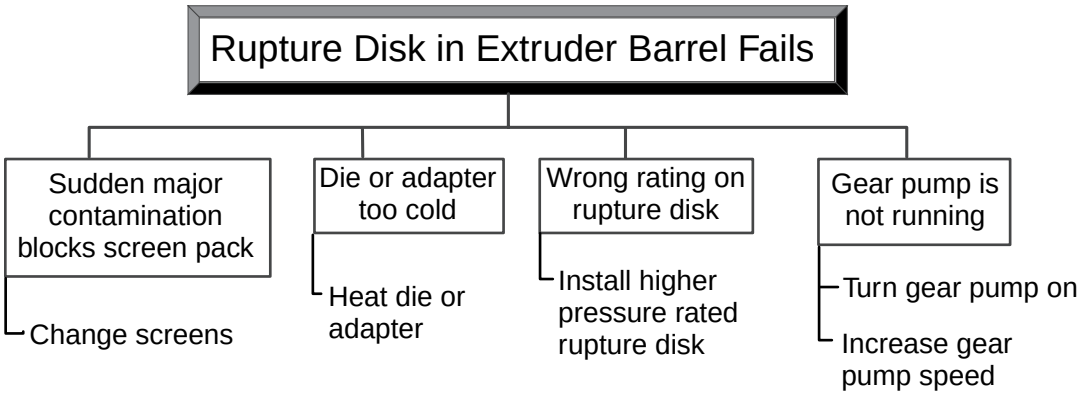

Figure 27.4. Rupture disk failure causes and corrective actions.

are typically 0.003"/foot. Equipment is available from Hamar Laser Instruments Inc.[2] to align the extruder components using a laser beam.

- Straightness of screw and barrel. Both the screw and barrel must be straight to prevent wear; if either is slightly bent, the screw drags on the barrel wall during each revolution, creating excessive wear in contact locations.
- Material being processed. Some materials are very abrasive (glass fibers, glass fillers, other fillers, and so forth), causing premature wear on flights in the feed throat or downstream wherever the abrasive material is fed into the extruder. Other materials (such as PVC and fluorocarbons) may be corrosive, causing corrosive screw wear. Selection of the optimum barrel and screw material is critical to provide long screw and barrel life while processing materials that cause either abrasive or corrosive wear.
- Incompatibility of screw and barrel material. Using materials for the barrel and screw that are incompatible can result in excess adhesive wear resulting from contact of the barrel and screw during operation.
- Improper barrel support. Improper barrel support allows the barrel to sag or be out of alignment, causing the screw to bind in the barrel. The barrel does not have to be level relative to the floor; however, the alignment of the barrel, feed throat, and thrust bearing must be in a straight line. If the middle or end barrel support is raised too high or too low, the barrel may exhibit a slight bend, causing the screw to bind during each revolution.
- Excessive loads on the discharge end of the extruder. Some dies are very large, requiring their own support. Heavy dies hung on the end of the barrel without proper support can cause the barrel to deflect, binding the screw in the barrel.
- Screw design. Using the wrong screw design to process a particular formulation may cause overheating in a specific section of the extruder, overfeeding the melting section of the screw (feeding more material than can be properly melted in the transition section of the extruder), excessive shear heating, and other problems. All of these situations can lead to excessive screw wear.

Figure 27.5 summarizes the potential causes and corrective actions to eliminate excessive screw and barrel wear.

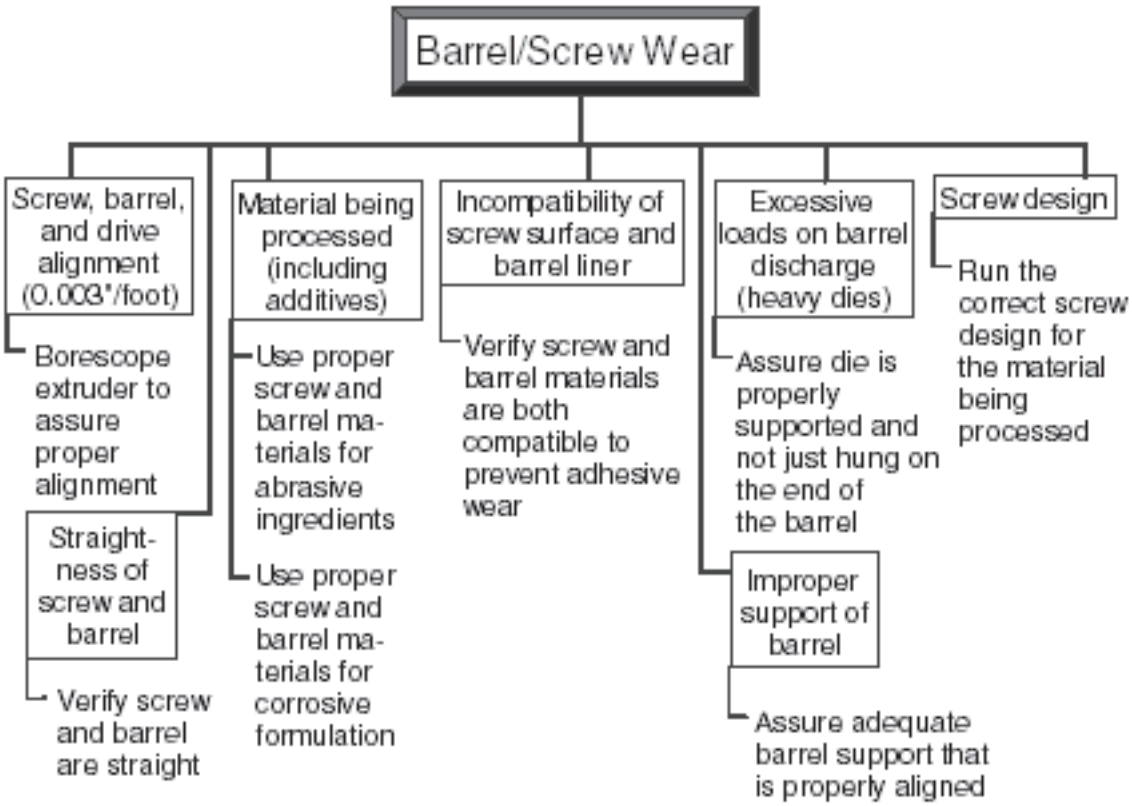

Figure 27.5. Cause and effect schematic for excessive screw and barrel wear.

27.6 Problem 6—Screw Turns and No Material Exits the Die

Based on the screw readout and the view of the shaft from the feed end of the extruder, the screw is turning at its set rpm, but no material is coming out of the die. Potential causes for this observation are given below:

- Feed throat is plugged or bridged. Material is present in the feed hopper and the screw is rotating; the possibilities are bridging in the feed throat or feed hopper or the feed throat is blocked. Push a plastic rod through the material in the feed hopper into the feed throat to determine whether the feed material is bridged or melted across the feed throat opening. Either rotating or moving the plastic rod up and down may break the bridge or blockage, allowing the material to flow freely to the screw. If bridging or blockage occurs once, it is likely to repeat itself, and a longer-term solution may be necessary for smooth operation.
- The feed hopper is empty. Verify formulation ingredients are present in the feed hopper of the extruder or the feed hopper of the feeder feeding the extruder.
- Feed throat cooling isn't working properly. The water may not be turned on to the feed throat cooling; or the inlet, outlet, or cooling channels around the feed throat may be partially plugged, resulting in insufficient cooling. The flow rate should be sufficient to cool the throat area so it feels warm to the touch but not excessively hot. Excessive cooling in hot, humid conditions can cause condensation to form in the feed throat area, introducing water into the extruder. Water flow rates around the feed throat need to be sufficient to prevent overheating of the feed throat while being low enough to prevent excessive cooling that can lead to condensation. A flow meter on the outlet side of the extruder provides information that can be related to cooling and temperature of the feed throat. Alternatively, the water temperature entering and exiting the feed throat can be monitored continuously.
- Polymer plug (melt-over) stuck to screw root. A plug of polymer is stuck to the screw root in the feed section. This is caused by excessive heat in that section, which allows the formulation to partially melt, generating high external friction between the material and the screw root and low external friction between the formulation and the barrel wall (discussed in detail in Part 1, Chapter 4). This can be caused by stopping the extruder with material left in the feed throat or by using the wrong temperature in barrel zone 1. Once a melt-over or melt plug occurs, it can be difficult to remove. Procedures to remove melt plugs were discussed in Part 1, Chapter 4.
- Zone 1 temperature is too high. If the temperature in zone 1 is too high, the melt film generated between the polymer and the barrel wall may occur too early in the extruder. The melt film can act as a lubricant, reducing the friction between the material and the barrel wall, resulting in surging, poor feed rates, or melt plugs.
- Wrong screw design. The wrong screw design for the material being processed can result in poor feeding and/or melting, limiting the amount of material exiting the die. This problem will not prevent polymer from exiting the die, but the output will be lower than anticipated.
- Broken screw. Under certain operating conditions, screws can break because of high torque or overfeeding. If the screw is broken, the shaft will continue to turn, but no material is extruded.

Figure 27.6 summarizes the potential causes and corrective actions to implement in the event the screw is rotating with either no material or limited material exiting the die.

27.7 Problem 7—Extruder Shuts Itself Off

The extruder is running and suddenly stops; potential causes and corrective actions to correct the problem are listed below:

- Backpressure is too high. The pressure transducer at the extruder discharge prior to the breaker plate is equipped with a high-limit alarm that sends a signal back to the extruder to stop the extruder motor if the predefined limit is reached. This is a safety device preventing pressure in the barrel from becoming high enough to rupture the barrel or blow the die off the extruder. Excessive pressure occurs from die blockage, contaminated screen packs, die or adapter temperatures that are too low, or a melt temperature that is too low.
- Gear box temperature is too high. Insufficient cooling is being provided to the gear box. Once the oil temperature reaches the setpoint limit, an alarm signal automatically shuts down the extruder drive.
- Motor or drive problems. The electrical circuits reach an overload condition, knocking out the circuit breakers. Determine the cause before restarting the extruder.

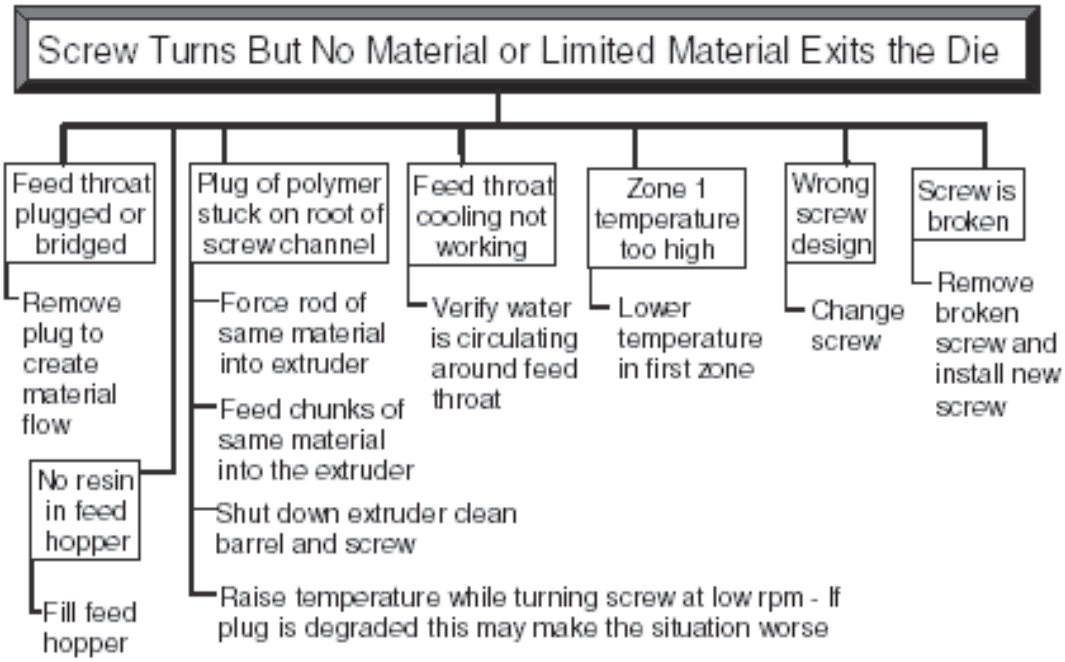

Figure 27.6. Corrective actions when the screw rotates and no material exits the die.

- Screw is locked up. Foreign material or contamination is fed into the extruder. A bolt or other metal part from a blender, regrind hopper, and so forth is inadvertently fed with the formulation to the extruder, jamming between the screw and the barrel wall, resulting in a motor overload condition. The wrong regrind (higher melt temperature material) is added to the blend and does not melt with the operating conditions being used for the current formulation, causing the screw to seize.
- Extruder drive is pulling too many amps. The maximum motor torque or load is reached, shutting down the extruder. This frequently occurs in a twin screw extruder by overfeeding the extruder at a given screw speed. Low screw speed in twin screw extruders does not provide enough power or torque to convey and melt the formulation. Excessive water cooling on the barrel caused by a defective solenoid that does not close properly can significantly increase the motor load until the extruder shuts down. High backpressure can increase the motor load by forcing material farther back into the screw channel.

Figure 27.7 summarizes the potential causes while showing some corrective actions to take in the event the extruder suddenly shuts itself off.

27.8 Problem 8—Melt Flows Out the Barrel Vent

Using a two-stage single screw extruder with the vent open or a vented barrel on a twin screw extruder, molten polymer occasionally flows out the barrel vents due to the wrong processing conditions being used or the screw design being incorrect. With a single screw extruder, the molten polymer flowing out of the vent or vacuum port is caused by the second stage not pumping as much material out the discharge as the first stage is delivering to the second stage. This is caused by the wrong temperature profile or excessive backpressure from the screen pack or die. This assumes that the screw is a two-stage screw with the decompression zone in the right location along the barrel.

In twin screw extrusion, polymer can flow out an atmospheric vent or vacuum vent if the screw design is improper, the die pressure is too high, or the vent is in the wrong location. Prior to the vent, the screw should have a melt seal or pressure generation device to build pressure, followed by high-pitch elements directly under the vent that reduce the melt pressure in a decompression zone. A vacuum or vent port, installed in a barrel section, that is too close to a die that creates a high die or backpressure has the potential for polymer to back up at the die and flow out the vent or vacuum port. The other potential problem is creating a high-pressure area in the screw design directly after the vacuum port, forcing molten resin to be pushed out the vent by the screw rotation.

Figure 27.8 summarizes the potential causes for vacuum or vent flow with potential corrective actions.

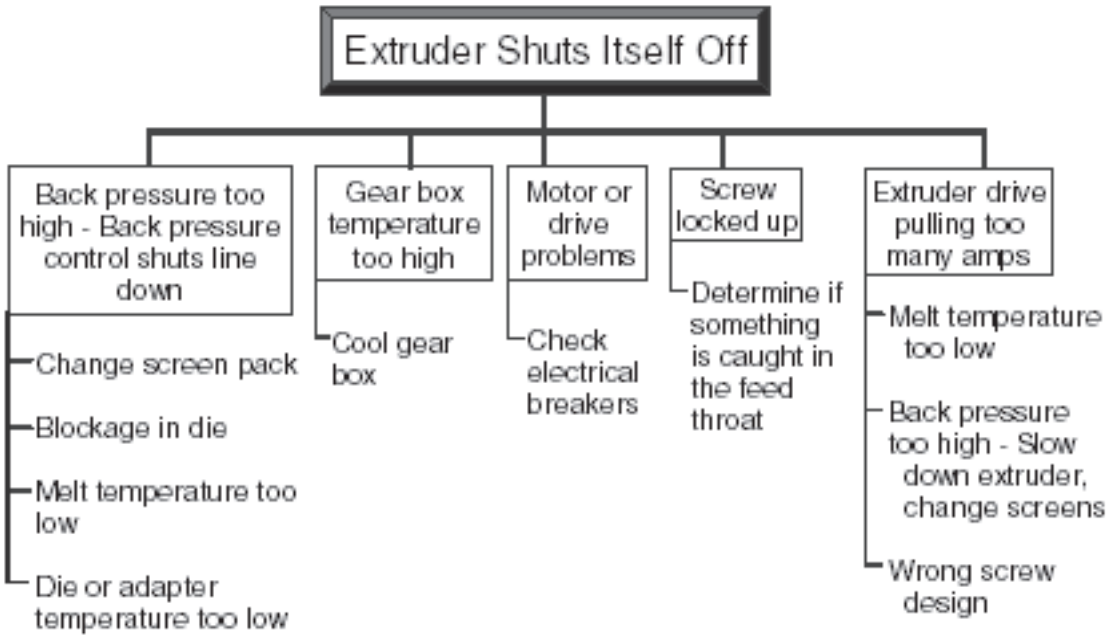

Figure 27.7. Corrective actions if extruder shuts itself down.

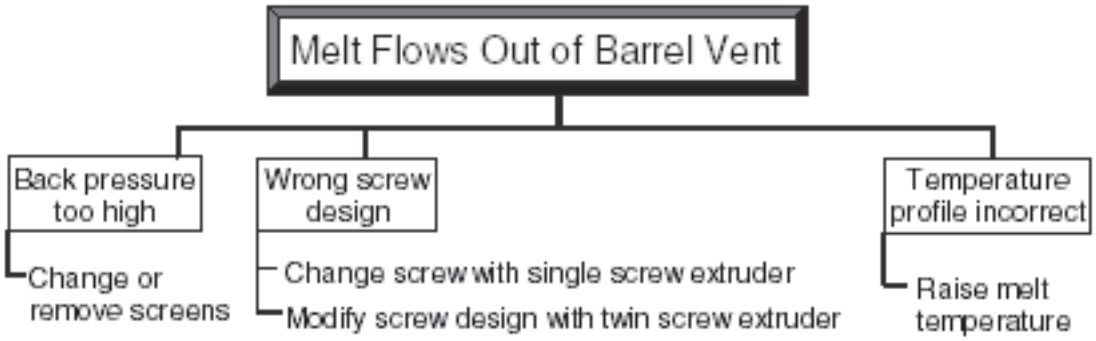

Figure 27.8. When polymer flows out the vent or vacuum port.

27.9 Problem 9—Leaking Polymer at Breaker Plate or Die

This problem creates a mess at the extruder head as molten polymer continually leaks very slowly from the extruder during the entire production run. In addition to creating a safety hazard from molten polymer dripping on wires or generating a tripping hazard, it wastes raw materials. Molten polymer leakage around the die or breaker plate area is caused by the following:

- Breaker plate is not properly seated. The breaker plate or sealing ring is not properly seated, allowing material to flow around the sealing surface.
- Breaker plate sealing surface is damaged. If hammers or pipe wrenches are used to remove the breaker plate from the extruder or screen changer, the sealing surface becomes damaged over time, containing dings, nicks, or scratches, preventing the surface from making a tight seal between the barrel and the die or adapter. Sometimes breaker plates are machined to remove surface defects, and after machining the breaker plate thickness may be reduced sufficiently to prevent a tight seal.
- Plugged screens. As screens become contaminated and clogged during a run, head pressure increases, generating more opportunity for leakage around the breaker plate.

- Die and adapter bolts are not properly tightened. Any time the die or adapter has been removed from the extruder or screens are replaced in a breaker plate, the bolts attaching the die and adapter must be uniformly torqued to provide a good seal. It is essential to retorque the bolts after all the equipment has been brought to operating temperatures before starting production.
- Breaker plate slide is not completely engaged. With an automatic screen changer, after installing new screens and a clean breaker plate, the slide must be completely across the extruder opening and properly seated to form a good seal.

Figure 27.9 summarizes potential causes and possible corrective actions to use in the event molten polymer is leaking between the extruder and die.

27.10 Problem 10—Extruder Throughput Rate Is Lower than Anticipated

From a mechanical view, there are numerous reasons why the extruder throughput might be lower than anticipated. A theoretical output calculation can be made based on equations presented in Part 1, Chapter 4. In addition, previous experiences with a given product using a specific extrusion setup may indicate a particular throughput rate is anticipated, and in the current production run the throughput is lower than expected. Possible causes and corrective actions for reduced throughput rates are listed below:

- Screw or feed throat is partially blocked. There may be a partial bridge in the feed hopper or in the feed throat due to overheating, material compaction, large regrind chunks, and so forth. In addition, the screw may have a partial melt-over or plug in the feed zone.
- Improper setup. Check the SOPs to verify that the extruder is properly set up with the correct polymer and melt flow, correct regrind level, proper die and barrel temperature profile, correct screw or screw design in twin screw extrusion, and the correct die and die assembly.
- Starve-fed process. If the extruder is starve fed, verify the feed rates are properly set on all feeders.
- Screen pack. The screen pack is assembled with finer mesh screens than were used previously or the screen pack is partially blocked. Both situations generate higher backpressure, resulting in lower throughput rates.
- Screw is worn. The last time this product was run was six months ago, and since that time there has been considerable screw wear, causing leakage over the flights in the metering section or flight deterioration in the feed section, resulting in less solid material being conveyed into the extruder.
- Die adjustments. The die has adjustable lips that need to be opened more to reduce the die pressure and allow a higher flow rate through the die.
- Screw is dirty. The screw has polymer build-up in one or more zones or a mixing section is dirty, preventing polymer flow. Screw contamination may be from a previous high-temperature run with the same screw, which was not properly purged. Lower melt temperatures leave higher-temperature material stuck to the screw root or in the mixing sections, partially blocking the screw flights.

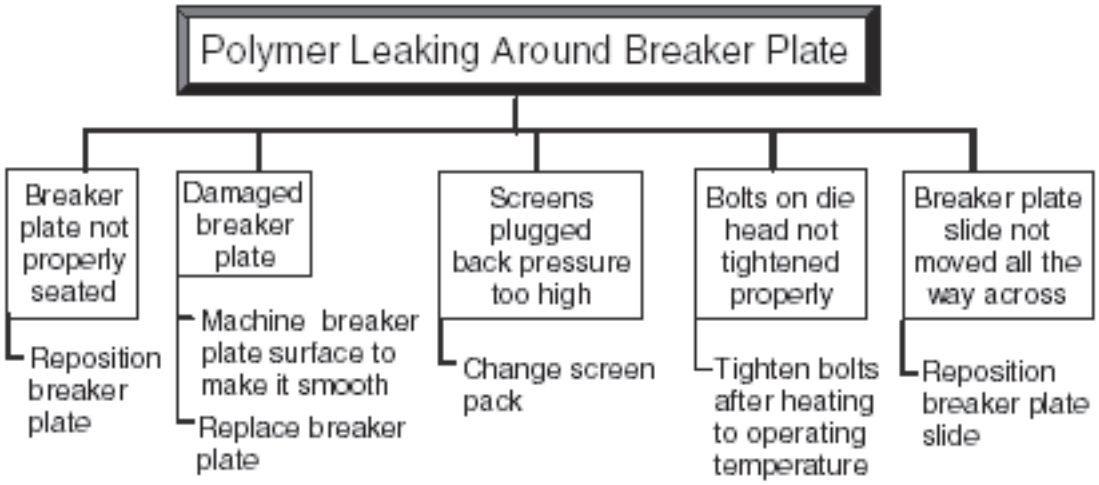

Figure 27.9. Corrective actions for polymer leakage around the breaker plate or die.

- Drive belts are slipping. The drive belts are worn out and slipping. This is normally accompanied by a squeal or some other noise.

Figure 27.10 summarizes potential causes and corrective actions to implement if the throughput rate is lower than anticipated.

27.11 Problem 11—Temperature Overrides Setpoint

Potential reasons with corrective actions for the actual barrel temperature exceeding the setpoint temperature are outlined below:

- Barrel zones are set too low. When a barrel zone temperature is too low, shear heating, produced by screw rotation, generates excessive heat, increasing the temperature in that zone. If a particular zone is overheating, increase the barrel temperature not only in the zone that is overheating, but also in the prior zones.
- Feed section is overfeeding. If the feed zone channel is too deep, more material may be fed into the extruder than can be melted in the transition zone. Excessive shear heat is generated at the entrance to the metering zone as unmelted solid is being forced into the metering section.
- Screw transition section is too short. The transition section in a single screw extruder is too short to be able to melt the plastic being fed by the feed section at a given screw speed. Similar to the situation above, this forces unmelted pellets into the metering zone, generating high shear heating. The barrel temperatures are high in the transition section due to the solid material compression. The melt density is higher than the bulk density, requiring less screw channel volume to handle the same mass of material in the melt compared to the solid. This increases the residence time for the melt, which may overheat from shear heating. It is also possible that there is too much solid material to melt compared to the energy put into the system through barrel heat and shear heat. This has the same effect as overfeeding solids or too short a transition section and is a result of poor process condition selection. Overheating will happen if there is enough torque and power to turn the extruder screw. The motor and screw supply the heat in the form of shear heat. The solution to this problem is to use a screw with a longer transition zone to provide sufficient time to melt the solid polymer being fed by the feed section.

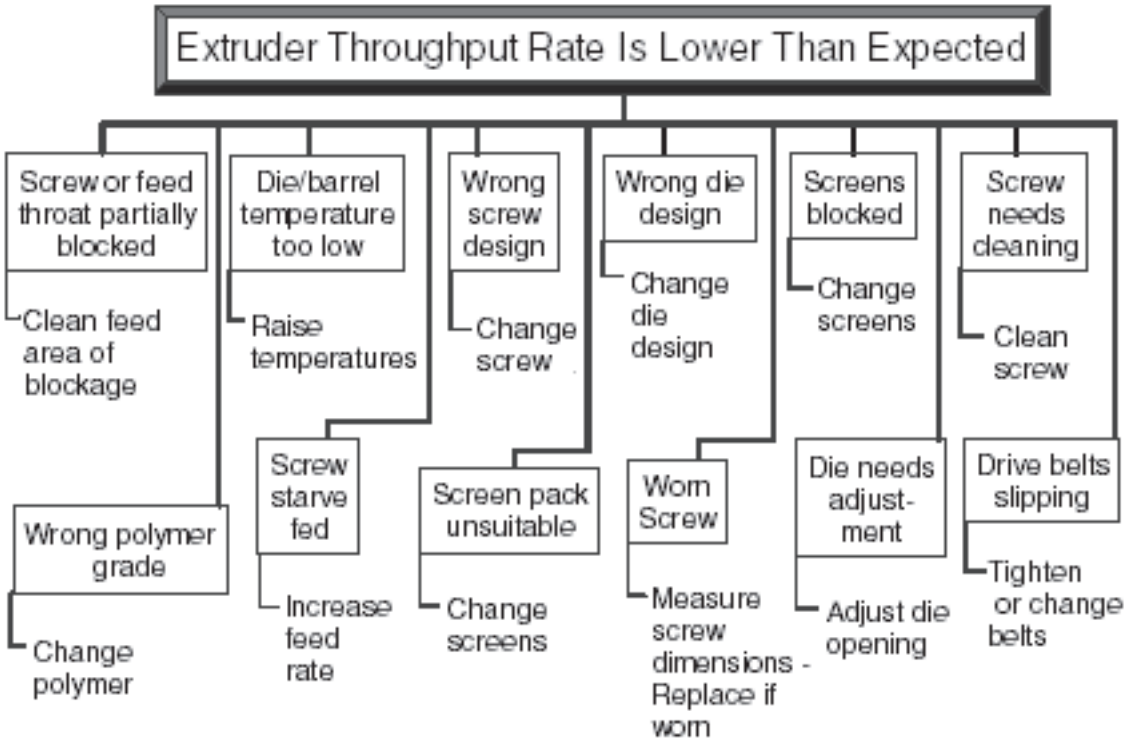

Figure 27.10. Corrective actions for extruder throughputs lower than anticipated.

- Screw is worn. Molten polymer is passing over the screw flights, minimizing the shear heat that can be generated by the screw in a given extruder location. This causes excessive heating by shear in other barrel locations by increasing residence time.
- Mixing section is too tight. The gap between the mixer barrier flights and the barrel wall in the mixing section is too small. The molten resin is forced through small gaps, creating excessive shear heating, leading to significant barrel temperature override.
- Excessive work in twin screw designs. Screw designs with a high number of kneading blocks, rearward conveying elements, rearward conveying kneading blocks, large neutral kneading blocks, blister rings, and other high-shear elements generate significant shear heat where they are located in the extruder. Using a rearward conveying element at the end of the melting section ensures that all the resin is melted before it leaves the melting section. However, this rearward conveying element generates high pressure and shear heat. Other areas of the screw with rearward conveying elements grouped with neutral, forward, and rearward conveying kneading blocks can generate excessive shear heat, leading to high barrel temperatures.

Figure 27.11 shows potential causes and corrective actions to use when the actual barrel temperatures are overriding the setpoint temperatures.

27.12 Problem 12—Extruder Surging Caused by the Equipment

Surging is the variation in throughput resulting from a cyclic behavior, with excessive extruder output followed by low extruder output, then excessive output, and so forth. Surging can be caused by factors related to both the product and the equipment; however, in this particular problem analysis, only those factors directly related to the equipment are addressed. Potential causes and corrective actions for surging due to the equipment are discussed below:

- Barrel temperatures are not controlling properly. If the barrel temperatures are going through wide swings, the resin viscosity is changing. Changes in viscosity are accompanied by changes in output. Normally, controllers control barrel temperatures within ±2˚F (±1˚C). If excessive shear heat is being generated in a particular zone, the temperature controller may not be able to control that zone.
- Poor die design. The die design may not have smooth transitions providing uniform flow. This can be compounded with nonuniform temperature across the die, resulting in some die areas being hotter than others. Temperature is directly related to the polymer viscosity, which is related to die flow properties. Die temperature fluctuations can affect the flow uniformity in the die.
- Worn screw or barrel. A worn screw or barrel does not generate uniform shear heating, feed characteristics, or conveying in the metering section. These can affect the polymer heating and flow characteristics.

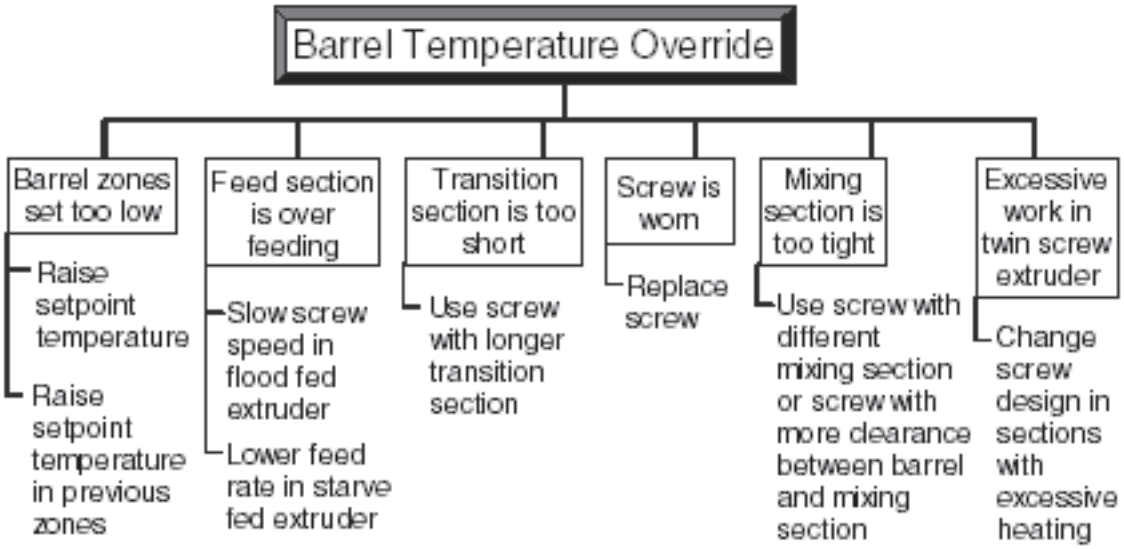

Figure 27.11. Corrective actions for potential causes for temperature override.

- Die/barrel temperature fluctuations. Not only is it necessary for the barrel and die heaters to control at their setpoints, but also the fluctuations between the maximum and minimum temperature values have to be small.
- Varying motor speed. If the motor speed is changing, the screw speed is pulsating, with more and less material being pumped out the extruder as the screw speed changes.

Figure 27.12 summarizes the potential causes and corrective actions to eliminate equipment-related surging.

27.13 Problem 13—Poor or Insufficient Polymer Mixing

Insufficient polymer mixing is characterized by improper dispersion or distribution of pigments or colorants, additives such as flame retardants or impact modifiers, stabilizers, reinforcements, fillers, and blends of two or more resins. The potential causes and corrective actions for improper mixing include

- Insufficient melt temperature. If the barrel and die temperatures are too low with insufficient shear heating, the polymer viscosity may be higher than normal, making it more difficult to disperse or distribute additives uniformly into the polymer melt.
- Insufficient die pressure or backpressure. Backpressure only marginally improves the mixing in an extruder. Material reaching the die or breaker plate with insufficient mixing normally requires other mechanisms to improve the mixing, such as a static mixer between the extruder and the die. Increasing the backpressure assists mixing if the dispersion and distribution are almost complete prior to reaching the screen pack and breaker plate. Die pressure is not easy to increase without drastically changing some other parameter in the process. Closing the die lip opening increases die pressure but may also create molecular orientation and other product effects that are not advantageous.
- Improper screw design. Whether the process is running in a twin screw or single screw extruder, changing screw designs to incorporate more mixing does increase the dispersion and distribution of the various ingredients in the melt. The proper mixing element (distributive or dispersive) needs to be added to the screw for the materials being processed. As an example, reinforcements need distributive mixing, while pigments and impact modifiers require dispersive mixing.
- Mixing head is plugged. The mixing head or mixing section may be dirty or contaminated with foreign material, preventing adequate mixing. Pulling the screw and cleaning the mixing head may improve the quality of mixing.

Figure 27.13 summarizes the potential causes and corrective actions to implement in the event of poor polymer mixing.

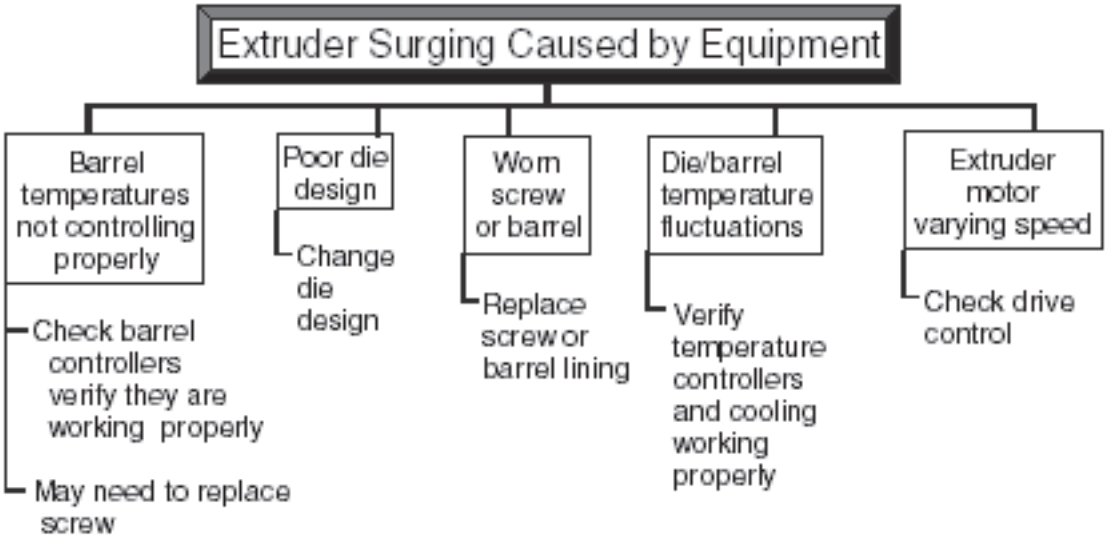

Figure 27.12. Potential corrective actions for surging related to the equipment.

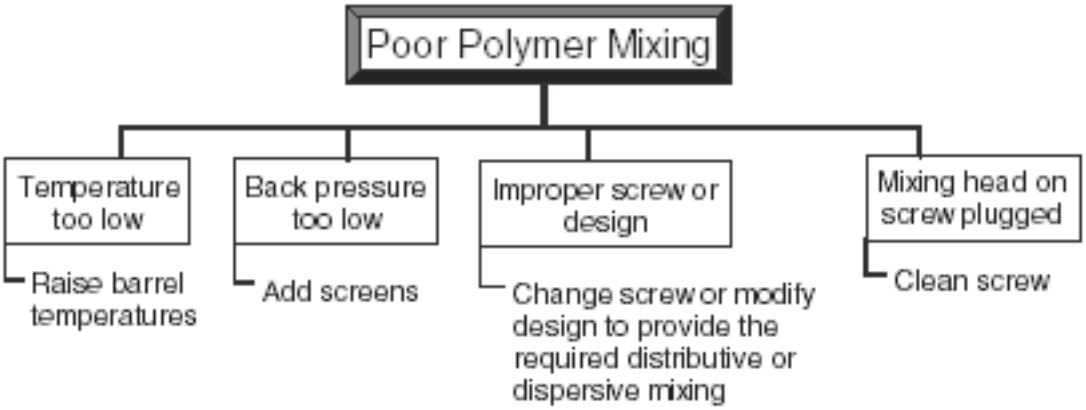

Figure 27.13. Corrective action steps for insufficient mixing in the extruder.

27.14 Problem 14—Throughput Variation Over Time

Potential causes and corrective actions for variations in throughput rate (normally a reduction in throughput) over time are summarized below:

- Screen pack contamination. As the screen pack becomes clogged or fouled with contamination from the formulation or foreign material, the head or backpressure increases, reducing the material flow out the die. The flow is composed of the drag flow and the pressure flow. As the pressure flow increases, the total throughput rate decreases.
- Extruder is coming to equilibrium. Depending on the extruder size, it may take anywhere from 20 minutes to three hours for the temperature and the entire system to come to equilibrium. Initially the output may be different from that attained when the system reaches equilibrium.
- Reduced output over time. Feed throat or die blockage will reduce the throughput over time.

Figure 27.14 summarizes potential causes and corrective actions to eliminate throughput variations over time.

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1. Womer, T. W., "Basics of Extrusion Troubleshooting," PowerPoint Presentation, Spirex Corp., Youngstown, OH, 1995.
2. Hamar Laser Instruments, Inc., 5 Ye Olde Road, Danbury, CT.

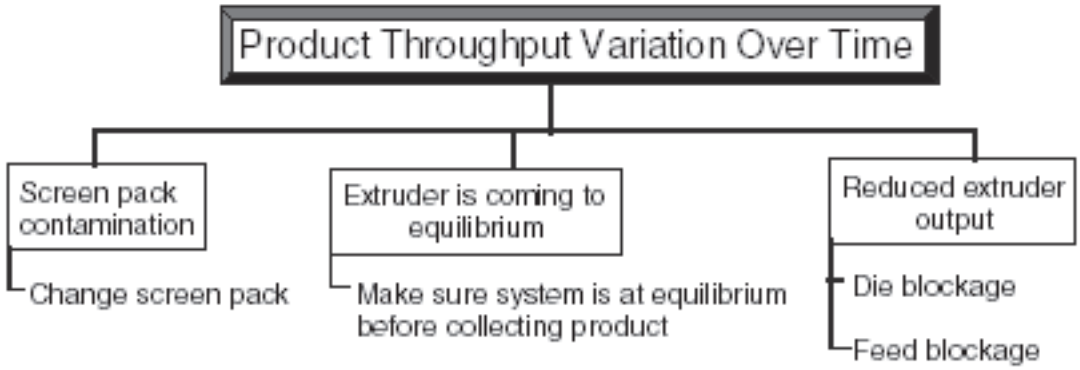

Figure 27.14. Corrective actions to use when variations in throughput occur.

Review Questions

1. Give five reasons why an extruder might shut itself off. What should be done in each situation?
2. Give six reasons why the extruder throughput might be less than anticipated. What is the corrective action to take in each situation to improve the output efficiency?
3. What drive train problems might be encountered in extrusion?
4. The screw is rotating in the extruder, but no material is exiting from the die. What are some potential causes?
5. Melt is coming out the vacuum port. What are some potential causes, and what corrective action is needed to fix the situation?
6. The polymer mixing is not sufficient to yield an acceptable quality product. What might be done to improve the mixing efficiency?
7. You are ready to start the extruder but can't get the screw to turn. What are some potential causes and corrective actions?
8. The product is changing dimensions over time. What are some potential causes that can be related to the equipment?
9. On start-up, the rupture disk blows. What is causing the problem?
10. Give seven potential causes of excessive barrel or screw wear.
11. On start-up, polymer is leaking around the breaker plate and die. What might be the problem?
12. The temperature on the barrel is exceeding the setpoints. Give possible actions to correct the problem.
13. Surging is seen in the extruder output. What are some equipment aspects to check to ensure that the surging is not equipment-related?

28 Troubleshooting Extrusion Product Problems

To solve or troubleshoot product-related problems in the extruder, it is important to understand polymer behavior in the extruder feeding, melting, mixing, and conveying sections. These areas were covered in Parts 1 and 2 for single and twin screw extrusion, respectively. Understanding polymer structure and polymer rheology is also important. Basic information on polymers was covered in Part 3. Bringing together extrusion mechanics and polymer behavior facilitates solving product problems. As in Chapter 27, various potential problems associated with products are identified with potential corrective actions to eliminate the problem.

28.1 Problem 1—Product Surging

Surging is identified as an extruder flow rate instability or inconsistent output rate. Surging is easily identified through die pressure, motor load, and product dimension variations. Throughput rate increases and/or decreases can be caused by factors such as using the wrong screw design, feed problems, or running the extruder too fast. In Chapter 27 in Part 4, problem 12 covered equipment-related surging problems, such as the barrel temperature not controlling properly, poor die design, worn screw or barrel, varying motor speed, and/or temperature fluctuations. An improper screw design can cause product problems that lead to surging; these include

- Poor feeding due to slip/stick (on the barrel wall) in the feed zone. This may be corrected by increasing screw cooling (assuming it is available) or decreasing barrel zone 1 temperature.
- Melting point instability, caused by the melting or softening area moving in the transition section. This is normally corrected by changing the transition zone barrel temperatures.
- Metering section channel depth is too deep, causing the screw to run partially filled. This is corrected by using finer mesh screen packs or a valve to increase the head pressure.

Replace the screw with a design suited to the polymer and application.

Ninety percent of surging problems occur in the feed and transition sections, yet most attempts to correct surging problems are usually concentrated on the metering. Potential causes for feed section surging are poor temperature control, short feed or transition section, deep feed section, bad feed mixture in the feed hopper, or nonuniformly preheated or dried materials. These factors lead to transition zone overfeeding, which prevents proper melting or softening before the polymer enters the metering zone. At the other extreme, if the feed zone flight channel is too shallow, the solids that can be fed are limited. Occasionally, a screw becomes fouled and then releases materials, causing flow fluctuations.

Other potential causes for surging and corrective actions associated with the product and screw are discussed below.

- Barrel temperature too low. If the feed or transition zone temperature is too low, complete polymer melting may not occur in the transition section. Unmelted pellets entering the metering zone are difficult to melt and may clog or interfere with the screw mixing section.
- Improper screw design. The screw runs satisfactorily at medium screw speed, properly feeding and melting the product in the transition zone. However, at high screw speed, melting is not completed in the transition zone, causing melt instabilities similar to those described above. If this problem cannot be overcome by modifying the temperature profile, it may be necessary to run at lower screw speeds and throughput rates to stabilize the process. Screws with stable operating conditions at low and medium speeds may become unstable at high throughput rates as the melting capacity is exceeded. To eliminate instability at high throughput rates, it may be necessary to change the screw design, incorporate barriers in the transition zone, or use screw designs to mix melt and solid polymer, such as the ET® or Wave® screw.
- Insufficient resin supply. In starve-fed extruders, insufficient resin feed relative to the screw speed can cause the discharge end to be incompletely filled. Surging can be caused by the oscillating metering section fill. Three potential solutions to this problem are to lower the screw speed, increase the feed rate, or increase the backpressure with finer screens or a valve.
- Partially bridged feed throat. If the feed throat is partially blocked, feed uniformity can vary. As the feed throat becomes blocked, the feed rate decreases, possibly even changing from flood fed to starve fed. As the blockage dissipates, the feed goes back to a flood-fed situation, filling the screw. Repeated cycles from no feed throat bridging or blockage to partial blockage are constantly changing the feed rate. A solution might be to redesign the feed hopper and/or feed throat.
- Irregular feed. Irregular or inconsistent feed to the extruder leads to inconsistent output or surging.

Slippage on the barrel wall resulting from reduced friction between the material and the barrel wall in the feed zone or extruder zone 1 can be caused by slippery additives such as lubricants, silicone, low-molecular-weight waxes, and so forth. Sticky additives in the formulation can lead to irregular feeding, caused by adhesion in the feed hopper or the feed throat leading into the feed zone. As material sticks to the walls, it blocks the flow into the screw, reducing the throughput. Throughput increases as large clumps that were sticking to the hopper or feed throat walls break free, increasing the polymer flow into the feed throat. A partially plugged feed zone screw channel reduces the material that can flow into the extruder. As the plug becomes dislodged, maximum flow is again obtained. Repeated material sticking and release decrease and increase the material entering the extruder. High filler loadings added with the resin in a single screw feed hopper can cause poor polymer feed.

Figure 28.1 summarizes some potential causes and corrective actions to eliminate extruder surging. Surging is more apt to be present in a single screw than in a twin screw extruder because the feed and melting mechanisms are different. Slippery additives and liquids added at the feed throat can cause problems in both extruders.

28.2 Problem 2—Variation in the Product Over Time

Variations in the product over time are a general problem associated with product changes that are not caused by other problems discussed in this chapter. This problem was partially addressed in Chapter 27 under problem 14, "Throughput Variation Over Time." Potential causes and corrective actions for variation in the product over time are

- Energy input to the plastic during processing has changed over time, or the energy required to melt the polymer has changed over time. This results in the polymer melting area moving in the extruder or the polymer viscosity in the mixing, conveying, or pumping section changing. Ultimately, the polymer does not process the same as it did previously; the product attributes, such as size, color, or surface aesthetics, have changed, and the product is not the same as when production started. Potential causes for energy variation are cold air flowing over the extruder (open overhead door in the wintertime), colder raw material feedstock, extruder temperature controllers not cycling properly, or wrong melt temperatures or pressures.
- Molecular weight has changed. The raw material supply has changed, with the polymer molecular

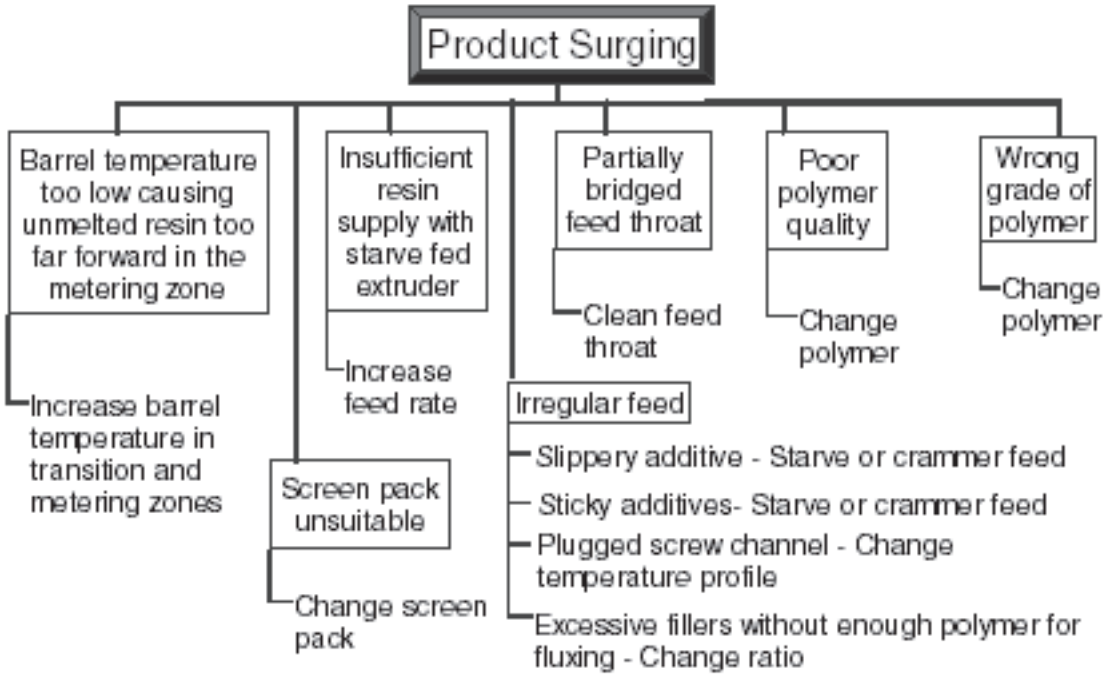

Figure 28.1. Potential causes and corrective actions for product surging.

weight being either higher or lower than earlier in the production run. This can be caused by a regrind that is not the same composition or molecular weight as that used previously.

Figure 28.2 summarizes variations in product over time with potential corrective actions to return the product to its original state.

28.3 Problem 3—Random Product Variation

Random product variation refers to the product varying inconsistently from time to time without any observable changes in the process. Whether the changes are attributed to appearance, physical properties, or compositions, this is a difficult problem to identify the cause and effect. Potential causes with corrective actions are listed below.

- Feed problems caused by sticky material adhering to the feed hopper or feed throat. Consider adding talc to lubricate the particle surfaces.
- Raw material variation due to regrind, a raw material lot change, or a supplier change. Check raw material properties.
- Cyclic factors such as screw speed, pressure fluctuations, and melt temperature variation can match a high or low simultaneously. It can be extremely difficult to identify the problem and correct it. Numerous processing changes may be necessary to correct the problem. The best approach is to monitor the process very closely and eliminate any cyclic variations. Those occurring over short times are the most difficult to identify. Long-time variation associated with temperature controllers is easier to identify.

Figure 28.3 summarizes potential causes and corrective actions to implement to eliminate random product variations.

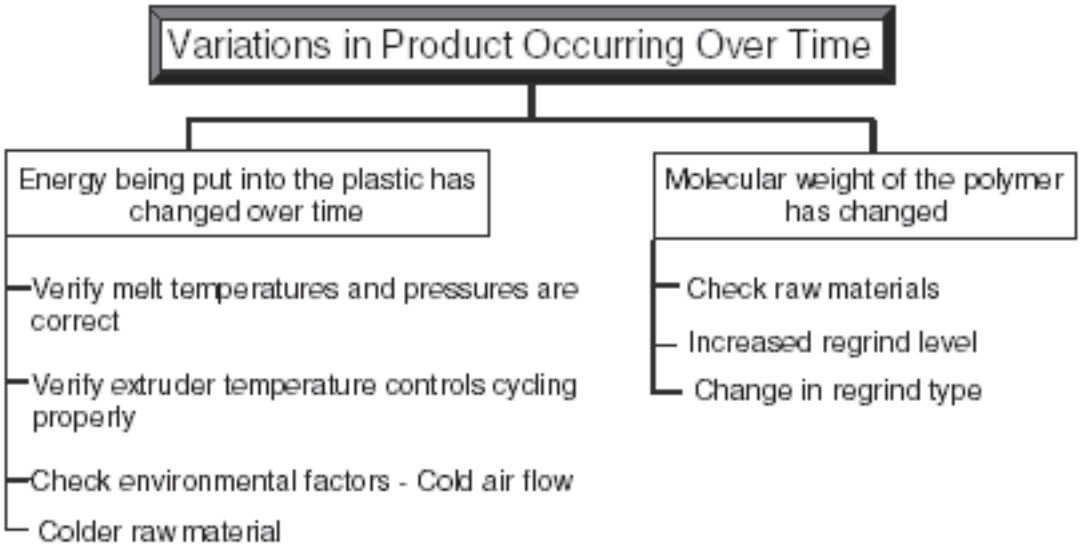

Figure 28.2. Corrective action for variation in the product over time.

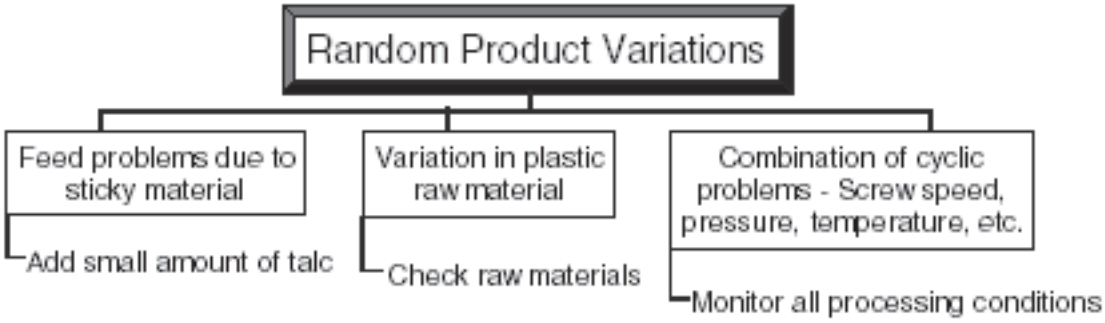

Figure 28.3. Potential causes and corrective actions for random product variations.

28.4 Problem 4—Streaks in the Product Caused by Poor Mixing

Streaks in the product that run in the machine direction or the direction the product is being pulled can be caused by poor color concentrate or other pigment mixing. Other potential causes with corrective actions for color streaks occurring in the product are given below.

- Extrudate nonuniformity caused by poor mixing. A short-term solution is to increase the head pressure or raise the barrel temperatures. Obviously, increasing barrel temperatures opposes the increased head pressure, as lower melt viscosity reduces the die and head pressures. A longer-term single screw solution is to change to a screw with a better mixing section, and for a twin screw incorporate more kneading blocks. Single screw mixing is improved by using a barrier flight in the metering section, a longer mixing head, or a static mixer in the transition pipe between the extruder and the die.

- Streaks that appear after production has been consistently producing good product. These may be caused by a change in raw material supply, adding more or less regrind, or a change in the pigment or colorant package. Other possible causes are viscosity changed due to a heater band burning out, cold air blowing on the extruder, some other factor that changes the energy input or removal from the system, a damaged screw, or deposit build-up in the die resulting in resin degradation.

- Unsuitable die. The die may not be suitable for the particular process being run. The die may not be assembled properly or may need adjustment.

Figure 28.4 summarizes the possible causes and corrective actions for streaks in the extrudate.

28.5 Problem 5—Variation in Product Strength

There is a change in the product strength as measured by impact, tensile, or flexural properties, compared to a previous production run. Potential causes and corrective actions to implement to eliminate changes in product strength are

- There is a molecular orientation change, resulting in a variation in the machine direction versus the transverse direction properties. Depending on the orientation change, the properties in the machine direction may increase or decrease, with the corresponding properties in the transverse direction varying in the opposite direction. In other words, if the tensile properties in the machine direction increase, they will decrease in the transverse direction, and vice versa. A potential cause for the change in molecular orientation is the draw ratio has changed between the extruder and the puller. This only occurs with a die with an adjustable gap, where the die lip setting has been changed. As the

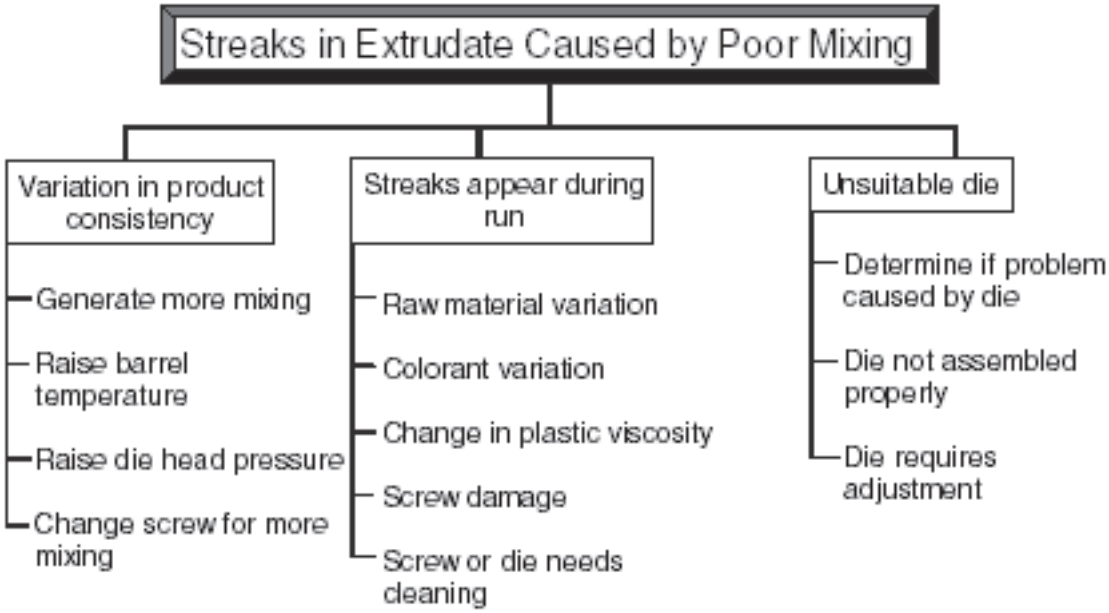

Figure 28.4. Possible causes and corrective actions for streaks in the extrudate.

die lips are opened, a higher draw ratio is required to maintain the final product dimensions. In processes with nonadjustable die lip openings, the product dimensions are changed as the puller speed or the extruder screw speed is changed. With an adjustable die opening, the product thickness exiting the die can be increased or decreased through die adjustments. As the die is opened, the puller speed is adjusted to maintain the product dimensions, causing the product draw ratio to be increased and changing the product orientation. Increasing the molecular orientation, as discussed in Part 3, increases the machine direction strength while decreasing the transverse direction strength. A second factor that can change orientation is puller take-up speed. Without a change in the die opening dimensions, this normally is accompanied by a change in product dimensions. A third potential factor influencing molecular orientation is polymer flow in the die. Small die openings with high throughputs lead to polymer molecules orientating in flow direction. Assuming this is the orientation source, it can be reduced by opening up the die. However, if the orientation is coming from the die land area, increasing these dimensions increases the extrudate size, requiring a higher draw ratio, inducing higher molecular orientation to obtain the correct product dimensions.

- Resin degradation in the extruder or die leads to product strength decreases in all directions. Resin degradation is normally caused by hot barrel or die temperatures, excessive screw speed in a starve-fed extruder, the wrong screw geometry for the polymer being processed, or a long residence time at high temperature. Polymer degradation can be accompanied by a color shift. Extreme processing temperatures from either barrel temperature or shear heating can consume the heat stabilization package, resulting in lower product strength over time in the application life cycle. Any system is particularly vulnerable in high-temperature applications where strength is required over a given time frame. Physical property deterioration normally occurs once the thermal stabilizer is consumed. In polyolefins, thermal stabilizers are free radical scavengers. Heat and ultraviolet light generate free radical formation. Once the stabilizer is consumed and free radicals form, the degradation reaction occurs much more rapidly.

- In crystalline polymers, strength can be related to the crystal size and percent crystallinity. Rapid product quenching reduces both the size and percent crystallinity. Generally, smaller crystal growth is accompanied by tougher parts. Another approach to obtain smaller crystal size is to add a nucleating agent to the polymer.

Figure 28.5 summarizes the potential causes and effects that lead to a change in product strength in extrusion.

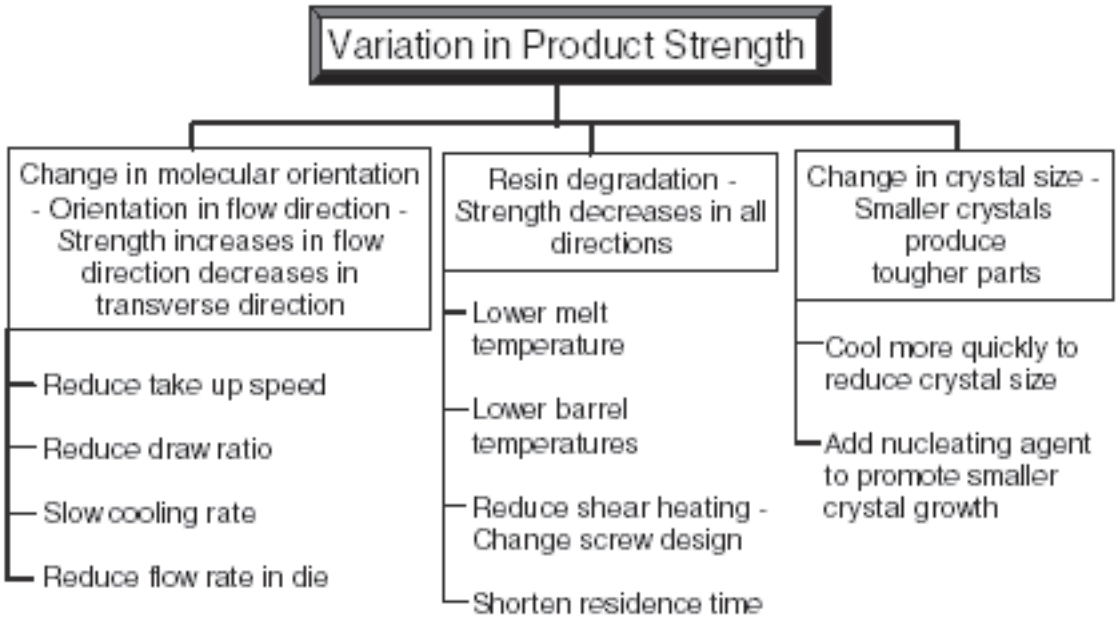

Figure 28.5. Potential causes and corrective actions for changes in product strength.

28.6 Problem 6—Product Degradation

There are many potential reasons for product degradation to occur in the extruder. They show up as discoloration, streaks, black specks, and/or property loss. Potential causes and corrective actions for product degradation are summarized below.

- Polymer melt temperature is too high, causing polymer degradation. Lower the melt temperature by changing the barrel or die temperatures, reducing the screw speed, or decreasing shear heating. To reduce shear heating, it may be necessary to change the screw design.
- Residence time in the extruder, transition pipes, adapters, or die is too long, allowing the polymer to degrade over time. Any dead spaces within the die or transition zones allow polymer to stagnate, resulting in potential degradation as the polymer sits at high temperature. All dies and transition pieces need to be streamlined to prevent any dead spaces where the polymer can build up and stagnate.
- Two or more noncompatible polymers or ingredients can react at processing temperatures, resulting in degradation. Verify all formulation additives are stable at the processing or melt temperatures being used. The wrong flame retardant additive may degrade at high processing temperatures. Additives designed for low-temperature polymers may not have the thermal stability required to be processed at high temperatures, leading to additive and/or resin degradation. Combining certain polymers (acetal and nylon, acetal and ionomer) leads to degradation reactions in the extruder.
- Black specks can be caused by a dirty screw, die, adapters, or transition pipes. Degraded polymer build-up in the extruder can break off or be scoured off by shear during processing, causing black speck contamination. Any contaminated area in the extruder has to be properly purged or, in extreme situations, disassembled and cleaned to eliminate black specks.
- The wrong polymer grade for the application or the wrong processing temperatures can result in polymer degradation. Regrind may degrade sooner than virgin materials due to its previous heat history and thermal stabilizer loss. Some polymer grades have more thermal stabilizer for use in more severe thermal environments. Adding thermal stabilizer to the polymer formulation can extend regrind use and other polymers in the formulation at high processing temperatures.

Figure 28.6 summarizes potential causes and corrective actions to use in the event polymer degradation is occurring during extrusion.

28.7 Problem 7—Dull Streaks

Dull streaks are defined as areas where the surface gloss or finish is dull. It can be a streak in one area where the surface appears to have a matte finish or over a large

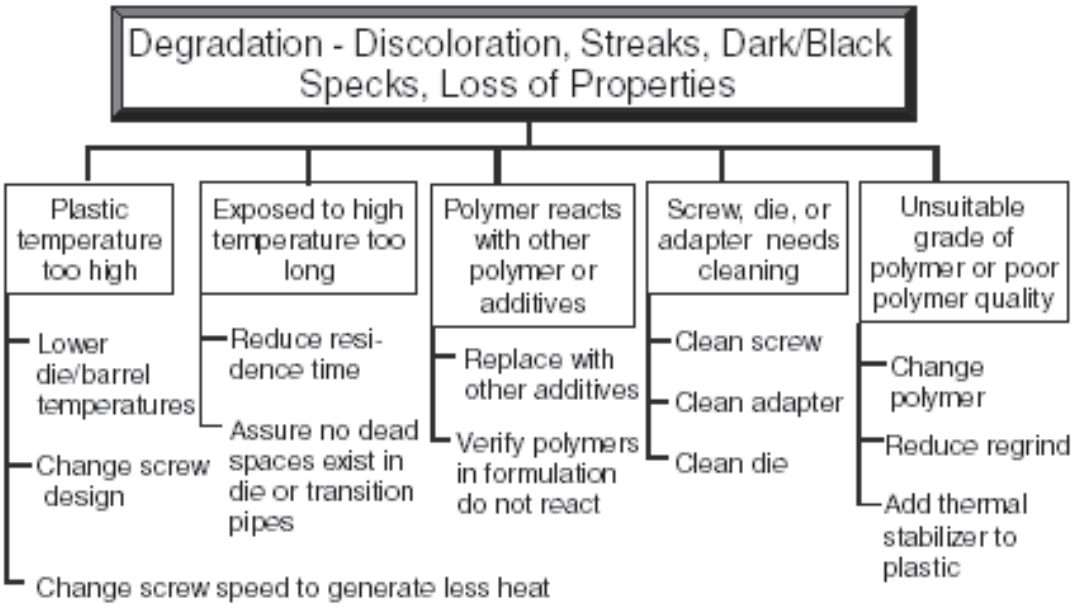

Figure 28.6. Potential causes and corrective actions for polymer degradation.

section of the extrudate. Potential causes with corrective actions for dull streaks are

- Polymer melt temperature is not uniform. If this is the problem, it is very difficult to measure the product melt temperature at the dull streak versus the other locations. A melt thermocouple in the die or a hand-held pyrometer outside the die measures an average polymer temperature around the probe. If the dull streak is a surface phenomenon, measuring the surface temperature versus the polymer directly below the streak is nearly impossible. An infrared (IR) pyrometer can measure the dull area surface temperature if it is fairly wide; however, if the streak is narrow, a very small focal spot size is required with the IR pyrometer, and radiation from adjacent melt may still interfere. An adjustable depth melt thermocouple placed in the transition pipe or entrance to the die can be used to measure polymer melt uniformity in the melt transfer pipe before entering the die. However, this raises the question, How is the melt temperature at different depths in the transfer pipe related to the dull streak? A temperature differential is present across the polymer flow channel, with the center usually higher temperature than at the wall. Adjustable depth melt thermocouples do measure temperature differences across the flow path. However, as the material exits the die, melt temperature differences are extremely difficult to measure or even locate. A cold spot in the die, resulting from a defective heater or thermocouple, can cause a dull streak in the product. Generally, if dull streaks occur and all barrel and die heaters are working properly, increase the barrel temperatures. A second potential solution is to increase the head pressure to induce more backward flow, resulting in more mixing.
- Nonuniform cooling can lead to dull streaks. This can be caused by either hot or cold spots in the die, preventing the extrudate from leaving the die with uniform melt temperature. If the extrudate temperature is not uniform, it is almost impossible to alter the cooling conditions to provide uniform cooling. Conversely, if the product does leave the die with a uniform melt temperature, it is essential to ensure uniform cooling. If water cooling is used, the product needs to be completely submersed in the uniform temperature coolant. With a water bath, the water enters the bath at the point farthest from the extruder and exits near the extruder discharge. The water temperature should increase uniformly from the entrance, with no localized hot or cold spots. The product must be uniformly submersed in the bath. If the product sticks out while the bottom is submersed, cooling around the product is not uniform. Cooling uniformity related to dull streaks is more critical in contact cooling devices such as chilled rolls, pipe corrugators, and so forth.
- Variations in the raw materials can cause dull streaks. Check the formulation for contamination that may have occurred during mixing or transfer, and verify that the melt flow index has not changed.
- When producing sheet and film, dull streaks can be caused by improper die lip adjustment.
- Build-up of additives (plate-out) on coating surfaces may also affect surface gloss. causing streaking.

Figure 28.7 summarizes potential causes with corrective actions to eliminate dull streaks.

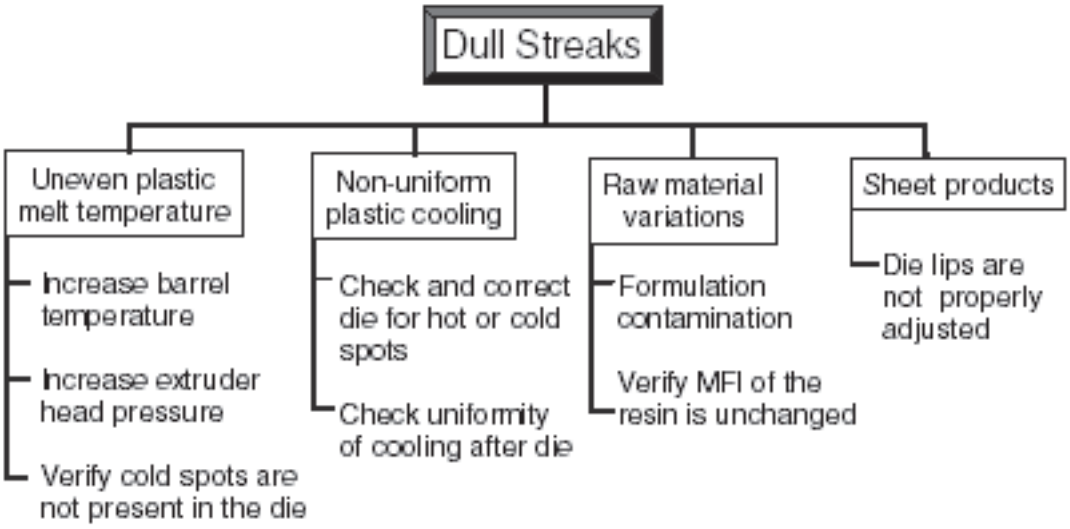

Figure 28.7. Potential causes and corrective actions for dull streaks.

28.8 Problem 8—Machine Direction Die Lines

Die lines are straight lines in the machine direction that remain in the same location on the extrudate during operation. Potential causes and corrective actions to eliminate die lines are

- Surface defects in the die land area can be caused by scratches, burrs, or foreign material in the die. The defect causes an indentation in the extrudate that remains in the same location over time. Any scratches in the die lip area have to be mechanically removed to eliminate the problem. If the problem is due to a foreign particle on the die lips (carbon deposit or other material located in the die, in front of the die lips, or on the die lips) or in the die (blown screen pack has left screen contamination in the die), remove the lips and clean the die to eliminate the problem.
- A machine direction streak may be caused outside the die by dragging the product across something in the sizing operation or on the take-up unit. This is usually fairly easy to detect and correct. If the scratch is not present as the extrudate exits the die but is seen partway down the line in the cooling section, the problem is not in the die. Determine where or what is scraping or damaging the part surface after it leaves the die and eliminate the problem.

Figure 28.8 summarizes the potential causes and corrective actions to employ in the event die lines or other lines appear in the extrusion direction and remain in a fixed location on the extruded part.

28.9 Problem 9—Color Contamination

Inadequately cleaning a previous formulation causes color contamination. Potential areas for color contamination are

- Extruder and die are not properly cleaned from a previous run. If possible, schedule production runs to go from lighter to darker colors. Assuming a light color product is followed by a dark color product, the darker color in most situations will mask any lighter color contamination. In the event a brown, black, blue, or other dark color is followed by a light color (white, cream, pale colors, or natural), any material left in the extruder or die takes a long time to remove. Depending on the extruder size, it is appropriate to clean the extruder and then run in order natural, light colors, medium colors, dark colors, and finally black. After the black run, the extruder can be shut down, the screw pulled, and the die cleaned. Any material stagnating or hanging up in the die, transfer pipes, screen, and breaker plate area or around the feed throat can lead to color streaks or color contamination in subsequent runs until it is completely purged.
- The feed system must be completely cleaned prior to any color changes, or color contamination will result. The feed system includes blender, hold hoppers, transfer lines, dryers, and extruder hopper.
- Improper premixing, feed ratios, or color ingredients added to the extruder can result in the wrong color. Formulations that are premixed and transferred to the extruder must be mixed in the correct ratio. With starve-fed extruders, the color formulation must be properly mixed and the feeder set at the correct throughput rate to generate the correct color. Color concentrate added at the wrong rate will result in the wrong color.

Figure 28.9 summarizes the potential causes and corrective actions to be taken in different situations for color contamination. Products color matched on one extruder with a given screw design, throughput rate, screw speed, and temperature settings may not give the same color on another extruder with a different screw design. Color

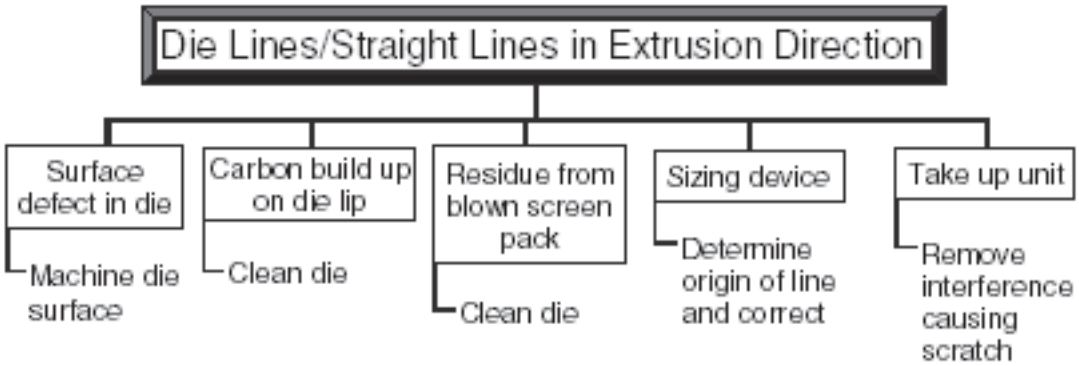

Figure 28.8. Potential causes and corrective actions for die lines.

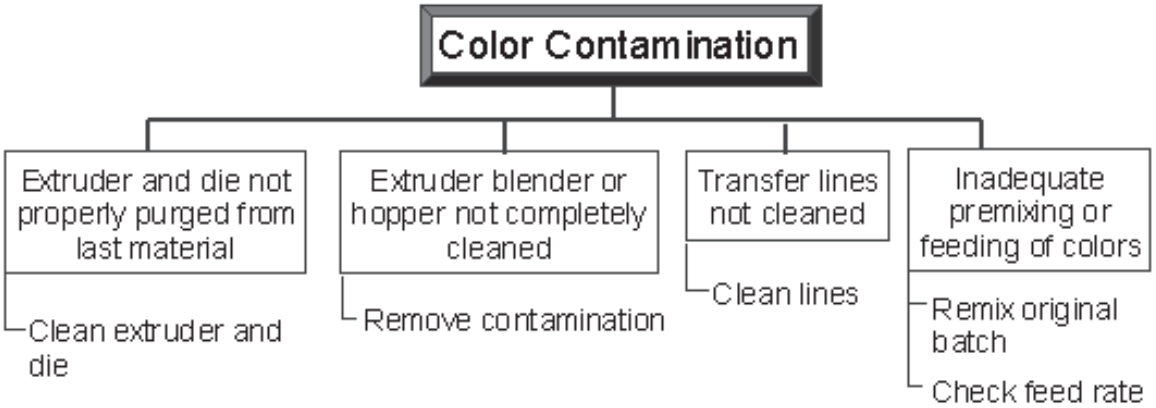

Figure 28.9. Potential causes and corrective actions for color contamination.

development can be dependent on extruder shear. Two extruders with different temperature profiles and shear rates may not produce exactly the same color product, even though the formulation is identical.

28.10 Problem 10—Holes in Extrudate

Holes in the extrudate can be caused by gases that are generated in the extrusion operation through ingredient degradation, steam, or trapped air that is not removed prior to the die. Some holes are vacuum voids that form in thick sections as the extrudate cools and shrinks. A vacuum void is centered in the extrudate, while gas generated in the extruder may blow a hole completely through the final part as the molten polymer exits the die. The trapped gas that is under pressure in the extruder expands rapidly once the pressure is removed. This rapid expansion may actually blow a hole in the extruded part. Potential causes and corrective actions to take in the event holes are present in the extrudate are

- Air trapped in the barrel must be removed prior to the die or holes will be blown in the final product. Normally air is forced out the feed throat in flood-fed single screw extruders. In twin screw extruders, the air is forced out the feed throat, a side feed extruder, atmospheric vent, or vacuum vent. The last screw flights in either single or twin screw extrusion have to be completely filled with polymer to prevent air from reaching the die. With a flood-fed single screw extruder, the air is forced back through the feed throat by compression and melting in the transition section. If a single screw extruder is starve fed, the screw speed has to be slow enough to prevent air from being pumped past the transition zone. Any air trapped in a single screw metering section will be forced out the die and blow holes in the final part. In a twin screw extruder, low-pitch conveying elements are used near the discharge to ensure the flights are full and force any air that may be present in the extruder out the vent prior to the die. Air exiting the die on a flood-fed single screw extruder may be an indication that all the material is not melted in the transition section, allowing unmelted pellets and air to penetrate into the metering zone. To correct this problem, raise the transition zone barrel temperature to ensure all melting occurs in the transition zone. Finer screen packs increase the backpressure and help fill the screw and prevent air from exiting the die. Another approach with a single screw extruder is to use a two-stage vented extruder to provide air with an exit path other than through the die. Trapped air can be more problematic when processing powder compared to pellets. Powder bulk density is lower and introduces more air into the extruder. A higher-compression ratio screw is used with powders, ensuring that the transition flights are full of melt, preventing air from entering the metering section.

- High melt temperatures can lead to resin or additive degradation that generates gas at the extruder discharge. With a vented extruder, gas can be removed from the extruder prior to the die. However, better approaches are to lower the processing temperatures, change screws to reduce shear heating, or decrease the throughput to minimize the melt temperature. While a vented extruder can remove generated gases, the resin is still experiencing some degradation, possibly reducing the final product properties.

- Resin contamination can cause a chemical reaction in the extruder that may generate gases. This is easily corrected by properly purging the extruder with a nonreactive resin and avoiding or removing any contamination from the product prior to feeding it to the extruder.

- Moisture introduced into the extruder with resin or additives is converted to steam. If the moisture is not removed prior to the die through vacuum or atmospheric venting, holes in the product can result. With predried resins still containing some moisture or moisture that has condensed on the pellet surface, rather than a hole, the moisture may cause surface splay on the extruded part. Splay indicates moisture or gases are present, but their concentration is lower than the critical amount required to blow bubbles in the product.

- Holes centered in thick product sections are due to polymer shrinkage on cooling. Thick extrudate sections quenched rapidly after leaving the die form a surface skin. As the material continues to cool, the skin continues to thicken. At a certain point, the molten material inside the thick area cannot pull the solidified outer surface in on the part, but it continues to cool and shrink. Since the outside is solidified, the walls cannot be pulled toward the center. Shrinkage causes the molten material on the inside to shrink toward the outside surface, forming a vacuum void in the center. Potential solutions are to cool the extrudate more slowly, allowing the outside to shrink toward the center. If cooling more slowly is not a successful solution, it may be necessary to redesign the part, minimizing the thick sections. Shrinkage is the same reason sink appears on the surface opposite bosses and ribs in profile extrusion. Figure 28.10 shows a shrinkage void in an extruded strand. Many strand compounded pellets have a hole in the middle. This is caused by shrinkage as the strands are rapidly quenched in a water bath. As the inside cools, the molten polymer cannot pull the thick skin toward itself and generates a vacuum void down in the pellet center.

Figure 28.11 summarizes potential causes and corrective actions to employ when holes appear in the extrudate.

28.11 Problem 11—Gels

Gels, which are often referred to as "fisheyes," are high-molecular-weight polymer particles that do not melt during processing. Gels may soften and elongate as they pass through a screen pack designed to filter or trap contaminants. However, they never really melt. After passing through the screens, they snap back or return to their original shape, forming small, hard particles in the melt. Others define gels as any contamination (dirt, other polymer, cross-linked polymer, and so forth) that passes through the die. In the final part, gels are a speck (may be clear) that forms a defect. In polyethylene, gels are occa-

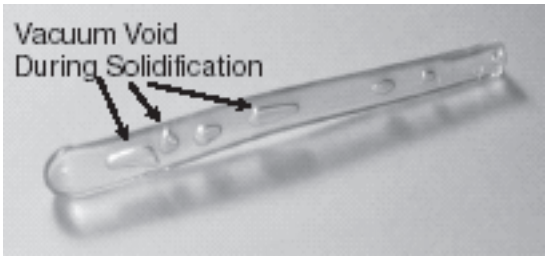

Figure 28.10. Formation of vacuum.

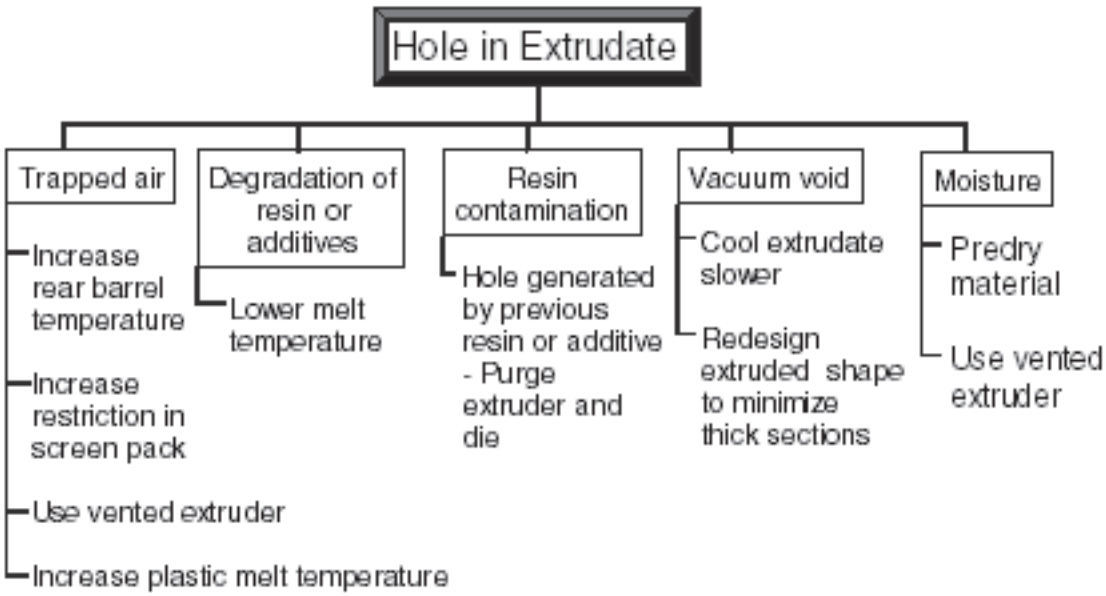

Figure 28.11. Potential causes and corrective actions for formation of holes in the extrudate.

sionally formed in the process as material hangs up or stagnates in the system. Polyethylene held at processing temperatures over time can cross-link and form high-molecular-weight particles that do not melt in the process. If gels are formed in the die, the die needs to be streamlined to prevent stagnation. To remove gels that are already present in the polymer, raise the melt temperature and/or use a finer screen pack to filter them out.

Figure 28.12 summarizes potential causes and corrective actions to use when gels attributable to high-molecular-weight material are present.

28.12 Problem 12—Gauge Variations

Gauge variations are defined as thickness nonuniformities in the final product. While this appears to be specifically aimed at sheet or film, the potential causes and corrective actions hold true for thickness variations in any products, such as profile, pipe, tubing, and wire coating. Changes in product dimensions are caused by nonuniform melt temperature, resulting in polymer viscosity variations that correlate with differences in extrudate flow out the die. Potential causes of nonuniform melt temperatures are

- Poor melt mixing, due to mismatch in screw design for the polymer being extruded, combined with low die pressure that reduces polymer backflow. Depending on the polymer and processing conditions, other situations can lead to insufficient mixing in the melt.
- An unevenly clogged screen pack allows more polymer flow through one part of the screen pack relative to another. This can lead to increased melt temperature in the area that is partially clogged versus the section where material is flowing more freely.
- Burnt-out heaters in the screen changer, adapter, or die lead to cold spots where the heaters are burned out and hot spots where the heaters are working, causing a differential in polymer melt temperature between the two locations. Loosely fitting heaters on either the adapter or die can create hot spots where the heaters are in contact with the die or adapter and cold spots where they are not.

Figure 28.13 summarizes the potential causes for melt temperature variation that can lead to product gauge variations and the corrective action steps for each situation.

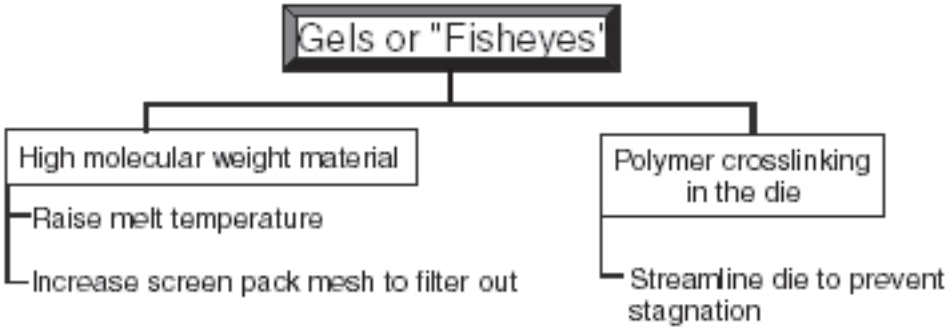

Figure 28.12. Potential causes and corrective actions for gels.

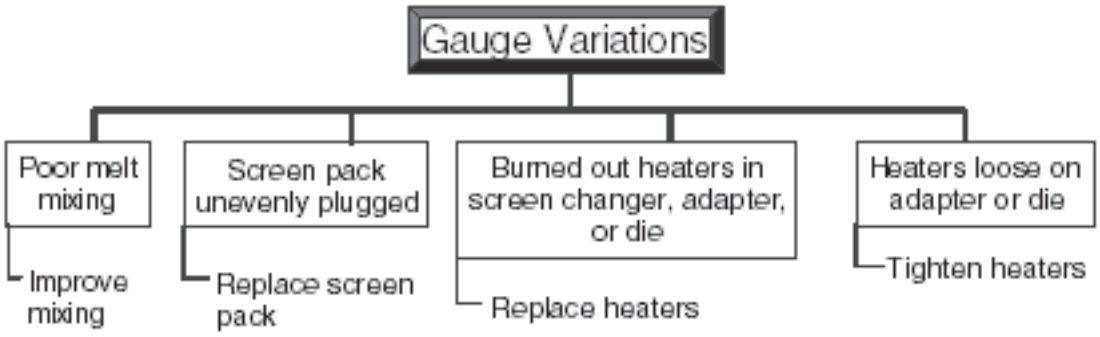

Figure 28.13. Potential corrective actions to eliminate gauge variations.

28.13 Problem 13—Weld Lines

Weld or knit lines are caused by poor polymer reentanglement after flowing around an obstacle in the die. Molten polymer molecules require the proper environment for molecular chain reentanglement at the flow front or the weld line. Poor entanglement results in lower strength at the weld line versus the polymer matrix. Dies containing mandrels normally have a spider to center the mandrel in the middle of the die. Polymer flows around each pin or spider arm holding the mandrel in place and then recombines just prior to exiting the die. Contamination or a foreign object such as a wire from the screen pack trapped in the die prior to or in the die lip area can cause a weld line to form in any extruded shape. Figure 28.14 shows how a weld line forms in a die. Possible corrective actions to improve weld line strength are

- Raise the barrel temperature, the extruder pressure, or the pressure in the die to encourage more molecular entanglement prior to the die exit. These three processing factors are all designed to raise the melt temperature, lowering the polymer viscosity and encouraging more molecular motion and entanglement when the flow fronts recombine.
- Increase the time the flow fronts are together in the die after the flow is recombined to increase the molecular entanglement at the flow front. This is accomplished by increasing the die length.
- Increase flow restriction in the die after the flow fronts recombine to increase the pressure on the melt at the flow front, driving the molecules together.
- Change the die design to incorporate mixing after the flow fronts recombine to ensure the weld line area is homogeneous.
- Change the polymer to a lower-molecular-weight polymer, reducing the viscosity while increasing the molecular motion in the die.
- Remove any foreign material or contaminants in the die that may be creating weld lines in the die lip area.

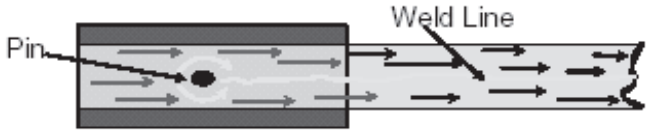

Figure 28.14. Weld or knit line formation.

Figure 28.15 summarizes potential corrective actions to pursue when weld line problems are present. The best practice is to design the die to minimize weld lines.

28.14 Problem 14—Weak Film or Coating

Potential causes for a weak film or a weak coating when using extrudate to coat another object such as paper, aluminum foil, another polymer, and so forth are given below.

- High polymer melt temperatures may produce insufficient melt strength, causing a film to break or to be hard to handle as it exits the die. High melt temperature can also cause resin degradation, lowering the molecular weight and making the resin coating brittle.

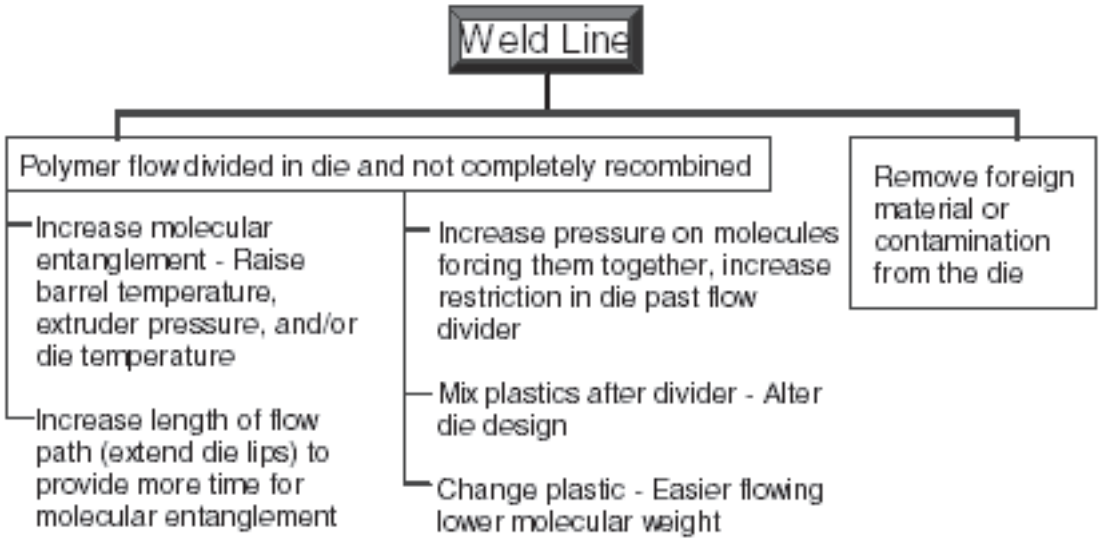

Figure 28.15. Potential causes for weld lines and corrective actions.

- Using the wrong resin for the application makes processing very difficult. Verify that the proper melt flow resin is being used to provide sufficient melt strength and end-use properties.
- If the polymer is mixed with other formulation ingredients in the extruder, improper additive mixing or dispersion provides sites for crack propagation and imperfections. Photo micrographs will verify the additive dispersion in the polymer. If mixing is insufficient, both the melt strength and final product strength will be inferior. It may be necessary to change the screw design to improve the mixing to obtain proper additive dispersion in the final product.

Figure 28.16 summarizes potential causes and corrective actions to pursue in the event the product melt strength or final strength is deficient in film or coating products.

28.15 Problem 15—Warpage

Warpage is caused by differential or nonuniform shrinkage in the extrudate cross sections. This is often referred to as internal stresses in the part resulting from the molecular orientation, molecular relaxation outside the die, or molecular orientation induced in the final product by drawing. Stresses are caused by some part sections shrinking differently from other sections. Factors that affect nonuniform shrinkage and internal stress development are

- Molecular orientation
- Reinforcing fibers and fillers orientation
- Molecular packing uniformity
- Solidification uniformity
- Asymmetric crystallinity from nonuniform cooling

Extrusion tends to align the polymer molecules more in the machine direction than in the transverse direction. Polymer molecules aligned in the machine direction have higher shrinkage than in the transverse direction. This shrinkage difference can cause the part to twist or deform. Assume one side is being pulled out the die faster than the other side; this leads to different molecular orientation across the part and different shrinkage, leading to part warpage. The slower side is stretched more, resulting in higher molecular orientation. Polymer molecules are oriented by stretching or pulling the polymers in a given direction. The stretch in any direction has to be equivalent to prevent warpage.

If the extrudate is filled with fiber reinforcements or fillers with an aspect ratio (filler particle length to diameter ratio, L/D) greater than 1, the reinforcement causes the resin matrix shrinkage to be anisotropic unless the filler or reinforcement is completely randomly dispersed. The pulling or drawing action in the die and outside the die aligns more fibers or fillers in the machine direction than in the transverse direction, resulting in a shrinkage difference between the two directions with a high probability of warpage. Since mineral fillers and reinforcing fibers have lower shrinkage than plastics (by at least one order of magnitude), shrinkage is now greater in the transverse direction compared to the machine direction. The shrinkage difference between the machine and transverse directions leads to warpage in the extruded part.

If the packing forces or pressure in the die are not uniform, some molecules may be more tightly packed in some areas compared to other areas. Areas with higher packing, or more molecules packed into a given volume, shrink less than areas with less molecules in the same volume. Highly packed areas have reduced shrinkage compared to sections with less packing pressure. This is particularly important in profile extrusion. Assume a window profile is being produced with numerous ribs and structure. The die and part design require uniform cross sectional areas with the correct taper and lead into the polymer flow channels in the die. This provides uniform flow pressure in the die to minimize shrinkage differences that can create warpage in the final part.

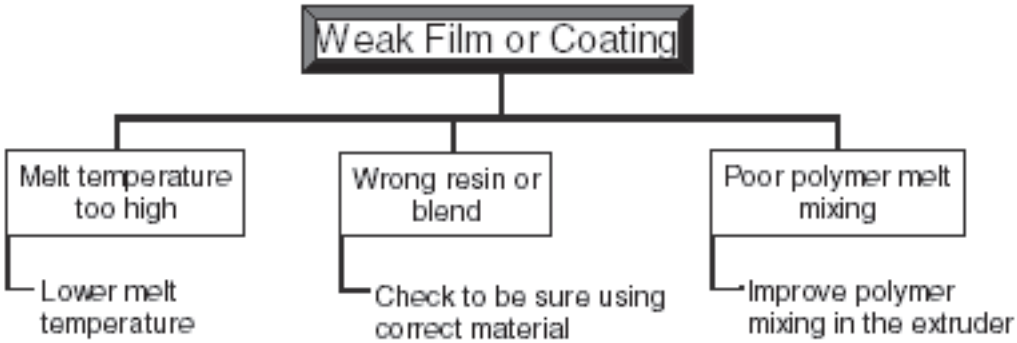

Figure 28.16. Potential causes and corrective action for weak films or coating.

All part surfaces have to solidify at the same time and rate to prevent warpage. A part solidifying more rapidly on the bottom than the top because the bottom is being cooled by water and the top by air will curl or warp toward the hotter side. Cooling rates have to be similar around the entire part to ensure the skin thickness develops uniformly in all areas. Figure 28.17 shows the effect of cooling more rapidly on one half of the part compared to the other. Assume the bottom is submersed in water and the top half is exposed to air. As the part exits the extruder and before it comes in contact with the water cooling, the part is expected to shrink the same throughout. As the part is submersed in the water, a skin is formed on the bottom side, and shrinkage starts to occur. The top half exposed to the air is still molten, and the polymer chains move as the part shrinks on the bottom, and shrinkage on the top has not yet been initiated. After some specific time, a skin forms on the top and shrinkage is initiated. Since both the top and bottom are expected to shrink a specific amount X, the bottom actually shrinks less compared to the top because the bottom was shrinking while the top was still molten. Assume the bottom shrunk by Y amount before the top solidified and started to shrink; the bottom actually shrinks X – Y while the top is shrinking X. Consequently the part has effectively more shrinkage on the top, causing the part to cup or warp toward the hot side. Figure 28.18 simulates warpage in the part just discussed.

If the melt temperature on one side is hotter coming out of the die than the other side, warpage can result even with uniform cooling, as the hotter side may solidify later in the process compared to the side with the cooler melt temperature. Uniform melt temperature criteria are important to minimize warpage.

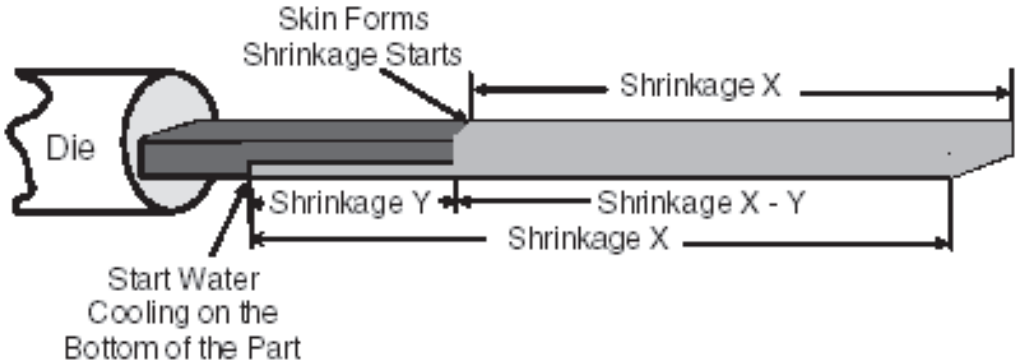

Figure 28.17. Effect of differential cooling on shrinkage and warpage.

Warpage is induced in the part if the extrudate exiting the die, entering the cooling fixture, or at any other location is dragged across a sharp surface. Molecular orientation is induced by the dragging action, similar to pulling a knife or scissor along a ribbon to make it curl. As discussed previously, high molecular orientation in one part of the extrudate causes different shrinkage from other areas, generating warpage.

Crystalline materials need to be cooled uniformly to generate the same crystallinity and crystal size. If the crystallinity is higher in one section compared to another due to extrudate quenching or cooling differences, the section with higher crystallinity will shrink more. This shrinkage difference induces warpage in the final part. Shrinkage is greater in crystalline polymers compared to amorphous polymers, leading to a higher tendency for crystalline polymers to warp.

Potential causes for warpage and possible corrective actions are summarized in Fig. 28.19.

28.16 Problem 16—Sink

Sink is defined as an indentation in the surface opposite a rib or boss. This is generally a problem only in profile extrusion, as sheet, film, blown film, pipe, tubing, wire coating, and so forth do not have ribs or nonuniform symmetry. Sink is caused by shrinkage. Where a flat area intersects with a rib, there is more material present in the cross section. As the material cools, more shrinkage occurs in the thicker section, causing a surface indentation opposite the rib. Indentation is minimized by decreasing the rib width and the radius going into the rib. However, a very narrow rib compared to the part's main thickness can lead to warpage due to the rib cooling and shrinking faster versus the rest of the part. Once the profile is designed with a rib in place, there is very little that can be done

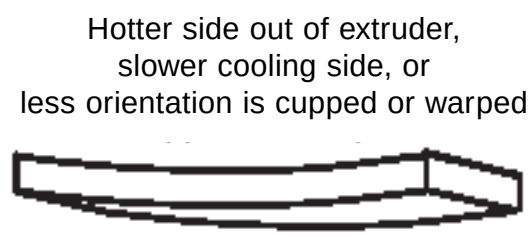

Figure 28.18. Warpage toward hot side, slower cooling, or oriented side.

by extrusion to remove the sink due to ribs. Sink is more of a design issue than an extrusion problem. If unacceptable shrinkage is present, it is necessary to change the part design rather than seek a corrective action in the extruder. Die design can be changed slightly to attempt to force more material into the rib area using higher packing forces. At best this is a stop-gap effort. One potential solution is to use a high density foam in the core of the part. While this may eliminate the sink, it may result in unacceptable property performance.

Review Questions

1. While running an extrusion process, black specks start to appear in the extrudate, and part testing reveals that the physical properties are below standard values. What might have caused the problem, and what potential corrective actions can be taken to eliminate the problem?
2. What causes holes to be formed in the center of the extrudate, and how can the problem be eliminated?
3. What are five potential causes of warped profiles? What is the mechanism for each cause, and what corrective action can be taken to eliminate each?
4. What causes surging in an extruder? How can it be eliminated?
5. What causes streaks and/or discoloration in an extruded part?
6. What causes lines on the extrudate surface in the machine direction?
7. If a resin is not properly dried prior to extrusion, how does the moisture manifest itself in the product? What corrective actions can be taken to eliminate moisture problems?
8. Where does the most surging originate in an extruder and why?
9. What is a gel, and how does it manifest itself in extruded products?
10. What are some potential causes that can result in product variations over time?
11. If a dull streak appears on one side of the extrudate, what corrective actions need to be considered?
12. What causes variations in the product strength as measured by changes in tensile properties over time?
13. What is a weld line? How is it formed? How can its effect on the product be minimized?
14. If the bottom side of an extruded pipe is hotter than the top and cooled more slowly, what direction will the pipe warp? Explain your reasoning.
15. What are some potential causes that produce weak film in an extrusion process?

29 Troubleshooting Sheet Extrusion

Sheet extrusion processes are discussed in Part 7, Extrusion Applications; this chapter covers how to troubleshoot potential problems encountered with the sheet processes and possible corrective actions. The process used to produce sheet is similar to that shown schematically in Fig. 29.1. The distinction between sheet and film occurs at 0.010 inch (0.254 mm); below this value the product is considered film and above it sheet. Problems associated with the sheet processes are centered on the die, three-roll stack, and take-off equipment. Other elements in the sheet process not shown in Fig. 29.1 are knives or slitting elements to trim the sheet edge to finished width dimensions. The edge trim is normally collected by a pneumatic conveying system that transports it to a cutting or dicing station where it is ground up and conveyed to another location for recycling.

A short description of key process elements is required to identify factors involved in problem analysis and corrective actions. After the material exits the die and before it passes through the first set of nip rolls in the three-roll stack, molten polymer builds up between the die exit and the entrance to the first nip roll. This excess polymer is referred to as a bead, a bank, a rolling bead, or a rolling bank of polymer. The bead builds up at start-up and then remains a constant size during operation, with the polymer exiting the die equal to the polymer passing through the rolls at process equilibrium. The correct bead size depends on the polymer and molecular weight or melt flow index.

Mechanical and product problems related to the extruder were addressed in Chapters 27 and 28, with their corrective actions. This chapter focuses on troubleshooting problems associated with flat sheet and the takeoff equipment to produce it. Potential processing problems are addressed first from a mechanical perspective, followed by material or product issues. Troubleshooting associated with coextruded sheet or film is covered in Part 6 under a separate section for troubleshooting.

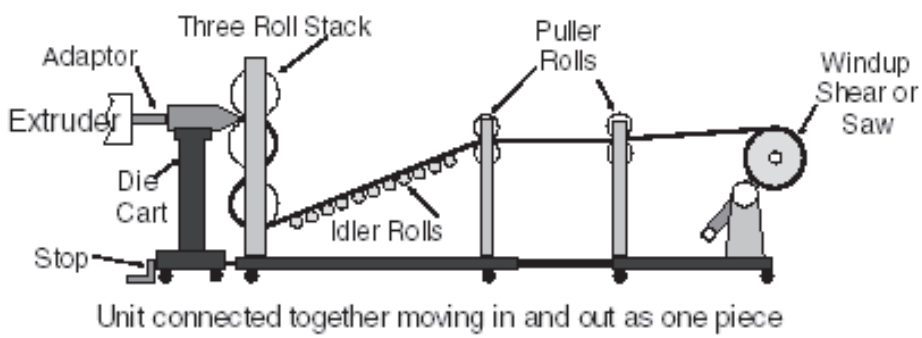

Figure 29.1. Three-roll sheet process.

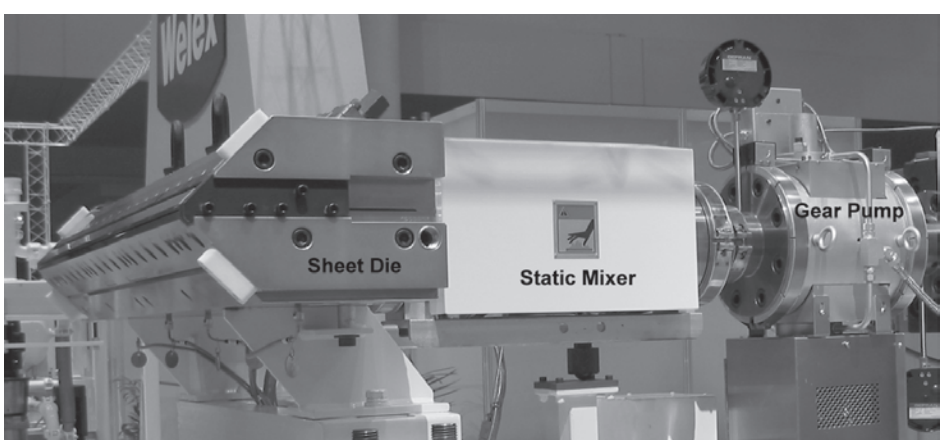

Figure 29.2. Sheet die with thermal mixer and gear pump.

29.1 Mechanical Problems and Solutions

29.1.1 Problem 1—Die Lip Damage Causing Defects in the Machine Direction

The die gap opening and condition of the die lips control the product quality exiting the die and entering the nip in a three-roll stack. Normally the bottom die lip is fixed, and an adjustable top lip creates the desired extrudate thickness. Figure 29.2 shows a sheet die and its adjustable lip. As the bolts on the flexible upper lip (bottom lip fixed) are turned up or down, the die gap opening increases or decreases. Potential problems caused by die lip damage are

- Die lines or straight lines in the product in the machine direction are caused by a scratch, nick, burr, or other imperfection in the die lip. Depending on which die lip contains the imperfection, the product defect appears on either the top or bottom of the sheet in the machine direction.
- Warped die lips create nonuniform polymer flow across the die, leading to transverse direction thickness variations and poor gauge control. Thickness

remains constant in the machine direction but is variable across the sheet. Pressure from the polymer exiting the die is not sufficient to bend warped die lips to a flat shape; consequently, die lip adjustments in operation cannot correct the problem. The damaged die lip has to be replaced. Die lip flatness with no material exiting the die is essential to obtain uniform sheet or film with tight thickness tolerances. Warped die lips can be created by repeatedly applying external heat with a torch to bring a die up to temperature quicker.

- Die lip finish has to be in good condition with the die lip to die face angle fixed at 90˚ to ensure a tight fit to the die body. In addition to the finish being smooth, polymer has to flow freely over the surface with minimal friction. Periodically the die lips have to be resurfaced and replated to ensure the surface quality. The purpose of the 90˚ angle is to reduce dead space, where material can stagnate and degrade over time.
- The top flexible lip adjustment bolts have to be clean and easily adjusted to create the proper product thickness. The adjustment screws have fine threads; however, small adjustments can still make significant changes in the thickness profile across the sheet. Decreasing the die gap in one location forces polymer to flow out an adjacent location. If an adjustment bolt is not easily rotated to open or close the die, it is difficult to control the product thickness in that section. In addition to the threads being clean and in good condition, the Allen head has to provide a good fit with the wrench.

Figure 29.3 summarizes the effect of die lip quality on the sheet and corrective actions to use in the event problems or issues exist in the sheet relative to the die.

29.1.2 Problem 2—Chill Roll Adjustment or Condition

Problems associated with the chill roll are gauge uniformity or thickness control in both the machine and transverse directions, lines on the product in the machine direction, nonuniform gloss, and sheet roughness. Potential corrective actions for these problems are summarized below.

- Poor gauge or thickness uniformity in the transverse direction can be caused by improper gap adjustment between the rolls. Typically the first gap is set at 1–2% greater than the final desired thickness, while the second gap in a three-roll stack is set slightly below the desired final thickness. Cocked rolls or uneven pressure across the roll surface can cause thickness variation across the product. As roll pressure is increased, rolls have a tendency to deflect in the center, leading to thicker product gauge in the center relative to the edge. For wide sheet, crowned rolls with the center having a slightly larger diameter than the edge are used in high-pressure situations to produce flat sheet. Another factor leading to gauge variation in the transverse direction is bearing wear on the roll shaft, allowing the roll to wobble slightly.
- Gauge variations in the machine direction can be caused by drive roll slippage or nonuniform roll temperatures. Proper tension and lubrication of the chain driving the rolls can prevent slippage. Temperature across the roll surface from the inlet to the outlet needs to be controlled within 2˚F (1˚C). If greater temperature variation is present, the sheet will not cool uniformly, creating different shrinkage effects across the roll and viscosity variations as the molten resin comes in contact with the

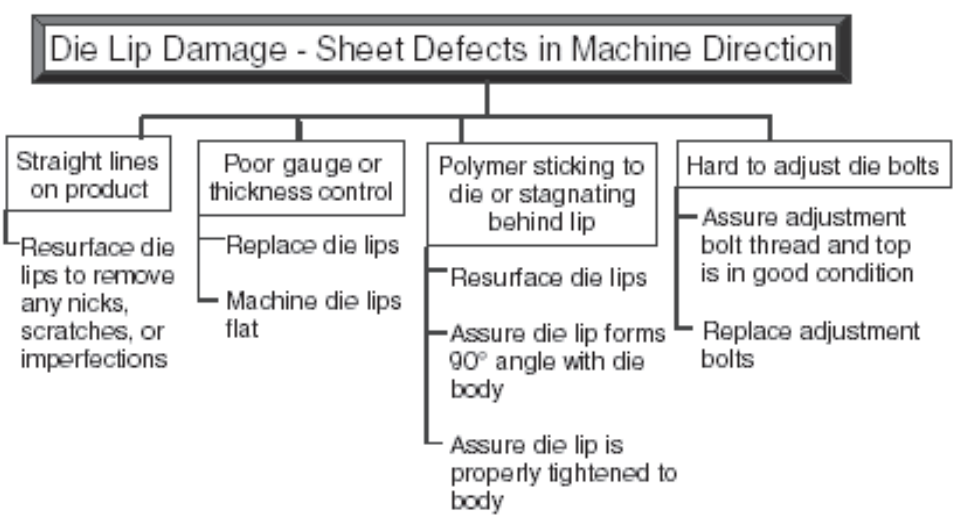

Figure 29.3. Problems caused by die lip damage and corrective actions.

roll surface. Nonuniform viscosity allows the polymer to flow differently across the roll as it is squeezed in the nip area.

- Lines, pits, and/or imperfections in a repeating pattern in the machine direction can be caused by scratches, dents, dings, and so forth in the roll surface. A scratch going all the way around the roll surface can be caused by moving the rolls too close to the die or having a metal object coming in contact with the rolls while they are rotating. A scratch of this nature will cause a line on the sheet in the same location over the entire production run. A damaged spot on the roll in one location leaves a mark or defect on the sheet on every roll revolution. To eliminate sheet or film imperfections caused by roll surface damage, the roll has to be removed and repaired or refinished. Deposit build-up is less severe because this can be removed during the production run. However, care must be taken to do it safely, ensuring hands, hair, loose clothing, and so forth are not caught in the rolls. Prior to start-up, the rolls are cleaned with cleanser or other material to remove any plate-out or deposits built up from previous production.

- Sheet warpage is when the sheet wants to curl, twist, or cup. As discussed in Chapter 27, this is caused by differential shrinkage. Using a down stack as shown in Fig. 29.1 (three-roll stack arrangement where the sheet or film is fed into the stack between the top two rolls with the sheet or film proceeding downward through the stack), if the resultant sheet curls or warps toward the bottom of the sheet, it means the bottom side of the sheet is hotter, i.e., cooling more slowly. To correct the warpage, either raise the top stack roll temperature or lower the middle roll temperature. If the product curls up in a down roll stack, the top is hotter and cooling more slowly than the bottom. To correct this warpage, lower the top roll temperature and/or raise the middle roll temperature. Running in an up stack configuration (polymer exiting the die goes between the bottom two nip rolls, proceeding upward and exiting over the top roll), change the roll temperatures in the opposite direction to that in a down stack operation. If the product is warping upward, the top side is too hot, so the middle roll temperature is lowered and/or the bottom temperature raised. Regardless of whether it is an up or down stack, the third roll temperature is always changed opposite the middle roll.

- Improperly adjusted rolls can lead to poor gloss and rough sheet. To obtain the best gloss, highly polished roll surfaces at high temperature with the correct pressure (PLI = pressure/linear inch) are required. A guideline for setting roll temperature is to raise the roll temperatures after the process is running until the sheet or film just starts to stick to the roll surface, and then lower the temperature approximately 10˚F (5.5˚C). The roll surfaces have to be clean with uniform pressure applied across the rolls. If the sheet has a rough surface and the gloss is poor at the correct roll temperature, verify that the rolls are clean and the gap is properly set for the thickness being processed. The sheet is polished by the roll temperature, pressure, and the roll surface quality with which the polymer is in contact or wrapping. Polish is easiest to obtain with the middle roll. The middle roll contacts the sheet top in an up stack and the sheet bottom in a down stack. The opposite sheet side polish is influenced by the first and third rolls in the stack.

Figure 29.4 summarizes the potential problems and corrective actions associated with the stack rolls.

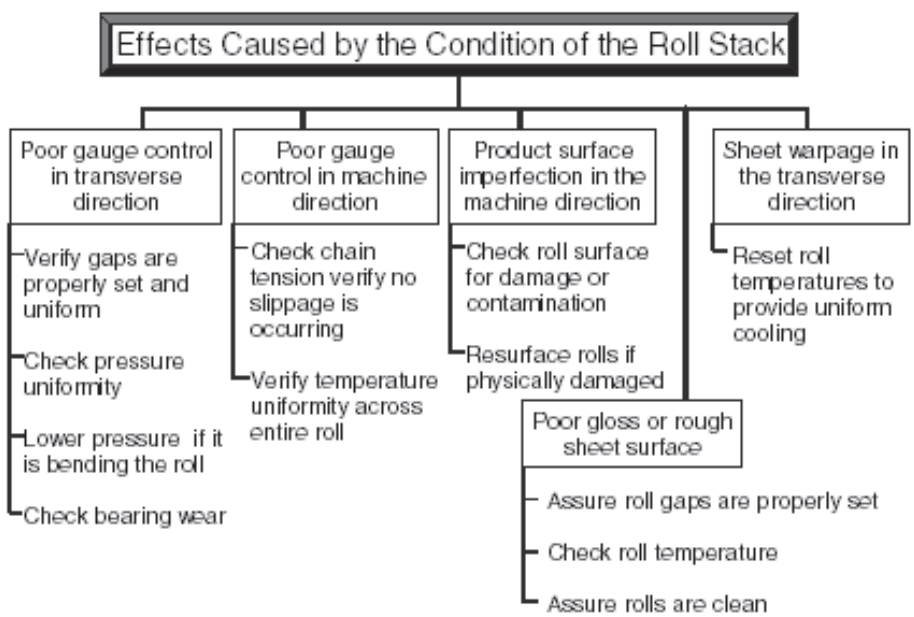

Figure 29.4. Sheet problems created by the roll stack and their corrective actions.

29.1.3 Problem 3—Product Variations Caused by the Puller Rolls or Winder

Problems can be created by the pull rolls or winder if they are not properly synchronized with the three-roll stack, either generating too much tension or pulling the product across the chill rolls. Potential problems associated with the winder or pull rolls overpowering the three-roll stack or chill roll are

- Nonuniform thickness is created in the transverse direction by excessive pulling of the winder or pull rolls, causing the sheet to slip and stick in the roll stack. When the sheet is slipping on the rolls, the resultant product necking is increased, and the product is more likely to be thin on the edges and thicker in the middle than when it is adhering to the three-roll stack and moving with those rolls. In some situations the product is slipping constantly, and lines are formed in the transverse direction by the apparent action of starting and stopping in the nip or chatter as it goes through the nip.
- Thickness variations in the machine direction occur because the product is not being pulled uniformly through the nip or across multiple cooling rolls. The gaps in the three-roll stack have to be properly set so the sheet cannot be pulled through them by the pullers. Slippage causes the resultant product to be thinner, while sticking to the rolls creates a thicker product with varying thickness in the machine direction.

Figure 29.5 summarizes the potential problems associated with the puller or winder speed not being properly synchronized with the chill roll speed, causing the product to be pulled through the nip rolls.

29.1.4 Problem 4—Static Charge Build-up on Sheet

Static charge can build up on the sheet as it is being processed either through a three-roll stack or over individual chill rolls, creating surface contamination by attracting airborne foreign material. Proper equipment grounding plus an alternative method, such as copper gauze running across the sheet and being connected to a ground, is essential to eliminate static electricity generation in the process. Housekeeping and cleanliness around the line help prevent airborne foreign material being attracted to the product in the event static electricity is present. A perfectly acceptable product for physical properties, gauge, color, and gloss can be made unacceptable by surface contamination resulting from static electricity attracting foreign materials to the product surface.

29.1.5 Problem 5—Other Equipment Problems

Other equipment problems can lead to product problems resulting in unacceptable or rejected products. Electrical problems associated with voltage variations can cause nonuniform thickness or gauge variations in the machine direction resulting from speed variations anywhere throughout the line. Voltage variations within any temperature zone in the extruder or die can lead to nonuniform melt temperature, creating melt viscosity variations resulting in gauge or thickness issues or, in extreme cases, changes in the physical properties.

Equipment alignment between the die, roll stack, and windup is critical. The roll stack or nip opening has to be parallel to the die lip opening, and the windup has to be directly in line with the roll stack. Normally the roll stack, puller rolls, and windup or stacker are bolted

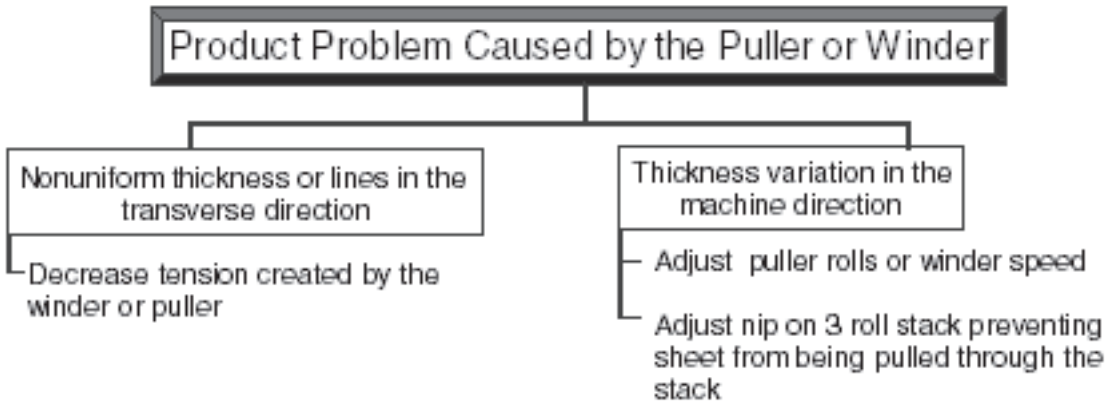

Figure 29.5. Potential problems created by the winder or puller running too fast.

together and moved as a single unit toward or away from the die on tracks. The tracks are at right angles to the die lips to properly align the nip with the die. Gauge nonuniformity can result if the die and rolls are not properly aligned. Warped, curled, or twisted sheet results when the winder and puller rolls are not properly aligned. This warpage is caused by one side of the sheet being stretched more than the other, changing the molecular orientation and shrinkage characteristics.

Equipment vibration can lead to gauge variations in either the machine or transverse direction by disrupting or disturbing the molten polymer as it leaves the die or in the rolling bank or bead. Any vibration resulting in a nonuniform feed to the nip area can create gauge variations. Winder or puller roll vibrations can create nonuniform pulling tension in the three-roll stack or across chill rolls, producing gauge variations. Equipment vibrations may signal defective bearing or other equipment abnormalities requiring maintenance.

29.2 Product Problems and Solutions

29.2.1 Problem 1—Sheet Product Contamination

Contamination falls into two categories: external or internal, with foreign material on either the outside or inside. Potential causes for each contamination and corrective action are given below.

- External contamination with foreign material on the sheet can be caused by static electricity attracting airborne objects to the sheet surface (discussed previously) or deposit build-up on the equipment that drops onto the product. If the polymer being used is prone to die deposits, die build-up will come loose and fall onto the molten web unless it is periodically removed. Assuming die deposits do fall on to the molten web, they are incorporated or forced into the product at the nip. Any other location along the line where deposits (plate-out on the rolls, foreign material on the idler rollers, and so forth) can build up and be released is a potential cause for contamination. Housekeeping and cleanliness in operation can eliminate all the potential sources.
- Internal foreign material comes from the extruder or die. Raw materials used in production may contain foreign materials, dirt may be introduced with regrind, the wrong regrind may be used, extruder feed transfer lines may not have been properly cleaned, gels or cross-linked particles can be formed in stagnant locations in the die, and so forth. (Review problems 6, 9, and 11 in Chapter 28.) Contamination generated in the extruder itself can come from ruptured screen packs or plating loss from the screw or die.

Figure 29.6 summarizes the potential sources for contamination and corrective actions to eliminate the problem.

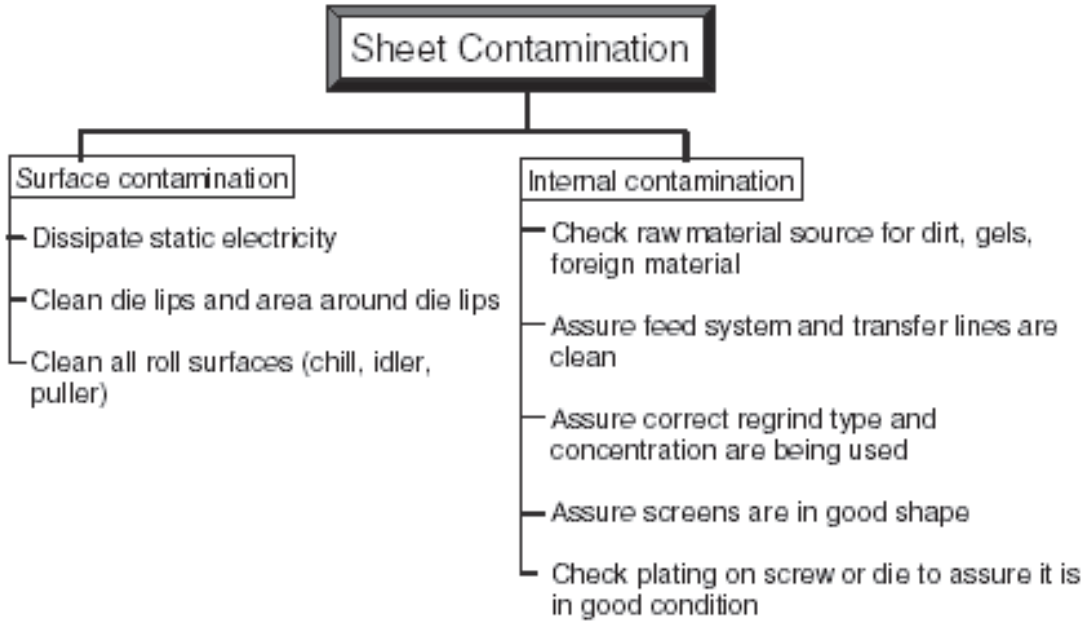

Figure 29.6. Potential causes for sheet contamination and corrective actions.

29.2.2 Problem 2—Poor and/or Nonuniform Gloss

Gloss measures the sheet surface reflectivity and is directly related to the polished roll surface quality and temperature. To obtain a matte finish, the polished rolls are replaced with matte-finished rolls. The matte-finished roll imparts a specific micro-roughness to the sheet surface. Potential causes for low and/or nonuniform gloss with potential corrective actions are given below.

- Poor or low gloss, assuming the roll surface quality is acceptable, can be caused by factors related to roll and polymer melt temperature. For best gloss, run the roll temperatures as hot as possible; a guideline is 10˚F (5.5˚C) below the temperature where the sheet just starts to stick to the polished rolls. Roll gaps have to be properly set to ensure uniform contact and pressure across the product. The melt temperature has to be uniform, which reverts back to good mixing in the extruder, with no hot or cold spots in the die, adapters, or transition pipes. To ensure uniform melt temperature, static mixers can be installed in melt pipes between the extruder and the die. This action is not normally necessary with proper screw design. Raise the melt temperature to reduce the resin viscosity, creating better flow and ability to duplicate the roll surface. Raise the last two extruder zones, adapter, and die temperatures to increase the melt temperature. The die temperature needs to match the melt temperature; polymer flow across the die opening has to be uniform.
- Nonuniform or inconsistent gloss across the sheet can be caused by nonuniform roll temperatures (discussed previously), inconsistent gauge or thickness, volatiles, or high melt temperatures. Inconsistent gauge can originate in the extruder through surging (see Chapters 27 and 28 for surging), in the die by improperly setting the die lip opening and/or improperly setting the choker bar, and/or inappropriate roll gap adjustment. Sources for volatiles include resin moisture caused by improper drying, excessive melt temperature leading to polymer or additive degradation, plugged extruder vent preventing gas removal, and a plugged or insufficient vacuum to successfully remove all the volatiles. Volatiles generated at high temperature as the extrudate exits the die can become trapped in the area between the roll and the sheet. Volatile defects show up as splay or other surface defects, destroying the gloss in that area.

Figure 29.7 summarizes the potential problems and corrective actions associated with poor and/or nonuniform gloss across the sheet.

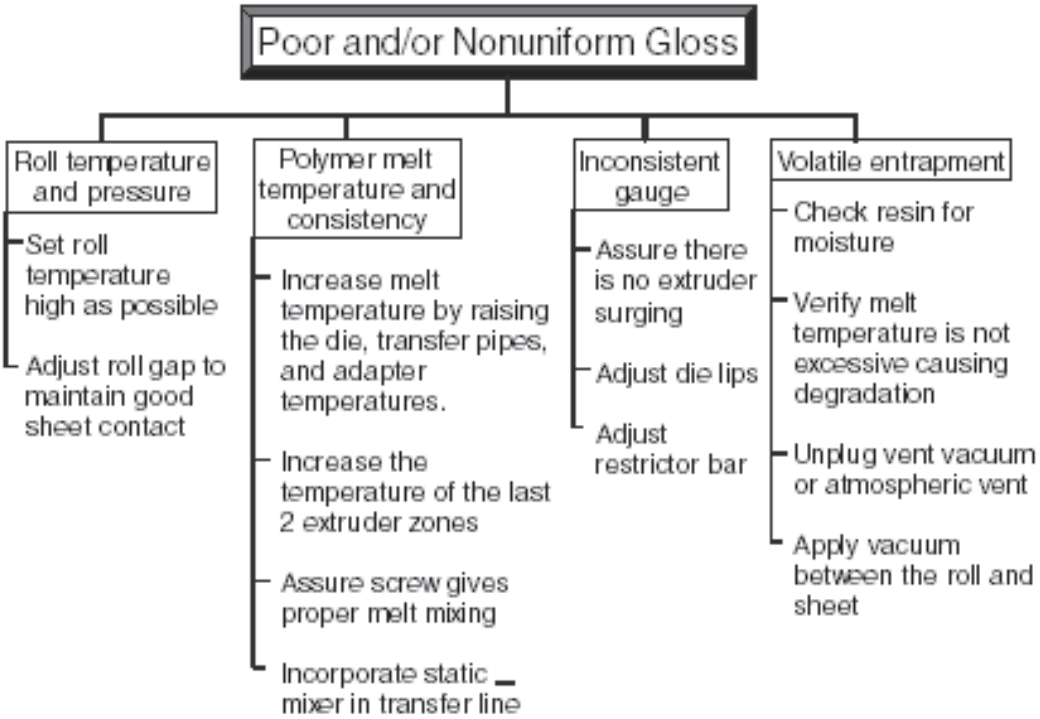

Figure 29.7. Potential causes and corrective actions for poor and nonuniform gloss.

29.2.3 Problem 3—Inferior Transparency

Transparency or optical clarity is important in amorphous materials such as polycarbonate, acrylic, and polystyrene. With crystalline polymers, melt has to be quenched rapidly to prevent high crystallinity and brought to room temperature quickly to ensure small crystal size. To ensure maximum transparency, processing conditions and equipment have to be operating at peak efficiency. Some factors that may lead to inferior transparency with potential corrective actions are

- Resin contamination from too much regrind, the wrong regrind, or foreign material can convert a transparent product to a hazy or opaque product. Contamination can be generated in the extrusion process if the equipment is not properly cleaned or dead spaces exist within the system where polymers can stagnate and degrade over time.
- Polished rolls are dirty or need to be refinished. Surface aesthetics, gloss, and optical transparency are controlled by the roll surfaces; product surfaces can never be better than the roll surface being used to produce them.
- Proper roll and melt temperatures are necessary to ensure good surface quality. Cooling rates with crystalline materials are critical, as they affect both crystal size and growth rate.
- Proper resin drying prior to extrusion and resin transfer in a dry environment between the dryer and extruder ensure resins do not pick up moisture as they are awaiting extrusion.

Figure 29.8 shows a schematic identifying potential causes for reduced transparency in sheet and potential corrective actions.

29.2.4 Problem 4—Color Shift or Wrong Color

Some potential reasons for product discoloration centered around degradation were presented in problem 6 in Chapter 28. Different shear configurations in an extruder can develop color differently. Consequently, if the same product run previously had a different color, in addition to reviewing the formulation data, the extruder setup and screw design need to be checked. Potential causes for color shifts with corrective actions are given below.

- The wrong color formulation or color concentrate quantity is being used. Check the formulation to see if all concentrations are correct. Check the blend to ensure all materials are feeding uniformly and the blend is properly mixed. If the color difference is from shear forces, it may be necessary to change the screw design or the formulation to produce the proper color.

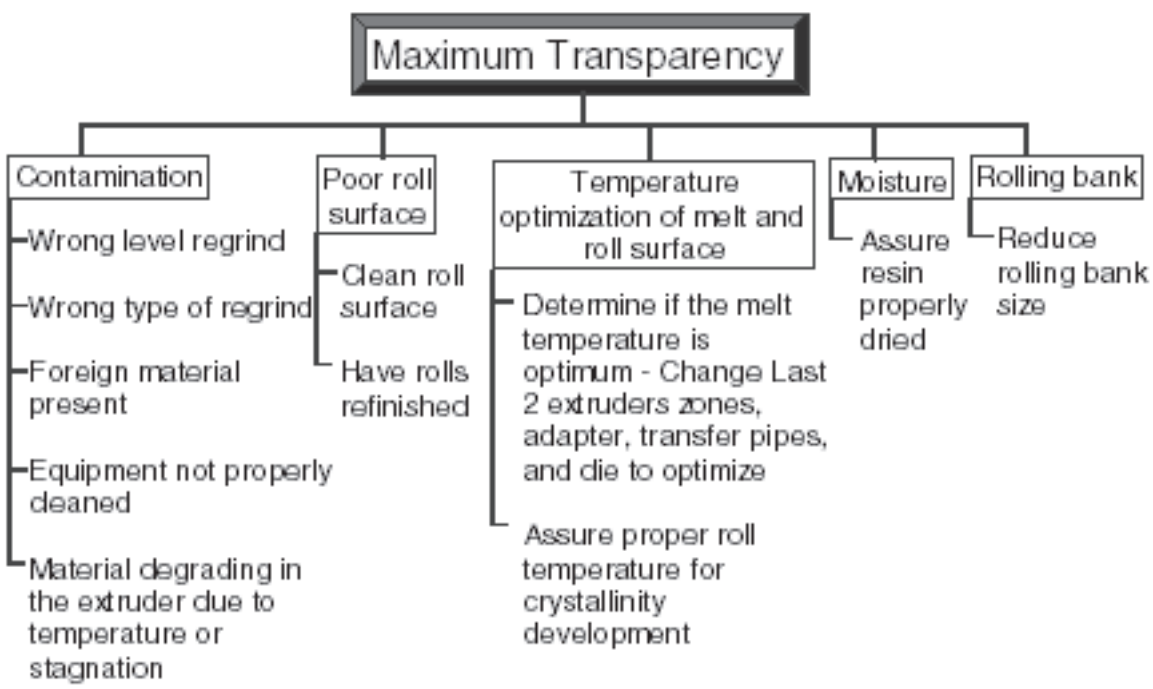

Figure 29.8. Potential causes and corrective actions to improve product transparency.

- Resin formulation or color concentrate degradation can occur due to overheating, long residence time, or combinations thereof. If deckle bars are in use to adjust the sheet width, verify that resin build-up behind the bars is not causing discoloration in the sheet due to resin degradation. Ensure all transition pipes have smooth transitions, which prevents resin stagnation and discoloration over time.

Figure 29.9 summarizes potential causes for color shifts or discoloration and corrective actions.

29.2.5 Problem 5—Sheet Is Brittle

Sheet ductility is essential for downstream handling and during the product life. Ductility loss can lead to brittleness, which is characterized by the sheet easily ripping or tearing in one direction or shattering when something is dropped on it. This can be caused by molecular orientation and resin degradation, which were covered in Chapter 28 in problems 5, 6, and 14. Potential causes for sheet brittleness are

- Using a low-molecular-weight polymer (high melt flow index) can result in the short chains aligning in the flow direction and not providing sufficient transverse direction strength. Adding too much degraded or lower-molecular-weight regrind can cause the overall formulation to have less melt strength.
- Additives may embrittle the resin or lower the melt strength, substantially reducing the processing window. Additives having a tendency to cause embrittlement or lower melt strength are regrind, colorants, fillers, flame retardants, and reinforcements. Conversely, impact modifiers and plasticizers can reduce resin embrittlement and increase melt strength.
- High machine direction orientation reduces the transverse direction strength, allowing sheet and film to crack easily in the machine direction. If cracks are found to propagate easily in the machine direction, molecular orientation is the number one potential cause.

Figure 29.10 shows the potential causes and corrective action to use when sheet has low strength. In addition, the corrective actions in problems 5, 6, and 14 in Chapter 28 need to be considered.

29.2.6 Problem 6—Surface Voids, Indentations, and Imperfections

Sheet surface imperfections are normally surface indentations and are called by numerous names, e.g., pits, dimples, pock marks. Build-up or deposits on the chill rolls or idler rolls were discussed previously. In this section, surface imperfections caused by improper processing conditions or setup are discussed. Potential causes for

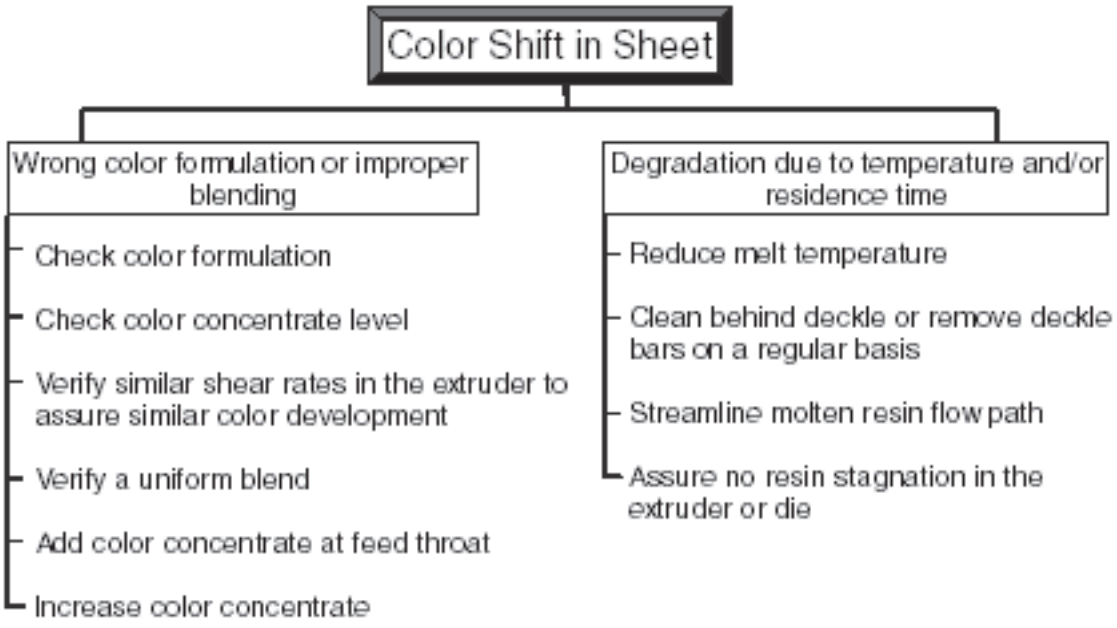

Figure 29.9. Potential causes and corrective actions for color shifts.

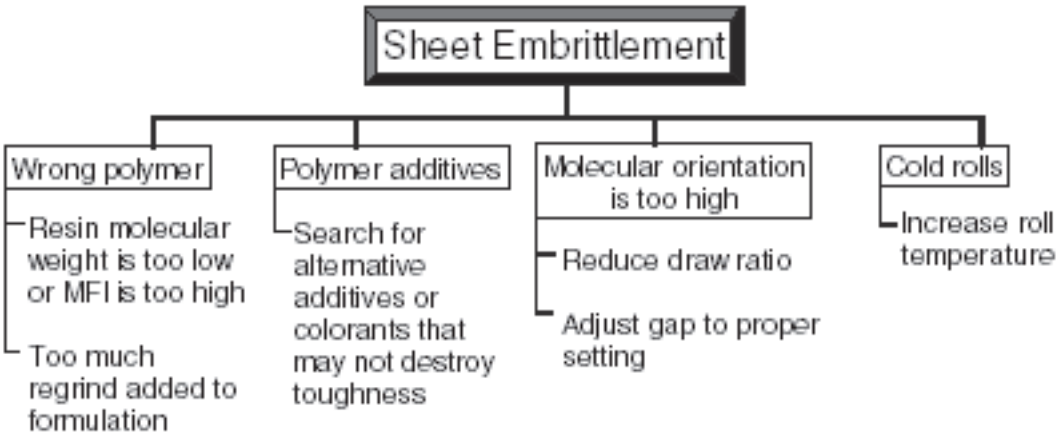

Figure 29.10. Potential causes and corrective actions for sheet embrittlement.

surface imperfections with their corrective actions are discussed below.

- Moisture that is not removed from the resin by either an atmospheric or vacuum vent exits the die as steam, causing splay, indentations, or, in the most extreme case, holes in the product. Moisture-sensitive products need to be monitored for moisture prior to extrusion to get the moisture content below the minimal acceptable limit before starting. For some resins it may be necessary to blanket the feed hopper after drying to keep the material dry prior to processing. If the moisture content is too high after drying, check the dryer.
- Volatiles generated during extrusion from resin outgassing or an additive has to be removed by vacuum or atmospheric venting prior to the die. As with moisture, volatiles exiting at the die can cause surface imperfections in the best situation and holes in the worst. Lowering the melt temperature may help to minimize volatile formation.
- A small bank between the die and the first nip roll can cause surface imperfections from too little material being compressed in the nip. The rolling bank or bead is increased by increasing the extruder output (screw speed in flood-fed extruder, feed rate in starve-fed extruder), decreasing the line speed, or decreasing the gap between the rolls. If the bead is not uniform across the sheet width, the die lips or choker bar might have to be adjusted to provide more uniform polymer flow from the die. The best way to tell if the bead is uniform is to open and close the nip rolls. This action allows the bead to pass through the nip, and the uniformity can be viewed by the operator. If the bank or bead is not uniform, die adjustments are required to produce a uniform bead.
- Air entrapment is sometimes blamed for sheet indentations. However, in most situations air is removed in the extruder and this is not usually the cause. Assuming air is identified as the cause, the screw must be filled with polymer prior to discharge and the air removed through a vent or vacuum port. In a flood-fed single screw extruder, air normally flows out the feed throat as the polymer is melted and compressed in the transition section. Air is not present in the metering section. In a twin screw extruder the last screw flights are designed to be full and pump material out the die. As the melt is compressed with smaller-pitch elements, the air flows backward out the vent.
- Cold roll temperatures can cause indentations by cooling the sheet or film too rapidly, causing a skin to form on the surface with additional shrinkage occurring on the inside later. If the roll temperature is below the dew point, moisture can condense on the roll surface, creating surface voids when the hot sheet comes in contact.

Figure 29.11 summarizes potential causes of voids or indentations in the sheet surface and corrective actions to eliminate the problem.

29.2.7 Problem 7—High or Inconsistent Molecular Orientation

Numerous factors can contribute to high molecular orientation in the sheet. As discussed previously, high molecular orientation leads to anisotropy or differential physical properties in the machine versus transverse direction. If these properties are required by the application, high orientation is advantageous; otherwise it may

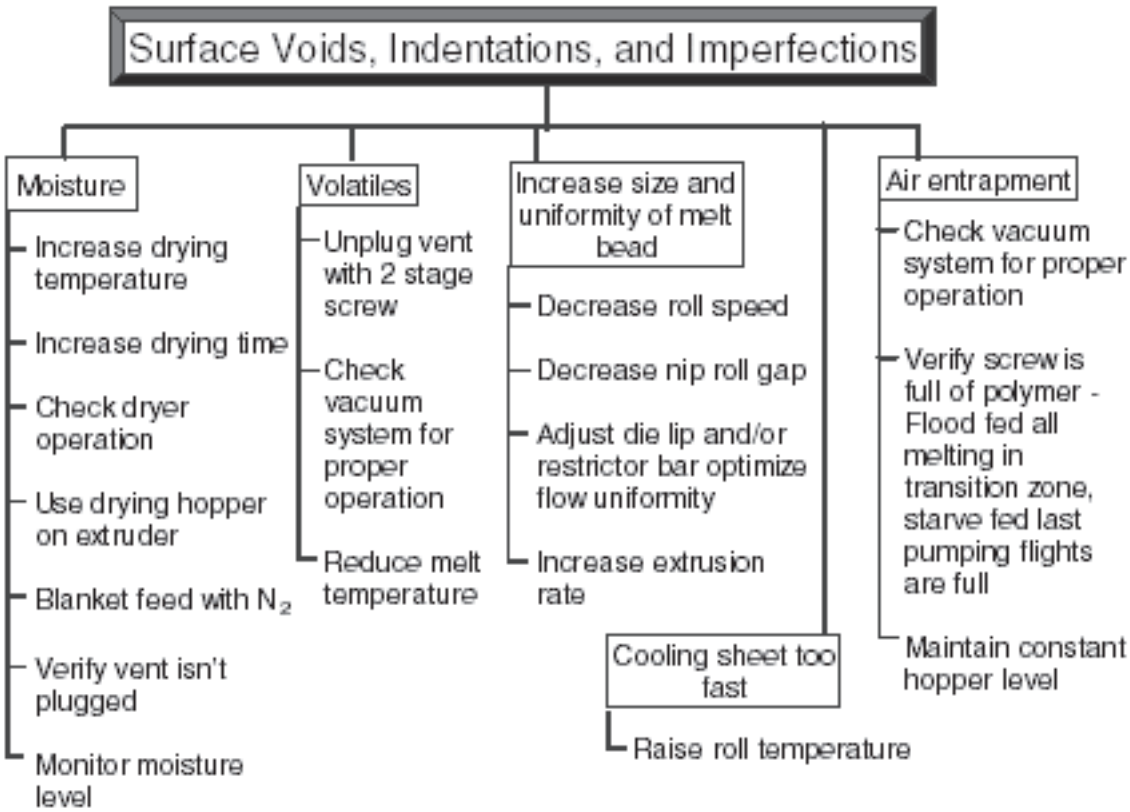

Figure 29.11. Potential causes and corrective actions for surface voids, indentations, and imperfections.

have a very negative impact in the application. Potential causes for high molecular orientation or nonuniform orientation are outlined below.

- Chill roll speed is high compared to the die throughput rate, causing the polymer to be drawn to high levels between the die and the first nip. Corrective action is to reduce the die gap. This lowers the draw down ratio and reduces the melt orientation.
- The pull roll or winder is running at significantly higher speed than extruder or die output, and high machine direction drawing occurs. This can also cause high tension, leading to cold drawing or drawing on the rolls. Corrective action is to reduce the puller or winder speed or increase the throughput rate and the chill roll speeds.
- Polymer cooling rate can affect the molecular orientation if the quenching rate causes solidification before all the drawing is completed. Stretching film in the semisolid state increases the molecular orientation, as the molecules do not have a chance to relax after drawing. The differential between the melt and roll temperatures needs to be kept as small as practical to prevent premature solidification and solid state drawing. An alternative approach is to increase the melt temperature to prevent the sheet from being quenched too quickly. This is a bigger issue with crystalline polymers because they have sharp melting points, while amorphous polymers remain rubbery above their glass transition temperature.
- Inconsistent or nonuniform molecular orientation can be caused by the die/nip distance not being constant. This is not normally a problem where tracks are attached to the floor and all equipment is properly aligned. However, if the die is closer to the nip on one side compared to the other, a draw ratio difference exists between the two sides of the sheet or film, resulting in inconsistent molecular orientation.
- Another factor leading to inconsistent orientation is a large rolling bank that is not uniform across the sheet width. In areas that contain a larger bank, less drawing is likely to occur, causing nonuniform molecular orientation.

Figure 29.12 summarizes potential causes for high or nonuniform molecular orientation and possible corrective actions.

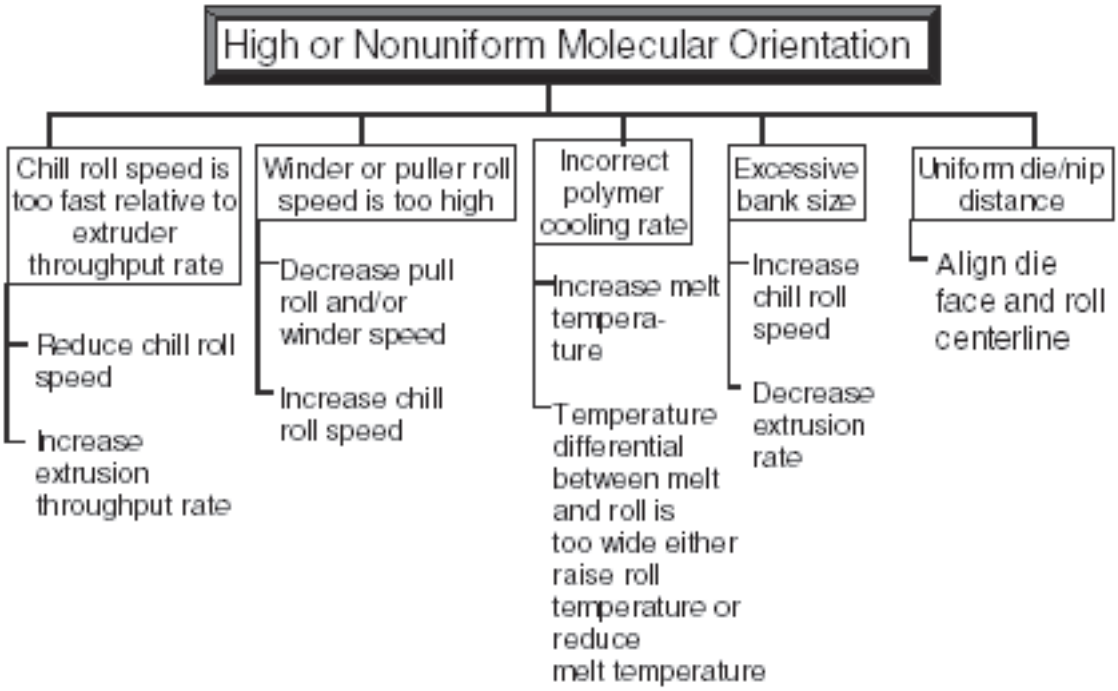

Figure 29.12. Potential causes and corrective actions for high molecular orientation.

29.2.8 Problem 8—Warped Sheet

Warped sheet has a curled, twisted, or cupped appearance caused by differential shrinkage due to molecular orientation, coating effects, crystallinity development, or packing. This is similar to problem 15 in Chapter 28. The direction and the shrinkage differences determine the warpage. Potential causes for warpage and corrective actions to eliminate warpage are given below.

- Causes for molecular orientation were discussed previously in problem 7. Assume sheet has a high molecular orientation in the machine direction on one edge versus the other edge due to an improperly adjusted die or distance between the die and the roll stack. The final product is going to show some warpage. This same phenomenon can be caused by an improperly set roll gap, where the product is drawn on one side more than the other.
- Uniform cooling is important in controlling warpage in sheet products. Both sides have to be solidified at the same time to prevent shrinkage differences from one side of the sheet to the other. This is controlled by the roll temperatures. The side the product is cupped toward is the hotter side, and the roll temperature contacting this surface has to be lowered or the roll temperature contacting the opposite surface has to be raised. Temperature nonuniformity across the roll can cause warpage in the transverse direction. This was covered previously in problem 2 under "Mechanical Problems" earlier in this chapter.
- Crystallinity development is related to cooling rates and cooling uniformity. In crystalline polymers, such as PE, PP, PBT, and so forth, higher crystallinity in one part of the sheet or film compared to another location can cause differential shrinkage, with the sheet shrinking more in the higher crystalline area. Cooling must be uniform across the entire sheet to develop uniform crystal size and degree.
- Packing is the final molecular factor that can cause differences in shrinkage. If the roll gap is smaller or the pressure higher in one area of the sheet or film compared to an adjacent area, more molecules can be packed into a fixed area or volume than in another location along the product. Areas with higher packing (density) will shrink less than a similar area of lower packing, leading to shrinkage differences and warpage.

Figure 29.13 summarizes potential causes of warped sheet and corrective actions to implement to reduce or minimize the warpage.

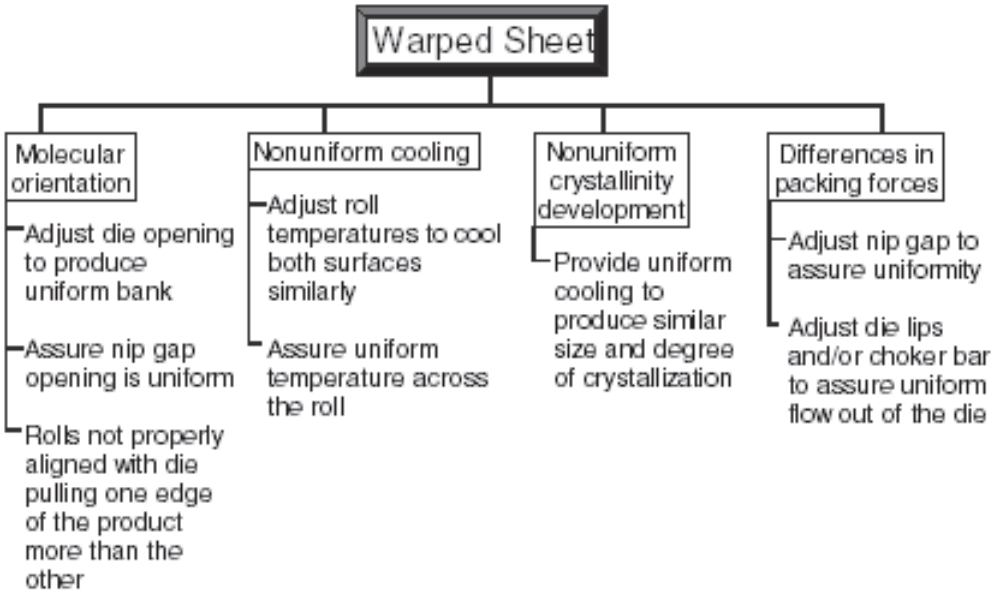

Figure 29.13. Potential causes with corrective actions for cupped, curled, and twisted sheet.

29.2.9 Problem 9—Thickness Variations

Thickness variations in either the machine or transverse direction are caused by a nonuniform bank, surging in the extruder, and sheet slippage in the roll stack due to excessive winder or puller tension. Potential causes with corrective actions for variations in thickness are given below.

- Die lip opening is an important control element for thickness variations, either across the sheet or in the machine direction. Small die lip opening changes can increase or decrease the thickness in one area while simultaneously affecting adjacent areas. As the die lips are closed, polymer flow is restricted and the sheet becomes thinner; however, in the adjacent area the sheet becomes thicker as the polymer flow is redistributed. The choker or restrictor bar is used to change the polymer flow to the die lips. In comparison, the choker bar provides a broad effect, while the die lip adjustment has a narrower effect.

- Die temperature also changes the polymer flow characteristics. Increasing a die zone temperature lowers the polymer viscosity in that zone, resulting in higher flow. In wide sheet dies, there are multiple temperature zones across the die to maintain uniform melt temperature across the die. Assuming the die lips and choker bar are properly set and the sheet or film thickness in one section is thicker than in another, the temperature is raised where the product is thin and lowered where the product is thick to create more uniform flow from the die. Excessive heat can cause hot and cold spots in the die. It is important to realize that uniform cooling is necessary to prevent warpage and to minimize molecular orientation differences from drawing in hot and cold areas, one must keep the temperature differentials across the die that are used to control polymer flow to a minimum. Depending on the environment and the higher exposed surface area, die edges are sometimes run hotter than the center to increase the polymer flow to the edge and decrease the polymer flow in the center.

- Surging in the extruder has to be eliminated to produce uniform sheet thickness. Even minor surging can result in sheet thickness variations in the machine direction. In some applications where the gauge is very critical, gear pumps are used to provide uniform polymer pressure and flow to the die.

- The rolls have to be properly gapped with uniform pressure across the entire roll width. Excessive pressure may cause the rolls to bend, creating thicker sheet in the center. Uniform temperature across the rolls provides uniform cooling.

Figure 29.14 summarizes the potential causes for sheet thickness variations and corrective actions to employ to eliminate problems associated with thickness variations.

29.2.10 Problem 10—Lines and Roughness on Sheet

Surface defects defined as roughness or lines on the sheet can be caused by the die, rolls, moisture, puller, and/or winder. Potential causes and corrective actions for lines or rough sheet are discussed below.

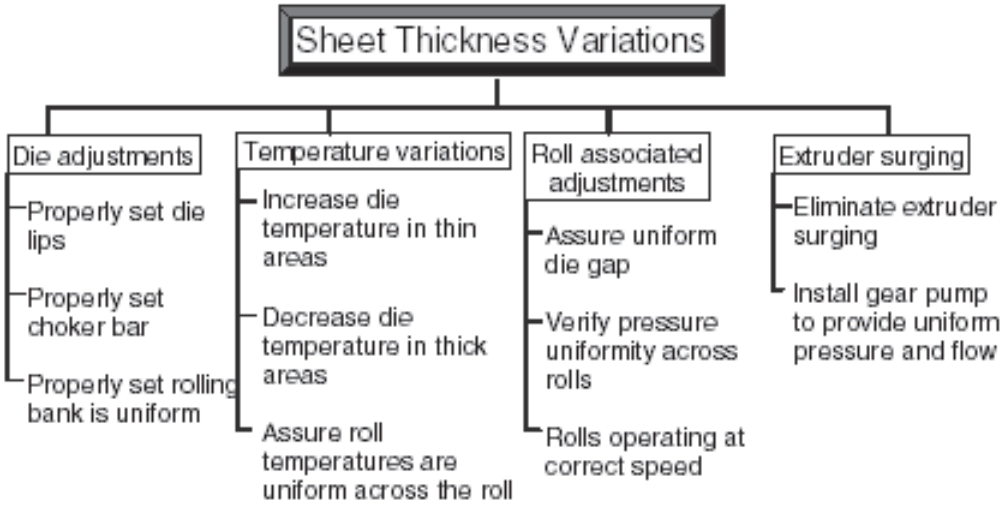

Figure 29.14. Potential causes and corrective actions for variation in thickness.

- Dirt in the die lip area caused by degraded polymer, blown screen pack pieces trapped in the die, or scratches on the die lips can generate surface lines in the machine direction. If the line is on both sides in the same location, it can be a weld line caused by something trapped in the die lip area that is disrupting the polymer flow and making it recombine. It may be necessary to remove the die lips or disassemble the die to remove the object creating the problem.
- Dirty or damaged roll surface can cause lines or roughness. The solution is to clean the rolls and, if necessary, have them refinished.
- Minor amounts of moisture can generate splay or a rough surface. Proper resin drying and/or venting is required to remove unwanted gases and volatiles.
- A puller or winder not properly synchronized with the chill rolls can create too much tension, pulling the sheet through the chill rolls. This can create a slip/stick situation, generating lines across the sheet. Corrective action is to properly synchronize the speeds and tensions to prevent chill roll slippage.

Figure 29.15 schematic summarizes the potential causes and corrective actions to implement to eliminate surface roughness and lines.

Troubleshooting associated with coextruded sheet is covered in Part 6 in a separate section.

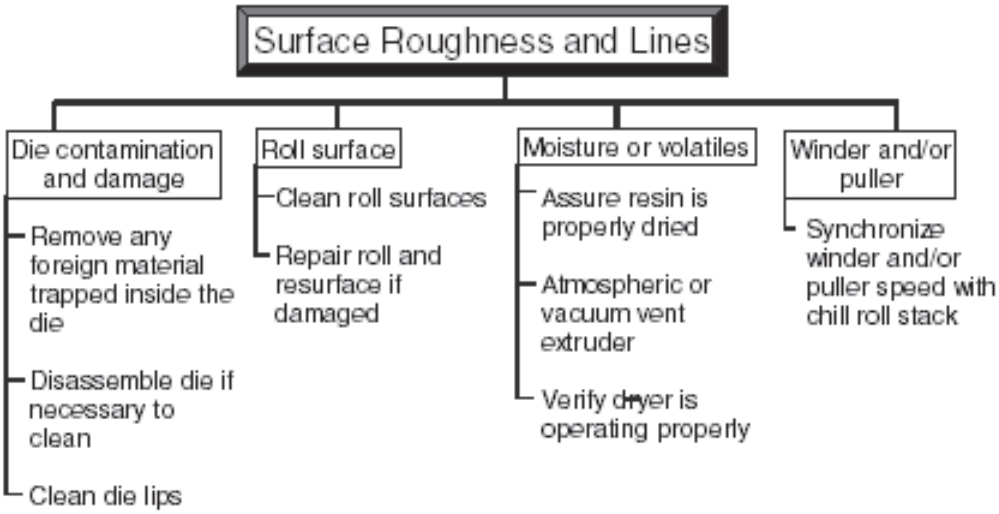

Figure 29.15. Potential causes and corrective actions for surface roughness and lines.

Review Questions

1. What causes twisted, curled, or cupped sheet and how is it eliminated?
2. What are some of the potential causes for inferior gloss or nonuniform gloss across the sheet surface?
3. What is a rolling bank or bead and how is it controlled?
4. What might cause rough surface on sheet and how can it be eliminated?
5. What chill roll adjustments are possible and what does each adjustment control?
6. What sheet variations are directly related to the puller rolls and winder?
7. What can lead to a color shift in the product and how is it corrected?
8. If static charge is present on a sheet, how can it be dissipated, and what are the consequences of static?
9. How can high molecular orientation in the polymer be eliminated? What are some potential problems associated with high molecular orientation?
10. What are some causes of sheet thickness variation and how can they be eliminated?
11. What causes sheet to be brittle and what are the potential corrective actions?
12. How can sheet transparency be improved?

30 Troubleshooting Cast Film

The cast film process is shown in Fig. 30.1. A vertical die lays a uniform thickness film (less than 0.010 inch [0.254 mm]) on a high-speed roll to cool the film. Depending on the rate, a second chill roll is used in some situations. After the second polishing roll or chill roll, an idler roll is used to remove the film off the high-speed roll on the way to the nip or puller rolls. After the puller, roll knives are located to slit the product to the desired width. A corona or flame treater station can be added to the cast line when the film has to have a printable surface. Both the corona and flame treatment process oxidize the nonpolar polyolefin surface to a polar surface that has better adhesion to inks and coatings. The edge trim is normally removed by a pneumatic system that conveys the film to a chopper, where it is diced and returned to the extrusion feed system or somewhere else in the plant as regrind. Cast film has better gloss, clarity, and gauge control versus blown film due to differences in the process and the rapid quenching provided by the chill roll, minimizing the crystalline size, growth, and concentration. Improved gauge control versus blown film is provided by the die and the cooling process. Cast film is produced at a higher throughput rate than blown film.

This chapter, similar to Chapter 29 discussing sheet, builds on concepts already introduced. General problems associated with mechanical or product aspects in the extruder were presented in Chapters 27 and 28. Concepts presented there also apply to cast film. Additional troubleshooting concepts presented here are centered on the process from the die to the windup; some problems associated with the die are presented in Chapter 29.

30.1 Problem 1—Molten Web Tears Easily When Exiting the Die

Assume low melt strength from using the wrong resin (too high an MFI or low molecular weight). If too much regrind or the wrong formulation is not the cause, potential causes and corrective actions are

- Polymer flow variations or fluctuations from the die can be caused by extruder surging, resulting from poor feed or other problems in the extruder (refer to problem 12 in Chapter 27 and problem 1 in Chapter 28).
- A dirty die caused by contamination near or on the die lips can create thin sections in the film due to flow restriction. Thin areas can cause the film to tear as it is being pulled away from the extruder on the polished roll. An improperly adjusted die lip opening creates thin sections in the film that can lead to holes or film tearing as it is drawn and pulled away from the die.
- Low melt temperature can limit the molten polymer flow as it exits the die and is pulled away by the high-speed rolls, causing the film to tear. Raising the melt temperature in the last extruder zones, adapters, and dies by increasing the setpoint temperatures can help alleviate this problem.
- Polished chill rolls are running too fast for the extruder throughput, pulling the molten polymer faster than it is delivered from the die.

Figure 30.2 summarizes the potential causes and corrective actions for the film tearing or breaking as it exits the die and is being laid down on the polished rolls. Other

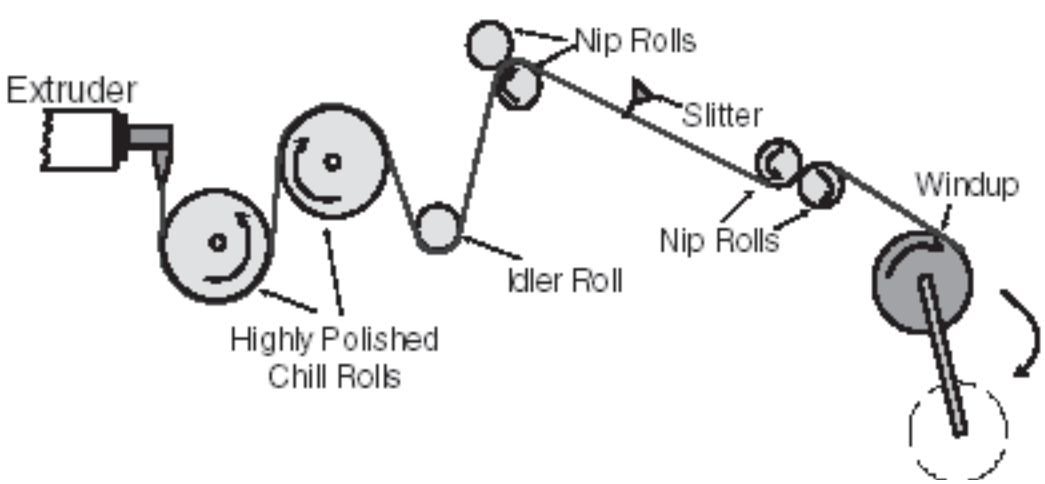

Figure 30.1. Simulated cast film process.

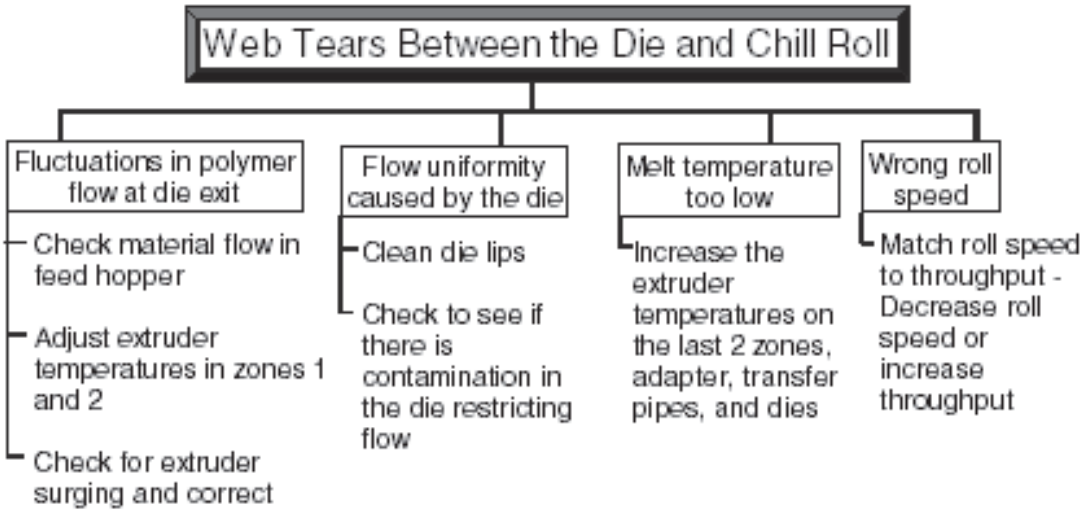

Figure 30.2. Potential causes and corrective actions to employ when the web tears.

potential causes, such as too much regrind, wrong polymer, molecular weight too low, wrong polymer formulation, and extruder surging, were covered in previous chapters.

30.2 Problem 2—Film Thickness Variation

Changes in film gauge uniformity across the roll or web over time can be caused by the die, poor melt quality, and surging. Potential corrective actions are summarized below.

- The main cause for variation in thickness across the web or in the transverse direction is the die lip opening. The die lip opening has to be parallel to the roll surface, and the adjustment bolts have to be properly set to provide uniform thickness. Dirty lips or contamination can restrict flow in particular areas, leading to gauge inconsistencies.
- Surging in the extruder and feed inconsistency can cause gauge variations. In a flood-fed extruder, the feed hopper has to be maintained at a relatively constant resin level during the run to ensure the same pressure on the resin entering the extruder. As the hopper level changes drastically, the resin actually being pulled into the extruder by the screw can vary. When producing cast film, slight variations in throughput can cause large differences in gauge. Poor flow characteristics in the feed hopper and the extruder feed section that are associated with regrind can create major gauge problems.
- Variations in the melt viscosity, resulting from changes in the melt temperature, regrind levels, or formulation changes, can cause dramatic changes in the film gauge. Hot or cold spots in transition zones, die, or adapter may lead to nonuniform melt viscosity, creating nonuniform flow from the die. Verify that the melt temperatures across the die are equivalent. If not, increase or decrease the die temperature selectively across the die to equilibrate the melt temperature and polymer viscosities.

Figure 30.3 summarizes the potential causes and corrective actions to employ in the event gauge nonuniformity occurs across the sheet or over time.

30.3 Problem 3—Lines, Streaks, and Foreign Contamination

Causes of film contamination, streaks, and lines are similar to those found in sheet. They are normally associated with contamination, poor mixing, or equipment damage. Potential causes and corrective actions are outlined below.

- Lines in the machine direction are normally caused by contamination in the die (degraded material, remnants from a blown screen pack, foreign material fed with polymer that can't pass through die opening, and so forth), a scratch or burr on the die lips, die lip build-up, a scratch on the roll surface, or something dragging on or contacting the film. To rectify die problems, the die must be purged; if the problem still exists after purging, the die must be disassembled and the die lips resurfaced to remove scratches and burrs. Causes for lines due to equipment damage outside the die have to be identified and the damaged equipment replaced or resurfaced to eliminate the problem. Die deposits

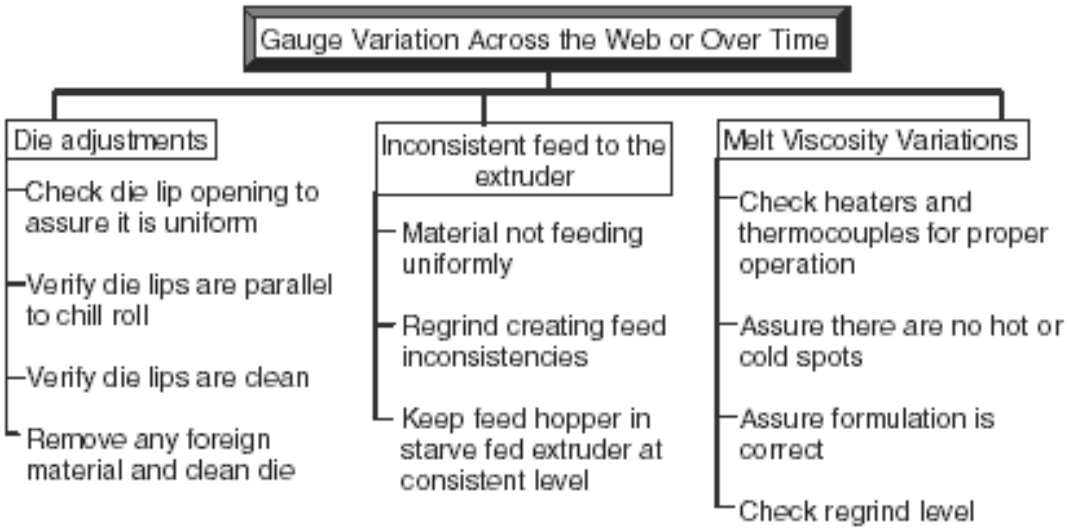

Figure 30.3. Potential causes and corrective actions for gauge variation.

have to be removed periodically to keep them from building up and affecting the film quality.

- Black specks can be generated over time if the process is running too hot and/or material is hanging up in the die and degrading. Black specks can result from shutdown and start-up operations when polymer left in the die or extruder is allowed to cool and degrade. Once degraded polymer is generated, the die either has to be disassembled or purged. Black specks normally appear at the beginning of a run.
- Contamination in the film can be coming from the feedstock, particularly if regrind is being used to supplement the feedstock. In polyethylene, gels (high-molecular-weight particles or cross-linked particles that don't melt) can create imperfections in the film. If gels are present, attempts are made to filter them out with fine screen packs. This approach is not always successful, as small gels may soften, elongate, and pass through the screens. It may be necessary to change the resin source to totally eliminate gels. High temperature in the die, combined with dead or stagnant areas in the die (behind deckles), can cause PE to cross-link and form gels.

Figure 30.4 summarizes potential causes for lines, streaks, or contamination in film with corrective actions to use in each situation.

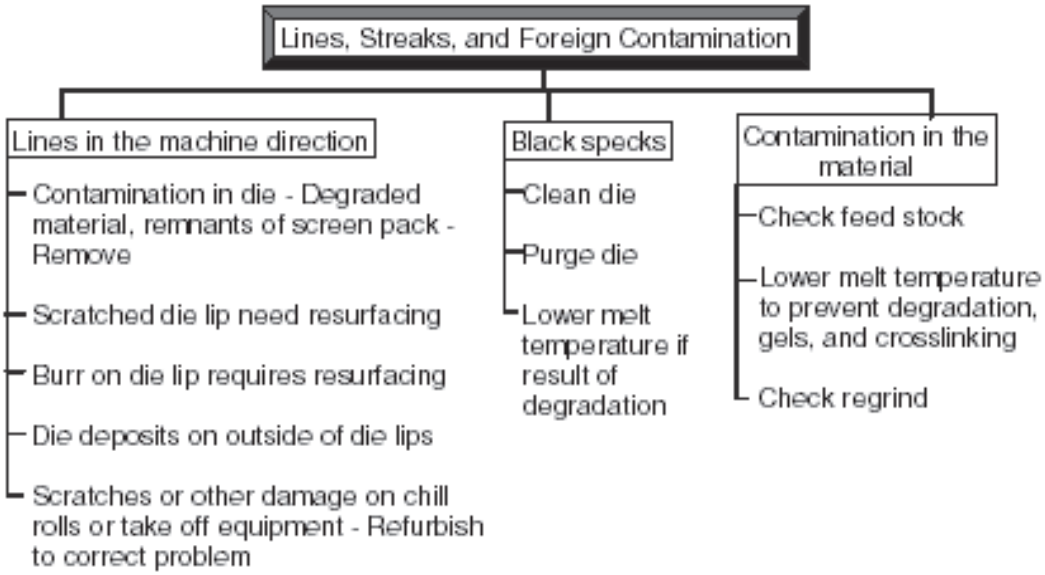

Figure 30.4. Potential causes and corrective actions for lines, streaks, and foreign contamination.

30.4 Problem 4—Extrudate Width Is Too Narrow

Potential causes and corrective actions for the film being too narrow are given below.

- The die width is too narrow to produce the desired film width or the deckles are not properly positioned. Film being laid down on a chill roll is drawn by the roll, requiring it to neck down and become narrower than the die width. When deckles are used, their length must account for the extrudate necking, while providing a little extra width on each side to provide for edge trim.
- Slow the chill roll to reduce the draw and reduce the necking. This change has to be accompanied by a change in the die lip opening or the film gauge will be incorrect.
- The polymer rheology is incorrect for the process being run. Melt viscosity is too low, generating too much neck down as the polymer is pulled away from the die by the chill roll. To increase the melt viscosity, lower the melt temperature or change to a lower melt flow index resin.

Figure 30.5 summarizes the potential causes and corrective actions to use if the film is too narrow.

30.5 Problem 5—Wrinkles Are Formed in the Film

Potential causes for wrinkles in the film and corrective actions to eliminate them are outlined below.

- The winding tension is too high, pulling the film across the polished casting or laydown roll. This can be caused by the winder and/or the puller speeds being too high. The problem is easily corrected by adjusting the puller and/or winder speeds and synchronizing the speeds with the chill roll.
- Fluctuations in polymer flow due to surging, blockage in the die, or improperly set die lips cause the polymer exiting the die to be thick and thin. This can cause extrudate sagging with excessive material that leads to wrinkles. Problems are easily corrected by cleaning the die, resetting the die lips, or eliminating the surging.
- Nonuniform cooling on the roll related to curl or warpage in thicker materials can cause the film not to track properly and wrinkle. Temperature differentials across the roll surface have to be within 2˚F (1˚C) to provide uniform cooling.
- Chill roll and die misalignment creates molecular orientation, leading to curl or twist normally associated with warpage in thicker parts. In film, wrinkles are created when the film does not track uniformly on the rolls. A casting unit is normally installed on tracks to ensure proper alignment between the rolls and the die.

Figure 30.6 summarizes the potential causes with corrective actions to use in the event wrinkles occur in a film operation.

30.6 Problem 6—Roughness on Film Due to Melt Fracture

A rough film surface similar to a sharkskin appearance is attributable to high friction in the die lip area. Potential causes for film roughness and corrective actions to eliminate it are outlined below.

- The problem can be eliminated by opening up the die gap and raising the die temperature. Increasing the die lip opening requires the line to be run at a higher speed with more molecular orientation to

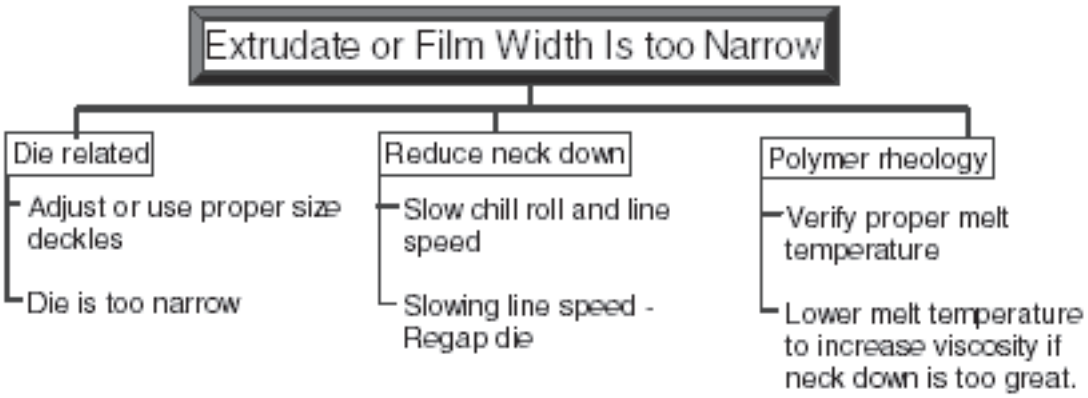

Figure 30.5. Potential causes and corrective actions if the film is too narrow.

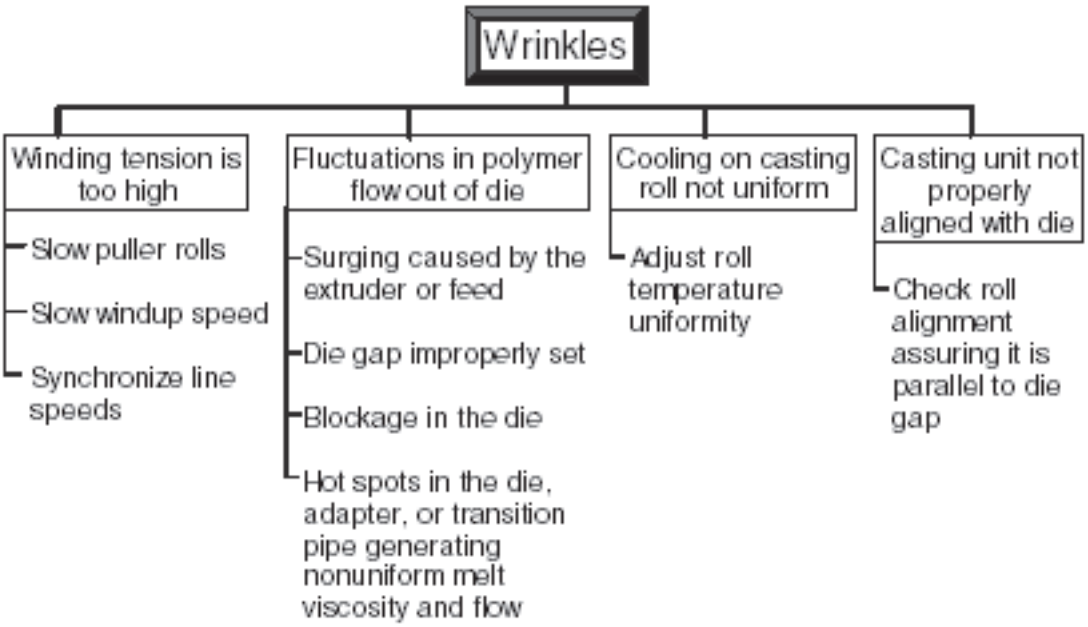

Figure 30.6. Potential causes and corrective actions in the event of wrinkles.

obtain the same gauge. This may make the solution to the problem impractical. Increasing the die temperature, attempting to heat up the die lip area, also may be impractical. New die lips with a smoother finish may be required to reduce the drag in the die lip area.

- Another approach is to increase the melt temperature to lower the resin viscosity, allowing the polymer melt to flow better through the die lip opening.

Potential corrective actions for melt fracture in films are shown in Fig. 30.7.

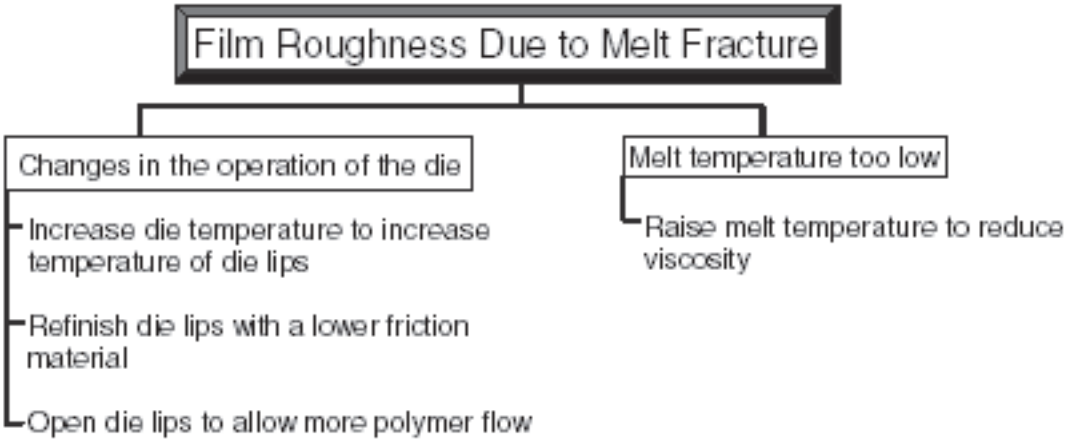

Figure 30.7. Corrective actions for film roughness attributed to melt fracture.

Review Questions

1. What causes film to wrinkle and how can each cause be eliminated?
2. Film is easily tearing between the die and casting unit; what corrective actions ought to be initiated to eliminate the problem?
3. What are some potential sources of lines or foreign contamination in film?
4. How can black specks be eliminated from film?
5. The film being produced cannot be properly trimmed because it is too narrow; what are some potential actions to widen the film?
6. What causes gauge variations in cast film and how can they be eliminated?

31 Troubleshooting Blown Film

Blown film is produced using an extruder in combination with a blown film die and cooling tower to cool film bubble. A small blown film line showing the extruder, die, and blown film from Rulli Standard[1] is pictured in Fig. 31.1. This is a small line compared to the larger production facility shown with the large tower and windup from Macchi[2] in Fig. 31.2. Blown film as a process is covered in more detail in Part 7, "Extrusion Processes." This chapter addresses troubleshooting problems in the blown film process centered on the die and downstream equipment. Chapters 27 and 28 focused on troubleshooting mechanical and product problems associated with the extruder.

31.1 Problem 1—Bubble Instability

Bubble instability is a process instability characterized by an unstable frost line in polyethylene film. Molten polyethylene exiting the blown film die is transparent, as no crystalline structure is present in the molten film. A short time or distance after the film exits the die, the resin starts to solidify and crystallize. As the resin crystallizes, the film changes from transparent to translucent at a point above the die called the frost line. Below the frost line, the resin is in its molten state and is amorphous, while above it the resin is solidifying and crystallizing. The frost line in a stable process is at a constant height, as it is controlled by the throughput rate, the film thickness, and the cooling rate. In a particular operation, these parameters are constant, providing a stable process with a stable bubble. There is no frost line visible on the bubble in Fig. 31.1 because the process is running an amorphous resin. Potential causes for bubble instability with corrective actions are summarized below.

- The melt quality, as measured by the melt temperature and pressure uniformity exiting the die, is not constant. Variations in the melt rheology or viscosity affect the polymer flow properties and the way the product is drawn as it exits the die. Verify that there are no hot spots in the die, melt transfer pipes, or adapters causing resin overheating and lowering the melt viscosity. A worn extruder screw can result in poor melt quality.
- Obstructions in the die caused by either improper die lip adjustment or degraded polymer can result in nonuniform flow out the die. Assuming there is a die flow restriction in one spot causing higher resin flow in another location, the frost line and cooling will differ from one part of the bubble to another. Corrective action is to remove any contamination from the die and verify that the die gap is uniform around the entire die.

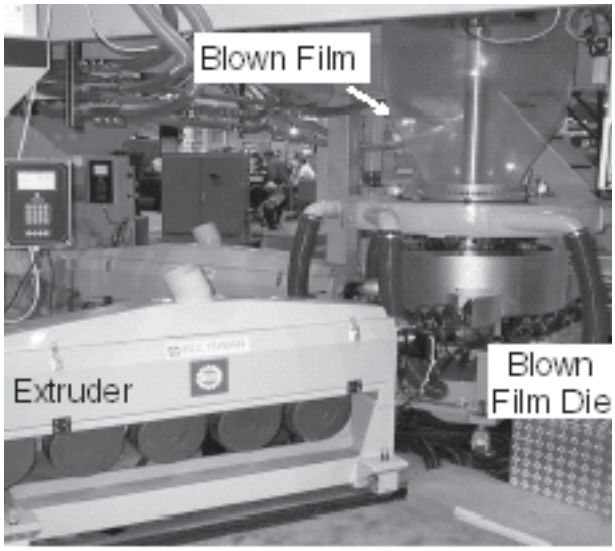

Figure 31.1. Rulli standard blown film line.

Figure 31.2. Macchi blown film tower.

- The air ring around the die providing cooling air has to be properly located and provide uniform air flow for cooling. Excessively high air velocity can move the bubble, resulting in instability as it is being pulled up to the nip at the top of the tower. A reduction in external air flow and cooling combined with internal bubble cooling leads to more stable bubble formation.
- The forming frame has to be properly aligned over the die center and set to collapse the bubble uniformly at the top of the tower.
- Low blow-up ratios (bubble-to-die-diameter ratio) can lead to bubble instability. Blow-up ratios of 2:1–3.5:1 provide a more stable process with better film properties (more isotropic) than lower blow-up ratios. Lower blow-up ratios have higher tensile properties in the machine direction compared to the transverse direction.
- Very high melt temperatures can cause bubble instabilities because the increased cooling distance allows the bubble to be more unstable.

Figure 31.3 summarizes the potential causes for an irregular frost line and bubble instability with some corrective actions to implement in each situation.

31.2 Problem 2—Gauge Variation Around the Circumference

Changes in film thickness or gauge around the bubble circumference can be caused by factors ranging from poor melt quality and surging to die and air ring misalignment. Potential causes and corrective actions are summarized below.

- The main causes for gauge variation are an improperly set or misaligned die gap, a dirty die, or a misaligned air ring. Check the die and air ring alignment to verify that the air flow around the bubble is uniform and the polymer flow rates at all points around the die circumference are equivalent. The die must be centered under the nip roll and collapsing frame to ensure uniform draw around the bubble circumference. Assuming the die gap is properly aligned and set and the polymer flow is not uniform, verify that the die does not have degraded polymer restricting the flow. In severe cases, it may be necessary to disassemble and clean the die to obtain uniform flow. Prior to disassembling the die, a purge material should be used to attempt to dislodge or clean the die.
- Poor or nonuniform air flow from the air ring produces nonuniform cooling, affecting both the draw and product orientation. This can allow part of the film to be pulled more than another, creating a

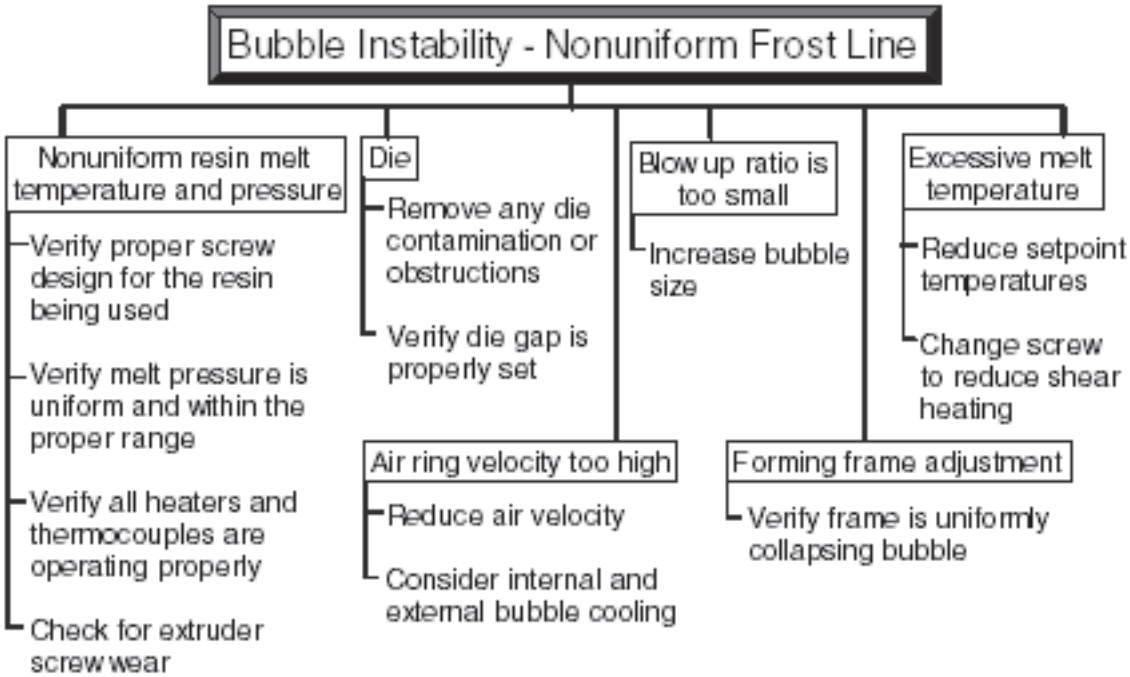

Figure 31.3. Potential causes and corrective actions for bubble instability.

gauge variation. Make sure the air ring is centered around the die, and clean the air ring if the flow is not uniform.

- Like bubble instability, poor melt quality where variations in melt temperature or pressure are present can cause throughput variations from pressure and resin viscosity changes associated with temperature. Extrusion processing conditions have to be stable and the process at equilibrium with the correct screw design to provide uniform melt temperature and pressure to the die. Pressure variations in the die over time cause changes in throughput, leading to variation in film thickness. Surging or inconsistent feed creates variations in the extruder output that are directly related to gauge variations. In a flood-fed hopper, the resin or formulation level needs to be maintained at a fairly constant level to provide a constant weight on the resin feeding the extruder. Large variations in the feed hopper can lead to minor variations in the feed rate, affecting film gauge over time. Review problem 12 in Chapter 27 and problem 1 in Chapter 28 for additional causes and corrective actions related to surging.

- Melt temperature fluctuations in the die are associated with changes in melt viscosity around the die circumference, creating differences in flow. Nonuniform melt temperatures are directly related to nonuniform viscosity, which is associated with nonuniform polymer flow from the die, creating differences in film thickness between hot and cold resin.

Figure 31.4 summarizes the potential causes leading to differences in film thickness around the bubble circumference with corrective actions.

31.3 Problem 3—Wrinkles

Wrinkles are caused by nonuniform melt cooling, improper tension from the winder, or misaligned nip rolls. Nonuniform film cooling is similar to warpage problems caused in other extrusion processes. If the shrinkage varies around the bubble, during winding the section that has shrunk more has to be drawn and elongated, or the area that has shrunk less wrinkles to eliminate the excess film over time. Other effects that lead to warpage in thicker parts cause film wrinkling in blown film. If the winder is not properly aligned, pulling more

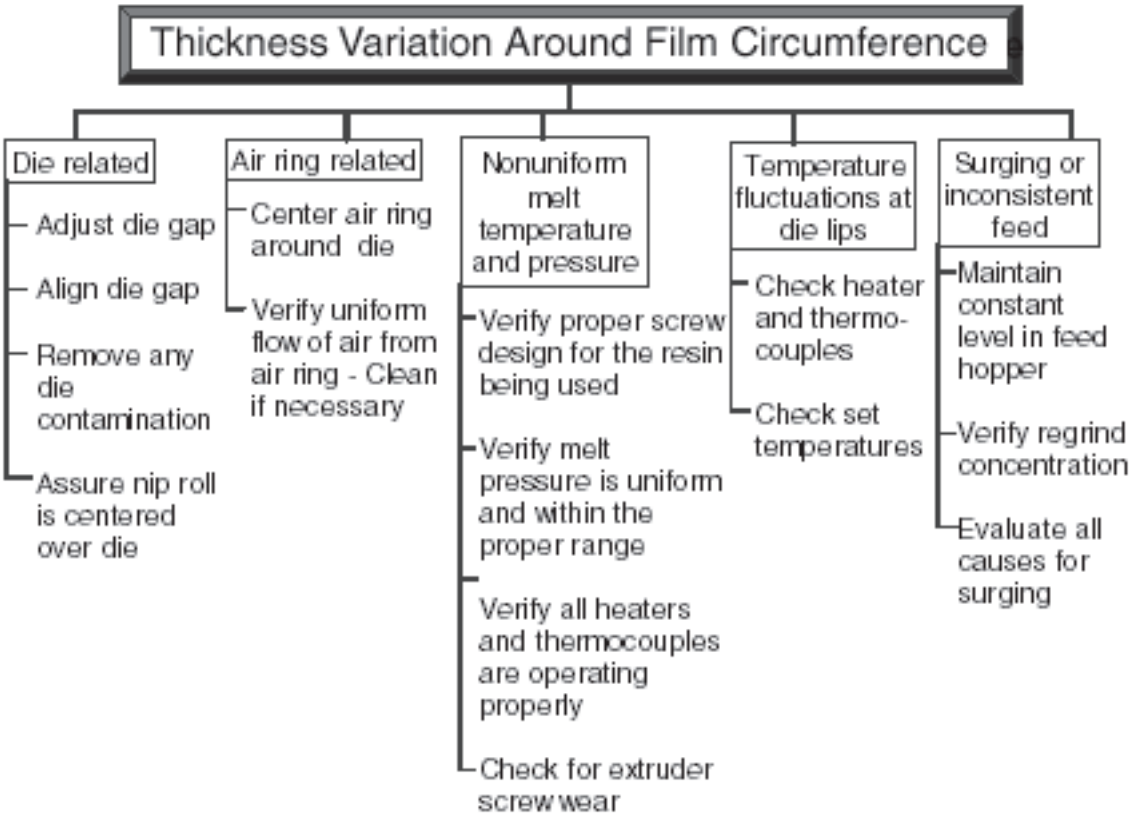

Figure 31.4. Potential causes and corrective actions for gauge variations.

tension on one side of the film compared to the other, the molecular orientation will be different, causing different shrinkage and warpage. Tension created by the puller or winder has to be uniform across the entire film web or around the bubble circumference to create uniform draw and molecular orientation. The bubble has to be properly centered in the tower between the nip and the die to prevent a higher tension or orientation on one side of the film versus the other in order to prevent differential shrinkage.

Figure 31.5 summarizes potential causes and corrective actions to implement in the event wrinkles are present. Other phenomena that create warpage in thick extrudate parts will generate wrinkles in film. The shrinkage around the circumference of the film has to be uniform.

31.4 Problem 4—Lines, Streaks, or Film Contamination

Lines, streaks, contamination, and black specks were covered under the troubleshooting sections for both sheet (Chapter 29) and cast film (Chapter 30). The potential causes for these defects in blown film are very similar to those in sheet or cast film.

- Contamination, foreign material, or black specks are generally caused by degraded resin or something added to the extruder feedstock, possibly through regrind.
- Degradation in the die can be from running the die too hot, having dead or stagnant locations in the die, and shutting down or starting up the line with material sitting in the die at high temperatures for a long time.
- Lines and streaks are normally caused by imperfections in the die lip area (burr, scratch, carbon build-up) or something on the takeoff systems that is scratching the product (burr or build-up on the nip rolls or damaged equipment). Lines are less likely caused by equipment defects in blown film compared to sheet and cast film, because the product is air cooled and is completely solidified before it is pulled across a nip or contacts the forming frame or other equipment.
- Lines can be caused in blown film by weld lines where the melt is split by a spider in the die and has to recombine after the spider leg. Lines attributed to weld lines are eliminated through higher melting temperature, longer residence time after the spider, or changing to a spiral-type blown film die.
- Streaks in the product are caused by poor mixing back in the extruder or improper die alignment.

Figure 31.6 summarizes the potential causes with corrective actions to eliminate streaks, lines, and contamination due to foreign material or resin degradation.

31.5 Problem 5—Rough Film Due to Melt Fracture

The potential causes for melt fracture in blown film are very similar to those in sheet or cast film. Melt fracture, discussed previously, is created in the die lip area by high shear stress.

- It can be eliminated by reducing the friction in the die lip area through polishing the die lips, using a lower friction coating, or lowering the resin viscosity.

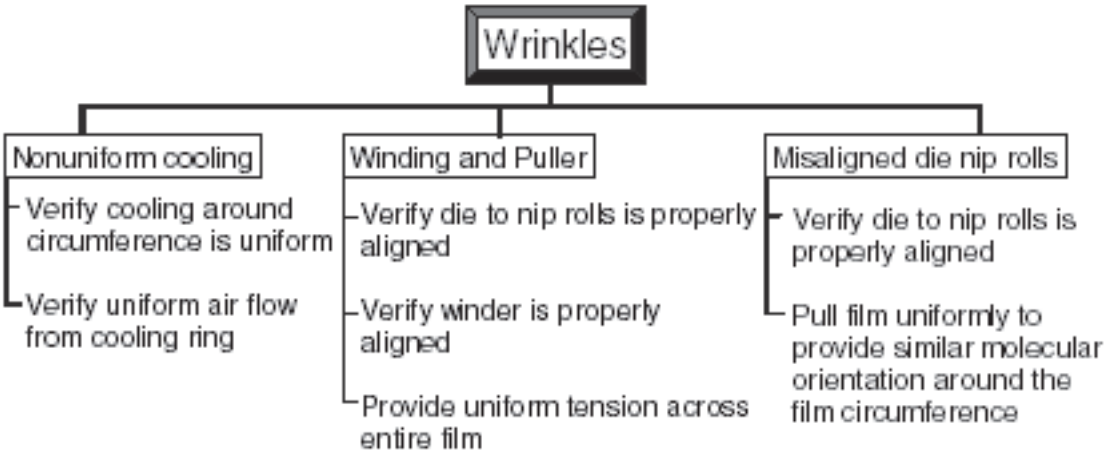

Figure 31.5. Potential causes and corrective actions to remove wrinkles.

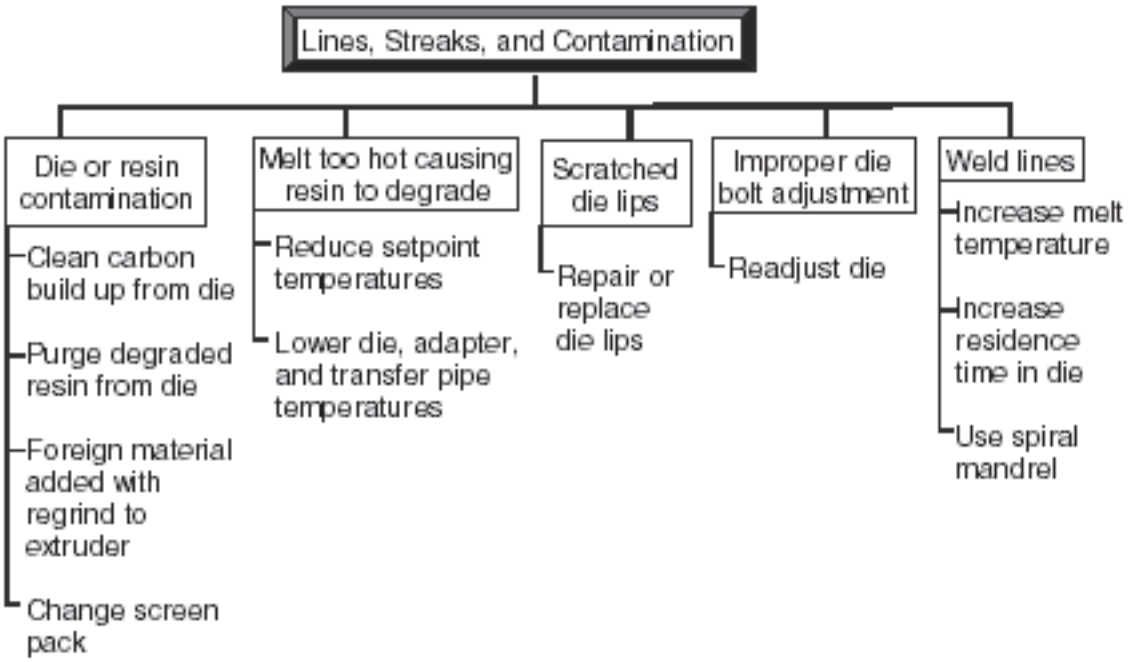

Figure 31.6. Potential causes and corrective actions to use with lines, streaks, and contamination.

- It can be caused by the die gap being too narrow. Unfortunately, opening the die gap increases the gauge, unless the line speed is increased to offset the higher throughput rate. Increasing the line speed may increase the molecular orientation in the film, unless the bubble size and lay-flat diameter are increased. Changing the die gap requires other process changes that may affect the final product properties and performance in a given application.

Figure 31.7 summarizes the potential causes and corrective actions to implement in the event melt fracture creates rough surface on the film.

PHOTO CREDITS

1. Rulli Standard, São Paulo, Brazil.
2. Macchi S.p.A., Varese, Italy.

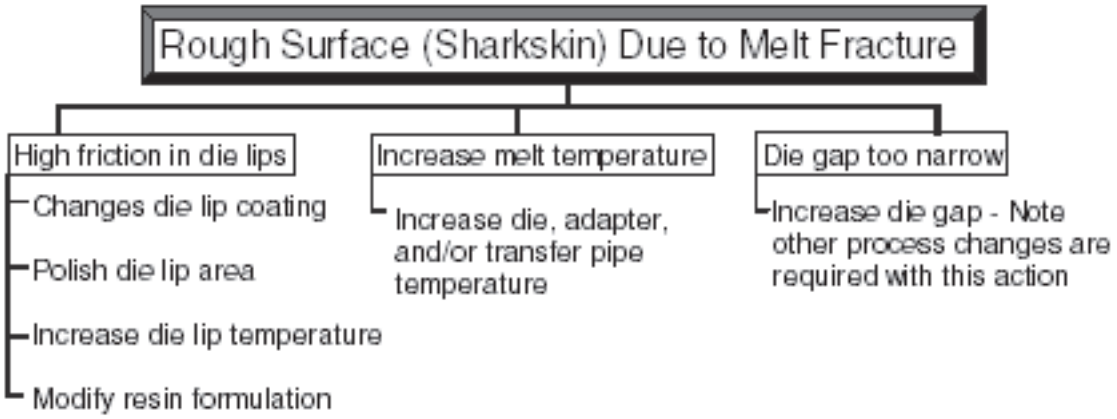

Figure 31.7. Potential causes and corrective actions in the event of film roughness due to melt fracture.

Review Questions

1. What actions need to be taken to address wrinkled film on the windup?
2. What problems can occur if the air ring is not operating properly?
3. Why is it necessary for the die to be lined up with the nip roll at the top of the film tower?
4. What actions need to be taken, and why, for rough film resulting from melt fracture?
5. Why is it important for the film to have uniform thickness around its circumference?
6. What are some potential causes for bubble instability and how is each one corrected?

32 Troubleshooting Profile and Pipe

A few potential problems exist in pipe, tube, and profile extrusion. Many of the typical problems encountered with these extrusion processes have been covered previously in Part 4 and will not be repeated here.

Pipe and tubing is a symmetrical profile extrusion. The process, shown in Fig. 32.1, has a special die to provide the desired shape, a cooling tank, puller, and saw or guillotine. The cooling tank comes in many shapes and sizes and is used to solidify the product into its final shape. For pipe and tubing, the cooling or calibration tank has sizing rings and one or more vacuum chambers to shape the pipe to the desired outside diameter.

32.1 Problem 1—Excessive Wall Thickness or Thinness

Wall dimensions that are much larger than anticipated can be caused by the wrong die lip opening, improper processing conditions, or slippage in the puller. Corrective actions for these problems are listed below.

- Profile dies do not normally have adjustable die lips to allow the die gap to be adjusted for the particular process being run. (Pipe and tubing dies may or may not have adjustable die lip openings, depending on the pipe or tubing diameter and the application.) Once the die is cut, the only way to thin the walls are to lower the screw speed and/or increase the puller speed. Speed changes are associated with a fixed molecular orientation, depending on the draw ratio required to obtain the desired product wall thickness. If the draw down ratio and molecular orientation are too high, inferior properties can result in the transverse direction; the only alternative is to cut another die with a smaller die gap. Each die gap opening is coupled with a fixed puller speed and throughput rate to produce an acceptable product with a specific draw ratio and cross sectional area. The draw ratio is defined as the die cross section divided by the final product cross section. If the die opening is twice as large as the final product, the draw ratio is 2:1. Higher draw ratios produce higher molecular orientation, resulting in greater differences between machine and transverse direction properties. If a pipe or tubing has high wall thickness on one side and low on the other side, the die bushing has to be properly centered on the mandrel to produce uniform dimensions around the pipe or tubing circumference. Nonuniform wall thickness in pipe or tubing induces warpage in the pipe, with the pipe curling or bending toward the thinner wall due to higher molecular orientation and shrinkage on the thin side. Proper die adjustments are required to produce both correct and uniform wall thickness.

- Improper processing conditions occur when the wrong draw ratio is used with the correct die gap opening. If the puller is running too slow for the extruder throughput rate, the wall thickness will be greater than anticipated. For each die gap opening, there is only one extruder throughput-to-puller speed ratio that gives the correct dimensions in the final product. This means the extruder screw speed in a flood-fed single screw extruder is nominally coupled with a given puller speed. The throughput rate in a flood-fed single screw extruder can be changed slightly at a given screw speed by changing the temperature profile and the melt viscosity. Any change in throughput rate must be accompanied by a change in puller speed to maintain thickness control. In general, higher melt temperatures lead to lower viscosity with higher resin flow, possibly generating higher throughput rates. Extruders are complex systems: as the barrel temperatures are changed, the feed characteristics may change (higher set temperatures in extruder zones 1 and 2); the conveying characteristics in the feed or conveying section might change; die pressure can be lower, reducing

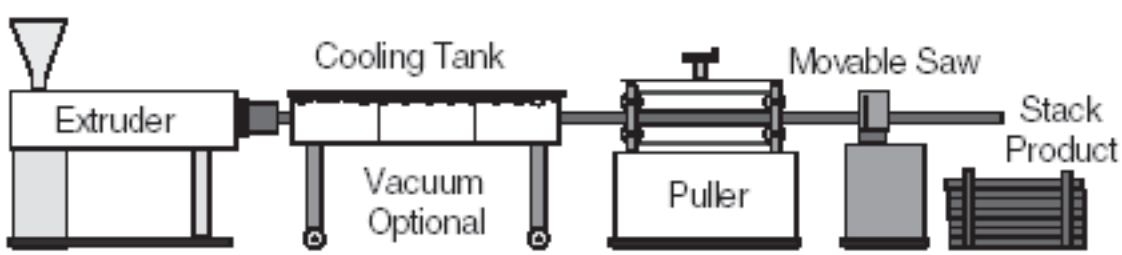

Figure 32.1. Schematic of a profile, pipe, or tubing extrusion line.

the backflow in the extruder; mixing characteristics change; and extruder pumping characteristics are different. Obviously the extruder is a fine balance of throughput rate, screw speed, and melt pressure and temperature. In absolute terms each fixed die gap opening has a fixed draw ratio based on the puller speed and extruder throughput. As the throughput rate goes up, the puller speed must be increased, and vice versa. At a given screw speed and operating conditions in a flood-fed extruder, the output rate is constant at a given melt temperature and pressure.

- Slippage in the puller allows the product dimensions to increase because the product is not being pulled away from the extruder at the proper rate. The caterpillar or puller pressure has to be adjusted to prevent the product from slipping without causing any distortion or damage to the final product. If the pipe, tubing, or profile is still wet from the cooling bath when it reaches the puller, slippage can be caused by moisture on the belts.

The same corrective action procedures outlined above are used when the wall thickness is thin, except the implemented procedures are opposite of those presented.

Figure 32.2 summarizes the potential causes and corrective actions to implement in the event the wall thickness is incorrect.

32.2 Problem 2—Wavy Surface Inside Pipe or Tubing

Sizing is normally done on the pipe or tubing outside diameter (*OD*) through the use of sizing rings or fixtures in the cooling tank. The inside diameter (*ID*) is controlled by air pressure combined with vacuum over the sizing tank to force the pipe to the final dimensions. The inside surface is controlled by the die and processing conditions. In some pipe operations, a plug is held in place by the die; as the pipe is pulled over the plug, air is used in combination with the plug to form a smooth pipe inside diameter. Viewing a hollow profile (pipe), the inside surface can be wavy and not nearly as smooth as the outside surface. Potential causes for excessive waviness on the inside surface with corrective actions are given below.

- The melt temperature is too low, preventing the polymer from flowing freely out the die lips due to the high melt viscosity. Corrective action is to increase the extruder temperature setpoints in the last one to two zones, the adapter, and the die to raise the melt temperature.
- Slippage in the caterpillar causes the profile to go through a cyclic action where the puller engages and disengages the profile. This slippage does not pull the molten material uniformly after it exits the die. One cause of slippage is a wet profile.
- The molten polymer is sticking or slipping on the die lips as it flows out the die. Changing the frictional characteristics between the melt and the die lips is necessary to eliminate the slip/stick phenomenon. An external lubricant may improve the

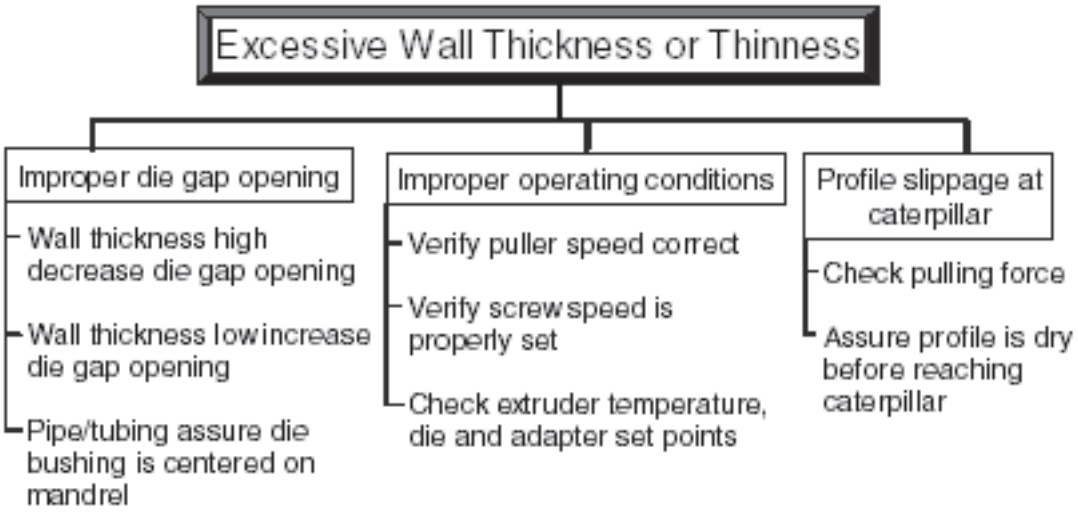

Figure 32.2. Potential causes and corrective actions for thick or thin wall dimensions.

resin flow characteristics over the mandrel. Alternately, change the coating on the mandrel or die to reduce friction.

Figure 32.3 summarizes potential causes and corrective actions to prevent a wavy inside pipe surface.

32.3 Problem 3—Weld or Knit Line Weakness

This problem has been identified and corrective actions proposed previously in problem 13 in Chapter 28. Weld lines in pipe or tubing are generated around the circumference in the machine direction by the spider legs used to center and hold the mandrel in place. If the die is not properly designed, weld lines can cause weakness in the transverse direction wherever a spider leg is located. If weld lines are a problem, the die length past the spider legs has to be increased to provide more time for the resin molecules to reentangle after they are split. Increasing the flow channel in the die after the spider legs builds pressure, encouraging the polymer molecules to reentangle. The other factor to consider is the polymer viscosity. Higher melt temperature lowers the polymer viscosity, increasing molecular motion. The polymer then has the potential for higher entanglement after the spider legs. Pipe and tubing processes create weld lines due to the dies used. Care must be exercised to ensure these weld lines have time to recombine and provide the strength required in the application.

Figure 28.15 summarizes the corrective actions required to implement improvements in weld line strength.

32.4 Problem 4—Nonuniform Resin Velocity from a Profile Die

The velocity profile from the die to each leg is critical in both simple and complex shapes. Assume polymer flow in one leg of a profile is faster than another leg; the resulting profile will be warped, curled, or twisted. In complex profiles (such as window profiles), all wall thicknesses need to have similar dimensions, with flow into each wall or leg being uniform. Internally the die needs a uniform taper into each leg or wall section to ensure uniform flow velocity and pressure generation across the entire die to produce a straight profile that meets the customer specifications. Potential causes for nonuniform die flow with corrective actions are given below.

- Legs in different sections have unbalanced flow due to the taper leading into the leg, the die land length, or die land gap dimensions. The die has to be modified or reworked to ensure uniform flow, velocity, and pressure in each section or leg in the profile.
- Flow is restricted in one leg or section by contamination, foreign material in the die, or degraded polymer. Carbon build-up in the die gap over time can restrict the flow in a given leg or section. Anything inside the die flow channel that restricts flow will affect the polymer velocity profile at the die exit. Initial corrective action is to properly purge the die; if that doesn't work, the die has to be disassembled, cleaned, and reassembled.
- Melt temperature fluctuations in the die can be caused by hot or cold spots in the die or nonuni-

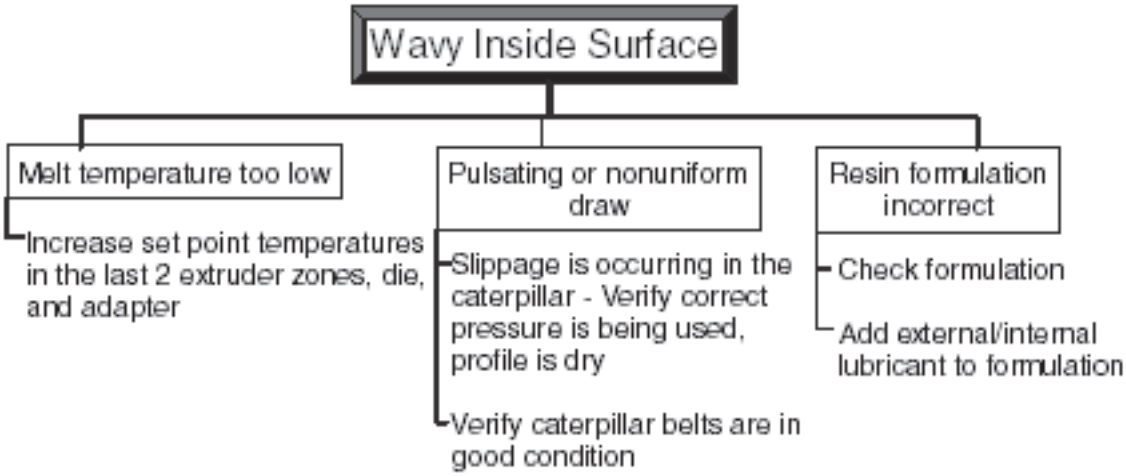

Figure 32.3. Potential causes and corrective actions for a wavy surface inside pipe.

form die heating that affect the polymer melt viscosity and polymer flow characteristics. All heater bands and thermocouples are checked to ensure the equipment is properly installed and functioning according to the specifications. If temperature fluctuations still occur after the equipment is checked, additional heater bands may be required or the die may need to be insulated to provide uniform temperature. Proper melt mixing prior to entering the die ensures a uniform melt temperature.

Figure 32.4 summarizes potential causes of nonuniform flow or velocity variations from different die sections with corrective actions to eliminate the problem.

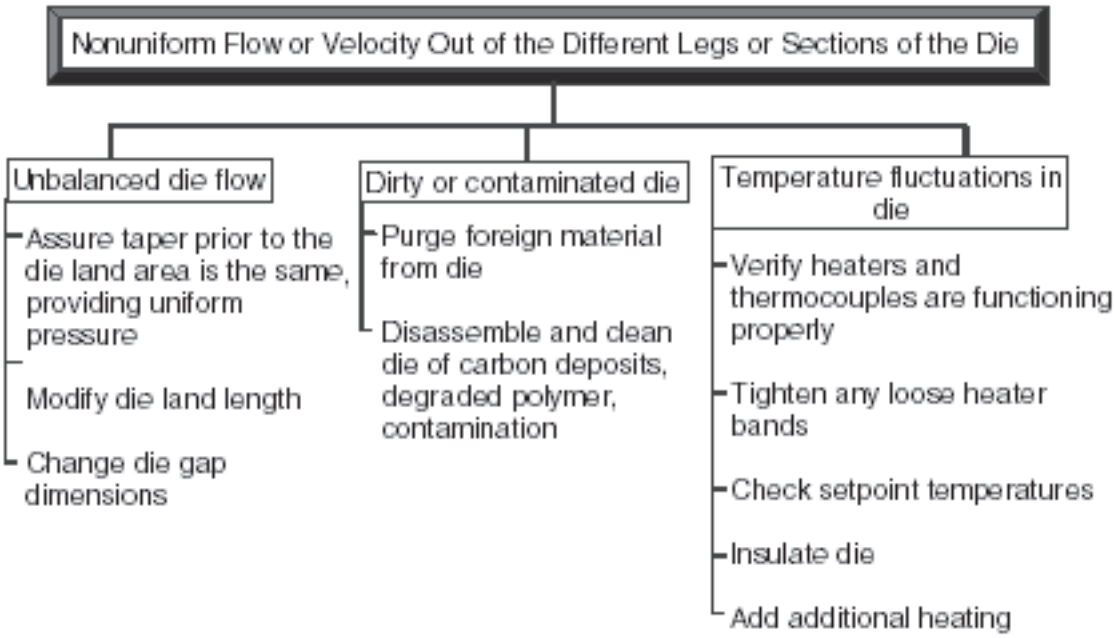

Figure 32.4. Potential causes and corrective actions for nonuniform resin velocity for flow out of the die.

Review Questions

1. What are the consequences of not having uniform flow out each die leg in a profile extrusion?
2. What are some causes and corrective actions for either thick or thin walls in profile or pipe extrusion?
3. What causes weld lines, and why might they create a problem in an application?
4. What are some methods to use to eliminate wavy surfaces on the inside of a pipe?

Part 5: Auxiliary Equipment

33 Feed Systems

Auxiliary equipment is defined as peripheral equipment that may be an integral part of the extrusion process to improve or optimize the extrusion process efficiency and ease of operation. It is not necessary to make the extrusion operation functional, but if properly designed and used, it can improve both the accuracy and precision of the operation. Optimum quality auxiliary equipment is required to run a world-class extrusion operation, producing quality products with tight tolerances in high yield. Auxiliary equipment is used to provide raw materials to the extruder in the correct ratio and concentration, with the proper moisture content, or modify the extrudate at the die end of the extruder through the use of an automatic screen changer or gear pump. Other auxiliary equipment may not be directly incorporated into the extrusion line but is an integral part of a successful extrusion operation. This equipment may include granulators, pulverizers, screw and die cleaning equipment, dryers, blenders, and purge material to use when changing from one formulation to another. Granulators and pulverizers are used to reclaim start-up scrap out of specification material and trim for reprocessing in future extrusion operations, recovering some raw material costs. Reclaimed material is also sold as regrind to other processors (extrusion and/or injection molding operations) or compact scrap for disposal. Screw and die cleaning equipment minimizes the amount of downtime required for cleaning. Appropriate purge material reduces the amount of downtime required to clean the extruder between product changes. Purge materials are particularly important during product changes using expensive resin systems. Dryers are required when processing moisture-sensitive materials to ensure the moisture content is below the critical concentration prior to feeding to the extruder. Depending on the number and type of feeders, blending equipment may be required to prepare the proper formulation or blend to be fed to the extruder.

Auxiliaries discussed in this part are general pieces of equipment that can be common to most extrusion processes. Unique pieces of equipment common to particular extrusion processes—such as a puller for profile, pipe, and tubing; a classifier for compounding; a vacuum tank for pipe and tubing; and a winder for sheet or film—are discussed in Part 7 in the particular extrusion process. Equipment covered in this part is common to all extrusion processes and includes feeders, dryers, blenders, gear pumps, screen changers, screw cleaning equipment, purge materials, chillers for water cooling the extruder or screws, and pellet conveyors to load feed hoppers or convey resin around the plant.

The first chapter in this part covers feed systems using volumetric, gravimetric or loss-in-weight feeders, liquid feeders, and blenders to premix materials off-line or directly above the extruder. The advantages and disadvantages of both volumetric and gravimetric feeders are compared.

33.1 Feed System

The feed system selected depends on many factors, such as

- Extrusion process and objectives
- Flood-fed single screw extruder or starve-fed twin screw extruder
- Extruder size or throughput
- The number of ingredients in the formulation
- Physical form of the feed (liquid, powder, pellets, and fibers)
- Feed stream bulk density (from <10 lbs/ft^3 [160 kg/m^3] to >60 lbs/ft^3 [962 kg/m^3])
- Resin moisture sensitivity
- Production run length (number of product changes)
- Physical space available
- Main feed stream concentration relative to other ingredients (additives at 0.5% or 30%)
- Cleanup time required between product changes

Evaluating the relative importance of each of these individual factors depends on the particular extrusion operation and the objectives to be accomplished. The **first factor**, concerning the extrusion process, depends on the process objectives. Is the extruder required to melt a virgin resin or precompounded pellet? To prepare an extrudate that can be formed into a final part such as film or sheet? Alternatively, is compounding part of the extrusion process where numerous feed streams are fed and melt blended in the extruder? Even if the process uses only a virgin or precompounded resin, two feed streams may be required if regrind is added to the formulation. If a precompounded or virgin pellet is simply combined with regrind, it may be possible to blend the materials off-line, transport the blend to the extruder, and add it directly to the extruder feed hopper in a flood-fed situation or with a volumetric feeder in a starve-fed extruder. If the extrusion objective incorporates compounding of colorants, lubricants, stabilizers, flame retardants, fillers, impact modifiers, or reinforcements as part of the extrusion process, either numerous feeders or preblending is required to prepare the raw materials for the extrusion operation.

The **second factor** to consider is the type of extruder being used and whether it is starve or flood fed. (Some single screw extruders are starve fed with different feeders, but most are flood fed.) In flood-fed extrusion, individual feeders can be used to make small batches as part of a gravimetric or volumetric blending system. The components are mixed and let down to the extruder in a continuous operation. Components from individual feeders are added to a small mixing chamber in specific weights or ratios to produce the desired formulation where they are mixed. Upon completion of the mix, it is delivered to the extruder, and a subsequent mix is started in a semicontinuous batch process. The feeders adding resin to the common mix chamber are either volumetric or gravimetric feeders. The mix chamber can be either a continuous or batch operation. In a continuous mix operation, materials are constantly fed into the chamber where an auger or cascade system is continually mixing the ingredients as they pass through to the feed hopper. In a batch operation, the ingredients are metered into a chamber where they are contained while an agitator or auger mixes the components for a predetermined period of time prior to dumping them into the feed hopper. The feed hopper size depends on the size of the mix. In a pellet/pellet mix, the mix can be relatively large where the separation of the blend into individual components prior to the extruder is unlikely to occur. However, in a pellet/powder mixture, the pellets and powder can separate; therefore, the quantity of material in the feed hopper directly above the screw in a flood-fed extruder should be relatively small, minimizing the potential for separation. This ensures the correct ratio of ingredients is being fed into the extruder. In starve-fed extrusion, individual feed streams are added directly to the extruder feed throat, where the screw rotation conveys the material forward into the extruder as soon as it contacts the screw. In this scenario no ingredient premixing is required (assuming there is a feeder for each component in the formulation), and the potential for ingredient separation in the feed hopper is eliminated.

The **third factor** to consider in selecting the proper feed equipment is the size of the extruder and the anticipated throughput rate. A 2.5-inch single screw extruder with a nominal throughput rate of 250 lbs/hr requires different size feeders than an 8-inch single screw extruder with a nominal throughput of 8000 lbs/hr. The throughput rate determines the following feeder factors:

- Hopper size
- Feeder size and speed
- Method of delivering resin to the feeder
- Space required above the extruder and feeder to install the delivery system for transferring powder/pellets to the feeder and from the feeder to the extruder

The **fourth factor** to consider is the number of components in the formulation. Assume a formulation with four ingredients—resin, colorant, lubricant, and thermal stabilizer—is being compounded as part of a profile or pipe extrusion line. What is the best method of feeding these ingredients to the extruder? The available options go from four individual feeders in a starve-fed situation to preblending the four components and adding the blend to the feed hopper in a flood-fed extruder. The sophistication selected depends on the run length and the available equipment. With four individual feeders, each feeder has to be sized properly and be the correct type to feed the materials being used. As an example, a single screw feeder feeds free flowing materials such as pellets very accurately, while a twin screw feeder is normally better for powders. Each feeder has to be properly sized for the desired feed rate per hour. In the example above, the formulation may be 95% resin, 3% color concentrate, 1.5% lubricant and 0.5% thermal stabilizer. If the total extruder throughput is 1000 lbs/hr, this requires 950 lbs/hr of resin, 30 lbs/hr of color concentrate, 15 lbs/hr of lubricant, and 5 lbs/hr of thermal stabilizer. The same feeder used to feed the resin at 950 lbs/hr is not appropriate to feed 30 lbs/hr of color concentrate, and the feeder used for the color concentrate may not be appropriate for the lubricant and thermal stabilizers if the later two additives are powder and the color concentrate is pelletized. Another way to feed this particular formulation to the extruder is to preblend the lubricant and thermal stabilizer and add the blend with a small powder feeder at the rate of 20 lbs/hr. A second feeder is used to feed the color concentrate in the correct ratio and a third feeder to add the resin. Another approach is to precompound the lubricant and thermal stabilizer in high concentration with the base resin to produce a concentrate or masterbatch. The concentrate is then mixed in the correct ratio with the color concentrate and added with one feeder, and the main resin is added with a second, large feeder. The relative proportions of each component in the blend depend on the concentrations of the thermal stabilizers and lubricants in the precompounded concentrates. Using a flood-fed extruder, the color concentrate and precompounded thermal stabilizer and lubricant concentrates can be mixed with the resin in a large blender off-line and the mixture air conveyed to the resin feed hopper. In this scenario no feeders are required.

The **fifth factor** to consider in determining the proper feeder is the physical form of the individual components in the formulation. Typically there are five physical forms, with a range of bulk densities for each form. The five physical forms are pellets, powder, liquid, fibers, and regrind (irregular granules in both size and shape). The different feed characteristics for powder, fiber, and pellets may require different feeder designs to provide uni-

form product flow. Designs available include single and twin screw auger, vibratory, and belt. Free flowing pellets, powder, and regrind can be uniformly fed on any of these three feeder designs. Nonfree flowing powders feed best using either a twin screw auger or belt. Powder feeding using belts can be dirty, as it is harder to contain the powder and dust than with auger-type feeders. Due to the compressibility, flowability, and bulk density of various powder ingredients, feeder trials are sometimes required to determine the best feeder options for a particular powder. Fibers are frequently fed with vibratory feeders to minimize fiber attrition. However, the nature of the flowability, that is, whether or not it is free flowing, and feeding precision will determine the appropriate feeder design. Liquids are added via a liquid feed pump at high pressure through an injection nozzle if added downstream in either single or twin screw extrusion, or by metering the material into the feed throat with a constant speed pump. (Liquid is not normally added to the feed throat of an extruder because it can interfere with the feed characteristics of the other components in the formulation, creating a feed problem. If a liquid additive is required in single screw extrusion, it is better to close the vent and pump it into the melt in the second stage of a two-stage extruder. Additional mixing may be accomplished with a mixing head on the extruder screw or with a static mixer in a transfer pipe between the extruder head and the die.) Liquid can be fed either gravimetrically or volumetrically, depending on the type of feeder available. (This is discussed in more detail later in this chapter under volumetric and gravimetric feeders.)

A **sixth factor** to consider when choosing the proper feeder is the additive bulk density. Assuming the bulk density is about 60 lbs/ft^3 (962 kg/m^3), feeding characteristics of free flowing materials are not affected by the material bulk density. However, if the bulk density drops below 20 lbs/ft^3 (320 kg/m^3), feeding issues may arise. The hopper size may have to be increased to be able to hold sufficient raw material to prevent constant hopper refilling during operation. If the bulk density is very low, powder tends to bridge in the feed hopper, and a special hopper or agitator may be required to eliminate bridging. Larger augers may be required in screw feeders to deliver the required rate. Low bulk density materials do not normally feed well with vibratory feeders.

A **seventh factor** to consider when selecting the feed system is the resin moisture sensitivity. Hygroscopic resins that require drying prior to extrusion, such as PET, PBT, PC, PA (polyamide or nylon), and ABS, need to be transferred from the dryer to the feed hopper using dry air and maintained in a dry air environment until being extruded. Hydroscopic resins that have been dried and transferred to an open feed hopper or extruder hopper will pick up moisture from the air. One solution is to blanket the resin hopper with dry nitrogen. A second alternative in large-scale production is to transfer the pellets from the dryer to the feed hopper pneumatically with dry air and then blanket the pellets in the feed hopper with dry air. A third approach used in single screw extrusion is to use a hopper dryer as the feed hopper. Resins previously dried in a remote plant location are conveyed with dry air to the hopper dryer, where they are maintained at temperature in a dry environment until extrusion. Anytime hygroscopic resins are extruded, special precautions are required to ensure that the product is properly dried and remains dry until it is extruded. In certain resin systems (PET, PBT, PA, PC, and so forth) moisture can have very detrimental effects on physical properties or the extrudate surface appearance.

The **eighth factor** to consider is the production run length. Short runs with numerous product changes require easy cleanup between runs to minimize change over time. A large part of the time involved in product changeover and cleaning between runs is the ease of disassembling, cleaning, and reassembling the feeder. Feeder accessibility can influence the time required for these tasks. Easy accessibility and adequate room around the feeders can facilitate cleaning and save time. Included in cleanup is cleaning the hoppers, transfer pipes, and the area around the feeders. During facility design, ensure that sufficient room is available to dismantle the feeders, water is available for cleaning, and proper drains for water removal are present (assuming the feeders will be washed with water). High-volume, long production runs use large equipment; cleaning between runs is just as important. Minimizing changeover time helps to improve operating efficiency. When purchasing feeders, consider both the ease and time involved in cleaning the feed systems and changing from one product to another, as this can be an integral part of the total production time, particularly with short production runs.

The **ninth factor** in properly selecting and designing a feed system was discussed briefly under factor eight. Ensure that there is adequate physical space around the feeders to work on them. This makes the job significantly easier during operation, product changeover, and routine maintenance. Adequate space aids housekeeping and minimizes cross-contamination. Putting feeders into an inadequate space creates production headaches that affect operation efficiency.

The **tenth factor** to consider when choosing the proper feed system is the feeder size. This was mentioned under factor three, dealing with the extruder size. Each feeder has a desired operating range. As an example, assume a feeder is being used with a throughput range of 0–500 lbs/hr (0–227 kg/hr), and the feeder is operating at 30 lbs/hr (13.6 kg/hr). The feeder precision suffers because it is running very slowly. The screws and dis-

charge chute size can be changed in a screw feeder to lower the overall throughput rate while maintaining the screw speed. With a belt feeder, the discharge gate from the hopper can be closed to increase the belt speed; this assumes the ingredient being fed flows freely through a smaller gate opening. Feeders are more accurate when they operate closer to their maximum throughput. A feeder operating near its maximum rate needs to have a large enough feed hopper to prevent constant refilling. The feeder has to be properly sized for the required throughput.

The **last factor** to consider in designing and selecting a feed system is cleanup and changeover time from one product to the next; this has been discussed in factors eight and nine above.

33.2 Blending Systems for Single Screw Extrusion

Single screw extruders are normally flood fed via a hopper above the feed throat that is filled to a controlled level. This improves feeding, as there is a constant downward force on the extruder screw, as shown in Fig. 33.1. If the extruder feed hopper level is allowed to cycle from almost empty to full, the throughput rate will change slightly with the resin force on the extruder screw. This is accompanied by a change in the product dimensions. In a flood-fed extrusion setup, the material level in the feed hopper needs to be kept at a fairly constant level. The level is kept constant by incorporating a vacuum loader, Fig. 33.2, on top of the feed hopper that adds material to the feed hopper during operation. With small extruders, material can be added manually using buckets. However, as the throughput increases, maintaining a constant feed hopper level manually with buckets is not practical. One alternative approach is to place a large storage bin above the hopper that maintains the hopper level, as shown in Fig. 33.3.

In single screw extrusion, individual feeders may not be used, as all formulation ingredients are premixed and added to the extruder hopper, as shown in Figs. 33.2 and 33.3, or with buckets. Depending on the batch size and the ingredients (pellet, powder, liquids, or fibers), preblending or mixing can be accomplished elsewhere and transferred to the extruder. Off-line blending systems include tumble blender, drum roller, cement mixer, paint shaker, ribbon blender, twin cone blender, vertical blender, continuous blender, and high intensity mixer. The ingredients are mixed for a specific time to uniformly disperse the ingredients throughout the blend. The blenders noted above are all batch operations; depending on the way components are weighed and added to the blenders, the potential for batch-to-batch variation exists.

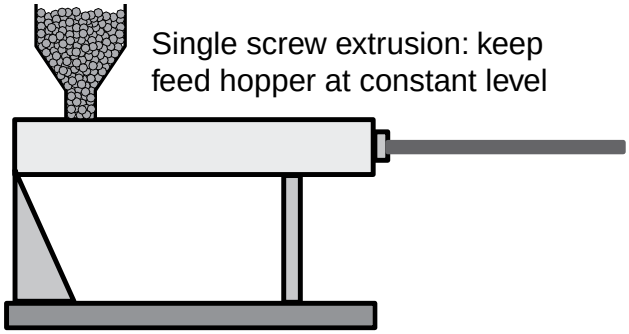

Figure 33.1. Single screw extruder feeding.

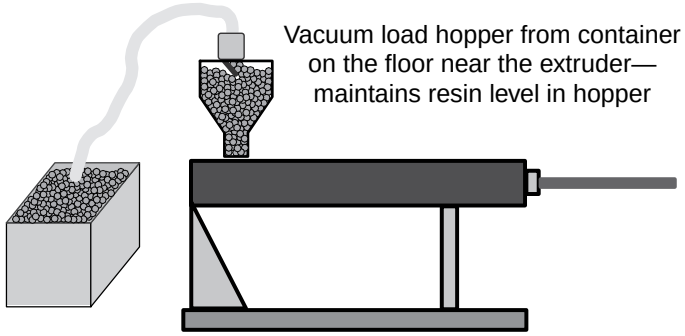

Figure 33.2. Vacuum loader to maintain constant hopper level.

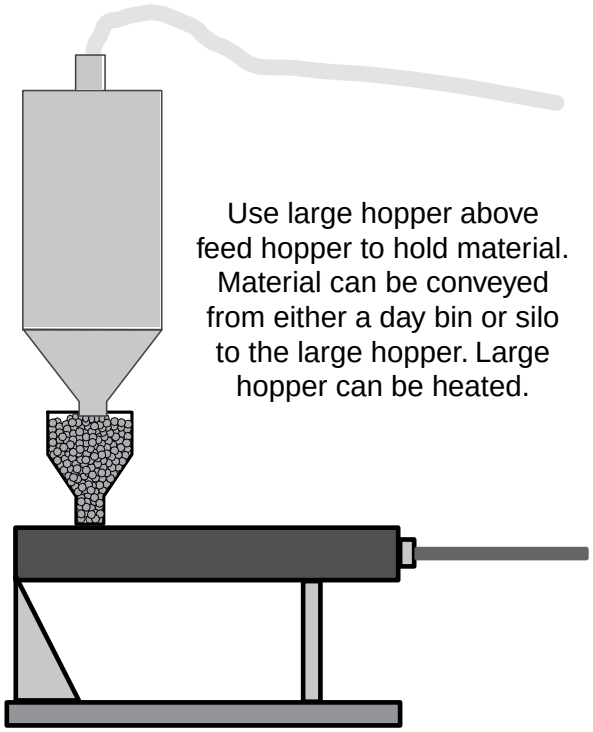

Figure 33.3. Extruder hopper fed by large hopper.

Potential causes for error in off-line blending occur through weighing errors, separation as the blend is transported to the extruder, and contamination resulting from improper blender cleaning prior to adding new ingredients. In a pellet/powder mixture, the powder can segregate from the pellets, resulting in the material on the top being richer in pellets and the bottom richer in powder. After transfer of a powder/pellet blend to the extruder feed hopper, additional separation is likely to occur as the blend flows into the extruder. Powder can filter through the moving pellets and fall to the bottom of the feed hopper, causing the initial extrudate to be rich in the powder component and the final extrudate to be rich in the pellet component. Pellet/regrind blends can separate during transport if there is a large particle size difference between the pellets and regrind. Once the blend is moved to the extruder, additional separation can occur, depending on how the mixture is transferred to the extruder feed hopper. The advantage of off-line blending is reduced equipment cost. Assuming blending is done at a higher rate than the extrusion operations, one blending station can supply numerous extruders. In computing the blending time, the cleanup time from batch to batch has to be factored into the equation to ensure that sufficient material can be blended for the number of extrusion lines requiring blender output.

Tumble blenders can blend small quantities in five-gallon pails, up to 1000-pound batches. Large tumble blenders are also used in compounding operations to postblend pellets after compounding to improve lot uniformity. Figure 33.4 shows a large, rotating, double cone tumble blender. The ingredients are weighed, added to the blender, and tumbled for a set time to provide a uniform mixture. This low intensity mixing generates little heat and does not affect the particle size as the blender rotates about its axis. The double cone can be jacketed to provide heat during mixing to dry mixtures or postdry compounded product. Jacketed tumble blenders, as shown in Fig. 33.4, can be used to dry resins prior to extrusion. Vacuum assist during the drying cycle increases drying efficiency while reducing the time. One end of the tumble blender has a slide gate to add the components in the mixture prior to mixing and to remove the blend once the cycle is complete. This is an excellent system for large-scale low intensity blending, both prior to and after compounding extrusion. Five- to 50-gallon tumble blenders use pails or drums to mix the formulation and are used extensively in small extrusion operations and for R&D to mix small batches. In small operations, any drying requirements are accomplished in another step, either prior to or after blending.

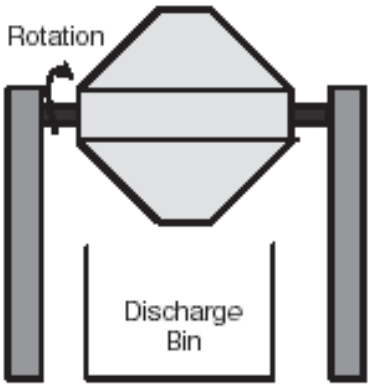

Figure 33.4. Large tumble blender

Cement mixers can be used in production to blend small batches near the extruder. These low intensity mixers are excellent for blending pellets and/or regrind. They are easy to load and unload, inexpensive, easy to clean, and readily available. Mix times are short and uniform pellet dispersion is easily obtained.

Drum rollers are low intensity mixers used for small batch blending. Ingredients are added to a drum, fiber pack, or bucket and placed between two corotating rollers. As the rollers turn, the drum or cylindrical container is rotated by the two driven rotating rollers. The drum is left on the rollers until the contents are properly mixed.

A **ribbon blender** is a low intensity mixer that can be jacketed for heating during blending. Normally a ribbon blender is made from a semicircular stainless steel tank with high sides and a set of rotating blades or paddles inside that mixes the formulation components. The ribbon mix blades come in a double helical configuration, shown in Fig. 33.5, which moves the material back and forth in the blender during the mixing operation. While the double helix is the standard configuration and will mix most materials, other blade designs are available for specific applications. Plow-style blades have plow-type feet on the end of individual shafts connected to the center shaft that push or plow through mixtures. Plow blades are particularly effective with materials that have poor flow characteristics or are sticky or tacky. Paddle-style blades are similar to plow blades except the plow is replaced by a paddle at the end of the shaft. Paddle blades are used with dense materials, slurries, and abrasive mixtures. Preweighed components are

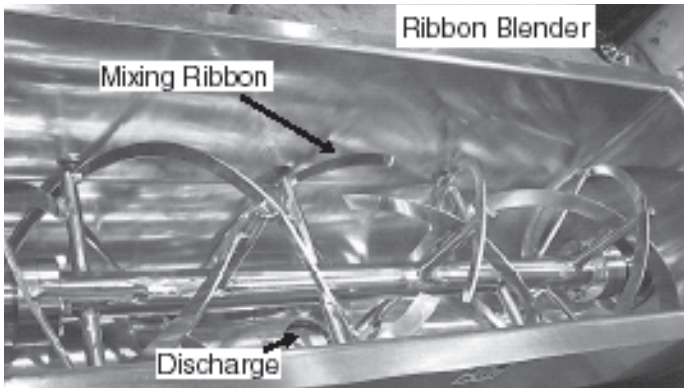

Figure 33.5. Top view of Ross ribbon blender.

added to the blender and mixed for a set time. After mixing is complete, the blend is discharged either through a center bottom port at one end or through multiple gates. Different discharge gate designs are available that are pneumatically or manually activated. The center shaft is driven by either a chain and sprocket (large blenders) or an in-line drive where the motor is coupled directly through a speed reducer. After the mix time is completed, the discharge port is opened and the rotating ribbon or blades force the material out through the blender discharge. The mixing ribbon mixes by moving material from the center out to the end and back to the center. Blenders are available from 20 pounds to thousands of pounds. Figure 33.5 shows the inside of a Ross[1] ribbon blender, and Fig. 33.6 shows a ribbon blender with both the end and center discharge. A big disadvantage of ribbon blenders is the time required for cleaning. The entire blend is not discharged with the blade rotation. When changing formulations, time for vacuuming or washing out the previous blend is required before a new mixture can be initiated.

A **twin cone blender**, or V-cone blender, operates similarly to a tumble blender. Two cones are joined together, as shown in Fig. 33.7, to form a V. As the shaft rotates, material is transferred back and forth between the two cones, mixing the formulation components at low intensity. After the mixing time is completed, the material is discharged out the bottom. Similar to a ribbon blender, the twin cone blender can be jacketed for heating and vacuum drawn for drying. Discharge is through the center of the bottom, with loading occurring through the top.

A **vertical blender** is cone shaped with an internal screw rotating on two different axes. The rotating screw orbits around the blender wall and moves material from the bottom to the top of the cone as it revolves around the center. The mix then cascades or flows back toward the bottom as fresh material is brought from the bottom to the top. A vertical blender requires less floor space and more overhead space than a ribbon blender. Similar to the double cone tumble blender and the twin cone blender, the vertical blender can be equipped with vacuum and a heated jacket to dry resins as they are being mixed. Due to its design, the vertical blender is very effective with small batches. Batches as small as 10% by volume can be effectively blended, while in a ribbon blender the mix needs to be at least 40% full to provide effective mixing. A vertical mixer is shown in Fig. 33.8. The blender wall slope and bottom discharge provide quick cleaning with no dead spots.

The last low intensity, off-line blending process discussed here is a continuous mixing operation. There are several continuous mixing designs, and they all operate off the same principle. Feeders continuously deliver formulation ingredients into a mixing chamber, where they are blended with an auger as the material moves through the mixing chamber. As long as the feeders contain for-

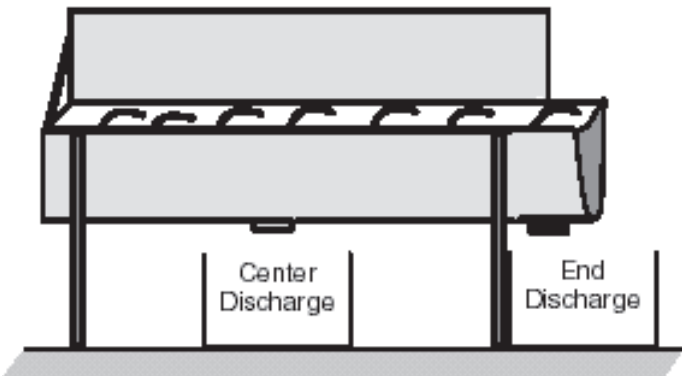

Figure 33.6. Schematic of a ribbon blender.

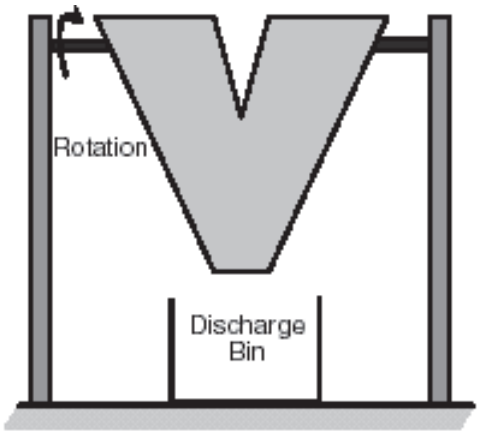

Figure 33.7. Schematic of a ribbon blender.

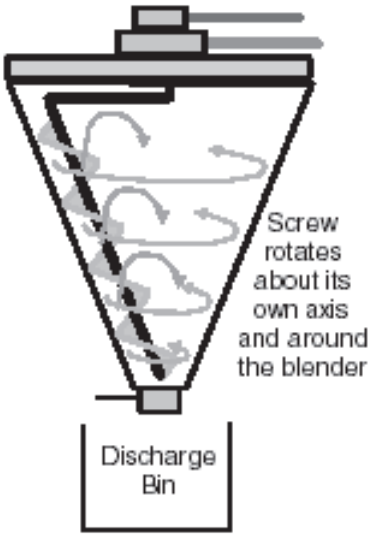

Figure 33.8. Vertical blender.

mulation ingredients, the process can run continuously. Figure 33.9 (supplied by Acrison, Inc.[2]) shows a continuous blender. A variable length trough (length is determined by how many ingredients are in the formulation) is equipped with a continuous length auger. Individual feeders along the blender deliver materials in the desired ratio to a rotating screw that mixes and conveys them forward to the discharge. After discharge the material can be fed either directly into an accumulation bin and transported to the extruder or to the extruder feed hopper, which can be located directly below the blender discharge chute. The single helix auger tumbles the dry components as it moves them forward along the trough, effectively and efficiently mixing the components. The trough is covered to prevent dusting and mixture contamination. Several feeders can be positioned on either side of the trough far enough from the discharge end to allow all components in the mixture to be uniformly distributed. Many variations of this scheme are possible. A number of feeders, mixing and conveying the components forward, can be starve feeding a large twin screw extruder or feed port. This same mixing principle is used in many in-line feeders, where individual feeders deliver the correct ratio of ingredients to a central chamber where the materials are mixed and transported to an extruder.

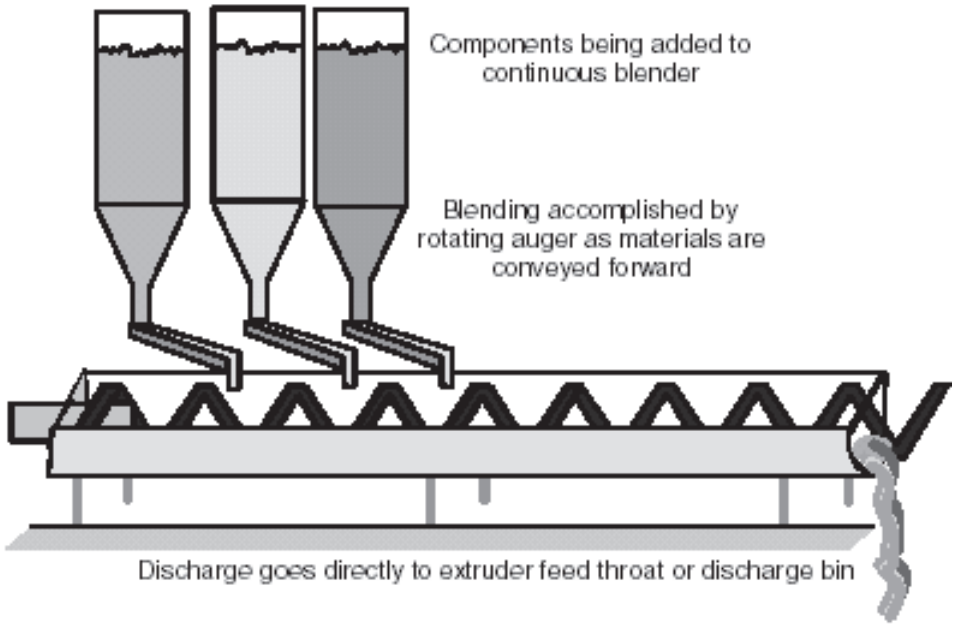

Figure 33.9. Continuous blender feeding either an extruder or discharge bin.

There are also **high intensity mixers**. These come in all sizes and shapes, ranging from the vertical model shown in Fig. 33.10 to horizontal models. The high intensity mixer shown in Fig. 33.10[3] is similar to a kitchen blender, where a blade in the bottom rotates at high speed, creating a vortex. The high intensity generates heat through shear. This provides relatively short mixing times compared to low intensity batch mixers. Low melting materials can melt in the process. These blenders are jacketed and in some situations cooled to remove excess heat generated in the mixing process.

Figure 33.10. Ferry Plas Mec high intensity mixer.

High intensity mixers are used to disperse powder/powder blends, intimately mix and disperse small concentrations of additives with powders, and generate masterbatches of additives with powder resin. High intensity mixers are used to premix stabilizers and plasticizers with PVC powder prior to extrusion. Enough heat is generated to soften the PVC particle, allowing the thermal stabilizers to adhere to the polymers and providing intimate contact during processing.

As mentioned previously, the advantages of off-line blending are lower capital investment and that one blend system can supply several extrusion lines. The disadvantages are excessive material handling, higher labor requirements, potential weighing errors, blend separation during transfer to the extrusion line, and contamination from blending different materials for several extrusion lines. In-line blending (blending directly above the extruder) eliminates a number of errors associated with off-line blending. In-line blending has the following advantages:

- There is less handling of material.
- There is less raw material usage, as there is no need to blend extra material. When production is stopped, the feeders are emptied, and any unused raw material is saved.

- If gravimetric feeders are employed, the correct ingredient ratio can be entered into the system, which eliminates potential weighing errors.
- The total extrusion system can be automated and monitored from feeder through puller speed.

The disadvantages to in-line blending are

- More capital is required for feeders and equipment.
- The extruder is not producing product during feed system cleanup.
- If only volumetric feeders are used, feeder calibration during changeover reduces extruder run time.

Many in-line blend systems are available. A batch blender can be located directly above the feed throat. Each feed hopper has its own auger or discharge mechanism metering a particular component into a small mixing chamber directly over the extruder feed throat opening, where the components are blended prior to entering the extruder feed hopper. Figure 33.11 shows a batch blending system. This is only one potential equipment configuration. Equipment manufacturers can provide many system modifications to the mixing chamber design, the feeding to the weighing hopper, and the feed hoppers used. However, all batch blenders operate on the same principle. Hoppers filled with formulation components are connected to a metering device that feeds the component ratio to a mixing chamber, where the components are mixed and the blend or mixture is discharged to the extruder. In Fig. 33.11 a load cell on the middle hopper weighs component A as it is discharged from the first hopper, followed by component B as it is discharged from the second hopper, and finally component C as it is discharged from hopper three. After the proper quantities are in the batch weighing hopper, the components are discharged to the mixing chamber, where blending occurs. During the blending cycle, the next batch is weighed into the batch weighing hopper. After the blending cycle is complete, the mixing chamber discharges the contents to the extruder hopper or a collection bin where they can be transported to the extruder, the next preweighed batch is discharged into the mixing chamber, and the cycle repeats. A gravimetric system can be based on either loss or gain in weight.

Volumetric delivery systems use screw speed and time to deliver the correct weight to the batch hopper. These systems are suitable for rates from 100–4000 lbs/hr (45–1816 kg/hr). The advantages of a continuous blender versus an in-line batch blending system are lower cost, simpler setup, and shorter material changeover time. Figure 33.12[4] shows two commercial gravimetric blend systems. Either system can be used as

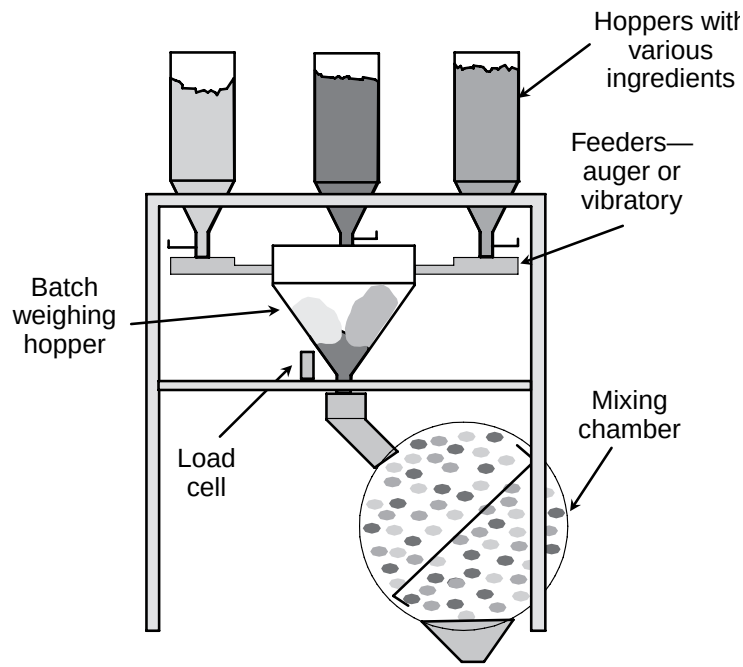

Figure 33.11. Batch blending system schematic.

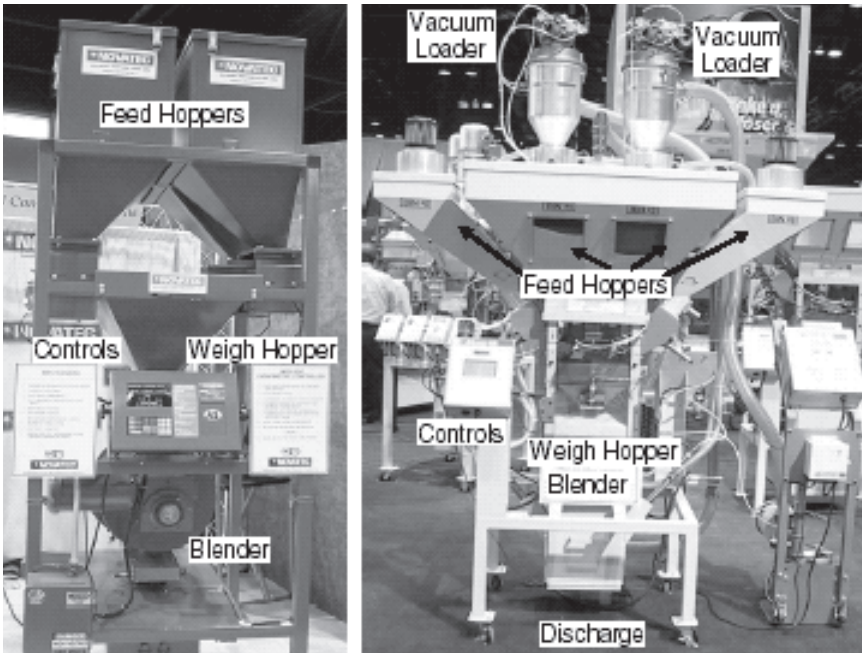

Figure 33.12. Gravimetric batch blenders.

a standalone unit or incorporated with an extrusion line. Novatec uses adjustable vibratory feeders and can store up to 100 recipes.

Figures 33.13 and 33.14 show two in-line gravimetric blend systems. Again, these general concepts are supplied by several companies with their special modifications. The systems are used with high throughput extrusion applications such as blown film, coextrusion, compounding, profile, and other large-volume systems. The blenders dispense all materials simultaneously in a preset ratio. Each material has its own hopper and load cell that control the delivery rate. Sophisticated controls on a single screw extrusion line can match the blender output to the extruder speed, which correlates to the material weight per unit length of product. A continuous system costs more than a batch system, as each ingredient has its own gravimetric feeder with a load cell, proportioning controller, and DC motor.

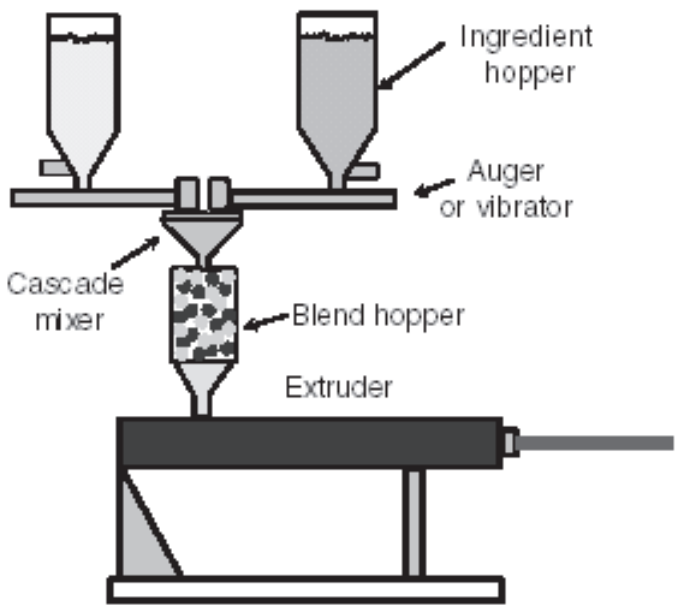

Figure 33.13. In-line continuous blend system.

Accuracy in blending depends on the equipment being used. Volumetric systems are not as accurate as gravimetric blenders. Volumetric is based on a fixed speed to deposit the correct amount of product in a blending chamber or through a continuous blender. If the calibration is incorrect, the formulation is incorrect. Calibration checks are critical to ensure reasonable accuracy. If time is unavailable for proper calibration checks and changes in settings between runs, volumetric blenders will waste more material producing the wrong formulation than is saved by not purchasing the more expensive gravimetric equipment. Gravimetric blenders, shown in Figs. 33.12, 33.13, and 33.14, are based on either weight gain or weight loss when making a batch. Weight-gain batch systems use augers or air-operated devices to dispense one material at a time in a desired amount into the hopper or mixing chamber, using one load cell to measure the mass of material added to the blend. Subsequently, other ingredients are added at a predetermined mass until all of the correct amounts of all formulation ingredients are added. After mixing is complete, the batch is dumped to a storage bin or extruder feed hopper. Loss-in-weight systems are based on each ingredient feed hopper having its own load cell to measure the amount of weight deposited by the individual feeder into a mixing chamber. In both loss-in-weight and gain-in-weight systems, the accuracy of the final blend depends on the accuracy and repeatability of the dispensing device, whether it is a screw, air actuated, a vibration device, or a rotating disk.

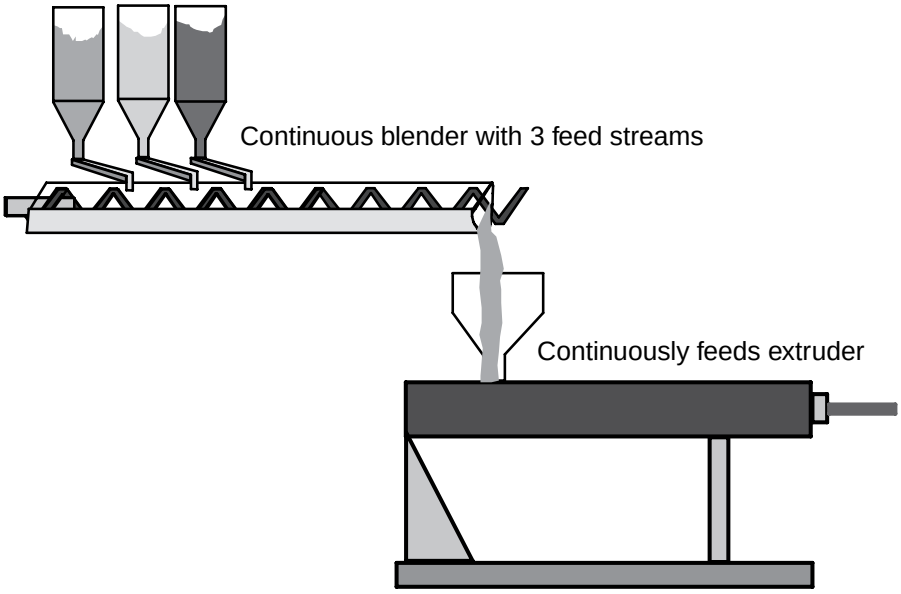

Figure 33.14. Continuous in-line blend system can be used to starve feed a twin screw extruder.

Before purchasing a blending system, carefully answer the following questions relative to the equipment to be installed or that is being upgraded:

- What is the anticipated maximum throughput rate?
- What is the maximum number of components to be blended in the different formulations?

- How are the materials going to be added to the blender or the feed hopper?
- What is the anticipated concentration range for each component?
- How accessible is the blender and feeders for product changes?
- Are all the formulation components free flowing? Are any of the formulation ingredients apt to bridge in a feed hopper?
- What is the physical form of each component, for example, pellets, powder, or regrind?
- Is the blender going to be located over the extruder throat or on a mezzanine, or at some remote location where blending is performed and then transported to the extruder?
- How many blends have to be made per shift and how many different components are in each blend? These factors affect cleanup time between the blends and the total throughput capabilities of the system.
- Is the blender easily cleaned in the location selected by vacuuming or washing? If the system is washed, can the water be easily disposed without draining on or affecting other equipment?
- Does the total blend or any component in the blend need to be dried prior to extrusion?
- Is each formulation and blend going to be printed out as a permanent record or part of the extrusion documentation?
- Is the blender going to be computer operated, whether it is portable and used in multiple locations or centrally located in one specific area?

An answer to each of these questions determines the type of system that is appropriate and defines its operation. A good supplier will seek answers to these questions before quoting or specifying a blend system.

In addition to determining the answers to the above question prior to buying a system, other criteria to address before purchasing a system are reliability, accuracy, ease of installation, user friendliness, system flexibility, safety in operation, customer support, whether additions to the system can be made easily, and price. To determine reliability, similar blend equipment can be run in the vendor's laboratory prior to purchase. Request a list of other installations and contact these companies about daily maintenance required, equipment availability, and formulation repeatability. Understand the manufacturer's warranty and ask about the number of field service people who can help in the event of equipment failure or downtime. What spare parts are required and are they stock items?

Batch-to-batch accuracy and precision is critical for repeatable operations and to producing the same product every time. If the vendor guarantees 1% accuracy, run tests in their lab with your formulation to verify 1% precision with your materials. Determine what is meant by 1% accuracy. Assume a blend is made with three components producing a 100-pound batch, and component A is added at 66 pounds or 66%, component B is added at 25 pounds or 25%, and component C is added at 9 pounds or 9%. If 1% accuracy means component C will range from 8 to 10 pounds or 1% of the total blend, this is unacceptable. Component C needs to be added at 1% accuracy or between 8.91 and 9.09%. Understand how often readings are updated in a gravimetric blender; a good rate is once every millisecond. With a large capital expenditure, it is essential to run trials on similar equipment with the ingredients to be processed in production to ensure blend accuracy and repeatability.

How long does it take to install the equipment and what construction trades are required to successfully complete the installation? Once the equipment is installed, is it easy to move the equipment to another plant location in the future as part of a plant expansion project? Are the installation instructions easy to read and follow? How difficult is the control wiring to install, and what are the power requirements? Is there adequate head space to properly install the equipment, maintain it, and operate it?

Once the equipment is successfully installed in your facility, how easy is it to operate? This can be partially determined in laboratory trials, during the equipment evaluation stage. Questions to ask concerning user friendliness are

- How easy are the controls to set up and change by either engineers or operators?
- Can recipes be stored electronically and brought back easily for future applications?
- Are the controls easily visible and understandable?
- Are there low-level alarms on the feed hoppers?
- Are sight glasses and viewing points present on all hoppers?
- Are the blender and feed hoppers easily disassembled and cleaned?
- Can blend recipes be locked out so they can be changed only by authorized personnel?

Safety during operation is a key factor in the purchase of any equipment. While all reputable equipment has proper safety electrical interlocks on doors and other moving parts, ensure that the equipment meets all local electrical codes. Verify that the frame is strong enough to properly support all the feeders and blender when they are full of resin.

Verify that the equipment uses standard software and hardware that is easy to replace in the event of component failure. Is the equipment expandable, can additional

feeders be easily added to expand the system at a reasonable price, or is the system at its maximum capability at the time of installation? If the system is expanded at a later date, can it be done relatively inexpensively or does it cost almost as much as the original system?

The final selection criteria are cost and customer support during the life of the equipment. Assume the equipment is being run in a three-shift operation. Is customer support available on weekends and evenings in an emergency? How long does it take to obtain replacement parts, and can they be installed locally or does it require a service technician to replace the part?

Proper mixing of formulation ingredients prior to extrusion is critical in any extrusion operation, whether it is done in-line or off-line. Reliable, accurate, and easy to operate equipment is essential to ensure that the same formulation is being added to the extruder every time a particular product is produced.

33.3 Feeders

The purpose of a feeder is to accurately control the flow of bulk solid material, regardless of its form—pellets, powder, regrind, fibers, flake, or mixtures of components—over a range of throughput rates in a continuous process, while minimizing waste at start-up or during operation. Feeder accuracy is critical to ensure the correct ratio of ingredients in the final extruded product. Feeder requirements include feeder stability, linearity over a prescribed throughput range, and consistency of measurement. Feeder stability is measured by the frequency the feeder needs recalibrating. In normal operation, once a feeder is calibrated for the type of material being fed, it should not require recalibration. Calibration checks can be performed during normal line preventative maintenance. The feeder has to demonstrate linearity over the throughput range being used. As the operator changes the setpoint, does the new setting produce the mass flow rate desired? This is more critical in volumetric than gravimetric feeders, which have feedback loops to measure and constantly correct the mass being delivered. Finally, the feeder has to produce consistent sample weights. Assuming 1-minute samples are collected over a 10- to 20-minute time frame, what is the sample standard deviation? Feeder manufacturers evaluate feeder accuracy in percent at 2σ, or 2 standard deviations, based on 1-minute samples. Assume 20 one-minute samples are collected and the standard deviation is 0.25 based on a throughput rate of 40 lbs/hr (18.2kg/hour); $2\sigma = 0.50$, and $[(0.50/40) \times 100] = 1.3\%$. Consequently, the accuracy of this feeder at a 40 lbs/hr (18.2kg/hr) throughput rate is 1.3% based on 2σ, or 40 ± 0.5 lbs/hr. If this precision does not meet the product requirements, this feeder is incapable of delivering the required accuracy needed.

There are two basic feeder designs commercially available. They are volumetric and gravimetric or loss-in-weight feeders. Volumetric feeders run at constant speed, while the gravimetric uses a load cell to deliver a constant mass flow. If the mass flow is below the setpoint, the feeder speed increases, increasing the mass flow; and if the mass flow is above the setpoint, the feeder speed decreases. Within these two basic feeders, many models, differing in size and shape, are available that use different mass transport concepts. Basic feeder designs are single screw, twin screw, belt, and vibratory. Volumetric feeders also come in disk and vane/slat designs. Volumetric and gravimetric feeders are discussed in detail below.

Factors to consider when purchasing feeders are

- Material type (pellets, powder, flake, regrind, fibers, etc.) being fed and its flow characteristics
- Material bulk density
- Maximum and minimum flow rates required for the application
- Particle size
- Feeder size
- Hopper dimension required to minimize cycles per hour
- Environment the feeder is being used in

Where is the feeder located; is it placed on a mezzanine above the extruder? Single screw extrusion is typically done via flood feeding. (The feed hopper on top of the extruder is full of the formulation to be extruded, and the throughput rate is determined by the extruder screw speed.) While individual feeders can be placed on a mezzanine above the extruder, as shown in Fig. 33.15, this is not the normal mode of operation for single screw extrusion but is quite common for twin screw extrusion.

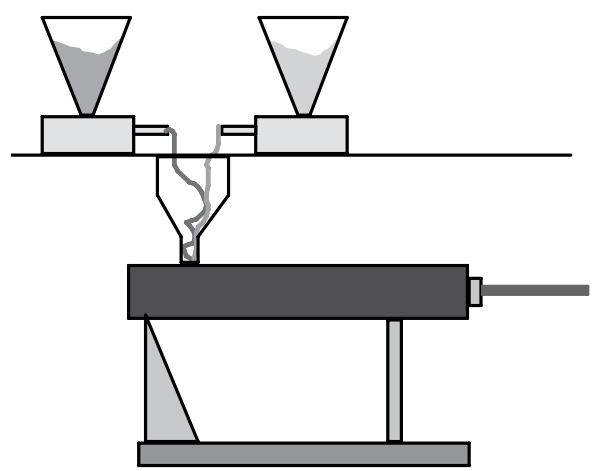

Figure 33.15. Feeders on Mezzanine above extruder.

Single screw extruders typically use in-line or off-line batch blending systems (previously discussed) to prepare the formulation for extrusion. While both single and twin screw extrusion can be accomplished through starve feeding, starve feeding is much more common in twin screw extrusion. (Starve feeding has no material in the extruder feed hopper as the extruder screw speed is set at sufficient speed to remove all materials being fed to the hopper. Throughput rate is determined by feed rate.) Individual feeders are placed on a mezzanine, suspended from a structure around the extruder, or mounted in some way above the extruder feed hopper. No blending is required prior to extrusion unless there are not enough feeders for each formulation ingredient. The feeders are preset to deliver the correct ratio of each component in the final blend to the extruder.

33.3.1 Volumetric Feeder

The most common volumetric feeder used for free flowing pellets is a single screw feeder, shown in Fig. 33.16.[5] The discharge chute is removed to show the helical screw, which conveys pellets or other materials. An agitator above the helical screw rotates inside the feed hopper to maintain material flow and prevent bridging (Fig. 33.17). The feed is assisted by the weight of material above the agitator and screw. As the material exits the feed hopper, it flows through a restrictive tube to the discharge point. Several screw geometries are available to accommodate different materials. Figure 33.18 shows a solid spiral screw. Figure 33.19 shows an open helical spiral conveying screw. Solid spiral screws with a center rod are used to feed powder and pellets, while open helical screws are used to feed pellets, fibers, and low bulk density fluffy materials. Both designs can have square-pitch or variable-pitch flights. The screw diameter and discharge chute can be changed to deliver larger or smaller volumes at a reasonable screw speed. Screw feeders tend to have a pulsating effect from the screw flight, regardless of whether the screw is an open flighted helix or a solid screw. As the screw flight passes the discharge opening, no or minimum material exits the feeder. Typically two-thirds of the material exiting the feeder per screw revolution is delivered in the first one-third of the screw rotation, while the other one-third of material exits during the final two-thirds of the revolution. This pulsing action can be minimized by changing the discharge chute so the material exits near the top of the discharge chute rather than through the bottom. By closing the end of the discharge chute and putting the exit opening at either an 11 or 1 o'clock position, the pulsing action can be minimized.

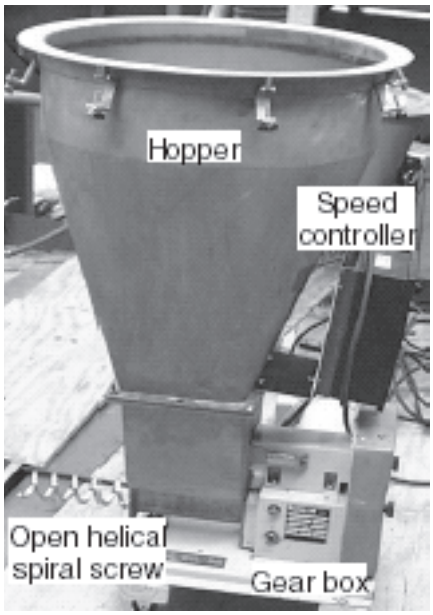

Figure 33.16. K-tron volumetric single screw feeder.

The screw speed is set with a potentiometer that is the speed controller setpoint for a motor drive. The motor is connected to the screw through a gear box. The feeder throughput rate is not linear over the entire speed range. This means that a volumetric feeder has to be calibrated for each material to determine output as a function of potentiometer setting or screw rpm. One- or two-minute samples are collected at different screw speeds, converted to lbs/hr or kg/hr, and plotted as a function of screw speed. A graph has to be generated for each material used in the feeder due to the different material flow properties for particle size, shape, moisture content, and bulk density. Other material characteristics affecting feeding are the adhesive or cohesive nature of the product. After calibration is complete, the proper screw speed for a given output is read from the graph. A final throughput rate check is run for two to five minutes to verify the calibration. If more than one volumetric feeder is being used, graphs and a weight check are required for each material and each feeder. When there are several volumetric feeders, it is sometimes difficult to obtain the desired component ratios. Every product change requires a new calibration, especially when there is a material change. The formulation accuracy and precision depend on the consistency and uniformity with which the agitator and screw can feed the product.

To improve the feed reproducibility and consistency with different materials, there are different agitators and screws available. (Two are shown in Figs. 33.18 and 33.19.) Agitators vary from the simple rotating blade shown in Fig. 33.17 to dual augers, concentric augers, triple augers, and flexible wall hoppers. Each is unique and may function differently with hard-to-feed materials. Free flowing pellets can be uniformly fed with most of the volu-

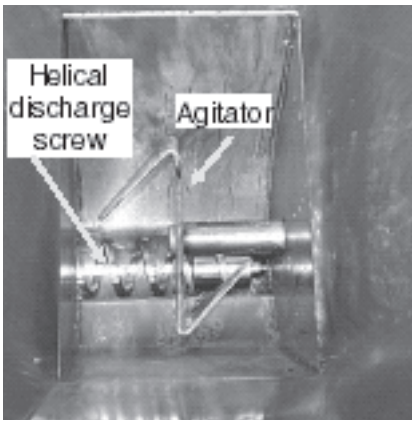

Figure 33.17. Agitator above discharge screw.

Figure 33.18. Spiral screw for single screw feeder.

Figure 33.19. Open helical spiral for single screw feeder.

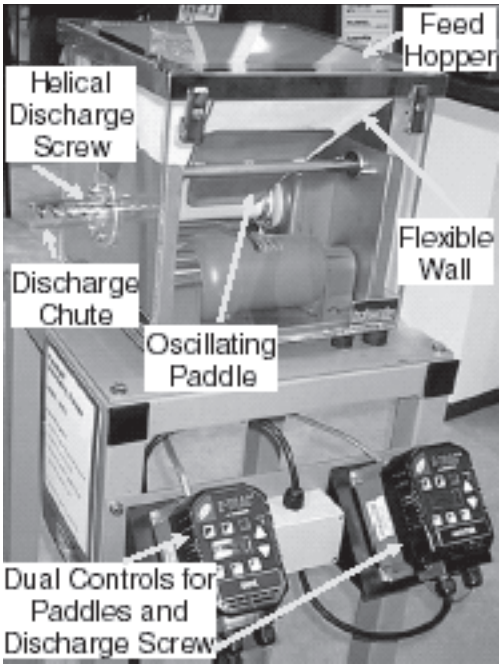

Figure 33.20. Brabender Technologie Inc. flex wall feeder.

metric feeders currently available. Powders, fibers, fillers, and adhesive or cohesive particles have to be tested on different feeders to determine which is most accurate, reliable, and consistent. A single auger feeder with an agitator is similar to that shown in Figs. 33.16 and 33.17. A double auger feeder supplied by Acrison Inc.[2] has a large spiral helical auger above the discharge screw to condition the material for the smaller discharge screw. The larger auger runs at a different speed from the discharge screw. Acrison Inc. also supplies a patented double concentric auger feeder, where the small discharge screw is centered in the middle of a large helical auger rotating at a different speed than the discharge helical screw. The function of the large auger is to condition the material and generate uniform discharge from the feeder by the smaller auger. Acrison Inc. also supplies a feeder with two large conditioning augers on each side of a smaller discharge auger. Similar to other Acrison feeders, the two large conditioning augers run at different speeds than the discharge auger. Some companies produce flex wall feeders, where paddles agitate the exterior flexible walls to prevent bridging and create uniform material flow to the discharge screw. Figure 33.20 shows a cutaway picture of a flex wall feeder from Brabender Technologie[6] with separate speed controls for the discharge screw speed and the oscillating paddles. Paddles move the flexible hopper walls in and out to provide uniform material flow to the screw. The flexible hopper walls are made out of molded vinyl or polyurethane that is replaceable. Vibra Screw Inc.[7] has another way to agitate feed the material so it fills the feed screw uniformly. Instead of using an agitator or paddles with a flexible wall, the bottom section of the feed hopper is vibrated to uniformly fill the screw. The vibrated feeder is used with nonfree flowing materials.

Some volumetric screw feeders use twin screws to convey material through the discharge chute and out of the feeder. Twin screw feeders are similar to single screws with an agitator over the discharge screws or a flex wall with paddles to condition the material and assist the flow uniformity into the discharge screws. Twin screw feeders can be equipped with different types of screws, depending on the material to be fed. Available designs are intermeshing corotating, and nonintermeshing corotating using either solid spirals or open helix. Figure 33.21 shows three solid screw designs with different pitches

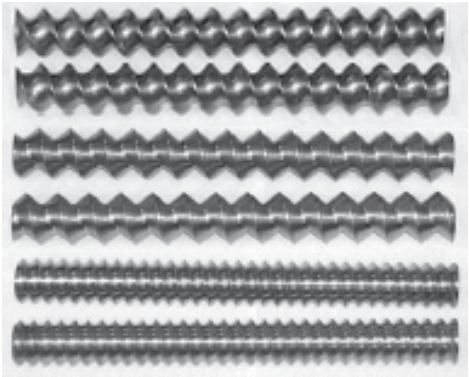

Figure 33.21. Three sets of solid corotating twin screw feeder screws with different pitches and flight depths from Brabender Technologie Inc.

used in twin screw feeders. Figure 33.22 shows the inside of a twin screw feeder hopper made by Brabender Technologie Inc. with a plow-type agitator above the screw to prevent bridging and condition the material for uniform flow to the discharge screw. Twin screw feeders are typically used to feed powders and other less free flowing materials that may not feed well in a single screw volumetric feeder. The bottom two screws in Fig. 33.21, with the small pitch, are used to feed powder and often small particulate materials. Twin helical spirals, Fig. 33.23, are used to feed pellets, fiber, and flake. These screws are less durable and can be easily twisted and deformed. The top two screw sets in Fig. 33.21 are large-pitch screws with deeper channels and can be used with pellets, granules, and flake.

Another volumetric feeder uses vibration rather than a rotating screw to deliver a constant volume per unit time.

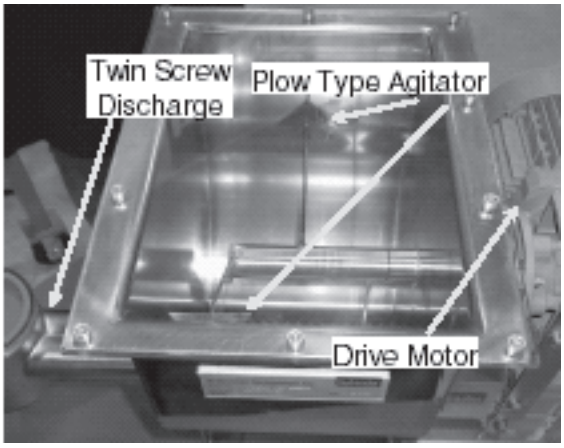

Figure 33.22. Plow-type agitator in Brabender Technologie Inc. twin screw feeder.

Vibratory tray feeders shake the delivery tray at a constant frequency. To change the feed rate, the vibration amplitude is changed. In some feeders, the hopper vibrates at the same frequency as the discharge tray. Arbo Engineering Inc.[8] has a patented system that moves the vibrating frequency of the tray to the resonant frequency. The feeder is designed to vary the frequency of the system to the resonance point while maintaining the amplitude frequency at the desired setpoint. Figure 33.24 shows two slightly different volumetric vibratory feeders. Similar to volumetric screw feeders, the feed rate depends on a constant vibration frequency and free flowing feed material. One advantage of vibratory feeders is there are no moving parts (such as screws) that might damage the feed material.

A third volumetric feeder has a belt running under a feed hopper with a gate to control the material deposited on the belt. The belt runs at constant speed. Like a vibratory feeder, there are no moving parts to degrade or damage the feedstock. The feed hopper is located over one end of the belt, while the other end of the belt deposits the product into the extruder feed hopper. Figure 33.25 shows a belt feeder. Belt feeders are low cost, have high throughput volumes, and are simply constructed. The rate of the belt speed and the size of the gate opening determine the throughput.

Extruders typically use either screw, vibratory, or belt feeders. However, two other volumetric feeders—the disk

Figure 33.23. Helical screws for twin screw feeder.

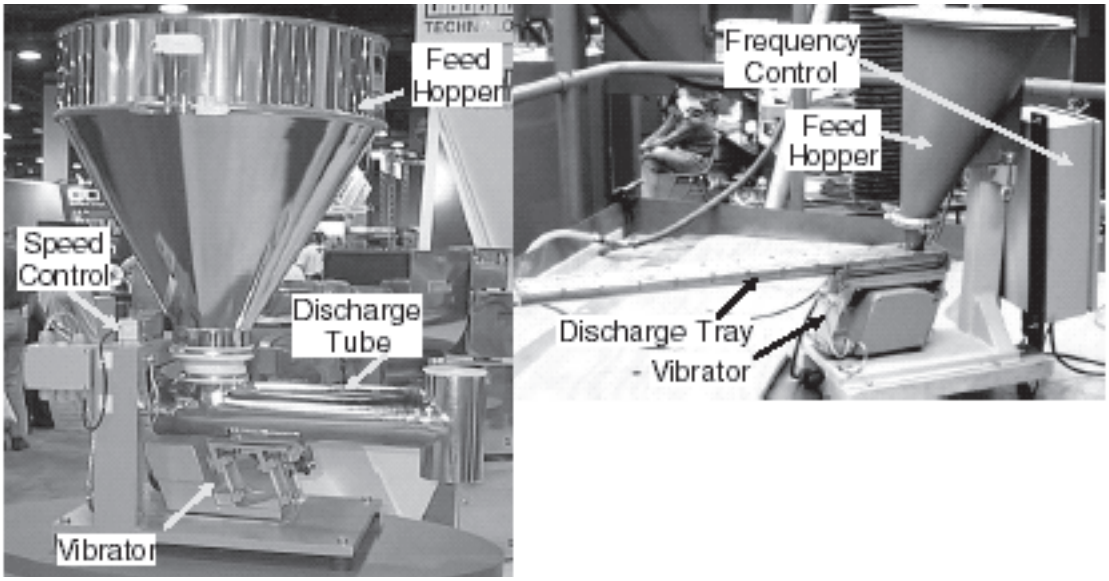

Figure 33.24. Brabender Technologie Inc. (left) and Arbo Engineering Inc. (right) vibratory feeders.

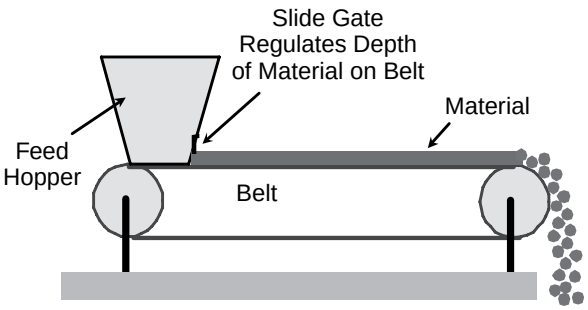

Figure 33.25. Belt feeder schematic.

feeder and the vane feeder—are available. The disk feeder, shown in Fig. 33.26, has a rotating disk that moves material from the feed hopper to the discharge chute into the extruder. A vane feeder, shown in Fig. 33.27, has slats that look like blinds at the bottom of a feed hopper. The slats are vibrated laterally to prevent material in the hopper from bridging across the opening. Throughput is controlled by the slat opening and the excitation vibration amplitude.

The advantage of volumetric feeders is low cost. The disadvantages are more output variation compared to gravimetric feeders, and recalibration is required every time the material or throughput rate is changed. A calibration curve (throughput rate versus speed) is required for each material being run in a given feeder. If several feeders are being used, it may be difficult to get them all properly set to provide the exact ingredients required in the formulation. If two resins are being blended in a 35/65 ratio, feeder calibration is probably not as critical. However, if one feeder is being used to add 1.2% of a thermal, oxidative, or UV stabilizer, the accuracy and precision are critical. A 0.5% variation in additive level may make the product unacceptable. Using volumetric feeders to add flame retardant additives may produce the wrong ratio and make the product have a UL94 V-1 flammability rating instead of V-0. One alternative is to add free flowing pellets with a volumetric feeder and other additives with gravimetric feeders. After calibrating the pellet feeder to determine the throughput rate, use this value to determine the setting of the other feeders to obtain the proper ratio of ingredients in the final products. In the past, most feeders were volumetric. To save time and money and to improve productivity, feeders are being converted to gravimetric or loss-in-weight feeders.

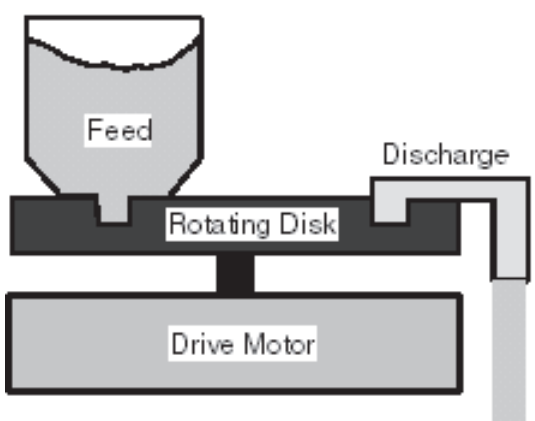

Figure 33.26. Schematic of disk feeder.

33.3.2 Gravimetric Feeders

Gravimetric or loss-in-weight feeders are more sophisticated than volumetric feeders.[9] The same general feeders used to feed materials volumetrically (single screw, twin screw, vibratory, and belt) are converted to gravimetric feeders by placing the feeder on a scale. As material is discharged from the feeder, the weight in the feeder is monitored and converted to a weight per unit time. Internal electronics compare the output with the setpoint and control the feed rate automatically to deliver a constant mass flow per unit time. Compared to volumetric feeders that run at a constant speed delivering whatever material is discharged, gravimetric feeders operate based on constant weight, with the speed being changed to deliver the desired mass per unit time. There are two main gravimetric feeder designs:

- A weigh belt that measures the weight and belt speed to deliver the correct weight
- A loss-in-weight feeder that monitors the loss in weight from the feeder over time

The screw speed or vibration amplitude controls the feeder rate. It is important not to touch a gravimetric feeder during operation unless the hopper is being refilled. Contact changes the measured weight, which will alter the output rate.

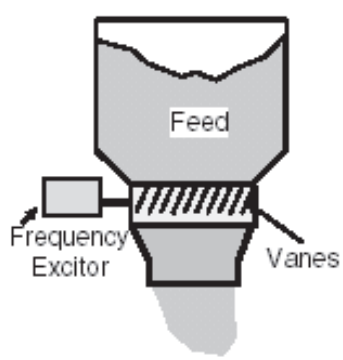

Figure 33.27. Schematic of vane feeder.

In operation, the feeder discharges material to the extruder or blend hopper as the weight decreases in the hopper. Mass flow is computed over time and compared to the setpoint. If a deviation between the actual measured value and the setpoint is present, the controller adjusts the speed to deliver the

correct product rate. Weight loss and time are used to calculate the mass flow rate. Figure 33.28 shows a representative curve of material mass flow being delivered versus time, with the refill cycle. Bumping the scale or adding weight, as in a refill process, causes the feeder to automatically transfer from the gravimetric to volumetric mode. After the hopper is refilled and the scale is steady, the feeder returns to its loss-in-weight mode. In Fig. 33.28, the sharp increase in weight is the refill time, when a large mass of material is rapidly added to the hopper. The slope of the loss-in-weight curve, $\Delta W/\Delta t$, is constant over a fairly long time cycle, representing the mass flow rate. Since operational accuracy depends on the time the feeder spends in the loss-in-weight mode, the material hopper needs to be large enough to minimize the refills per hour, and the refill delivery system needs to be sized to deliver the required material to fill the hopper quickly. If a high mass flow is being processed and the feed hopper is relatively small, it may be necessary to refill the hopper every 10 minutes. Therefore, six times per hour the feeder goes into volumetric or constant speed mode. If the refilling takes one minute and the feeder is in the volumetric mode for 1.5 minutes, each hour of operation the loss-in-weight feeder is operating nine minutes or 15% of the time in the volumetric mode, significantly reducing the accuracy of the feeding operation. For this reason, vacuum loaders are not hooked directly to loss-in-weight hoppers. A better refill method is to have a storage hopper over the feed hopper that is automatically filled by a vacuum loading system. As the feed hopper is emptied, a low-level sensor automatically opens the storage hopper gate, delivering material to the feed hopper until a high-level sensor shuts the storage hopper gate. After the feeder returns to the gravimetric mode, it may take a short time to return to the specified mass flow rate.

There are two common refill modes and they depend on the equipment sophistication and size: manual and automatic. In the manual mode, a bag or some other container is used to pour material into the feeder, while in the automatic mode low- and high-hopper-level sensors control resin transfer from a storage bin. During refill, the feeder switches to volumetric mode and continues feeding material to the extruder as material is being added. After returning to the gravimetric mode, the derivative of mass per unit time is calculated and appropriate corrective actions taken to keep the output at setpoint. One can envision during this cycle that the feeder may not deliver the desired mass flow. So some time is required for the system to reestablish gravimetric control. For this reason the feed hopper size is critical to prevent excessive refills. As mentioned previously, a vacuum loading system should not be installed directly onto a loss-in-weight hopper continuously loading material. A vacuum loader in this environment will constantly disrupt the loss-in-weight feeding operation, forcing the gravimetric feeder to operate in volumetric mode.

If a disturbance is caused by somebody bumping or leaning on the feeder, the feeder goes into volumetric mode for a short time. This leaning impacts the mass flow calculation. The method used to detect feeder disturbances depends on the feeder supplier.

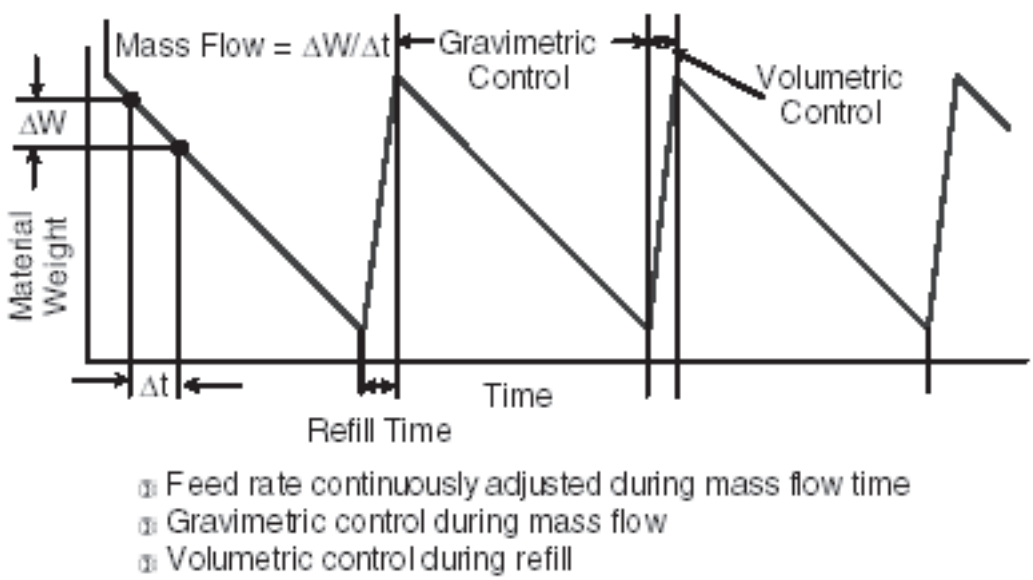

Figure 33.28. Gravimetric feeder operation over time.

The advantages of gravimetric feeders relative to volumetric are

- Saves time between product and formulation changes.
- Recalibration is not necessary between product and formulation changes.
- Faster setup than volumetric feeders.
- Product formulation is more accurate.
- Saves money with high-cost, low-volume additives, as it is easier to meet formulation specifications.

Gravimetric feeders' biggest disadvantage is the higher cost compared to volumetric feeders. Other disadvantages are that more things can go wrong, controllers are more complicated, and the feeders should not be touched during operation.

Figure 33.29 shows a small twin screw K-tron Soder feeder.[5] The square box under the feeder is the scale. Figure 33.30 shows a larger twin screw feeder from Brabender Technologie Inc.[6]; the discharge chute is not attached. Rather than a scale under the feeder, a load cell is attached to the frame to measure the loss in weight from the hopper as material is discharged. The solid screws shown in Fig. 33.21 are used in this feeder.

A weigh belt feeder is shown in Fig. 33.31. The weighing mechanism is under the belt, between the hopper slide gate and the feeder discharge. The mass is weighed as it crosses the scale. If the discharge rate does not meet the setpoint, the belt speed is adjusted. This feeder is similar to the volumetric belt shown in Fig. 33.25 with the added scale under the belt.

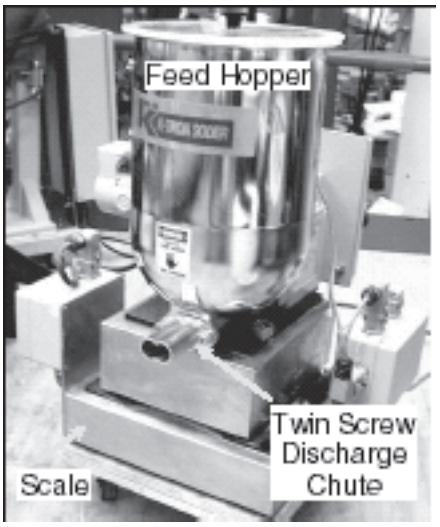

Figure 33.29. K-tron gravimetric twin screw additive feeder.

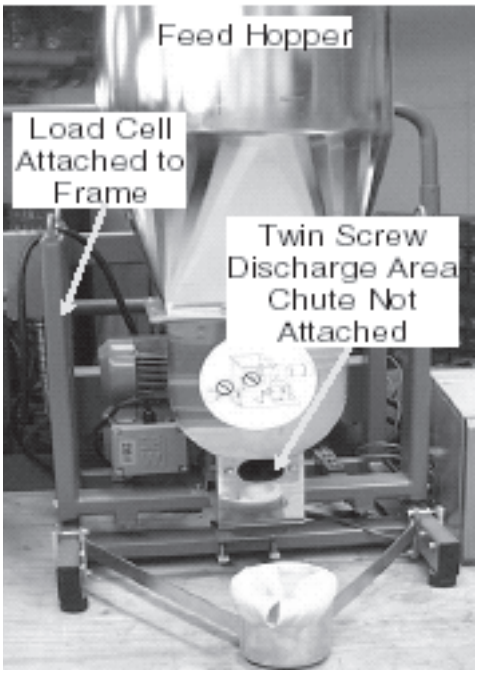

Figure 33.30. Brabender Technologie Inc. twin screw gravimetric feeder.

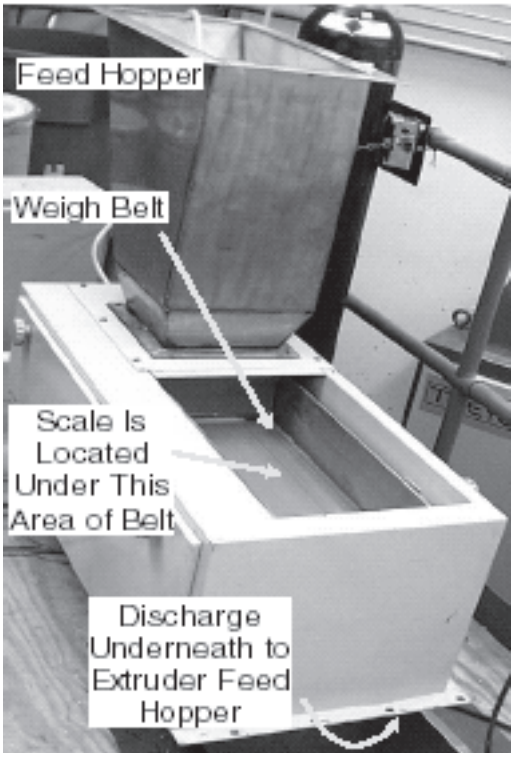

Figure 33.31. K-tron weigh belt feeder.

33.3.3 Liquid Feeders

Liquid feeders are used for colorants, liquid additives, plasticizers, and flame retardants. Liquid is fed with a positive displacement pump based on a constant throughput rate versus rpm. Similar to volumetric feeders, liquid feed pumps have to be calibrated to determine the appropriate feed rate. A graph of throughput in lbs/hr or kg/hr versus pump rpm is generated, and then the appropriate pump speed is selected to produce the desired throughput. As long as the liquid viscosity and density remain constant, the pump will deliver a constant, uniform flow and throughput rate. However, if the environmental temperature changes, the liquid viscosity and density will change the flow and throughput rate. Consequently, the liquid temperature when the calibration curve is generated has to be the same as the liquid temperature during operation.

To overcome problems with viscosity and density changes associated with temperature, the liquid feeder can be placed on a scale and the weight loss monitored versus time to create a gravimetric liquid feed system. Similar to other gravimetric systems, the liquid feed rate is regulated using the pump speed. The liquid feed output is no longer dependent on the liquid viscosity and density.

In high-volume, continuous extrusion applications, the positive displacement pump may be replaced with a gear pump to provide better durability and pump life. Gear pumps are also useful when pumping high-viscosity liquids into a high-pressure environment.

High-viscosity liquids may have to be heated to reduce the viscosity and provide sufficient flow for a positive displacement pump to operate correctly. Heated liquid temperature has to be controlled very accurately and the temperature has to be uniform throughout the container, or the feed rates will be inaccurate.

When specifying liquid pumping systems, the liquid viscosity, rate, and head pressure caused by the extruder are very important in obtaining the correct pump.

In both single and twin screw extrusion, it is best to introduce liquid additives downstream if possible. Adding liquid downstream through an injection nozzle in an open twin screw barrel section is straightforward. Sufficient mixing through a variety of screw elements can be provided to properly mix the liquid into the polymer melt. In single screw extrusion, liquid injection into the system is not as easy. If liquid is introduced in the feed throat, it may allow the solid resin to slip on the barrel wall and stick to the screw root, stopping all polymer feed to the extruder. A second scenario might have the liquid cause a slip/stick situation in the feed throat, leading to surging and output inconsistencies. In a two-stage extruder, the liquid can be added downstream through the vent port. The difficulty in this scenario is the screw must be properly designed to provide sufficient mixing in approximately one-third of the available *L/D* to properly disperse the liquid into the melt. A second factor hindering liquid dispersion is the liquid viscosity may be low compared to the plastic melt viscosity, which is compounded by the addition of a room temperature liquid onto a hot polymer, cooling the polymer and increasing its viscosity. An injection nozzle can be added downstream in a single screw extruder at a location other than the vent, but this requires drilling a hole in the barrel, mounting a liquid injection nozzle, installing special heater bands to go around the liquid injection nozzle, and developing a special screw design to properly mix the liquid into the melt.

Figure 33.32 shows a picture of a liquid displacement pump and controller. This is a volumetric feed system based on a constant pump speed. It can be used to feed liquids to either a single or twin screw extruder.

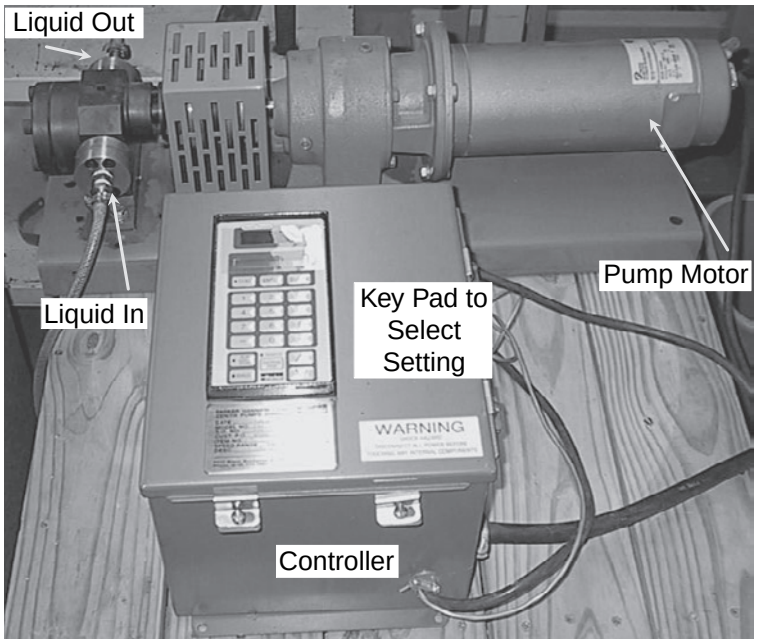

Figure 33.32. Zenith liquid feed pump.

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2. Acrison Inc., 20 Empire Blvd., Moonachie, NJ 07074.
3. Ferry, 4445 Allen Road, Stow, Ohio 44224-1093.
4. Novatec Inc., 222 E. Thomas Avenue, Baltimore, MD, 21225.
5. K-tron International Inc., Routes 55 & 553. Pitman, NJ 08071.
6. Brabender Technologie Inc. Mississauga, ON L5T 1Z6 Canada
7. Vibra Screw, 755 Union Boulevard, Totowa, NJ 07512.
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9. 9 Block, Debbie G., PT Auxiliaries, *What to Look for in a Gravimetric Blender*, March 1997, p. 25.

Review Questions

1. What is the difference between a volumetric and gravimetric feeder?
2. What are three different types of gravimetric feeders?
3. What is the difference between an in-line and off-line blend system?
4. Why is a gravimetric blend system more expensive than a volumetric blender?
5. What are the different steps in setting up a volumetric feeder versus a gravimetric feeder?
6. What are 10 factors to consider when purchasing a feed system?
7. What is the significance of each factor listed in Question 6?
8. List four types of off-line blend systems.
9. How does a high intensity mixer work, and what materials are mixed in it?
10. What is the difference between a double cone and V-cone blender?
11. What are some disadvantages of in-line blending?
12. What are some key questions to be answered before purchasing a blending system and why?
13. How is the mass flow controlled in a gravimetric feeder?
14. What happens to a gravimetric feeder when the hopper is being refilled?
15. What are the different types of screws used in a screw feeder, and why is more than one type necessary?
16. What is a vane feeder and how does it function?
17. What are the different types of materials fed with vibratory, screw, and belt feeders?
18. How does a belt feeder work?
19. What is a flex wall feeder and why does it have two speed controllers?
20. What types of screws are available for single and for twin screw feeders?

Certain polymers are hygroscopic in nature, meaning they absorb moisture from the air until their equilibrium moisture content is reached. The polymer chemical and molecular structure determines

- The moisture level that can be absorbed
- The effect moisture has on the polymer during extrusion
- The permissible moisture content allowed prior to extrusion

In general, polar polymer molecules containing oxygen and/or nitrogen as pendant atoms off the polymer backbone tend to absorb moisture through hydrogen bonding. The hydrogen atoms in water have a partial positive charge, while the oxygen has a partial negative charge, setting up a dipole moment in the molecule. Polymers with pendant oxygen or nitrogen atoms have partial negative charges on these atoms, which attract the positively charged hydrogen atoms. The attraction between the positive and negative charge sets up a weak bond called hydrogen bonding. While hydrogen bonds are weaker than the polymer molecular bonds, they are strong enough for the polymer to absorb and hold moisture at an equilibrium level based on the polymer structure. Once the moisture is absorbed, it can interfere with extrusion and polymer properties if it is not removed. Moisture can turn into steam in the extruder, having deleterious effects on the polymer structure and/or the extruded product. Figure 34.1 shows examples of hydrogen bonding in polyester-based polymers, while Fig. 34.2 shows hydrogen bonding in polyamide-type polymers. In polyamides, the pendant hydrogen on the nitrogen atom in the polymer backbone has a slight positive charge because the nitrogen atom is more electronegative than hydrogen. This pendant hydrogen is attracted to the negatively charged oxygen in water molecules. Most hydrogen bonding in polyamides occurs through the pendant carbonyl oxygen, and the possibility exists for hydrogen bonding with the pendant hydrogen on the nitrogen atom. The relative moisture absorbed by various polyamide polymers is shown in Table 34.1. As CH_2 groups are added to the backbone, the equilibrium moisture content decreases. This is from fewer pendant hydrogen atoms on the nitrogen atoms and fewer pendant oxygen atoms on the carbonyl groups, which decrease the relative locations available for hydrogen bonding.

Polar polymers containing oxygen, nitrogen, and sulfur are likely to absorb moisture due to hydrogen

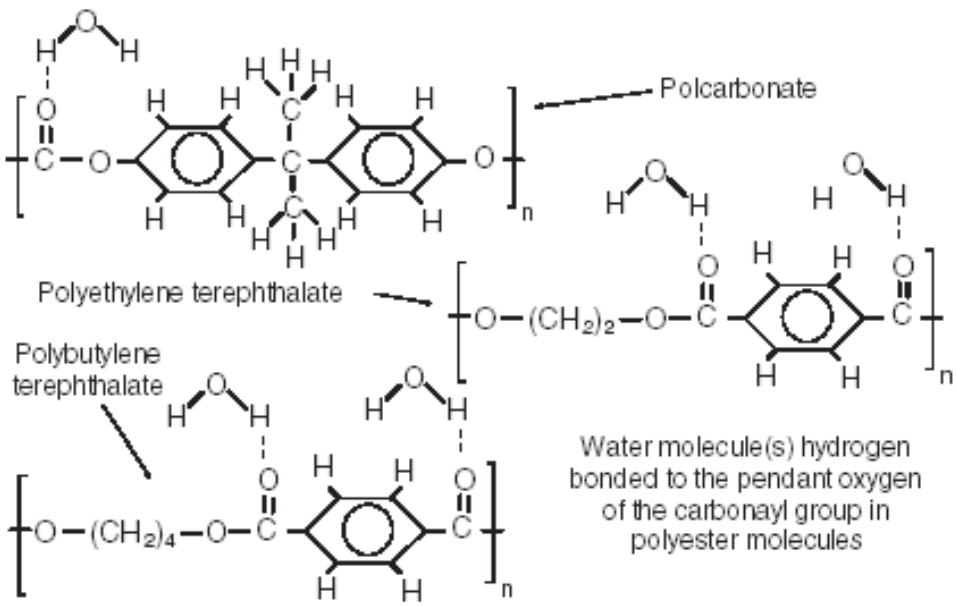

Figure 34.1. Examples of hydrogen bonding in polyester polymers.

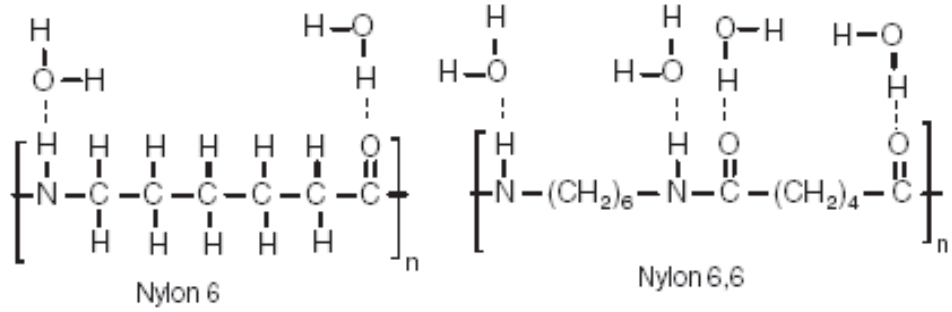

Figure 34.2. Potential hydrogen bonding in polyamide polymers.

Table 34.1. Polyamide Moisture Content

Nylon	Equilibrium Moisture, 50% Relative Humidity	Equilibrium Moisture, % Saturation
6	2.7	9.5
6,6	2.5	8.0
6,10	1.5	3.5
6,12	1.3	3.0
11	0.8	1.9
11	0.8	1.9
12	0.7	1.4

bonding and may need to be dried prior to extrusion. Polar polymers containing halogen atoms (fluorine and chlorine) do not absorb moisture through hydrogen bonding and do not normally need to be dried due to moisture absorption. Nonpolar polymers such as polyolefins and polystyrene do not absorb moisture and do not require drying prior to extrusion unless the pellets or powder are wet due to surface moisture. A partial list of polymers requiring drying prior to extrusion (this assumes the pellets or powder have not been packed and shipped dry in moisture-proof containers) is given in Table 34.2.

Polymers synthesized through condensation reactions are particularly susceptible to hydrolysis reactions at high temperature. In the synthesis of condensation polymers, monomer A is reacted with monomer B to produce polymer C, with water as a reaction byproduct.

Table 34.2. Some Polymers that Absorb Moisture and Require Drying Prior to Extrusion

Acrylonitrile Butadiene Styrene (ABS)	Polycarbonate Blends and Alloys
ABS Blends and Alloys	Polyesters
Acetal	Polyetheretherketone (PEEK)
Acrylic	Polyetherimide
Acrylonitrile Copolymers	Polyethersulfone
Liquid Crystal Polymers	Polyphenylene Ether/ Oxide Based Resins
Polyamide	Polyphenylensulfide
Polyarylate	Polysulfone
Polycarbonate	

As the reaction is an equilibrium reaction, higher molecular weight polymer is produced by removing water from the reactor. In the presence of heat and moisture, the reaction reverses, and the polymer backbone breaks, resulting in lower molecular weight and poorer properties. Condensation polymers include polyesters (polybutylene terephthalate [PBT], polyethylene terephthalate [PET], polycarbonate [PC], and many other polymers with a –COO– group in the backbone of the polymer) and polyamides. Polyamides (PA, nylon) are unique in that moisture plays a dual role. At high moisture levels and high temperature, water can break polymer backbone bonds, resulting in lower molecular weight. However, moisture also acts as a plasticizer for polyamide, improving its flow and processability. If the moisture level is too low, nylon becomes more viscous and is difficult to process. While polyesters need to be dried to below 0.02% moisture for processing, nylon 6,6 is normally processed between 0.07 and 0.2% moisture. Below 0.07% moisture, nylon 6,6 loses the plasticizing effect and the viscosity increases.

Moisture in noncondensation polymers can cause the following problems:

- At low concentrations, moisture can cause splay and surface defects.
- At medium concentrations and thicker extruded shapes, moisture turns to steam and acts as a blowing agent, generating a foamy structure. If the steam is uniformly dispersed, the product may contain small holes.
- At higher concentrations, moisture can cause a hole to be blown in the part as the steam expands upon exiting the die. In the extruder and die, the moisture turns to steam, which is under pressure and remains in a confined space. As the steam pocket exits the die, it expands rapidly, possibly blowing a hole in the extrudate with a popping sound.
- In very high concentrations, moisture can generate significant steam in the barrel and die, causing the molten polymer to be blown out the die exit. This is a dangerous situation, as molten polymer can be blown onto the operator, causing burns. Wet resin with downstream feed in twin screw extrusion can be accompanied by steam exiting through the downstream feed port.

All hygroscopic materials and nonhygroscopic materials where moisture has condensed on the pellet surface or the polymer is wet for other reasons need to be analyzed for moisture content before extrusion. In many operations polymers are dried for a specified time and temperature, and the material is never checked for moisture content prior to extrusion. The

only safe way to determine if the resin is suitable for extrusion is to run a moisture check. Several test methods are available to check moisture content. Arizona Instruments[1] has a moisture analyzer that measures weight loss versus time as it heats the polymer to a specified temperature. In 10 minutes or less, the moisture content can be measured and the suitability for extrusion determined. While this instrument really checks volatiles and not just moisture content, it is sufficiently accurate to determine the suitability of a hygroscopic resin for extrusion.

Some polymers containing moisture are processed in vented extruders to remove the moisture as steam prior to the die. With a single screw extruder, it is necessary to use a two-stage screw with the proper temperature profile to prevent molten polymer from bleeding out the vent. In a twin screw extruder, a barrel section one or two sections before the die can be opened for atmospheric or vacuum venting. A melt seal before the vent and large-pitch elements under the vent lower the polymer pressure and help the steam to escape. A vacuum pump attached to the vent increases the steam removal. Another advantage of processing with a vacuum vented extruder is that volatiles generated during extrusion processing are removed, too. Condensation polymers such as polyesters need to be predried rather than allowing the steam to be removed in a vent. As these polymers melt in the presence of water, polymer chain scission occurs through a hydrolysis reaction before the moisture can be removed.

Materials that have been properly dried and are ready for extrusion need to be kept dry during processing. With single screw extrusion, a hopper dryer above the feed throat can be used to keep pellets dry. Sometimes the hopper is blanketed with dry nitrogen to prevent moisture absorption during processing. Some hygroscopic resins are dried by the manufacturer and shipped in moisture-proof containers. These containers are opened as they are ready to be used and only minimum amounts sit in feed hoppers and extruder hoppers. After a moisture-proof bag has been opened, unless it is properly resealed the resin will absorb moisture with time and have to be dried prior to its use.

Table 34.3 lists polymers that are not hygroscopic and typically do not require drying prior to extrusion.

Table 34.3. Non-hygroscopic Polymers

Polyethylene	Polyvinylchloride
Polyethylene Copolymers	Polystyrene
Polypropylene	Fluorinated Polymers
Polypropylene Copolymers	

34.1 Drying Definitions and Factors Affecting Drying

Drying occurs when a vapor pressure differential exists between the pellet moisture and the surrounding air. Moisture migrates to the medium with the lowest vapor pressure. If the air is drier than the pellets, moisture migrates from the pellets to the air. On the other hand, if the air moisture is greater than in the pellets, water will migrate to the pellets. Removing moisture from pellets is a diffusion process and requires time. As moisture migrates from the pellet surface to the air (assume the air has lower moisture content), moisture from the pellet center diffuses to the pellet surface, where the moisture content is lower. Heating the pellets increases moisture diffusion through the pellets. This also increases the moisture migration from the pellet surface to the air.

Terms used to describe drying factors are

- Relative humidity
- Dew point
- Moisture weight percent in the plastic

Relative humidity is the actual air moisture compared to air saturated with water at that temperature. The higher the air temperature, the more moisture the air can hold. Hot summer air can hold significantly more moisture than cold winter air. Similarly, hot air in a drying oven can hold more moisture at higher temperature. If the hot air has a high humidity, plastics can easily absorb the moisture from the air, increasing their moisture content. The air dew point determines the air moisture. The dew point is the temperature where moisture condenses out of the air. Lower moisture in the air correlates with a lower dew point. A dew point of –4˚F (–20˚C) means that the air must be cooled to –4˚F (–20˚C) before moisture will condense out of the air. If the dew point is –4˚F (–20˚C), the relative humidity in 250˚F (121˚C) air is very low, and this assists moisture migration from pellets to air. The weight percent of moisture in plastic is given by Eq. (34.1):

$$\text{Weight \%} = [(\textit{Sample Weight} - \textit{Sample Weight Dry}) / \textit{Sample Weight}] \times 100 \tag{34.1}$$

Five percent moisture means each 100 pounds of plastic resin actually weighs 95 pounds, as it contains 5 pounds of water.

Factors affecting the drying rate are

- Air temperature surrounding the pellets—At higher temperatures, air can hold more water and the diffusion rate in the pellets is faster.

- Air dew point—Lower air dew point lowers the air moisture content and provides a larger driving force for water migration from the pellet to the air. Combining high temperature with a low dew point generates a greater pressure differential between the moisture in the plastic and the air.
- Weight percent moisture in the plastic—Higher moisture content requires more time to remove or reduce the moisture to acceptable concentration.
- Air flow around the pellets—Air flow around each pellet is critical to remove wet air (air where moisture has already migrated from the pellet to the air) and replace it with dry air. New dry air maintains a vapor pressure differential between the air and the pellet surface. Air flow required for proper drying depends on the plastic heat capacity and the air drying temperature. (Heat capacity is the amount of heat required to raise the temperature of a specific mass of plastic one degree.)

Time and temperature required to dry various polymer resin systems vary greatly, depending on the heat resistance, melting or softening point, and thermal resistance of the plastic. Overdrying or drying at excessively high temperatures can have deleterious effects. Nylon 6,6 dried at 200˚F (93˚C) for four hours or more can cause thermal oxidative degradation, leading to yellowing. Polycarbonate is typically dried at 250˚F (121˚C) for four to six hours; drying ABS at the same temperature results in a large solid ball because this is above the softening point (T_g) of ABS. As the pellets soften, the surface becomes tacky and the pellets stick together, so a solid mass results on cooling. PET is very moisture-sensitive in processing, requiring the moisture content to be below 0.01%. To attain this level, PET is first crystallized to prevent the pellets from softening and sticking together. Once PET is crystallized, it is heated to 300˚F (150˚C) to remove moisture and finish the drying process. Excessive drying temperature or time consumes thermal stabilizers that may be required during processing, reprocessing, or in the application if the part is used at elevated temperatures. After the thermal stabilizer is consumed, the resin degrades more rapidly. Drying conditions and moisture content guidelines for various resin systems are provided by suppliers. Table 34.4 presents some guidelines for moisture content and drying conditions.

A general guideline for dryer air flow is 1.0 cubic foot per minute (cfm) of air per pound of plastic being processed per hour.

Table 34.4. Some Typical Resin Drying Conditions

Resin	Moisture Content, %	Drying Temperature, ˚F (˚C)	Typical Drying Time hours[1]
ABS	<0.02	185 (85)	4
Nylon 6,6	0.2 – 0.08	170 (77)	4 – 6
Nylon 6	0.08 – 0.02	180 (82)	4 – 6
Polymethyl methacrylate	>0.05	175 (79)	2 – 3
Polycarbonate	0.02	250 (121)	4
Polybutylene terephthalate	0.02	275 (135)	2 – 4
Polyethylene terephthalate	0.01	300(150)	2 – 6
Polyphenylene sulfide	0.02	300 (150)	4
Acetal	0.02	210 (99)	2
Styrene acrylonitrile	0.02	190 (88)	3

1. Depends on moisture content when drying is started. High moisture contents take longer time to dry to processing levels.

34.2 Types of Drying Equipment

Drying equipment ranges from general purpose ovens with trays to sophisticated dehumidifying dryers, drying thousands of pounds per hour with dry air conveying systems to transport resin from dryers to feed hopper without exposing the resin to the environment. Drying equipment can be divided into three categories:

- Ovens
- Hopper dryers
- Central drying systems

There are three general drying systems used for hopper dryers and central drying systems:

- Hot air
- Refrigerant dehumidifying
- Desiccant dehumidifying

Each drying system is discussed in detail below. Dryers are used principally to

- Remove surface moisture of both hygroscopic and nonhygroscopic resin systems
- Preheat pellets prior to extrusion to increase the throughput rate
- Dry hygroscopic resins to acceptable moisture content for processing

34.2.1 Oven Drying System

Oven drying is the least sophisticated system and is generally used for small lots in product development or laboratory-type applications, where small samples have to be dried simultaneously. There are many oven sizes and designs that satisfy different applications. Ovens available include hot air, vacuum, and dehumidifying with regenerative desiccant beds. A hot air oven, containing samples for drying, is shown in Fig. 34.3. The blower circulates air through a heater, where the air is heated to the desired setpoint and then flows through the oven from bottom to top. In some ovens the air flows from side to side. In most designs the air is recycled back through the blower and heater. If the air relative humidity entering the oven is high, the oven drying capacity is limited. At a given temperature, the air can absorb moisture only as long as the moisture in the pellet is greater than the air moisture. At some point the water no longer migrates from the resin to air. Unless the dew point or air moisture content is lowered, hot air continues to circulate around the pellets, but no additional moisture is removed. Heating the air hotter allows the air to absorb more moisture as the relative humidity drops; however, the higher temperature may cause the resin being dried to become tacky, melt, or degrade as the thermal stabilizers are consumed.

To improve the drying efficiency, a desiccant bed is used to lower the air dew point or a vacuum is used to increase the pressure difference. A desiccant oven is identical to that shown in Fig. 34.3 with the exception that a desiccant bed is added between the blower and the heater. With lower moisture content in the hot air, the water partial pressure differential between the pellets and air is greater, improving the drying efficiency. Moisture transfers from the pellets to the air, and the moisture in the recycled air is removed in the desiccant bed before reheating to the setpoint. Normally there are at least two desiccant beds, one drying the air while the other is regenerating. In the regeneration cycle, the desiccant is heated to drive out the moisture. After regeneration, the dry bed is rotated into the air stream to remove the moisture while the other desiccant bed is regenerating. A second approach to lower the air moisture content is to apply a vacuum, lowering the oven pressure. As the water partial pressure differential is increased between the pellets and the surrounding air, water transfers to the lower pressure. The moisture leaving the pellets is removed by the vacuum pump.

To improve oven drying efficiency, the pellet bed thickness in the drying pans needs to be as thin as possible. It is more efficient to use more trays with less resin in each tray, as the circulating air does not flow between pellets. Oven shelving is perforated to provide good air flow through the oven. Solid shelving restricts air flow and reduces oven efficiency. Another factor that may lead to poor drying is overloading the oven. This restricts the air flow and may generate excessive moisture in the oven as it diffuses from the pellets. Once all the material in an oven is dried, addition of new material with higher moisture content may affect the dried material in the oven. If the air moisture content becomes higher than the moisture content of the dried pellets, moisture migrates from the air back into the previously dried material.

Without adding a desiccant bed to circulating dry air or a vacuum to remove moisture, ovens do not provide acceptable means to dry hygroscopic resins. If dry air is not used, hygroscopic resins like PET and PBT can actually act as desiccants, with moisture migrating from the air to the resin. Properly dried hygroscopic resins placed in a standard hot air oven without a desiccant dryer or vacuum will increase in moisture content over time. Standard hot air ovens do not lower the dew point of the air, which is required to dry hygroscopic resins.

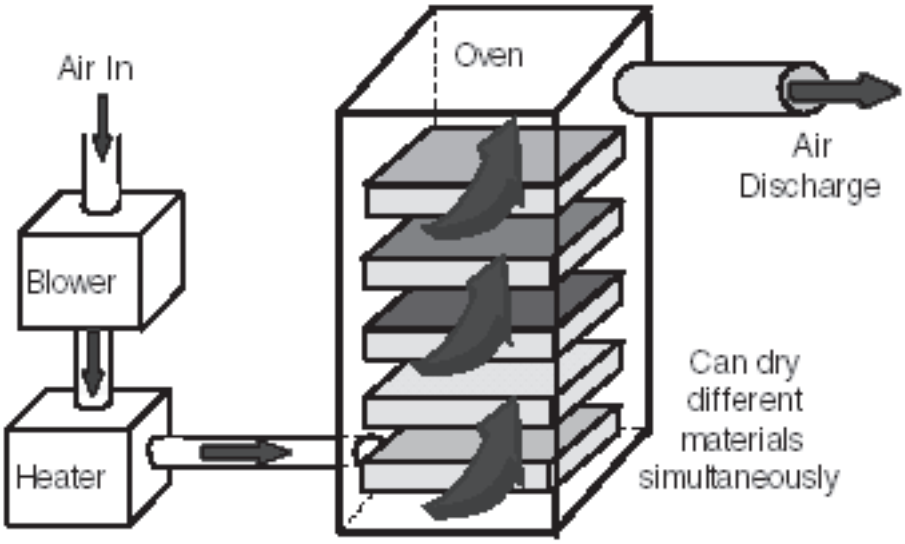

Figure 34.3. Drying oven schematic.

34.2.2 Hopper Dryers and Central Drying Systems

There are three types of hopper dryers:

- Off-line
- In-line
- Portable

An off-line dryer is used in a central drying system where dried materials are pneumatically conveyed to the extrusion equipment. An in-line dryer is used on a single extrusion line. In single screw extrusion, the dryer may sit directly above the feed throat, feeding hot pellets to the extruder. An in-line dryer may be coupled with an off-line dryer if the off-line dryer is supplying many extruders. Large off-line dryers can dry resin for multiple extruders; dried resin is conveyed from the off-line dryer to the in-line hopper dryer with dry air. The in-line dryer either maintains or lowers the resin moisture content as the resin awaits extrusion. The third hopper dryer is a portable dryer that can be moved around the plant to the extrusion line requiring dry resin.

Hopper dryers are designed for plug flow, meaning the first material added to the dryer is the first material out. In normal operation, as material is removed from the bottom, additional resin is added to the top through an automatic conveying system. It is important to scale the dryer to the equipment throughput requirements. A very significant production problem occurs when extruder capacity exceeds drying capacity. Assume a PET production run is being planned and a 1000-pound off-line dryer is available to dry the resin. One thousand pounds are added to the dryer and dried at 300˚F (150˚C) for four hours. The moisture content is checked and found to contain 0.008% moisture, which is below the moisture limit required for processing. Resin is transferred from the dryer to a 250-pound in-line drying hopper, where it is held for extrusion. As resin is transferred to the in-line dryer, new resin is added to the off-line drying hopper. Production is started and running at a throughput rate of 225 lbs/hr. After two hours, the product meets all specification and is running at very high efficiency. As a result, the rate is increased to 300 lbs/hr, and everything runs smoothly with product meeting specifications. The operation continues in this mode for three hours, with all processing and product parameters meeting specifications. Sometime during the fifth hour of production at 300 lbs/hr, the process starts to change, and sometime in the sixth hour the physical property impact is low. The question is, What has changed or gone wrong after five hours and gotten worse in the sixth hour? While extrusion conditions and resin lots will probably be the first parameters investigated, the cause of this problem is that after hours of production, the resin in the dryer did not have its full four hours of drying time. The 300 lbs/hr production rate exceeded the drying capacity, which was 250 lbs/hr based on a 1000-pound dryer requiring four hours of drying time. As the run progressed past four hours, the drying time continued to decrease and the moisture content increased, causing degradation in the extruder. One alternative may be increasing the drying temperature, assuming this does not lead to other problems. Sizing the dryer to the throughput rate is critical when specifying dryers or designing drying systems to work in tandem with the extruder.

The dryer hopper has to be properly insulated to provide uniform temperature around and across the dryer. The resin and air temperature at the walls need to be the same as the center temperature in the dryer to promote uniform drying. Insulation is important both for energy efficiency and safety. Hot hoses can cause burns if touched. Supply and return lines have to be properly designed for the drying temperatures being used. Above 200˚F (93˚C), high-temperature silicone hoses are recommended for both air supply and resin discharge. Above 250˚F (121˚C), high-temperature silicone hoses are required on the air lines. Coolers on the return air reduce the air temperature, improving the desiccant efficiency to lower the moisture content. After-coolers are used when drying temperature is greater than 160˚F (71˚C). Figure 34.4 shows percent desiccant loading as a function of desiccant temperature.[2] You can see that above 160˚F (71˚C) the desiccant holds much less water than below 160˚F (71˚C).

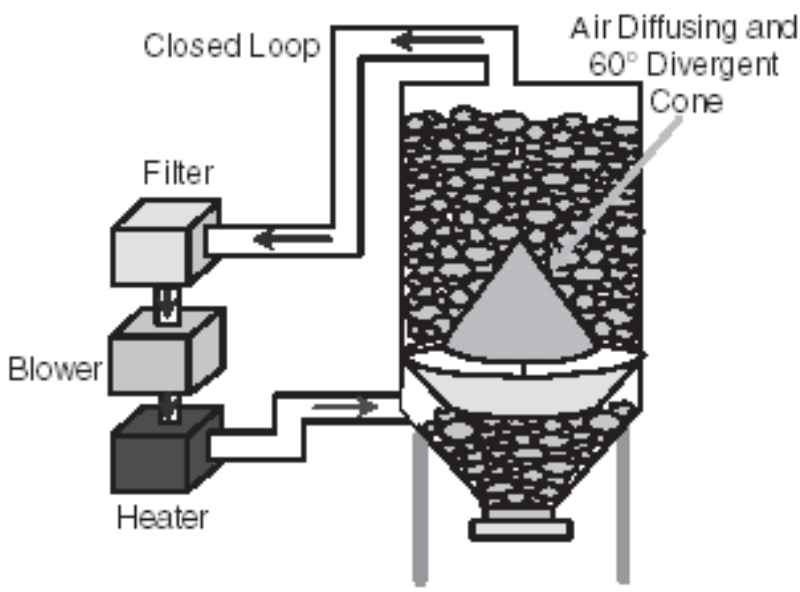

Figure 34.4. Percent desiccant loading vs. temperature.

As mentioned previously, dryers work on the principle of plug flow with a 60˚ divergent cone in the center of the dryer to encourage plug flow through the dryer. Plug flow, with its constant residence time for each pellet, provides uniform drying.

The three general drying systems are—

- Hot air
- Refrigerant dehumidifying
- Desiccant dehumidifying

—and are all used for drying plastics, with the most common being the desiccant dehumidifying dryer. A hot air dryer is shown in Fig. 34.5. The schematic shows a closed loop system; air exiting the dryer passes through a filter and then is blown through the heater, where the air is reheated to the desired drying temperature. The hot air enters the cone-shaped area at the bottom of the dryer and exits the top. Resin at the very bottom of the dryer cone below the hot air entrance may not be properly dried in start-up mode, as the hot air does not properly circulate through this resin. It is sometimes necessary to remove the resin from the bottom of the dryer after the initial drying is completed and recycle it to dry it properly.

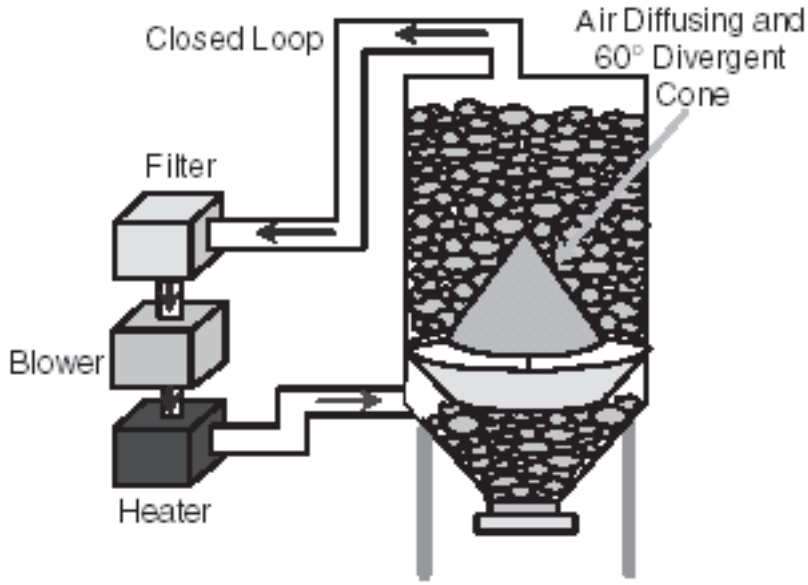

Figure 34.5. Hot air dryer schematic.

The hot air dryer shown in Fig. 34.5 cannot be used to dry hygroscopic materials such as polyesters and polyamides. As the moisture migrates from the pellets to the air, the air can only absorb moisture until equilibrium is reached between the pellets and the air. Once a steady state is attained, no additional moisture is removed from the pellets, regardless of the time the pellets are exposed to high temperature. To remove more moisture, drier air has to be introduced into the system.

In an open system, new air is introduced into the heater through a blower and exits the top of the oven into the atmosphere. Steady state in an open system depends on the air dew point entering the blower. If the dew point is low (low humidity on a cold winter day), the air circulating through the oven can absorb more moisture, drying the resin more quickly and to a lower moisture content. Similar drying efficiency is not obtained in the summertime when high humidity is present. The biggest disadvantage of an open versus closed loop operation is the energy inefficiency, as heat is given up to the surroundings.

To improve the efficiency and flexibility of hot air dryers, incorporate a moisture removal system prior to heating the air to provide a lower air dew point for drying the resin. Two systems used to remove moisture from air entering the heater are refrigerant dehumidifying and desiccant drying. In refrigerant dehumidifying, air flows across a refrigeration coil, where the air is cooled to its dew point. Moisture condenses on the coil and drier air goes to the blower and heater. Figure 34.6 shows a refrigerant dehumidifying dryer. The dryer shown is an open system; in practice the system will be designed to recirculate hot air from the dryer back through the evaporative and condenser coils to remove the moisture from the air.

The purpose of the filter in the closed lines is to remove fines and dust particles exiting the dryer. The dust can create a safety hazard if it is recirculated back through the blower and heater. Filters need to be replaced or cleaned in a regular preventive maintenance program. A plugged filter reduces air flow through the oven, decreasing the drying efficiency. Filter cleaning depends on the materials being dried and the fines. Resins with significant fines require more frequent filter cleaning and replacement than cleaner resins. Other maintenance includes blower motor and bearing lubrication and heating element inspection.

Hot air dryers can be used with nonhygroscopic resin systems to preheat resins prior to extrusion. If the resin is heated near its melting or softening point, less heat is required for processing from either conductive or shear heat in the extruder. Preheating resins can increase the throughput as more resin can be melted. Assume polystyrene is being extruded at a melt temperature of 450˚F (232˚C). The energy required to heat the resin to that temperature is given by Eq. (34.2):

$$Q = \dot{m} \times C_p \times \Delta T \tag{34.2}$$

where Q = Energy (cal or Btu or watts)
m = Mass flow in mass/time
C_p = Heat capacity (cal/g ˚C or Btu/lb ˚F or kJ/kg ˚C)
ΔT = Temperature rise

At 250 kg/hr, heating polystyrene from 25˚C to 232˚C requires 18.7 kW from the following calculation (C_p = 1.3kJ/kg ˚C for polystyrene and 1 Joule = 1 watt-sec):

$$Q = (250\text{kg/hr})(1.3\text{kJ/kg}^\circ\text{C})(232–25^\circ\text{C}) = 6.73 \times 10^4\text{kJ/hr}$$

$$Q = (6.73 \times 10^4\text{kJ/hr}) \times (1\ \text{hr}/3600\text{sec}) = 18.7\text{kW}$$

If the resin is preheated to 90˚C (10˚C below polystyrene glass transition temperature), the energy required by the extruder to raise the melt temperature to 450˚F (232˚C) requires 12.8 kW, or about 33% less energy.

$$Q = (250\text{kg/hr})(1.3\text{kJ/kg}^\circ\text{C})(232–90^\circ\text{C}) = 4.62 \times 10^4\text{kJ/hr}$$

$$Q = (4.62 \times 10^4\text{kJ/hr}) \times (\text{hr}/3600\text{sec}) = 12.8\text{kW}$$

A desiccant dryer is shown schematically in Fig. 34.6. This is the most common dryer in the plastics industry. Air is passed through a desiccant bed to remove moisture and lower the dew point, heated, and blown through the dryer hopper, where moisture transfers from the resin to the air. In the closed system shown, the air is recirculated back through the blower and desiccant beds to remove moisture absorbed in the dryer. Normally dryers have at least two desiccant beds, one in the drying mode while the second one is being regenerated. When the desiccant bed becomes wet and the dew point starts to rise, the dry bed is automatically rotated into the process air stream and the wet bed is regenerated. The desiccant is heated, driving the moisture off and returning the bed to a dry state where it is ready for use. In high temperature operations, the air exiting the dryer is cooled prior to passing through the desiccant. Cooling the air makes it easier to remove the moisture. Remember, hot air has a greater capacity to hold water compared to cool air. The filter at the dryer discharge is used to remove fines and dust, preventing them from entering the desiccant bed. As with the other dryers, resin in the dryer discharge cone is not dried the same as the resin in the hopper at start-up because the hot air passes over top rather than through this resin. Before feeding resin to the extruder, resin is drawn off the bottom and recycled to the top.

Dryers are normally top loaded with a vacuum loader that automatically adds new resin as dry resin is removed from the bottom. Dried resin is conveyed with dry air from the dryer to the extruder feed hopper. With hygroscopic resins such as polyesters, the holding hopper above the extruder is a hopper dryer with dry air or is blanketed with dry nitrogen. After the resin is dry, it is essential to keep it dry until it is used.

34.3 Purchasing a Dryer

When purchasing a dryer for an extrusion operation, questions to ask are

- Is the dryer machine mounted, beside the extruder, or a central system?
- Are the controls individual or microprocessor-based?

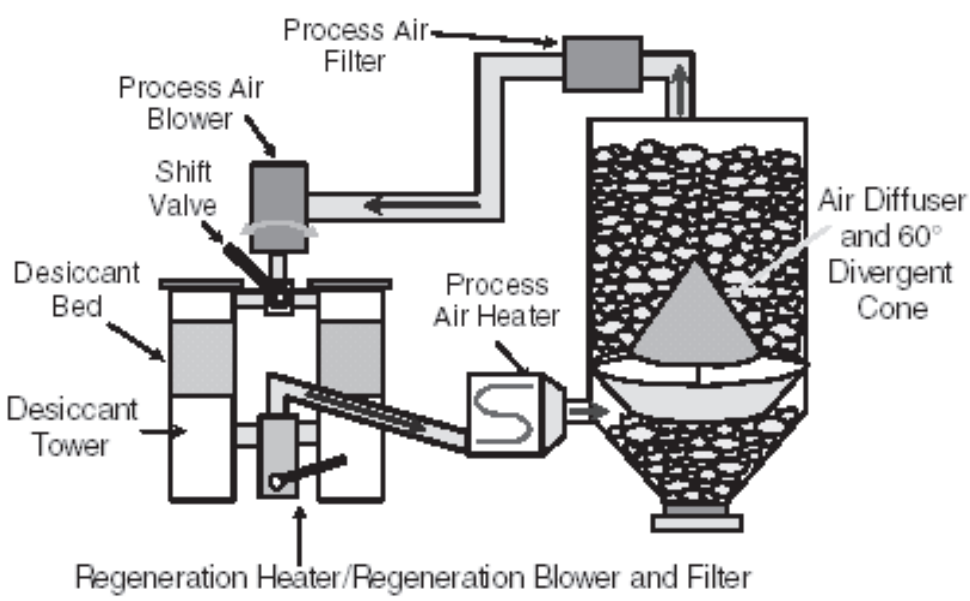

Figure 34.6. Schematic of a refrigerant dehumidifying dryer.

The dryers have already been discussed in detail. Individual controllers have LED readouts with temperature selection, a high-temperature alarm, and an on/off switch. While this simple control is reliable, it does not provide operational flexibility. A start-up timer for starting a loaded dryer early provides dry material when the extruder is ready to start and can save significant time. Microprocessor-controlled dryers monitor dew points, inlet and outlet air temperature, air flow direction during desiccant bed regeneration, automatic start-up and shutdown, and

can communicate with other systems. Some models have optional cycles to save energy by regulating the air flow and the process heaters. Alarms specify what is wrong with the dryer. This helps in troubleshooting when there is a malfunction.

Novel dryer designs have been reported to use less energy through improved dryer insulation and by not cooling down the return air with water.[3] A heat exchanger installed between the hopper exit air and the desiccant bed can be used to capture some of the heat previously lost by cooling the return air.

Another option to consider is how many desiccant beds are needed and their size. Generally the option is for two or more desiccant beds. Multiple-bed dryers are reported to use the desiccant more efficiently.

Transferring the resin from the dryer to the extruder in a closed system with dry air is important when processing highly hygroscopic resins at high throughput. This keeps the resin dry. If room or compressed air is used to convey resin from the dryer to the extruder, it can contain moisture that will be absorbed by the resin. Be sure that any compressed air used for conveying is dry.

Regardless of the number of desiccant beds, the type of controller, or the dryer location, the most important factor for a continuous operation is dryer size. Proper sizing provides sufficient residence time at maximum throughput to supply the extruder with properly dried resin. Therefore, the dryer size must equal the maximum extruder throughput (in lbs/hr or kg/hr) times the number of hours it takes to properly dry the resin.

REFERENCES

1. Arizona Instruments Inc., Tempe, AZ.
2. Stoughton P., *The SPE Guide on Extrusion Technology and Troubleshooting*, Society of Plastics Engineers, Brookfield, CT, 2001, Ch. 5.3, pp. 5–23.
3. Wortberg, Johannes, Walter, Andreas, and Holler, Klaus, Mod. Plastics, *Novel Dryer Designs Seek to Cut Energy Consumption*, Feb. 1998, p. 81.

Review Questions

1. Why is it important to use a desiccant or refrigerant dryer to remove moisture from hygroscopic resins?
2. What polymers are hygroscopic and why?
3. Why should polyamides not be dried too much?
4. Other than resin degradation, what other effects does wet resin have on extrusion?
5. What is meant by dew point and why is it important?
6. Describe the resin drying process and the factors driving the drying process.
7. What resins are not hygroscopic?
8. Why might it be necessary to dry a nonhygroscopic resin?
9. Why might oven drying not be successful with polyester or polyamide resin systems?
10. What are the different types of hopper dryers?
11. How does a desiccant dryer work?
12. How does a refrigerant dryer work?
13. Why is the dryer size important in actual operations? What happens if the dryer is too small or too large?

35 Screen Changers

Automatic screen changers are used in continuous processes where contamination can clog normal screen packs in a relatively short time compared to the production run time. Generally, if a production run can be completed without requiring a screen pack change, an automatic screen changer is not required. There are three basic screen holders:

- Part of sealing system between the extruder and die
- A hydraulic system that switches breaker plates when the extruder is stopped
- An automatic screen changer that supplies fresh filtering media during normal operation

The screen changer or filtering system chosen depends on the process, the throughput rate, the run length, the contamination, and the extruder size. Filtration systems available include

- Breaker plate with screen pack
- Hydraulic slide plate, either fast or slow movement
- Double bolt
- Ribbon-type screen or continuous breaker plates
- Rotary disk filter

Each of these screen changing systems are discussed in this chapter with their advantages and disadvantages.

There are many designs and sizes for wire mesh screens. Most screens are made from steel wire, with stainless steel available for special high-pressure or corrosive applications. A 20-mesh screen has 20 holes or wires per inch, while an 80-mesh screen has 80 holes or wires per inch. As the mesh increases, the hole size decreases, providing more filtration capacity and higher pressure drop, which reduces the extruder output per rpm. Finer filtration is accomplished with smaller screen openings, such as 300 or higher mesh. The screen pack normally consists of more than one screen. If a 120-mesh screen is used by itself, the pressure build-up in front of the screen will blow holes through the screen pack at the breaker plate holes. To prevent polymer from blowing holes in the screen, coarse screens are placed between the fine screen and the breaker plate holes to support the fine screen. Assuming the polymer has to be filtered through a 120-mesh screen, it is common to use a 20/60/120/60/20 screen pack, where the first 20- and 60-mesh screens filter coarse particles, preventing them from accumulating on the 120-mesh screen, and the back 60- and 20-mesh support the 120-mesh screen.

Many screen designs along with mesh sizes are available for different filtration requirements. Some screen weaves are shown in Fig. 35.1. Using the appropriate mesh, a square (A) or twill weave (B) can provide filtration down to 100 microns (μ). Dutch weave (C) filters out particles between 40 and 80 μ in diameter, while a Dutch twill weave (D) filters particles between 8 and 35 μ in diameter. The advantage of Dutch weave is the precise wire separation, minimizing the pressure drop across the screen pack while removing smaller particle sizes from the melt.

In filtration applications, more than one type of screen may be used to provide better filtration. Haver and Boecker[1] recommend up to five filtration layers to make up a given screen pack in Gneuss automatic screen changers. Depending on the screen type and mesh size, one can remove particles from 500 μ to 5 μ.

As screen packs become fouled, the pressure drop increases, the extruder throughput decreases, and the melt temperature increases. Figure 35.2 shows this relationship for a slide plate screen changer. The throughput is at its operating rate with the new screen pack. As the screen pack becomes contaminated, the head pressure

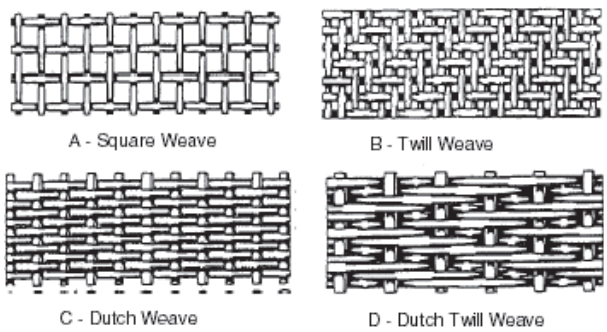

Figure 35.1. Types of weaves for filtration screens.

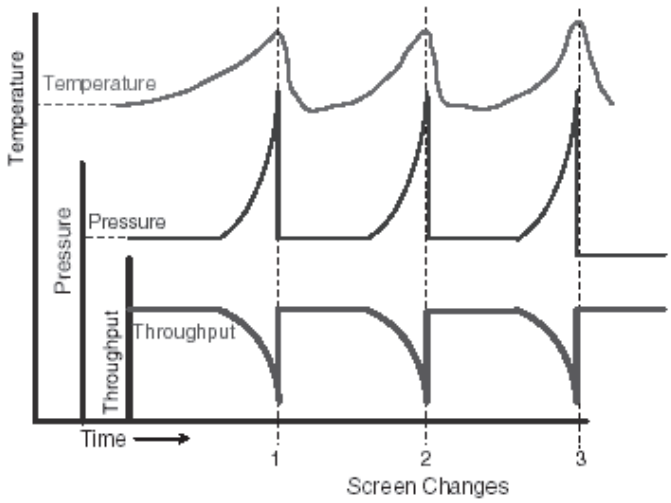

Figure 35.2. Pressure, temperature, and throughput relationship in discontinuous screen changer.

increases, and the throughput starts to decrease. This creates more pressure flowback into the extruder, creating more work, which increases the melt temperature. As the screen pack plugs, the pressure builds rapidly, further decreasing the throughput and increasing the melt temperature. After a new screen pack is inserted, the throughput and pressure return to normal. The melt temperature takes longer to recover, as the extruder barrel needs more time to equilibrate.

Figure 35.3 shows actual pressure increase versus time as highly contaminated material is extruded through an automatic screen changer. The melt pressure versus time shows the rapid pressure increase as the filter becomes contaminated. This extruded material is recycled oil bottles that contained paper and metal. A coarse mesh, square weave screen (12 mesh) clogged in approximately 12 minutes.

The screen changing system required depends on the contamination level. If it takes approximately 10 minutes to plug a screen pack a continuous screen changer is required or the extruder operation will never reach equilibrium and throughput rates will not meet expectations. Screens that clog in 10 minutes are either too fine or the extruded polymer has too much contamination. This operation definitely requires a continuous screen changing system. Screens that plug between 2 and 16 hours can benefit from automatic screen changers. However, if the screen clogs in 12 hours and a normal production run is less than 12 hours with cleanup between runs, then a discontinuous screen changer will meet the process requirements. An economic study is required to determine the value of the downtime saved versus the screen changer costs. If the feedstock is clean with long runs and minimal screen plugging, the payback period will probably not justify a continuous screen changer. However, depending on the extruder size and the downtime required to change screen packs, using a continuous screen changer may be justifiable.

35.1 Breaker Plate with Screens

A breaker plate's function in single and twin screw extrusion is to hold the screens while providing a seal between the extruder and die. Scratches, dents, or marred surfaces on the breaker plate can lead to leakage. If the breaker plate has been machined to remove defects, verify that it is still thick enough to act as a seal. Figure 35.4 shows a breaker plate with and without screens. To change a screen pack in a breaker plate, it is necessary to remove the die, the breaker plate, and screens. After new screens are inserted in the breaker plate, the extruder piping is reassembled. This is the most time-consuming screen pack system. Depending on the extruder size and the die removal difficulty, downtime can range from 15 minutes to an hour or more.

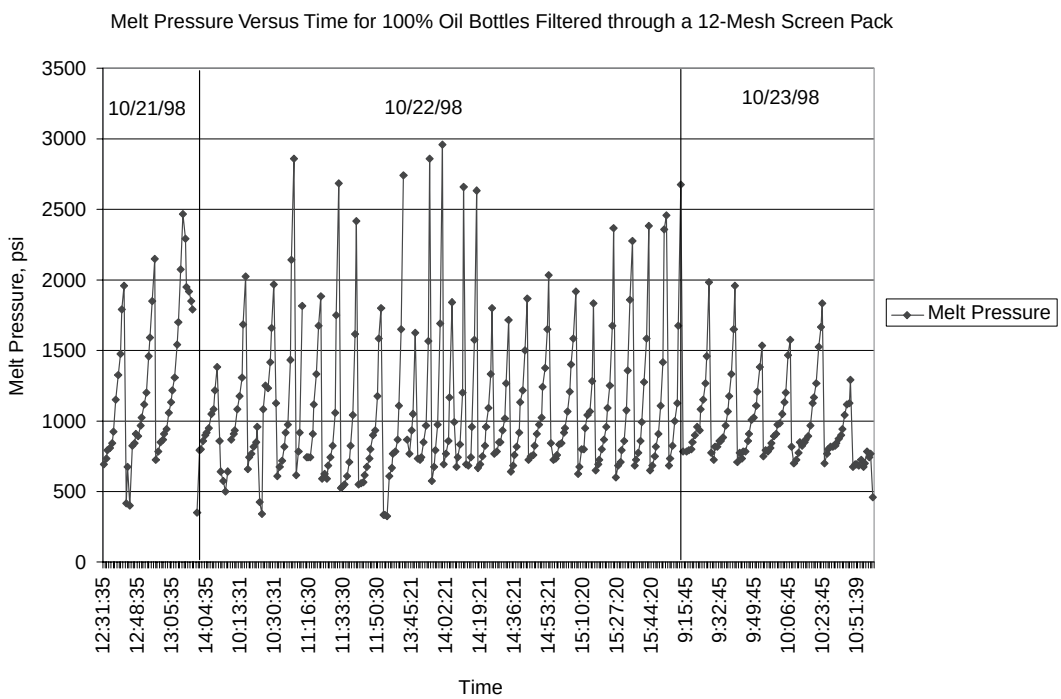

Figure 35.3. Melt filtration of recycled oil bottles with automatic screen changer.

35.2 Manual Screen Changer

Manual screen changers are available with two breaker plates mounted in a metal plate close to one another. With a long handle, one breaker plate and screen pack is moved out of the melt stream while the second one is inserted into the melt flow path. This system is inexpensive and works very well in a discontinuous operation. Figure 35.5 shows a Berringer screen changer[2] with a clean breaker plate out of the melt stream and the handle to move it in line with the extruder.

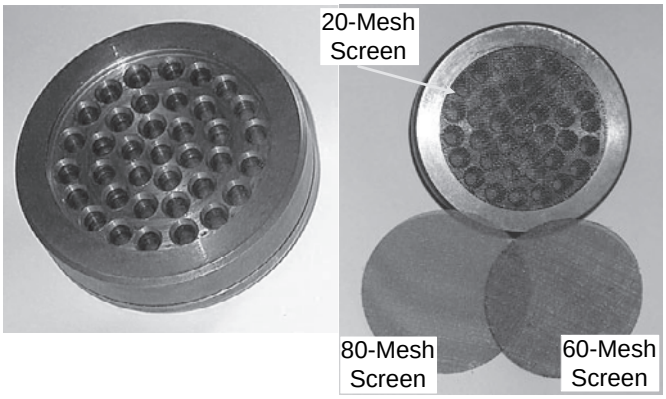

Figure 35.4. Breaker plate and screens for 2.5 inch single screw extruder.

35.3 Hydraulic Screen Changer

A hydraulic screen changer can function as either a continuous or discontinuous changer, depending on the transfer time from one breaker plate and screen to the other. In a slow transfer operation, the extruder has to be shut down while the slide plate is moving between positions, creating a discontinuous operation. The dead space between the two breaker plates during slow transfer could generate high head pressure and blow the rupture disk if the extruder ran during transfer. If the transfer is rapid from one position to the next and the distance between the breaker plates is less than the breaker plate diameter, the process can continue to run in a continuous, non-steady-state operation. When the dead space between the breaker plates is small, the clean breaker plate starts into the melt stream before the dirty one is completely removed. This prevents rapid pressure rise in the extruder. Figure 35.6 shows a hydraulic screen changer with a clean screen and breaker plate next to the hydraulic piston. When the active screen becomes contaminated, the hydraulic cylinder is activated. It pushes the clean screen breaker plate into the melt stream, and the dirty one exits from the other side. The dirty screen pack and breaker plate are removed and replaced with a clean screen pack and breaker plate. The configuration in Fig. 35.6 shows a horizontal hydraulic cylinder moving from side to side. Depending on the extruder configuration and downstream equipment, the hydraulic screen changer can be installed vertically. The melt stream is free from any dead space that might cause resin degradation over time. All adapters and flanges have to be heated and securely tightened to the extruder and die to prevent polymer leakage. The screen changer body is heated with cartridge heaters to maintain temperature. It is possible to use a liquid heat transfer medium if required.

Figure 35.7 shows a rapid transfer, continuous operation hydraulic screen changer made by Dynisco®.[2] The large cylinder is necessary to supply the hydraulic fluid to rapidly move the slide plate from position 1 to position 2. The entrance to the screen changer is the same diameter

Figure 35.5. Berringer manual screen changer.

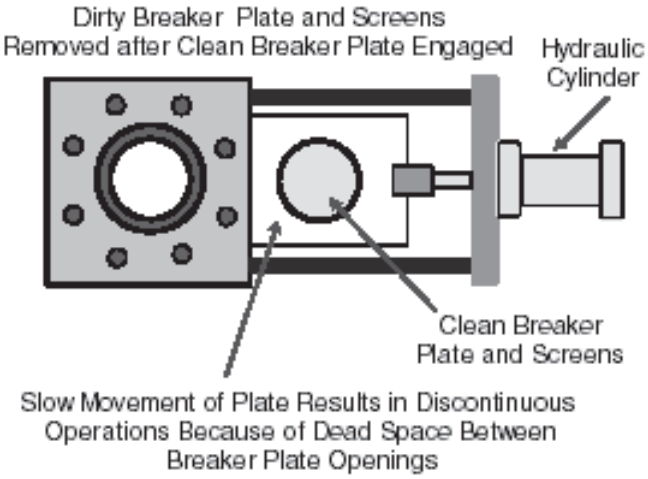

Figure 35.6. Schematic of hydraulic screen changer.

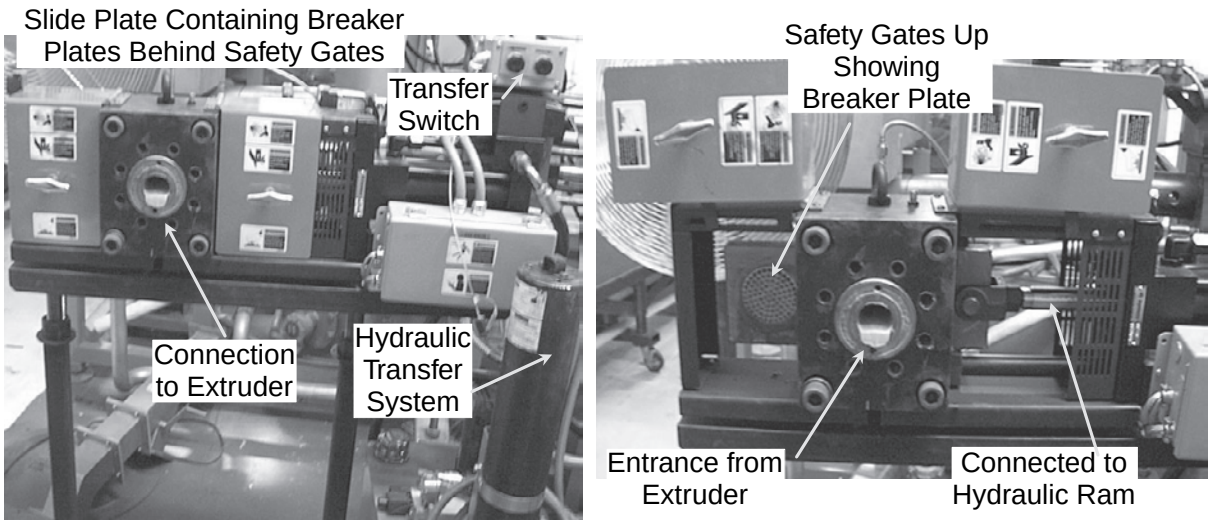

Figure 35.7. Continuous Dynisco hydraulic screen changer.

as the extruder. The exit side is normally smaller than the extruder bore to maintain the resin velocity in the transfer pipe to the die. Figure 35.8 shows a reduced bore diameter on the downstream side of the screen changer.

35.4 Double Bolt Screen Changer

A double bolt continuous screen changer has two round bolts or hydraulic cylinders with breaker plates and screen packs that move in and out of the melt stream. Options are available for the melt to flow through both bolts simultaneously or individually. Figure 35.9 shows a double bolt screen changer.[3] In both pictures the cylinders are completely retracted with no flow through either bolt. The breaker plate opening and screens are visible. The melt flows through the adapter and into the flow channel in the double bolt housing. The melt passes through the screen and into another adapter connected to the die. The screen changer housing is heated to maintain the polymer temperature and prevent freeze-off if the extruder stops. A pressure transducer, mounted in the adapter between the extruder and screen changer, indicates when the screens are contaminated and need to be replaced. In Fig. 35.10 both screens are in the polymer flow path.

Two operation modes are available. In the first mode, one filter is used while keeping the second as a spare to replace the first when it becomes dirty. The second operational mode uses both screens simultaneously. In the first operational mode, there is less filtration area, leading to more rapid screen plugging. One filter becomes dirty; the second bolt with a clean screen is slowly brought in line, allowing the air to escape from the bolt along with some polymer melt. This provides continuous operation and prevents trapped air in the bolt from exiting the die. The dirty screen is removed from bolt one and replaced with a new screen pack. Melt is pumped through the second screen and bolt until that screen becomes dirty. Once the second screen is dirty, the first bolt with a new screen is slowly transferred back into line and the second one removed and cleaned.

In the second operational mode, both screens become clogged simultaneously. Through the use of a back flush operation, screens can be cleaned one at a time by forcing polymer back through the screen to remove contaminants. As the first screen is transferred to the back flush position, polymer from the second slide bolt flushes contaminants from the screen. The first screen is transferred to its normal operation position while the second screen is back flushed to remove contaminants. The advantage of using both filters simultaneously is having twice the filtration area.

During the transfer operation, excess polymer is cleaned off the slide bolt. This is usually done with copper gauze. As the bolt is transferred back on line, it is stopped before complete engagement to allow the air to escape. Polymer fills the breaker plate cavity, forcing the air out. After some polymer weeps out, the cavity is completely filled and the slide can be completely engaged. If this procedure is not followed, air flows to the die, where holes form in the extrudate and possibly cause material breaks.

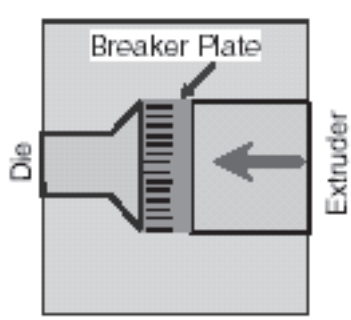

Figure 35.8. Screen changer cross section.

In a continuous operation, an option is available to back flush the screen pack to remove contamination. This procedure is

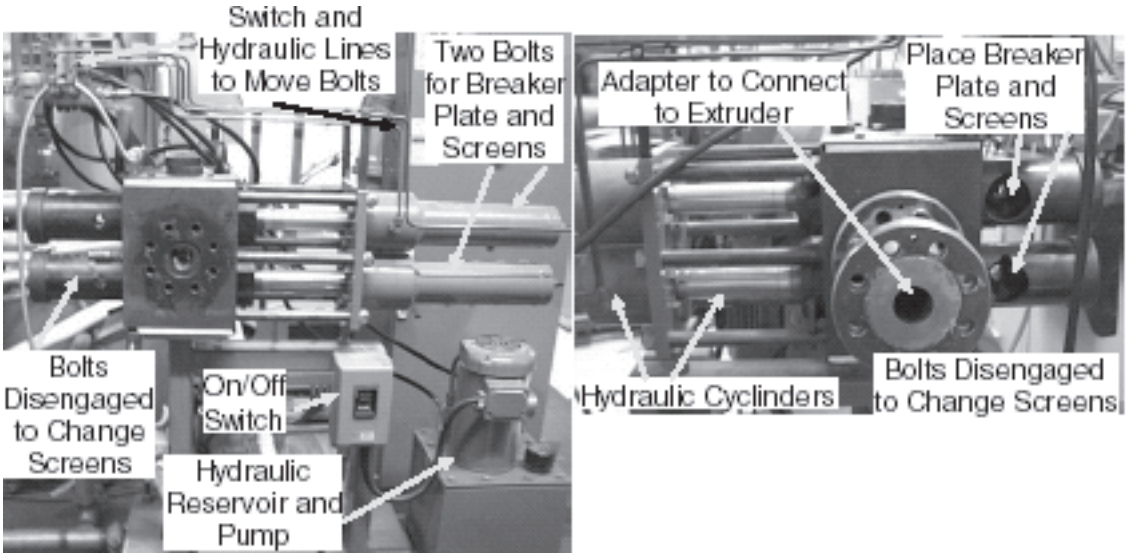

Figure 35.9. LCI double bolt continuous screen changer.

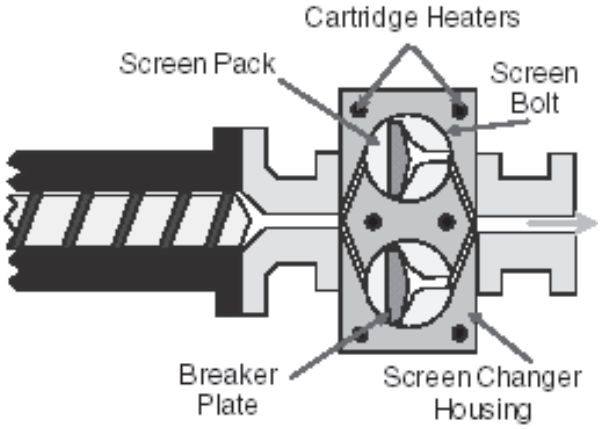

Figure 35.10. Schematic double bolt continuous screen changer.

used in applications requiring long life or in recycle applications where the screens foul rapidly (see Fig. 35.3). With the slide in the back flush position, polymer flows through the screen from the back side, loosening the contamination and conveying it out of the cavity. After the back flush operation is complete, the cleaned screen goes back on line. Back flushing can be done many times and it uses little material. Typical material loss in a back flush operation is 1 to 1.5 times the slide cavity volume. Programmable controls are available to let the operator know when the back flush is complete and approximately how many back flushes can be done.

Figure 35.11 shows the double bolt screen changer slide positions as the screens are changed and reengaged. Positions 2 and 3 allow for venting of air and the polymer

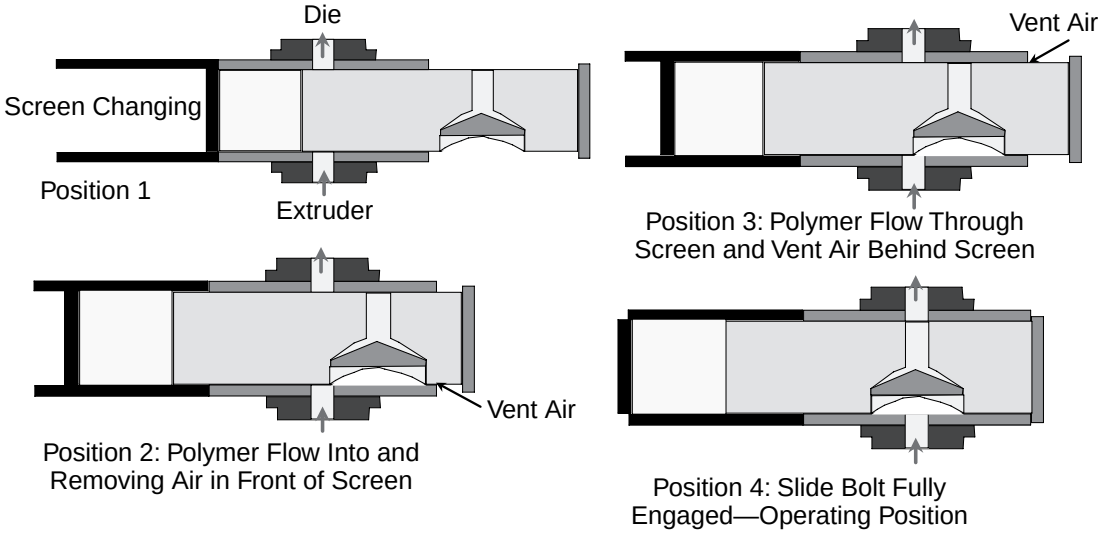

Figure 35.11. Slide bolt positions—changing screens and reengaging the bolt.

to prevent any air from passing through the die. Figure 35.12 shows the slide bolt position in a back flush operation. In the back flush operation, both slide bolts are used simultaneously, with polymer from one flushing the other. Figure 35.13 shows the process for changing screens with a slide bolt designed for a back flush operation.

35.5 Ribbon-Type Screens

Ribbon-type screen changers use long or continuous screen material that moves through the polymer flow channel. The screen is pulled by a hydraulic clamp or pushed through the melt by the melt pressure. As the melt pressure increases, it pushes the screen forward to provide clean screen for filtering. This is a semicontinuous process, with the melt pressure acting as the driving force to provide fresh screen. The screen is a continuous ribbon that is sealed to prevent polymer leakage by cooling the ribbon where it enters and exits to form a polymer seal. These systems provide continuous filtration with minimal moving parts and a simplified system. Figure 35.14 shows a ribbon-type screen system. Throughput, pressure, and melt temperature variations are less than shown in Fig. 35.2, which is more typical of hydraulic breaker plate units.

Disadvantages of these systems are that a contamination surge can clog the filter and, if the seals fail, large amounts of material will flow onto the floor.

35.6 Rotary Disk Screen Changer

A rotating disk has 10 to 12 screens in a ring pattern around the disk; they are indexed in steps as small as one degree as screens become contaminated. The indexing

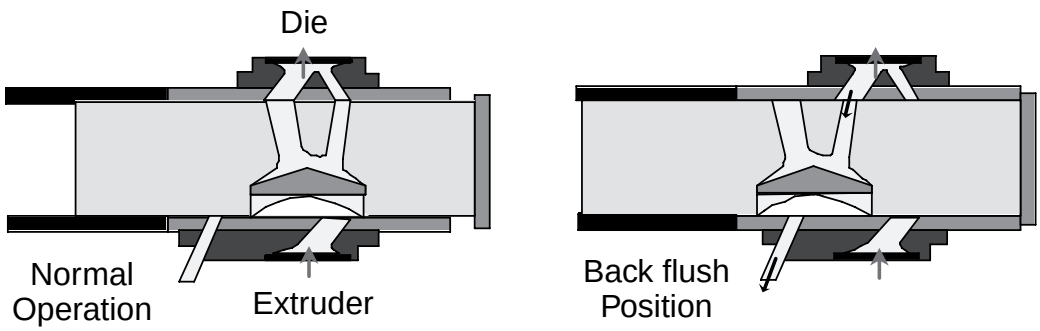

Figure 35.12. Back flush operation—polymer is fed from second bolt.

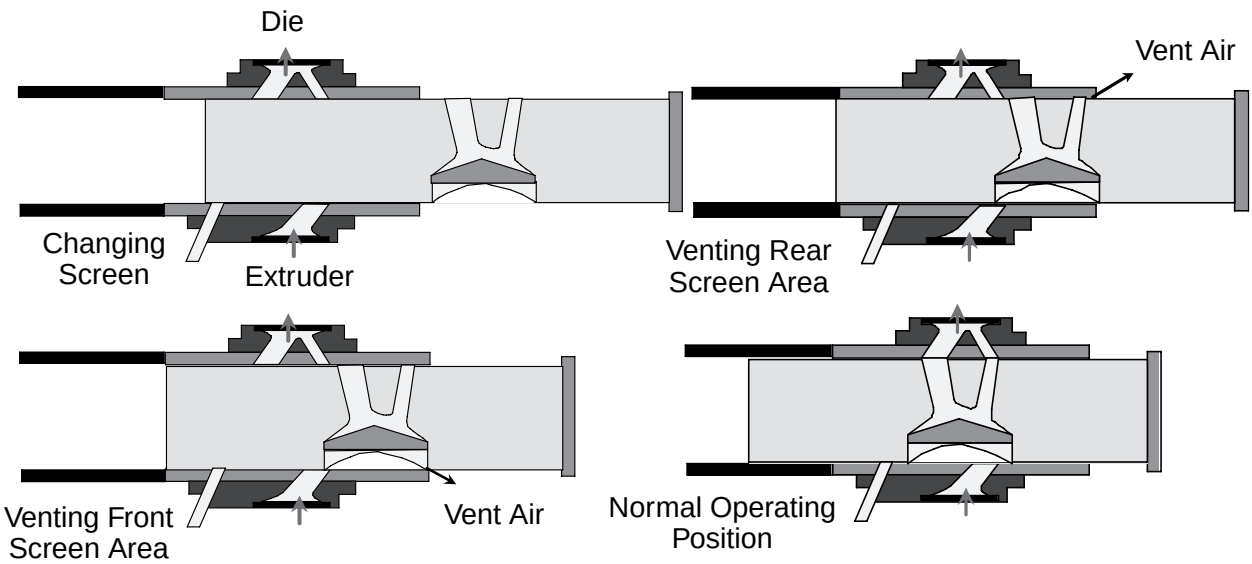

Figure 35.13. Changing screens in a back flush slide bolt screen changer.

sequence is very rapid. Figure 35.15 shows a rotary disk screen changer.[4] Constant melt pressure is obtained during operation, through indexing the rotary disk to supplying clean screens. The microprocessor control determines the index rate by monitoring either the pressure before the screens or the pressure differential across the screens. With a continuous rotary screen changer and microprocessor control, the graph shown in Fig. 35.2 produces three flat lines for throughput, pressure, and melt temperature, resulting in a stable process.

In operation the melt passes directly from the extruder through the screens to the die or adapter in the shortest possible path. The flow path is uninterrupted as the screens index. Screen packs are fitted over breaker plates in the rotary disk to provide filtration down to 6–8 μ. As new cavities in the rotary disk move into the melt stream, the air is vented as the screen cavities are filled to prevent air getting into the polymer melt, which would create downstream extrudate interruptions. A back flush channel forces small amounts of filtered melt back through the screen to purge contaminants from previously used screens. The purging valve stays open for a short time (1–2 seconds per stroke), minimizing the lost polymer while keeping the screens clean. The screen area purged is equal to the contaminated screen indexed forward. Rotary screen changers without a back flush require the dirty screens to be removed and replaced as the screen rotates.

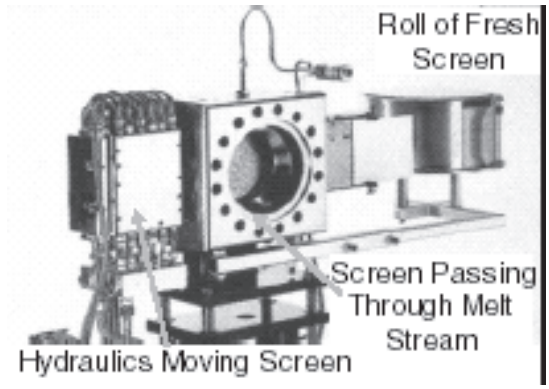

Figure 35.14. Ribbon screen changer.

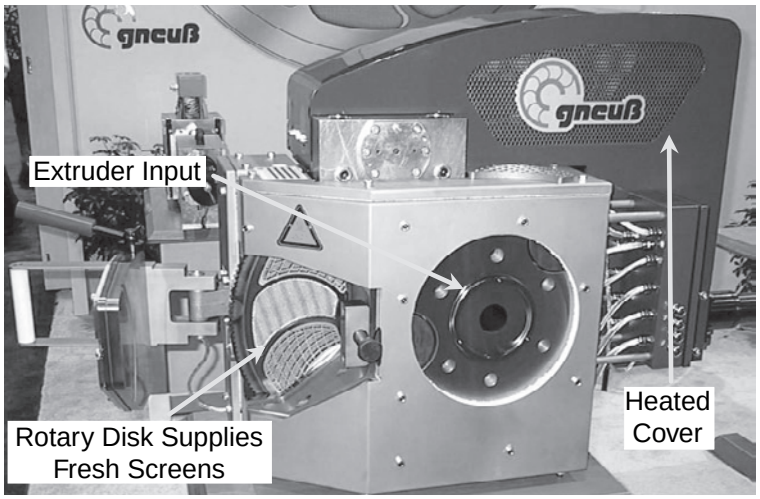

Figure 35.15. Gneuss continuous rotary filtration.

Heating the screen changer is accomplished with cast heaters and heater cartridges strategically located around the screen changer. Individual heater zones provide better heating efficiency and uniformity. Heat-transfer fluids can be used if required for the application. Disk indexing is accomplished through either a hydraulic or a pneumatic drive. Machining precision prevents these units from leaking, even at high pressures.

While the rotary disk screen changer, shown in Fig. 35.15, has continuous screens and cavities around the circumference of the disk, some disks contain distinct breaker plates with dead spaces between them and are indexed at discrete distances to provide a completely new filter medium for each index step. In this scenario, the throughput, pressure, and melt temperature graphs will be similar to Fig. 35.2 and not relatively flat as obtained with the continuous rotary disk screen changer. Between these two scenarios are rotary disks with discrete breaker plates and screen packs with minimal dead space between the breaker plates. As the disk is indexed, new filter medium is available, so the throughput, pressure, and melt temperature curves are flat even though distinct breaker plates and screens are used. The throughput, pressure, and melt temperature curves depend on the dead space between the breaker plates.

Continuous rotary disk screen changers are used in high-quality applications requiring gentle treatment of the melt. Potential application areas include foam sheet extrusion, fibers, compounding and PVC processing, tight tolerance finished parts, and processes requiring only limited pressure drops, such as resin polymerization and fiber production.

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2. Dynisco®, Franklin, MA.
3. Kreyenborg GmbH Maschinenfabrik Coermühle 1-5 48157 Münster/Germany.
4. Gneuss Inc., 921-A Matthews Mint Hill Road, Matthews, NC.

Review Questions

1. What is the difference between a continuous and a discontinuous screen changer?
2. What are some types of continuous screen changers?
3. What types of continuous screen changers produce changes in melt temperature, melt pressure, and throughput as the screens become clogged?
4. Why do some rotary disk screen changers show no increase in melt temperature or melt pressure or decrease in throughput as screens become dirty?
5. How does a double bolt screen changer work?
6. What is the purpose of the back flush in double bolt and rotary disk screen changers?
7. What are some different screens and how do the weaves differ?
8. What is a ribbon-type screen changer and how does it operate?

36 Gear Pumps

Gear pumps provide consistent polymer flow, which results in more consistent product dimensions. Potential advantages associated with gear pumps include

- Higher product yield per pound of material from more consistent gauge control, allowing an overall reduction in the average product dimensions.
- Increased extruder output attributed to reduced pressure flow backward into the extruder caused by high die pressure. Equation (36.1) shows that extruder output decreases as pressure flow from high head pressure increases.

$$\textit{Extruder Output} = \textit{Drag Flow} - \textit{Pressure Flow} \tag{36.1}$$

- Potential increased regrind use by eliminating extruder surging from nonuniform regrind feed.
- Start-up time reduction as the gear pump automatically adjusts the screw speed while providing the desired constant die pressure.
- Energy reduction as the extruder operates at lower backpressure or pressure flow. The lower pressure flow generates higher throughput at lower screw speed and less shear heat. Lower shear heating generates less heat that has to be removed through cooling.

A gear pump is a positive displacement pump consisting of two gears, four bearings, internal heating, shaft seals, and very precise speed control. There are multiple sizes and speed ranges that provide multiple output ranges. They deliver a uniform and homogeneous extrudate to the die at a constant pressure. The gear pump is used to provide the required pressure and uniformity rather than the extruder and die. A gear pump produced by LCI[1] is shown in Fig. 36.1.

Gear pumps are located between the extruder and the die. If an automatic screen changer is used, the screen changer is installed between the extruder and the gear pump. A screen changer installed after the gear pump compromises the gear pump stability. As the screen's filter contaminates, the pressure drop to the die exit increases. This changes the melt temperature and the load on the gear pump.

Heated adapters and/or transfer pipes are required to connect the equipment together. When no screen changer is present, a short adapter is placed between the extruder and the gear pump, with a second adapter possible between the gear pump and die. In some situations, the die is connected directly to the gear pump. Transition or transfer pipes should be kept as short as possible. Figure 36.1 shows an adapter or transition piece at the entrance to the gear pump, allowing the pump to be connected to a twin screw extruder. A short adapter has a similar center bore on one end to match the extruder diameter, with the bore at the opposite end matching the gear pump diameter. The adapter between a twin screw extruder and the gear pump transitions from a figure 8 at the extruder to a circular shape at the gear pump. This is commonly referred to as an 8–O transition. Depending on the die entrance, a simple round shape on each end of the transition piece may suffice. It is critical to prevent dead spaces and that all sharp transitions within the adapters are properly machined to minimize the potential for polymer build-up and degradation over time. If the transition pipe or adapter is long, the center bore is reduced in size with a smooth transition from the extruder exit diameter to the adapter diameter, and from the adapter diameter to the gear pump entrance diameter. The transition pipe diameter is reduced to increase the polymer velocity in the transition pipe, reducing polymer hold-up time.

Installing a gear pump after an automatic screen changer may require three adapters or transition pieces between the extruder and screen changer, the screen changer and gear pump, and the gear pump and die. Three pressure transducers are required to monitor the pressure before the screen pack, before the gear pump, and before the die. The first pressure transducer monitors the screen pack pressure and tells when the screens need changing. The second transducer monitors the pressure entering the gear pump. As the pressure increases or

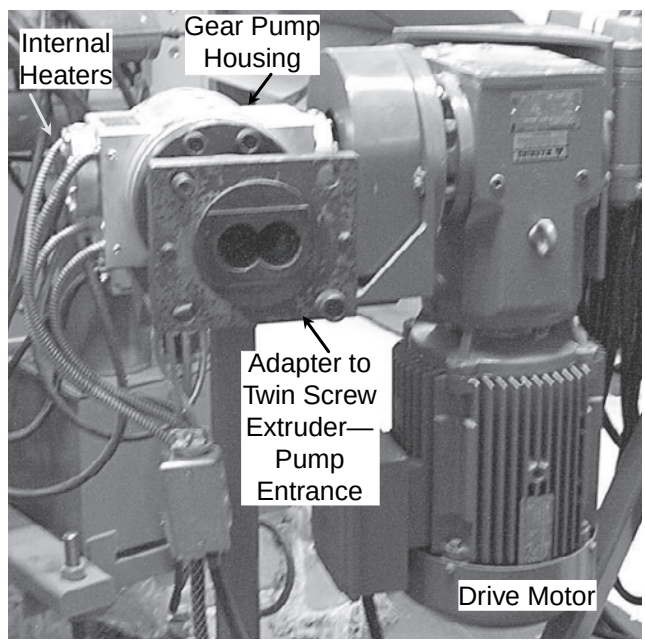

Figure 36.1. LCI gear pump.

decreases, the gear pump sends a signal back to the extruder to increase or decrease throughput. In a single screw extruder this is accomplished by changing the extruder screw speed, while in a starve-fed twin screw extruder the signal increases or decreases the feed rate. The second transducer monitors the melt as it is being delivered to the gear pump. The third pressure transducer monitors the gear pump output pressure or the pressure entering the die. Figure 36.2 shows an extrusion setup with an automatic screen changer, gear pump, and adapters.[2]

36.1 Gear Pump System

A gear pump system is composed of

- Pump
- Drive system
- Pump controller that interfaces with the extruder
- Safety interlocks to protect the pump and operator

The pump housing is normally fabricated from standard 4140 carbon steel, stainless steel, or carbon steel alloys. Stainless steel is used with corrosive materials; it costs more and has lower strength, so it is only used in specialized applications where required. As a gear pump is a pressure vessel, the housing thickness is dependent on the anticipated application pressures. In addition to pressure, the housing is operated at high temperature with torsional forces provided by the rotating gears. Generally, the heavier the housing, the higher the forces the pump can withstand.

The pump has to be heated to prevent polymer from solidifying in the gear pump during a process upset and process start-up. In normal operation shear heat is generated by the rotating gears, minimizing the heat required. Unlike an extruder, the gear pump has no external air or water cooling to remove excess heat. Consequently the heat required during the process depends on the heat lost to the environment through conduction and radiation and the shear heat generated by the rotating gears. Pumps are normally heated with electrical resistance heaters located at the four corners and passing through the pump housing, as shown in Fig. 36.3. Larger pumps may require more heaters to provide uniform heat. In very large pumps or where heating is critical, fluid heating is used to provide more uniform heat. Circulating fluids have the advantage of either adding or removing heat, as required to keep the system in equilibrium. Internal coring for heater fluid or jacketing is essential for processes where material temperature sensitivity is critical and heat has to be removed from the process. Obviously fluid heating requires an auxiliary oil heater and hoses to heat and transfer the fluid between the pump and oil heater. Other heater alternatives include band or cast aluminum heaters. These need to be properly insulated to prevent operators from burning themselves as they work around the equipment.

Each pump has two gears, one that is driven and one that follows and is called the idler gear. The end of the drive gear is shown in Fig. 36.3. This shaft goes completely through the pump and is connected by a spline or keyway to the motor or drive. Normally the gear is an integral part of the shaft, as shown in Fig. 36.4. The idler gear is driven by the drive gear. There are four bearings, one at the end of each shaft, that are lubricated by molten poly-

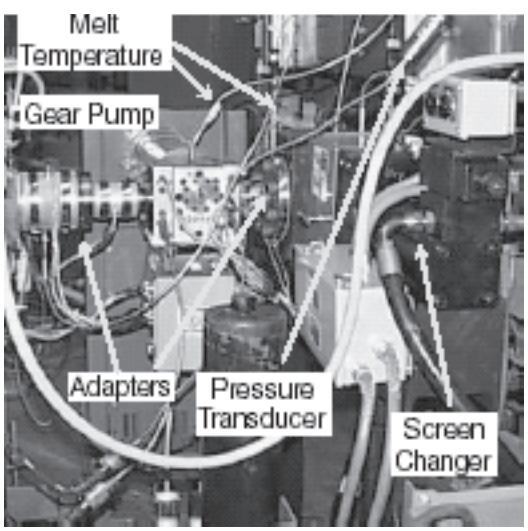

Figure 36.2. Dynisco gear pump and screen changer.

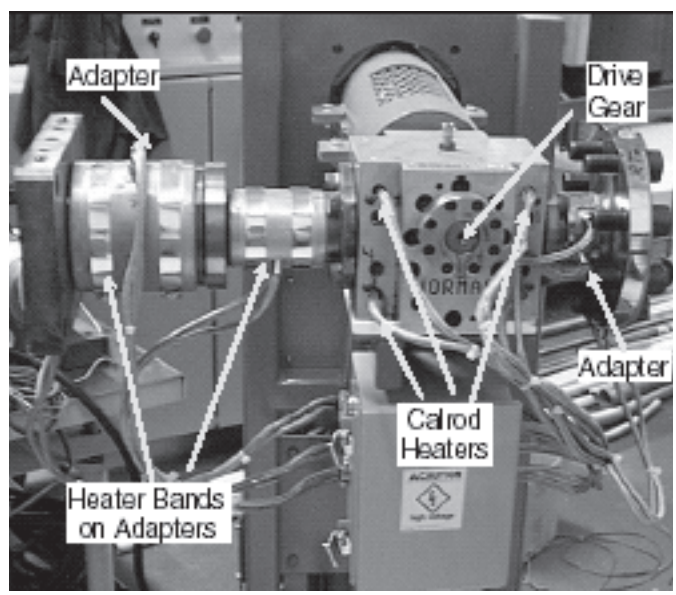

Figure 36.3. Calrod heaters and adapters on Dynisco gear pump.

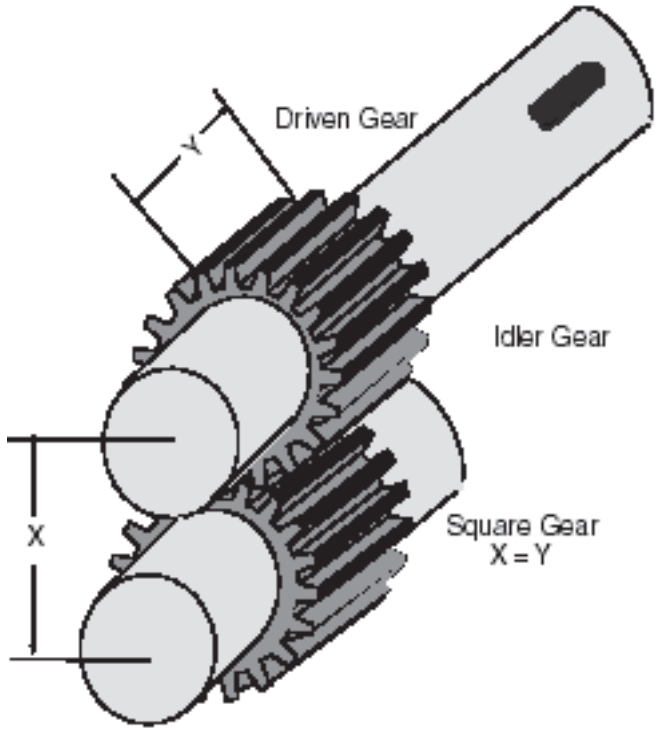

Figure 36.4. Square gears on a shaft with a key.

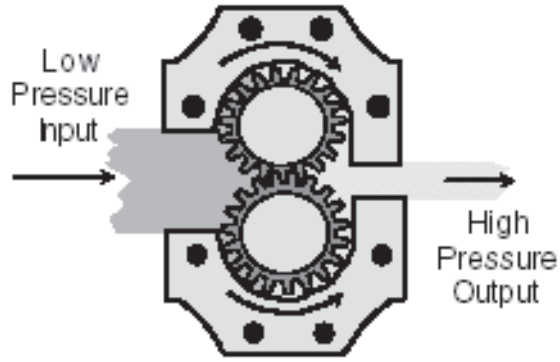

Figure 36.5. Cross section of gear pump.

mer. Operating the system without polymer will damage the bearings and the pump. Figure 36.5 shows a pump cross section with polymer present. The gears rotate in opposite directions, picking up the polymer and positively conveying it to exit around the outside of the gears. Normally the input side is at low pressure and the output at high pressure. As the polymer passes around the outside of the gear, lubrication grooves convey molten polymer through the bearings and return it to the low-pressure side of the pump. This provides the necessary lubrication to prevent bearing wear.

The spur gears shown in Fig. 36.4 are referred to as square gears, where the center distance, *X*, between the drive gear and the idler gear, equals the gear width, *Y*. In most extrusion applications, square gears are used. The primary gear tooth patterns are either spur, shown in Fig. 36.4, or helical. Spur gears are more volumetrically efficient, cheaper to manufacture, and load the bearings more evenly. Spur gears have minor pulsation as the material exits the pump. This is caused by the gear teeth and is similar to the spiraling action of molten polymer exiting the extruder due to the screw flight. While pulsations are an interesting discussion point, in actual operation this is a minor problem. A 12-tooth gear pump running at 100 rpm translates to 1200 pulses/minute or 20 per second. At this rate, the pulses tend to level themselves out. If pulsations are a problem, helical gears can be used in place of spur gears. Advantages of helical gears are better wiping action of the gear root and lower pulsations. Helical gears do provide higher and uneven loadings on the thrust bearings, used to keep the gears in the housing. Figure 36.6 shows typical pulsations. The pulsation frequency is the same (assumes both tests run at the same speed), while the amplitude is lower with the helical gear. Using four helical gears rather than two lowers the pulsation amplitude farther. Note the time scale shown in Fig. 36.6 is very short. Figure 36.6 is a graphical representation of a 12-tooth gear run at 50 rpm, resulting in 600 pulsations per minute or 10 per second. This pulsation frequency and amplitude is normally smoothed out in the die and never visible in the final product.

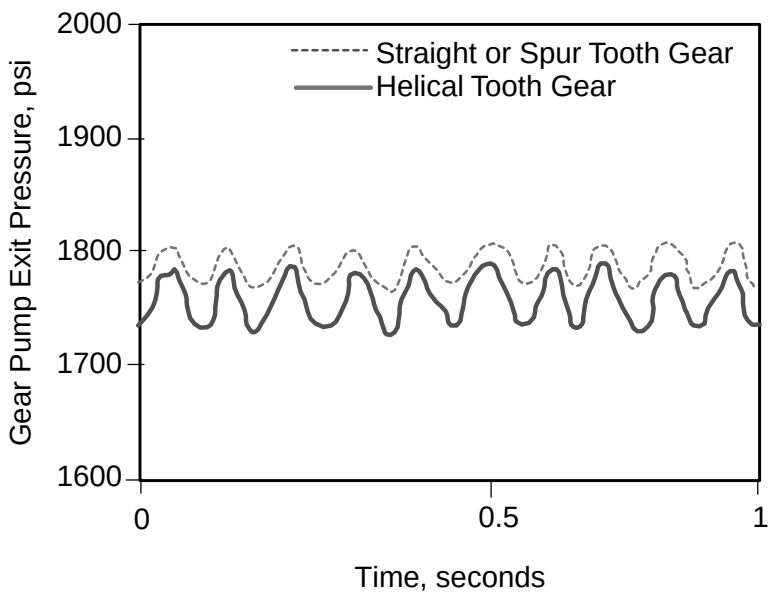

Figure 36.6. Typical pulsations from a gear pump with different types of gears.

In operation it is critical to fill the gear teeth with molten polymer. As the gear rotates, suction draws polymer into the top area between the teeth, while melt compression is required to completely fill the area between the teeth. A compression zone at the pump intake forces the polymer to completely fill the gear gap between the teeth.

For most plastic applications, square gears are used with either a spur or helical tooth configuration. For applications where the pressure differential between the inlet and outlet is more than 4000 psi, non-square gears may work better. The gear width or dimension *Y* in Fig. 36.4 is decreased, reducing the bearing pressure and the gear teeth load. Nonsquare gears are more expensive to build and less energy efficient.

Most gears and shafts are one-piece design to provide better reliability in the high-torque environment where they are used. Gears are either case hardened, through hardened, or a combination of both case and through hardened. Characteristics of each gear are

- Case-hardened gears are produced by applying a thin skin of a hard material over a soft metal. Processes such as ion nitriding produce a thin, hard metal coating of approximately 0.015 inch thickness over a soft metal. Nitriding provides a hardness of approximately 74 Rc over a gear with a base hardness of approximately 35–40 Rc.[3] Once the nitride coating is worn away, the soft metal will wear very quickly. If the hardened coating is broken, it can chip or break off easily because of the softer metal underneath. Case-hardened gears provide excellent torque transfer with less tendency for breakage under load.
- Through-hardened gears have typical Rockwell hardness of 55–65 Rc.[3] While these gears have improved wear resistance without being subject to the surface cracking of the case-hardened gears, through-hardened gears are more brittle.
- Case hardening hardened gears provides the optimum gear. The base material may not be as hard as the through-hardened gears but have higher hardness than the case-hardened gears. Base metal hardness might be in the range of 52–58 Rc. This design makes the gears more resistant to surface cracking while providing the maximum wear resistance.

Sleeve bearings on each shaft hold the shafts in place, with recirculating polymer providing lubrication. Gear pumps should never run dry because without polymer lubricant, the high bearing loads can lead to galling and bearing failure. In operation, plastic melt is recirculated from the high-pressure to the low-pressure side through the bearing as a thin lubricating film. A combination of the bearing design, gear rotation, and pump differential pressure produces the necessary polymer flow through the bearing sleeve. Bearings are usually hardened tool steel. Polymer melts containing abrasive materials can accelerate bearing wear, requiring the correct bearing specification for pumps processing abrasive polymeric compounds.

36.2 Drive and Control System

The drive and control system is designed to interface the gear pump operation with the extruder screw speed to provide constant pressure and output from the die. Pump drives use either a DC or AC motor. As with extruders, DC drives dominate the market, but AC drives are becoming more common. Either drive delivers the accuracy needed to provide constant output. DC drives are available with and without brushes. The more common DC drive uses brushes, as this has been the industry standard for years. The only negative of this system is the periodic need to replace the brushes. Brushless DC motors are more expensive and there are fewer of them in the field. AC drives with digital feedback provide very good control, are low cost, and are finding a home in pump applications. Basically any motor provides adequate control, assuming it is properly sized and has the accuracy required. Motor selection depends on purchaser preference. Motors are positioned in various locations around the gear pump based on available space and the space required by the operators. Some motors are underneath the gear pump; some are close to the gear pump, hanging off the gear pump on the nonoperating side (Figs. 36.1 and 36.3); and some are near the floor (Figs. 36.7[4] and 36.8[5]). The optimum motor location depends

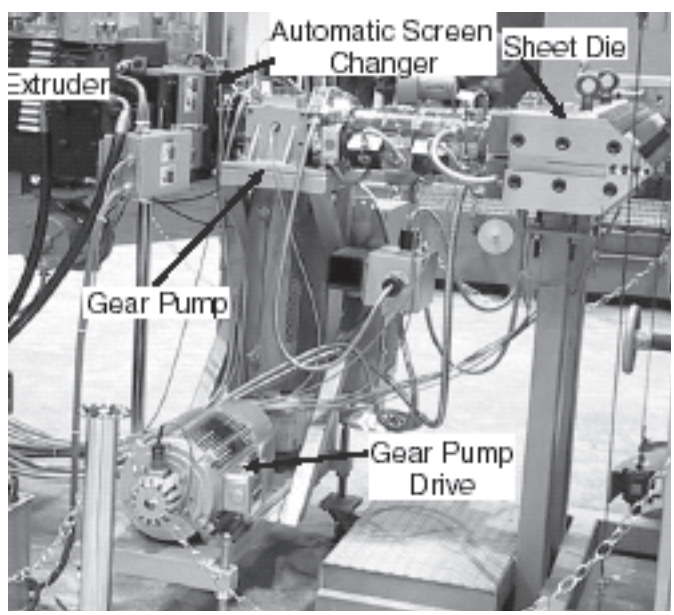

Figure 36.7. HPM extruder connected to screen changer, gear pump, and sheet die.

on the particular extruder and downstream equipment configuration.

The control system acts as the interface between the gear pump and the extruder. In single screw extrusion, the gear pump entrance pressure is controlled. An increase or decrease in pressure sends a signal back to the extruder drive controller to decrease or increase the extruder screw speed to maintain a constant delivery pressure to the pump. The gear pump speed maintains a constant volume output to the die. Figure 36.7 shows a single screw extruder, automatic screen changer, gear pump, and sheet die.

Twin screw extrusion requires the inlet pressure signal to go to the feeders, increasing or decreasing the feed rates. With multiple feeders, this requires numerous signals to be tied together. In addition, the time required for the increased pressure from higher feed to get from the feeder to the gear pump can be minutes. Significant increases in feed rates can cause the extruder to exceed the torque limits and shut down. In some twin screw extrusion applications, the gear pump may not be tied back into the extruder or feed loop control.

Corotating, intermeshing twin screw extruders do not generate high head pressures, so processes that require high die pressure sometimes use gear pumps to provide adequate pressure and flow uniformity. Gear pumps are commonly found on corotating, intermeshing twin screw extrusion lines used to simultaneously compound resin formulations and produce sheet or film products. In this situation, the gear pump may not be tied back into the extruder or feed system, but may be a standalone unit in the line to provide appropriate and constant pressure in the die to ensure better gauge control. Figure 36.8 shows a co-extrusion sheet line on a twin screw extruder with a gear pump to provide uniform flow from the main extruder to the feed block.

The control system can use discrete controllers to control individual equipment, as in Fig. 36.8. Figure 36.9 shows a discrete controller for a Dynisco automatic screen changer and gear pump. Each controller runs independently, and it is up to the operator to synchronize the equipment operation to make the process run correctly and efficiently. A microprocessor-based computer control interconnects the gear pump and extruder, similar to the system described above for single screw extrusion and shown in Fig. 36.10. Discrete controllers cost less and are easy to install, but

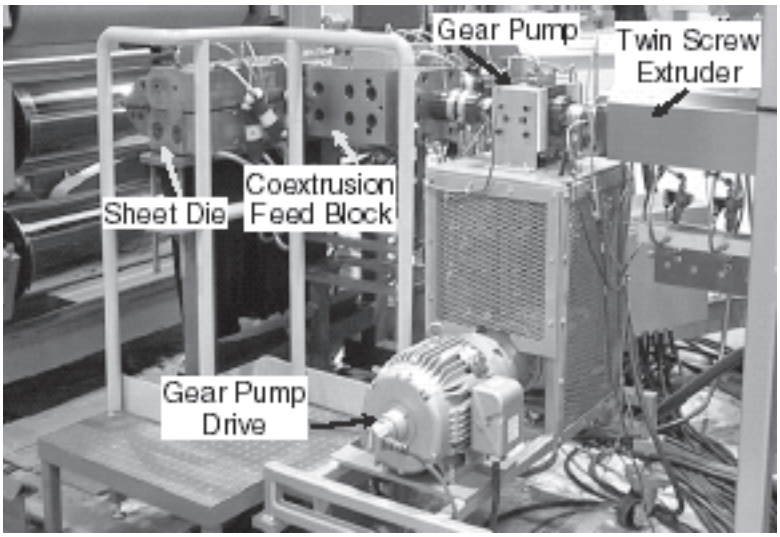

Figure 36.8. American Leistritz twin screw extruder with gear pump and sheet die.

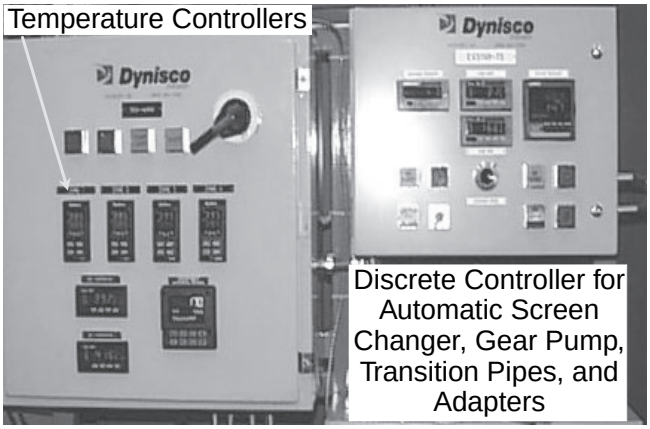

Figure 36.9. Discrete Dynisco controller.

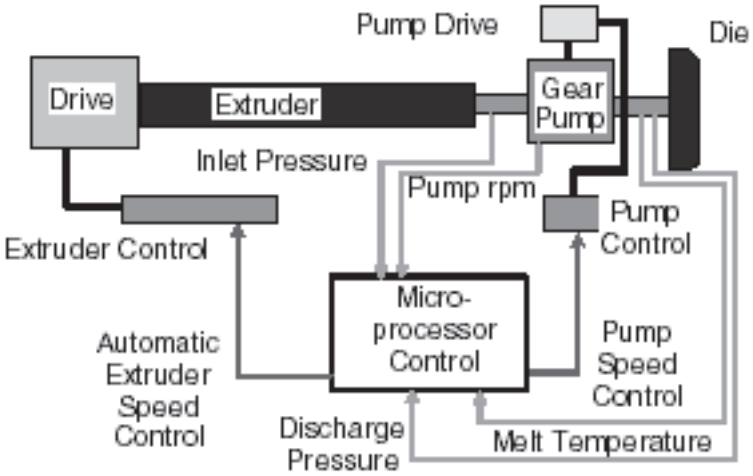

Figure 36.10. Gear pump microprocessor control schematic.

the system is not expandable. Advantages of microprocessor-based controllers are

- Closed loop control with an automatic switch that allows the computer to operate both the extruder and gear pump at the desired output
- Data collection for statistical process control
- Recipe storage for different processes, reducing product conversion time
- Better system control from one location with multiple process readouts available

In addition to controlling the extruder and gear pump, the microprocessor can collect real time data, analyze it, and print out meaningful statistical process control data.

36.3 Potential Gear Pump Problems

Potential problems that can lead to premature gear pump failure are

- Processing abrasive materials
- Losing molten polymer lubrication
- Damage caused by foreign material
- Overpressure
- Excessive heat

Abrasive wear is caused by small, hard particles or fillers in the melt stream. This wear is first evident as a dull appearance at the root between the gear teeth. The shaft may appear dull, as if it were sandblasted. As polymer circulates through the bearings, wear may become evident in the sleeve and bearings. If abrasive formulations are run on a pump, the pump manufacturer needs to know this information before purchase so appropriate alloys and hardness can be used to maximize wear resistance. Any metal particle contamination can create very rapid abrasive wear. Screen packs in front of a gear pump remove gross contamination that may create abrasive wear. Foreign contamination may be a problem when using recycle as part of the feedstock.

Pump wear due to poor lubrication can be quite rapid and severe in the bearings. Gear pumps have to be run with molten polymer present, as this lubricates the bearings. Starting up a new pump is done at low rpm until polymer flows freely out the pump exit. Once polymer is circulating through the pump, the speed can be increased. During shutdown, the line can be run empty, the pump shut down and disconnected from the extruder, and then the extruder can be purged as necessary to remove any residual polymer.

Damage from foreign material can be quite dramatic if it breaks a gear tooth. This is a problem in gear pumps used to feed liquids to an extruder. With proper screening, most contamination that may cause severe damage is trapped. Metal in the feedstock normally will not pass through the extruder to the gear pump. Damage occurs to the screw or barrel back near the feed throat. However, on large extruders with a portion of the feed stream being recycled, care needs to be exercised to prevent nuts, bolts, or other foreign metal objects from entering and passing through the extruder to the gear pump. A broken or damaged tooth is expensive to repair.

Excessive operating pressures or excessive temperatures relative to the design criteria can cause premature bearing failure.

36.4 Applications

Figure 36.11 compares pressure build-up in an extruder with and without a gear pump.[6] Being able to control the pressure with the gear pump provides better product dimensional control through more consistent pressure. While most extrusion processes can benefit by incorporating a gear pump, processes such as monofilament, sheet, film, wire and cable, coating, and fibers benefit significantly from gear pumps. One main gear pump advantage is output consistency and less product variation. This comes from reducing pressure variation. Figure 36.11 shows that gear pumps give tighter statistical process control limits.

The shear generated in the gear pump running at high speed with tight clearances raises the polymer melt temperature. Melt temperature increase in a gear pump depends on size and pump speed. The increase may be 20–40˚F (11–22˚C). The overall melt temperature increase with a gear pump is normally less than this because the lower extruder screw speed and head pressure shear heats the polymer less than if there was no gear pump. In closed-loop single screw extrusion, this can actually lead to a melt temperature reduction. However, in twin screw extrusion the gear pump is not normally connected to the extruder screw speed; and pressure flow is not a major component in the melt temperature determination as the screw, except for the last few flights, is not normally full of polymer (due to starve feeding). Gear pumps in twin screw extrusion are used principally to generate uniform and sufficient pressure to the die. Consequently, melt temperature increases attributed to the gear pump are in addition to the melt temperature generated in the extruder under normal operation.

A few case studies have shown the following advantages resulting from the incorporation of gear pumps into the line:

- Polystyrene sheet produced on a 4.5-inch extruder using a gear pump in-line with the extruder resulted in more consistent gauge, allowing the reduction in sheet thickness by 20% while increasing the

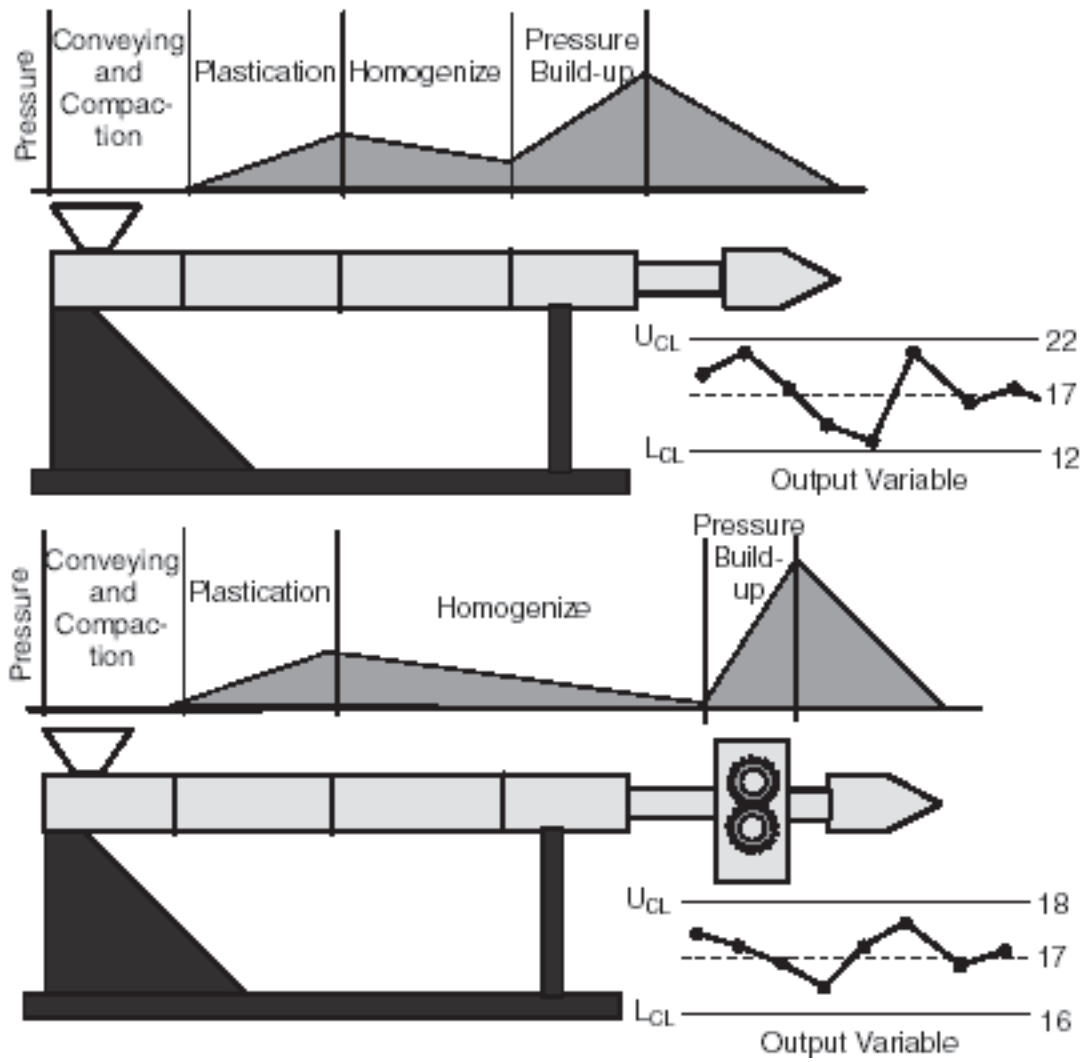

Figure 36.11. Pressure generation and control in a single screw extruder.

product per pound of resin by 15%. The energy consumption to produce the sheet was simultaneously reduced by 13%.

- PET cast film production on a 6-inch extruder reduced the head pressure from 1800 psi to 650 psi. This led to a 15% (3400–4000 lbs/hr) extruder throughput increase at the same extruder screw speed, with gauge variation decreasing 2–3% to ±0.5%.
- Installing a gear pump on a 2.5-inch wire coating application increased the throughput rate by 25% (150–200 lbs/hr) while reducing the scrap rate from 5% to 1%. This improvement is associated with eliminating the screw beat and surging associated with feeding.
- Adding a gear pump to a PP strap production line reduced the extruder head pressure by 1300 psi. This resulted in a 28% increase in throughput rate (250–350 lbs/hr). A secondary advantage was reduced sensitivity to viscosity variations and different viscosity materials being processed.
- Incorporating a gear pump into a PS foam sheet production line reduced the head pressure and the premature expansion of the pentane foaming agent while increasing the throughput rate 20% (800–1000 lbs/hr).

Other applications where gear pumps are particularly applicable are in monofilament production, where uniform polymer pressure and flow is required across the entire die face to ensure uniform flow rate per hole. Small variations in flow across the die or any type of surging generate filaments with unacceptable cross sectional uniformity. Similar to monofilament extrusion is fiber production, which requires uniform die flow for the same reasons.

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Review Questions

1. What are the advantages of using gear pumps in single and twin screw extrusion?
2. How is a gear pump installed in an extrusion system?
3. Explain the options for controlling a gear pump.
4. How does a gear pump work?
5. Why is polymer flow through a gear pump critical to its operation?
6. What are the different gears used and the advantage and disadvantage of each type?
7. What are some potential causes for gear pump failure?
8. Describe polymer flow through a gear pump.
9. What are the different types of gear pump heating? What determines the type of heating used with a particular gear pump?

37 Granulators

Almost all extrusion systems have granulators to recycle off-specification, start-up, or edge trim material. Sheet or film edge trim can be reground and fed directly back into the extruder hopper to improve material efficiency. Multiple-layer sheet is reground and fed back into the center or major layer. In pipe or tubing production, dimensionally off-spec product can be ground up and either recycled in current production at a low level or used in another product. Regrind is also sold on the open market for injection molding or other plastic processing as a lower-grade material at reduced price. A critical auxiliary system is an appropriately sized granulator or shredder to reduce the part to a chip or flake that will feed into an extruder, injection molding machine, or other thermoplastic processing equipment.

Large blobs or plops from extrudate running to the floor can be sawed or shredded into smaller chunks and granulated. All quality extrudate, whether blobs, edge trim, or off-spec parts, can be reduced to a chip or flake. If clean, this regrind can be used as part of a recycle stream or at 100% in a different part with less stringent end-use requirements.

Granulators come in many sizes and shapes designed to granulate specific parts or plastic chunks. The granulator may be integrated into the extrusion system, grinding edge trim for recycling to the extruder. Using an air conveying system, the edge trim is fed into a suction nozzle, where it is conveyed to a chipper or small granulator that cuts the trim into flake and conveys it to a hopper to be reintroduced into the extruder. Some plants use a separate process where scrap (start-up waste, off-spec parts, start-up parts) is taken and reduced in size to chip. Black concentrate can be added to colored and/or natural products and extruded into a black product.

Figure 37.1 shows granulator components. Plastic for recycle is added to the hopper. As the rotor rotates, the rotating knives and bed knives cut the plastic. Just below the rotor is a screen; when the plastic chips from the granulating operation are small enough, they pass through the screen and are collected in a collection bin below the screen. Screen size defines the maximum particle size. Flake, chips, or particles smaller than the screen holes pass through the screen and are collected in the bin below. The rotor is massive, with rotating blades that crush the material between the stationary blades. The rotor may have smaller blades that pass between stationary plates for cutting large chunks.

Granulator cleaning between different colors requires removing all fines in the chamber, around the rotor, in the collection bin, and the hopper, if keeping the colors separated is important. Fines adhere to the rotor, screens, hopper, and collection bin through static electricity. Therefore, product change should be considered when purchasing a system. A central granulator that may process all the product in the plant has different requirements than an individual unit near a packaging station that may process the same product all the time. Cross-contamination between products and/or colors is a potential problem. How the granulated product will be used dictates the system required. Light colors can be mixed and extruded into a darker color to minimize any color contamination. Similar end-use products produced with incompatible resin systems cannot be ground on the same granulator without cleaning the granulator between products. Re-extruding incompatible resin can lead to delamination and defective products.

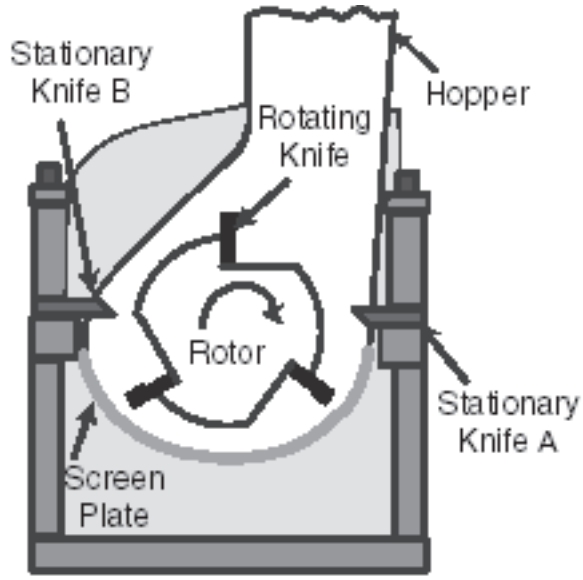

Figure 37.1. Granulator schematic.

Figure 37.2 is a commercial Conair granulator with the hopper tilted back to show the rotor, screen, and discharge bin. The hopper normally has a large opening with a heavy flexible cover to prevent chips and dust from exiting the feed hopper opening.

The rotor with the cutting knives is the only moving part in most granulators. There are many rotor designs available with different blades, depending on the granulator size. The simplest rotors have one-piece knives that extend the full length of the rotor. These blades are usually arranged helically so the entire knife does not pass the bed knife at the same time. This reduces the horsepower required. This rotor, called an open rotor, creates a scissor or slicing action as it passes the bed knife. Depending on the granulator size, two and three bolt-on knives are common on small granulators, and five to

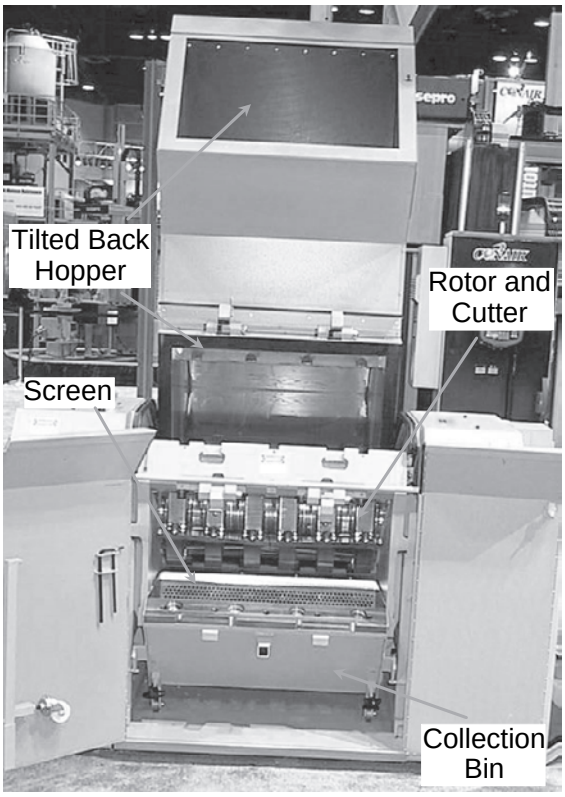

Figure 37.2. Conair granulator.

nine blades on large granulators. Figure 37.3 shows a single piece blade passing the bed knife. Another rotor design uses segmented sections to form a helical rotor. Each small segment can have either bolt-on knife tips or be machined directly into the segment. Figure 37.4 shows a typical rotor segment. Multiple segments are bolted together to make up the rotor. A third rotor design uses cassette type knives that bolt on the rotor. Individual cassettes are staggered, reducing the force and horsepower required during granulating. Figure 37.5 shows bolt-on knife, staggered cassette-type cutters on the rotor. Staggered rotors are available with either open blade rotors or cassette cutters.

Small granulators for granulating up to 350–400 pounds per hour may use any rotor and blade configuration. Higher throughputs require wider rotors and more horsepower. The number of rotating knives depends on the rotor design and throughput. Cassette knives have three blades on each cassette with two to six cassettes. Segmented rotors have three cutting edges on each segment with the number of segments dependent on grinder design. Helical one-piece cutters use between 5 and 12 segments, while the number of segments with bolt blades is between 5 and 8, as the segments are wider. Open rotors with four blades use either the conventional one-piece cutter style, shown in Fig. 37.3, or a staggered rotor requiring twice as many blades. All small granulators use two fixed knives (stationary knives in Fig. 37.1). As the granulator size and capacity increases, the rotor width increases, with more disks combined.

Larger and medium size granulators use open rotors with either a conventional or staggered configuration. The bed knives may increase to three. The designs are similar to those discussed above, with the rotors being wider with significantly higher horsepower. The blades on the rotor increase to five, seven, or nine. Large granulators are purchased for a centrally located granulating operation that can treat all the plant scrap from start-up

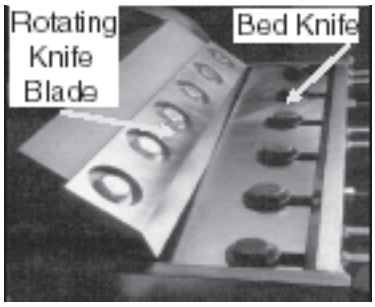

Figure 37.3. Helical single rotating blade on open rotor.

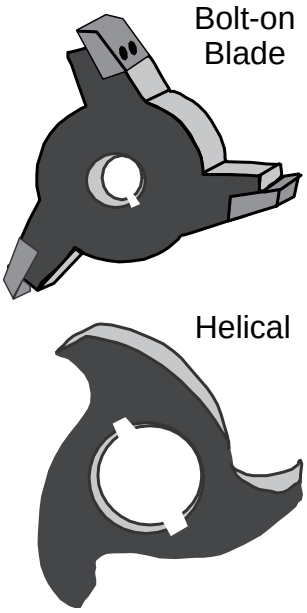

Figure 37.4. Segmented sections.

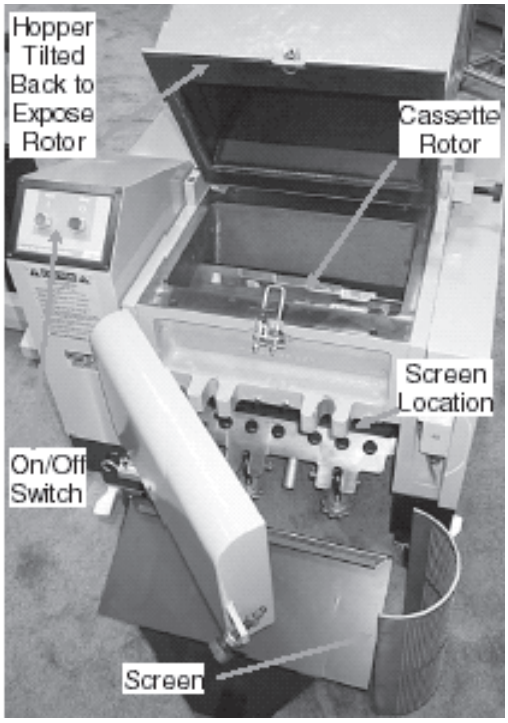

Figure 37.5. Conair granulator—cassette staggered rotor and screen.

and off-spec materials. Throughputs can range from 500 to 5000 pounds/hour. Central granulators are typically isolated in a different plant location because of—

- The noise, dust, and potential for contaminating first-grade products
- The area required to stage the materials to be ground
- The space needed for the granulated product

The knife design depends on the materials being granulated. PE and PP are soft and withstand high impact but can be easily cut. High-impact resins like ABS and PC do not fracture easily but can shatter when exposed to high force. These materials require greater blade toughness with a recommended knife angle of 50 degrees. Other resins such as phenolic, acrylic, and PS are friable, breaking apart easily, resulting in dust. High-shear knives are used with energy absorbing and friable materials. The blade is flat on the bottom with compound angles on the top. As shown in Fig. 37.3, plastics tend to cut with a scissor action.

Screens come in various sizes, depending on the particle size required in subsequent processing. Typical screen hole sizes are

- 1/8 inch (3 mm)
- 1/4 inch (6 mm)
- 5/16 inch (8 mm)
- 3/8 inch (10 mm)
- 7/16 inch (11 mm)
- 1/2 inch (12 mm)
- 5/8 inch (17 mm)
- 1 inch (25 mm)

The smaller the diameter selected, the more time and energy required to accomplish the desired particle size.

Granulated product collection depends on the quantity and how it is going to be used. Product collection can range from a bin under the granulator screen, to a Gaylord placed under the granulator, or, in large grinding operations, to pneumatic transfer to a storage bin where it is packaged and shipped to other plant locations or off-site. In setting up a grinding operation, attention to product disposal and handling after grinding is essential to an efficient, safe operation.

A film and low-gauge sheet granulator has a different configuration than the granulators discussed to this point. Two crimping rolls pull the edge trim to the cutting chamber and hold it back as the helical cutting knives cut the film against a bed knife. When the trim slices are chopped to the proper size, they pass through the screen into a collection chute, where a blower conveys the flake into a bin or back to a recycle feed hopper feeding the extruder. Hygroscopic materials can be collected, chopped into flake, and recycled back to the extruder feed hopper without the need for redrying.

Off-specification film rolls can be processed through the same system. A dancer or tensioning arm on the front of the machine controls the film tension being unwrapped from the roll. Gathering spools or a funnel narrows wide film prior to the crimping rolls. The driven crimping rolls complete the film folding begun in the gathering spools. Crimping roll tension compresses the folded film, making it easier to cut with the helical cutter against the bed knife. Flake is gathered as described previously in a surge hopper, bin, Gaylord, etc., for later use.

Granulators are rugged pieces of equipment designed to last a long time, provided they receive proper maintenance and care. Common granulator problems include the following:

- Knife wear is the most common and expensive maintenance problem.
- Knife damage can result from chipping caused by foreign material (principally metal) in the granulator feedstock.
- Improperly sharpened or balanced knives lead to premature machine wear and poor granulation efficiency and quality.
- Jammed rotors can lead to belt wear and, in the most severe case, burned-out motors.

A proper preventive or predictive maintenance schedule provides

- Knife quality checks for wear on the rotating blades and the bed knives
- Rotor bearings lubrication
- Motor lubrication
- Screen condition

One major problem in regrinding is foreign material getting into the product before granulation. When the material introduced to granulators is clean, routine wear is not severe. However, if the resin contains abrasive materials such as minerals or fiberglass, normal knife wear is accelerated. Dull knives need to be replaced routinely, as these are inexpensive and good knives reduce the overall wear and tear on the granulator. Likewise, if the product to be ground contains metal or other foreign objects, knives can become chipped and damaged. In a granulation operation, it is important to keep the raw material clean and the knives sharp.

In operation, failure to empty the hopper or properly convey the granulated product away can cause the regrind to build up in the granulator. Without proper evacuation, the regrind can fill the collection bin, preventing product from going through the screen. A build-up above the

screen causes the cutters to cut and recut the product many times, leading to excessive wear. In severe cases, the rotor can jam, causing the motor and/or the belts to burn out.

Proper granulator preventive maintenance to ensure long granulator life includes

- Grease the rotor bearings as recommended by the manufacturer.
- Check the belt tension regularly, retensioning the belts as required.
- Keep the regrind clean.
- Routinely check rotating blade and bed knife wear.
- Replace blades or bed knives on a routine schedule or as required, based on the materials being granulated.
- Sharpen knives to the proper angle.
- Check all safety switches and interlocks to verify that they are operating properly.
- Keep a maintenance record of all work and part replacement.

Review Questions

1. What are the different rotor designs used in granulators?
2. What determines the size of the granulated particles?
3. Where are granulators used in extrusion?
4. What are four maintenance issues associated with granulators?
5. What needs to be checked and serviced in a good preventive maintenance program for granulators?
6. What is the difference between a granulator used to regrind film compared to a granulator designed to regrind profiles, pipe, and heavy stock?
7. What are the differences in knives required for different resins?

Chillers can be used to control the temperature of water-cooled extruders, the extruder feed throat temperature, and the cooling medium temperature for profile, pipe, underwater pelletizing, and tube extrusion. Cooling tower water in a closed-loop system is used to perform the same functions as a chiller. Heaters can be used to control chill roll temperatures for sheet and film, water temperatures in monofilament, profile, pipe, and tubing.

This chapter provides information on portable chillers and temperature control units. Cooling towers are outside the scope of this book. Actual use of chillers or temperature control units with specific extrusion processes is covered in Part 7.

38.1 Chillers

Chillers are used to control the feed throat temperature and barrel temperature for water-cooled extruders. Downstream cooling associated with vacuum sizing tanks for profile, pipe, and tubing, and other cooling processes can use chillers. Chillers provide a closed-loop water system to cool either the equipment or product. Advantages versus once-through water are

- Constant water temperature
- Reduced water cost
- Reduced water waste
- Lower water temperature control point possible

In a large extrusion facility with many extruders and water cooling systems, it is common to have a cooling tower. If no cooling tower is available, a chiller provides an efficient way to recirculate treated water through the feed throat and extruder barrels if water cooling is being used. One chiller can cool a number of feed throats if that is all the cooling required. The chiller has to be properly sized to match the heat load.

The most important factor to consider when selecting a chiller is to choose the proper size. The extruder manufacturer can supply the flow rate through the feed throat and the approximate flow rate to control the barrel in a water-cooled extruder. Using Eq. (38.1), the heat load can be calculated and the chiller properly sized.

$$Q = V \times \rho \times C_p \times \Delta T \tag{38.1}$$

where

Q (heat flow) = BTU/hr (Kcal/hr)
V (volumetric flow rate) = gallons/hr (liters/hr)
ρ (density of cooling fluid) = pounds/gallon (kg/liter)
C_p (specific heat) = BTU/pound˚F (cal/gm˚C)
= 1 for water
ΔT (temperature change) = ˚F (˚C).

Recognizing that the cooling fluid density changes with temperature, assume 60˚F (15.6˚C) where water density is 8.33 pounds/gallon (1.0 kg/liter). Assume the feed throat flow rate is one gallon/minute at 100˚F. (Without a measurement of the water temperature entering and exiting the feed throat, ΔT is not known; therefore, assume the water temperature increased 10˚F to maintain the feed throat at 100˚F.) The heat load (Q) calculation is given by Eq. (38.2):

$$Q = 60\frac{\text{gal}}{\text{hr}} \times 8.33\frac{\text{lb}}{\text{gal}} \times 1\frac{\text{Btu}}{\text{lb} \times {}^\circ\text{F}} \times 10^\circ\text{F} = 4998\frac{Btu}{hr} \times \frac{1\text{watt}}{3.41\frac{\text{Btu}}{\text{hr}}} = 1.47\text{kW} \tag{38.2}$$

Therefore a chiller capable of removing 1.47 kW or 4998 BTU/hr is needed to keep the cooling throat at the proper temperature. Cooling is given in BTU/hr (kcal/hr) or tons. To convert from BTU/hr to tons, use the conversion 1 ton of refrigeration = minute/200 BTU. Equation (38.3) converts the results from Eq. (38.2) into tons:

$$Q = 4998\frac{\text{Btu}}{\text{hr}} \times \frac{1\text{hr}}{60\text{ min}} \times \frac{1\text{ min}}{200\text{ Btu}} = 0.41\ \text{Tons} \tag{38.3}$$

In chillers the heat load is described in BTU/hr, kW, or tons capacity. Table 38.1 gives common guidelines for basic chiller tonnages.[1]

Table 38.1. Typical Heat Loads in Extrusion

Load	Chiller Tonnage
Fee Throat	0.5
Gear Box	1
Screw and Barrel	1 ton/inch screw diameter
HDPE, LDPE, PP	1 ton for each 50 lb/hr
PS	1 ton for each 75 lb/hr
PVC	1 ton for each 80 lb/hr

Obviously the calculated load depends on whether the chiller is required to cool just the feed throat or the feed throat plus barrel zones. If the barrel zones are water cooled, factors affecting barrel temperatures are

- How much heat supplied by viscous shear heating has to be removed
- Extruder size
- Processing temperature window size

The more cooling required, the greater the chiller cooling capacity and the larger the chiller must be. If the extruder manufacturer can supply an appropriate heat load for different resin systems, this data can be used to properly size the chiller.

In sizing chillers for extrusion, the following information is required:

- Extruder diameter and length
- Number of zones
- Extruder horsepower
- Connection sizes
- Throughput
- Materials being processed
- Melt temperature
- Cooling required if it is known in either BTU/hr or tons

The second issue in selecting a chiller is the choice between an air-cooled or water-cooled unit. Cooling is required to dissipate the heat removed from the process. Water-cooled units are generally used in applications where tower water or some other inexpensive water source is available to take away the heat. Air-cooled units are used where water is contaminated, an inexpensive water source is not available, or the heat being dissipated is part of a heat recovery system for heating the plant. Water-cooled models use condensers, while air-cooled models have fins with an air-circulating blower to remove the heat. Water-cooled units are 10 to 15% more efficient than air-cooled units. The advantage of the air-cooled model is it becomes more portable because a water source and drain or return are not required at each location. Portable chillers in extrusion operations are generally used when cooling tower water is not available. If cooling tower water is available to remove the heat from a chiller, it is also available to hook directly to the extruder. In this scenario, a portable chiller may still be required for cooling water in a vacuum tank or other downstream equipment to a temperature below cooling tower water.

The third selection criterion is the control system, with a choice between a digital temperature controller and a fully programmable, self-diagnostic microprocessor controller. A digital controller is inexpensive and provides continuous operating temperature readings, with a single button for starting and stopping, easy entry of the process temperature setpoint, low-liquid mode signal, and status indicators for on/off, compressor, alarm, and low flow. The microprocessor controller supplies some of the same information available with the digital controller plus additional information and more detail. Microprocessor controls can provide

- Setpoint and actual return temperatures
- Percent load or tons
- Process flow rate in GPM or LPM
- Refrigerant pressures
- Run status
- Diagnostic information
- Adjustable alarm limits to display low level, high temperature, and low temperature
- Password entry to prevent accidental or unauthorized changes in settings
- Automatic fine tuning of proportional and integral and derivative (PID) temperature control parameters to provide uniform temperature
- Setpoints in degrees F or C

Options available include communications ports for RS232, RS485, or RS422 with remote control switches and sensors.

A schematic of the operation sequence in a chiller is shown in Fig. 38.1. Hot water in the hold tank from the process is pumped through the refrigerant system to remove the heat and properly cool the water going to the process. If the water in the hold tank becomes too hot, cool water from the refrigerant system can be pumped through the process bypass valve, sending cool water directly to the holding tank to cool the water in the holding tank. Make-up water is brought into the system through the refill valve at the top of the holding tank.

The pump size is critical for good heat transfer characteristics. Turbulent flow through the process system is required for good heat transfer. The pump size and flow rate are just as critical as the circulating water temperature. Higher flow rates result in higher heat transfer rates.

Some options available on chillers include

- Nonferrous materials in contact with the water to prevent rust
- Pump motor overload protection
- Stainless steel tanks
- Brass shut-off valves
- Brass pump
- Audible and visual alarms
- Water filtration

One advantage of using a chiller versus once-through water is the water savings and the cost associated with the water. Using 1 gallon of water per minute for 24 hours a day, 5 days per week, for 50 weeks consumes 360,000 gallons per year. With many water-cooled extruders, the consumption significantly exceeds 1 gallon per minute.

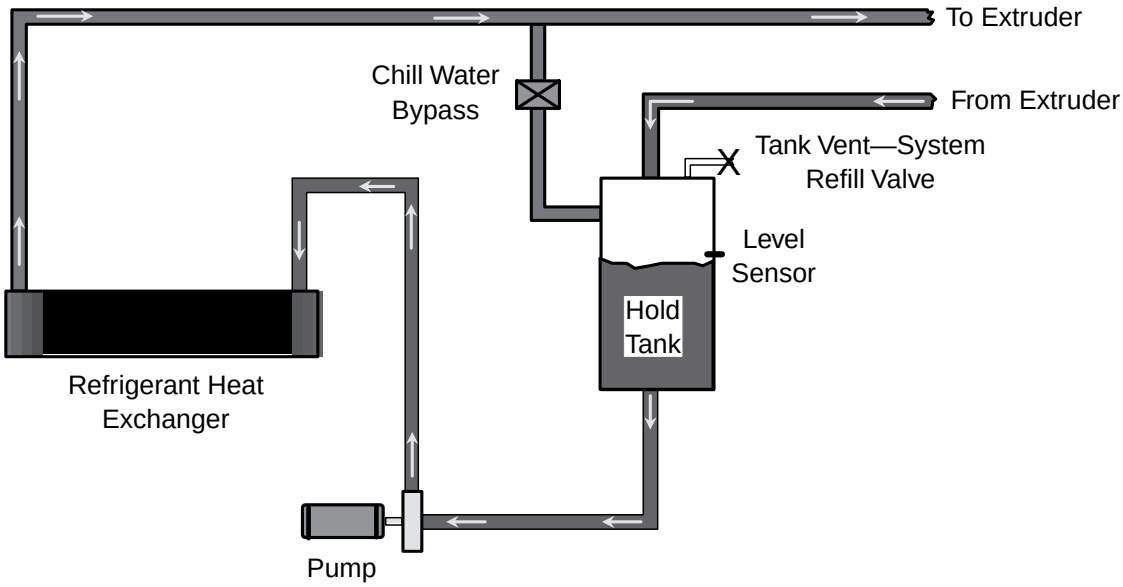

Figure 38.1. Schematic of fluid flow in a chiller.

Payback for chillers based on water savings alone is relatively short.

Chillers are also used to control the water bath temperature in profile and vacuum sizing tanks and to control the cooling air temperature in blown film operations.

38.2 Heaters

Heaters are more properly called temperature control units that provide higher temperatures than chillers. A temperature control unit in most situations heats equipment up to processing temperature prior to start-up and maintains temperature during operation and work stoppage. During actual operation, the main function may be to control at a high temperature where the hot plastic will process efficiently. Temperature control units are similar to portable chillers, except they circulate heat transfer fluid at higher temperatures. Some extrusion applications requiring high temperature are heated roll stacks in sheet and film extrusion; cast film rolls; heated water baths to control the rate of extrudate cooling in profile, pipe, and monofilament; heating molds in large-part blow molding; and fluid-heated dies, gear pumps, and screen changers. A heater works on the same principles as a chiller; however, instead of having a cooling unit, a heater is present. The heat transfer fluid is normally water or oil, depending on the temperature requirements. Controls are similar to those used in chillers.

Figure 38.2 shows a portable water or oil heating system. It does not contain all the relief valves and solenoid valves and shows only the heat transfer fluid flow. The cooling heat exchanger cools the heat transfer fluid if it comes back from the process at a higher temperature than it entered the process. This can occur with the heated rolls in a three-roll sheet stack. Heat is supplied to the rolls to bring them up to temperature prior to the run or to maintain the roll temperature during work stoppage in the middle of a run. During the run, heat from the hot polymer is transferred to the roll, requiring the temperature control unit to remove heat from the process to maintain a constant temperature. The cooling water entering the heat exchanger circulates through a coil inside the heat exchange unit, cooling before exiting the process as cooling water out. The heat transfer fluid is cooled below the temperature required by the process before leaving the heat exchanger and passes through the pump to the heater. In the heater, the heat transfer fluid is reheated to the desired process temperature.

As with chillers, there are different control packages with different levels of sophistication. The simplest controls have password protection, PID temperature control, and digital display of actual and setpoint temperatures. As controls become more sophisticated, added options include communication ports (RS485), auto-start capability, remote mount controls, phase detection to ensure the pump is rotating in the correct direction, purge valve control, choice of GPM or LPM, choice of degrees F or C, status of different control functions, and alarms.

Three ways to control water temperature are

- Direct injection of water into the cooling system
- Closed-loop cooling
- Isolated cooling

With direct injection, water is directly added to the processing water as required. This system is typically used only when treated water or cooling tower water can be added to the system. The second system is a closed-looped process, where water is added initially to the recommended level. Subsequent addition of water occurs when the process water is low or a low-level alarm sounds. The third system is an isolated system, where the cooling water is held in a reservoir separate from the process fluid (system shown in Fig. 38.2).

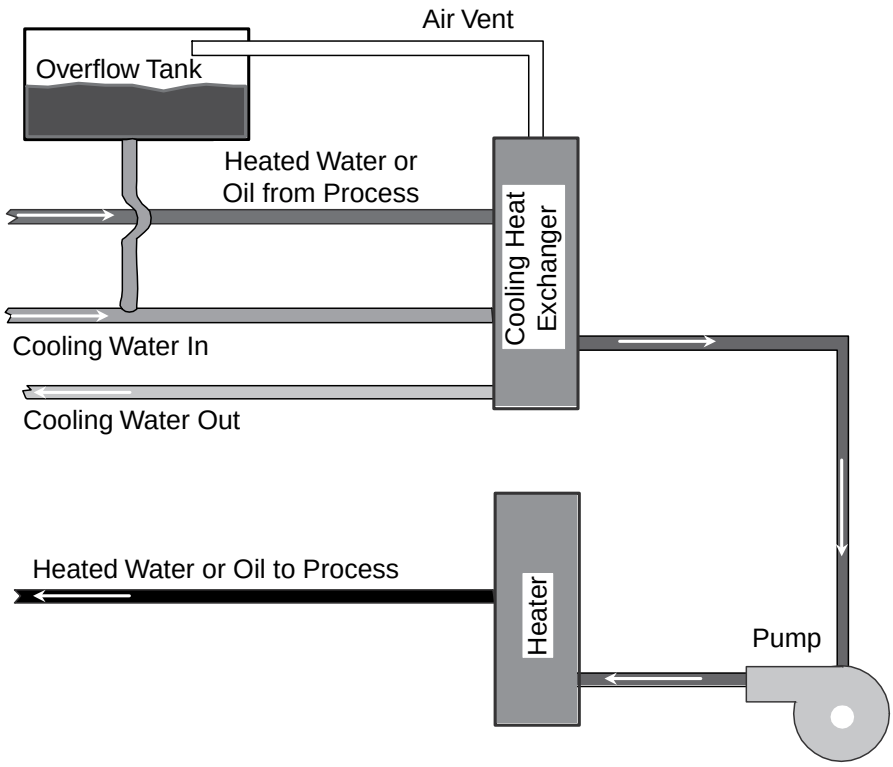

Figure 38.2. Schematic of portable heating system.

REFERENCE

1. Fosco, Al, "Chillers and Mold Temperature Controllers: A Guide to Usage," *Plastics Auxileries*, March 1999, p. 24.

Review Questions

1. What is the purpose of chillers in extrusion processing?
2. Calculate the heat load and specify the tonnage cooling capacity needed for cooling the feed throat on a twin screw extruder using water at a rate of 1.5 gallons/minute and a temperature differential of 12˚F between the entrance and the exit of the feed throat.
3. What is the purpose of the heat exchange unit in a temperature control unit used to heat a three-roll sheet stack?
4. What are two different types of controllers used with chillers and how do they differ?
5. What are three questions that have to be answered when selecting a chiller?
6. Why is the pump size critical in chillers and temperature control units?

39 Screw Cleaning and Purge Compounds

Many options are available for cleaning the extruder screw, barrel, breaker plate, and die between production runs or during routine maintenance. The first step in cleaning the extruder is to empty the hopper, screw, and barrel. Follow this with an appropriate purge material to remove any residual material from the extruder. At this stage, another production run can be started or the screw pulled to change the screw in single screw extrusion or modify the existing screw configuration in twin screw extrusion. After removing the die and any adapters, transition pipes, gear pump, and automatic screen changer, the screw is either pushed or pulled out of the extruder barrel. The barrel, feed throat, and die are cleaned and a new or modified screw installed in the extruder. Finally the screw that was removed has to be cleaned of any polymer or polymer residue left on the screw in preparation for its next use. While the screw is out of the extruder, it should be measured for wear with a micrometer.

This chapter covers purge materials and methods of cleaning the barrel, die, breaker plates, and any transition pipes or adapters.

39.1 Purge

Many purge materials are commercially available. It is the processor's job to identify the appropriate purge to use with the resin being processed. Purge materials work by one of four mechanisms:

- Mechanical purge
- Abrasive mineral filled material
- Chemical purge to break down resins and contaminants
- Hard resin filled with surfactants

Mechanical purges are stiff materials, normally polyethylene (PE) based, containing cleaning and release agents. Fractional melt flow high density polyethylene (HDPE) functions very well as a purge material over a wide temperature range. Low viscosity abrasive mineral or glass filled materials will force most resins out of an extruder while scouring the screw, barrel, and die. Cast acrylic is an abrasive purge, as it does not completely melt in the extruder. Due to the high melt viscosity of cast acrylic, it is better to remove the die prior to purging the screw. Typical amounts required are 5 to 10 pounds/inch of screw diameter. Once cast acrylic is in the barrel, it has to be purged out or the screw pulled and mechanically cleaned. The third purging material uses a surfactant that penetrates and loosens residue on the screw, barrel, and die, dispersing it in the melt. The surfactant is mixed with a melt flow resin that is 0.1 to 0.3 times that of the original melt to provide maximum purging effectiveness. The fourth material is a chemical purge that attacks the material left in the barrel. This may be through plastication of the resin in the barrel, lowering the viscosity and making it easier to force out of the barrel, or the additive may actually cause polymer depolymerization into lower-molecular-weight components. Table 39.1 contains a list of polymers and recommended purging agents.[1][2] The majority of purge materials listed in Table 39.1 all work by mechanism 1 or 2, where the

Table 39.1. Recommended Purge Materials for Various Resin Systems

Material to Be Purged	Potential Purge
ABS	Cast Acrylic, Polystyrene, Natural ABS, SAN
ABS/PVC	General Purpose ABS (do not use FR ABS)
Acetal	Polystyrene, HDPE (do not contact with PVC)
Acrylic	Clean Acrylic regrind
ASA Blends	Cast Acrylic, ASA, HDPE
Filled/Reinforced Resins	Cast Acrylic
Fluoropolymer	Cast Acrylic followed by HDPE
LCP	PP
Nylon	Polystyrene, Low MFI HDPE, Cast Acrylic, PP
PBT	Polystyrene, HDPE, Next material to be run
Polycarbonate	Cast Acrylic, Polycarbonate regrind, Polystyrene
PC/ABS	Cast Acrylic, Polystyrene, HDPE
PC/PBT	HDPE, Polystyrene
PEI	HDPE, Glass Reinforced PC
PET	HDPE, Cast Acrylic, Low MFI HDPE
Polyester Alloys	HDPE
Polyolefins	HDPE
PPO/PS	Cast Acrylic, Polystyrene
PPS	HDPE
Polysulfone	Polycarbonate, Low MFI PP
Polysulfone/ABS	Polycarbonate, Low MFI PP
PVC Flexible	HDPE
PVC Rigid	ABS, Acrylic, Polystyrene
TPE	HDPE, PP

resin is cleared out of the system by a higher-viscosity resin. In many cases these are general purpose resins.

Mechanical purging commonly uses a fractional melt flow HDPE. The stiff material pushes the resin being purged out in front of the HDPE. Fractional melt flow HDPE is cheap, works well, and has a wide processing temperature range. After the resin being purged is out of the extruder, the barrel temperatures can be lowered and more HDPE added to remove the earlier purge material. As the screw is being pulled from the extruder, any residual HDPE can be brushed off the screw with a brass wire brush. Alternatively an electric or air-powered brass wire brush can be used in cleaning. Several commercially available materials are produced as purge compounds based on mechanical purging.

Dyna-Purge® is produced by Shuman Plastics Inc.[3] in nine different commercial grades. Dyna-Purge® is designed to soften but not melt in the extruder while scrubbing the screw and barrel walls. It is easy to use, requiring no mixing with other resins or ingredients. Its nonabrasive, nonmelting character loosens carbon and resin residue, which is flushed out of the extruder. It is appropriate to use in product or color changes and also before manually cleaning the extruder. It contains no chemicals or hazardous materials. The different Dyna-Purge® grades for extrusion and their temperature ranges are shown in Table 39.2. There is also a high-temperature grade for engineering resins that are not shown in the table.

Pax Purge™ from Paxon Polymer Company and UniPurge™ from Union Carbide are mechanical purges based on polyolefin resin. Pax Purge™ is based on high-molecular-weight HDPE polymer with polyisobutylene and no abrasive or chemical additives, relying on its scrubbing action to remove foreign materials from the extruder. UniPurge™ works on the same principle of mechanically scrubbing the extruder. It does contain some specially formulated cleaning agents to scavenge and clean degraded and/or oxidized polymer out of the extruder. Ultrahigh-molecular-weight PE can also be used as a scouring, mechanical-type purge compound.

Sun Plastech Inc. produces a product called ASACLEAN® that forces polymer residues out of the barrel ahead of it.[4] Using the appropriate temperature grade of ASACLEAN® for the resin being processed, remove all resin from the extruder and clean the feed throat and hopper of all resin and fines. With unfilled resin grades, the die, breaker plate, and screens can be left in place. Add the equivalent of two barrel volumes of purge to the hopper and run it through the extruder at low screw speed until it exits the die, then increase the screw to the maximum safe speed. Stop the screw for 5 to 10 seconds and then increase the screw speed to maximum rate again. Continue the stopping and starting action until all the ASACLEAN® is removed from the barrel. In the event filled materials are being purged from the extruder, filled grades of ASACLEAN® are required. The die temperature needs to be raised 30–50˚F (15–30˚C), while remaining in the temperature range for the purge grade being used. Table 39.3 shows the various grades of ASACLEAN® and their temperature ranges. Grades EG and EX are filled purge materials.

HiTech Polymers Inc. makes a range of PolyPurge™ thermoplastic polymer purging compounds that scour and push contaminants out of the barrel.[5] There are several commercially available grades (see Table 39.4) that can be used with different thermoplastic resin systems. All grades are easy to handle, nontoxic, and do not give off any noxious odors during the purging operation.

Abrasive-type purges are glass filled or mineral filled materials that scour and scrub the screw and barrel as they pass through the extruder. Based on the abrasive additive particle size, screen packs may have to be removed to prevent them from clogging. Depending on the resin system and the filler, abrasive systems can be costly, such as glass filled polycarbonate (PC), which is an excellent purging compound to clean PC or polyetherimide out of an extruder. Unfortunately glass filled PC is an expensive material to use for a purge resin. Cost-effective grades of filled ASACLEAN® can be used as purge with abrasive action. Polychem USA makes a multipurge abrasive compound that scours the screw and barrel.[6]

The third class of purge materials is chemical cleaning agents. There are materials that react with the chemical residue in the extruder, either depolymerizing it or penetrating the residue to lower the viscosity so it can be purged out of the barrel. Whereas the mechanical and abrasive purges force material out with a scrubbing action, using chemical purges, the feed hopper and throat must first be

Table 39.2. Dyna-Purge® Temperature Ranges and Applications

Grade	Temperature Range, F (C)	Application
F	320–420 (160–215)	Cleaning single film and sheet extruder and die; will pass through barrel, screen pack, dies without requiring equipment dismantlement
SF	320–575 (160–301)	For multilayer extrusion; similar to F with wider processing window
X	320–525 (160–274)	Use for sheet, profile, pipe, tubing, and blow molding
V	620–400 (160–204)	Designed for use with PVC

Table 39.3. Grades of ASACLEAN®

Grade	Temperature Range, F (C)
SA	350–680 (180–360)
SPL	340–570 (170–300)
SV	320–450 (160–230)
SX	570–735 (300–390)
E	350–530 (180–270)
EG*	350–630 (180–330)
EX*	390–680 (200–360)

* Filled grades

Table 39.4. PolyPurge™ Grades

Grade	Temperature, F (C)	Materials to Purge
PolyPurge C	350–550 (177–305)	Colored Resins
PolyPurge H-10	425–600 (218–316)	Nylon, PC, polyester, polyphenyleneoxide, other engineering resins
PolyPurge L-20	350–450 (177–232)	PS, polyurethane, acrylic, polyolefins, ABS
PolyPurge V-10	<400 (<204)	PVC, TPE, polyurethane

cleaned of all resin to be purged and the barrel run empty. The chemical purge is added to the barrel and run at low speed until it just starts to exit the extruder. The screw is stopped as the purge sits in the barrel, reacting with the polymer residue. After approximately five minutes the screw is restarted and then stopped a couple of minutes later. This cycling of the screw continues until the purge has removed all the previous resin. Some of the chemical purges that are commercially available are Rapid Purge™ from Rapid Purge Corporation, Novapurge and SuperNova™ from Novachem, Z Clean from LarSan Chemical Company, XPD8™ from RBM Polymers Inc., and Purge USA from Purge USA, a division of PTY Enterprises Inc.

Rapid Purge Corporation[7] makes several chemical cleaning grades to purge a wide variety of thermoplastic materials. Very small quantities of Rapid Purge™ react with the polymer to be cleaned, causing resin depolymerization to a lower-molecular-weight, low-viscosity species that can easily be forced out of the extruder, die, and transfer pipes. Rapid Purge™ is mixed with the next resin to be processed through the extruder in the percentage given in Table 39.5. The mixture volume required for complete purge is approximately three to four barrel volumes. As the material starts to foam out of the die, shut the extruder off for approximately 10 minutes; longer soak time will continue to remove more black specks and degraded resin. Slowly purge the Rapid Purge™ mixture out of the barrel and then stop the screw for another five minutes to allow the barrel to soak. Continue this process until the purge material has been completely consumed. Add the new resin to the barrel and run until all Rapid Purge™ is removed from the barrel. In severe contamination situations, it may still be necessary to pull the screw. Running purge through the extruder first makes the screw pulling process easier.

Novachem produces chemical purging compounds that are blended with carrier resins to form concentrates or used directly as purchased. Novapurge, based on ammonia cleaning, comes in three grades to use directly in purging different resins systems, as a concentrate in powder (NP-500), or in pelletized form (NP-600). The correct

Table 39.5. Rapid Purge™ Percentage

Material to Be Purged	% Rapid Purge™ by Weight
ABS	20–25
Acetal	15
Acrylic	15
Nylon	15
Polyester	15
PC	15–30
PE	15
PP	15
PPO	15
PPS	30
PS	15
SAN	15

Table 39.6. Novapurge and SuperNova™ Purge Characteristics

Purge	Temperature, F (C)	Resin Systems
Novapurge NP-1500	< 425 (<218)	EVA, flexible PVC, PS, LDPE
Novapurge NP-2500	<550 (<288)	ABS, acetate, acrylic, acetal, EVOH, Kraton, PPO, PA, PC, polyester, PE, PP, PS, PU, PVC, SAN TPE
Novapurge NP-3000	<800 (<427)	Fluoropolymers, fractional melt, HDPE, PPO, PEEK, PPS, PEI, rigid PVC, LCP
SuperNova™ LT		EVA, flexible PVC, PS, LDPE
SuperNova™	<550 (<288)	ABS, acetate, acrylic, acetal, EVOH, Kraton, PPO, PA, PC, polyester, PE, PP, PS, PU, PVC, SAN TPE
SuperNova™ LT	<800 (<427)	Fluoropolymers, fractional melt, HDPE, PPO, PEEK, PPS, PEI, rigid PVC, LCP

grade for each application is displayed in Table 39.6.[8] SuperNova™ grades of chemical purge concentrate are ammonia-free and safe for incidental food contact applications, as defined by the Food and Drug Administration. Three grades are available for purging low-, normal-, and high-viscosity resins as either concentrates to be mixed with resins or in a ready-to-use form. In addition, SuperNova™ CX is used for purging transparent resins such as PC and acrylic, while SuperNova™ XB is used to purge extrusion blow molding equipment; it is sold only in a ready-to-use form. Processing temperature ranges for the different grades are shown in Table 39.6. Purge quantities required are approximately two extruder barrel volumes. The ready-to-use purge is mixed at approximately 45% by weight with PE. Resins processed at low temperature (EVA, flexible PVC, PS, and LDPE) use either LDPE or LLDPE as the carrier resin. Resins processed at temperatures up to 550˚F (288˚C) use HDPE as the carrier, while higher-temperature processes use high-molecular-weight HDPE (fractional melt flow PE).

Several grades of Z-Clean® chemical purging material, sold by LarSan Chemical Company,[9] are available to purge all types of thermoplastic resins. Table 39.7 shows the grade, base resin mixed with the purge, temperature range, and materials that can be successfully removed from the extruder with each grade. Two to three barrel volumes of purge are recommended. The purge procedure is—

- Clean the feed hopper of all dust and particles.
- Run the barrel empty.
- Add the appropriate purge volume to the feed hopper.
- Turn the screw on at 20–25 rpm.
- Run the extruder until purge is exiting the die.
- Run approximately five minutes.
- Turn the extruder off and allow the material to sit in the barrel for five minutes.
- Start the extruder and run for two to three minutes.
- Stop for five minutes.

The last two steps are repeated until the barrel is clean. New resin is added to the hopper and new production started.

RBM Polymers Inc. supplies odor-free pelletized purging compounds called XPD8™.[10] These purge compounds use heat-activated chemical reactions to lower the contaminants' molecular weights. In zones where there is potential contamination (dies, sharp angles), temperatures are increased to activate the purge. Other areas are left at their normal operating temperatures. At the end of the purge cycle, XPD8™ is easily removed from the barrel and die. Table 39.8 shows available grades and their temperature ranges with a description of the purge material.

The fourth purge mechanism is a thermally stable surfactant-type additive that attaches to any residual polymer or polymer build-up to loosen it from the barrel, screw, and die, dispersing it into the polymer melt. The surfactant function is to bond the anionic or nonionic surface-active agents to the degraded polymer or gels, softening and loosening the particle from the metal, allowing it to be carried out with the viscous polymer melt. The purge is sold as an additive or concentrate to be added to existing resin or to be mixed with PE, PS, or cast acrylic as a carrier. The purge formulation is added to the extruder until it starts to exit the die; the screw is turned off and then, after 10–15 minutes, the next production is started. If it is feasible, the barrel temperatures are lowered 50–75˚F (28–42˚C) prior to starting the purge procedure. Purge passes through screen packs, so it is not necessary to remove them from the extruder. Purge-USA, supplied by PTY Enterprises Inc.,[11] is a liquid that is sprayed on to a carrier resin at approximately 3–5% by weight to coat the carrier resin. The liquid spray is water based with anionic and nonionic surface-active agents, defoamers, and cleaning additives.

TexCarPurge™ is a surfactant purge compound supplied by Texas-Carolina Chemicals Corporation.[12] A pellet concentrate is added to a carrier resin at approximately 15%. Two different purging procedures are used, based on the cleaning required. One procedure is to fill the barrel, transfer pipes, and die with the purging compound (concentrate plus carrier resin), stop the extruder, and allow the material to sit in the barrel for 15 minutes. This method is used when the extruder requires more intense cleaning due to severe product contamination. The second procedure is used in continuous operations when color changes or other product changes are required and the barrel, screw, or die

Table 39.7. Grades of Z-Clean® with Base Resin and Temperature Range

Material	Base Resin	Temperature Range, F (C)	Resin to Be Purged
A1	PP	320 - 572 (160 - 300)	PP, PE, ABS, TPE
S-28	PE	350 - 570 (177 - 299)	PC, PMMA, PS, SAN, rigid PVC, ABS, PP, PE
S-29	PE	482 - 752 (250 - 400)	PPO, PPS, PA, PEI, LCP, PEEK,
E3001	LLDPE	350 - 570 (177 - 299)	PE, PP, PA, ABS, PPO, PVC

Table 39.8. Operating Temperatures for Some XPD8™ Grades

XPD8™ Grade	Operating Temperature	Description
"LQ"	240–650˚F	Easy use, no mix or mess liquid
"PG"	350–700˚F	Superconcentrate in pellet form
"SP"	380–650˚F	Redi Mix in pellet form
"D8"	400–700˚F	Salt and Pepper Redi Mix in pellet form
"SA-400"	440–800˚F	Styrenic Redi Mix in pellet form

does not have black speck accumulation. In a continuous operation, when the old formulation is complete, TexCarPurge™ is added to the extruder and run for 5–10 minutes before starting the new product; the extruder is never shut down. Table 39.9 shows the carrier resin and blend ratios to use with the different purging procedures.

As with all commercially available materials, the product mix and performance characteristics change with time. When looking for new materials, it is better to compare available products and test them to confirm that they will do the job you desire.

39.2 Mechanical Cleaning

Purging procedures are used when changing from one color or product to another and/or prior to shutting down to mechanically pull the screw and thoroughly clean the extruder. Proper purging prior to pulling the screw or disassembling and cleaning the die can make the cleaning job easier. Running purge through the extruder and die (assuming the purge material is the type that will pass through the breaker plate and any small openings in the die) removes processed polymer, replacing it with a potentially higher-viscosity resin that has less adhesion to the barrel, screw, and die. Purge the extruder, transition pipes, adapters, and die. Remove the die and clean out any purge and degraded polymer adhering to the flow surfaces. Breaker plates are removed while hot and cleaned with brass scrapers and brass wire brushes. Any screens are removed from the breaker plates and discarded. Prior to removing the screw, start the extruder motor with the die and breaker plates disconnected to remove any material still in the barrel. The screw or screws are ready to be pushed or pulled out of the extruder. Single screw extruders have a bolt on a threaded adapter that attaches to the back of the extruder. A threaded rod is run through the adapter by turning the extruder on at low rpm. A nut or other type of flange bolted to the opposite end of the thread rod prevents the rod from turning. As the extruder rotates, the threaded rod is slowly turned into the extruder, pushing the screw out. In twin screw extruders, the screw tips are removed, a pull handle is attached to the end of the screws, and the screws are mechanically pulled out. As the screws are being removed, use a brass brush to remove any excess polymer still on the screws. While the barrel is still hot, an appropriate size brush wrapped with copper or wire gauze attached to a long rod is used to clean the barrel. Attach the rod to a suitable size drill and rotate the brush in and out of the barrel several times; the copper or brass gauze removes any polymer residue. As the wire gauze is consumed in the cleaning operation, it is replaced by wrapping new gauze around the brush. Once the barrel is clean, blow any wire gauze or other debris in the barrel out the barrel's open end. Verify that the keyway in single screw extruders is clean. In conical twin screw extruders, the screws have to be pulled out of the feed throat end of the machine.

Some methods to clean the die, screw, and breaker plate after removal from the extruder include

- Mechanical cleaning
- Heat to burn off any residue or carbonized material
- Dry ice blast
- Solvent cleaning

The advantages and disadvantages of these methods are discussed below.

Table 39.9. TexCarPurge™ Applications and Procedures

Process	Material Being Purged	Resin to Blend with Purge	Concentration of Purge
Pipe, Tubing	HDPE, flexible PVC	HDPE	15%
Blown Film	LDPE, LLDPE, PA	LDPE or LLDPE	15%
Blow Molding	HDPE, PVC, PC	LDPE or LLDPE	15%
Sheet	PC, PMMA, PS	LDPE or LLDPE	15%
General	PVC, PP, PC, PS	LDPE or LLDPE	15%

Mechanical cleaning is done using brute force, with the aid of wire brushes attached to either air or electrical drills, and torches to spot heat areas. Proper screw and barrel purging prior to screw removal can minimize the problems with this procedure. While the screw is hot and as it is being removed from the extruder, brush any residue off the screw. Pull the screw slowly and only clean small sections at a time; otherwise, the screw will cool and any molten polymer sticking to the screw will be difficult to remove. Cleaning dies and breaker plates has to be done when they are hot or it is extremely difficult to remove the polymer. The obvious advantage of using this method is no additional equipment is required. However, it can be very labor intensive and require more time to clean everything properly. If the parts cool below the crystalline melting point (T_m) for a crystalline polymer or below the glass transition temperature (T_g) for an amorphous polymer, it will be difficult to remove. It can be extremely tedious and difficult work, depending on the amount of polymer to be removed. Damage to extruder parts is common because of the handling and cleaning methods. Some methods lead to shorter life and earlier part replacement. From a safety perspective, mechanical cleaning creates the greatest hazard potential from hot plastic, fumes, torches, and mechanical equipment such as hand-held polishing wheels.

The most common method of cleaning dies, breaker plates, and screws is to burn contaminates and polymer residue off with high temperature. Advantages of this method are

- Minimal labor is required.
- There are no chemicals to dispose of.
- It is relatively rapid.
- Many methods and pieces of equipment are commercially available to do the job.

Heating extruder components between 800–1000˚F (427–538˚C) removes all carbon deposits and residue; only inorganic materials remain, and they are easily brushed off the parts once they cool to room temperature. Dies, breaker plates, and twin screw elements are easily cleaned with this method. Single screws, however, are long, single-piece objects and require long ovens to burn off ingredients. In many applications, equipment is purchased to burn polymer off the smaller equipment pieces, with single screw extruder screws being mechanically cleaned. As with any equipment, the cost has to be justified versus time saving and productivity to determine if the payback time is acceptable.

Furnaces available to clean screws are gas fired, electric, or a combination of gas and electric. Pyrolysis of the polymer residue at high temperature removes all organic material. Pyrolysis smoke and gases generated through the heating operation can be quite significant in cases where a large amount of polymer is removed. Units to control the off-gases include afterburners and scrubbers. Afterburners consume the pyrolysis gases at 1400˚F (760˚C) in a secondary chamber. As the polymer being removed at one time increases, pyrolysis product venting to the atmosphere must be controlled. Pollution Control Products Company[13] makes a number of furnaces large enough to clean screws. The large furnaces have tracks to roll a screw-holding cart in and out. Their controlled pyrolysis procedure heats the parts to 800–900˚F (427–472˚C) in an oxygen-deficient environment, generating volatile off-gases that are passed through a combustion chamber and burned, generating exit gases that are primarily CO_2 and H_2O. Polymer that melts during the heat-up cycle is drained into a cooled collection chamber where the polymer can easily be removed. After cleaning, any inorganic pigments, glass, fillers, or other inorganic residue remaining from the polymer has to be removed from the screw by wire brush or ultrasonic cleaning (mainly for monofilament dies), or dipped into dilute acid (to remove TiO_2). Problems with this technology are fires and fume removal. Fires are controlled through reduced oxygen during the heating step, and fume emission is controlled by afterburners or scrubbers.

Rieter Corporation[14] produces a vacuum pyrolysis unit that cleans parts in a four-step process. Initially the parts to be cleaned are slowly heated to allow excess polymer to melt and drain away from the part into an unheated chamber, where the residue can be collected. Then the parts are heated further, and the pyrolysis decomposition products are drawn off under vacuum and cooled with a water mist in a secondary chamber. Third, any remaining organic material is oxidized, followed by cooling of the equipment parts. Large parts can be placed on dollies to roll in and out of the oven.

Small part high-temperature cleaning can be accomplished using a muffle furnace set for one or two hours of exposure to temperatures of 800–1000˚F (427–538˚C). Overheating die components, breaker plates, and screw elements can cause damage to the parts through the loss of hardness. Tempyrox Company Inc.[15] makes thermal cleaning systems for the plastics industry composed of relatively small ovens that eliminate the smoke from pyrolysis through the use of electric, gas, and oxidizer technologies. Gases generated are heated to 1300–1500˚F (704–816˚C), breaking down any residual pyrolysis products to CO_2 and H_2O. After pyrolysis, parts are cooled to room temperature, brushed, and are ready to be reused in the extruder.

An older technology is a molten salt bath, where inorganic salts are heated to their molten state. Screws, screw elements, breaker plates, and die components are placed in a wire basket and immersed in the hot molten salt,

vaporizing the organic materials. The advantage of this method is cleaning speed, as the salt baths are kept hot. Disadvantages are the potential for the hot molten salts to splatter, and the bath can explode if parts contain any liquids, particularly water, that react violently with the molten metal salts.

A technology similar to the molten salt bath without the negatives is a fluidized bed, where hot aluminum oxide particles are kept in motion by air or nitrogen fed through the bed. With the proper air flow, an alumina bed becomes fluidized and exhibits liquid characteristics. Similar to the molten salt bath, parts are suspended in the bath in a wire basket. Advantages of this approach are the fluidized bed has no boiling point, odors, or vapors; the procedure is very rapid with good temperature control and heat transfer from the alumina surrounding the parts. The disadvantage of this procedure is the control of the pyrolysis products generated in the cleaning. A combustion chamber may be required above the fluidized bed to ignite the gases. Procedyne Corporation[16] produces fluidized beds for cleaning extrusion parts and screws. Alumina dust sticks to the metal parts and has to be removed.

A third cleaning method is cold jet dry ice blast. Pellets of dry ice are blasted onto the polymer-coated screw and extruder parts, much like sand blasting. CO_2 blast cleaning uses solid dry ice particles in high-velocity airflow to remove contaminates from surfaces. Dry ice pellets are accelerated through high-velocity nozzles to impinge upon the surface being cleaned. The thermal shock effect separates the residue from the surface; the residue falls away from the surface and is easily swept up. The dry ice pellets vaporize at impact, leaving no other cleaning byproduct and therefore eliminating the added cost and inconvenience of secondary waste treatment and disposal. Dry ice blast cleaning extends the useful life of equipment by facilitating regular, thorough cleaning. And because it is nonabrasive, dry ice blast cleaning does not wear on parts or contribute to surface erosion.[17] The thermal shocks cause microcracks in the polymer, causing it to break up as the CO_2 expands. The process has to be optimized for the particular applications through nozzle size and dry ice pellet size. Advantages of this process are no gases are generated that require proper disposal, and polymer residue is easily swept up and disposed of. Disadvantages are the need to obtain pelletized dry ice supplied by Cold Jet Inc. and the development of the proper nozzle and spray technique for the specific cleaning requirements. Dry ice pellets are produced locally and supplied by container to coincide with specific cleaning requirements.

The final cleaning method is solvent cleaning, where contaminated or fouled parts are placed in a solvent bath and the polymer is dissolved. This method has the advantage of minimal equipment, as only a tank filled with solvent in which to place the parts being soaked is required. There are many disadvantages:

- Disposal of contaminated solvent
- Need for different solvents for different resins
- Solvent flammability
- Need for protective clothing and breathing apparatus
- Lengthy dissolution process
- Carbon deposits are not likely to be dissolved by solvent.

Solvent can be used with mechanical cleaning where the entire screw is not placed in a solvent bath. Solvent or cleaning solution applied with rags may be used to clean dies.

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Review Questions

1. What are the four different mechanisms found in polymer purge systems?
2. What is the typical procedure to use when purging an extruder with a chemical purge?
3. What is the typical procedure to use when purging an extruder with an abrasive purge?
4. Why are purge materials developed with different temperature processing ranges?
5. What procedures are used with a surfactant-type purge?
6. How does a mechanical purge function to clean an extruder?
7. Why and when are purge materials used?
8. What are four methods of cleaning an extruder screw?
9. What are the disadvantages of heating a screw in a furnace to burn off any polymer or residue?
10. How is dry ice blast used to clean extruder parts?
11. What are the disadvantages of solvent cleaning?
12. What are the disadvantages of mechanically cleaning extruder parts?
13. What is one method of cleaning the extruder barrel?

Part 6: Coextrusion

40 Coextrusion Applications

Many extrusion processes use coextrusion to optimize product performance by combining multiple resins with different properties to produce products with combined features that are not feasible using a single resin system. Many coextrusion applications are in the areas of packaging, where specific resins with attributes such as

- moisture resistance
- oxygen barrier
- film for toughness
- improved heat sealability
- improved appearance

are combined to produce a product that can be used in multifaceted applications. The driving force is an improved product at lower cost. Cast, blown, and oriented film processes are used to produce coextruded products that are used in stretch wrap, decorative films, medical applications, trash bags, agricultural films, shrink film, shipping bags, and food packaging films for meats, cereal, snack foods, bakery goods, frozen foods, and so forth. Other packaging applications include sheet that is thermoformed and multilayer blow molded bottles. Coextruded sheet is used in construction, food packaging, and medical applications. Coextrusion is used in large-part blow molding to produce products such as gas tanks. In pipe applications, coextrusion is used to apply a stripe to one or more of the pipe sides, provide multilayer pipes that may have one color material on the inside and another color or material on the outside—particularly in corrugated pipe—and multilayered pipes for automotive fuel lines. Wire coating can be coextruded with a foam layer and a solid insulative layer or two layers of insulation for use in power cables. Profile coextrusion is used to make siding for homes, apply an elastomeric or flexible strip onto the edge of a solid profile, produce a profile with two different colors, and so forth. Most extrusion processes, discussed in Part 7, use coextrusion in some form to produce products that cannot be made with just one resin source. Part 6 covers the different aspects of coextrusion. Chapter 40 covers some of the applications and materials used in the different processes; the remaining chapters cover the coextrusion process, equipment criteria to define the resins for coextrusion, and troubleshooting.

Coextrusion applications are driven by economics, properties, and market needs. Coextrusion economics allow cheaper products to be produced by using recycled or reprocessed material or regrind in an inner layer of the structure to reduce resin costs. Pigment costs can be reduced by putting the desired pigment only in the appearance layers, using natural resins in the inner layers. Less expensive resins can be used in the inner layers even if they are not compatible with the other resins in the structure through the use of tie or adhesive layers to bond the noncompatible resins together. Market drivers are the combination of appearance plus functionality. Different color resins can be used on the inner and outer surface to provide a color contrast that is more visually pleasing. Different functionality can be applied to the inner surface compared to the exterior surface, such as a tackifier on the inside of shrink film or stretch film. A printable or glossy surface can be used as the exterior layer for improved printability and/or high gloss. Property improvements mentioned above include barrier properties to the transmission of moisture, carbon dioxide, and oxygen; improved heat resistance; better heat sealability in packaging applications; improved thermoformability; improved strength and rigidity; low odor; and low taste.

Successful coextrusion is dependent on the feedblock, die, and material selection. If the resin is properly melted in the extruder and delivered to the feedblock or die at the proper melt temperature, where it is combined with other resins in the melt state, coextrusion can be straightforward. The different extruder rates are adjusted to provide the correct resin ratio for the desired layer thickness in the final structure. If a feed block is used, the proper flow channel settings are required to feed the correct amount of each ingredient to the die. As the polymer flow is laminar, there is no mixing between adjacent layers. To assure a stable coextruded product, the shear stress in each layer has to be below a critical level to provide a uniform flow condition. Once the extrudate leaves the die, downstream equipment is the same as that used for extruding a single resin.

40.1 Film

Typical resins used in coextruded film applications include

- Low density polyethylene (LDPE)
- Linear low density polyethylene (LLDPE)
- High density polyethylene (HDPE)
- High molecular weight polyethylene (HMWPE)
- Ethylene vinyl acetate (EVA)
- Ionomer
- Polypropylene (PP)
- Polyamide (PA)
- Polycarbonate (PC)
- Polyethylene terephthalate (PET)
- Ethylene vinyl alcohol (EVOH)

Table 40.1 summarizes the attributes each of these resins makes to a coextruded film structure.

The majority of coextruded films go into packaging applications where the important criteria are

- Barrier to the transmission of gases such as oxygen, nitrogen, and carbon dioxide
- Moisture barrier
- Barrier to flavors and aromas
- Barrier to UV radiation
- FDA approval

Structures are generated based on the following requirements:

- Polymer in contact with food
- Oxygen and moisture barrier requirements
- Food flavor change
- Package sealing
- Transparency
- Printability on the outer surface

Since some of the resins systems that provide the unique properties are not compatible, tie layers or adhesive layers between noncompatible resins are necessary to tie the structures together.

Table 40.2 provides some general compatibility guidelines for coextruded resins that will adhere to each other and those that require tie or adhesive layers to provide appropriate adhesion in packaging applications. In general, polar resins adhere to polar resins and nonpolar resins adhere to nonpolar resins. Polar and nonpolar resins repel each other, providing no adhesion. As an example, coextruding polycarbonate (polar resin) and polypropylene (nonpolar resin) and combining the two resins in the melt state in either a feedblock or coextrusion die produces a two-component product with no adhesion between the two components. The resins, although combined in the melt state, where the materials

Table 40.1. Attributes of Various Resin Systems

Polymer	Attributes	Contribute to Coextrusion
LDPE	Processability, transparency, moisture resistance, low cost, readily available, numerous types, weldable	Large MFI range to match viscosity of numerous resin systems, use in combinations with itself
LLDPE	Readily available, inexpensive, transparency, strength, weldable	Higher processing temperature, use in combination with LDPE
HDPE	Strength, rigidity, toughness, large number of flow grades, higher processing temperature than LDPE	Use in combination with LDPE
EVA	Puncture resistance, weldable, use as a tie layer, good adhesion to PP	Limited temperature range, use as a welding material
Ionomer	Puncture resistance, weldable, good adhesion to polymide	May need to use special steel in feedblock, die, and melt contact areas
PP	High stiffness, wide processing window, many grades, tough, readily available, good moisture resistance	Processes at higher temperatures compared to PE
Polyamide	Flavor and aroma barrier, good chemical resistance, higher temperature	Nylon 6,6 processes at higher melt temperatures than nylon 6
PC	Optical transparency, temperature resistance, surface gloss, impact, puncture resistance, high cost	High processing temperature, speciality applications
PET	Puncture resistance, transparency, inexpensive high temperature resin	Combine with polyvinylidene chloride to produce moisture and oxygen resistance barrier
Polyvinylidene Chloride (PVDC)	Inexpensive, oxygen barrier, clear, tough, clings to container in use	Limited temperature processing range
EVOH	Oxygen barrier, clear, reasonable moisture resistance, numerous ethylene contents available, flavor retention	Larger processing range than PVDC but limited high heat stability

Table 40.2. Resin Compatibility

Material	LDPE	LLDPE	HDPE	Ionomer	EVA	PP	PA6	PET	PC	EVOH	PS
LDPE	GA										
LLDPE	GA	GA									
HDPE	GA	GA	GA								
Ionomer	V	V	V	GA							
EVA	GA	GA	GA	TL	GA						
PP	TL	TL	TL	TL	GA	GA					
PA 6	TL	TL	TL	TL	TL	TL	GA				
PET	TL	TL	TL	TL	TL	TL	TL	GA			
PC	TL	TL	TL		TL	TLTL	TL	TL	GA		
EVOH	TL	TL	TL	TL	TL	TL	GA	TL	TL	GA	
PS	TL	TL	TL	V	TL	TL	TL	TL	TL	TL	GA

GA = Good Adhesion
TL = Need for Tie Layer or Adhesive to Provide Bonding
V = Variable Adhesion Depends on Resin Type

could bond and react, exit the die in molten state as two separate layers. By the time the film is cooled on the casting roll or is passed through a three-roll stack, the two films peel or fall apart with no adhesion between the layers. To produce a coextruded PC–PP structure, a tie layer is required that bonds to both the PC and the PP to act as an adhesive holding the structure together. Concepts of compatibility and tie layers are being discussed as part of coextruded film; however, the need for tie layers in film is the same as pipe, profile, wire coating, and other coextrusion applications. Packaging applications typically contain a more diverse range of polymers because of the specialized requirements associated with the packaging industry. While the example above considered cast film and sheet, the same product diversity exists in blown film applications, where 11-layer coextruded products are becoming commonplace.

Several tie layers or adhesives can be used with different resin systems to provide adequate adhesion between layers. Tie layers are functionalized resins that will bond to adjacent nonpolar and polar resins to act as an adhesive, bonding different resins together. Figure 40.1 shows typical structures where two tie layers are used in a five-layer structure to coextrude PE with EVOH. Adhesion between the resins and the tie layer is controlled by the chemistry involved. At the polymer

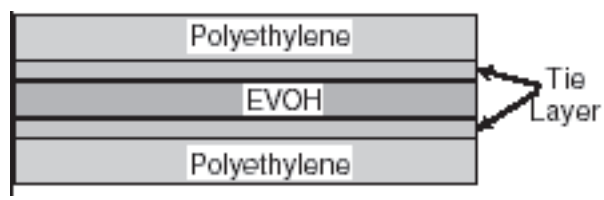

Figure 40.1. Five-layer structure with tie layers.

interface, adhesion occurs through molecular chain entanglement in the melt phase, Van der Waal forces, and co-crystallization. This occurs in the melt state at the interface of the polyethylene with the backbone of the functionalized polyethylene tie layer. At the incompatible interface between the EVOH and the tie layer, a chemical bond, hydrogen bond, or dipole-dipole interaction is occurring to provide the adhesion mechanism.[1] While more than one adhesion mechanism may occur at the incompatible interface, the driving reaction is chemical bonding between the functionalized portion of the polyethylene molecule and the polar material.

Covalent bonds are formed by the reaction of pendant groups in the tie layer and pendant groups in the adjacent resin. See Fig. 40.2 for chemical formulas and simplified structures. Many tie layers are made by grafting maleic anhydride [$C_4O_3H_2$] onto polyethylene. The anhydride

group [RCO_3CR] will react with the hydroxyl group [OH] of EVOH (ethylene vinyl alcohol) to provide an ester linkage [RCOOR]. Similarly, the maleic anhydride modified polymers will react with the amide group [$RCONH_2$] on nylon. Hydrogen bonding is a secondary or weaker bond than the covalent bond. Hydrogen bonding is the attraction of a hydrogen atom on a polar molecule such as an organic acid [COOH], a hydroxyl [OH], an amine [NH_2], or an amide [$CONH_2$] to the oxygen atom on the [C=O] maleic anhydride modified polyolefin.

In coextrusion processes, the adhesion level is driven by processing factors, assuming an appropriate tie layer is employed as an adhesive between the two resins being bonded together. Adhesion is affected by

- Tie layer thickness
- Increased functionality in the tie layer
- Melt temperature
- Contact time[2][3]

Increasing thickness provides a thicker region to distribute the stress between the two substrates while providing more bonding sites. Thickness provides increased adhesion only until a critical tie layer thickness is reached, where the adhesion levels out. Increased functionality improves the bonding characteristics by providing more sites for covalent bonding between the incompatible resin and the tie layer. Contact time is similar to a time-temperature effect. Increasing the contact time in the molten phase between the tie layer and either resin allows more time for secondary effects such as molecular chain entanglement with the compatible resin or hydrogen bonding and dipole-dipole interactions to occur with the noncompatible resin. Obviously higher melt temperatures result in lower viscosity and more molecular motion. Increased molecular motion allows for more polymer chain entanglement with compatible resins. Similarly, the higher molecular motion increases the probability that the functionalized part of the chain will come in contact with the correct chemical species in the incompatible layer to form a covalent bond or establish other secondary interactions. Higher melt temperatures increase the rate of reaction or the kinetics of the bonding reaction.

Coextrusion processing effects that can reduce tie layer adhesion are tie layer orientation and line speed. High molecular orientation causes the polymer chains to become aligned in the machine direction. This stretches the polymer chain, reducing the possibility of molecular entanglement with the compatible resin. Higher line speeds reduce the time at temperature and the contact time, lowering adhesion. In addition to reduced contact time in the die or feedblock, line speeds also relate to cooling rates. Once the product is solidified, additional molecular entanglement does not occur and covalent bonding is complete. At temperatures above T_g, some minor molecular motion still occurs that can improve secondary bonding, such as hydrogen bonding or dipole-dipole interactions.

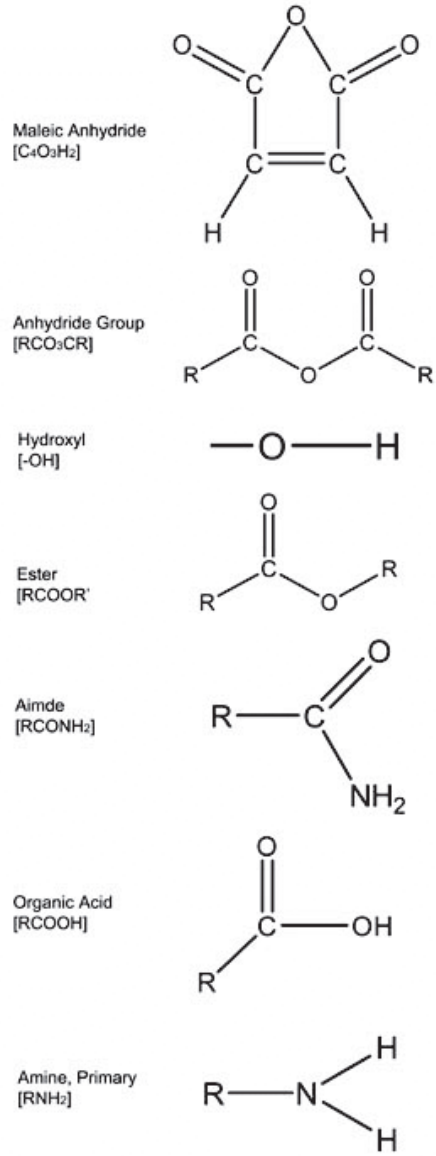

Figure 40.2. Chemical structures.

In blown film coextrusion, the blow-up ratio can decrease adhesion between the tie layer and the adjacent layers. This occurs for the same reason molecular orientation decreases adhesion. A higher blow-up ratio results in greater orientation. Adhesion is increased with the amount of time it takes to get to the frost line.[3]

Proper tie layer selection is most dependent on the required adhesion level between the two adjacent resins. While many tie layer resins are available, they do not all have the same functionality grafted onto the polymer

backbone and are not compatible with all resins. The principal criterion in the selection of a tie resin is that it will bond equally well with the adjacent resins in the coextruded structure. The second factor is the adhesion based on the tie layer thickness. While thicker resin layers provide better adhesion, what is the application adhesion requirement? If the adhesion requirement is less than maximum strength, the tie layer thickness can be reduced. Adhesion requirements are defined for the worst possible scenario the application may experience to ensure that the adhesion between layers meets all environments the application experiences. Optical clarity is required in some applications. Resins for the structure as well as the tie layer must be carefully selected to meet the overall optical clarity requirements. Economics are always an issue. This comes back to the minimum thickness required to provide adequate adhesion. Melt viscosity of the tie layer at process melt temperature has to be similar to the melt viscosity of the adjacent films.

Materials used as tie layers include

- EVA
- Maleic anhydride modified LDPE
- Maleic anhydride modified HDPE
- Maleic anhydride modified LLDPE
- Maleic anhydride modified PP
- Acid modified ethylene acrylate
- Maleic anhydride modified ethylene acrylate
- Anhydride modified EVA
- Acid acrylate modified EVA
- Ethylene methyl acrylate
- Ethylene ethyl acrylate

A list of tie layer resins for different structures is given in Table 40.3. Some of the tie layers are identified as adhering to metal and paper. A two-layer coextruded structure of PE/tie layer can be extrusion coated onto paper or foil to provide excellent adhesion between the

Table 40.3. Tie Layer Materials with Resin Adhesion

Material	Adhesion to Resins
Plexar®[1] Anhydride Modified EVA	EVA, PP, PA, EVOH, Ionomer, PA, PET, PS, PC, metal, paper
Plexar®[1] Anhydride Modified HDPE	HDPE, EVA, PP, PA, EVOH, Ionomer, PA, PET, PS, PC, metal, paper
Plexar®[1] Anhydride Modified LDPE	LDPE, EVA, PP, PA, EVOH, Ionomer, PA, PET, PS, PC, metal, paper
Plexar®[1] Anhydride Modified LLDPE	LLDPE, EVA, PP, PA, EVOH, Ionomer, PA, PET, PS, PC, metal, paper
Admer®[2] Functionally Modified Polyolefin	LDPE, LLDPE, HDPE, PP, EVOH, Ionomer, PA, PC, PS, PET, PS, metals
Bynel®[3] Acid Modified EVA	HDPE, LDPE, LLDPE, PP, PS, PVDC, PET
Bynel®[3] Acid Modified Ethylene Acrylate	PA, Ionomer, HDPE, LDPE, LLDEP, PP, Ps, PET, paper, metal
Bynel®[3] Anhydride Modified Ethylene Acrylate	EVOH, PP, HDPE, LDPE, LLDPE, Ionomer, PA, PET, PC
Bynel®[3] Acid/Acrylate Modified EVA	Ionomer, PVDC, HDPE, LDPE, LLDPE, PP, PA, PET, PS
Bynel®[3] Acid Anhydride Modified EVA	PA, Ionomer, HDPE, LDPE, LLDPE, PP, PS, EVOH, PET, PC
Bynel®[3] Anhydride Modified HDPE	PA, HDPE, LDPE, LLDPE, EVOH
Bynel®[3] Anhydride Modified LLDPE	PA, HDPE, LDPE, LLDPE, EVOH, PP, Ionomer
Bynel®[3] Anhydride Modified LDPE	PA, HDPE, LDPE, LLDPE, EVOH, Ionomer
Bynel®[3] Anhydride Modified PP	PA, EVOH, PP

1. Tie material supplied by Equistar Chemicals, Houston, TX
2. Tie material supplied by Mitsui Chemicals America, Inc., Purchase, NY
3. Tie material supplied by E. I. du Pont de Nemours and Company, Wilmington, DE

polyethylene and the paper or foil substrate. Table 40.4 shows the physical property characteristics of the HDPE, LDPE, and LLDPE modified polyolefins plus the physical property characteristics of EVA.

Work continues in the development of new tie layers to allow different resin systems to be brought together. A tie layer was made by reacting polypropylene graft maleic anhydride with poly(methylmethacrylate-co-hydroxymethylmethacrylate) to produce a tie layer for a polypropylene copolymer containing EPDM and polyvinylidenefluoride.[4] The adhesion of PP to PA 6 with a tie layer produced by melt blending PP with PP grafted with maleic anhydride was characterized.[5] Results show that higher levels of the PP graft maleic anhydride in the tie layer yield better adhesion. In addition, the adhesion to PA 6 was better than adhesion to EVOH.

The strength of the adhesive bond between different layers in a coextruded structure is measured using a T-peel test.[6] The strength required to pull the layers apart using a universal stress-strain tester is measured. Two different tie layer failure mechanisms are adhesive failure and cohesive failure. In adhesive failure, the tie layer separates from one of the resin interfaces. In cohesive failure, the tie layer bond to the resin substrates is greater than the tie layer strength, and the tie layer physically separates. The two failures are shown in Fig. 40.3. Factors affecting peel strength are

- The tie layer itself
- How it bonds to each of the substrates
- Coextrusion processing variables (discussed previously)
- Testing variables—temperature and speed of test

Testing is done off-line after 48 hours of conditioning at constant temperature and humidity. Replicates are tested to determine variability in the sample.

Tie layer adhesion can change over time. Under optimum processing conditions, adhesion between the substrate and the tie layer can continue to improve for up to two weeks. Changes in bond strength are dependent on the adhesion mechanism. Covalent bonding between the tie layer and substrate will occur almost instantaneously. However, dipole-dipole interactions, hydrogen bonding, molecular entanglement, or Van der Waal forces can continue to change over time until an equilibrium level is attained. Molecular motion in the polymer chains continues to occur above T_g. With both PE and PP, the polymer T_g is below room temperature; consequently, at room temperature there still is substantial molecular motion in the amorphous phase, allowing some chain entanglement and co-crystallization at room temperature. While the molecular motion is limited, the process continues until the substrate reaches an equilibrium crystallinity level.

Tie layer adhesion is also dependent on the end-use application of the film. Some films are used at high temperature, others in boiling water or in retort (steam sterilization environments). Tie layer or peel strength has to be tested in the environment in which the product is used to ensure it meets minimum product requirements while providing some minimal acceptable safety factor.

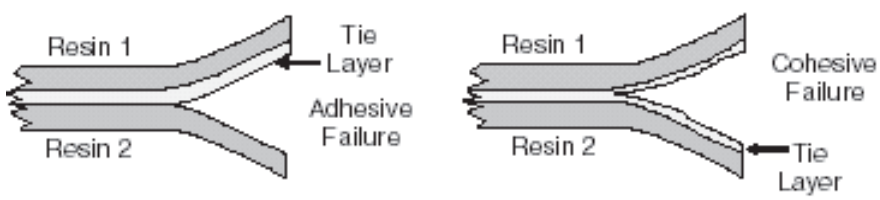

Figure 40.3. Tie layer adhesive and cohesive failures.

Barrier resins are normally incorporated in the middle of packaging film structures to provide resistance to oxygen (O_2), moisture (H_2O), carbon dioxide (CO_2), or nitrogen (N_2) penetration from outside the package to the inside or inside the package out. Oxygen penetrating food packaging from the outside can cause food to spoil,

Table 40.4. Characteristics of Modified Polyethylene Tie Layers[6]

Base Material	Characteristics
HDPE	Good melt strength, high melt point of 257˚F (125˚C), highest stiffness, good chemical resistance, high tensile properties
LDPE	Better clarity than HDPE, easily processed, lower modulus than HDPE
LLDPE	Good heat sealability, tough, puncture resistant, best tear resitance, high elongation, best environmental stress crack resistance
EVA	Elastomeric, good adhesion to polar substrate, best environmental stress crack resistance, barrier properties, tough at low temperatures

degrade, lose flavor, and in some instances change color. Permeation of gases from inside packaging structures to the outside can allow carbonated soft drinks to go flat or retard spoilage of some foods packaged under a particular environment needed to maintain freshness and product shelf life. In addition to preventing gas permeation, barrier films are used to preserve flavors while also preventing extraneous flavors or aromas from penetrating the package and being absorbed by the contents. The gas transmission rate is given by Eq. (40.1):

$$GTR = \frac{P \times A \times \Delta p}{L} \tag{40.1}$$

where

GTR = Total gas transmission rate (cc/24hr)

P = Permeability of the resin (cc-mil/100 in^2/24hr at 1 atm)

A = Package area

Δp = Pressure difference across the barrier

L = Layer thickness

The temperature at which the test is performed has a direct bearing on the gas transmission rate. At higher temperatures, the rate increases substantially. Other factors in Eq. (40.1) that have a direct bearing on the transmission rate are pressure difference across the barrier and the effect of moisture on both sides of the barrier. As the relative humidity increases or air moisture content increases, the oxygen permeability of some barrier resins increases. The last factor in Eq. (40.1) that plays a significant role in the barrier properties of the film is the barrier resin thickness. Obviously as the barrier layer thickness increases, the transmission rate decreases because thickness (L) is in the denominator. Table 40.5 shows some oxygen transmission rates for various commercially available barrier resins. As seen in Table 40.5, EVOH has the lowest oxygen transmission rate, followed by PVDC, acrylonitrile methacrylate copolymer (Barex® 2109), MXD6 nylon, oriented PET, and nonoriented nylon 6. Olefinic polymers have a high oxygen transmission rate.

Relative humidity either inside or outside of the package can affect the oxygen transmission rate. Increasing the

Table 40.5. Oxygen Transmission Rates at 0% Relative Humidity[8]

Material	5°C[1]	23°C[1]	35°C[1]	50°C[1]	5°C[2]	23°C[2]	35°C[2]	50°C[2]
EVOH – 27% ethylene	0.022	0.095	0.231	0.637	0.001	0.006	0.015	0.041
EVOH – 38% ethylene	0.090	0.395	0.946	2.600	0.006	0.025	0.061	0.167
EVOH – 48% ethylene	1.034	1.800	2.700	6.110	0.067	0.116	0.174	0.394
PVDC[3] – Saran MA	0.093	1.240	4.464	19.80	0.006	0.080	0.288	1.280
AN[4] – Barrex 210	2.325	12.40	31.00	95.00	0.150	0.800	2.00	6.129
MXD6 nylon	0.670	2.325	4.430	14.26	0.043	1.150	0.280	0.920
Oriented PA6	7.590	25.59	51.15		0.490	1.780	3.300	
Non-Oriented PA6	22.30	78.74	154.9		1.439	5.080	10.00	
Oriented PET	10.23	35.64	79.04	260.0	0.660	2.300	5.100	16.78
HDPE		2325	4448			150	287	
LDPE		8586	11547			554	745	
Oriented PP		2526	3146			163	203	
PS		4030				260		

1. Units = cc. 25μ/m2 * 24hr * atm
2. Unites = cc. mil/100 in.2 * 24hr * atm
3. PVDC = Polyvinylidenechloride or Saran
4. AN = Acrylonitrile methacrylate copolymer

relative humidity from 10 to 90% increases the oxygen transmission rate by greater than 10-fold in EVOH and by 8 times in MXD6 nylon, while the oxygen transmission rate remains unchanged with PVDC and Barex® 210. The relative ratings of these four materials, from lowest to highest oxygen transmission rate at low relative humidity, are

- EVOH
- PVDC
- MXD6 nylon
- acrylonitrile-methacrylate copolymer

At 90% relative humidity the ratings are

- PVDC
- acrylonitrile-methacrylate copolymer
- MXD6 nylon
- EVOH

The oxygen transmission rate of Selar® PA3426 amorphous nylon from DuPont improves at higher relative humidity. As an example, the oxygen transmission rate at 0% relative humidity is approximately 2.6 cc-mil/100 in^2/24 hours at 1 atmosphere, and at 40 to 80% relative humidity the oxygen transmission rate is approximately 1.3 cc-mil/100 in^2/24 hours at 1 atmosphere.

Other factors affecting the gas transmission rate are temperature and gas. Higher temperatures increase the gas transmission rate in all barrier resins. As the temperature increases, the polymer molecular motion expands the structure, which provides more space for gas molecules to pass through the structure. Transmission rates of other gases through the structure will differ from the oxygen transmission rate. The relative ratings of different barrier materials from best to worst for CO_2 at 10% relative humidity are

- EVOH
- PVDC
- acrylonitrile-methacrylate copolymer
- amorphous nylon
- copolyester

At 100% relative humidity, the relative ratings for CO_2 barrier from best to worst are

- PVDC
- amorphous nylon
- acrylonitrile-methacrylate copolymer
- copolyester
- EVOH

For nitrogen at 10% relative humidity, the lowest to highest rates are

- EVOH
- PVDC
- MXD6 nylon
- acrylonitrile-methacrylate copolymer

At 100% relative humidity, the lowest to highest transmission rates for nitrogen gas are

- PVDC
- EVOH
- acrylonitrile-methacrylate
- MXD6 nylon

What all these data suggest is the type of coextruded structure to use in an application depends on the environment and packaging requirements. What is the relative humidity where the packaging material is going to be used? What gases need to be kept from the product or in the product? Obviously, in food packaging oxygen must be kept from contact with some products to prevent oxidation and spoilage of the product. In other situations, gas must be kept in the product, such as carbonation in soft drinks, requiring good CO_2 barrier properties.

One final important barrier property is moisture vapor transmission rate. High moisture vapor transmission rates can cause ingredients in a package to dry out or, conversely, allow the ingredients to pick up moisture and become soggy. Regardless of which direction the moisture transmission occurs, the product is ruined. Some moisture vapor transmission rates are given in Table 40.6. Based on the moisture transmission rates, resins that provide good barrier properties to gas transmission do not provide good barriers to moisture. Consequently, structures requiring both good gas transmission barriers and moisture vapor barriers need at least two different barrier resins to meet the product requirements. To provide barriers to both moisture and oxygen or nitrogen transmission, the structure might require a resin layer of both EVOH and PP. From Table 40.2 it is obvious that EVOH and PP do not bond to each other without the addition of a tie layer. Picking Bynel® PP maleic anhydride modified tie layer for the adhesive layer, the structure is now three layers composed of EVOH/Bynel®/PP. A fourth film is necessary to cover and protect the EVOH barrier resin. If the fourth film is PP, another layer of adhesive is required to attach the PP. The structure becomes a five-layer structure designated as PP/Bynel®/EVOH/Bynel®/PP, which requires at least three extruders to produce. With the incorporation of a moisture barrier, the effects of relative humidity on the EVOH is reduced.

"Smart" packaging is being developed to replace the passive packaging that has been used in the past. Passive packaging is defined as packages that were designed to

Table 40.6. Moisture Vapor Transmission Rates[8]

Material	g 25μ/m²/24 hrs	g mil/100 in²/24 hrs
BOPP[1]	5.9	0.38
HDPE	5.9	0.38
PP	10.7	0.69
LDPE	17.7	1.14
PET	20.2	1.3
UPVC[2]	46.5	3.0
PS	131.8	8.5
PC	170.5	11.0
EVOH 38% ethylene	32.6	2.1
PVDC	3.4	0.22
AN – Barex®[3]	94.6	6.1
MXD6 nylon	50	3.2

[1] BOPP = Biaxially Oriented PP
[2] UPVC = Unplasticized PVC
[3] AN = Acrylonitrile-methacrylate

keep oxygen, moisture, nitrogen, and other gases either inside the product or outside the product through physical means of selecting the correct barrier resin combinations. Smart packaging employs a scavenger for oxygen, CO_2, moisture, or ethylene in one of the packaging layers. The scavenger is an active ingredient that will absorb a specific gaseous component that can cause food spoilage. The scavengers are activated after the package has been sealed and can be used to absorb oxygen that penetrates the structure or is generated over time by the package contents, leading to longer shelf life.[10]

There are many different coextruded structures with different layer thicknesses. Figure 40.4 shows several structures with different layer thicknesses. Each resin layer has a letter designation to differentiate the different components. As an example, the first coextruded structure has two layers designated as A and B, where A is the main component and B can be a top layer incorporated into the structure for improved heat sealability, gloss, antistatic, color, or a number of other reasons. Figure 40.4 shows a balanced structure (ABA), an unbalanced structure (AGB), and two- through seven-layer structures. Unbalanced structures are more likely to curl or warp than balanced structures because of the differential shrinkage across the structure. The coextruded structure selected depends on the end-use requirements for barrier properties, moisture resistance, weldability, toughness, use temperature, chemical resistance, and numerous other characteristics.

Many different coextruded structures are currently used in the packaging industry for a variety of reasons. Structures go from two layers made with two extruders to 11-layer blown film structures made with five or more extruders feeding the die. Some typical coextruded film structures and their attributes are summarized below.

Two-layer structures

- LDPE/LDPE: Used in blown film to prevent pinholes in one layer from collapsing the bubble and to create films with different inner and outer colors. Applications are in general packaging.

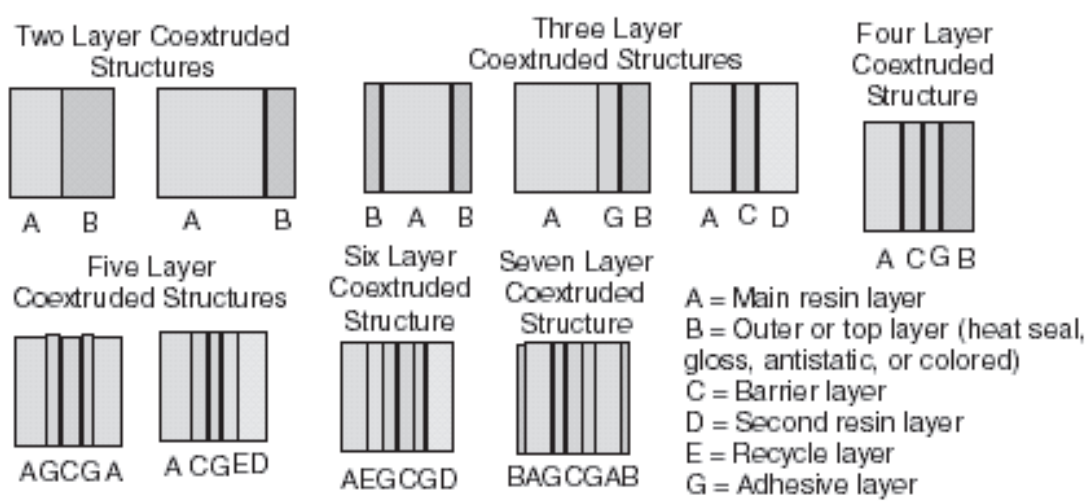

Figure 40.4. Potential coextruded structures for various end-use requirements.

- HDPE/LDPE: Combines the strength of HDPE with the transparency and weldability of LDPE. Applications include food packaging and baked goods.
- LLDPE/LDPE: High elasticity and is used in stretch film applications.
- LDPE/EVA: Excellent weldability and can be sterilized. Films are used for stretch packaging, medical packaging, and heavy duty bags.
- HDPE/EVA: More strength than its LDPE counterpart with the same sterilizability. Other advantages are low-temperature heat sealability, high contact transparency, good gloss, and flow in the heat seal area. Used in food packaging such as cereal and snack foods, medical packaging, and blood plasma bags.
- EVA/LLDPE: High surface adhesion, elastomeric characteristics, and good puncture resistance, making it a good fit in the stretch film market.
- Ionomer/LDPE: Good weldability, clarity, toughness, and puncture resistance and is used to package medical instruments, dairy packaging, food packaging, and other general packaging applications.
- Ionomer/EVA: Resistance to greases and oily-type foods with outstanding sealability, low temperature toughness, and good toughness. Applications include packaging greasy-type products and seafood.
- PA/EA copolymer (Nucrel®10 from Dupont[11]): Used as a "cook in package" film to cover poultry and ham. Nylon (PA) has excellent adhesion to Nucrel®, which is used to provide good adhesion to the foil container, while nylon provides the heat resistance for the cooking cycle. Nucrel® has excellent adhesion to foil and paper, allowing the two-layer film to be extrusion coated onto a foil or paper surface.
- LDPE/EA copolymer: Used for extrusion coating of a foil to form juice boxes. Two layers of LDPE/ethylene acid copolymer are extrusion coated on both sides of a foil with a paper layer bonded to the LDPE. The juice containers are then formed from the combined structure.
- EVA/PP: Good low-temperature welding, excellent flex crack resistance, and is used for packaging cheese, drinking cups, and laminating film.
- PET/EVA: Good puncture resistance, low welding temperature, and good toughness. Used in medical and cheese packaging.
- PET/ionomer: Good toughness and clarity. Used in meat, seafood, and cheese packaging. Ionomer welds quickly to other ionomers at low temperature, providing good weldability. This is particularly beneficial in dog food packaging, where PET/ionomer is the top film bonding to the ionomer that is the top layer in the bottom half of the package. Ionomer bonds well to paper and foil, providing the opportunity to extrusion coat this film onto a paper or foil substrate with good adhesion.

Three-layer structures may be formed from three compatible resins or two noncompatible resins and a tie layer that acts as an adhesive.

- LDPE/HDPE/LDPE: Weldable on both surfaces due to the LDPE, while adding the higher strength of the HDPE core. The balanced structure is less likely to curl. The product is used for food packaging, tomato concentrate, pet foods, cereal, and bakery goods.
- EVA/PP/EVA: Good weldability on both sides with good elastomeric surface layers, and a balanced structure to prevent curl. Used in similar applications to the LDPE/HDPE/LDPE structure above.
- EVA/HDPE/EVA: Similar to the EVA/PP/EVA structure.
- PA6/EVOH/PA6: Excellent gas barrier resistance through a broad relative humidity range, as the polyamide surface layers prevent the moisture from reaching the EVOH and changing its gas transmission effectiveness. The film is used in packaging, providing resistance to gas permeability while acting as a flavor and aroma barrier.
- EVA/PVDC/EVA: Good heat sealability due to the EVA, excellent impact and puncture resistance properties, good low-temperature toughness, and good resistance to flex cracking. Applications are in meat and poultry packaging.
- EVA/ionomer/EVA: Easily thermoformed, has good puncture resistance, low-temperature stability, and resistance to oils and greases. Due to their thermoformability, these structures find use in the formed base of medical pouches.
- OPP/LDPE/EVA: Used to package produce. The oriented polypropylene provides strength and good rigidity, with the EVA and LDPE providing toughness and moisture resistance.
- LDPE/HDPE/EVA: Used for bakery goods and food packaging. It provides good weldability, toughness, and rigidity.
- LDPE/tie/PA or LLDPE/tie/PA: Used to package meat, sausage, ham, fish, and frozen meals. The nylon provides an aroma barrier and water barrier.

The tie layer, required to bond the LDPE to polyamide, must be compatible with LDPE and nylon. A functionalized maleic anhydride modified LDPE tie layer will provide the necessary adhesion.

- PA/tie/EVA: Used to package processed meats. The EVA bonds well to the bottom of a thermoformed package, while polyamide has good temperature range, chemical resistance, and aroma and gas protection, and is receptive to surface printing.

Four-layer structures

- PA/EVOH/tie/LDPE: Used as stretch shrinkable film.
- PA/EVOH/tie/EVA: Used for packaging meat and processed meat. The EVOH provides a gas barrier along with the polyamide (nylon), while the EVA provides good weldability and cling characteristics.
- OPP/tie/HDPE/ionomer: Used to package peanuts. The ionomer provides good welding characteristics plus oil and grease resistance, while the HDPE and OPP provide the package strength and durability.
- PS/styrene butadiene/tie/PETG: Used for vacuum packing due to its good gas barrier properties, aroma resistance, and sealability.

Five-layer structures

- LDPE/tie/PA/tie/LDPE: Used to package cheese, meat, sausage, and ready-made meals. This structure provides resistance to gas and water permeability while keeping the aroma inside the package. Polyamide prevents moisture absorption. The structure is balanced, minimizing film curling, and the LDPE surfaces provide weldability of both sides of the film. In some instances, the LDPE is replaced with LLDPE.
- LDPE/tie/EVOH/tie/LDPE: Used to package meat, fish, and other foodstuff where gas permeation must be limited. The EVOH is incorporated as a gas barrier, with the LDPE acting as a moisture barrier and to provide strength. In addition, the LDPE provides low-temperature weldability.
- HDPE/tie/EVOH/tie/ionomer or HDPE/tie/EVOH/tie/EVA: Used as cereal box liners. The EVA or ionomer provides the heat sealing surface. EVOH provides the gas barrier to keep the cereal fresh, while the HDPE provides both strength and a moisture barrier.
- PP/tie/EVOH/tie/PP: Used for prepared meals and fruit juice containers. It is a good gas, moisture, and aroma barrier. It can be sterilized prior to packaging.
- Ionomer/tie/EVOH/tie/PA: Used as the bottom half of a hot dog package. Polyamide provides the moisture barrier, structural integrity, and thermoformability; EVOH is the gas barrier to keep the hot dogs fresh; and ionomer is the heat seal and chemically resistant layer.
- EVA/tie/PA/tie/EVA: Similar performance and application to the LDPE/tie/PA/tie/LDPE structure discussed above.

Other film types used in multilayer structures include fluoropolymer (Aclar® from Honeywell[12]), which is used in medical applications as barrier blister packaging, laminates, and lid stock. The film is clear, scratch resistant, and flex crack resistant. Ionomer resins are used in packaging because of their clarity, good weldability, thermoformability with deep draws without creating pinholes, and oil and grease resistance. Amorphous nylon films are used for their transparency, barrier properties, chemical resistance, and high-temperature performance. The amorphous nylons can be extrusion blended with EVOH to provide barrier properties, thermoformability, relative humidity insensitivity, and good processing characteristics.

The extrusion system required for coextruded film depends on the number of different resin components in the structure. Each resin requires an extruder; if a regrind layer is incorporated as a separate layer in the final coextruded structure, it requires its own extruder. The extruders are arranged around a die or feedblock with each extruder feeding a separate port in the feedblock or die. (Feedblocks and dies are discussed in more detail in Chapter 41.) Figure 40.5 shows a number of configurations with multiple extruders feeding one die. The die can be a sheet, a flat film, or a blown film die. Melt transfer pipes are used to connect the extruders to the die. The melt transfer pipe length is dependent on where the extruders are placed relative to the feedblock or die. As the number of resins and extruders increases, it becomes more of a challenge to place the extruders in a compact pattern. It is much easier to position three extruders around a die than five or seven extruders. With blown film, the extruders can be located in a circle around the die, with the extruders located as spokes in a wheel around the hub. The hub is the blown film die with the tower located above. In a cast film or sheeting operation, the extruders have to be located behind the die, similar to a fan shape where the die is the base of the fan and the extruders are located in a 150° arc behind the die.

Since the layer thicknesses in the final structure vary, the different extruder output rates will not all be the same. Assume the following structure is being fabricated:

- LDPE/tie/EVOH/tie/LDPE

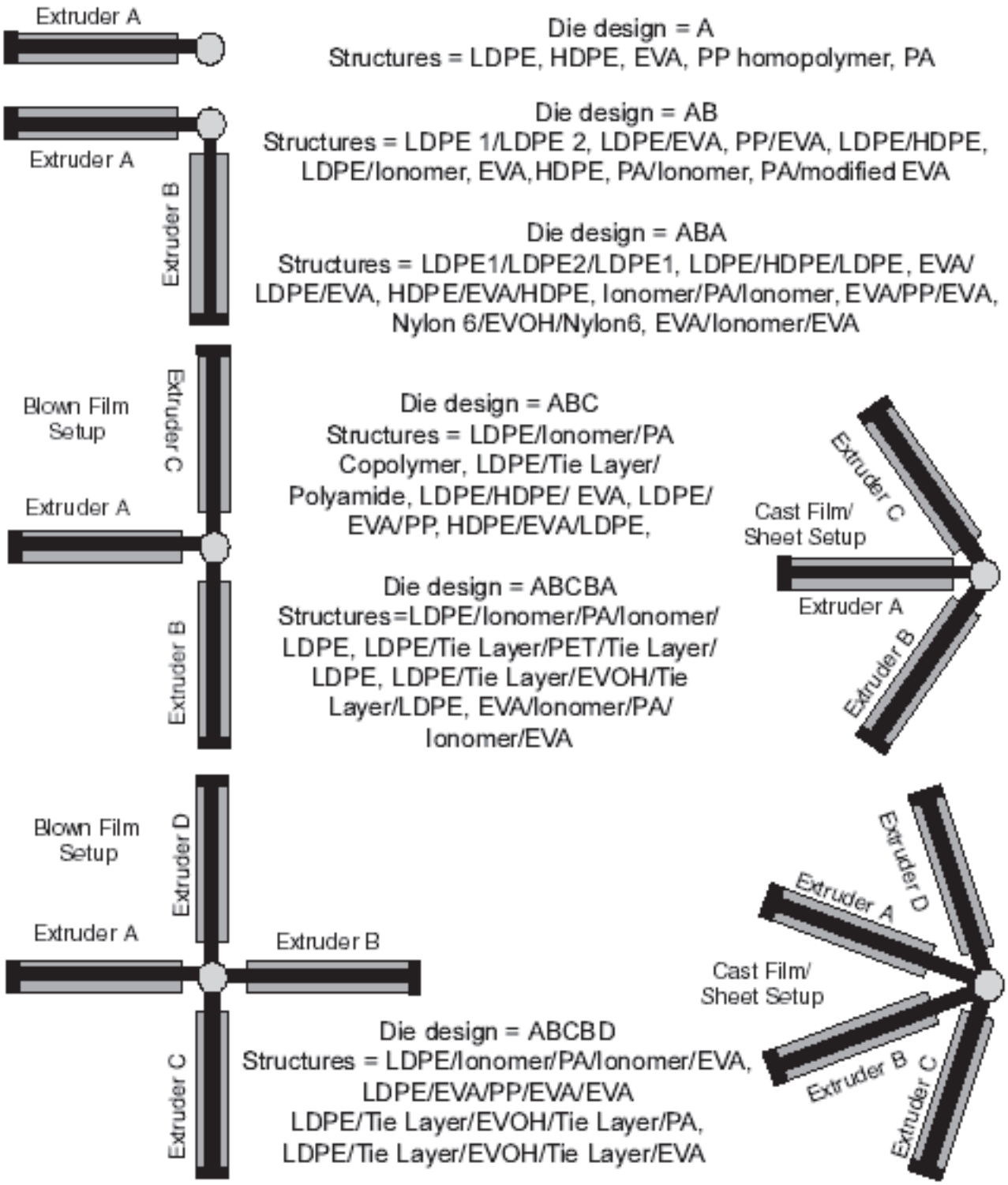

Figure 40.5. Extruder configurations for different coextruded structures.

The tie layers are the thinnest layers, with sufficient thickness to provide the required adhesion to the coextruded structure; the next layer in thickness is the EVOH. EVOH is relatively expensive compared to LDPE; consequently no more is used than is necessary. The thickest layers are the LDPE, as these are providing the structural integrity. The extruder feeding the LDPE is the largest, with the extruders feeding the tie layer and EVOH being approximately equal in size but smaller than the LDPE extruder. Total output for the tie layers and EVOH extruders will be approximately equivalent because there are two tie layers in the structure and only one EVOH layer.

40.2 Sheet

Coextruded sheet is used in a number of thermoforming applications, such as tubs and showers, spas, and packaging. Exterior sheet applications for spas use coextrusion to apply a weatherable surface over a durable, cheaper substrate to provide resistance to weather and chemicals such as chlorine that are required to keep the water clean. Coextruded sheet is used to thermoform packaging containers used for margarine, yogurt, pudding, beef stew, and other food stuffs.[13] Figure 40.6 shows a coextrusion setup to make sheet or film with two extruders and a feedblock to combine the two melt streams. This particular configuration shows a twin screw as the main extruder and a single screw extruder as a satellite extruder supplying the minor layer. This configuration can be used to make AB or ABA structures with the A resin being fed with the twin screw extruder. Each extruder needs to supply the resin to the die at similar melt temperature and viscosity.

Another processing method used to produce sheet is called side-by-side coextrusion, where multiple resins are extruded side by side in a sheet rather than on top of each other in layers. This is useful for making plastic structures of different colors side by side. A typical example is a clear styrene and black styrene resin side by side. The sheet is later thermoformed into packaging containers with a clear top and a black bottom. These are used to provide snacks in airplanes, package salads in delicatessens, etc.

40.3 Pipe, Tubing, and Profile

Coextrusion is used to—

- Stripe pipe
- Produce pipe with a different color on the inside versus the outside
- Customize a pipe with multiple layers for specific applications

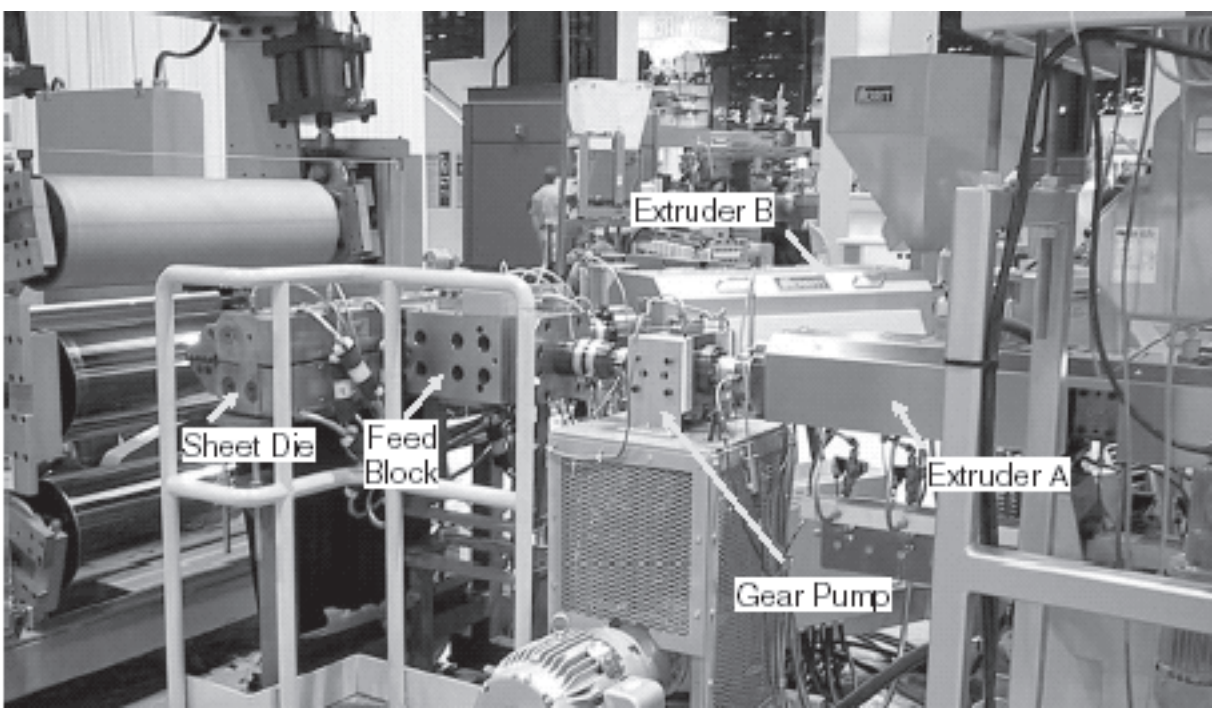

Figure 40.6. Leistritz[14] twin screw extruder and Merrit Davis single screw extruder combined to produce coextruded sheet using a feedblock.

To place a stripe on a pipe, a small extruder is used for the striping resin with a large extruder supplying the melt for the overall pipe. The stripe resin is normally the same resin used for the pipe with added color or pigment. A tie layer is not required as the striping resin is the same as the base resin. The two resins are compatible, and molecular entanglement will provide the necessary adhesion. If the stripe is at only one location, as shown in Fig. 40.7, the die is less complicated than if the stripe is required to be in two, three, or four places. The presence of more than one stripe requires a resin distribution system at the back of the die to provide two, three, or four distinct paths for the stripe resin to combine with the base resin stream. If a mandrel die is being used, the spider legs cannot interfere with the stripe pathway, disrupting the flow, or the stripe will be spread out. Figure 40.8 shows a coextrusion operation that is making pipe with the smaller extruder used to apply the stripe to the pipe.

Pipe coextrusion is used to apply a different color resin or resin with different chemical compatibility to the pipe core. Figure 40.9 shows corrugated pipe with the pipe inside surface a different color from the outside. If a different resin is used in the center of the pipe for purposes of chemical resistance, the structure may require a tie layer. Assume the outside is a high-strength, ductile material such as HDPE or ABS, and the inside is PVC, nylon, PBT, or some other resin providing good chemical resistance to the liquid being transported. A three-layer coextrusion is required with an ABC-type structure, where A might be HDPE, B is an appropriate tie layer, and C is polyamide for good chemical resistance.

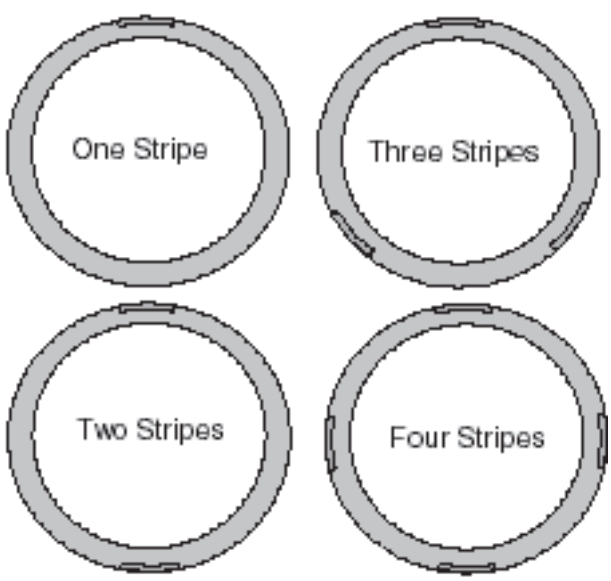

Figure 40.7. Pipe cross sections with different numbers of stripes.

Multilayer pipe can have an aluminum center section covered and/or lined with polyethylene. Unicor MSR pipe[16] has a coextruded inner layer made of HDPE plus an adhesive layer and an outer layer of HDPE. The pipe has the advantages of corrosion resistance, elimination of gas diffusion, limited longitudinal expansion due to the presence of the aluminum, and is light weight. Other coextruded structures found in pipe are

- EVOH/tie/cross-linked polyethylene: Used in heating pipes.
- PE/tie/steel: Used in oil and natural gas pipes.
- Cross-linked polyethylene/tie/aluminum: Used to protect aluminum piping.

Tubing for medical applications, fuel lines, and hoses are only a few of the many applications requiring coextrusion to provide the optimum balance of properties and performance. HDPE/tie/plasticized PVC small diameter tubing is one of many coextruded products used in medical applications. The tubing is clear so fluid flowing through the tube is obvious. Some typical coextruded tubing applications are

- LDPE/tie/nylon: Used in cosmetic packaging. The LDPE provides the flexibility to squeeze while resisting cracking, and the nylon has the required chemical resistance to the cosmetic ingredients.
- LDPE/tie/EVOH/tie/LDPE: Used in toothpaste tubes and medical applications, providing both moisture and gas barrier properties while supplying the toughness and handling characteristics required.
- LDPE/tie/nylon/tie/LDPE: Used in the food industry.

Fuel lines in cars are being developed with coextruded tubing made of fluoropolymers and less expensive ingredients. Two- to six-layer structures are based on ethylene tetrafluoroethylene (ETFE Tefzel®) as the barrier layer, combined with nylon 12 for the chemically resistant layer. Other structures include polytetrafluorethylene (PTFE) and nylon 12, or polyvinylidene fluoride (PVDF) with polybutylene terephthalate (PBT), polyphenylene sulfide (PPS), or elastomer. The key to this technology is the development of tie layers to bond the resins together. Currently sequential extrusion with plasma treatment between the individual extrusion steps is required to provide the necessary adhesion.[17]

Profile extrusion can use coextrusion to attach another resin to an existing profile, change the color of different profile sections, or add functionality to the profile. This is accomplished through incorporation of an elastomeric edge or shape to the edge of a rigid profile to act as a weather strip; see Fig. 40.10. The smaller profiles in Fig. 40.10 show an elastomeric section added between

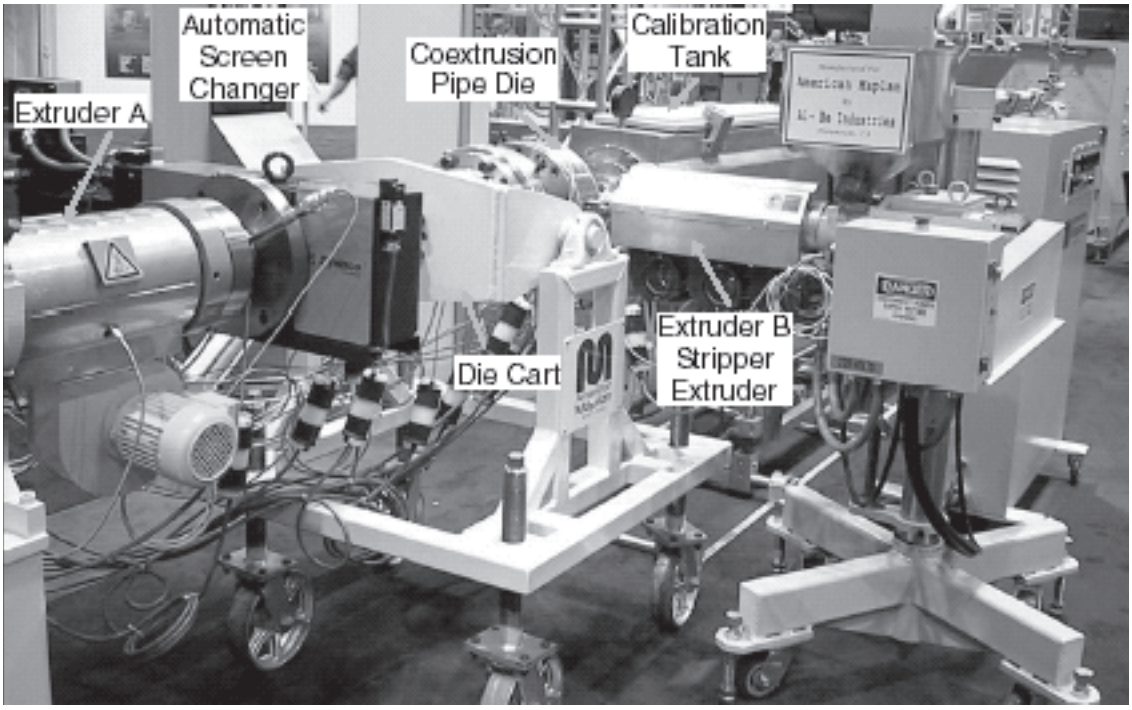

Figure 40.8. American Maplan[15] pipe coextrusion with stripper extruder.

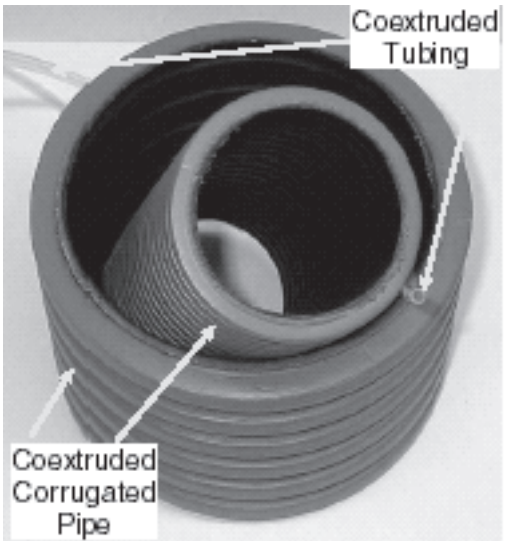

Figure 40.9. Coextruded pipe and tubing.

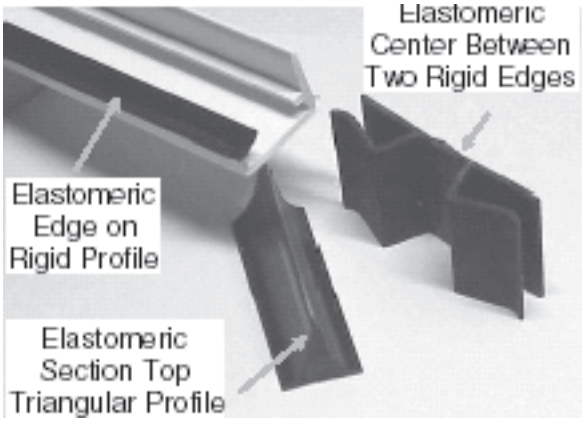

Figure 40.10. Coextruded elastomeric sections on rigid profiles.

two rigid sections and an elastomeric point added to a triangular profile. Other than adding elastomer, the color of different profile sections can be changed. The advantage of coextrusion in profile extrusion is to add functionality to a part in a one-step operation, eliminating the need for secondary operations that may add cost and yet not be as strong or resilient as the coextruded part. As with all finished plastic parts, the objective is to add as much functionality as possible in a given operation to eliminate secondary operations that may be more costly and not as efficiently performed, while minimizing part handling and increasing the part utility.

Profile extrusion used to make siding employs coextrusion techniques to apply a weatherable film over a nonweatherable substrate. More commonly, a thin, darker color layer will be coextruded over a lighter color core as a method of adding color to the profile without having to use costly pigments throughout the entire profile.

40.4 Wire Coating

Coextrusion in wire coating is similar to pipe applications; a second extruder might be added to apply a stripe down the side of the wire or to add a second resin to change the color or the insulation characteristics. In wiring applications, different color wires and different coating configurations are required in large wire bundles to distinguish wires used in applications such as

- Telephone wire
- Electronic applications
- Control panels in industrial equipment
- Underhood applications in cars

Different color-coating configurations make it easier to trace wires and ensure that wires are connected at each end to the correct electrical circuit. Color-coded wires are used for wire identification. While single colors can be added with one extruder to provide a range of coated wire colors, coextrusion is required to add a second, third, or forth color. A different extruder is needed for each color applied. Typically, the same resin is used for each color, eliminating any need for tie layers.

40.5 Large-Part Blow Molding

Coextrusion is employed in large-part blow molding to add functionality to large, hollow parts. Examples of coextruded large-part blow molding applications are

- Gasoline tanks
- Parts with foam centers between two thick skins
- Parts with recycled centers
- Multilayer 30- to 55-gallon drums

Many blow molded parts are made using coextrusion to meet specific applications. The biggest challenge in coextrusion blow molding is to use the scrap, which can be a mixture of different resins with lower property performance than any of the individual components. As an example, consider blow molded gas tanks that are made with up to six different layers. Most tanks are based on HDPE with a barrier added to eliminate hydrocarbon loss through the tank wall. Typical structures are HDPE/tie/EVOH/tie/HDPE. Nylon is used in some applications along with fluorocarbon polymers.

Coextruded drums[18] have the advantage of incorporating additives and pigments in the outside layer, where they are unaffected by the drum contents. Thirty to forty percent of the total drum thickness can be made of recycled resin placed in a center layer away from both the container contents and exposure to either the interior or exterior environment. Most drums are currently HDPE or high-molecular weight HDPE. Other ingredients that are being considered in high-performance drums are amorphous nylon, EVOH, PP, and reinforced layers with either mineral or glass reinforcement. The coextruded drums are made with either accumulator heads or continuous extrusion blow molding. Blow molding processing is discussed in more detail in Part 7.

Large coextruded blow molded parts can be made with a foam core with two thick outer skins. This adds stiffness to the application through increased wall thickness without increasing the part weight. In addition, the foam core can be made with recycled resins to reduce the part cost. The need for a tie layer depends on the core–surface layer compatibility. If the core is the same resin with a chemical blowing agent, no tie layer will be required to provide adhesion between the layers. However, if the core is a recycled resin containing multiple feed streams, a tie layer may be required for optimum structural performance.

The coextrusion process is used in many extrusion processes to bring two or more extruder outputs together to form a single multilayer product that has

- Improved performance
- Product differentiation
- Cost savings
- Added functionality
- Reduced handling
- High throughput

Coextrusion is based on feedblock design, die design, and material selection to generate a new product with improved properties and performance.

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Review Questions

1. What is coextrusion?
2. What is the purpose of a tie layer? Explain how a tie layer works.
3. How many extruders are required to produce each of the following coextruded structures, and how many layers are in each structure: AB, ABC, ABCBA, ABCBDBA, ABA, and AEBC?
4. Of the structures in question 3, how many are balanced? What is the advantage of producing a balanced structure?
5. Where in an extrusion process are the various molten layers combined to form a multilayer structure?
6. What is a barrier layer, and what properties are important in differentiating barrier layers?
7. How do relative humidity and temperature affect properties in some barrier resins?
8. Which of the following resins, when combined in the melt state, require a tie layer: PC/HDPE, HDPE/EVA., EVA/nylon, LDPE/EVOH, PP/EVOH, nylon/LDPE, LLDPE/LDPE, PET/HDPE, PET/nylon, and PVC/HDPE?
9. What extruder configuration is required to make an ABCBA structure in cast film versus blown film?
10. Why is coextrusion used in pipe and tubing applications?
11. What does a striper extruder do in either pipe or wire coating applications?
12. Why is coextrusion used in large-part blow molding?
13. What are some coextrusion applications used in profile extrusion?
14. What is the difference between adhesive and cohesive tie layer failure, and how is it detected?

41 Feedblocks and Dies

There are three methods to combine melt streams from the individual extruders:

- Feedblocks connected to a single manifold die
- Multiple manifold dies with no feedblocks
- Feedblocks combined with multiple manifold dies

This chapter focuses on the different methods used to combine the melt streams from the various extruders to produce a coextruded structure. Feedblocks are used to combine the melt streams into a sandwich or layered structure prior to the die. In the die the individual layers are spread out to the full die width. Multimanifold dies combine the individual extrudate streams together after they have taken the die shape. In a sheet or film die, individual layers are combined after they enter the die and the melt is distributed across the entire die surface. In a blown film or pipe die, the melt is distributed around the circumference of the die before the individual layers are combined. To produce flat sheet or cast film with seven layers, feedblocks are combined with multiple manifold dies to introduce all the resins into the desired geometry.

41.1 Feedblocks

Feedblocks are junctions where the different melt streams are brought together prior to the die. The die is then attached to the feedblock to produce the coextruded structure. The advantage of a feedblock is it allows the different melt streams to be brought together in a location before the die, allowing the die to be simpler and less expensive. Feedblocks are used principally in flat sheet and cast film processes.

Polymer melt flow is laminar, meaning once the different layers are combined they do not mix. Each layer maintains its individual identity through the rest of the process. Pure laminar flow is maintained if the melt viscosities are similar and no secondary effects are present. A melt viscosity mismatch produces secondary effects (concepts are discussed in Chapter 42), which can cause one layer to encapsulate another layer or can produce unusual flow patterns. For purposes of this chapter, viscosity and secondary effects are ignored with the polymer melt having laminar flow.

A feedblock is made with openings for each layer in a coextruded structure. Consider the simulated feedblock shown in Fig. 41.1; there are three openings in which up to three different polymer melts can be introduced. The number of extruders and different melt configurations depend on the application and the coextrusion objective. The feedblock shown in Fig. 41.1 is connected to three different extruders to produce an A/B/C/B/A structure. The melt stream from extruder A is split in two as it enters the feedblock, with half of the material flowing to the top slot and the other half to the bottom slot. Likewise, polymer melt B from another extruder is split to form an upper and a lower layer in the structure. Material C comes from the third extruder and may be fed into the feedblock where it flows into the center channel. Melts from extruders A and B enter the feedblock from opposite sides where they are each split into their appropriate channels. The extruder size required for each feed stream depends on the relative concentration or layer thickness of each component in the final coextruded product. If B is an adhesive or tie layer bonding resin A to C, its extruder may be smaller, as the tie layer is usually 5–10% of the total flow. Extruder A might be the largest extruder if the outside layers are the bulk of the structure, while the inside layer may be a relatively thin barrier resin.

The feedblock openings are designed relative to the individual layer thickness in the final structure. As an example, assume a five-layer 1.25 mm (0.050 in or 50 mils) film is extruded in the same A/B/C/B/A structure discussed above, with the layer percentages in the structure being 32/10/16/10/32; the relative dimensions of the feedblock openings are the same. In practice, a more generic

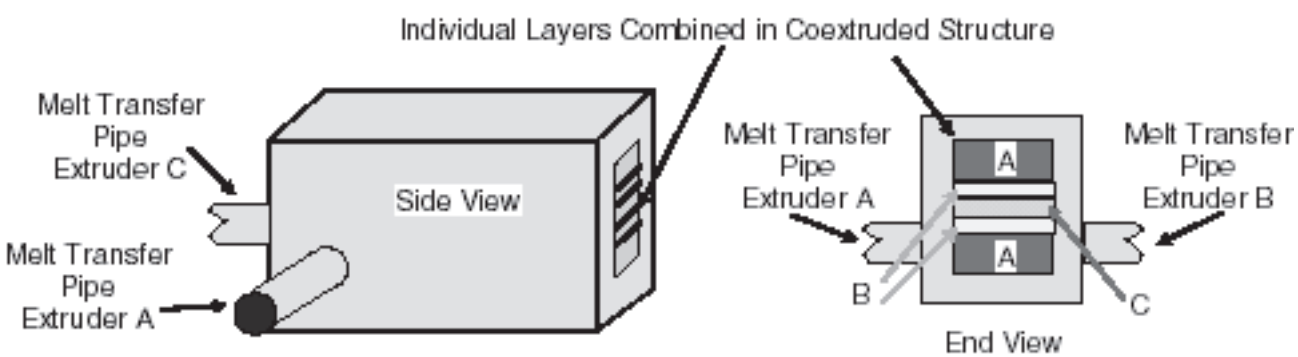

Figure 41.1. Schematic of a slotted feedblock.

feedblock might be used, with each layer thickness controlled partially by the extruder throughputs. At 500 pounds/hour throughput rate, extruder A has to deliver 320 pounds/hour, extruder B 100 pounds/hour, and extruder C 80 pounds/hour. A general feedblock geometry for this structure might have relative openings of 30/10/20/10/30 for the different resin slots. Feedblock opening thickness is adjustable using different techniques, depending on the feedblock manufacturer. Different inserts can be placed in standard feedblock openings to change the individual openings for each specific resin layer.

With the five-layer feedblock shown in Fig. 41.1, a number of different coextruded structures can be produced. The structure discussed so far is a three-component A/B/C/B/A structure. Other options include a five-layer structure A/B/C/D/E made with five extruders, unbalanced structure A/B/C/D/A using four extruders, and A/B/A/B/A made with two extruders. Using different inserts in the feedblock, layer thicknesses can vary between structures. Smaller feedblocks may be used for three layers to produce an A/B/C structure or an A/B/A structure. Larger feedblocks can produce structures with seven or more layers. A seven-layer feedblock can be used to produce an A/B/C/D/C/B/A symmetric structure or numerous asymmetrical structures, such as A/B/C/B/D/B/A where B is a tie layer bonding the various materials together.

While the feedblock shown in Fig. 41.1 has fixed rectangular channels, some feedblocks are made with circular melt pipes feeding a variable geometry feedblock that allows the individual opening sizes to be adjusted to accommodate different layer thicknesses. Figure 41.2 shows a variable geometry five-layer feedblock produced by Cloeren Inc.[1] The feedblock geometry can be adjusted automatically without changing slot widths or heights by using the adjustable vanes to change the melt velocity flowing from the manifold. Adjustable feedblocks are used to better control thickness uniformity and flow stability in the final coextruded structure. The feedblock, shown horizontally in Fig. 41.2, is configured to feed a single manifold sheet die. If the feedblock was rotated 90˚ it would be in a position to feed a cast film die. Melt channel A comes from the primary extruder feeding the end of the feedblock, while melt channels B and C come from two satellite extruders. Melt channels B and C could each be supplied by two separate extruders so the structure produced would contain five different layers. In the configuration labeled in Fig. 41.2, the coextruded structure produced is C/B/A/B/C.

Extrusion Dies Inc.[2][3] uses a "Selector Spool" technology that provides extrudate distribution to different feedblock locations. By changing the selector spool configuration, the melt stream can be moved to a different location within the feedblock to produce a different coextruded structure. This allows the coextruded layer sequencing to be established upstream of the actual feedblock. The spool technology is designed for quick structural changes, allowing the coextruded melt configuration to be easily converted from one configuration to another without necessitating a feedblock change. In addition, both the height and the width of the feedblock channels can be adjusted to produce the desired layer ratio in the coextruded structure exiting the feedblock. This technology is designed to eliminate any dead spaces in the feedblock that might be associated with other adjustable feedblock technology utilizing pins, spools, or plates.

Experimental work was done at Dow Chemical Company[4] to evaluate the effects of melt transfer line shape (rectangular, square, or circular) and feedblock slot width. Five different feedblock slot widths were evaluated with each melt transfer line configuration. The feedblock geometry was decreased from the entrance to the feedblock exit by a fixed amount. All five configurations had the same width opening of 1 inch (25.4 mm), with the exit being 35% (0.35 inch or 8.9 mm), 50% (0.5 inch or 12.7 mm), 67% (0.67 inch or 15.9 mm), 80% (0.80 inch or 20.3 mm), or 100 % (1 inch or 25.4 mm) of the initial opening. The effect on the layer thickness profile in the final coextruded structure was evaluated based on these feedblock process changes. Results showed the following:

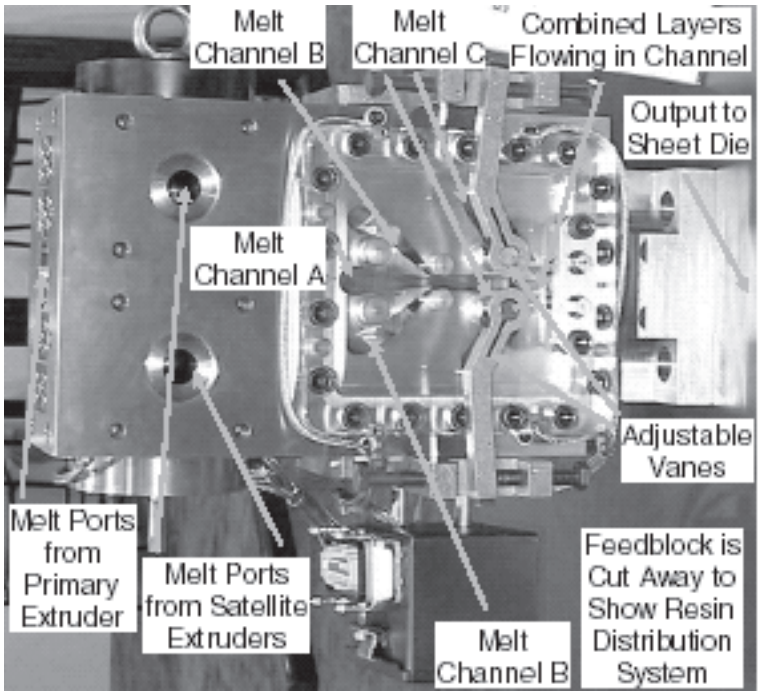

Figure 41.2. Cutaway view of five-layer variable geometry Cloeren Incorporated feedblock.[1]

- Effects of feed slot shape and melt transfer line shape are the same for both polystyrene and polycarbonate (only resins tested).
- Transfer line shape had no effect on the final layer thickness in the final profile.
- Slot width is an effective method of changing the layer thickness profile in the final structure. The largest change in the thickness profile is observed in the straight channel slot going from 100% width to 67% width. In other words, the slot is only 67% of the final width of the coextruded structure.

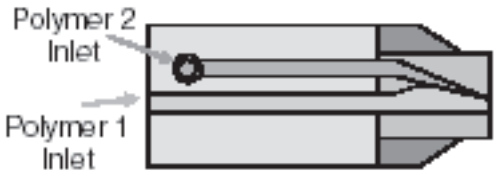

Figure 41.3. Schematic of two-layer coextrusion die.

41.2 Multimanifold Dies

A second method of combining melt streams from different extruders to form coextruded structures is with multimanifold dies. At the back or side of the die, ports exist to introduce the melt from the primary extruder and all the satellite extruders. Manifolds inside the die distribute the melt to the desired shape and location in the structure, where it is combined with the other melt streams to form the coextruded structure.

Figure 41.3 shows a two-layer die where the primary polymer enters the die back and the secondary or cap layer is introduced into the side. The resin distribution within the die is through a coat hanger, T-slot, or fishtail design for each polymer. After the resin is evenly distributed across the die, the melt streams are brought together near the die exit. Figure 41.4 shows a two-layer coextrusion die from Cloeren Inc.[1] to produce two-layer vinyl siding. The cap layer can be a different color from the substrate, or, with a different resin system, the cap layer could be a weatherable cap. The manifold for the primary extrudate is shown entering the back of the die, while the secondary or cap layer from the satellite extruder is shown entering from the side. Both the top and bottom halves are T-slot sheet dies with a deep manifold channel across the back of the die to uniformly distribute the melt. After the melt streams are spread across the die, the two layers are combined just prior to the die exit.

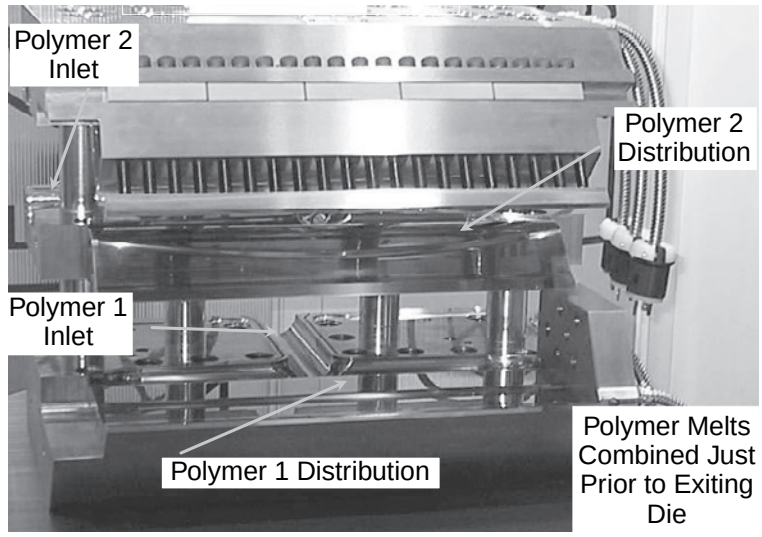

Figure 41.4. Cloeren two-layer vinyl siding die.[1]

Modifications of Fig. 41.3 allow the introduction of more than two melt streams into a die. It is common to connect three extruders to a die with multiple manifolds, as shown in Fig. 41.5. The streams from three extruders can be split to make a five-layer structure. Or in the example shown in Fig. 41.5, the two melt streams A and B could have come from one extruder to produce a B/C/B structure.

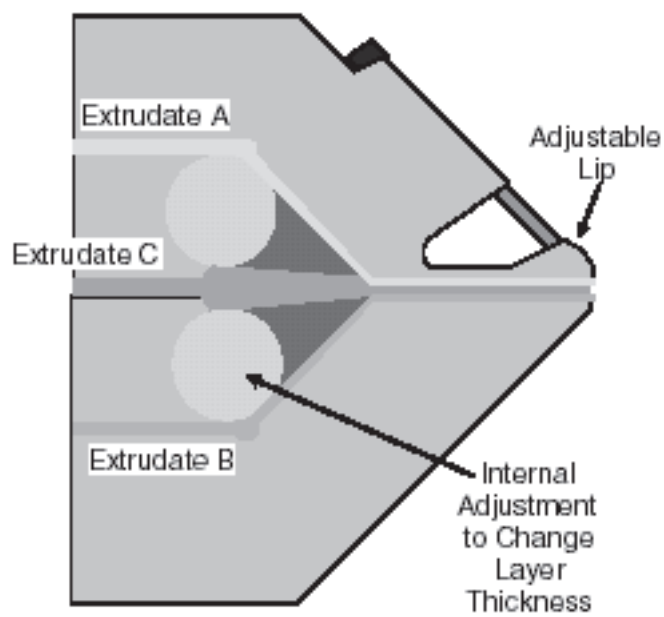

Figure 41.5. Multimanifold die.

Design work continues on manifold and preland areas to obtain the optimum layer distribution with a coextruded structure. The teardrop manifold was replaced with a rectangular manifold to improve the distribution. Recent studies[5] have suggested changing to a modified rectangular manifold with a circular entrance may provide a better layer distribution with some polymers. Die and feedblock design is as critical as selection of the optimum processing conditions and polymers to provide uniform layer distribution and thicknesses in the final coextruded structure.

Coextrusion blown film dies do not use feedblocks. The extruders are arranged in a circular pattern about the die, as shown in Figs. 40.5 and 41.6. Coextrusion die technology has continued to evolve in blown film with capabilities to make coextruded structures with 2 to 11 layers. A spider-type coextrusion blown film die is shown in Fig. 41.7. Spider dies are used very seldom in coextruded blown film as they have been replaced with spiral mandrel dies to eliminate weld lines and more recently by stacked flat plate dies to add versatility.[6] The spider die in Fig. 41.7 shows how the different resins are combined in a coextruded blown film die. Fig. 41.7 shows a five-layer structure A/B/C/B/A made from three extruders. In the structure, resin A, introduced near the bottom of the die, has significantly more contact with the die metal surface as it flows out of the die than the other resins in the structure. This can affect the melt stability and resin distribution in the various layers. Consider the introduction of the second resin from the bottom; this is in contact with the mandrel for only a short time before it is totally encapsulated by resin A in the outer layer and resin C in an inner film layer. Likewise, as resin C is introduced into the coextruded structure, it is in contact with the die for only a short time before it is encapsulated by resin B on both sides. Resin A on the outside surface of the structure is in contact with the stationary die body the entire time. The encapsulated resins are moving with the other layers. Instabilities can be generated as a result of the difference in melt environment of the outer layers versus the inner layers in the structure.

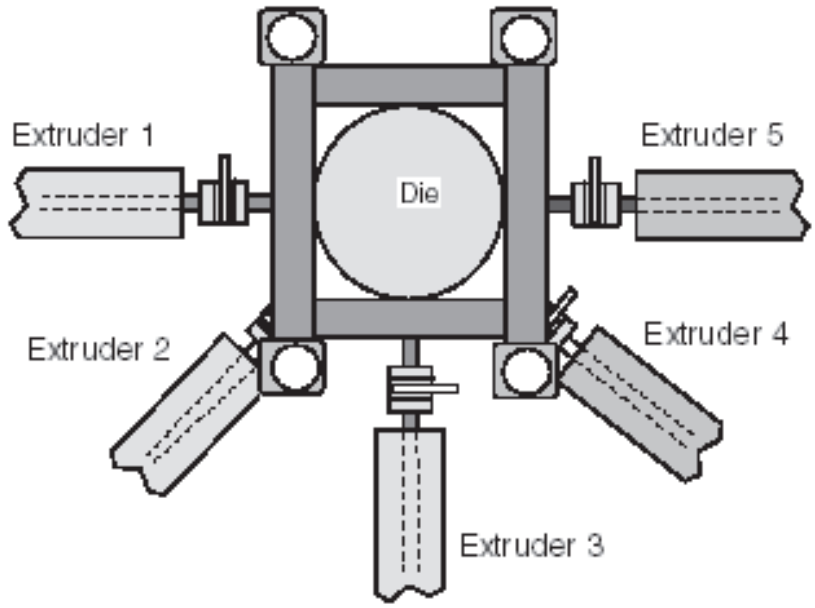

Figure 41.6. Five-extruder arrangement for coextruded blown film.

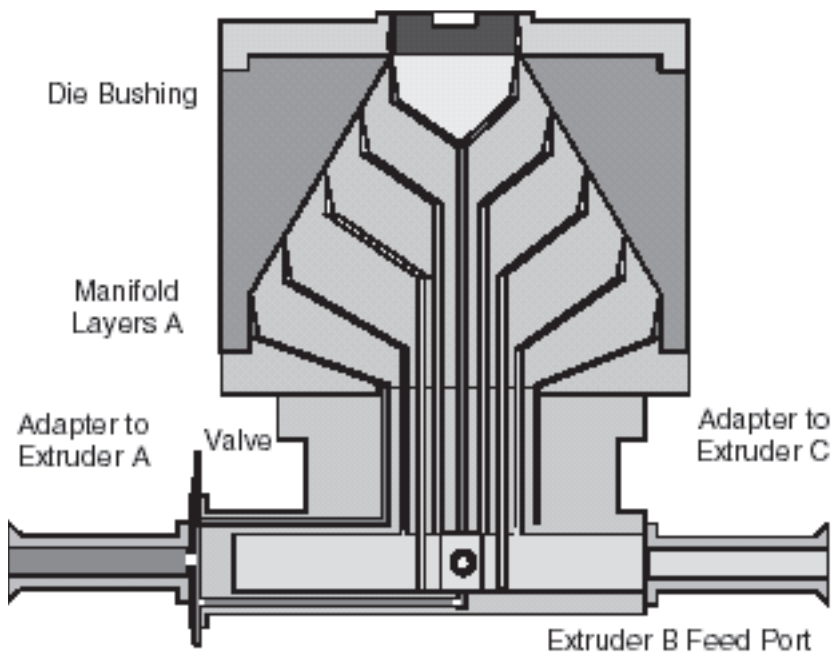

Figure 41.7. Spider mandrel coextrusion die.

Figure 41.8 shows a spiral mandrel blown film coextrusion die. The spiral melt channels are center fed through the nested spirals. This eliminates problems associated with the spider legs disrupting the melt stream, causing weld lines and flow interruptions in the coextruded layers in the mandrel die shown in Fig. 41.7. However, there are still shortcomings associated with a spiral mandrel die. The die flexibility is extremely limited; in the example shown in Fig. 41.8, it is difficult to add another

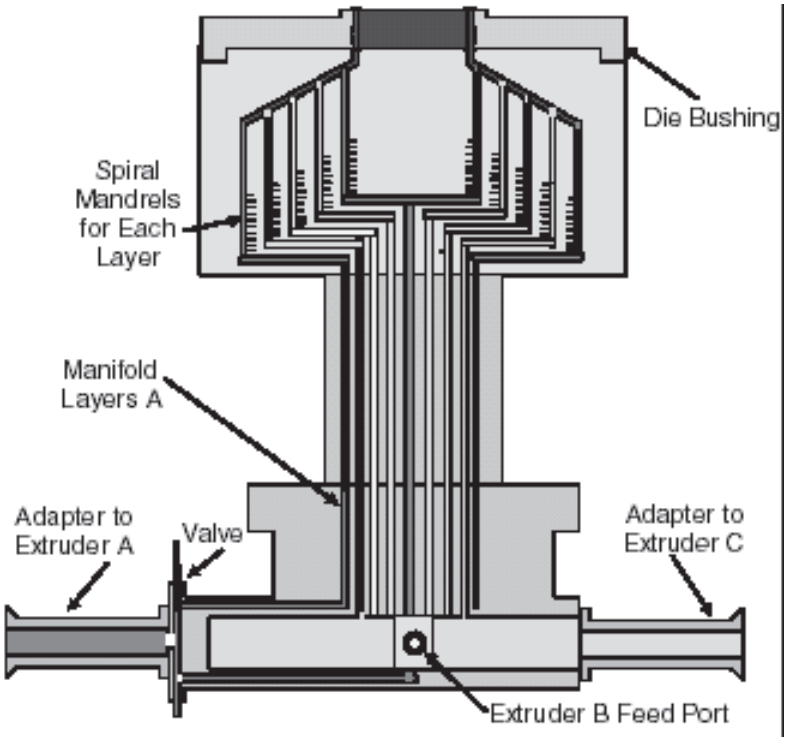

Figure 41.8. Spiral mandrel coextrusion blown film die.

layer or extruder to the existing die. Dies designed for five layers cannot be changed to seven layers without major rework. In addition, the layer thicknesses in the die cannot be easily changed. Assume the layer thicknesses in the die shown in Fig. 41.8 were 25/20/10/20/25; the layer ratio cannot be changed to 40/5/10/5/40. Once the die is fabricated, it has to be used for the coextruded structure for which it was designed or the die has to be disassembled and reassembled with different spiral mandrels. A second problem is heating of the inner mandrel compared to the outer mandrel. Heat is applied to the outside of the die body; to heat the innermost mandrel the heat has to radiate or diffuse through the other mandrels and the insulative polymer films to reach the center most nested spiral. This can be particularly troublesome if greatly different melt temperature polymers are used that have to be at the same temperature to match polymer viscosities. A third problem is the die residence time. The outer spiral diameter is much larger than the inner spiral; consequently, the polymer melt has to travel farther to exit the die through the outside layer compared to the inside layer. Longer time at temperature can lead to resin degradation and, over time, black specks. The fourth problem is the long time required to purge the outer spirals compared to the inner spirals during product changeover or shutdown.

Due to the problems associated with nested spiral mandrel coextrusion dies, they have been replaced in many blown film coextrusion applications with the flat plate stacked dies shown in Fig. 41.9. The stacked die is designed to feed one layer from each extruder to produce a five-layer structure of A/B/C/D/E. The flat plates all have the same diameter, allowing them to be stacked one upon the other. Polymer is fed into the side of each plate, where it is evenly distributed around the center. Spiral channels are incorporated in the stacked dies to provide better mixing, thermal uniformity, and polymer distribution. The spiral channels are machined in the flat plates to deliver polymer melt at uniform flow rates around the die. The spiral channels can be cut in the bottom plate, top plate, or split into both plates. Serpentine channel design is touted to reduce the large residence time that can occur in the spiral-type channels, caused by the difference in melt flow length in the bottom of the channel versus that flowing across the barrier at the top of the channel.[6] Some stacked dies are bottom fed rather than side fed from the extruder. Proper manifold design between the extruder and die allows a given extruder to feed more than one layer in a stacked die. Consequently, the die shown in Fig. 41.9 could be fed with three extruders to produce an A/B/C/B/A structure. The advantage of the modular die structure is that it provides flexibility for easy layer addition or subtraction to produce different coextruded structures from the same die.

A new addition to the flat plate die is the incorporation of a dual spiral system that allows the polymer

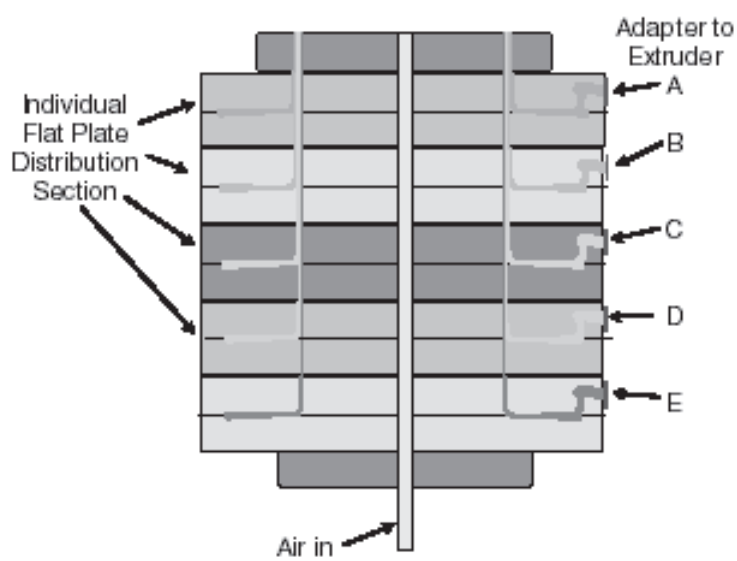

Figure 41.9. Side fed stacked flat plate blown film coextrusion die.

stream to be split to form two individual layers.[7] Each layer is split into two separate layers, allowing the incorporation of more layers in the structure without the addition of more extruders. Instead of including one thick layer of an expensive material such as EVOH, the layer can be divided into two thinner layers, improving the barrier properties at reduced polymer cost.

The use of stacked flat plate dies with spiral distribution has expanded the capabilities of coextruded blown film structures. Dies commonly used can produce 7, 9, and 10 layers in blown film. This increases the functionality of the film and allows market penetration into new areas that were previously not possible. Blown film coextrusion technology has been expanded to include multilayer coextruded structures with double and triple bubbles to produce biaxially oriented PP film and sausage tubes, respectively.[8]

Coextruded pipe and tubing dies are used for applications such as

- Water pipes where an oxygen barrier, thermal stability, and durability are required
- Medical tubing requiring barrier properties, resistance, and FDA compliance
- Nylon tubing for fuel lines requiring UV stability, barrier properties for hydrocarbons, and antistats
- Cosmetic tubing used in packaging that requires barrier properties, printable surface, and smooth inner surface
- Pipe with foam cores
- Multiple layer garden hoses
- Corrugated pipe where multiple colors are required

For multilayer tubing and pipe, much of the die technology has been expanded from the blown film industry. Figure 41.10 shows a typical five-layer, multiple plate die without the spider legs commonly found in mandrel pipe and tubing dies, which generate weld lines. The coextruded pipe structure produced with the die geometry shown in Fig. 41.10 is an A/B/C/B/A structure.[9] This die could be changed to an A/B/C/D/E structure if the melt streams from the extruders are not split and two other extruders are available. The die shown is similar to the stacked die shown in Fig. 41.9 for coextruded blown film except it has been turned horizontally and is attached directly **attached directly or via a transfer melt pipe to the end of extruder A**. Other extrudate is fed through the sides of the individual die segments. Similar to the stack die, this die configuration can be easily changed to add or reduce layers. However, unlike a blown film die, which can produce a variety of laydown widths and bubbles from a single die, the pipe or tubing die is used to produce essentially one pipe or tubing size or diameter. Modifying the die bushing opening and the pin at the end of the die provides capability for some adjustment to the inside and outside diameter and wall thicknesses.

Die design must accommodate the following requirements to produce a five-layer structure using five different extruders:

- Melt is sequentially combined in the desired order to produce the correct structure.
- There has to be sufficient room around the die to accommodate the extruders. Some extruders may be larger than others, depending on the relative throughput rates to provide the final structure layer thicknesses. If some extruders are relatively small, space may be available to suspend them from above the die.
- There must be no stagnant areas where polymer can build up and degrade over time.
- Uniform, undisturbed flow through the die is required to minimize instabilities.
- There must be uniform pressure across the die exit.

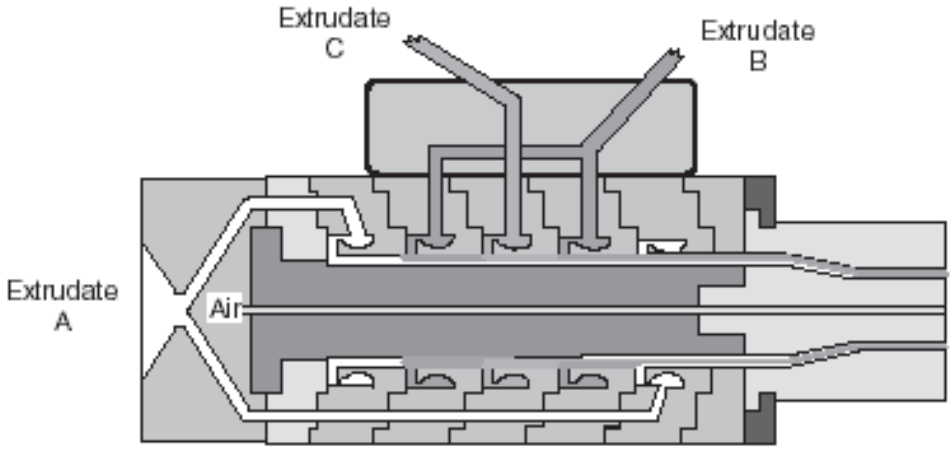

Figure 41.10. Pipe or tubing coextrusion die.[9]

Special die heads are used to stripe pipe. Melt is introduced into a spider mandrel die or a stack die in one location. The melt is not uniformly distributed around the circumference of the mandrel. It is distributed to the die locations where the secondary layer is required. As an example, the secondary or stripe layer may come in contact with the primary resin in one or more individual streams around the die wherever the stripe is required.

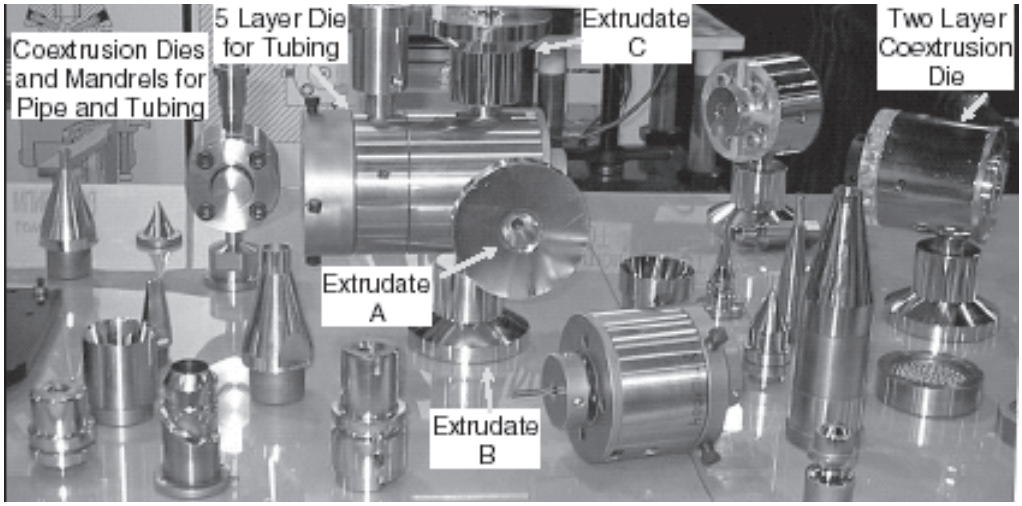

Figure 41.11. Coextrusion dies for pipe and tubing by Guill Tool & Engineering.[10]

In some applications, a second layer is applied to the pipe surface using a secondary extrusion process downstream. As an example, a smooth surface applied to a corrugated pipe is accomplished after the corrugation step and the pipe is cool. A secondary extruder with a crosshead die applies the extrudate to the corrugated pipe surface in a continuous operation as the corrugation passes through the die. This produces a multilayered structure and is done in a two-step operation.

Coextrusion dies used in wire and cable are similar to those used in pipe and tubing.[10] Striper dies similar to those in the pipe industry are used to apply continuous stripes to wire for identification purposes in end-use applications. Secondary operations applying wire coating to bundles of previously coated wires is quite common. This is not coextrusion but a continuous process with multiple crosshead dies in the line. Other coextrusion applications in the wire and coating industry are used to apply an insulation layer, followed by a foam core and a protective cover.

Coextrusion is used in large-part blow molding to provide barrier layers and adhesive layers between noncompatible resins in parts such as automotive gas tanks. Another application is to extrude two materials in the same part, such as a hard and a soft segment to produce an HVAC pipe for automotive use, Fig. 41.11. Multiple extruders and/or accumulators feed melt to the blow molding head, where the parison is dropped prior to clamping and blowing to fill the cavity. Figure 41.12 shows one method of producing multilayer structures in coextrusion. Coextrusion blow molding does not use a feedblock; instead, the die head has to be properly designed to uniformly distribute the different polymeric materials in the correct ratio around the flow channel as the parison is forced out of the die head. Some of the same technology used in coextrusion blown film is used in large-part blow molding except the die openings are larger, the die is inverted, and the melt is supplied by an accumulator on demand. The parison is programmed to provide extra thickness where the draw is the greatest in the part. In sequential blow molding, the equipment is programmed to add one material then a second before returning to the original resin.

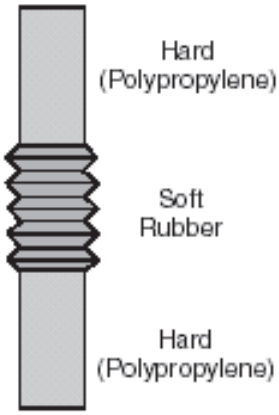

Figure 41.12. Sequential coextrusion of two different materials in one part.

In profile extrusion, a second extruder may be used to add an elastomeric bead to a specific area in the profile, as shown in Fig. 40.10. This is applied with a small extruder, similar to those used to apply stripes to pipes. The extrudate from the satellite extruder is added only in one part of the die, near the exit of a large die. Both resins have to be compatible with the correct melt viscosity to establish a strong bond while producing a good appearance part.

41.3 Feedblock Combined with a Manifold Die

The last process for producing coextruded structures is to combine a manifold die and feedblock to add additional melt streams and layers to the final structure. Figure 41.13 shows a five-layer feedblock combined with a three-layer manifold die to produce a seven-layer structure. This structure requires seven extruders. Splitting various layers to produce a balanced structure (A/B/C/D/C/B/A) requires only four extruders.

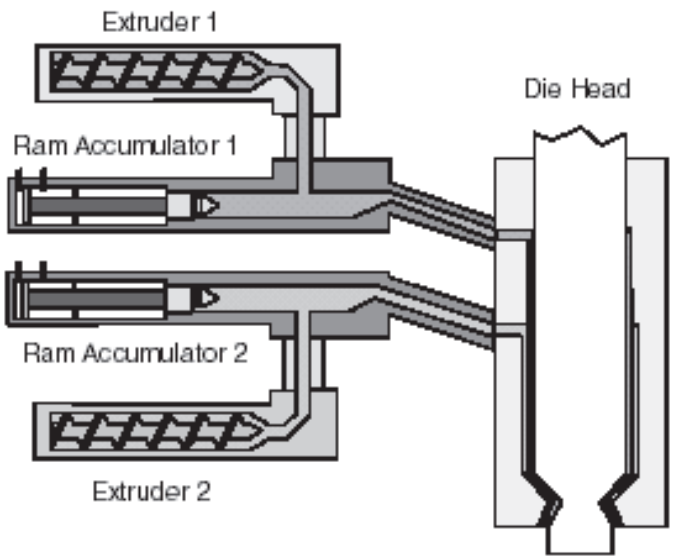

Figure 41.13. Large-part blow molding coextrusion.

Multimanifold dies combined with feedblocks are used in sheet, cast film, and pipe coextrusion. Pipe coextrusion with a feedblock is shown in Fig. 41.10, where the one stream in the center is split in the feedblock and fed to separate ports in the die.

There are many methods of combining melt streams using dies and feedblocks in all types of extrusion processes to add value to the product through enhancement of physical properties, functionality, durability, barrier properties, and chemical properties. Combining the proper equipment, resin layer orientation, and proper materials selection, coextruded products add value to many products. The next chapter deals with proper material selection and extrusion conditions to produce acceptable structures.

41.4 Computer-Aided Design

Computer software to assist the coextrusion die designer, based on finite element analysis, is available from a number of sources, such as Compuplast International Inc.[11] and PolyDynamics Inc.[12] The computer programs design the die and simulate the process based on information such as material types, viscosity, layer thickness, coextruded structure, melt temperature, and extru-

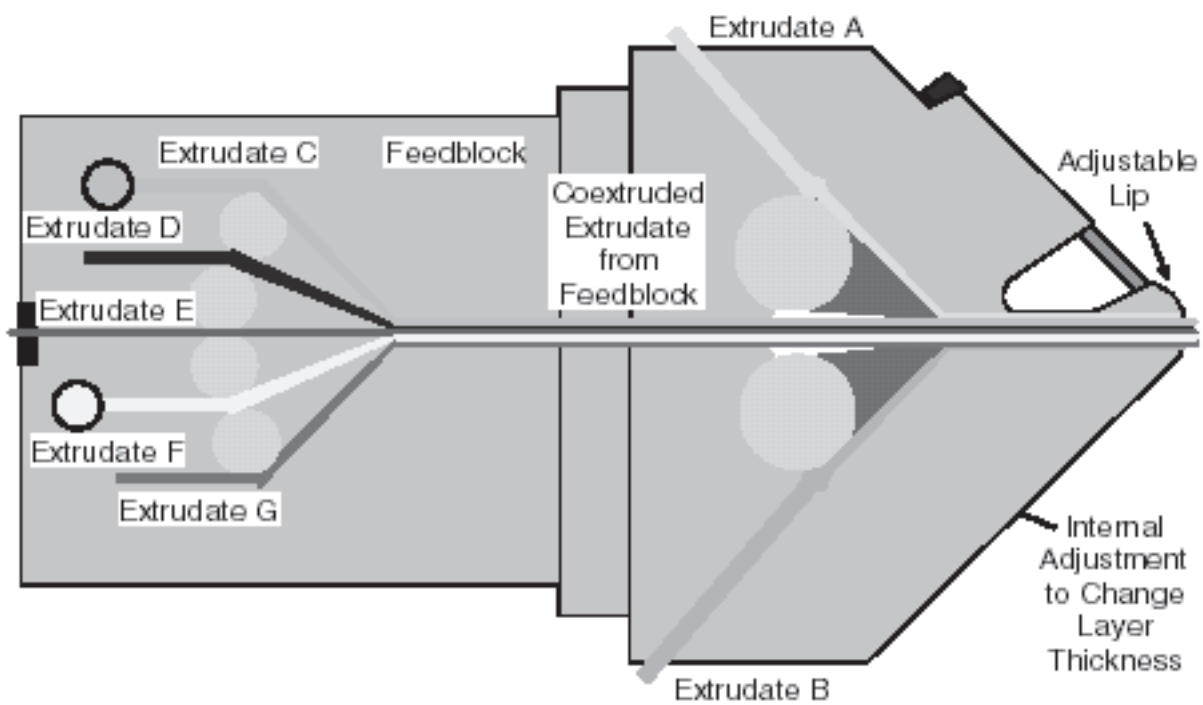

Figure 41.14. Seven-layer structure combining feedblock technology with a multiple manifold die.

date shape. In addition to assisting the die design geometry, the software is used to simulate flow in the die, measuring pressure drops, polymer velocity, and shear stress, which can be used to infer the presence of layer-to-layer interfacial instability in the coextruded structure. Flat sheet or film dies to annular-type dies used in blown film, pipe, tubing, and blow molding can be modeled using this type of software prior to the actual cutting of steel.

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Review Questions

1. What are three equipment techniques used to produce coextruded structures?
2. How does a feedblock function to produce coextruded structures?
3. What types of extrusion processes lend themselves to coextrusion?
4. What is the difference between a nested spiral mandrel and a stacked flat plate blown film die?
5. How does a striper die work?
6. What is the difference between a multimanifold die and a feedblock?
7. What is the purpose of combining feedblocks and multimanifold dies in the same process?
8. How does a tubing or pipe coextrusion die function, and what other type of die is it modeled after?
9. What is the problem with a nested spiral mandrel die for coextrusion blown film?
10. What is the difference between an adjustable and a fixed dimension feedblock?
11. How is large-part coextrusion blow molding accomplished?

Review Questions (continued)

12. What type of equipment might be selected to make the following cast film structures? How should the extruders be arranged?
    a. A/B/C/B/A
    b. A/B/A
    c. A/B/C/D/B/A
    d. A/B/C/D/C/B/A
13. How many flat plate stack dies need to be combined to make 8- and 10-layer blown films? Explain.
14. What is the purpose of producing multilayer pipe and tubing?
15. What might be the advantage of producing A/B/A sheet?
16. How are sequential blow molded parts produced?
17. What is the difference between side-by-side coextrusion and layered coextrusion?

42 Polymer Selection for Coextrusion

To produce an acceptable coextruded structure, the proper resin systems have to be used with the correct die and/or feedblock technology. Just having the proper extruder, feedblock, and die design will not guarantee successful coextrusion. Assume an elastomeric edge is to be applied to a rigid vinyl profile. The rigid PVC grade being used is probably fixed, based on other formulations being extruded in other profiles in the plant. The elastomeric edge is a flexible PVC that will bond directly to the rigid PVC without a tie layer. What are the resin requirements, or can any flexible PVC grade with the correct color and durometer be added to the satellite extruder to produce the desired coextruded profile? Proper resin selection for this or any other coextrusion is dependent on matching resin melt viscosities at the processing melt temperature. Once the coextruded structure is defined and the equipment is in place, proper resin selection is critical to produce quality parts with the proper adhesion, clarity, layer thickness, and interfacial stability to produce a commercially viable product with the desired performance properties. This chapter focuses on the resin requirements and processing conditions required to obtain quality coextruded products.

42.1 Melt Viscosity Matching

One of the most critical steps in producing quality coextruded parts is matching the layer melt viscosities as they are combined in the die or feedblock. In principle, polymer melt flow is laminar, meaning it flows in discrete layers. As polymer layers are combined inside the die, the laminar flow characteristics keep each polymer flowing in its own discrete area, preventing the different polymer layers from mixing. If turbulent flow was present in either the die or feedblock, the polymers would mix and be distributed in a random orientation throughout the melt. If the polymers in the different layers all have the same viscosity when they are combined and have the same velocity after combination, the structure should be uniform exiting the die. Factors affecting laminar flow are viscosity mismatch, which can be caused by selecting the wrong polymers, poor melting in the extruder, nonuniform melt temperature as the polymers are combined, or different melt temperatures due to different shear conditions during melting and conveying in the extruders, melt pipes, and adapters.

Melt viscosity mismatch can result in encapsulation and interfacial instability. Encapsulation happens when a low-viscosity material flows around and encapsulates a high-viscosity material. Three different A/B/A coextruded structure are shown in Fig. 42.1, assuming different layer viscosities. The first schematic in Fig. 42.1 assumes polymer A and polymer B viscosities are equal, with all three layers well defined and distinct. In the second scenario, where polymer A viscosity is less than polymer B, the lower-viscosity material encapsulates polymer B. In the third scenario, where polymer A viscosity is greater than polymer B, polymer B flows out from the center of the structure and up to the top and the bottom of the structure, encapsulating polymer A. The polymer viscosity ratio determines the encapsulation level and the individual layer redistribution. The greater the viscosity ratio, the greater the rearrangement and encapsulation in the final structure. Another factor influencing this phenomenon is where the different layers are combined, affecting the residence time in the die. If the layers are combined in a feedblock and then go through the die, there is more time and longer distance for the individual polymer layers to rearrange than if they are combined near the die exit. After the structure exits the die and starts to cool, no more layer rearrangement occurs. The coextruded structure, shown in Fig. 42.1, is very simple and the viscosity effects easy to understand. Expand this structure to five different materials that include different base resin layers, tie layers, barrier resins, and recycled layers, and the viscosity issue becomes more complex.

If feedblock technology is being used to generate the coextruded structure and known viscosity mismatches are going to be present in the polymers being processed, modifications can be made to the feedblock slot geometry to compensate for the viscosity differences. Figure 42.2 shows two feedblock geometry corrections that can be made to help compensate for polymer viscosity differ-

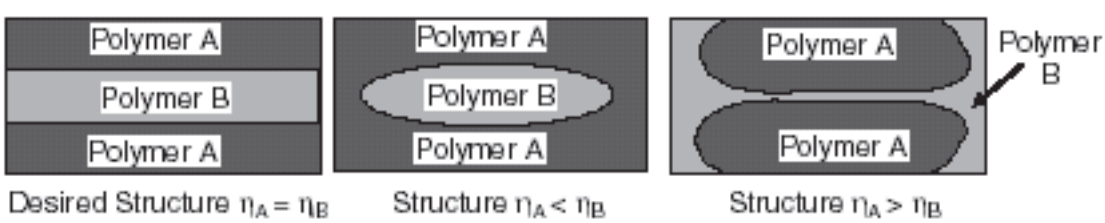

Figure 42.1. Effect of melt viscosity mismatch on coextruded structures.

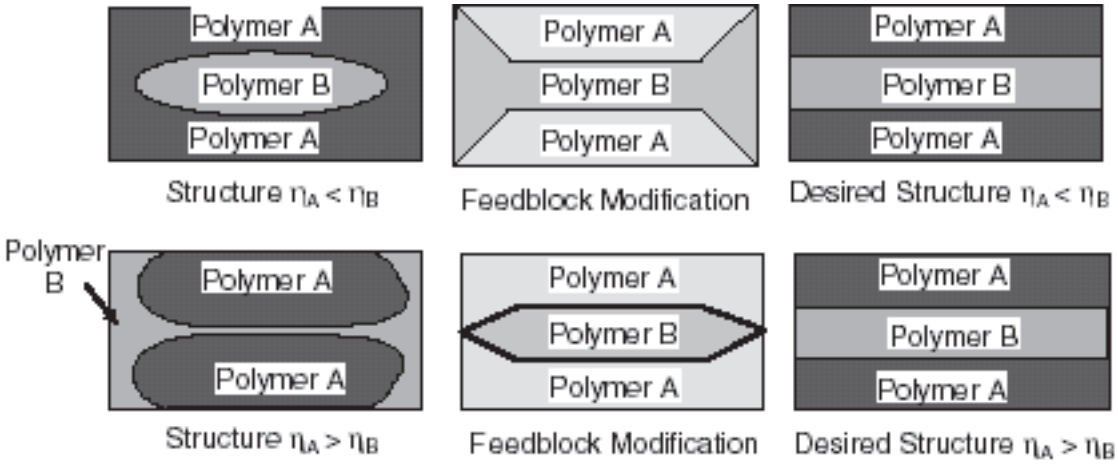

Figure 42.2. Feedblock modifications to produce flat layers with viscosity mismatches.

ences. Ideally it is best to use similar viscosity materials. While feedblock modification is a potential solution to the problem in wide sheet or film, it is likely the final product will exhibit layer thickness variation from edge to edge. Thickness variation, particularly with expensive polymers, can use more resin than necessary, which inflates the product cost and reduces profits.

In flat sheet or film extrusion where polymers are being brought together in the die, slight polymer viscosity mismatches can still produce good products. The examples shown in Fig. 42.1 have large viscosity ratios. If the polymers joined in the die have slight viscosity differences, insufficient residence time is available to totally encapsulate one polymer layer or the other, as shown in the second and third scenarios in Fig. 42.1. Instead minor layer rearrangement may occur. If layer rearrangement is only minor on the edge of the sheet or film and that part is being trimmed to produce the final product dimensions, the product may be totally acceptable. One of the biggest problems with layer rearrangement due to viscosity differences is the potential for nonuniformity in layer thickness across the coextruded structure.

Another problem related to viscosity is caused by nonuniform melt temperature across the polymer melt, which affects the polymer melt viscosity. It is common in melt transfer pipes between the end of the extruder and die or feedblock to have melt temperature gradients of 20–30˚F (11–17˚C). (The effect of melt temperature on viscosity is shown later in this chapter in Tables 42.1 and 42.2.) Figure 42.3 shows the potential effect of melt temperature variation on the individual layer uniformity. Melt temperature changes are accompanied by thickness variation across the web as thick and thin sections develop in an individual layer, depending on the resin viscosity. Thick areas in the machine direction are caused by the higher melt temperature and lower viscosity. This is similar to the previous discussion of encapsulation. Where the polymer flows better (higher melt temperature, lowers viscosity), that section of the individual layer is thicker than where the melt temperature is lower and the viscosity is higher. If polymer layer uniformity is an issue, melt temperature differences can be minimized through improved mixing in the extruder or using a static mixer in the melt transfer pipe.

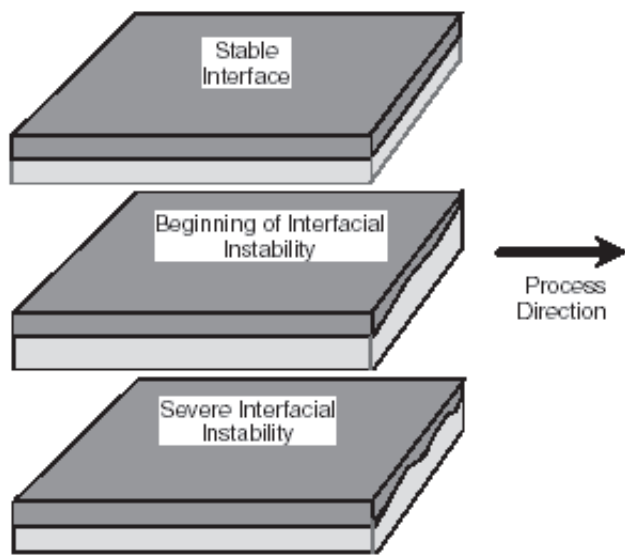

Figure 42.3. Interfacial instability due to melt temperature variations.

Layer uniformity differences in the machine direction are caused by nonuniform feeding of the melt to the die as a result of surging or other pressure variations in the extruder. Figure 42.4 shows layer variations due to pressure variations at the end of the line. As discussed in Part 4, Chapter 6, surging is caused by poor feed characteris-

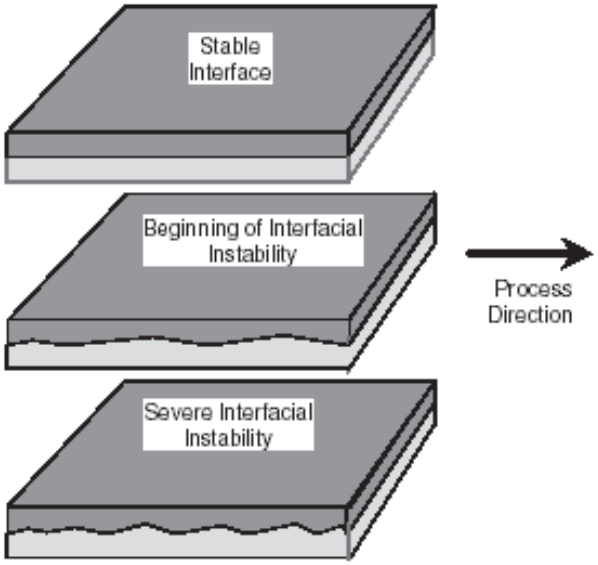

Figure 42.4. Variation in layer thickness due to pressure variations.

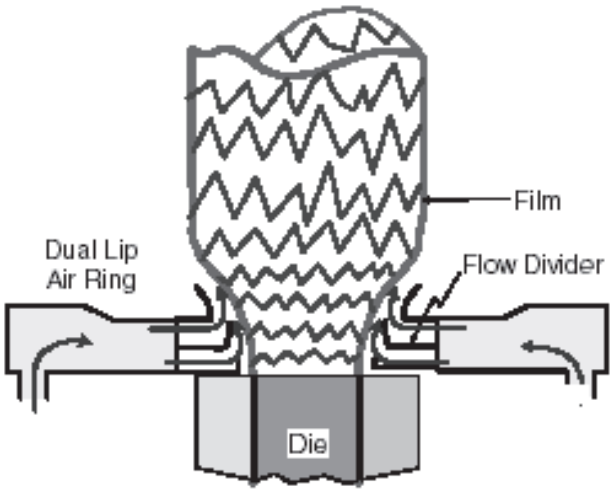

Figure 42.5. Zig-zag interfacial instability in coextruded blown film.

tics in the extruder feed section, bridging in the feed throat, too high of a feed throat temperature, improper feed throat water cooling, wrong or plugged screen pack, or the wrong screw design. One method of overcoming surging problems is to use a gear pump.

42.2 Interfacial Instability

In addition to viscosity mismatch effects, interfacial instability in multilayer systems can create appearance effects that render the product totally useless. Interfacial instability is caused by two or more layers coming together with different viscosities at different velocities, producing an unstable interface. Interfacial instability is typically classified into one of three types:

- Zigzag pattern
- Scattering
- Wave pattern

The patterns appear across the entire structure.

In addition to poor surface quality, the product surface may be very rough, making it totally unacceptable. Zigzag instabilities appear as a group of V-shaped patterns across the product, as shown on coextruded blown film in Fig. 42.5. The zigzag pattern is thought to initiate in the die land area as a result of polymer melt instability at the die exit, similar to melt fracture discussed in Chapter 4. A critical shear stress developed at the interface between two layers is thought to cause interfacial instability. Above a certain critical shear stress value, interfacial instability manifests itself in the zigzag pattern. Shear stress is defined as the tangential force on the melt divided by the area over which it is applied. Equation (42.1) relates the shear stress to the viscosity and shear rate.

$$\tau = \eta \times \dot{\gamma} \tag{42.1}$$

where

τ = Shear stress (in psi or MPa)
η = Viscosity
$\dot{\gamma}$ = Shear rate

Ways to eliminate zigzag instability are

- Match resin viscosities.
- Reduce the flow rate.
- Change the layer ratios.
- Increase melt temperature (depends on effect temperature has on viscosity).
- Change die geometry.

Relative shear stress is shown in Fig. 42.6, where the highest level is at the die-polymer interface. Intermediate shear stress occurs between polymer resins A and B. This is the location where interfacial instability occurs when the critical shear stress is exceeded. Figure 42.7 shows the development of interfacial instability in a coextruded sheet.[1] From Eq. (42.1), the shear stress is directly related to the shear rate. Polymer flowing through a narrow gap, such as a die or manifold, shears thin due to the increased shear rate. If the flow through a small area (feedblock or die) is increased, the shear rate also increases, causing the polymer to shear thin.

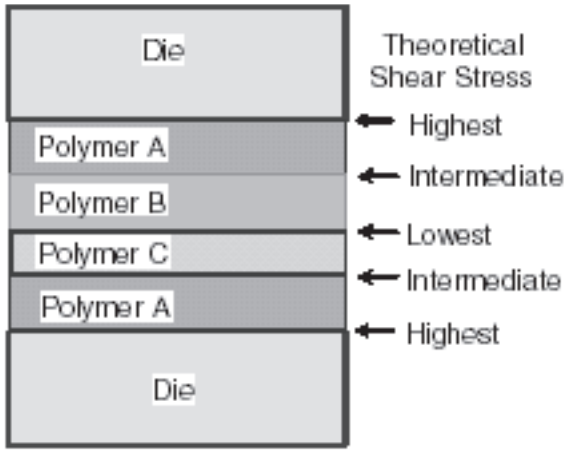

Figure 42.6. Relative shear stress levels.

Coextruded polyethylene blown film studies have shown a difference in interfacial instability based on molecular weight distribution. A broad molecular weight distribution polyethylene resin is more likely to produce a zigzag interfacial instability than a narrower molecular weight distribution resin. With the broad molecular weight distribution, the only way to eliminate the interfacial instability is to increase the thin layer thickness to which it is in contact or change the material. With the narrow molecular weight distribution polyethylene resin, interfacial instability can be eliminated by increasing the temperature or increasing the die gap.[2]

Viscosity matching to eliminate encapsulation and zigzag instability is accomplished by analyzing shear rate versus viscosity data at three different temperatures. Viscosity at relatively low shear rate data are required because the shear rates for the resins exiting a feedblock or in the die are relatively low. Viscosity versus shear rate data for coextrusion are measured using oscillating plate or capillary rheometry. If capillary rheometry is used, viscosity data have to be collected in the low shear rate range. The shear rate in a rectangular die channel is given by Eq. (42.2):

$$\textit{Shear Stress } (\dot{\gamma}) = \frac{6Q}{W \times h^2} \tag{42.2}$$

where

W = Channel width
H = Channel height
Q = Volumetric flow rate given by Eq. (42.3)

$$Q\,(\dot{\gamma}) = \frac{W \times H^3 \times \Delta P}{12 \times L \times \eta} \tag{42.3}$$

where

ΔP = Pressure drop across the channel
L = Channel length
η = Polymer viscosity

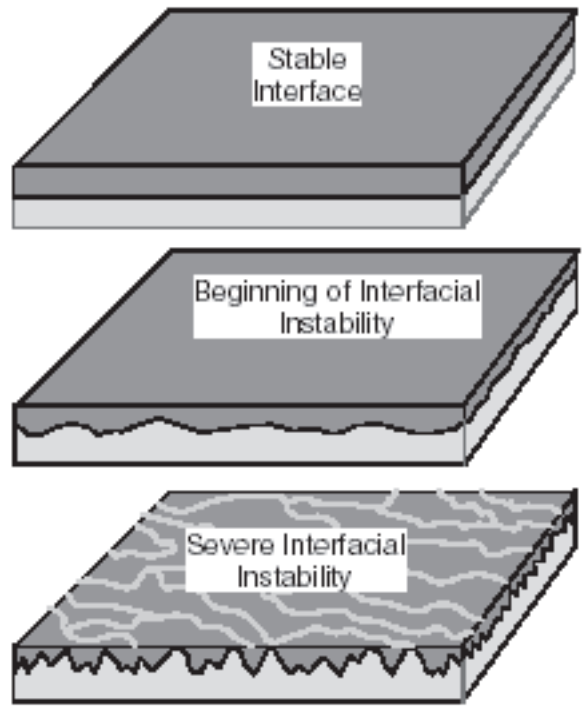

Figure 42.7. Degrees of interfacial instability.

Calculations of shear rate from these equations will give 20–100 sec^{-1} in the die lip area. At these shear rates, the viscosities of most polymers do not change drastically. Therefore, matching resin viscosities at 20 or 50 sec^{-1} will provide reasonably good results.

If viscosity data are available at three temperatures, the log of the viscosity data at a given shear rate is plotted versus $1/T$, where T is temperature in Kelvin. This is called an Arrhenius plot and usually gives a straight line. Similar plots for other polymers in the structure are used to determine the temperature where the melt viscosities are equivalent. This then becomes the melt temperature where the polymers are brought together with equivalent viscosities. As an example of this technique, assume a coextruded polycarbonate/tie layer/polypropylene structure is to be produced. What is the appropriate melt temperature to run the process, and what melt flow resins are required to be combined at a specific melt temperature to provide equivalent viscosities? The objective is to select resins that will have equivalent melt viscosities and the melt temperature where that occurs.

The first step in the process is to obtain the melt viscosity data for polycarbonate (PC) at three temperatures and relatively low shear rates. The viscosity data are given in Table 42.1 and plotted as the log of the viscosity versus shear rate, shown in Fig. 42.8. Viscosity data can be supplied in tabular form, as shown in Table 42.1, as a

Table 42.1. Polycarbonate (Lexan 121) Viscosity Data

Shear Rate, sec^{-1}	235° C, Pa.sec	250° C, Pa.sec	265° C, Pa.sec
0.6	3706	1738	898
1.0	3736	1789	920
10.0	3444	1683	907
40	2952	1540	869
100	2269	1298	787
250	1614	1018	661

Table 42.2. Polypropylene Viscosity Data

0.5 Melt Flow PP Homopolymer			
Shear Rate, sec–1	**235° C, Pa.sec**	**250° C, Pa.sec**	**265° C, Pa.sec**
0.1	8254	7981	7254
1.0	7353	6425	5754
10.0	2976	2576	2386
100	760	660	671
5. 0 Melt Flow PP Homopolymer			
0.1	4107	3022	2233
1.0	2755	2222	1648
10.0	1327	1190	911
100	432	434	328
12.0 Melt Flow PP Homopolymer			
0.1	1791	1397	731
1.0	1302	957	657
10.0	705	547	409
100	265	221	178
35 Melt Flow PP Homopolymer			
0.1	519	238	184
1.0	465	221	181
10.0	323	180	140
100	158	102	82

curve fit equation for easy interpolation, or taken directly from a shear rate versus viscosity curve, as shown in Fig. 42.8. From the PC viscosity data and the graph, it is obvious the viscosity is relatively Newtonian (viscosity does not change as a function of shear rate) in the shear rate region between 1 and 100 sec^{-1}. Obviously from these data, small increases or decreases in PC melt temperature during coextrusion can alter the melt viscosity dramatically, resulting in a viscosity mismatch between layers.

The second step is to obtain the viscosity data for polypropylene (PP) at different temperatures and shear rates. Viscosity data for different melt flow PP homopolymer resins are provided in Table 42.2 and plotted in Fig. 42.9. Unlike PC viscosity, melt temperature changes do not have a dramatic effect on the viscosity, and shear rate changes do. Small changes in PP melt temperature do not have the same effect on viscosity as seen with PC. Note the curves in Fig. 42.9 are in groups of three, with the highest viscosity in each group being at the lowest temperature, 235˚C, and the lowest viscosity being at 265˚C. Proceeding down the graph to lower viscosities, the PP melt flow index increases.

To determine which PP resin to coextrude with PC and the melt temperature to use, pick a shear rate that is appropriate for the process being run and generate an Arrhenius plot. Using the viscosity at a selected shear rate, plot the log of the viscosity versus $1/T$ (T is in Kelvin). This produces straight line graphs that can be extrapolated. Extrapolate both the PC and PP lines until they intersect and determine the melt temperature. This is the melt temperature where the viscosities are

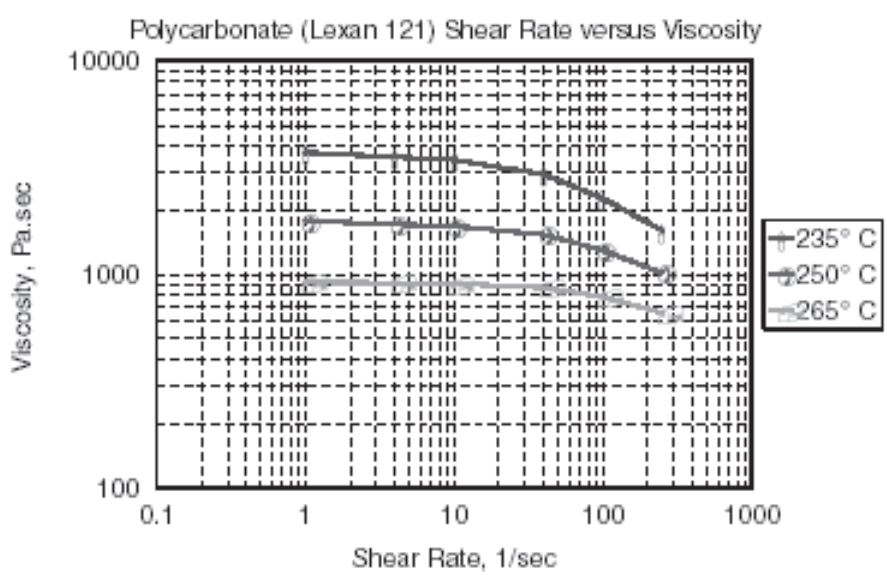

Figure 42.8. Shear rate versus viscosity data for polycarbonate.

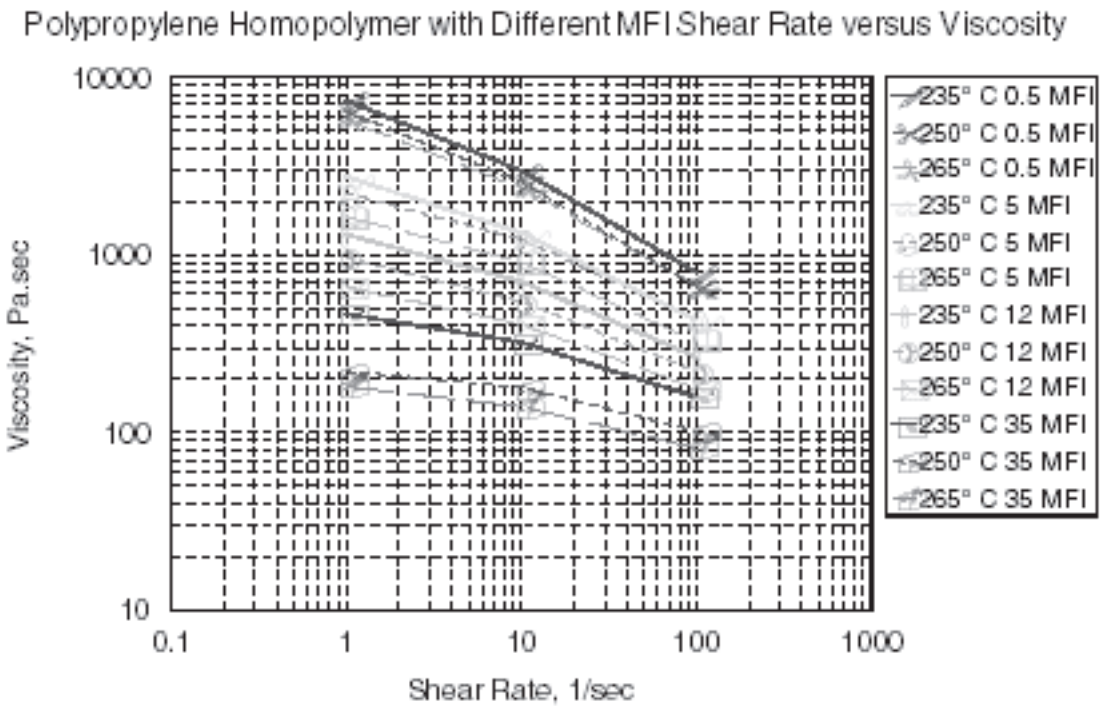

Figure 42.9. Shear rate versus viscosity data for different MFI PP Homopolymers.

equivalent. Calculations for the purpose of this example are made at a shear rate of 10 sec^{-1}. Table 42.3 shows the viscosity data that are entered into the Arrhenius plot graphed in Fig. 42.10. Based on Fig. 42.10, 0.5 melt flow PP has the same viscosity as PC at approximately 236˚C (457˚F). While extrusion of PP is feasible at 236˚C, extruding PC at reasonable rates at that melt temperature is impractical. Shear heating from the screw will result in higher melt temperatures than 236˚C or the extruder will require too much torque, creating a motor overload situation. Consequently, it is unlikely that Lexan 121 PC can be extruded at that low a melt temperature. The 5 MFI PP resin viscosity curve intersects the PC curve at 261˚C (502˚F), and both PC (Lexan 121) and PP (5 MFI) will have similar viscosities at that temperature. Since this is a reasonable processing melt temperature for both PP and PC, these two resins could be coextruded at that temperature and have similar viscosity. The 12 MFI PP resin curve intersects the PC curve at 300˚C (572˚F). While this is a good melt temperature to process PC, it may be too hot for PP, leading to resin degradation over time. The 35 MFI PP resin does not intersect the Lexan 121 resin at reasonable processing temperatures for PP. From these data, it is obvious the resin and temperature have to be carefully selected to ensure resin viscosity matches. Once the proper processing temperatures and resins are selected, attention to the proper tie layer to use to bond the two resin layers together is needed. The tie layer needs to have similar viscosity to both the PP and PC at the selected processing temperature. Based on the data presented, the tie layer has to have a viscosity of approximately 3000 Pa/sec at 261˚C (502˚F) to provide a viscosity match with both the PC and PP.

Table 42.3. Viscosity Data to Generate Arrhenius Plot

Shear Rate = 10 sec^{-1}			
Resin	**1/(265+273)**	**1/(250+273)**	**1/(235+273)**
0.5 MFI PP	2386	2576	2977
5 MFI PP	911	1219	1326
12 MFI PP	409	547	705
35 MFI PP	140	180	323
Lexan 121	907	1683	3444
Shear Rate = 100 sec-1			
0.5 MFI PP	671	660	796
5 MFI PP	328	435	433
12 MFI PP	178	221	265
35 MFI PP	82	102	158
Lexan 121	787	1298	2269

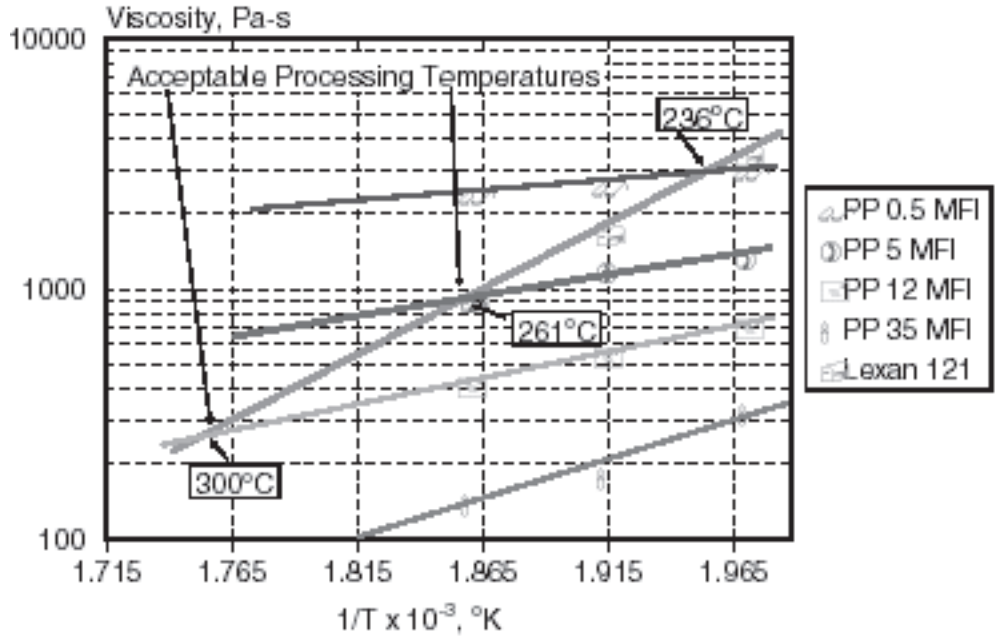

Figure 42.10. Arrhenius plot for example at 10 sec^{-1} shear rate.

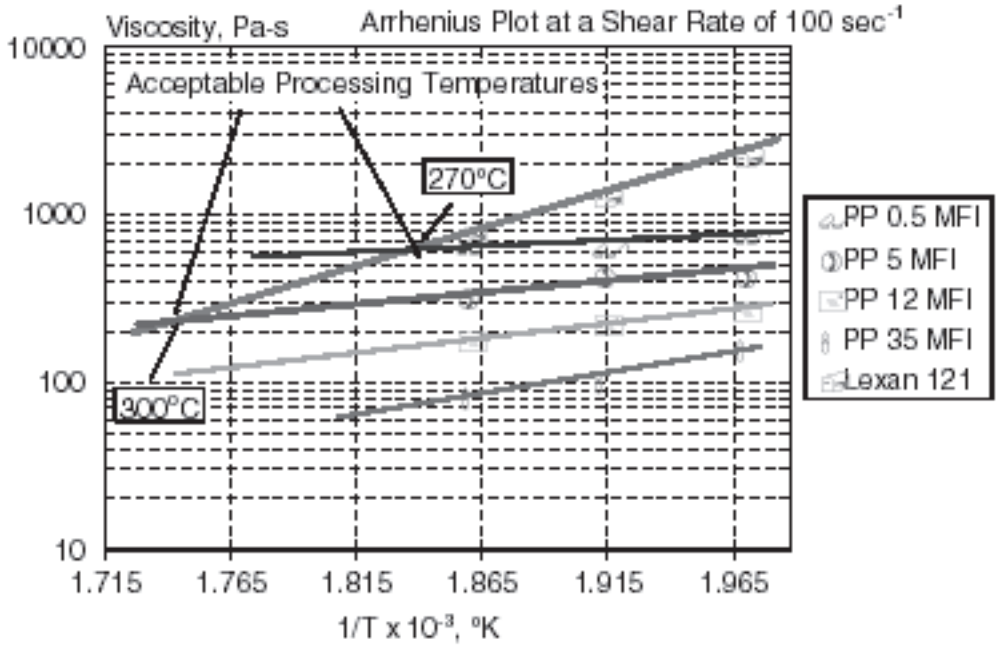

Figure 42.11. Arrhenius plot for example at 100 sec^{-1} shear rate.

What happens to the curves if the shear rate in the die or feedblock is increased to 100 sec^{-1}? Figure 42.11 shows the Arrhenius plot based on a 100 sec^{-1} shear rate. To match melt viscosities at this shear rate, PC has to be coextruded with 0.5 MFI PP at a melt temperature of 270˚C (518˚F) or 5 MFI PP resin at 300˚C (572˚F). These temperature conditions and resin choices are entirely different from those required at 10 sec^{-1} shear rate to match melt viscosities. For successful coextrusion without interfacial instability due to shear stress, it is important to select the correct resin viscosities and melt temperature to ensure that the viscosities are the same at processing conditions and shear rates the materials are exposed to in producing the coextruded structure.

Figure 42.12. Wave instability.

Zigzag type instability can occur regardless of the number of layers present and the thickness of the layers. Zigzag instability has been observed even when adjacent resins are the same resin system. This can be caused by two different viscosities of the same resin or differences in pressure or melt temperature.

A second interfacial instability, called wave instability, is observed as a set of relatively parallel parabolas in the machine direction, as shown in Fig. 42.12. Wave instability occurs at the interface and is attributed to large differences in the extensional viscosity[3] between adjacent layers. As coextruded structures enter a die from separate melt channels, the extensional viscosities can be relatively high as the polymer is elongated from a small transfer pipe to a wide die. If the extensional viscosity ratio between resins in adjacent layers is high, wave-type instabilities are common. Wave instabilities can be generated in either the feedblock or die. Extensional or elongation viscosity is a measure of the resistance to stretching. Resins with higher extensional viscosities can be stretched more in the polymer melt without breaking. This is a very positive factor in most extrusion operations because it provides better melt strength as the molten polymer exits the die. However, it can be a negative factor if wave pattern interfacial instabilities are present in coextruded structures.

In the past few years, significant emphasis has been placed on layer thickness variation caused by secondary flows. Studies coextruding different color resins with the same molecular formulation and viscoelastic properties to study the flow patterns associated with secondary flow effects have been very effective in showing this behavior. Flow patterns are dependent on the flow channel geometries and the resin elastic properties. These observations have some far-reaching implications, because flow patterns suggest layer rearrangements may occur in coextruded structures even though the viscosities are well matched. In one study done with polystyrene, polycarbonate, and polyethylene, the results suggest rearrangement in flow patterns is directly related to the polymer elastic nature, with polystyrene having the most layer rearrangement, followed by polyethylene, and essentially none with polycarbonate.[4] It appears secondary flow effects are generated normal to the main channel flow. Another study using two different colored resins in a two-layer structure investigated the effect of the channel shape on secondary flows. Square, circular, and teardrop shape channels were evaluated for secondary flow effects because these are the most commonly found shapes in feedblocks and dies. Layer rearrangement was observed in the channel after the melt streams were combined. Rearrangement was most severe in the square channel, followed by the teardrop design, with the least rearrangement taking place in the circular channel. The longer the polymers flowed together in the channel, the greater the layer rearrangement became, even though the resin viscosities were identical.[5] This has some significant implications for feedblocks and wide dies where the flow path after the melt streams are combined can be quite long.

Secondary flow effects have been measured experimentally for polyethylene and polystyrene in a square channel and shown to be about 0.33% of those in the principal flow direction.[6] Although these velocities are very low compared to the main machine direction flow, they are enough to disrupt the coextrusion layer geometry, particularly when long flow paths are present after the polymers have been combined. As secondary flows have been linked to elastic effects, both the melt viscosity and elasticity are critical when producing coextruded structures with uniform layer thickness.

Secondary flow effects were modeled with two identical polymers and found to be attributable to the polymer melt elasticity characteristics and flow channel geometry. Computer simulations using finite element analysis have been able to successfully predict the presence of secondary flow patterns.[7]

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Review Questions

1. What causes layer encapsulation and how can it be eliminated?
2. How can melt temperature affect instability between layers? How are melt temperature instabilities visible in coextruded structures?
3. What is zigzag interfacial instability, and what causes it?
4. How does extruder surging affect coextruded structures?
5. Why is it important to use viscosity versus shear rate data at three different temperatures when matching viscosities?
6. How is the polymer melt temperature and melt flow index selected to ensure a viscosity match between layers?
7. What interfacial instability is related to extensional viscosity, and how does it appear in coextruded structures?
8. What are secondary flow effects?
9. How do secondary flow effects manifest themselves in the final coextruded structure?
10. How is the shear rate determined where the melt fronts combine?
11. Why is melt velocity important for interfacial instabilities?
12. If melt temperature variations cause layer thickness variations, how can the melt temperature of the different layers be made similar? What are common melt temperatures in a melt transfer pipe?
13. Name the interfacial instability that occurs when shear stress between the layers is a key factor. Where is the shear stress the highest in a five layer A/B/C/B/A structure?

43 Troubleshooting Coextrusion

Troubleshooting and problem solving extrusion problems were covered in detail in Part 4. Troubleshooting relative to coextrusion deals only with coextrusion problems and more specifically with problems associated with feedblocks and dies. Melt temperature uniformity, surging, mixing, warpage, and a host of other problems were covered in Part 4. Various coextrusion problems are covered individually with potential causes and corrective actions to implement to eliminate the problem.

43.1 Problem 1—Nonuniform Cross or Transverse Direction (CD or TD) Layer Thickness

Layer thickness variation is one of the most common problems associated with coextrusion. It is identified as a layer thickness variation across the extrudate. Nonuniformity can be in either the machine direction (MD), transverse (TD) or cross-machine direction (CD), or both directions simultaneously, depending on the cause. Potential causes for thickness variation in a layer in the transverse direction with potential corrective actions are summarized below.

- The die gap is not properly adjusted, causing thickness variations in the product in the TD or CD. Assuming the die lips are not damaged or warped, this is easily corrected by readjusting the die gap opening. In a sheet die, the choker or restrictor bar has to be properly set to distribute the melt flow uniformly. With a pipe or tube die, verify that the mandrel is properly positioned in the middle of the die opening.
- Improper die design or defective equipment such as warped die lips can create TD thickness variations. More serious is a poor die design that does not provide uniform flow across the entire die. If flow is higher in one area of the die versus another, this may be partially corrected by changing the melt temperature. It is more likely a better-designed die is required to eliminate the problem.
- One of the biggest causes of layer nonuniformity in the TD is caused by melt temperature variations that affect resin viscosity from one location in the die to another. Melt temperature variations can originate from different locations in the process, starting in the extruder. Melt temperature exiting the extruder is dependent on the shear rate, barrel temperatures, throughput rate, screw speed, and screw design. As the melt enters the transition or transfer pipe, resin in the pipe center is flowing at higher velocity than that near the walls. It is not uncommon to have a difference of 20–30˚F (11–17˚C) in the melt temperature from the center to the pipe wall. Material at the wall matches the pipe temperature. Variations in the melt temperature can be caused by the feedblock and the die. The polymer melt in the center of a coextruded structure is insulated from the die heat and transition pipe heaters by insulative plastic layers. Melt temperature differences entering the feedblock or die should be less than ±3˚F (±1.5˚C). Accurate melt temperature measurements at the end of the transition pipes are necessary to match viscosities. Melt thermocouples have to protrude into the melt stream to produce accurate melt temperature readings. Melt thermocouples near the wall measure the metal surface and not the melt temperature. This can be misleading and result in not addressing the correct factors when trying to solve problems. Melt homogeneity can be achieved in melt transfer pipes with static mixers to break and recombine the melt streams. Other corrective actions to change the melt temperature are to change the screw design, change the barrel temperatures, reduce the backpressure, lower the screw compression ratio, modify the screw mixing section, and with, fluid cooled screws, increase or decrease the fluid circulation.
- Melt flow from the die may be uneven due to melt temperature differences across the die. This can be caused by the die heaters not working properly or being improperly set so the polymer is heated in the wrong location, lowering the viscosity. If there is higher flow in one section of the die compared to another, lower the die temperature where the flow is greater, and raise the die temperature where the flow is lower.
- Viscosity mismatch between adjacent layers can lead to interfacial instability that causes layer thickness variations. This can be magnified by different flow velocities between layers. Lower-viscosity polymers will flow out to the edge and, in extreme cases, encapsulate higher-viscosity polymers.

Figure 43.1 summarizes the potential causes and corrective actions to implement in the event layer uniformity variations occur in the transverse or cross-machine direction.

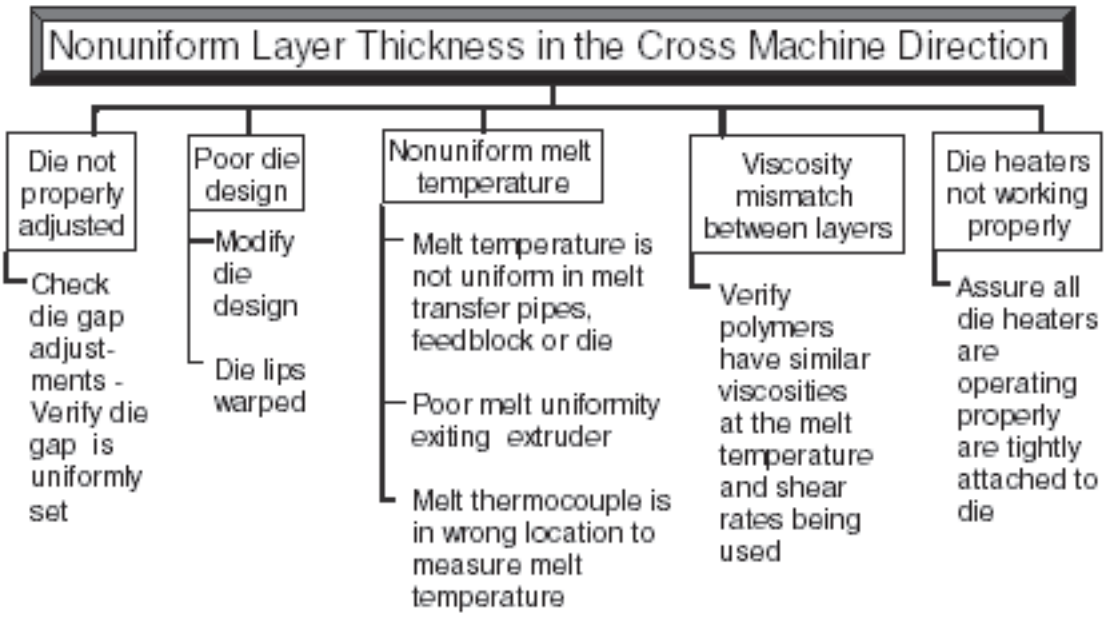

Figure 43.1. Potential causes and corrective actions for layer thickness variation in the cross-machine direction.

43.2 Problem 2—Nonuniform Machine Direction (MD) Layer Thickness

Potential causes for nonuniform layer thickness in the machine direction (MD) with potential corrective actions are presented below.

- If the die or feedblock is not properly designed or the flow channels prevent uniform flow, thickness variations will occur in the MD as the flow fronts are not uniform and consistent.
- Melt pressure variations as the layers are combined can be associated with surging or nonuniform flow velocity as the polymers combine. Melt pressure fluctuations will cause variation in the MD thickness. Surging, discussed in detail in Part 4, is frequently caused by inconsistent feed due to a partially blocked feed hopper, slipping and sticking on the screw in the feed section, the wrong extruder zone 1 temperature, poor feed characteristics caused by low friction between the material and the barrel wall, and improper feed throat cooling. Surging in the feed throat and zone 1 is frequently eliminated by using the correct temperature profile in extruder zones 1 and 2. Surging can also be caused downstream by incomplete melting in the screw transition zone, solid bed breakup, poor melt conveying, and the wrong screen pack or block screens. Regardless of why there is surging, it has to be corrected to eliminate MD thickness variations. Other potential causes for surging are the wrong screw design, and/or the wrong size diameter melt transfer pipe. The best way to eliminate surging is to install a gear pump either directly before or after the transfer pipe. Unfortunately this is also one of the most expensive ways to eliminate surging. Improved layer uniformity reduces the expensive barrier or tie layer resin required. This reduction in resin cost may justify installing a gear pump.
- If the takeoff equipment is not functioning properly, either product slippage or excessive tension can cause layer thickness variation. Slippage in pull rolls, a puller, or nip rolls allows the product to essentially start and stop, causing thick and thin spots in the MD.
- Improper screw design with the wrong compression ratio can lead to surging and nonuniform melt pressure.
- Melt conveying is insufficient.

Figure 43.2 summarizes the potential causes and corrective actions for eliminating MD layer thickness variations.

43.3 Problem 3—Interfacial Instability

Depending on the degree of interfacial instability, the structure may have minor layer thickness variations in mild cases to severe layer thickness variations in all

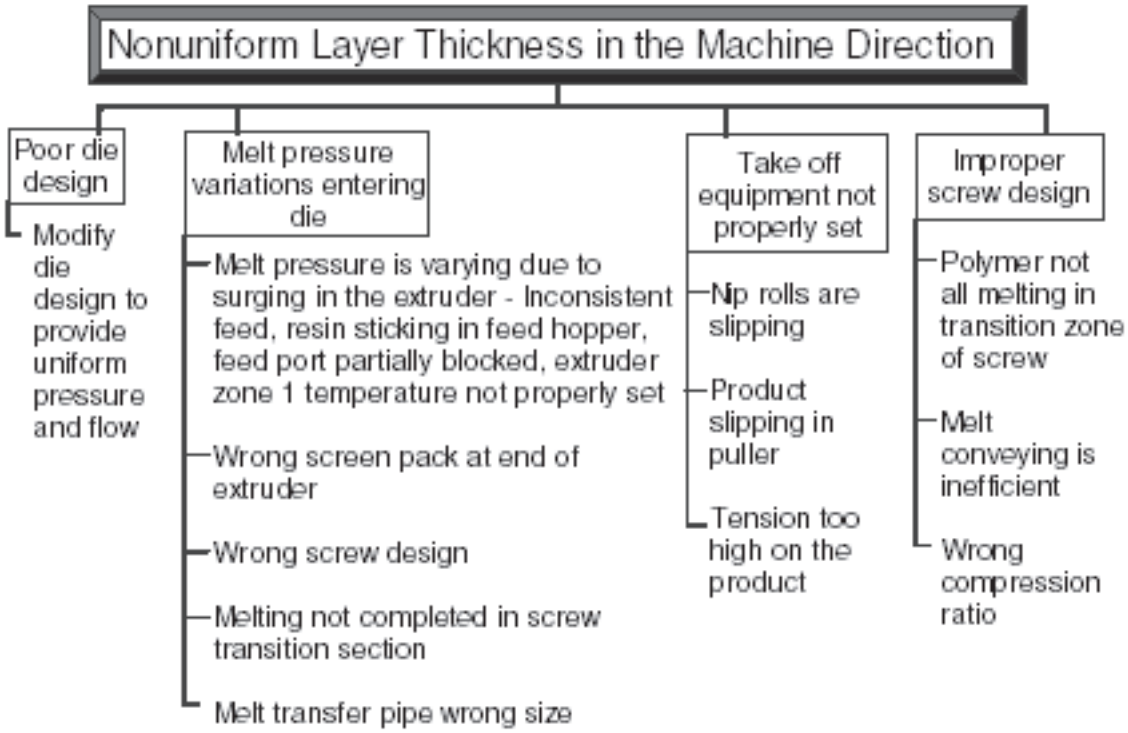

Figure 43.2. Potential causes and corrective actions for nonuniform layer thickness in the machine direction.

directions in extreme cases, rendering the structure totally useless. Potential interfacial instability causes, with corrective actions to eliminate the problem, are presented below.

- The greatest cause of interfacial instability is viscosity mismatch between adjacent layers in the coextruded structure. The different layers' melt viscosities must be similar at the melt temperature and shear rate used in the die and/or feedblock when the layers are combined and through the rest of the process. Melt temperature can affect the layer viscosities differently, so the melt temperatures have to be kept constant after the layers are combined. Procedures were described in Chapter 42 to select the proper melt flow resin and melt processing temperatures to ensure the resins have similar viscosities when the melt streams are combined. To guarantee that the viscosities are the same, the melt temperatures must be accurately measured prior to entering the feedblock or die. This means the melt thermocouple has to protrude into the melt stream, measuring the resin melt and not the pipe wall temperature. In addition, the melt temperature has to be uniform throughout the entire polymer melt. Melt temperature variations in melt transfer pipes can be homogenized or minimized through the use of static mixers.
- High interfacial shear stress can lead to zigzag-type instability, described in Chapter 42. Certain process changes can be made to reduce the shear stress and alleviate the zigzag pattern. Increasing the melt temperature will lower the resin viscosity, reducing the shear stress. Increasing the die gap and lowering the throughput will reduce the shear rate and shear stress. Increasing layer thickness or adding internal lubricants to the resin is another process change to minimize shear stress. Both actions may change the structure properties. Finally, the polymer viscosity may have to be changed to a lower-viscosity grade to eliminate the problem.
- A wave pattern instability appearing as numerous parallel parabolas in the machine direction is caused by melt elasticity differences between layers. To minimize a wave pattern problem, increase the layer thickness or change the resins to ones with more closely matched melt elasticity.

Figure 43.3 summarizes the causes of different types of interfacial instability and potential corrective actions to eliminate the problem.

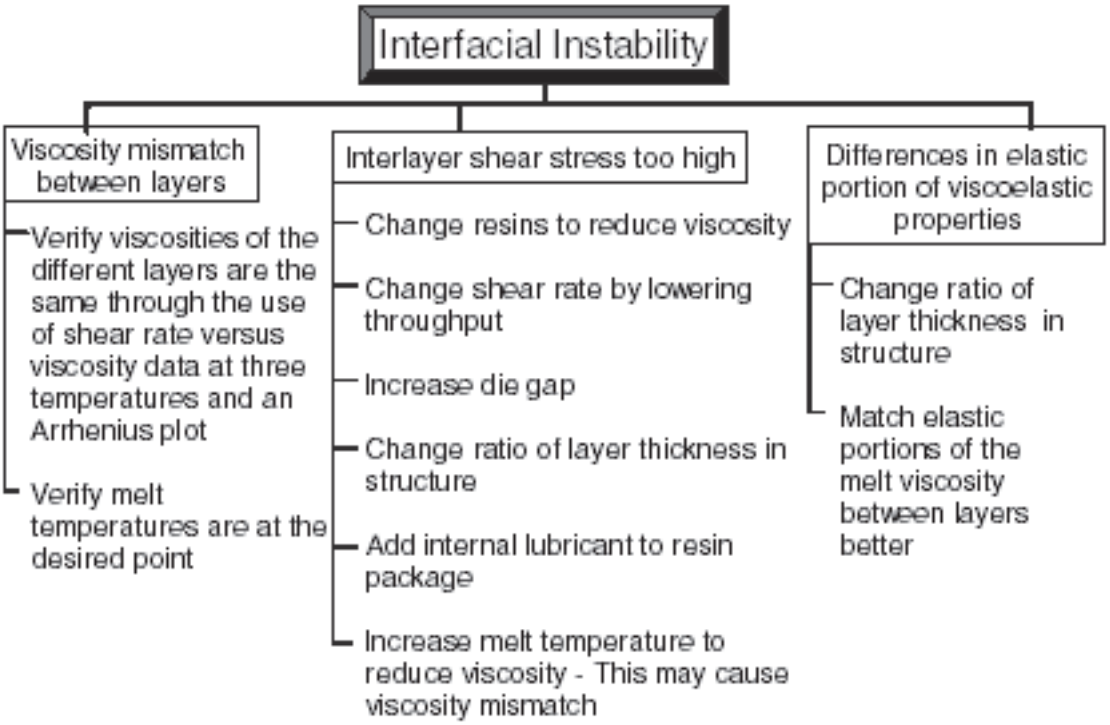

Figure 43.3. Potential causes and corrective actions for interfacial instability.

43.4 Problem 4—Repetitive Layer Thickness Variations

Potential causes with corrective actions to eliminate thickness variations in the MD that occur in a repeating pattern are presented below. The solutions are similar to those presented in problem 2 that were associated with nonuniform layer thickness in the MD. This problem varies slightly because it is reoccurring. The problem is caused by pressure variations entering the die, surging, or takeoff equipment.

- Pressure variation exiting the melt transfer pipe at the end of each extruder feeding either a feedblock or die has to have pressure variations of less than ±1%. If pressure variations are greater than ±1%, the extrusion process has to be modified through screw design changes, temperature profile modifications, throughput rate, or backpressure to reduce the difference between the maximum and minimum pressure readings.
- Surging that is periodic in nature can cause repetitive swings in layer uniformity. Surging is often created in the extruder feed throat by poor feeding characteristics associated with material sticking to the screw root; high friction between the material and the hopper wall, preventing the polymer from flowing freely into the extruder; the wrong temperature profile in extruder zones 1 and 2, causing the polymer to slip on the barrel wall and stick to the screw root; and/or improper feed throat cooling. The wrong screw design can cause surging by not melting all the polymer in the transition section, inefficiently pumping the melt to the die, and/or having the wrong compression ratio for the resin being processed. A worn screw can also lead to surging; regardless of whether the screw is worn or the design is wrong, changing the screw can eliminate the problem. Surging problems associated with only one of the multiple extruders feeding the feedblock or die can effectively ruin the coextruded structure, even if all the other extruders are functioning properly.
- Another way to build melt pressure uniformity is to use tighter screen packs that create higher head pressure in the extruder or through smaller melt transfer pipes that restrict polymer flow. Both of these approaches can reduce the throughput rate while increasing the melt temperature, which lowers the melt viscosity. Whatever corrective actions are taken to eliminate a problem has effects on other aspects of the process. Eliminating one

problem may create another. Consequently, it is important to understand the ramifications of any process or material change taken to eliminate a problem.

- Takeoff equipment can be responsible for a repeating pattern in the MD if the product is not being pulled in a continuous, uniform manner. Thickness variations will correspond to the nip rolls or puller operation uniformity. Slippage either in the puller or by the puller will cause nonuniform thicknesses. At the other extreme, excessive tension results in the product being pulled too strongly from the extruder and can lead to thin sections.

Figure 43.4 shows potential causes and corrective actions to eliminate repetitive thickness variations in the machine direction.

43.5 Problem 5—Layer Rearrangement

Layer rearrangement in coextruded structures can be caused by viscosity mismatches, secondary flow effects, and too large a bank in sheet and film three-roll extrusion. Corrective actions to eliminate these causes are summarized below.

- Viscosity mismatch without the presence of interfacial instability can result in the low-viscosity polymer encapsulating the high-viscosity polymer. Since the low-viscosity resin flows better, it actually flows around the high-viscosity resin, resulting in thinner layers for the low-viscosity and thicker layers for the high-viscosity resin. If the low-viscosity resin is in the cap layer, it flows toward the edge and down along the edge, preventing the higher-viscosity resin from reaching the edge of the product. If the higher-viscosity resin is in the cap layer, the second layer will flow out of the middle and replace some of the cap stock along the edge. It may be possible to reduce the high-flowing resin viscosity by increasing the melt temperature, and increase the poor-flowing resin viscosity by decreasing the melt temperature. To correct this problem better, temperature control and viscosity matching of the polymers at processing temperature are required.
- Extensional viscosity differences can cause layer rearrangement when viscosities are properly

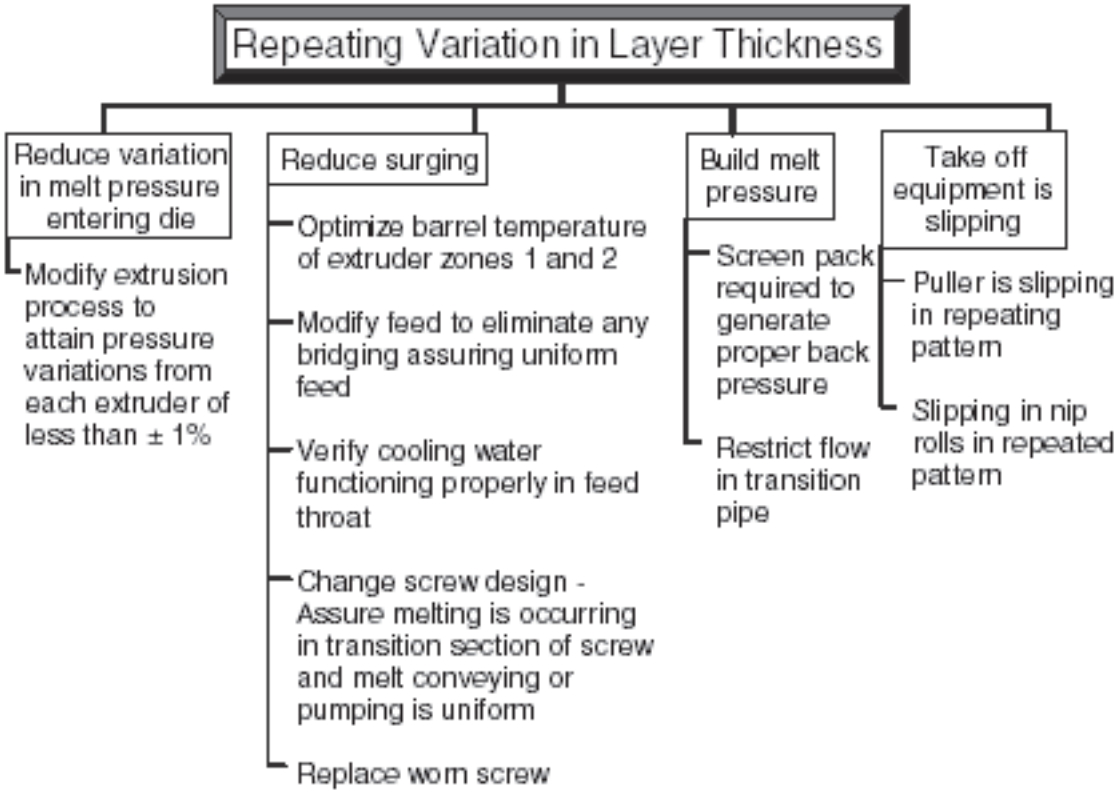

Figure 43.4. Potential causes and corrective actions to use for variations in layer thickness that repeat in the machine direction.

matched. In this event, changing the flow channel geometry after the resins are brought together in the feedblock or die can minimize layer velocity differences that lead to layer rearrangement. Shorter flow lengths where the velocities between layers are different can reduce effects attributed to secondary flow effects.

- Excessive rolling bank or bead size in coextruded sheet or film can lead to layer rearrangement as the rolling bead tries to pass through the three-roll nip. Reducing the bead can minimize the layer rearrangement.

The schematic in Fig. 43.5 summarizes the potential causes and corrective actions when there is layer rearrangement.

43.6 Problem 6—Layers Aren't the Correct Thickness

A coextruded structure is designed to have a layer thickness ratio of 26/16/11/16/26; however, after production the actual layer ratio is 13/30/14/30/13. Potential causes for the wrong layer ratio with corrective actions are summarized below.

- Viscosity mismatches that create higher flow of one polymer relative to another can cause layer thickness ratio variations. As with other viscosity matching problems, it is essential to have uniform viscosities across the entire structure.
- The leading cause of layer thickness ratio differences from that designed is the wrong throughput rate from the various extruders making up the structure. In an A/B/C/B/A structure with A layers being thicker than B layers, extruder A has to have a higher throughput than extruder B. If the extruder B rate is greater than extruder A, the layer ratios will be incorrect. The throughput rates from each extruder have to be matched to the layer thickness ratio. Single screw extruder throughput rate is dependent on screw speed, while starve-fed twin screw extrusion throughput rate is dependent on feed rate.
- Feedblock geometry is incorrect for the layer ratios desired. The feedblock geometry has to be changed to the correct ratio. Viscosity mismatches with the correct feedblock geometry can produce the wrong layer ratio due to viscosity effects.

Figure 43.6 summarizes the potential causes and corrective actions when the layer thickness ratio is incorrect.

43.7 Problem 7—Poor Adhesion Between Layers in Structure

Potential causes and corrective actions to use to overcome poor adhesion between different layers in a coextruded structure are summarized below.

- If adjacent polymers in the coextruded structure are not chemically compatible or only have minimal compatibility, the adhesion will be nonexistent or marginal at best. A tie layer that is compatible with both resins is required to provide the maximum adhesion between layers. The tie material functionality is selected based on the chemistry of the adjacent resins. To maximize adhesion, it may be necessary to use different tie layers.

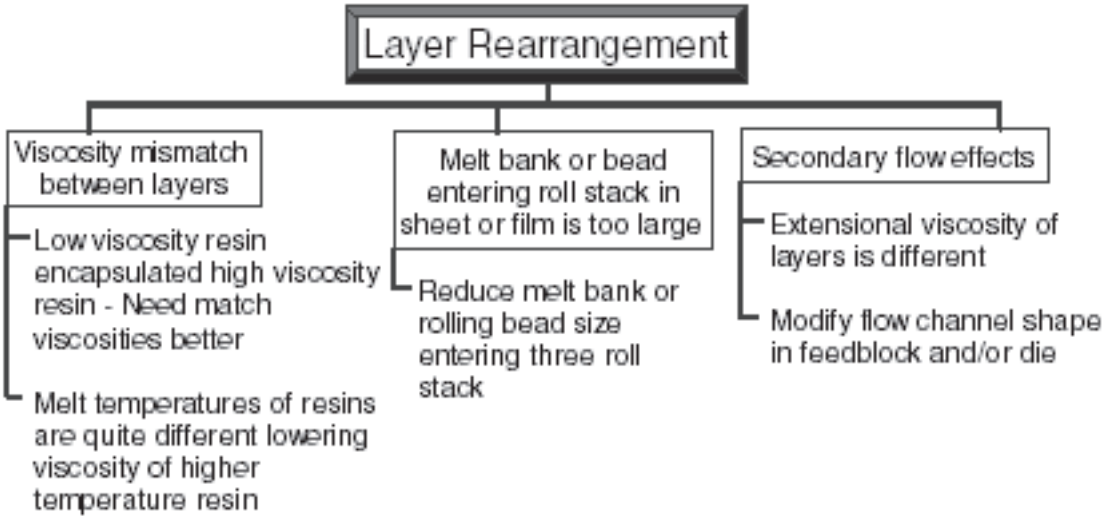

Figure 43.5. Potential causes and corrective actions for layer rearrangement.

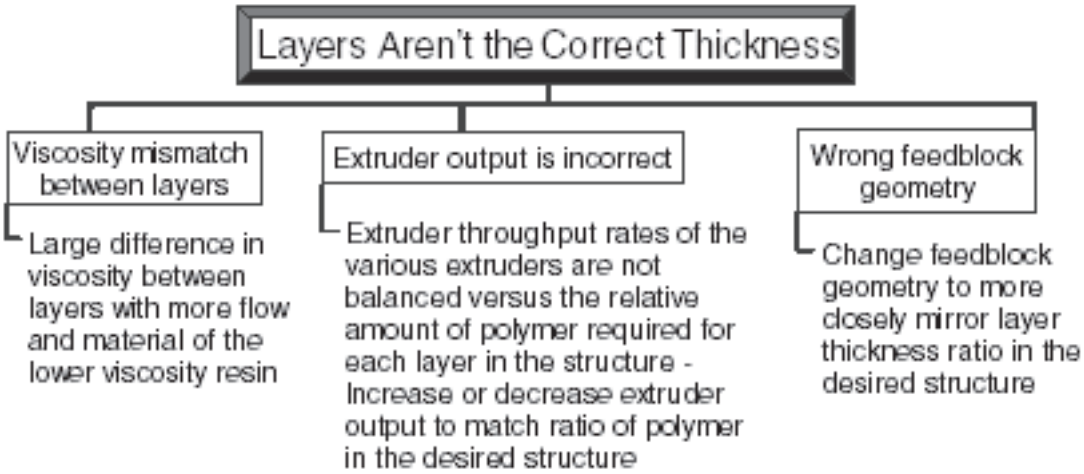

Figure 43.6. Potential causes and corrective actions for the wrong ratio of layer thicknesses.

- Processing conditions affect adhesion between layers and the performance of any tie layer material. Process changes that improve adhesion are higher melt temperature, longer residence time in the melt for the polymer to react, increased die temperature, less molecular orientation (slower puller speed or lower blow-up ratio in blown film), increased die land length, and higher temperature in the cooling state (roll temperatures, higher frost line height, warmer water temperature in pipe and tubing coextrusion, and so forth).

- Increasing the tie layer thickness improves adhesion between adjacent layers.

Figure 43.7 depicts potential problems and corrective actions to implement in the event of poor adhesion between layers.

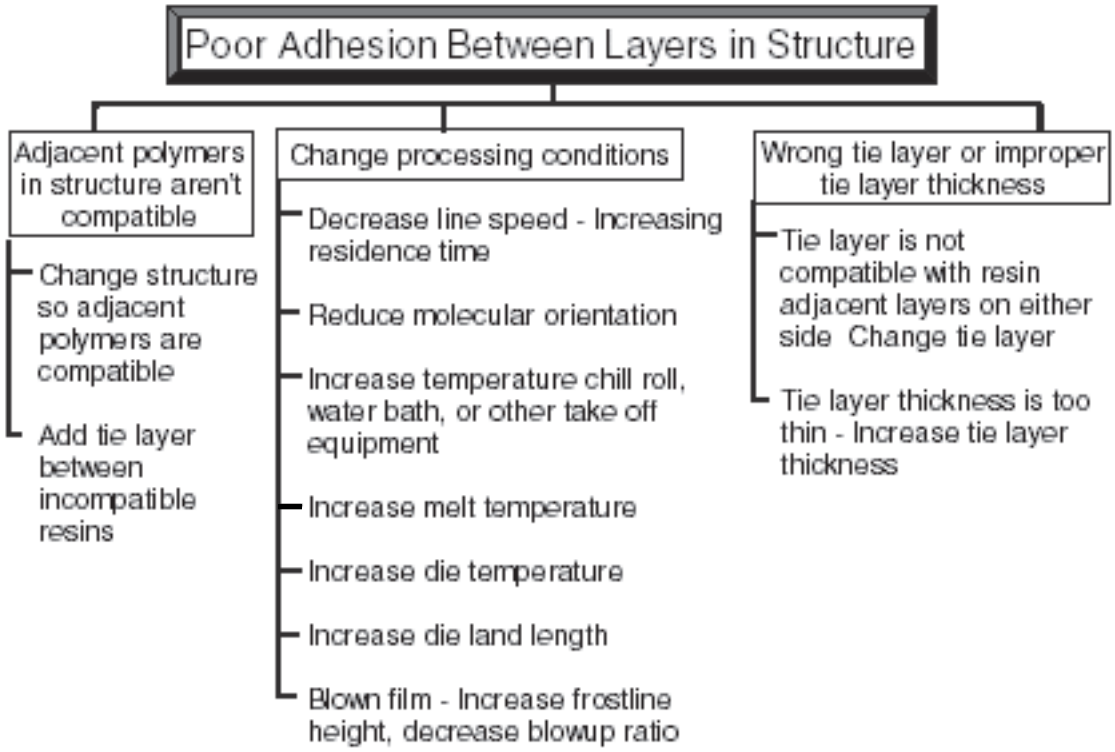

Figure 43.7. Potential causes and corrective actions to implement for poor layer adhesion.

43.8 Problem 8—Structural Integrity Problems from Recycle

Recycle from edge trim, start-up material, scrap from off-specification product, and shutdown material is frequently used as a filler layer in the final coextruded product. In many situations, the components in a coextruded structure are not compatible, and a tie layer is needed to bond different resins together. The polymer ratios in the structure determine the compatibility and structural integrity. As an example, assume the regrind is PP/tie/PC (60/10/30); the structural integrity of this mixture is very poor because PP and PC are totally incompatible. A 60/40 PP/PC blend delaminates; adding 10% tie layer is not a sufficient compatibilizer to make the extruded mass provide structural integrity. Potential causes and corrective actions when regrind is incompatible are summarized below.

- If the regrind contains high concentrations of incompatible resins, the challenge of using regrind as an individual layer in the coextruded structure may be particularly tough. The first step is to provide sufficient melt mixing to uniformly disperse the discontinuous phase in the continuous resin phase. The tie layer acts as a compatibilizer. However, tie layers are usually a small percentage in the coextruded structure, and there may not be enough present to do the job. Consequently it may be necessary to incorporate more compatibilizer into the blend to provide sufficient structural integrity. A second approach is to add virgin resin of the cheaper component to reduce the concentration in the regrind of the incompatible, noncontinuous phase so it is easier to disperse it in the continuous phase, while bonding it to the continuous phase through the use of a compatibilizer. The third alternative is not to put the regrind back into the coextruded structure. The regrind can be collected off-line, treated in some manner with other resin, compatibilizer, filler, reinforcement, and so forth, to create a value-added product that can be sold into a different application or used in another project.

- Coextruded structure properties that can be affected by regrind concentration and regrind compatibility are clarity (transparency), contact clarity (transparency based on the contents being in contact with the packaging), physical properties, adhesion to other layers within the structure, and overall extrusion processability. The solution to these problems is to modify the ingredient levels in the regrind composition, decrease the regrind concentration in the structure, or eliminate the use of that particular regrind in the final structure.

Figure 43.8 shows potential causes for loss of structural integrity due to incompatible regrind and actions to alleviate or eliminate the problem.

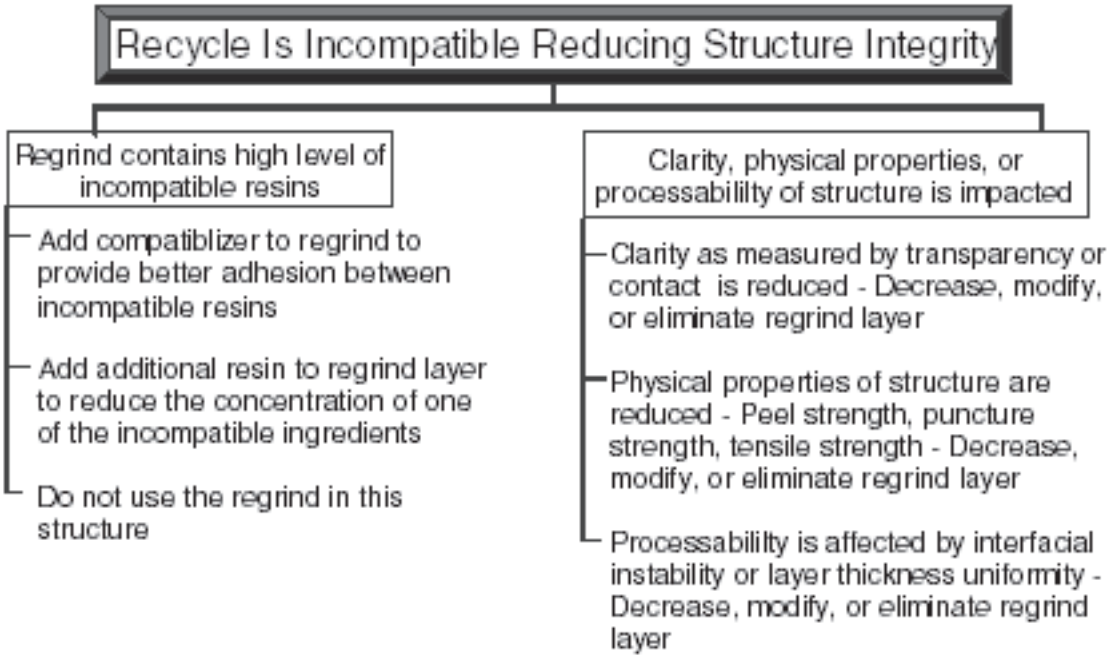

Figure 43.8. Potential effect of using regrind that is incompatible.

43.9 Problem 9—Haze Level Is Higher than Anticipated

High interfacial shear stress can cause interfacial instability between different layers. While the instability may be minor, the differences in shear stresses are sufficient to cause optical distortion between layers. This lowers the structure transparency and increases haze. Two fixes are to modify the polymer or change the process.

- Polymer modifications include matching the viscosity between layers more closely or changing the melt elasticity between layers. Melt viscosities can be changed by altering the melt temperatures. However, it may be necessary to select another melt processing temperature plus a different melt flow resin to properly match the viscosities at the process temperature and shear rate. Melt elasticity is more difficult to change. It may be necessary to change the melt flow rate of both resins or change one resins to more closely match the adjacent layer.

- The second way to reduce the shear stress between layers is through process modification, such as changing the layer thickness, increasing the die temperature or die gap, or lowering the throughput rate. Any process modification that lowers the viscosity or decreases the shear rate will lower the shear stress.

Potential causes of increased haze due to low levels of interfacial stability are shown in Fig. 43.9 with corrective actions to eliminate the problem.

43.10 Problem 10—Product Discoloration

Product discoloration is usually a yellowing or darkening. Following are potential causes and corrective actions to eliminate the problem.

- Excessive shear heating in the extruder can lead to resin degradation and yellowing in any layer within the structure. Shear heating can be minimized by changing the screw design to a lower-shear screw in the melting and/or mixing sections. While this change can lower shear heating, the screw must still melt the resin in the transition section plus provide adequate mixing to homogenize the melt temperature and any other ingredients in the mix and pump molten resin to the die. Shear heating can be reduced by lowering the screw rpm and throughput rate in single screw extruders. However, the increased residence time may still cause yellowing. Raising the barrel temperature may lower the melt viscosity and the degree of viscous shear heating.

- Resins that require drying prior to extrusion can yellow if overheated or dried for too long in the presence of oxygen. Every resin has a unique time-temperature curve where the resin degrades in both the solid state and melt. Higher temperatures for shorter times can have the same deleterious effects as lower temperatures for longer times.

- Excessive backpressure caused by too tight a screen pack, restrictive flow in the melt transfer pipe, or restrictive die flow can lead to higher melt temperatures, longer residence times, and resin degradation. As the backpressure increases, the

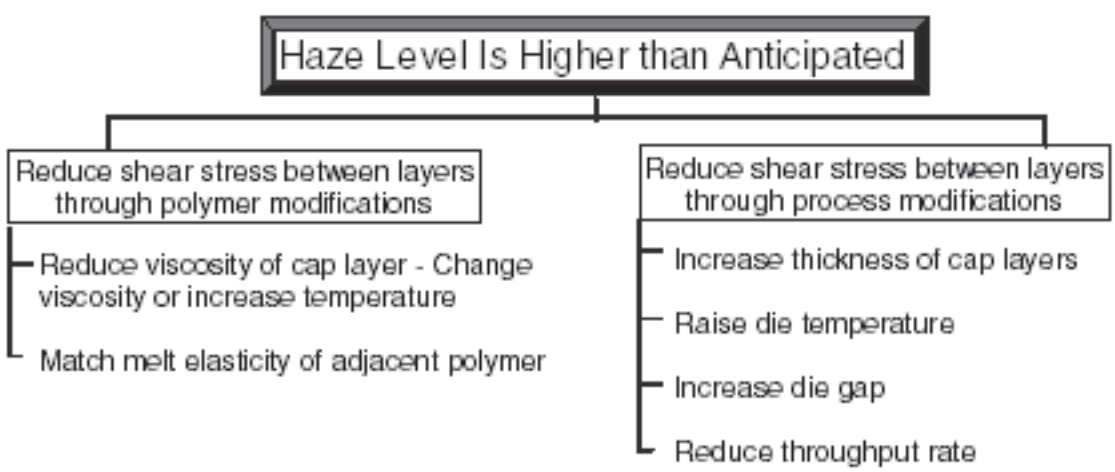

Figure 43.9. Potential causes for higher than anticipated haze level.

throughput decreases due to material flow back toward the feed throat. To overcome high pressure drop from the extruder to the die exit, the screw speed is increased, which generates higher shear heating and higher backpressure.

Corrective actions if yellowing or discoloration occurs are summarized in Fig. 43.10.

43.11 Problem 11—Gel or Lump Formation

In the strictest sense, gels are high-molecular-weight particles that don't melt. However, gels have been defined as almost any contamination, such as paper, bugs, foreign materials, and so forth. Cross-linked polyethylene acts similarly to a gel particle because it does not melt. High-molecular-weight particles that don't melt can be elongated and pass through screen packs. After passing through the screen pack, the elastic portion of the viscoelastic melt pulls it back into a tiny hard speck that shows up in the coextruded sheet as defects, referred to as fisheyes. Potential causes for gel or lump formation and corrective actions are summarized below.

- Polymer gels pass through the extrusion system because there is insufficient time and temperature to melt them. Increasing the polymer melt temperature or using a tighter screen pack can help eliminate the problem. Stagnant areas in the die, feedblock, or melt transfer pipes promote gel formation. High temperature and long times can cross-link polyethylene into higher-molecular-weight particles or gels, or degrade over time. Regardless of which mechanism takes place, the material generated is deleterious to the process.
- Paper contamination from the packaging and foreign pellets from dirty transfer lines, blender, or hopper causes lumps. The dirty regrind from contamination picked up off the floor, oil, and grease from equipment cause lumps, too. If the contamination source that causes lumps can be identified, it is usually easy to eliminate. In coextrusion, contamination in only one of five extruders from the transfer lines or other sources can make the entire structure useless. Therefore, it is important to eliminate all contamination across all the lines.

The potential causes for lumps and gels in a coextruded structure and the corrective actions are summarized in Fig. 43.11.

43.12 Problem 12—Warpage

Warpage is a twisting or bending of a coextruded structure that is anticipated to be flat or straight. Potential causes and corrective actions to implement to eliminate warpage are summarized below.

- One of the main causes for warpage is an unbalanced structure where layers on different sides of the structure have different shrinkage characteristics. In a round structure, product symmetry in a

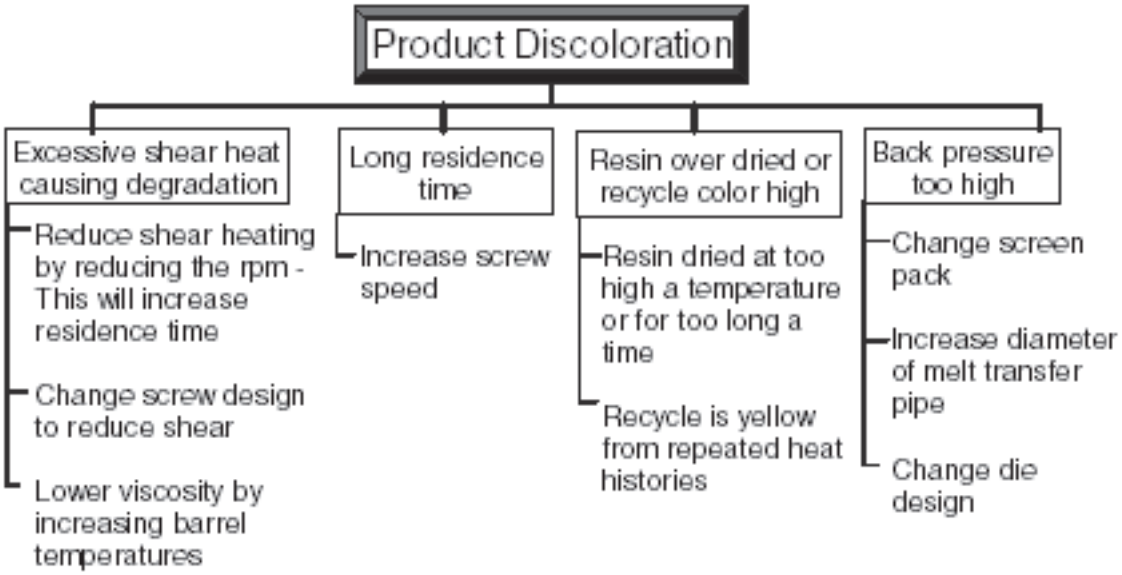

Figure 43.10. Potential causes and corrective actions to implement in the event of product discoloration.

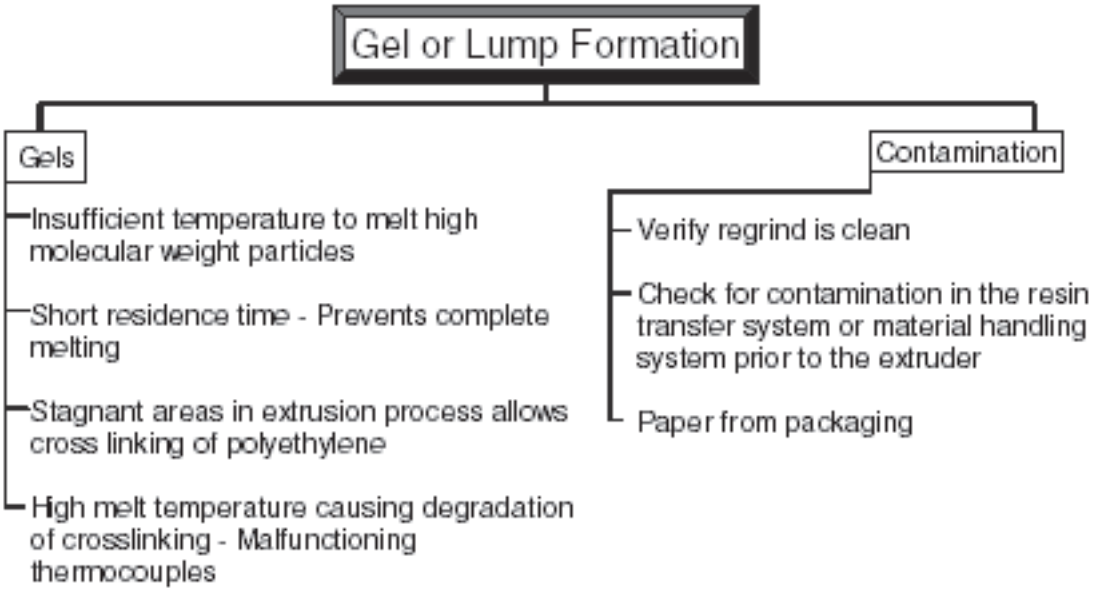

Figure 43.11. Potential causes and corrective actions to implement for gel formation.

pipe or tube prevents warping, even when the layers have different shrinkage characteristics. However, a sheet of profile with a different resin on one side is quite likely to warp, bow, or curl in the direction of the resin with the greater shrinkage characteristics. By controlling the cooling, it may be possible to minimize the unbalanced structure effect. Warpage is always likely to be present. Structures with crystalline layers on one side and amorphous layers on the other that do not have a center of symmetry (such as a pipe or tube) are likely to bow or curl toward the crystalline side because crystalline resins shrink more than amorphous resins. Anytime an unbalanced structure, such as A/B/C, A/B/C/D, A/B/C/B/D, and so forth, is used and the product is not symmetric in shape about a central axis, warpage is likely to be present. Unbalanced thin film structures are likely to curl.

- Nonuniform layer thickness in either the machine or the transverse direction can cause warpage due to differential shrinkage. Assume a coextruded balanced structure of A/B/A, where resin A has different shrinkage characteristics compared to B. If B is thicker in one section of the product and A is thicker in another section, as shown in Fig. 43.12, the shrinkage will vary across the part, leading to warpage that is caused by differential shrinkage from one position in the part to another. Nonuniform cooling, where the top solidifies either faster or more slowly than the bottom, can induce warpage even in balanced structures by generating different shrinkage characteristics within the structure. Warpage due to this phenomenon was discussed in detail in Part 4, Chapter 28, problem 15, and Chapter 29, problem 8.
- Molecular orientation differences in different layers can lead to warpage. Assume a balanced structure of A/B1/C/B2/A, where the thickness of B1 does not equal B2; the thinner layer may be drawn more than the thicker layer, creating a molecular orientation difference between B1 and B2. Molecular orientation correlates with different shrinkage properties in one direction versus the other. As with other extruded materials, differential

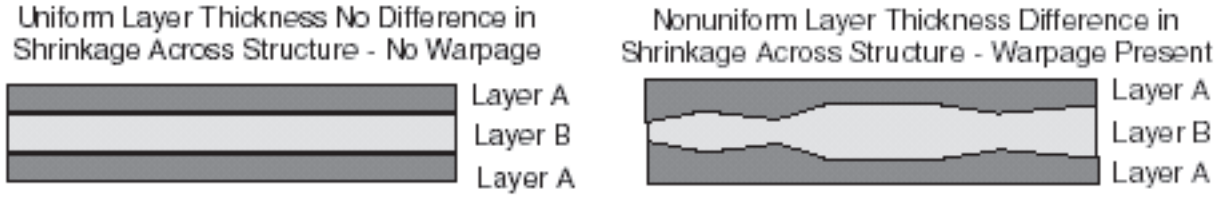

Figure 43.12. Warpage caused by nonuniform layers creating different shrinkage characteristics across the structure.

shrinkage across the part will lead to warpage in the final product.

A summary of the potential causes and corrective actions in the event of warpage are given in Fig. 43.13. A complete explanation of warpage, as it relates to extruded parts, is summarized in Part 4, Chapter 28, problem 15.

43.13 Problem 13—Regrind Utilization

Regrind from coextruded structures can create a special challenge because the resins making up the regrind may not be compatible. (Regrind issues were partially covered in problem 8.) The tie layer resin may help minimize the problem, or the regrind may not have the structural integrity or physical properties required. Potential ways to utilize regrind from start-up, scrap product, and trim are summarized below.

- The most important factor in using regrind is the mixture's chemical compatibility and the regrind physical properties. If the resins are compatible, they can be put back as a center layer with a tie layer on each side if necessary to provide the proper adhesion to the rest of the matrix.
- If the regrind is not compatible, choices for use of the regrind create more of a challenge. Generally the properties are inferior to any of the other individual resins in the structure. Consequently, the regrind may not be able to be used within the coextruded structure. Other applications have to be identified for the regrind. These may be to use internally or to sell outside to a scrap dealer. The other alternative is to identify additives that can be added to the system as compatibilizers, fillers, reinforcements, other resins, and so forth, to upgrade the regrind so it can be used in a value-added application. For incompatible polymers, it may be possible to add the major component, making it dominant. This disperses the other resin(s) as a discontinuous phase in the dominant phase. There are a number of options available to modify regrind; however, some product development work may be required to identify a viable use or modification procedure to make the product useful.

Potential procedures to use regrind in coextruded structures or to identify other applications for the product are summarized in Fig. 43.14.

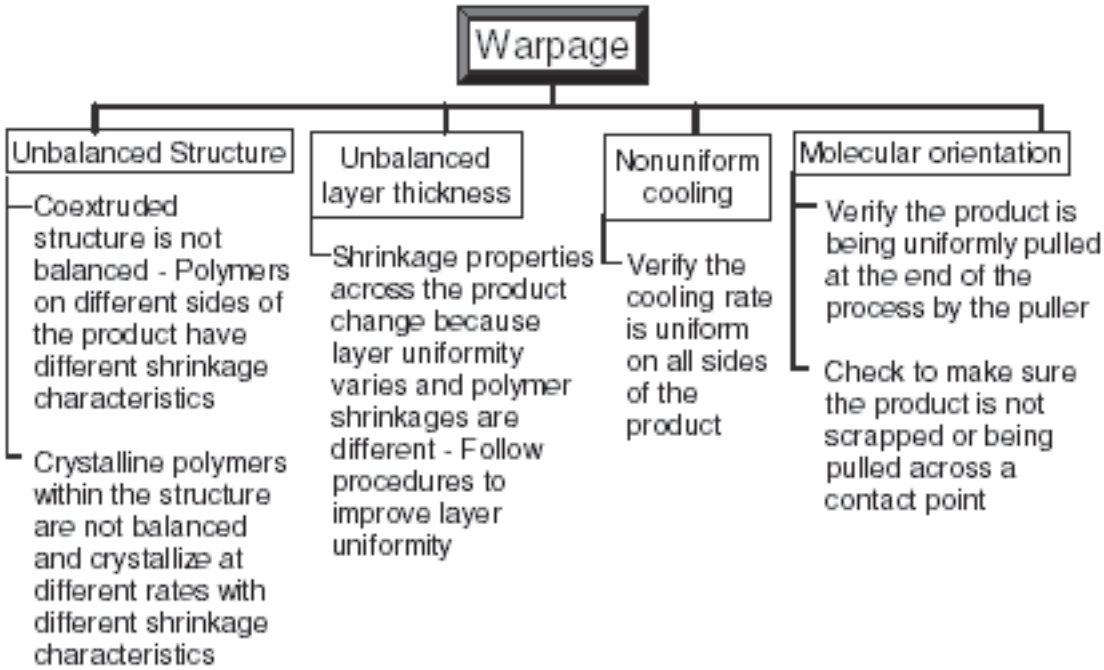

Figure 43.13. Potential causes and corrective actions to implement in the event of warpage.

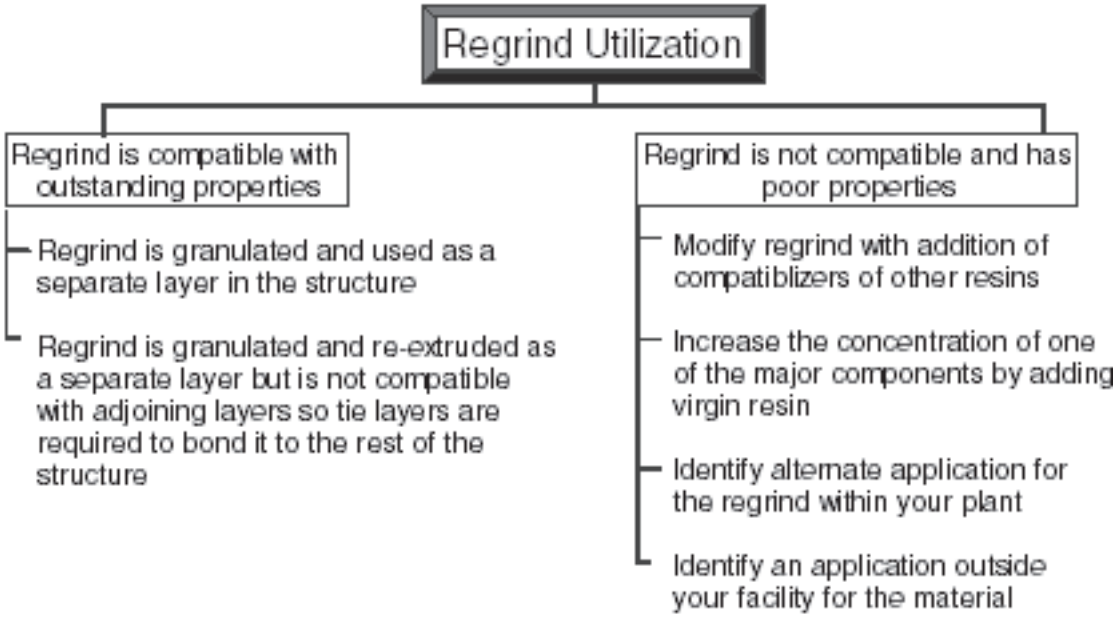

Figure 43.14. Potential uses for coextruded regrind.

Review Questions

1. In the event warpage is a problem with a coextruded structure, what corrective actions might be implemented to eliminate warpage?
2. What causes nonuniform layer thickness in the machine direction, and how can this problem be eliminated?
3. What causes polymer layer rearrangement, and what corrective actions should to be taken to eliminate the problem?
4. Why is viscosity matching between layers important?
5. What steps might be taken to improve the color of a coextruded product?
6. What causes nonuniformity in layer thickness in the cross-machine or transverse direction?
7. How is interfacial instability eliminated, and what causes the problem?
8. If extensional viscosities are not matched, what kind of problems might be anticipated?
9. What problems are anticipated if elastic properties are not matched?
10. In matching viscosities, what factors are important to ensure that the viscosities will match in the final structures?
11. What are the best methods of treating regrind from coextruded products, and what different regrinds are generated?
12. What factors may cause a repeating layer nonuniformity pattern in the machine direction?

Review Questions (continued)

13. What might cause layers in a coextruded structure to be the wrong thickness, and how is the problem corrected?
14. What corrective actions might be implemented to lower the haze level in the final coextruded structure?
15. What process variables can be changed to improve adhesion between layers?
16. What product variables can be changed to improve adhesion between different layers?
17. What actions might be taken if gels are present in the final product? What are gels?
18. What can cause contamination in the final product?

Part 7: Extrusion Applications

44 Compounding

The extrusion process, from equipment through processing, has been covered previously. Chapters 44 to 54 look at different processes and downstream equipment. The extruder and auxiliary equipment associated with feeding and supplying a properly conditioned melt to the die are only one part of the overall extrusion process. To produce a world class, quality product for the marketplace with tight product specifications and high production yields, the extruder and downstream equipment have to be operated properly and at peak efficiency. Understanding how an extruder functions and operates is not sufficient. The extruder and the melt it delivers have to be matched with the downstream equipment.

Each extrusion process has its own unique downstream equipment and operation relative to the extruder. Chapters 44–54 cover the following extrusion processes from both an equipment and an operation perspective:

- Compounding
- Sheet
- Cast film
- Blown film
- Wire coating
- Profile
- Pipe and tube
- Monofilament
- Large part blow molding
- Extrusion coating and lamination
- Rod extrusion
- Reactive extrusion
- Reprocessing

Chapter 44 covers the compounding process, equipment, and operation.

Compounding operations are used in conjunction with other processes to be discussed later. Compounding can occur prior to sheeting, film, profile, and other extrusion processes if more than two materials are fed into an extruder. Compounding as defined in this chapter is the mixing of two or more components in an extruder to produce a pellet that can be used in other processes, whether it is extrusion, injection molding, or rotomolding.

44.1 Pelletizing Process

The compounding process is defined as melt mixing different components in either a single or twin screw extruder to form a new material. This can be as simple as mixing a colorant in the form of a liquid, powder, or concentrate into the polymer melt to change the resin color, or as complicated as blending or alloying two or more different resin systems while adding a filler, reinforcement, colorant, flame retardant, and/or stabilizercs to produce a radically different formulation with its own unique properties and end-use performance. Figure 44.1 shows a compounding process with feed system, extruder, strand die, water bath, air knife, and pelletizer. Molten polymer exits the die as a strand approximately 1/8 inch (3.2 mm) in diameter. Strands pass through a water bath to remove the heat and solidify the strand, which will be cut to length in the pelletizer. Between the pelletizer and water bath is an air knife or air stripper to remove any surface moisture attached to the strand. A rougher strand surface makes it more difficult to remove the moisture after the strands exit the water bath. How many strands exit the extruder depends on the extruder size and throughput rate, with larger diameter extruders producing more strands and higher throughput rates.

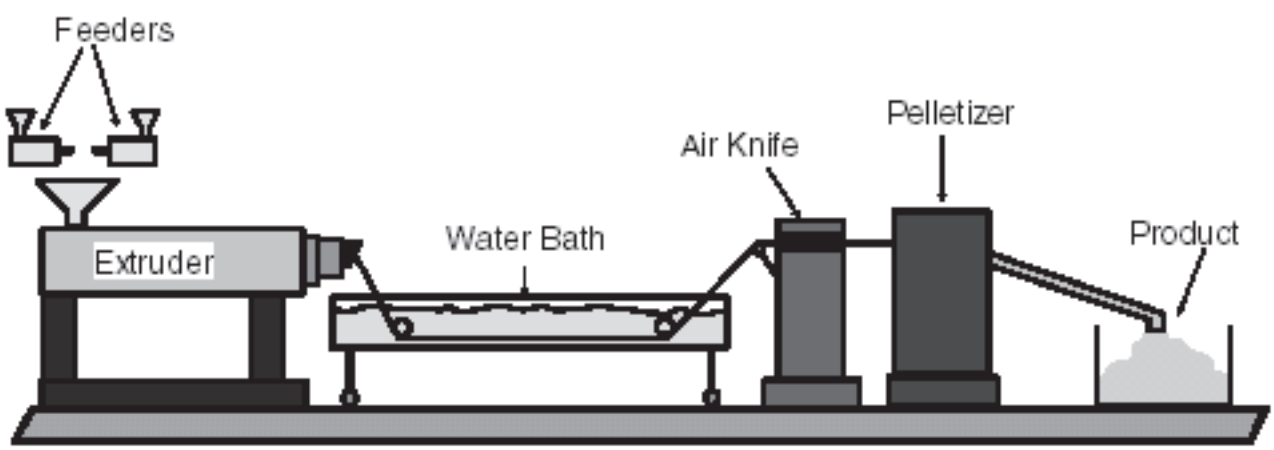

Figure 44.1. Compounding process using a water bath and strand pelletizer.

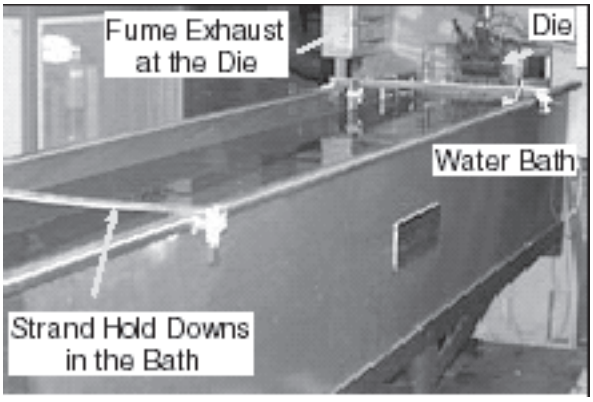

Figure 44.2. Berlyn water bath.

Figure 44.2 shows a water bath and extruder die with a fume hood surrounding the die to remove volatiles coming off the strands before they enter the water bath. Cold water enters the water bath at the tank bottom farthest from the extruder and exits from the tank top next to the extruder. This allows the warmest water to be removed from the tank. It is common to have the water bath water flow countercurrent to the strands. This results in the water temperature being hotter at the strand entrance than at the strand exit. If insufficient cold water is supplied to remove the heat, the water bath temperature can build up over the run, necessitating higher water flow to keep the strands cold enough so they can be pelletized.

Strand throughput is measured in pounds/hour/strand. Typical strand throughput rates range from 25 to 125 pounds/hour for a 1/8 inch (3.2 mm) diameter strand and 12 to 70 pounds/hour for a 3/32 inch (2.5 mm) diameter strand. A given water flow rate is required to remove the excess heat and cool the strands sufficiently for pelletizing. As the throughput rate per die hole increases, strand or water baths have to become longer or a serpentine path is required to ensure sufficient time for cooling the strand. A lower throughput per strand rate requires more strands run at slower speeds to produce the same extruder throughput rate. The time a strand spends in the water bath is easily calculated based on the polymer density and throughput rate. Assume a polycarbonate strand with a density of 1.21 grams/cm3 is running at 100 pounds/hour/hole, and the strand diameter is 1/8 inch. How long does the strand spend in a 50 foot long water bath? The following calculations show how to determine the time in the bath. Equation (44.1) gives the pounds/minute/hole throughput rate.

$$\frac{100\,\text{lbs}}{\text{hour}-\text{Strand}}\times\frac{hour}{60\,\text{minutes}}=\frac{1.66\,\text{lbs}}{\text{minute}-\text{Strand}} \quad (44.1)$$

Using a 1/8 inch diameter (0.125 inch or radius r = 0.0625 inch) and a 1 foot or 12 inch length, Eq. (44.2) calculates the volume of a 1 foot polymer strand.

$$\text{Vol}(1\text{ft})=\pi\times r^2\times L=3.1416\times 0.0625^2\times 12=0.147\,\text{inch}^3 \quad (44.2)$$

With the density, it is possible to calculate the weight of the 1 foot strand, Eq. (44.3).

$$Wt(1\text{ft})=0.147\,\text{inch}^3\times 1.21\frac{\text{gm}}{\text{cm}^3}\times\frac{2.54^3\,\text{cm}^3}{\text{inch}^3}=2.91\text{g} \quad (44.3)$$

At 1.66 pounds/minute/hole, there are (1.66 pounds/minute × 454 grams/pound) equals 753 grams/minute/hole. With each foot of strand weighing 2.91 grams, each strand in the process is running at the rate given by Eq. (44.4).

$$Speed=753\frac{\text{g}}{\text{min}}\times\frac{1\text{ft}}{2.91\text{g}}=259\,\text{FPM} \quad (44.4)$$

In a 50 foot long bath, each strand spends 11.6 seconds in the bath, Eq. (44.5), to cool from a 590°F (310°C) melt temperature to 200°F (93°C), which is below the glass transition temperature.

$$Time\,in\,Bath=50\,\text{ft}\times\frac{60\,\text{sec}}{1\,\text{min}}\times\frac{\text{min}}{259\,\text{ft}}=11.6\,\text{sec} \quad (44.5)$$

The cooling time required to adequately cool the strand depends on the strand diameter, the water temperature, and the temperature difference between the polymer and the water. The heat required to be removed (Q) is given by Eq. (44.6).

$$Q = \dot{m}\times C_p\times\Delta T+\dot{m}\times\Delta H_{fusion} \quad (44.6)$$

where

$\dot{m}$ = Mass flow rate in pounds/hour (kg/hr)

C_p = Heat capacity of the resin in BTU/pound °F (kJ/kg °C)

ΔT = Temperature difference between the melt temperature exiting the die and the strand temperature at the end of the bath in °F (°C)

ΔH_{fusion} = Heat of fusion of crystalline polymers as the polymer changes from a liquid to a melt in BTU/pound (kcal/kg). ΔH_{fusion} of amorphous resins is zero.

(If the result is calculated in BTUs, to convert this to other units, 1 BTU = 252 calories = 1055 joules, = 0.293 watt hour.) In the previous example, the amount of heat (Q) that has to be removed per hour per strand is calculated in Eq. (44.7). The heat capacity of polycarbonate is 0.3 BTU/pound °F.

$$Q=100\times 0.3\times(590-200)=11{,}700\frac{\text{Btu}}{\text{hr}} \quad (44.7)$$

This is then multiplied by the number of strands to determine the total heat to be removed. The heat removal efficiency depends on the bath water temperature. Heat transfer occurs more rapidly with lower water temperature due to the large temperature difference between the material and the water. The water must be recirculated using either a chiller, cooling tower, or once-through water to maintain the water bath temperature at the correct setting. If the water flow through the bath is insufficient, the water bath can heat up over time until the water actually boils. Water bath temperature can be critical to the processability of the strands. Water baths that are too hot may not provide adequate cooling when the strands enter the water to form a thick enough skin layer on the strand. This leads to periodic strand breakage. Cold water chills the strand skin rapidly. This forms a thick skin while the center is still molten. Strands where the skin layer becomes thick very rapidly may have a vacuum void down the center. This gives pellets with holes in the center. As the center cools, the outside skin cannot be drawn in toward the center as the center material is shrinking. Continued cooling and shrinkage of the center leads to a vacuum void in the center of the strand due to the resin shrinking away from itself, filling the constant volume formed by the outside skin. Holes in the center of pellets caused by vacuum voids is not a product detriment. Optimum cooling bath temperature accompanied with the proper strand diameter can eliminate vacuum voids.

The second factor influencing vacuum voids is the polymer thermal conductivity. Thermal conductivity measures heat transfer by conduction. Higher thermal conductivity allows more rapid heat transfer from the strand center to the surface, where it is transferred to the cold water. Higher conductivity materials are less likely to have vacuum voids as the entire polymer cross section is cooling relatively simultaneously and shrinking at the same time, without the formation of a thick solid skin. In practice, polymers are great insulators with poor thermal conductivity, leading to vacuum void formation. As with other plastic parts, the larger the strand diameter, the more likely the strand will have a vacuum void in its center.

All of the factors mentioned are important in determining the water bath size required to properly cool the product. Depending on floor space available, a wider shorter bath can accommodate the same total extruder throughput, running more strands at a lower throughput per hole rate. Fewer strands with higher throughput per hole may be easier to string up from the extruder die to the pelletizer.

After the water bath, an air knife or air wipe removes moisture on the strands. An air knife blows air across the strands as the strands are bent over a bar. The air knife needs to be positioned so it is not blowing moist air toward the die face and cooling the die. Depending on the water bath length, this may or may not be a problem. If the air is blowing toward the die, an air deflector can be installed over the water bath or the air knife can be turned at a slight angle so the air blows to either side of the die. (Preferably the air will blow to the nonoperator side.) An air stripper operates using vacuum. The break angle and strands are pulled through the air stripper in a similar fashion to an air knife, but air is pulled across the strands, pulling the excess water into the equipment instead of blowing it away. Water is collected and run into a drain. The advantage of an air stripper versus an air knife is that moist air is not blown into the room or toward the die. When the moisture being removed is large, an air knife can cause a wet spot on the floor, which is a safety hazard. Figure 44.3 shows an air knife. High velocity air impinges the strands, blowing the water off. Figure 44.3 shows the top separated from the bottom; however, one side is connected to the bottom and the other side is open to string up the strands through the equipment. An air stripper is very similar in appearance to an air knife except the water is pulled off the strands into the equipment via vacuum.

After exiting the air knife or air stripper, the dry strands enter the pelletizer, where they are chopped to length. Two feed rolls, designated as an upper and a lower feed roll, are used to pull the strands into the cutting section at constant speed. The lower feed roll is carbide or elastomer coated (depends on application) and driven by gears, belt, or chain. The upper feed roll is covered with different durometers elastomers. The material processed and its temperature determines the roll covering. A gap is set between the upper and lower feed roll based on the strand diameter. In larger pelletizers, the upper feed roll is gear driven in synchronization with the bottom roll. In smaller pelletizers, the top roll may be an idler roll, forcing the strands against the bottom, driven feed roll, with the moving strands driving the upper roll. Either a mechanical device (in smaller

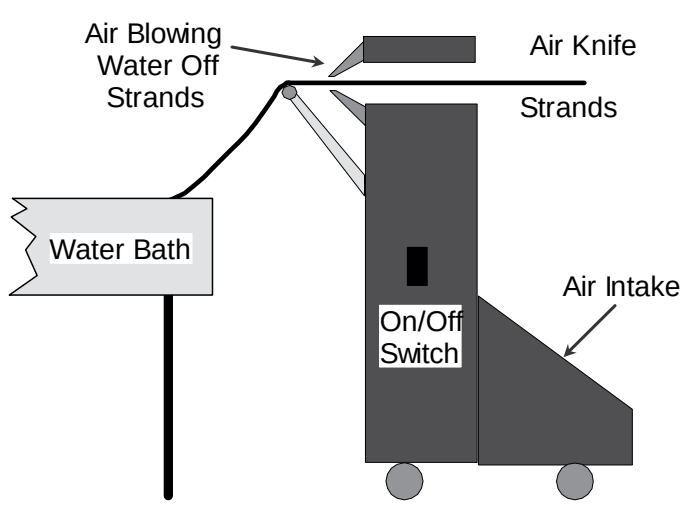

Figure 44.3. Air knife.

pelletizers) or an air-actuated cylinder raises and lowers the upper feed roll. In operation, air is normally used to hold the top roll down at the specified gap opening, with the strands being pulled by the rollers at constant speed as they pass between the rollers. Upper-roll wear can cause strands to slip in the pull roll, leading to dropped strands and processing problems. Roll wear is observed as grooves in the rolls or nonuniform wear across the roll where the edges may have a larger diameter than the center.

The pull roll speed is critical to the pelletizer operation; it has to be properly synchronized with the throughput rate or the process will continuously produce unacceptable strand quality or dropped strands. High pelletizer speed draws the strands away from the extruder too fast, generating significant postdie drawing. This results in a smaller strand diameter than desired. Excessively high pull roll speed can cause the molten polymer to break as it exits the die. Low puller roll speed causes the strands to be too large in diameter. This results in significant quantities of rejected product because the pellets are too large. Excessively slow speed can cause adjacent strands exiting the die to touch each other and become a double strand. If the strands touch each other coming out of the die, quite frequently a large mass will be produced that cannot go through the pelletizer or air knife, breaking down the entire process.

Figure 44.4 shows an open view of a pelletizer with upper and lower feed rolls, bed knife, and rotor. The cutting action is a scissor cut between the bed knife and rotor. The bed knife and feed table are slightly below the feed roll centerline, pulling the strands into the pelletizer; this ensures a clean pellet cut. Rotors either have individual bolt-on blades or come as one-piece helical rotors with the number of cutting blades dependent on the rotor size and anticipated throughput or cutter speed. Small or laboratory size 4-inch diameter rotors have 4 or 8 blades. Six-inch diameter rotors use between 14 and 36 blades, whereas 12-inch diameter rotors have between 48 and 72 blades. Helical cutter blades do not cut all the strands simultaneously as the blade progressively cuts strands across the rotor face. This reduces the energy required to cut the pellets while providing a cleaner cut with less fines and longer blade life. Cutting blades are normally made of high speed steel, D-2 tool steel, or carbide and stellite coating to extend blade life. Figure 44.5 shows an opened pelletizer where one can see the upper feed roll and the bolt-on blades on the rotor.

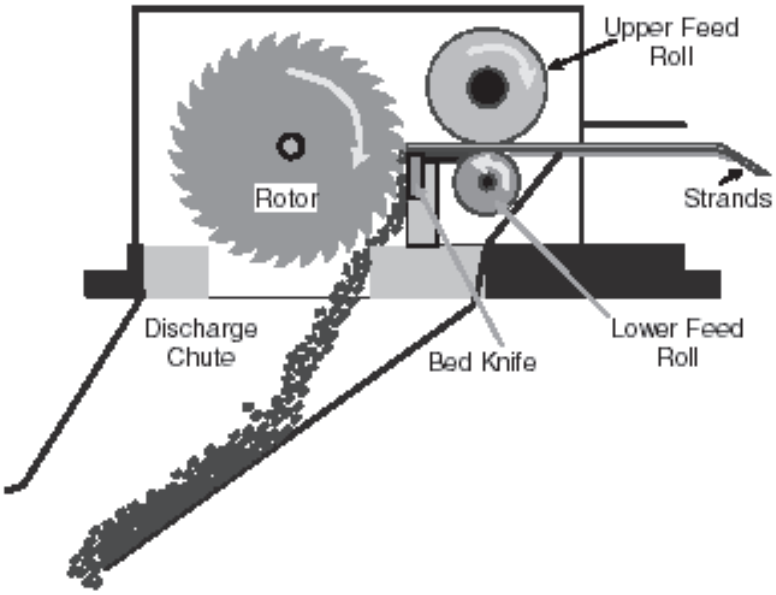

Figure 44.4. Pelletizer schematic.

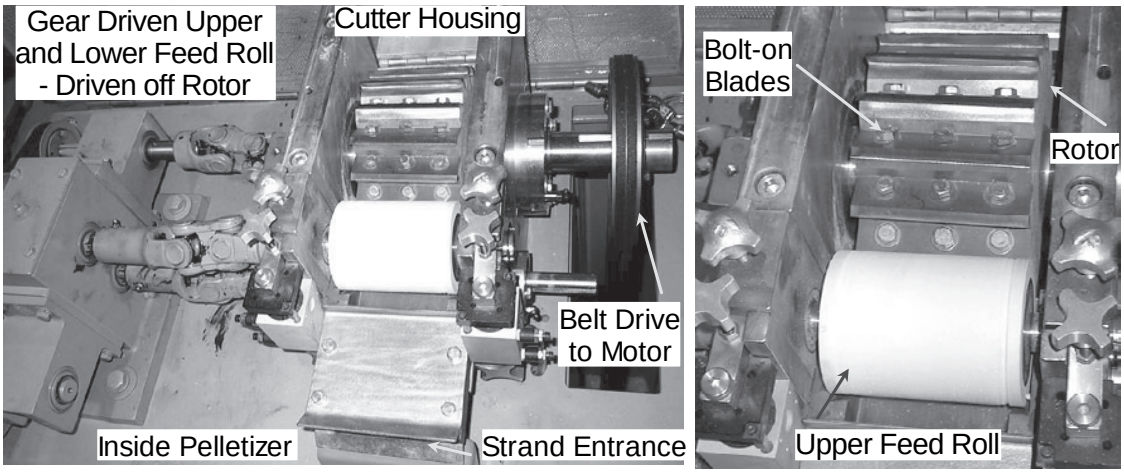

Figure 44.5. Inside a Conair 6-inch pelletizer. Left shows drive belts and gear housing to transfer speed from rotor to upper and lower feed rolls, right shows cutter roll and upper feed roll housing.

The bed knife can be rotated as it wears, allowing all four cutting edges to be used before it has to be resharpened or replaced. Rotor blades, whether replaceable or part of a helical cutter, are sharpened on the rotor to ensure that the length each blade extends from the center of the rotor is exactly the same. Bed knife rotor clearance is set between 0.001 and 0.005 inch; if the rotor blades are not all sharpened simultaneously and properly set, the cutter blade and the bed knife clearance can vary from one cutter blade to another, resulting in poor cut quality or incomplete cutting. If the clearance is too high, strands may not be properly cut, resulting in sections of plastic ribbon being generated that can clog and jam the cutter discharge chute.

The number of blades required are given in Table 44.1[1] for 1/8 inch (3.2 mm) and 3/16 inch (2.5 mm) strand diameters, based on the throughput/hole/hour.

To change the pellet length, some pelletizers have independent feed roll speed and rotor speed controls. As the feed roll speed relative to the rotor speed is increased, the pellet length becomes longer. Pellet length is decreased by reducing the feed roll speed relative to the rotor speed. In practice the feed roll speed is dictated by the extruder throughput and the strand size. After the feed roll speed is properly set to provide a stable, consistent process, the rotor speed is increased or decreased relative to the puller roll speed to change the pellet length.

After the pellets exit the pelletizer, they pass through a classifier to remove fines or large pellets created in the pelletizing process. Large pellets are minimal if the pelletizer is run at the correct speed, the cutter blades are in good condition, and the gap between the blades and bed knife is properly set. Screens are available to segregate pellets according to the product specification. Generally two different screen meshes are utilized, with the top screen mesh allowing all the pellets meeting the upper size limit specification to pass through, where they drop on to a second screen. The second screen mesh size allows all pellets below the lower limit to pass through. In this manner, oversized pellets are collected off the top screen, first-quality product is collected from between the screens, and fines and undersized pellets are collected from the bottom of the classifier. All off-spec pellets can be recycled back into the product at a lower percentage or sold as second-grade product.

There are basically two classifiers that work on the principles of either vibration or oscillation to move the pellets across the screen to the exit port for collection. One classifier is rectangular and sits at a slight angle. Pellets are dropped on the highest point and vibrated down the incline. Screens make up the incline plane, allowing pellets to pass through the screens to separate the product. The second classifier is circular. The pellets are dropped onto the middle of the screen. As pellets drop through the screens, they fall onto a funnel-shaped section that delivers the pellets to the center of the next screen. Vibration moves the pellets from the center to the edge, where they are discharged into a collection chamber. Figure 44.6 shows cross sections of both classifiers.

Table 44.1. Cutter Blades and Rotor Speed for Various Throughputs

Throughput, lbs/hr/strand (kg/hr/strand)	Required Number of Blades	Rotor (Surface) Speed, ft/minute
1/8" (3.2 mm) Diameter Strand		
72 (33)	72	225
60 (27)	60	190
48 (22)	48	150
36 (16)	36	112
24 (11)	24	75
3/16" (2.5 mm) Diameter Strand		
30 (14)	75	165
25 (11)	60	140
20 (9)	48	110
15 (7)	36	85
10 (5)	24	55

Once pellets exit the classifier they are conveyed by vacuum, air, or mechanically to a holding bin for packaging in bags, fiber packs, or gaylords, or transferred to tank truck or rail cars. Depending on the throughput rates, gaylords can be filled directly from the classifiers.

The final product is tested and certified to meet product specifications prior to shipment. If the plant has a statistical process control (SPC) program, product can be certified based on the control charts and minimal physical property measurements, such as melt flow index and color.

An underwater or die face pelletizer can replace the water bath, air knife, and pelletizer. An underwater pelletizer is typically used with higher throughput operations with resins that process below 625°F (329°C). This eliminates handling many molten strands and the potential for broken strands to interfere with production. Extrusion processes running at many thousand pounds/hour have greatly improved efficiency when they use underwater pelletizers. Visual pellet inspection identifies the process used to produce them. Round cylindrical pellets are produced using a water bath or water slide technology (discussed later), while a tear drop shaped pellet is manufactured with an underwater pelletizer.

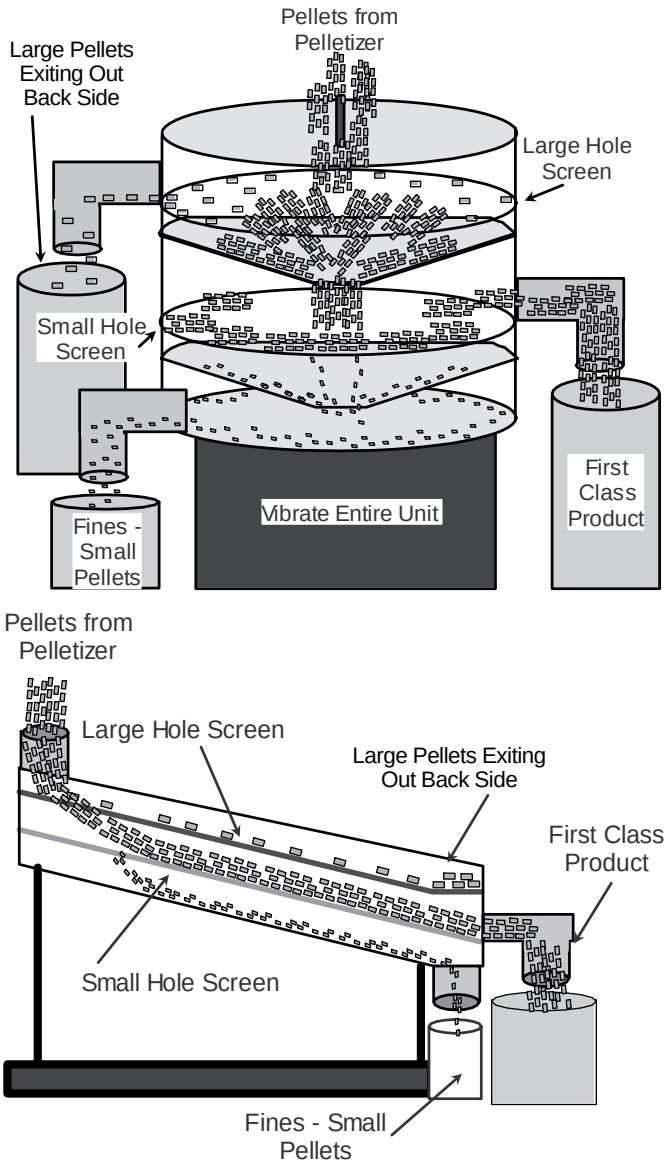

Figure 44.6. Two types of classifiers to segregate pellets.

The die in an underwater pelletizer is round, with die holes arranged in a circular pattern. Rotating knives fit flush against the die face in a water chamber; as the extrudate passes though the die, it is cut by the rotating knives underwater. Water cools the pellets, forming a thick skin, as the rotating blade slices the molten polymer flush with the die face. The water flow carries the pellets to a centrifugal dryer where the water is removed and the pellets are dried. The entire process is a self-contained operation; pellets exiting the dryer are ready for packaging. Figure 44.7 shows an extruder equipped with an automatic screen changer, gear pump, and underwater pelletizer.[2] An eight-blade cutter head is shown on the cutting unit. In operation, the cutting unit is moved forward on rails into the die, ensuring proper alignment between the die face and the cutter blades. Water flows into the cutting chamber bottom and out of the top. The water cools the cut pellets and conveys them away from the die. A sight glass in the exit pipe allows viewing the cut pellets as they are transported away from the die. The key to underwater pelletizing is to have molten polymer on one side of the die and water on the other without cooling the die enough to freeze off the die holes. This is accomplished by high flow rate through the die.

In operation, the die is bolted directly to a diverter valve; polymer flow is established through the die, and the cutter is engaged. The extruder is started with the polymer flow running to the floor through the diverter valve. Once adequate polymer flow is established, the diverter valve is closed and polymer is purged through the die with the cutter pulled back, allowing the polymer to exit as spaghetti-type strands through the die. This establishes that the die is not frozen off and that polymer is flowing freely. The diverter valve is opened, the cutter head is engaged with the die, and the cutter is rapidly brought up to cutting speed. The diverter valve is closed and water is started. The polymer flows through the die into the moving water, where it is cut into pellets by the rotating cutter head. Logic for this sequence is programmed into the underwater pelletizer, requiring the pushing of a single button to establish the proper sequence of the diverter valve closing, water flow starting, and the cutter head rotating at the proper speed. Pellets are conveyed in the water to the dryer, where the water and pellets are separated. Initially the pellets may be oversized, as the cutter head is coming up to speed. The oversized pellets are separated from the standard product through a diverter chute. Water is recirculated through the underwater pelletizer while the pellets are dried and packaged. Water temperature and die throughput are critical to the operation to prevent polymer from freezing off in the die. If some of the die holes freeze off, due to insufficient polymer flow, the extruder throughput needs to be increased or the number of die holes decreased. Die hole freeze off results in slightly large pellets, as higher throughput/hour/hole is occurring through the other die holes at the same rotor speed. Cutter head speed dictates the pellet size. Slower speed with fewer blades results in larger pellets at the same overall throughput. Micropellets are made with an underwater pelletizer using smaller die holes and more cutter blades.

Before any cutting occurs, the cutter blades have to fit flush against the die face. Blades are run for a short time with no polymer present to confirm that the blades sit flat against the die face. The die face must be very smooth with round holes containing no damage or deformation around the hole exit. Defects in any die hole can lead to tails on the pellets, as the polymer is pulled away instead of being cut clean.

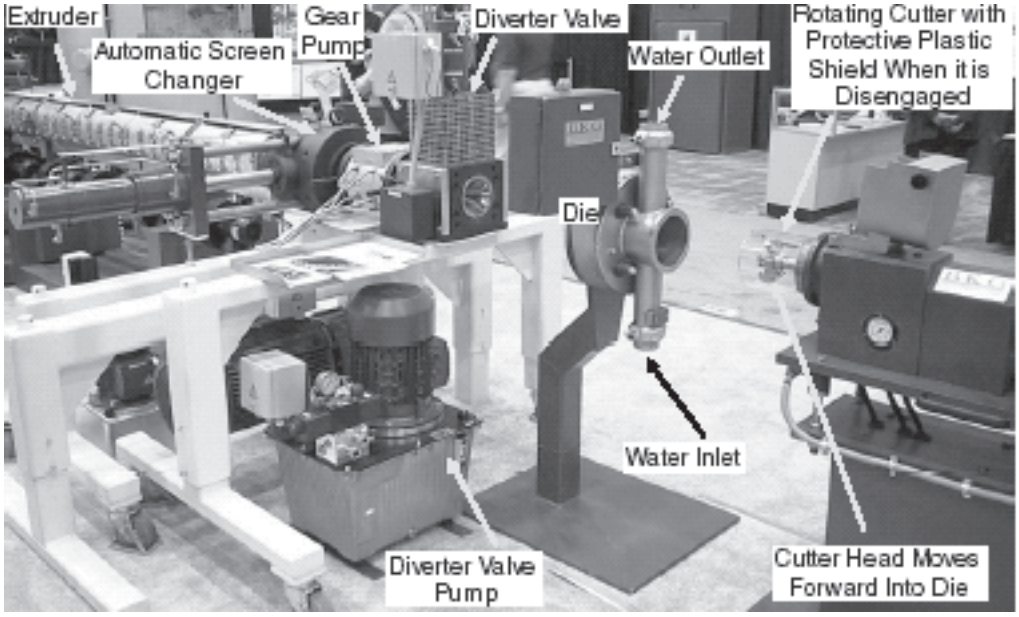

Figure 44.7. BKG underwater pelletizer connected to the end of an extruder equipped with an automatic screen changer and gear pump.

Advantages of underwater pelletizing systems are

- Once a die face pelletizer is up and running, the process is self-contained from the entrance to the extruder to the pellets exiting the process.
- There is no possibility of dropping strands or having slow die holes interrupt the operation.
- Product changeover from one product to another is fairly rapid, with no need to clean the pelletizer. Purge between items can be removed through the diverter valve at the end of the extruder.
- A tempered water system and dryer can be located away from the extruder, reducing the overall space needed for water baths and pelletizers.
- There is less downtime because cutter blades are self-sharpening
- Minimal maintenance is needed. Preventive maintenance requires blade replacement and bearing lubrication.
- The noise level is low, even at high pelletizing speed.
- Small or large pellet sizes are possible.
- Start-up is simple using the automatic sequencing.
- The system has the flexibility to cut all types of products, ranging from hot melt adhesives to thermoplastics to thermoplastic rubber.

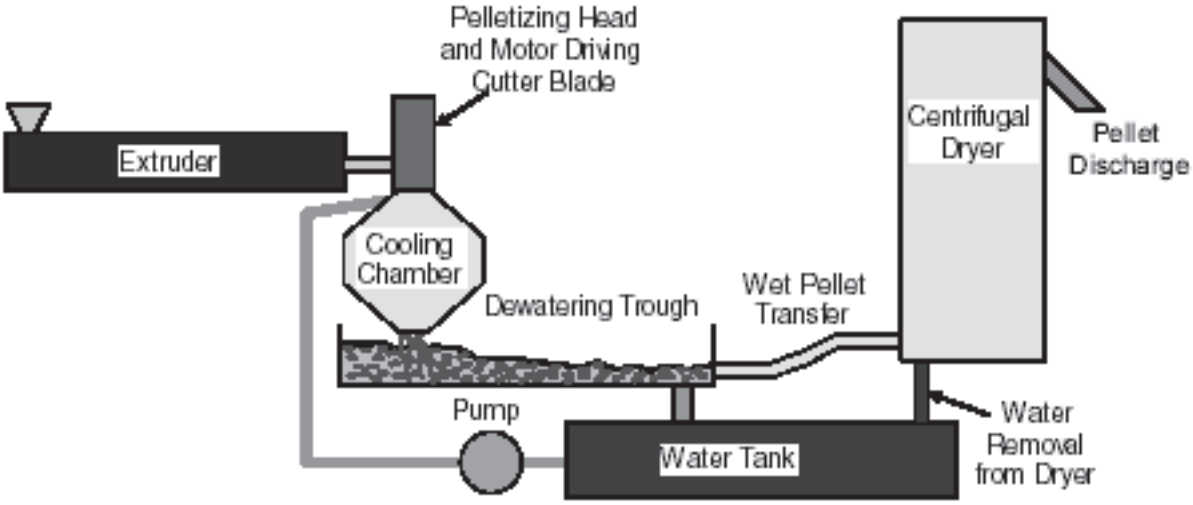

Figure 44.8. Process flow in Beringer water ring pelletizing system.

A third pelletizing system, supplied by Beringer Division of John Brown Plastic Machinery, is a water ring pelletizer, which is similar to underwater pelletizing in that the molten polymer is cut at the die and submerged immediately in water, creating a slurry.[3] Figure 44.8 shows the water ring process. Polymer exiting the extruder passes into a pelletizing chamber, where the polymer flows through an annular die equipped with a flexible blade that cuts the molten polymer as it exits the die. After cutting, molten pellets are thrown into a ring of falling water in the cooling chamber, where the polymer solidifies. The cooling chamber shape and high water flow prevent the molten pellets from agglomerating before cooling. Based on the die plate and pellet requirements, pellets can be formed as cylinders, spheres, or lens shaped. The pellet/water slurry is transported to a dewatering unit where excess water is removed from the pellets. Final drying is done with a centrifugal dryer, similar in operation to that used with an underwater pelletizing system. The water is recycled back to the pelletizing head and cooling chamber. Water temperature is maintained via a chiller or replacing hot water with cold water.

Advantages of a water ring system compared to strand pelletizing are

- Less floor space required
- Cheaper cutter blades
- Less energy required to cut
- Probably most significant, no strands to breaks

Relative to underwater palletizing, the advantages of a water ring system are

- There is reduced die freeze off because water is not running against one side of the die.
- Control system is less complicated because the automatic start-up cycle is not required.
- A flexible blade eliminates operator blade adjustment.
- Less power is required because cutting occurs in the air rather than underwater.

A fourth cooling system is a water slide, which uses a typical strand die. The strands lay down on a stainless steel incline slide containing water. At the end of the slide, the strands with water go into the pelletizer to be cut. Pellets and water from the cutting chamber are transported to a dewatering unit and centrifugal-type dryer for drying. Figure 44.9 shows the water slide pelletizer process.

44.2 Operation

Previous information on process and equipment discussed downstream equipment used in compounding. Compounding is the mixing of two or more components

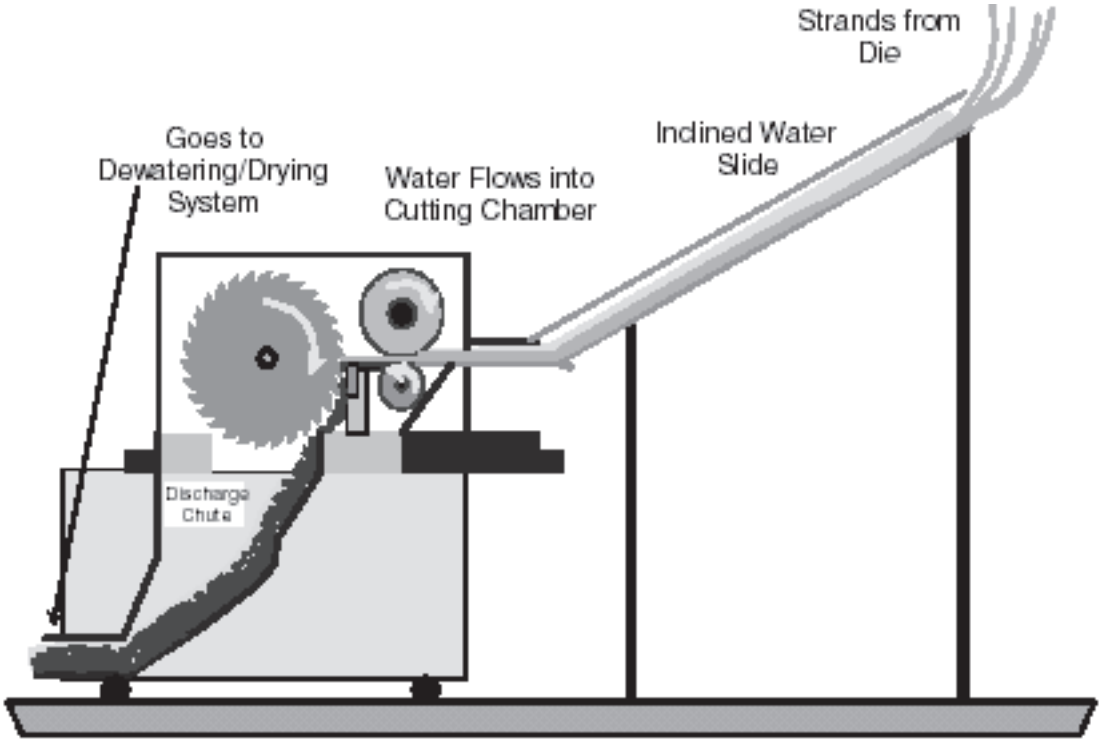

Figure 44.9. Water slide pelletizer.

in an extruder to produce a new product. The equipment discussed to this point explains options available to handle the product once it exits the extruder. Compounding takes virgin resins containing thermal stabilizers for processability and upgrades the resin by adding other ingredients. The ingredients usually melt or become plasticated in the extruder and mix as two or more viscous liquids. Sometimes the polymer melts and other ingredients remain in their solid state to be dispersed or distributed in the melt. Physical properties are determined by the particle size, distribution in the continuous phase, and adhesion to the polymer matrix. When blending two or more polymers in the melt state, one polymer becomes the continuous phase and the other polymers the noncontinuous or disperse phase. The polymer present in the highest percentage is not always the continuous phase. The compound properties depend on which polymer is the continuous phase and the dispersion and distribution of the noncontinuous phase.

Compounding operations are used for the following extrusion applications:

- Blending two or more resins together in the melt state.
- Alloying two or more polymer resins together in the melt state. (An alloy is a new material with one T_g, where a blend is a mechanical mixture of resins with at least two or possibly more T_g's, depending on how may resins were mixed.)
- Adding impact modifier to a resin to make it tougher.
- Adding flame retardant additives to improve flame resistance.
- Adding stabilizers to meet the following special requirements:
  - — improved UV stability
  - — improved thermal stability in processing
  - — improved long-term heat aging
  - — improved oxidative stability
- Adding release agents to provide either internal or external release characteristics.
- Adding mineral fillers or extenders to reduce the cost.
- Adding colorants, blowing agents, heat stabilizers, or other additives to produce a concentrate that can be let down in secondary operations to provide product benefits.
- Adding reinforcements to improve the strength and toughness, such as glass, carbon, and aramid fibers.
- Adding conductive fillers to provide electromagnetic properties.
- Adding natural fillers and fibers to thermoplastic matrices.
- Adding plasticizers to change the flow characteristics or stiffness.
- Reactive compounding, where additives are chemically bonded to the polymer backbone to make products like maleic anhydride grafted PP or nylon.
- Incorporating nanoparticles to improve barrier or stiffness.

Compounding different ingredients with polymers has produced customized formulations to meet the needs of particular applications. Currently there are more than 27,000 different polymer formulations available to meet the needs of different applications, and the list is growing to meet new, more stringent applications.

REFERENCES AND SUPPLIERS

1. Conair Jetro, "Plastic Strand Pelletizers," 400 Harry S. Truman Highway, Bay City, MI.
2. BKG Bruckmann & Kreyenborg, Granuliertechnik GmbH, Hessenweg 3, 48157 Münster/Germany.
3. Beringer Division of John Brown Plastic Machinery, "Water Ring Pelletizing Systems, P. O. Box 485, Marblehead, MA.

Review Questions

1. What are 4 different ways to pelletize polymeric formulations?
2. In a strand process, what is the most critical step for good strand and pellet quality?
3. How is water introduced and removed from a water bath and why?

Review Questions (continued)

4. How much heat has to be removed when processing PP at 700 pounds per hour with a melt temperature of 475°F if the water temperature entering the pelletizer is 128°F? The heat capacity of PP is 0.478 BTU/lb °F and the ΔH_{fusion} = 43.9 BTU/lb.
5. What is the difference between an air knife and air stripper and what is the function of each piece of equipment?
6. What is the advantage of underwater pelletizing?
7. What is the purpose of a diverter valve in underwater pelletizing?
8. What is the purpose of the upper and lower feed rolls in a strand pelletizer?
9. Explain the cutting mechanism in a strand pelletizer?
10. What are the two types of classifiers; how do they differ?
11. What is the function of a classifier in a compounding line?
12. What is the difference between underwater pelletizing and water ring pelletizing?
13. How does a water slide work?
14. What is the function of a compounding line? What are some materials that are produced?

45 Sheet and Cast Film

The convention between sheet and film is that film is less than 0.010 inch (0.254 mm) thick and sheet is greater than 0.010 inch (0.254 mm) thick. The key components in a sheet or film line are the die, roll stack or cooling rolls, puller or nip rolls, and windup or sheet stacking system. Cast film is made using either one or multiple cooling rolls, while sheet is typically produced using a multiple roll stack. Depending on the thickness, film is wound in continuous rolls and sheet is cut to length and stacked. Figure 45.1 shows a three-roll stack used to cool sheet and film. The roll stack, puller roll, and winder are connected together in a common unit. A motorized drive moves the entire unit toward or away from the extruder die on rails to guarantee the roll stack and winder are perfectly parallel to the die opening. The unit is moved away from the die for ease of cleaning and die adjustment prior to running and moved into close proximity with the die during operation. In operation, the distance between the die exit and the entrance to the three-roll nip is adjusted to maximize the processability. Under most operations, the die is situated as close to the entrance of the three-roll stack as physically possible. A positive stop is employed to guarantee the unit cannot make contact with the die as the roll stack is being moved toward the die. Contact between the die and rolls can result in serious damage to the highly polished chrome rolls in the three-roll stack.

Sheet operations are used to make glazing for doors, windows, bullet proof sheet, protective sheet such as floor mats around desks, and so forth. There are many sheet applications in the construction industry other than glazing, such as protective covers over walkways and graffiti-proof sheet; it is used as sound barriers along highways, in lawn and garden, and other commercial and industrial applications. Large volumes are used in other plastic processes, such as thermoforming to produce signs, appliance parts such as refrigerator liners, plastic food containers, and many other products; decorative panels where film, fabric, or other materials are laminated to one face of the sheet; and lamination applications where the sheet is melted to provide a uniform layer of resin.

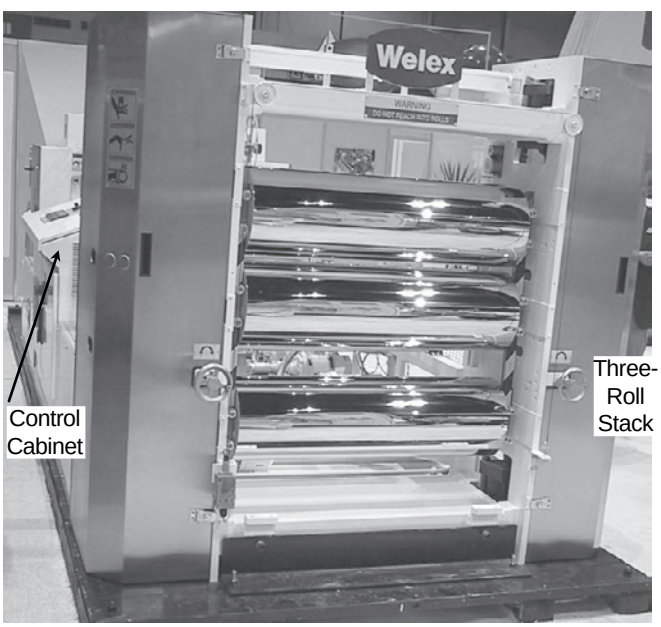

Figure 45.1. Welex, Inc. sheet/film three-roll stack.

Cast film is used in packaging, food wrap, substrate for coating, protective film, agricultural film for weed control, general purpose polyethylene film as a protective barrier to prevent scratching of parts during shipment, and many other applications. This chapter discusses the equipment required to produce both sheet and cast film and its operation. Sheet and cast film are discussed separately because the equipment and operation can be slightly different, depending on the product thickness.

45.1 Sheet

A sheet extrusion line is shown in Fig. 45.2. The key components are

- Die
- Three-roll stack
- Conveyor to support sheet
- An edge timing device
- Two sets of nip or puller rolls
- A winder or stacker

The line speed is synchronized to prevent the pull rolls from drawing the sheet away from the three-roll stack too rapidly, causing excessive molecular orientation between the stack and puller rolls, or excessive tension between the puller rolls and windup. A turret-type winder is shown, which automatically indexes the product to the next roll when one roll is complete. Each piece of equipment and its operation is discussed below. Troubleshooting sheet processes was presented in Part 4, Chapter 29.

45.1.1 Die

The sheet die should uniformly distribute the polymer melt across the entire die width. This is to provide a uniform melt thickness and polymer flow to the three-roll stack. Die widths typically range from approximately 24–120 inches (0.6–3 meters). Computer-aided design programs have moved die design from art to science. These programs provide good die designs for different polymers and rheologies. Uniform polymer flow and

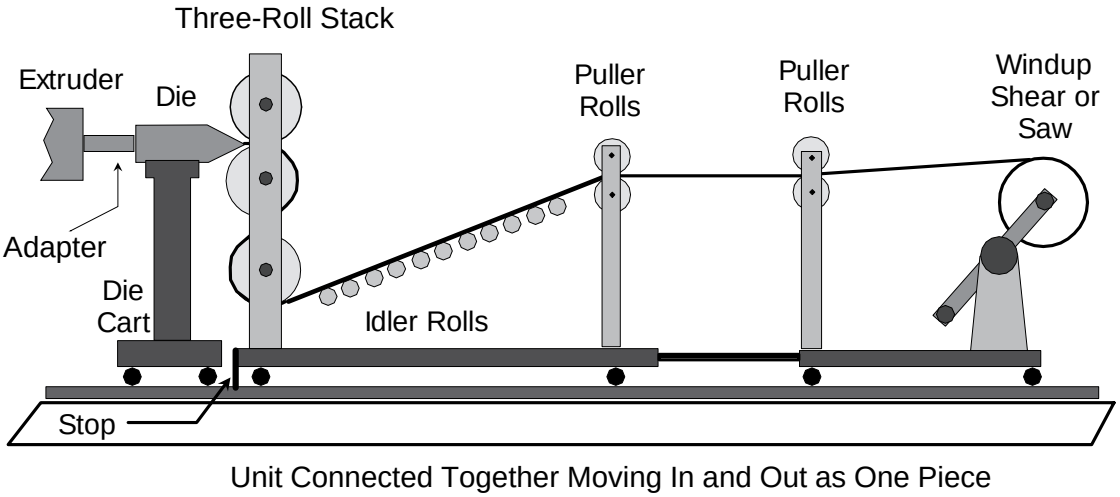

Figure 45.2. Schematic of a sheet line with a down acting roll stack.

thickness from the die is a critical step in guaranteeing uniform sheet thickness in both the machine and transverse directions. Three common die designs are

- Coat hanger die
- T-slot die
- Fishtail die

The most popular design for sheet is the coat hanger die. These designs are shown in Figure 45.3. The polymer enters the die in a circular or racetrack (rectangular with curved corners) channel that is normally on the die back. As sheet dies are large, they require their own die stand to support their weight. A melt or transfer pipe connects the die to the extruder. The transfer pipe inside diameter is smaller than the extruder diameter and closely matches the die entrance hole diameter. The pipe cross sectional area is reduced to maintain the polymer flow velocity and prevent dead spots where polymer can build up over time and degrade. Band heaters are uniformly spaced along the transfer pipe, covering as large a surface area as physically possible. Each transfer or melt pipe has its own temperature controller and thermocouple feedback loop. Unlike the extruder heating zones, no cooling is provided to remove excess heat. The heaters are designed to increase, decrease, or maintain the polymer temperature based on conductive heating, as there is minimal shear heat generation as the molten polymer flows through the adapters and melt pipes. Polymer flow in transfer or melt pipes is non-Newtonian laminar (shown in Fig. 45.4), with the velocity at the pipe center higher than at the edge. The flow velocity profile across the channel depends on the polymer rheology and the shear rate. Figure 45.4 shows the transfer or melt pipe connected to the extruder; the other end is connected to the die. In twin screw extrusion, the adapter from the barrel to the transfer pipe is an 8–O transition, with the end connected to the extruder

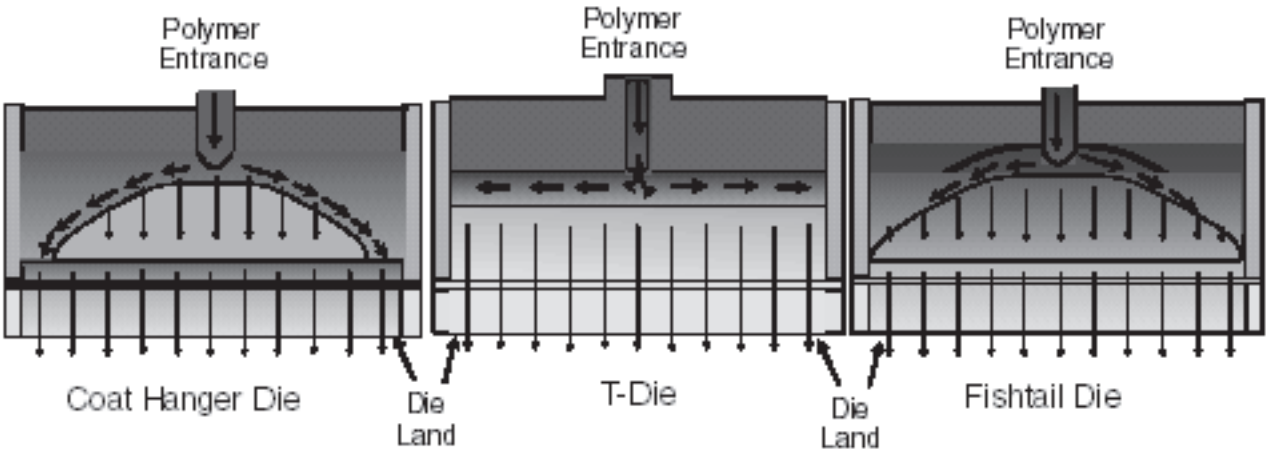

Figure 45.3. Top view of different sheet dies.

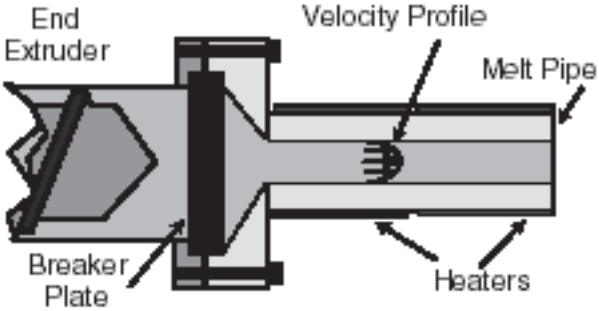

Figure 45.4. Polymer velocity flow in the transfer pipe.

having a figure 8 configuration and the end connected to the transition pipe having a circular configuration.

Figure 45.5 shows a sheet die on a die stand. The flexible upper die lip is used to adjust the die gap with the die lip adjustment bolts. The larger bolts above the die lip adjustment bolts are attached to the choker or restrictor bar. The choker bar is a large internal bar that can be raised or lowered across the die to assist in controlling the polymer flow. When the choker bar is fully retracted and the die gap is uniformly set, there should be heavy center flow at the standard line speed. The choker bar should be closed uniformly until the exit flow is reasonable uniform. Then fine adjustments are made with the adjustable die lip. If there is heavy edge flow when the choker bar is fully retracted or heavy center flow when the choker bar is fully inserted, then there is a die design problem and a new die or different polymer is needed. When adjusting either the restrictor bar or the die lips, small changes are usually sufficient to bring the sheet dimensions into control. Large movement at any given time of either the choker bar or die lips can cause major flow changes across the die. As the flow is restricted in one area, it redistributes differently across the die as there is no change in extruder throughput. Likewise, a flow increase in one area by opening the die lips or choker bar is accompanied by decreases in other areas across the sheet. If the gauge is fairly uniform across the sheet, small die gap adjustments will normally bring the gauge into specifications. If large die gap adjustments are required because the gauge uniformity is completely skewed, it is sometimes more expedient to shut the line down, withdraw the three-roll stack, shut the extruder off, and regap the die using a feeler gauge to obtain a uniform die opening before restarting.

The three dies shown in Fig. 45.3 have different resin distribution systems (manifolds) to promote uniform polymer flow at the die exit. In the coat hanger die, the channel curving along the back of the die like a coat hanger has a generally circular cross section and is connected directly to the die entrance, where polymer enters the die from the extruder. This circular channel decreases in depth as it moves from the polymer entrance from the extruder toward the front of the die. The area directly in front of the channel is referred to as the preland; this is a higher shear rate region than the circular channel depth and flat, partially restricting the polymer flow as it moves toward the die lip land area. The purpose of the circular channel is to evenly distribute polymer entering the die from the extruder to the die lips. The channel volume is designed by computer to ensure a uniform distribution of polymer melt. Progressively decreasing the channel size as it moves from the die entrance to toward the die lips helps to prevent the polymer from flowing heavily along the back channel and exit at the edge of the die. The choker or restrictor bar is raised or lowered into the polymer melt as it flows across the land in front of the circular channel. Just prior to entering the die lip area, a small groove in the bottom of the die is matched to a half-round piece protruding from the top of the die. This provides a secondary polymer distribution manifold prior to polymer entering the primary die land area. The length of the die land depends on the thickness of the product being produced. Higher thickness requires longer land lengths with sheet gauge above 0.100 inch (2.50 mm), needing a land length > 2 inches (51 mm). At the other extreme, sheet between 0.010 and 0.030 inch thickness (0.25 to 0.75 mm) requires a die land length of 0.50 inch (13 mm). All these factors combine to enhance the polymer flow uniformity across the die to ensure polymer flow uniformity exiting the die. Finally the die lips are gapped and adjusted in operation to produce the correct thickness product. Variations in channel

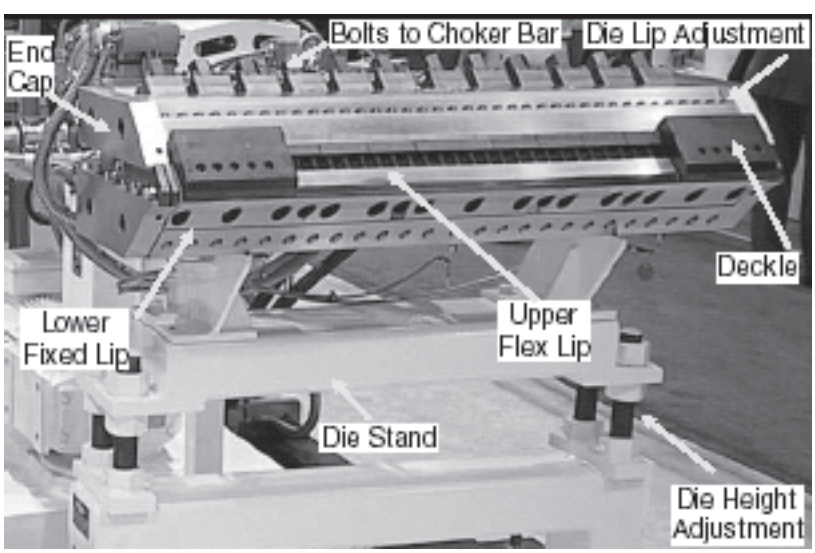

Figure 45.5. Welex, Inc. sheet die.

depths by adjusting the choker bar and die lip across the die are designed to provide uniform polymer flow exiting the die; otherwise the sheet gauge will be nonuniform.

The T-die shown in Fig. 45.3 is similar to the coat hanger die, except the circular channel connected to the die inlet to distribute the polymer uniformly goes straight across the back of the die instead of being tapered toward the front of the die. The land in front of the distribution channel goes completely across the die body. Unlike the coat hanger die, the channel has uniform length and height across the back of the die. The third type of die shown in Fig. 45.3 is called a fishtail die. This is very similar to the coat hanger die, except the distribution channel is not circular and completely disappears near the front of the die manifold into the land area.

Figure 45.6 shows a die cross section. The channel in the back is used to evenly distribute the polymer flow. In practice, this is assisted with the choker bar that can be closed or opened to provide a coarse flow adjustment. Due to improvements in die design calculations, many modern dies are designed without choker bars. The last adjustment is the flex lip bolts that adjust the die gap. The bottom lip is fixed; a specific range of sheet gauges or thicknesses are possible with a specific bottom die lip and flexible upper lip. To change gauge thickness outside this range, the bottom lip is changed to make the neutral gap wider or narrower. The gauges possible with a given bottom lip depend on flex lip adjustability. An alternative to flex lip dies is rigid or fixed lip dies designed for one thickness. Regardless, the lips have to be in good condition and provide a uniform gap when properly set with no product present. Warped die lips do not provide adequate gauge control and have to be machined flat or replaced.

Typical die body materials are prehardened P20 tool steel with a Rockwell hardness of RC 36–39. If harder material is required, 15/5 stainless steel can be used with a Rockwell hardness of RC 38–42. Factors determining the steel to use include strength versus flexibility for the die lip and metal purity for finishing. Dies are hard chrome plated to a thickness of 0.0010–0.0015 inch (0.025–0.038 mm or 25–38 microns). Chrome plating provides a very smooth, durable, hard surface that is not easily scratched during cleaning, while providing a good wear surface with low friction and high corrosion resistance during processing. Today there are many alternatives. Die surface finishes such as nickel plating, nitriding, and polymer coating depend on the application. These improve die pressure drop and die lip cleanliness.

The temperature zones across a die depend on the die width. For narrow dies, up to 12 inches for one temperature zone is quite common. Between 24 and 45 inches for three temperature zones is common, with one zone in the center and one on each edge. As the die gets progressively wider, the number of temperature zones increases. The purpose of the temperature zone is to provide uniform temperature or, more important, uniform polymer melt viscosity across the die. While restrictor or choker bars and die lip adjustments are used to create uniform flow rates across the die, temperature can also be a factor in controlling the polymer rheology. Higher melt temperature reduces the polymer melt viscosity, increasing the polymer flow. As an example, consider the flow profile across a 60-inch wide die (shown in Fig. 45.7). Flow from the die is slower in zone 1 compared to zones 3, 4, and 5. In operation this is observed by allowing the sheet to flow out the die to the floor. A line scribed across the sheet as the material exits the die will show the line in zones 3, 4, and 5 traveling faster away from the die than material from zones 1 and 2. If the zone temperatures are set the same, this phenomenon can be caused by cold air from a fan in the summertime or a door opened in the wintertime, cooling the zone 1 side. Assuming the restrictor bar and die gap are properly set, corrective action is to increase zones 1 and 2 temperatures, with a larger temperature increase in zone 1. The lower the resin viscosity in zone 1, the higher the flow rate. In the example, zone 4 flow rate is slightly less than zones 3 and 5. This can be corrected by increasing the temperature of

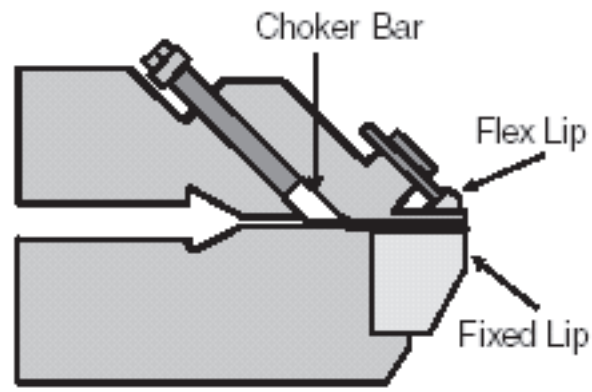

Figure 45.6. Sheet die schematic, side view.

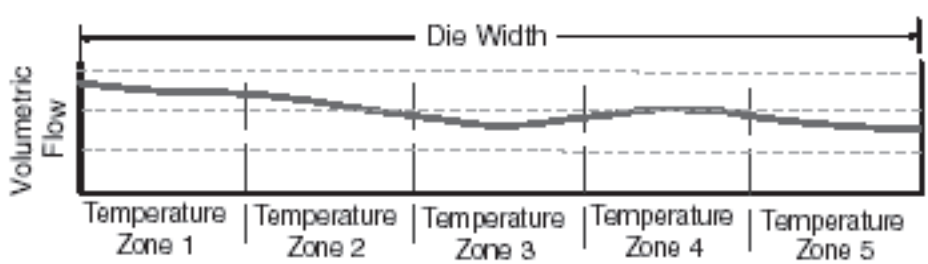

Figure 45.7. Example of volumetric flow out of the die.

zone 4 or through slight adjustments with the choker bar. An alternative approach to this example is to maintain the temperature in zone 1 and lower the other temperatures to reduce the flow in the other zones. Which approach to use depends on the

- Resin temperature sensitivity
- Melt temperature across the polymer web
- Die pressure
- Throughput rate

Good die design provides uniform pressure drop across the die body. Computer-aided design needs the proper rheological data and process conditions to generate an effective design. Equation (45.1) is the die pressure drop.

$$\Delta P = Q_v \cdot R \cdot \eta \tag{45.1}$$

where

ΔP = Pressure drop
Q_v = Volumetric flow rate
R = Flow resistance
η = Polymer viscosity

Figure 45.8 shows the goal for pressure drop and volumetric flow across the die face.[1] The pressure drop in the primary land section is calculated using Eq. (45.2).

$$\Delta P = \frac{12\eta \times Q_v \times L}{W \times H^3} \tag{45.2}$$

where

L = Channel length
W = Slot width
H = Channel height

Figure 45.9 shows a typical pressure drop curve[1] through the different areas of the die. The boundary conditions dictate that the entrance pressure and exit pressures are the same for all flow paths in the die. If the die design is good, the pressure drop in the manifold or coat hanger section is balanced with the pressure drop in the preland, so the exit flow to the final land is uniform. If the die geometry is poor, the flow will distribute itself so that there will be an imbalanced flow.

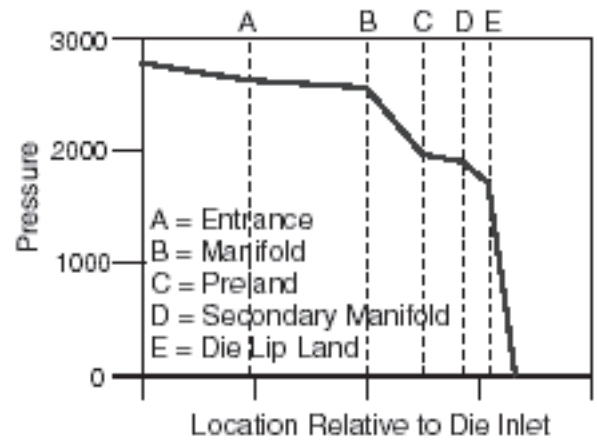

Figure 45.9. Typical pressure drop through a die.

Automatic dies are available to adjust the die lip opening through heated bolts. Initially the die gap is set close to a uniform gap. The gauge profile across the sheet is monitored on-line, and a signal fed back through a microprocessor indicates whether to heat or cool the die adjustment bolts. Cartridge heaters next to the bolts heat the bolts, allowing them to expand and close the die gap. If the gauge is too thin, cooling air cools the die lip adjustment bolt, causing it to shrink and open the die gap.

To change sheet width, deckles, as shown in Fig. 45.5, are installed. Molten polymer exiting a sheet or film die necks down along the edge, as shown in Fig. 45.10 This is caused by the roll stack pulling the polymer away from the die. The result is reduced sheet width compared to the die width. Consequently the die has to be wider than the final width desired. As the edge quality is not equivalent to the rest of the sheet, edges are normally removed in a trimming operation. Consequently with a 30 inch (762 mm) wide die, the maximum sheet width possible will range from 26 to 28 inches (660–711 mm). Using the same die to produce a 20 inch (508 mm) wide sheet generates significant waste and lost production, assuming seven inches are

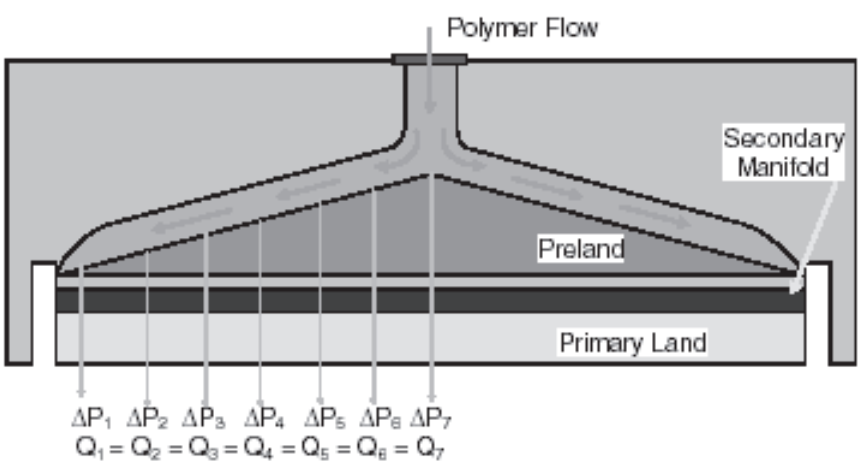

Figure 45.8. Pressure drops associated with polymer flow through the die produce equivalent flow.

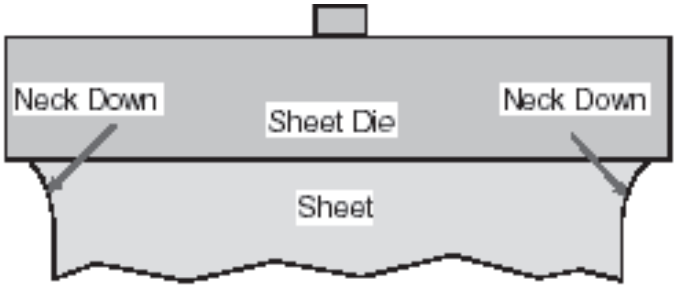

Figure 45.10. Top view of sheet showing neck down along edges.

removed and either recycled or discarded (assumes 3.5 inches [89 mm] trim per side and 27 inch [686 mm] sheet entering the three-roll stack). To minimize the edge trim and waste, deckles are used to reduce the sheet width. To prevent leaks at the deckles, they must be properly installed. The downside of using deckles is dead spots behind the deckles in the land area and polymer stagnation in the polymer flow channels, where material can degrade in the die. Deckles work well with thermally stable resin systems. Use of deckles with rigid PVC is discouraged, due to the poor thermal stability of PVC.

If a die has been taken apart for cleaning and reassembled, the bolts need to be properly torqued after the die has been brought up to temperature and allowed to heat soak prior to start-up. Proper torquing after bringing the die to operation temperature is required to prevent polymer leakage during operation.

45.1.2 Three-Roll Stack

The die is designed to fit between the rolls in a three-roll stack and lay the sheet down on the roll just prior to the nip point. The three-roll stack, shown in Fig. 45.1, is wider than the die so the die, including the end-caps, can fit. Roll diameters generally range between 8 and 16 inches (203–406 mm). Roll stack designs can be simple up or down stacks to configurable stacks that can be changed to control the entry and exit angle and the total wrap. Higher wrap provides more cooling as the rolls have more contact time to remove heat. The rolls control the

- Cooling rate
- Final thickness
- Surface finish

Figure 45.11 shows the die in front of the three-roll stack and sheet passing over the bottom roll in a down stack. The nip rolls are open and the roll stack pulled back from the die in the view shown.

An up stack has the sheet passing between rolls one and two, making an S wrap around rolls two and three, and exiting the stack over the top roll. Conversely a down stack feeds the sheet between rolls two and three, makes an S wrap around rolls two and one, and exits under roll one, as shown in Fig. 45.12 A and G. Fig 45.12 B, C, D, E, F, H, I and J show variations on the downstack system where the rolls are not vertical or in a straight line. The middle roll position is fixed with both the top and the bottom rolls moving up and down for cleaning and stringing up the stack. The pressure applied between each roll is given in pounds/linear inch or PLI. The force applied by the hydraulic cylinders is the nip contact pressure times the sheet width in contact with the rolls. In operation, an up stack with the sheet coming off the top roll has the idler rolls between the top roll and the puller slanting down. Cleaning the rolls or looking at the sheet quality is done underneath the rolls. In a down stack operation, the idler rolls are slanted up to the puller rolls (shown in Fig. 45.2) with a bridge across the idler rolls to clean the rolls or evaluate the sheet quality. The advantage of an up stack is the bottom roll can be cleaned in operation without concern that the cleaning material will fall onto the product, causing contamination. Using a down stack, as shown in Fig. 45.2, the top nip is usually closed, and the bottom nip between rolls one and two may be opened or closed, depending on the operation, film surface requirements, and gauge control.

Figure 45.12 shows some of the possible roll configurations and wrap angles possible with various roll setups. The wrap depends on the product thickness and the heat to be removed. Configurations A and G are typical down and up stack configurations with three vertical rolls. Configurations G and J have minimum wrap angles and are most commonly used for thicker sheet products. Configurations C, D, and I have large wrap angles on two rolls, providing good gloss and finish on both sides of the

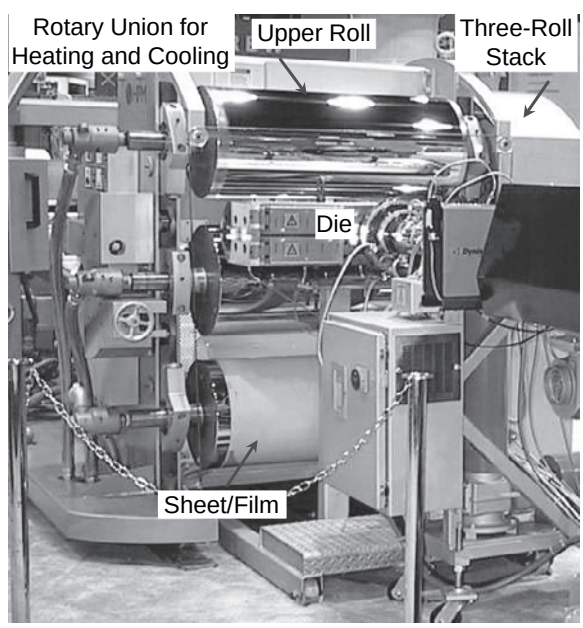

Figure 45.11. Three-roll stack operating in a downward configuration.

sheet. To obtain high gloss and surface smoothness, the rolls are run as hot as possible without allowing the product to stick to the surface. Optimum roll temperatures for surface quality are approximately 10°F (5.6°C) below when the polymer first starts to stick to the roll surface. To determine this temperature in operation, increase the roll temperature until the polymer just starts to stick and then back off 10°F (5.6°C). Using configuration F as an example, the top is polished with roll number two and the bottom with roll number three. Roll number one helps to set the gauge and initiates the surface skinning on the bottom and initiates the surface quality. If the top is to have a textured appearance and the bottom smooth, roll number two would have a matte or textured surface, with rolls one and three being polished.

Water or oil circulates through the rolls and controls the temperature. Temperature uniformity across the roll surface is critical to produce warp-free sheet with uniform surface aesthetics and release from the roll. Differential side-to-side or edge-to-edge cooling can lead to warpage in cut and stacked sheets. A skin forms on the sheet surface as it is cooled. The wrap angle determines the skin thickness on both sides. If one side has a thicker skin and is cooled more rapidly than the other, the sheet can be expected to cup or bow toward the hotter side. If this happens, lower the roll temperature on the side where the bowing occurs. If surface gloss is not an issue, the roll temperatures can be run colder to remove heat faster. The rolls remove heat from the product. The hot oil or water heats the rolls and maintains temperature when no product is running. During operation, the polymer melt keeps the rolls hot, and the heat transfer fluid removes heat from the roll surface. The roll diameter and fluid flow rate are designed to remove heat and maintain a constant temperature across the entire roll surface.

Before start-up, the roll gaps between the middle stationary roll and the upper and lower rolls are set. The first gap is normally set approximately 5% less than the sheet gauge, and the second gap is set at the desired sheet thickness. Rolls are gapped before starting with a feeler gauge. The pressure is adjustable, with the pressure being applied to the ends of the rolls. At high PLI pressure, the rolls can bow in the middle, causing the middle of the sheet to be thicker than the edge. To compensate for bowing, roll manufacturers can produce crowned rolls that are bigger in the middle than on the edge. As pressure is applied and the rolls bow, a uniform gap is obtained across the entire roll width. There is also a ContraBend™ roll where the journal design allows the roll to bend opposite the way a roll normally bends.

Nip roll pressure can be easily measured by passing Pressurex[2] nip-impression film through the nip rolls. Stresses imparted to the film are read optically by a camera and translated with Windows-based software into a pressure profile across the nip rolls. The time required to obtain a steady-state pressure profile is minutes.

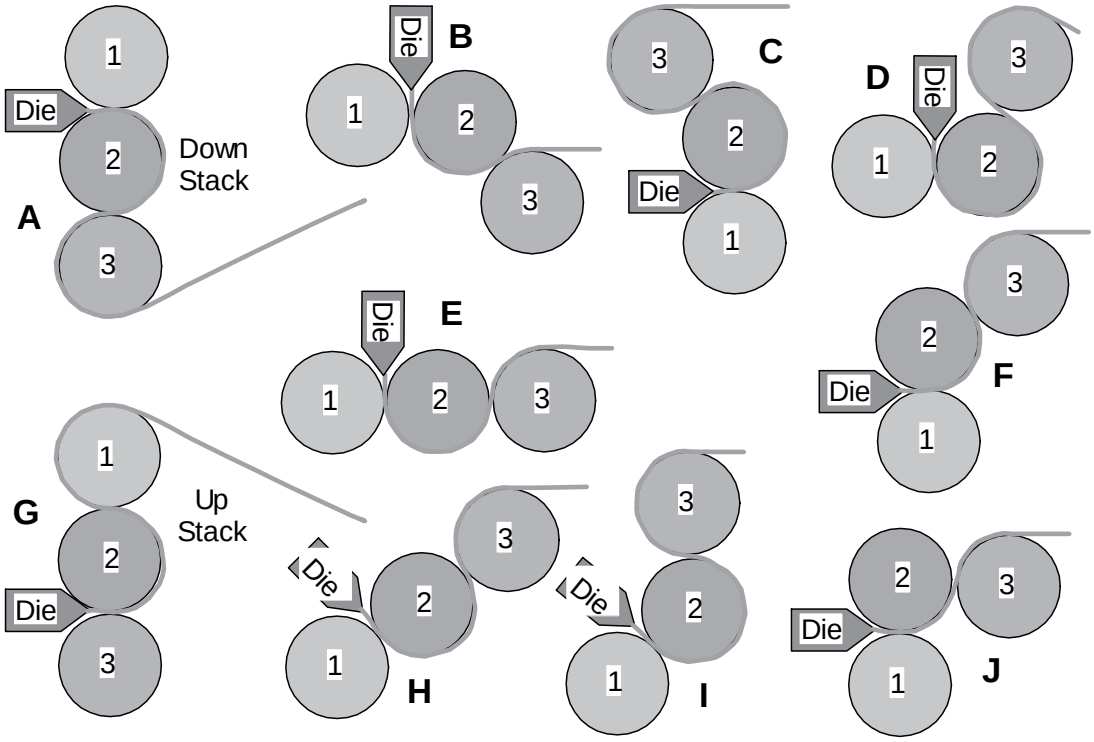

Figure 45.12. Potential three-roll configurations for sheet fabrication.

Roll stacks are driven by either one motor using a chain or gears, or through multiple motors with one for each roll. Chain drive with one motor is the most versatile system, easiest to maintain, and easiest to operate. It is used to produce most sheet or film rolls used in packaging and disposable sheet. Gear-driven rolls are used for thick and wide sheet products. It is more costly than the chain-driven rolls, all rolls have to be the same size, and it is harder to maintain. The advantages are the rolls can be used for calendaring and the ability to transfer high power. The third type of roll stack has motors on all the rolls. These are used for optical quality and thick sheet. They are the most expensive and hardest to run. Speeds have to be properly matched to produce quality sheet. Unlike the gear-driven systems, the rolls can vary in size. Motors on all the rolls produces the lowest chatter and marking.

45.1.3 Puller Rolls

Idler rolls support the sheet between the three-roll stack and the puller rolls, where it continues to air cool (see Fig. 45.2). The puller roll gap is normally set approximately 5% percent below the sheet thickness to provide uniform pressure and good pulling. The puller speed is synchronized with the roll stack speed to prevent any drawing or slippage on the roll stack or between the roll stack and the puller roll. Puller and roll stack speeds play a large role in determining the final product thickness. Consequently, good speed control with fine adjustment is required. If the product is pulled away from the die too rapidly, the sheet will have excessive necking and the gauge may be too thin. Conversely, if the product is pulled away from the die too slowly, the rolling bead (discussed later) may become too large and cause the process to break down; or, in extreme cases, the roll stack does not remove the polymer fast enough, causing it to wrap the rolls and build up on the die. The puller speed has to be matched to the die opening and the extruder throughput to yield the correct product gauge, width, and molecular orientation. Puller speeds have to be constant with no sheet slippage; otherwise gauge variations in the final product will result.

Puller rolls are usually rubber or elastomeric coating, such as polyurethane, to provide positive pulling characteristics. As the roller surface becomes worn, the roller has to be resurfaced or machined smooth to provide uniform sheet contact.

Between the three-roll stack and the puller roll, knives cut the sheet to the proper width. Edge trim knives come in many forms, ranging from utility knife blades fixed in place on a holder to rotating knives above and below the sheet. Edge trim is removed and chopped or ground for either recycle or disposal.

45.1.4 Windup or Stacker

Sheet less than 0.03 inch (0.76 mm) thickness is normally wound in continuous rolls, similar to film. The winder (windup) handles large-diameter rolls at high speeds while pulling at constant web tension as the roll diameter increases. Turret winders are used to maintain a continuous operation. A turret winder has one or two winding positions to accumulate the finished product. In a dual winder system, when the first roll is complete, the winder automatically indexes the empty core to start another roll in a continuous operation. During winding, the completed roll is removed and an empty core placed back into the winder. The cores can be held by either mandrels or core chucks. Mandrels are typically air actuated to prevent the cardboard tube, forming the center of the sheet roll, from rotating or moving on the mandrel during winding. After the mandrel is inserted into the tube, air pressure applied at the end of the mandrel expands areas of the mandrel, preventing tube rotation or slippage. To remove the mandrel, air is released, relieving the tension on the inside of the tube. Below speeds of 1000 feet/minute (305 meters/minute), steel mandrels are used, while composite air shafts are used at higher speeds. Core chucks can be used with metal cores.

Three winder designs are classified as center, surface, and center/surface. Figure 45.13 shows the difference between center wind and surface wind operations. Center/surface winders are used mainly with stretch film and slippery materials. Sheet is typically wound using

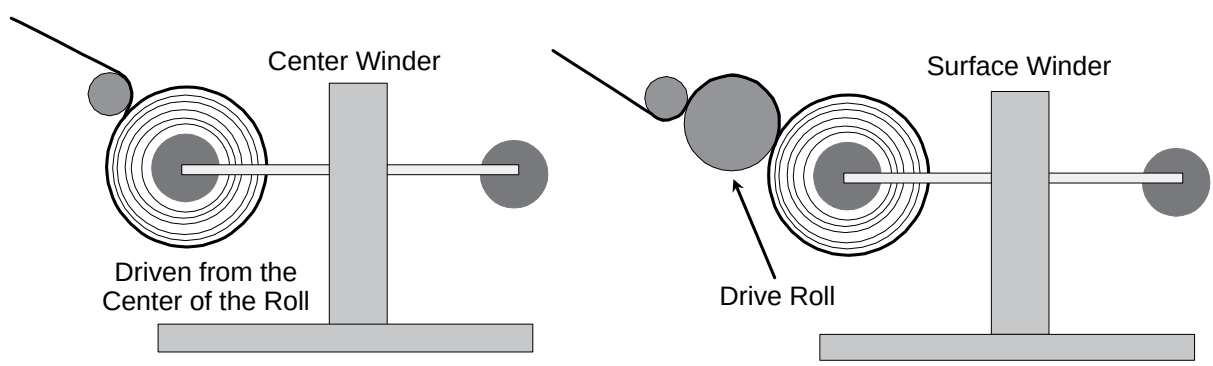

Figure 45.13. Comparison of center and surface winders.

either surface winders or center winders. Surfaces winders are simpler technology, used with large-diameter reels or rolls and broad speed ranges. A driven roll in contact with the surface runs at constant speed to wind the roll. As the roll or reel becomes larger, the surface speed remains constant and the revolutions per minute (rpm) decrease. If winders run at constant rpm, the surface speed increases substantially as the reel becomes larger. Obviously the surface speed cannot change without a change in product dimensions, assuming a constant throughput rate. Center winders work based on the center torque providing constant web tension. The winder torque changes as a function of roll diameter. Decreasing the torque as the roll gets larger produces a firm nontelescoping roll that handles well after it is removed from the line.

For sheet thicker than 0.30 inch (0.76 mm), winding is not an option. The sheet may be too stiff or winding may impart a set that is difficult to remove later. A stacker coupled with a guillotine or shear cutter cuts the sheet to length. Based on the line speed, the guillotine or shear is activated in a time sequence that momentarily stops the product and cuts it. If the shear or guillotine moves with the product during the cutting operation and then back to its original location, the sheet continues to lie flat and run smoothly. If the guillotine is in a fixed location and clamps the sheet, a buckle in the product occurs between the puller rolls and the shear. To prevent this buckling activity from traveling back to the roll stack, significant distance is required between the pull roll and the shear or guillotine. In most situations, the shear cutter moves with the sheet to produce a smooth operation.

Once the product is sheared, it is stacked. Stacking can be done by picking up the product with suction cups and moving it to another location. In other instances, a table is installed directly after the guillotine that allows the sheet to slide onto the table, with the shearing action taking place immediately prior to the final location of the sheet. As the number of sheets and weight of the product increases, the table automatically lowers, keeping the product stacked and neat. A potential problem with this operation is scratching the sheet as the next sheet slides over it.

Another cut-to-length option is a moving saw. The sheet is clamped in place, with the clamp moving at the speed of the sheet line. As the sheet moves, the cutting saw transverses the sheet, cutting it in a straight line between the clamped locations as the saw moves with the sheet in the machine direction. Moving saws provide a clean cut and flat edge. After the cut is complete, the clamps release and the saw and clamps return to their original positions in preparation for the next cut. The saw travel distance is determined by the sheet line speed. Higher line speeds require longer machine direction travel or faster cutting speeds for the cutting operation to be completed.

45.1.5 Thickness Measurement

On-line thickness gauges provide gauge profile, average thickness, and specific thickness at various locations across the sheet. In addition to thickness, these gauges can also determine the basis weight (ounces/foot^2 or grams/meter^2), sheet density, color, and opacity. Measurements of basis weight can determine the weight per unit area of the entire sheet or specific layers in coextruded sheet. Options for measurement go from a traversing, noncontact source and detector to mechanical offline measurements at specific locations across the sheet. Other options are to use stationary sensors at various fixed locations across the sheet or a portable device that can measure either the machine or transverse direction profile. Typical sheet thickness variations are caused by the following, among other factors:

- Extruder surging
- Melt instability
- Line speed variations
- Cold or hot air drafts
- Incorrect nip roll setting
- Incorrect rolling bead or bank
- Uneven die flow

On-line thickness measurements make thickness variation detection and control much easier.

The advantages of stationary sensors are[3]

- They cost less than continuous scanning measurement systems.
- Two measurement points may be sufficient to monitor the process and make necessary corrections.
- They are useful in determining machine direction variations.
- They are easier to maintain and support than a continuous system.

The advantages of continuous scanning measurement systems are

- They provide transverse and machine direction profiles and averages.
- They can scan wide sheet or film in a timely manner.
- A feedback signal facilitates automatic gauge adjustment.
- They can be moved out of the web path during start-up and calibration.

Most continuous, noncontact gauges assume a constant sheet density. Whether a single source or scanning device is purchased depends on the cost/performance ratio required for the application. Sensor types are based

on gamma/x-ray for single-point measurements in heavy gauge sheet. Infrared is used in specialized applications. Ultraviolet (UV), ultra-sound, collimated light reflectance, magnetic resonance, inductance, capacitance, air pressure, or beta gauges are for scanning applications. Regardless of the measurement technique used, the technique must provide both accuracy and repeatability. Accuracy is how close the measurement is to the true value. Repeatability is producing the same value if the same sample is measured many times. Changes in sheet characteristics such as composition, thickness, basis weight, density, temperature, and opacity can affect both the repeatability and accuracy. Environmental factors can have a pronounced effect on the measurement.

In scanning operations, sources and detectors are mounted on an "O" or "C" frame and constantly transverse the sheet as they measure weight per unit area. This is converted to thickness using the sheet density. Figure 45.14 shows a source and detector on an O-frame. A C-frame is similar, except it is open on one side. In backscattering gauges, such as a gamma backscatter gauge, the gamma ray source and detector are in the same instrument on one side of the sheet. The detector is calibrated by measuring the gamma signal in the air, which establishes the zero point; a known thickness sample is used to establish the signal as gamma rays are applied to the film or sheet surface and the detector measures the gamma rays reflected. Thicker sheet reflects more gamma radiation. Connecting the signal to a recorder provides real-time data.

Unlike a backscatter gamma gauge, a beta thickness gauge has a source on one side of the web and a detector on the other, as shown in Fig. 45.14. This can be in a stationary position; however, in operation it is better to have it mounted to a moving platform that scans the web to provide real-time web thickness data. Beta particles are fast moving electrons emitted from a radioactive source. Beta gauges are not very sensitive to the material formulation or sheet composition. These rays do not travel far in the air and are relatively safe as long as operators are not exposed to them in large doses. Promethium 147 is a suitable beta source for basis weight up to 275 grams/meter^2. Krypton 85 is used for basis weights between 150 and 1500 grams/meter^2, and strontium 90 for sheet with a basis weight of approximately 1000 to 8000 grams/meter^2. Read and follow all manufacturer's directions very closely when using beta, gamma, or x-ray sources and detectors. Thickness is determined from the particles passing through the sheet to the detector versus those absorbed. As sheet thickness increases, more radiation is absorbed by the sheet, with less passing through to the detector. Measurement precision is improved by increasing the source size, lengthening the measurement time, and reducing the air gap between the source and the detector. Beta gauges are excellent for resolving thin or thick streaks in the sheet and fast scan speeds.

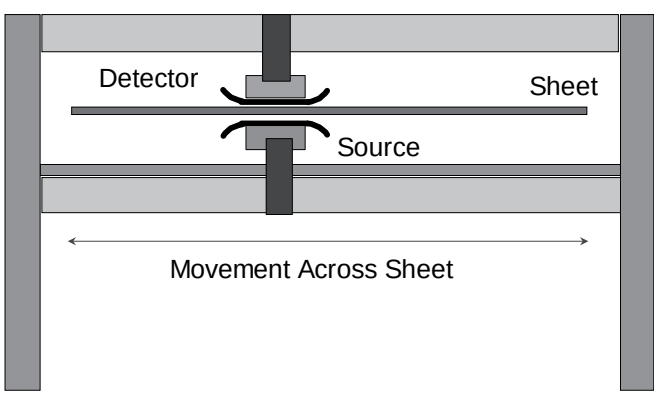

Figure 45.14. O-frame with scanning source and detector.

Gamma and x-ray instruments are used in thick sheet applications where beta gauges are not appropriate. Photon energy is used to make the measurement. Signal interference can be caused by additives or colorants in the formulation.[4]

O-frame (shown in Fig. 45.14) or C-frame scanners provide the current technology in continuous, high-speed measurement systems. The frame comes in widths to accommodate any size sheet. Frames are either I—beam or U-beam construction with roller bearings on the sensor and detector for easy movement. The sensors and detectors are small and durable instruments that move in tandem.

Another gauge is a nonnuclear, noncontact device that measures the polymer capacitance between a sensing element and a ground plate.[5] The polymer dielectric properties are constant and a function of the polymer. Recording the difference in voltage provides a way to determine thickness.

Infrared gauges can be used where radioactive instruments are not allowed or are difficult to use. Infrared works better for identifying composition, additives, and specific materials present than it does in determining thickness.

In addition to the previous techniques that measure basis weight and covert it to thickness, thickness can be measured directly with lasers, pneumatics, or optical interferometers. Each method has its pluses and minuses, with the technique used depending on the sheet thickness, material, transparency, surface smoothness, precision required, and line speed. Pneumatic gauges are difficult to use with foam applications where contact with the web may cause surface defects. Pneumatic thickness gauges do work with heavy gauge sheet at low line speeds. Lasers pass light through the sheet and measure the deflected beam as a function of the top side and bottom side of the web. Lasers do not work well on rough sheet and transparent materials such as polycarbonate

and acrylic. Optical interferometers are expensive and best suited for thin films.

45.1.6 Operation

As the extruder, die, and roll stack come up to temperature, the product standard operating procedure (SOP) or specifications are checked. Verify that the following equipment parameters are properly set before starting:

- Extruder temperatures, both setpoint and actual, and die temperatures are appropriate.
- Transfer pipe or adapter temperature, setpoint and actual, are appropriate.
- Make sure the screen pack is clean and has the proper mesh.
- Gauge the die lips and set if required. If the correct gap is not known, set the die lip openings approximately 20% thicker than the product to be extruded. Then close the gap to bring the gauge into control.
- Set the nip roll gaps.
- Set roll temperatures.
- Clean the roll surface.
- Remove any sheet or film left from a previous run.
- Prepare the line for stringing up the new product.

There are many ways to thread up the line. If the product is wound on a reel, unwind film back through the process so it hangs over the first nip roll in the three-roll stack. Molten polymer from the die that adheres to the film or sheet is slowly pulled through the process by the winder onto a scrap roll. After the process is running, the winder is indexed to collect product. Stringing up heavy gauge sheet can be done using cloth strung through the process and allowing the molten web exiting the die to bond to the cloth. After the product is being pulled by the nip rolls and is either stacked or wound on a reel, the line speed can be increased.

Operating a sheet line as well as other extrusion equipment requires matching takeoff speeds to the throughput rates. Uniform melt temperature and pressure provide a constant throughput rate and a quality melt to the roll stack. Die adjustments provide a uniform sheet thickness. At start-up, the molten polymer flows to the floor at a low rate. Scribing a line across the melt with a sharp object shows the flow uniformity exiting the die. When the scribed line uniformly flows to the floor, this indicates the polymer flow across the entire die is uniform. If part of the scribed line flows to the floor at a different rate than another part, either a choker bar or die lip adjustment is required to make the flow more uniform. Once uniform flow is attained, the roll stack is moved toward the die and the molten web is bonded to the string-up web or material going through the entire line. The three-roll stack is open until the new molten material passes through. After the new web is through the three-roll stack, closing the stack nip rolls pulls the sheet away from the die. Once the new sheet passes through the puller rolls, the rolls are closed and the product is either sheared or wound on a reel. Finally the edge trim knives are placed in the sheet to cut the sheet to the proper width. If the sheet is excessively wide, it may be necessary to start edge trimming before the product can be properly wound on the winder due to its width.

Verify that the distance between the die and the three-roll stack is correct. Gradually increase the extruder and roll stack speeds simultaneously. If the extruder increases faster than the roll, polymer build-up between the die and roll stack will either wrap a roll or flow back onto the die face. If the roll stack and takeoff equipment speed is increased too rapidly, the web between the die and the stack rolls can be drawn so thin it breaks. Consequently it is important to increase both the takeoff equipment and extruder throughput simultaneously while watching the process to ensure they stay synchronized.

After the process is up to speed, note the area between the die exit and the first nip in the roll stack. A rolling bead or bank of material (slight build-up of molten polymer) is required to have the roll nip area fill uniformly. Either the extruder screw speed or puller roll speed is increased or decreased slightly to adjust the molten polymer bank. Proper extruder output and puller roll speed balance will maintain a constant bank size. It is possible to view the rolling bead looking across the nip roll from the side. Opening and closing the nip rapidly allows the bead or bank to pass through the nip area. If the die is properly adjusted, the bead will be uniform in size across the entire sheet. If the bead or bank does not exist in some areas and there are other areas where there is too much material, the die lips are adjusted to provide a uniform bank. After the bead is uniform, a sheet sample is measured physically or an on-line thickness gauge is used to make further corrections. Figure 45.15 shows a rolling bank or bead.

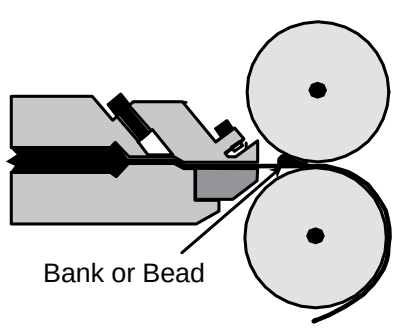

Figure 45.15. Rolling bank or bead.

The puller roll speed has to be properly set to produce acceptable product. If the die gap greatly exceeds the final sheet thickness, drawing or necking down to the final product thickness may generate too much molecular orientation. High machine direction orientation can lead to significantly higher machine direction properties compared to the transverse direction. In extreme situa-

tions, the product may become brittle and crack in the machine direction.

Troubleshooting guidelines for sheet are given in Part 4, Chapter 29.

45.2 Cast Film

Cast film applications include

- Food packaging
  - — Baked goods
  - — Candy
  - — Cheese
  - — Meat
- Medical applications to package disposable medical supplies
- Embossed and normal film used in diapers
- Labels
- Stretch wrap
- Boil-in bags for cooking
- Geomembranes

This market is rapidly expanding, particularly in packaging and stretch wrap. Multilayer films are used to preserve freshness while offering convenience.

Cast film is produced by depositing molten polymer directly from the die on a highly polished, high-speed chilled roll, where the film is cast. In less than one revolution, the chill roll solidifies the product as it draws it down to the correct thickness. Figure 45.16 shows a cast film line with two polished rolls, puller rolls, treatment center, slitter, and windup using a two-roll turret-type winder. Roll number 1 in Fig. 45.16 is the casting roll, and it is larger than roll number 2. Molten polymer, laid down across the roll surface by the film die, is rapidly quenched due to the roll temperature and the film thickness preventing or controlling crystallization of semicrystalline resins. This process provides the optical clarity required in packaging film produced using cast film technology with high gloss, high production rates, and good gauge control. In comparison to blown film, gauge variation in cast film with self-adjusting dies is ±1–2%, where blown film gauge is normally ±3–4%.

Components in the cast film extrusion process, as shown in Figure 45.16, are

- Extruder, either single or twin screw, equipped with a film die
- One or more casting rolls
- Polished chrome rolls for cooling and polishing the film
- Vacuum to remove air from behind the film to assist laydown on the casting roll
- Puller rolls to maintain constant tension on the film
- Treatment center for corona, flame, or plasma treatment of polyolefin films to improve adhesion for printing
- Slitter to cut the film to width
- Tensioning rolls to maintain tension through the slitter and treatment center
- High-speed winder to wind up the film

The different line sections will be discussed in the rest of this chapter.

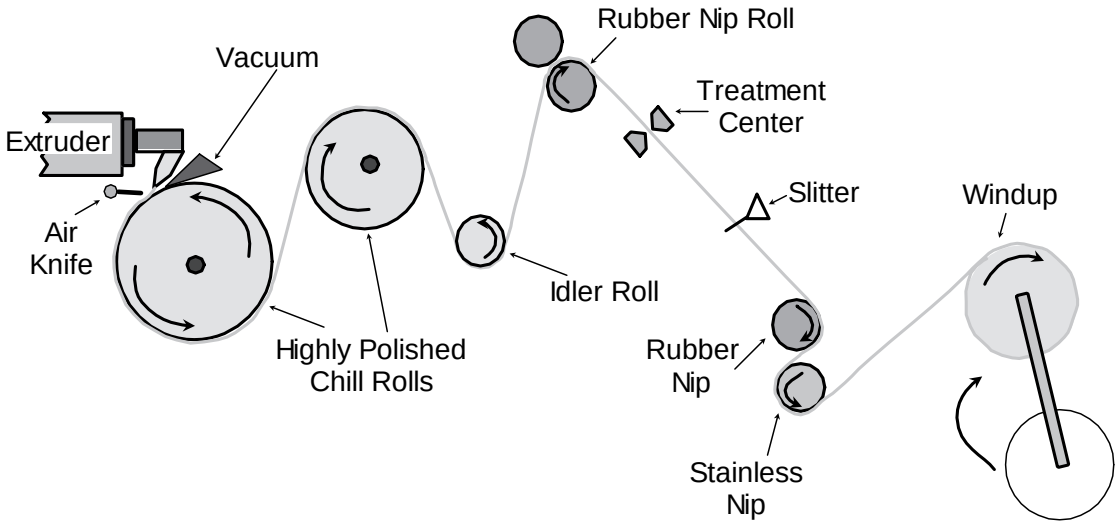

Figure 45.16. Schematic of cast film process.

45.2.1 Die

Cast film dies are similar to sheet dies, presented earlier in this chapter. The differences are

- The die lip on the bottom half of the die body is an integral part of the die, rather than a fixed lip that can be changed for different thickness.
- The slot die geometry does not have a restrictor or choker bar.
- The die opening is vertical, laying the extrudate directly onto the roll, rather than the horizontal configuration used with sheet dies and a roll stack.

The slanted part of the bottom of the fixed die lip is parallel to the roll surface in an attempt to place the molten polymer exiting the die as close to the roll surface as possible. Positioning the die in a vertical configuration prevents any sagging film from coming in contact with the bottom die lip, where it might stick and break the web. Film exiting a vertical die comes in direct contact with the roll surface, where it is pulled away from the die. The film thickness provides for almost immediate quenching as it contacts the roll surface, minimizing the crystallinity and opacity developed in a semicrystalline resin. A vacuum behind the die plus an air knife or air impingement jet on the top side of the film prevents air being between the roll and the film. This process ensures intimate contact between the roll and the film.

Like sheet dies, film dies are computer designed to provide uniform flow across the die width. There are two common designs. One is a coat hanger die, as shown in Fig. 45.8, designed for the polymer viscosity and throughput rate. The second is a long, transverse slot with a low pressure drop relative to the pressure drop across the land, similar to a T-slot die shown in Fig. 45.3.

Figure 45.17 shows a cross section of a cast film die with automatic die adjustment. Automatic die adjustment can be done electromechanically, hydraulically, or thermally. Thermal systems are the most common. A control signal from the gauging system heats or cools the die bolts to adjust the film gauge. Automatic systems will reduce the film gauge variation to less than one-half that associated with manual die bolt adjustments. Automatic die control can rapidly pay for itself with improved gauge control. Individual electric heaters are used on each die lip adjustment bolt to provide control. Insulation between the heater block and the die body prevents the heat from affecting the polymer viscosity. Some dies are made with internal deckling to change the film width. The die surface in contact with the molten polymer can be chrome, electroless nickel, polymer-impregnated chrome, or nickel plate.

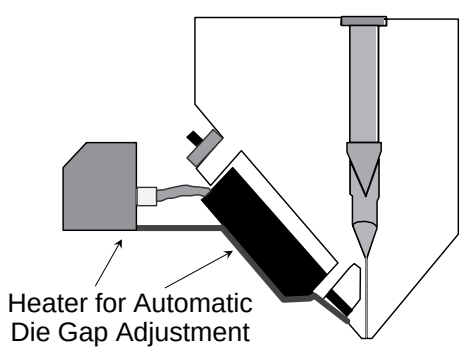

Figure 45.17. Cast film die with automatic die adjustment.

In combination with the automatic die adjustment feedback loops, feedback loops and microprocessor controls can be used to change drive speeds and temperatures to get very accurate film gauge control.

45.2.2 Rolls

Film drops directly from the die onto the casting roll, where it wraps approximately 180 degrees before passing to a second polished roll or cooling roll. The second roll can be used to polish the backside of film, if polishing on both sides is required. A matte finish rather than a glossy finish is obtained by replacing either one or both polished rolls with rolls having a matte surface finish. A third option is to use an engraved roll with a rubber-backed nip roll to produce an embossed film with a particular pattern. It is also possible to adjust the surface gloss with resin blends and additives.

As the film exits the die, it is drawn down by the casting roll moving at higher speed than the resin exiting the die. As a result, the film is experiencing a number of forces. The film exiting the die is thicker than the final desired gauge and is drawn down (both in thickness and width) by a factor normally greater than 2.5:1 up to very high ratios. One key to properly stabilizing the film on the roll is to remove the air from behind the film with a suitable vacuum system. As the casting roll rotates at high speed, it pulls air near the roll surface around with the roll. If the air is not removed, air becomes trapped between the roll and film, which prevents uniform drawing. One or two vacuum systems are used to remove the air and get good film-roll contact. On the film side, one or multiple air knives are used to pin the film to the roll. A single air knife may be used across the entire film width and two smaller air knives or air jets may be used on each film edge to reduce necking. A vacuum and air knife system is shown in Figure 45.18. Vacuum level control in each vacuum chamber provides very good film stability and laydown. By controlling the primary vacuum, the operator can control the contact of the melt film to the roll. Manipulating the secondary vacuum removes the air and any foreign material or airborne contaminants from the area behind or around the film. The optimum vacuum level depends on the film being processed. A

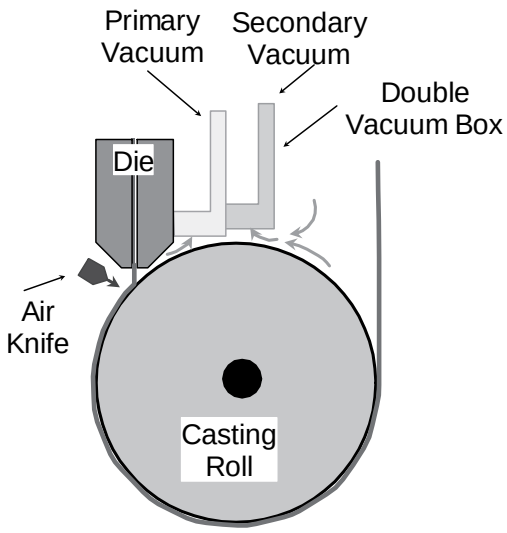

Figure 45.18. Vacuum and air knife system to remove air, pinning film to the casting roll.

double vacuum system, compared to a single vacuum box, provides additional process flexibility; in some situations it may be appropriate to use only the secondary vacuum. Die position is critical to ensure a stable film process. Either a vertical position, shown in Fig. 45.18, or a slight die angle, shown in Fig. 45.16, can be used satisfactorily. If the angle is too severe, the film may drag or hang up on the bottom die lip.

Using a wide die to produce wide film, large roll widths are required, and changing from one roll setup to another is a time-consuming job. Davis Standard Egan, Film and Coating Systems, is touting a roll conversion station that contains both cast and embossing rolls with the option to move the unit from one location to another, minimizing the downtime associated with product changes while increasing production flexibility. Figure 45.19 shows a roll system that goes from one setup to the next; the entire unit is rolled forward or backward to provide the proper location for the introduction of the film into the equipment.

To produce optical quality film with superior gloss and smoothness, the first casting roll has to have a highly polished, chrome plated surface. Multiple chill rolls can be used to obtain the desired polish and gloss while providing the proper cooling. For high smoothness and gloss, the casting roll temperature needs to be set approximately 10°F (5.5°C) below the temperature when the film just starts to stick to the roll. Temperatures below this will remove heat too fast from the film. The final film temperature at windup depends on the line speed, the number of chill rolls to remove heat, and the roll temperatures. Surface temperature across the rolls needs to be maintained within a 3°F (2°C) range to provide product uniformity. If a crystalline resin is being processed, the frost line or crystallization line on the first casting roll needs to be relatively straight across the roll surface. If the crystallization line is not straight, there is either a temperature fluctuation in the roll or the film is not the same gauge across the entire roll. Thin sections cool first and crystallize before thicker sections.

A second factor influencing the haze and gloss (for semicrystalline resins) in the final product is the resin melt temperature at the casting roll. Higher melt temperatures result in higher gloss and lower haze (in semicrystalline materials). However, the melt temperature has to be balanced versus other properties and the line cooling capacity. Higher melt temperatures are associated with reduced film tensile yield and break strength. Increasing the chill roll temperatures will increase the gloss in all resin systems and the haze in crystalline resin

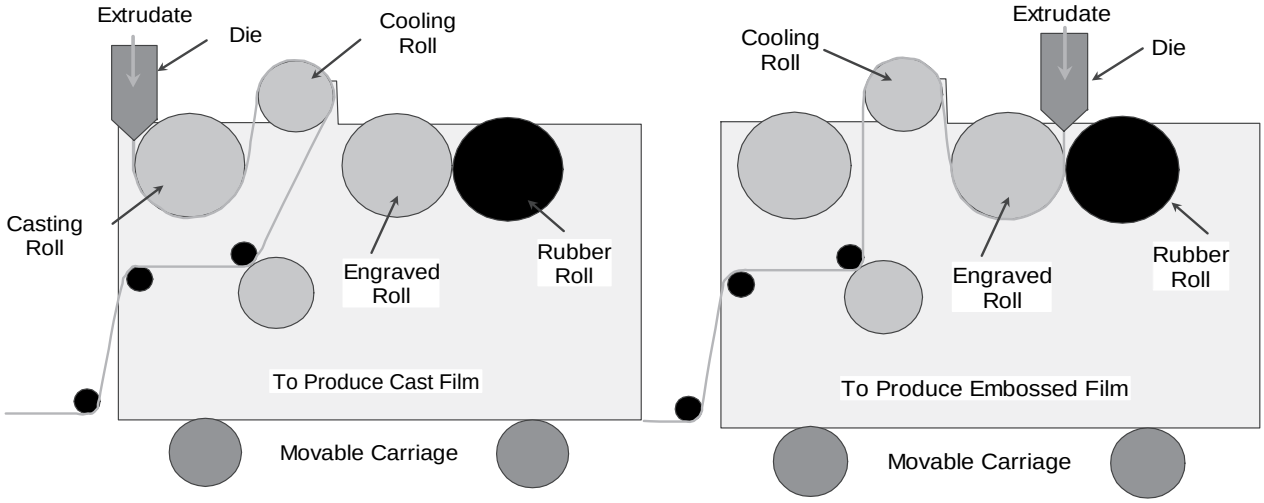

Figure 45.19. Movable carriage by Davis Standard Egan to produce both embossed and cast film.

systems, as the resin is quenched more slowly, allowing large crystals to grow.

Ideally the film will have an S wrap around all rolls to provide maximum contact for polishing and cooling. Cooling is more efficient when film is in contact with rolls than in the air between rolls. Since the film needs to be near room temperature for winding, larger contact angles provide more cooling.

Another option to produce cast film is to use a water bath, as shown in Fig. 45.20. The guide shoe or roll in the bottom of the tank is polished to provide surface gloss. Water in the tank is circulated to remove the heat. The principal advantage of roll cast films compared to water cast films is optical clarity, with higher gloss and less haze. In addition, roll cast films have the potential of being stiffer and having higher throughput rates. In water cast systems, the cooling tank water temperature is between 70° and 150°F (21° and 66°C). As might be anticipated, the water temperature affects the properties, and experimental studies are required to optimize the water temperature for the process and material being run. Generally, low water temperature leads to reduced blocking with improved slip. In addition to optimum temperature control, it is important to eliminate dirt or foreign particles in the water that can scratch the film or lead to other contamination in the final product. The air gap between the die and the water bath controls the film necking, with larger distances resulting in more neck down. Air gaps are generally set in the range of 1–5 inches (25.4–127 mm).

45.2.3 Film Treatment

The film treatment station is not essential to all film lines. It depends on the film application and any other secondary operations required. Polyolefins have poor surface wetability, good chemical resistance, and are nonpolar. While these properties make them very good packaging materials, these same surface characteristics make them difficult to print, decorate, coat, or adhere to other materials such as aluminum foil. As a result, different surface treatments are used to alter the surface polar characteristics to improve the surface reactivity and adhesion between substrates applied in a secondary operation and the film surface. Surface treatment methods available to alter the surface characteristics and improve adhesion are

- Flame plasma treatment
- Corona discharge or electrical surface treatment
- Atmospheric or vacuum plasma treatment

Flame plasma treatment changes the surface chemistry of the polyolefins by forming free radicals on the surface followed by the formation of hydroxy groups. A gas plasma is an extremely reactive ionized gas

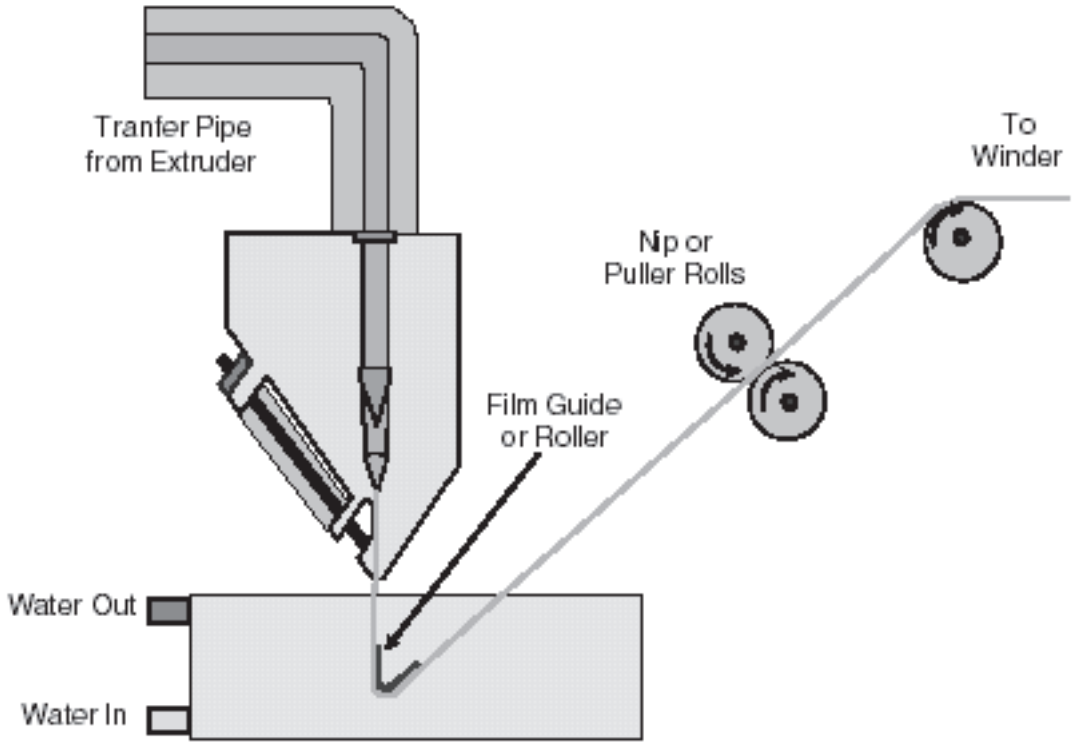

Figure 45.20. Schematic of film casting with water.

mixture generated at the film surface to alter the surface chemistry and subsequent characteristics. Many methods are available to generate this surface altering plasma. One way is a natural gas flame that is mixed with air, causing O_2 molecules to dissociate into oxygen ions, generating plasma that is made up of oxygen, nitrogen, free electrons, positively charged oxygen, and other polar organic functional groups. The film exposed to this plasma is rapidly conveyed forward through the system over a water-cooled temperature control roll. As the flame plasma comes in contact with the film, it alters the surface chemistry of the film, transforming it from a nonpolar to a polar character. The burner is designed to provide the correct contact time between the plasma and web to provide the required treatment. This contact time is part of the system design, based on the throughput rates and the web speed. Treatment is a surface process and penetrates only several nanometers into the film without affecting the film or the backside surface.[6][7] Flame plasma surface modification is long lasting, uniform across the entire surface, reliable, and safe with no environmental effects.

A second way to generate the plasma is through electric discharge, referred to as corona treatment. The electric discharge can be a low frequency discharge, suppressed spark discharge, glow discharge, or high frequency discharge. These systems have the electrodes in different arrangements, making some very compatible with film treatment and others not. High frequency discharge uses electrodes on both sides of the film, with the film passing the electrodes. High frequency discharge is uniform and continuous at both high levels and high speed. The other three processes all have either equipment or process limitations that make them unsuitable for use in treating films. Corona discharge generates ozone as part of the process, and it must be captured and treated. Other factors associated with corona discharge are unwanted backside treatment, pin-holing, and a lower concentration of the polar groups on the film surface than with flame plasma.[8] Lower concentration of polar groups provides fewer bonding sites for inks and other substrates to be attached to the film.

Atmospheric or vacuum plasma treatment is a new development that promises to provide greater versatility and efficiency in treating film.

45.2.4 Slitter

Either prior to or after the treatment station, the film edges are slit to the desired product width. With wide dies, internal or external deckling is used to reduce the film width close to the final product specification width. The film exiting the die is always wider than the desired final width to account for the neck down as the film is pulled away from the die. Attempts are made to keep edge trim to a minimum to reduce the scrap or recycled material. Knives are normally used to trim the film to the final dimension. Edge trim is pulled through a vacuum system to another location, where it is ground into flake for recycling or reuse, or sold as regrind.

45.2.5 Gauging

Automatic thickness gauges are identical to those discussed earlier. With an automatic adjustable die, feedback from a thickness gauge to a microprocessor is required to make on-line adjustments to the die. Based on the thickness profile, heat is supplied or withdrawn from the die bolts to close or open the die gap.

45.2.6 Winders

Winder technology has changed in recent years to higher speeds, larger diameter rolls, and wider rolls in response to the demands in the film industry. Speeds of 2000 feet/minute with roll diameters of 5 feet and widths up to 20 feet are possible. Many cast film producers require small rolls to be produced rapidly, while stretch wrap film producers are seeking larger rolls at higher winding speeds. Large film rolls at lower speeds are normally produced with surface winders, while the smaller

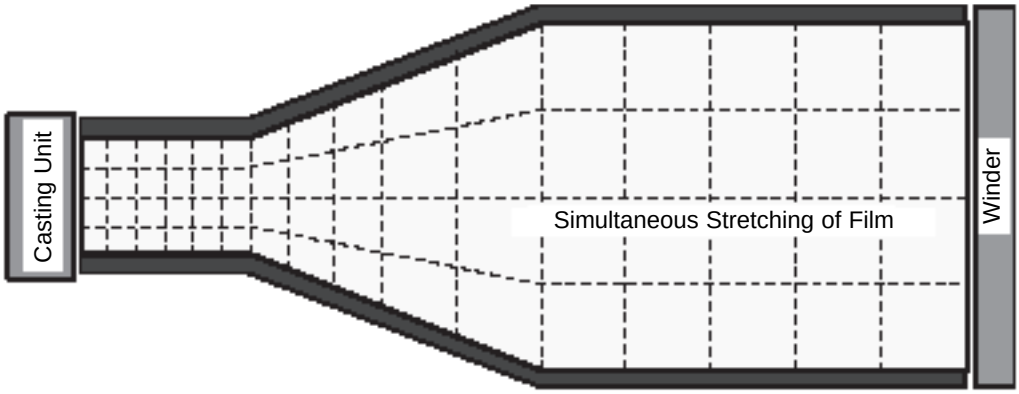

Figure 45.21. Simultaneous stretching to produce biaxially oriented film.

rolls run at high speeds on turret winders use center wind technology. A surface winder has a driven roll that lays the sheet on the winder at a constant speed, while a center-driven winder keeps constant tension on the film through the winder roll itself. With a center winder, the roll rpm changes as the roll size increases to maintain constant web tension. In a surface winder, the windup roll also decreases in speed as the roll gets larger, and it is driven by a constant speed roll contacting the surface of the roll being wound. High-speed stretch wrap operations use a combination of center and surface wind technology. The center/surface winding combination is also used with slippery material, turret winders requiring high web tension, and when in-line slitting is present. The center wind helps to prevent the film from slipping when winding large rolls of slippery film. With center/surface winders, the web tension is independent of the winding tension. At the roll core, the winding tension can be as much as 50% higher than the web tension. As the roll builds, the winding tension decreases compared to the web tension. New technologies added to winders include the size, motor and drive designs, composite air shafts for winding, bowed roll development for high-speed winders (>1500 feet/minute), and composite lay-on rolls.

AC vector and servo drive motors are replacing DC motors to provide more uniform winding. Dual turret winders are replacing single turret winders. Single turret winders can have up to three winding stations. At winding speeds greater than 1000 feet/minute, composite air shafts replace the steel air shaft for reduced weight and lower deflection at a given diameter, although the cost is higher. For the same reasons the lay-on roll is changed from steel to composite.

Winding systems can vary from being very simple to automated systems with high sophistication. Options available in winder designs are

- AC flux vector drives
- PLC with touch screen control
- Rapid transfer from one roll to another
- Swing-out shafts for easier roll removal with automatic inflation
- Adhesiveless transfer systems from one roll to the next
- High-speed indexing

Biaxially oriented polypropylene (BOPP) film production drives winder technology. BOPP film is produced by drawing cast sheet in both the machine and transverse directions. Other resins systems, such as PET, PVC, PS, polyamide, and so forth, can also be stretched or oriented in both the machine and transverse directions, providing improved properties in both directions. As the films are stretched, the molecular chains are elongated in that direction to produce higher strength. Orienting in both the machine and transverse directions produces film with similar properties in both directions versus mono-oriented PP film, which has strength only in the orientation direction. Drawing the product in both directions means the winder has to be wider than the casting roll, and the winder has to run at higher speeds than the casting roll. Figure 45.21 shows a top view of a simultaneous stretching operation.

45.2.7 Operation

To produce film with good gloss, surface qualities, gauge control, and that is pucker-free, the system must be operated with the correct melt temperature, roll temperatures, and speeds. Excess casting speed or winder speed causes neck down, which can affect gauge and width. Casting roll temperature can affect product smoothness, gloss, and clarity. The vacuum level behind the die can alter the film laydown characteristics on the roll. High puller roll or winder speed can change the molecular orientation between the casting roll and the windup, affecting properties in both the machine and transverse directions. Critical to the operation is a uniform polymer flow from the die, at a constant and correct melt temperature and pressure across the entire die face. Troubleshooting guidelines for the cast film process are given in Part 4, Chapter 30.

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Review Questions

1. What is the thickness convention that distinguishes between film and sheet?
2. What is the purpose of the rolling bead or bank in sheet extrusion?
3. At what gauge is sheet no longer easy to wind on a roll and has to be cut and stacked?
4. How are the optimum roll temperatures determined to produce a sheet product with the best gloss?
5. What is the difference in the takeoff equipment used for sheet and cast film?
6. What is the definition of PLI? Where is it used and why is it important?
7. What is the purpose of deckles and how are they used?
8. What die adjustments are present on a sheet die and a cast film die, and what is the function of each adjustment?
9. How does a coat hanger die differ from a T-slot die and a fishtail die?
10. How can temperature be used to promote uniform polymer flow out of the die?
11. What is neck down?
12. What is the difference between an up stack and a down stack roll operation?
13. How is a sheet line started up?
14. What is the purpose of the treatment center in a cast film operation? Can it be used in a sheet process?
15. What controls the gauge in cast film and sheet production?
16. How does a beta gauge work and what does it measure?
17. How are the gaps set in a sheet or film die? In a roll stack?
18. What is the purpose of the vacuum box in cast film production?
19. What is the difference in process between cast film made on a casting roll and that made using a water bath?
20. What is the difference between a center winder, a surface winder, and a center/surface winder, and where is each one used?

46 Blown Film

Blown film is another process used to make thermoplastic film in high volume. The process is different from the cast film process discussed in Chapter 45 in the die design and cooling system. Sheet and cast film are produced with both single and twin screw extruders. Blown film is almost exclusively produced on single screw extruders. Some are equipped with grooved feed throats to increase the extruder capacity. Instead of extruding the film out of a flat die and casting it onto a cooling roll or in a water tank, the extrudate flows through an annular die, forming a bubble or tube that is pulled away from the die vertically (can be horizontal or downward) as the polymer is air cooled. Upon exiting the die, the bubble can be blown to different diameters, allowing different film widths to be produced with the same die. An air ring at the die exit and possibly internal bubble cooling can be used to solidify the polymer melt. The cooling rate is determined by the air flow, film speed, and temperature difference. Bubble stability is critical for gauge control and uniformity. After cooling, the bubble is collapsed in a forming tent where two nip rolls collapse the bubble into two flat layers. After the nip are two puller rolls, similar to those used in cast film and sheet, to pull the film away from the tower. A treatment center to modify the polymer surface polarity for printing or adhesion to another substrate can be incorporated between the nip rolls and the windup. If the product is wound flat and not slit, the film is packaged as a continuous tube. Slitting the film on one side will produce double the lay-flat width, and slitting on both sides will produce two lay-flat width rolls. Figure 46.1 shows a blown film process in operation. Figure 46.2 identifies the different components within the process that will be discussed in more detail in this chapter.

Following are the advantages of blown versus cast film.

Cast film has

- Better optical properties as measured by transparency or haze and gloss
- Higher throughput rates
- Stiffer films as a result of instant quenching
- Better gauge control (variation 1 to 2% versus 3 to 4% for blown film)

Blown film has

- Better property balance between the machine and transverse direction
- One die can make many different widths without using deckles or significant trimming

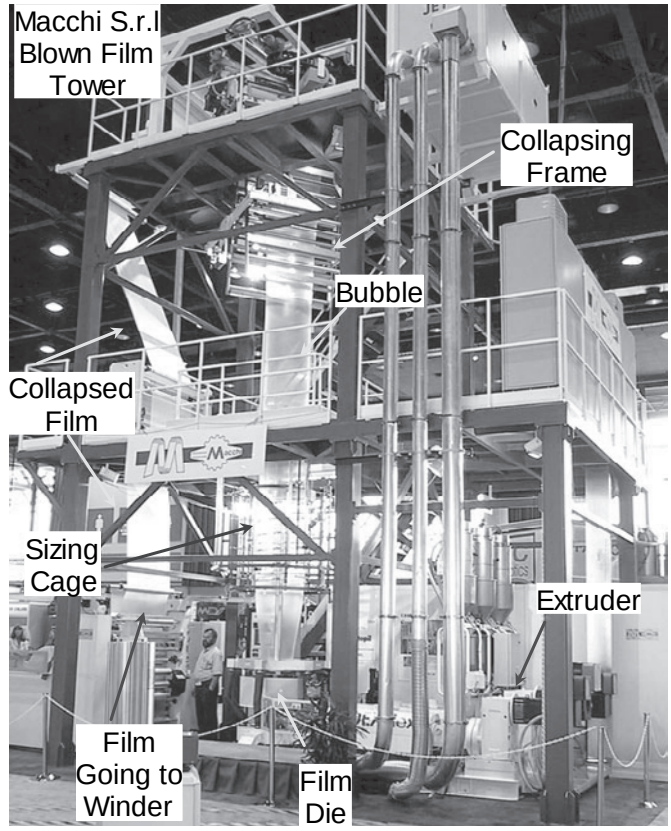

Figure 46.1. Maachi S.r.l. blown film line.[1]

- Less scrap with no edge trim removal
- Equipment cost (approximately 50% of a cast line)

When making bags, blown film can be sized so that only one heat seal at the bag bottom is needed. Bags made from a cast film process require three heat seals—one on the bottom and one on each side of the bag.

Extruders used in blown film lines are normally close to the floor, compared to other extrusion operations, because the takeoff equipment is above the extruder. With the extruder close to the floor there is approximately four more feet for cooling. As a result, dies are supported on roll-around carts with the die entrance on the bottom. A melt or transfer pipe from the extruder is connected to the die. Different dies can easily be rolled into the center of the blown film tower and connected to the extruder to either change the blown film die size or the material being processed. Blown film dies are unique in that many different size products can be made on the same die without changing the die size.

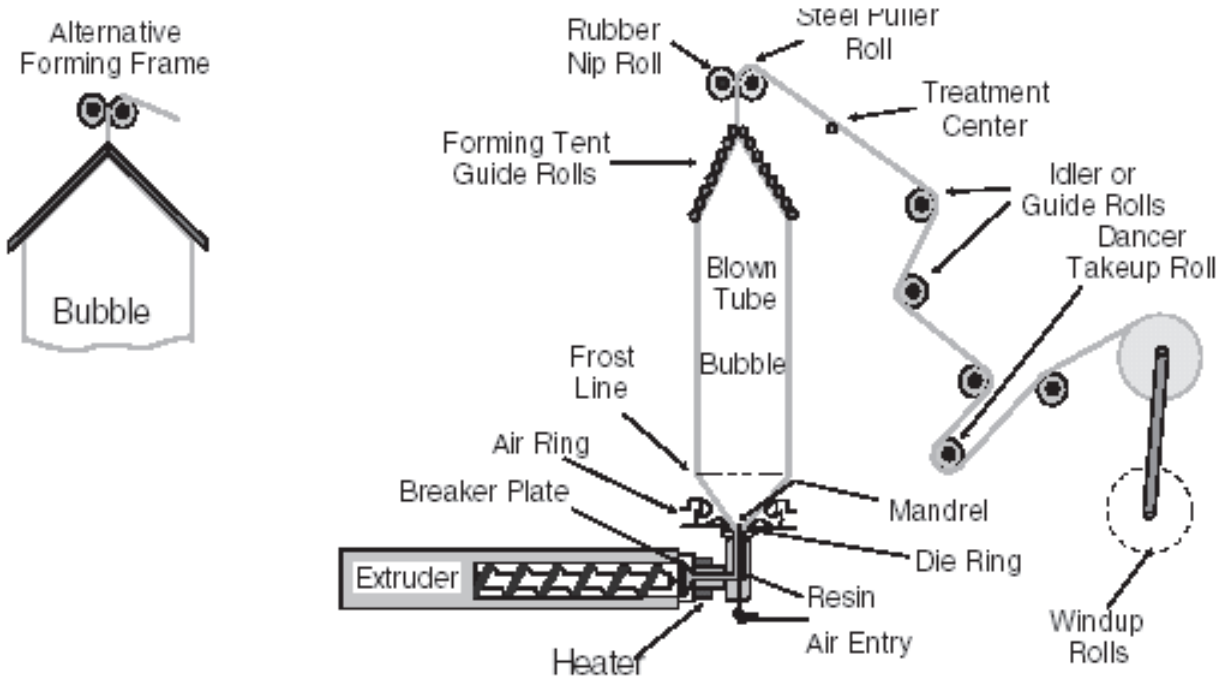

Figure 46.2. Schematic of blown film process showing two types of forming frames.

46.1 Die

Over the years, there have been three types of dies commonly used to produce blown film. The first two, a bottom-fed spider die and a side-fed die, have almost been replaced by the bottom-fed spiral die, which provides more uniform resin distribution and eliminates weld lines caused by the spider holding the mandrel. Air passes through a center hole in the die and blows up the bubble or is used for internal cooling. While die technology has advanced considerably, requirements of the die have remained constant. A good die design must accomplish the following objectives:

- Produce uniform thickness
- Be streamlined for uniform resin flow with no dead spaces where polymer degradation can occur
- Minimize the residence time
- Optimize the shear rates for the resins being processed
- Have a robust design that withstands the rigors of production in day-to-day operations

Three die designs are shown in Fig. 46.3. Some die designs use a spider ring to support the mandrels in the center of the die. Spider support arms disrupt the polymer melt flow through the die, causing weld or knit lines. Weld or knit lines can cause lower strength where they occur. To minimize the effect, sufficient residence time and temperature are needed after the polymer recombines to allow the molecules to reentangle in the melt state.

The spiral type die shown in Figure 46.3 has flow channels around the mandrel with narrower land areas in between. Polymer enters the bottom, passes through the channel and over the land to reach the die exit. Figure 46.4 shows the parts of a bottom-fed spiral mandrel die in more detail. Polymer melt enters the bottom and is distributed around the mandrel. The mandrel is held in place by a spider ring. Air to inflate the bubble is passed through one of the spider ring legs. The smear device refers to the deep channel spirals and land areas between the spirals going around the mandrel circumference as the melt moves to the die exit. Spirals are deep at the bottom, becoming progressively smaller as they approach the end, building pressure in the die. Just prior to the exit is a final, slightly deeper melt reservoir used to remove melt pressure variations prior to exiting the die. Heated die lips are used on some blown film dies to reduce melt fracture caused by high shear stress at the die lips. The die centering ring is used to adjust the die gap for uniform melt film thickness around the entire die.

Polymer exiting the extruder and passing through the melt pipe has a temperature gradient across the melt pipe caused by the plug flow within the pipe. A radial adjustable melt thermocouple in a transition pipe may measure melt temperature differences of 20–40°F (11–22°C). This corresponds to melt viscosity differences entering the die. Temperature uniformity is

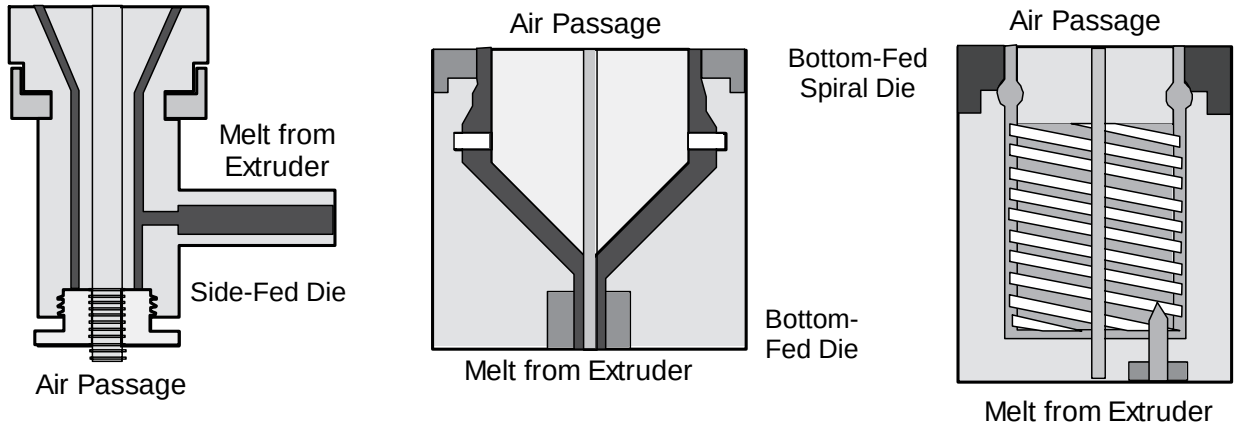

Figure 46.3. Schematic of three types of blown film dies.

improved by dividing the melt into two to four streams at the base of the mandrel. As polymer passes through the spirals and land areas on the mandrel, melt temperature differences are reduced. This helps the polymer exiting the die to have a more uniform temperature profile and viscosity. All melt distribution occurs in the spiral area of the die. Blown film die design is extensively done on computers to provide the proper design. Variables used in die design include die geometry, polymer rheological properties, and flow through the die. After the design is complete, simulation programs are used to predict the flow, shear rates, residence time, and melt temperatures in the die. If the design does not meet the specifications, the die design is modified, followed by a new simulation. A proper die design and simulation is completed before the die specifications are sent to the manufacturing floor. To purchase a properly designed blown film die, it is only necessary to know the resin, extruder size, anticipated throughput range, and the die diameter.

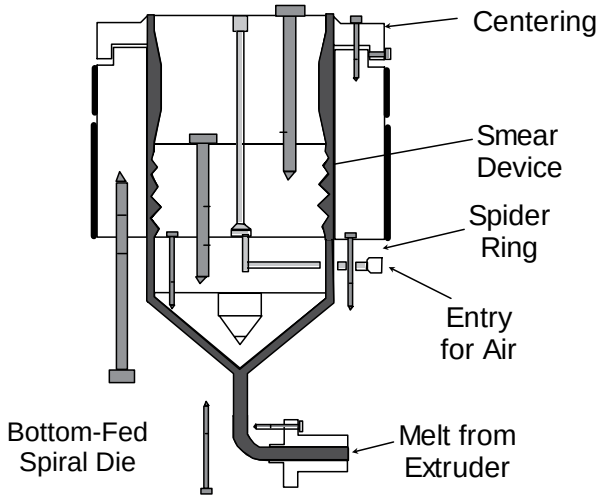

Figure 46.4. Components of a bottom-fed spiral mandrel blown film die.

All dies must be designed with sufficient robustness to withstand the operational and mechanical requirements plus a safety factor for unexpected events in the die life cycle. The design has to address surface smoothness for flow with no dead space or stagnant areas where polymer residence times could be excessive, causing degradation; anticipated pressures generated by the polymer flow; thermal expansion; uniform heating in all areas of the die; and, above all, it has to be practical and easy to use by production personnel. Dies are made of AISI-4340 or 4140 machine steel. The surfaces are either chrome or nickel plated to optimize surface smoothness and flow.

Figure 46.5 shows a spiral mandrel-type die from Akron Extruders. The air ring around the top of the die is for cooling the film as it exits the die. This die is

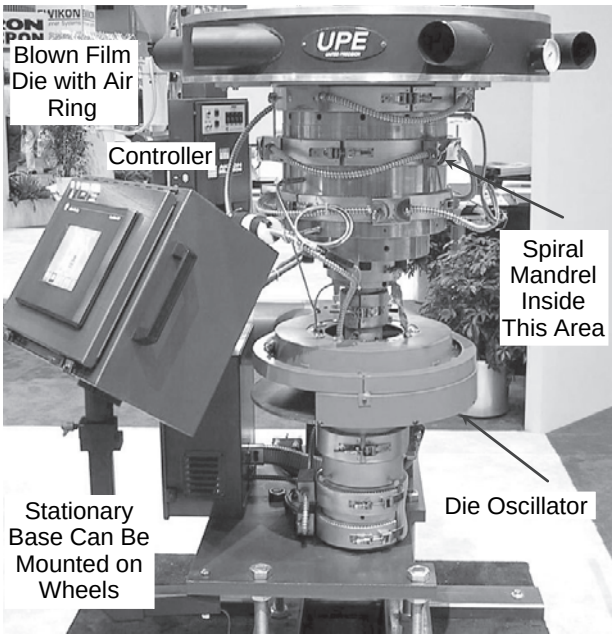

Figure 46.5. Akron extruders blown film dies.[2]

mounted vertically for the bubble to be blown upward, as that shown in Figure 46.1. It is becoming more common to have bubbles blown downward or horizontally. While the internal workings of the die do not have to change to accomplish this, the die's physical configuration has to be altered to accommodate blowing in directions other than up. To improve gauge uniformity, the die or forming tent is oscillated 180–360 degrees or rotated. This prevents a thick or thin section exiting the die from occurring in the same location on the wound roll throughout the run. The die oscillator is shown in Figure 46.5.

As the technology has advanced and the applications have become more demanding, blown film has changed from single layer to multilayer film using coextrusion, covered in Part 6. Die complexity increases dramatically with the addition of other polymers, film layers, and more extruders. Ten-layer films can be produced through coextrusion technology and blown film dies.

46.2 Cooling

The two ways the bubble is cooled are

- External cooling with an air ring
- Internal bubble cooling (IBC) combined with an external air ring

Figure 46.6 shows the air ring around the die. Crystalline polymers, such as PP, HDPE, or LDPE, exiting the die in the amorphous state are clear and transparent. As the film cools, it starts to crystallize, becoming translucent, hazy, or opaque. The point where opacity begins is called the "frost line." Before the frost line, the melt can be deformed. After crystallization, it is much more difficult and requires more energy to deform the film. The frost line height is controlled by the air flow, film speed, and temperature difference. Of these factors, the most important is the air velocity and the way it impinges on the bubble. The air ring is specified based on cooling capacity, bubble stability, and the uniformity of the air stream. Air rings are available with either single or dual lips. Figure 46.6 shows a dual air ring.

IBC plus the air ring cools the film from both sides. This added cooling allows for higher line speeds. In addition to better cooling, IBC provides better bubble stability. This allows higher air velocities in the cooling ring without influencing the bubble stability. Figure 46.7 shows how the fresh air is brought into the bubble and removed. Without IBC, air is initially added to blow up the bubble; however, once the bubble is the correct size and the film is passing through the nip in the collapsing frame, no additional air is added to the system or the bubble size will change. Changes in bubble size affect laydown width and gauge. IBC reduces the air temperature

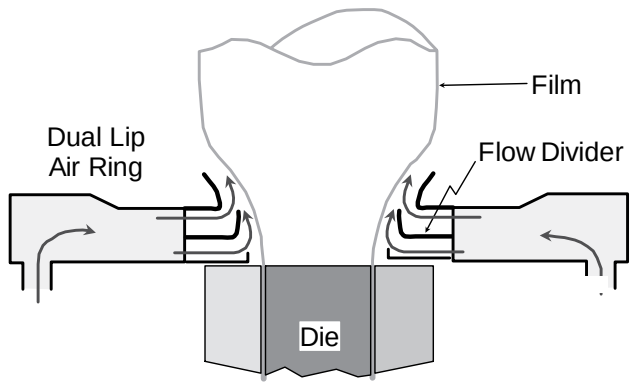

Figure 46.6. Dual lip air ring.

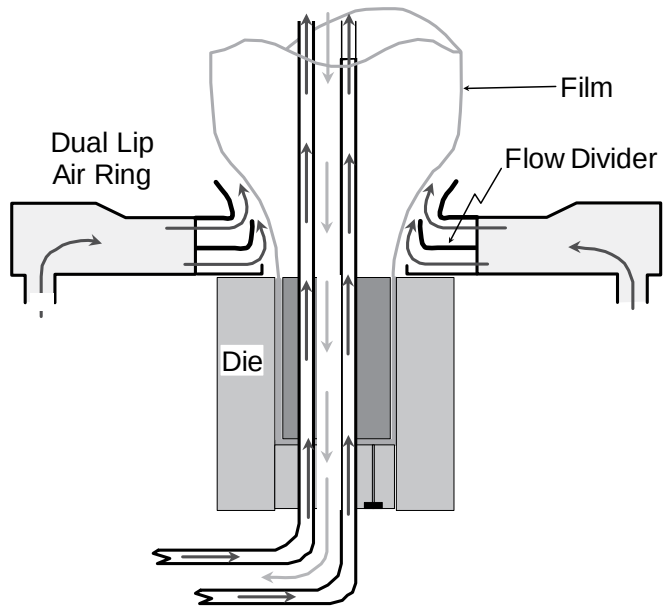

Figure 46.7. Internal bubble cooling.

inside the bubble while maintaining the bubble diameter. Many options are available for IBC, with patented designs from several suppliers. Air is brought into the bubble and dispersed by mechanical means at the base of the film as it exits the die body and along the inner surface as the bubble is formed. The hot air rises with the bubble and is returned through the center tube. Figure 46.8 shows a spiral mandrel blown film die from Addex Inc. with an air ring and internal cooling.

Brampton Engineering Inc.[4] has a 4+1 Pancake IBC system that offers improved cooling efficiency. It has five adjustable orifice pancakes, and each can deliver a different air flow. With this process flexibility, the unit is used to process many resins in either single or multilayer films. The blower system regulates the air flow into the bubble through the motor speed. Lower air flow can be provided to the bubble bottom, where the melt film has

the lowest strength. Air flow velocity is gradually increased in the upper pancakes to improve the heat transfer and provide more cooling. An air flow regulator plate provides greater overall air exchange while providing profiled air flow to each individual pancake. The control system precisely controls the air exchange to provide outstanding bubble stability. Figure 46.9 shows a picture of the 4 + 1 Pancake IBC.

46.3 Blown Film Tower

The blown film tower supports the

- Nip rolls
- Forming tent at the top
- Web oscillation system (if used)
- Bubble stabilizer

It provides distance to cool and solidify the film being processed at the throughput rate.

The tower is rigidly constructed so that vibrations are not transmitted through the frame to the bubble. Vibrations produce transverse lines in the bubble as the molten polymer leaves the die. The collapsing frame or tent at the top of the bubble can be made out of rollers, wood slatts, or perforated metal. Whatever collapsing frame surface is used, it must be highly polished and free of nicks, burrs, or other imperfections that can scratch the film or lead to other film appearance defects. The collapsing frame is adjustable to the bubble diameter, allowing the frame angle to be changed based on the bubble diameter. Both the nip and the collapsing frame have to be centered over the die to provide uniform draw. Normally one steel and one rubber-coated roll are used to prevent air from escaping past the nip and interfering with winding. Nip rolls need to be easily opened and closed to facilitate line operation.

Bubble stabilizers or sizing rings are used to control the bubble size. A cage inside the tower prevents the bubble from being expanded too much. This assists both gauge control and bubble stability. Brampton Engineering Inc.[4] produces a stabilization cage that prevents the film from dragging over the arms in the cage by providing an air layer to eliminate drag while preventing bubble distortion and surface imperfections due to dragging. Figure 46.10 shows a blown film tower from Maachi S.r.l.[1]

If an oscillating haul-off is used, the collapsing tent and nip rolls become part of the oscillating unit. Oscillating 180 or 360 degrees distributes die defects across the roll. A variable speed motor and drive control the oscillation rate. After leaving the tower, pull rolls pull the film through idler rolls to provide uniform film tension.

Between the nip roll and the winder, the film can be slit on both sides to form two separate films, slit on one side and opened up to form one large film, or wound as a tube to be used in bags. Between the slitter and the windup, a treatment station similar to that discussed in Chapter 45 for cast film can be installed to modify the surface chemistry, making the film more suitable for printing and bonding to other substrates.

Figure 46.10 shows a blown film tower with the film being drawn down between the idler rolls and wound into relatively large rolls.

Blown film technology has advanced to a double bubble line from Reifenhauser[5] and a triple bubble line to make sausage tubes from Kuhne.[6] In the Reifenhauser line, a

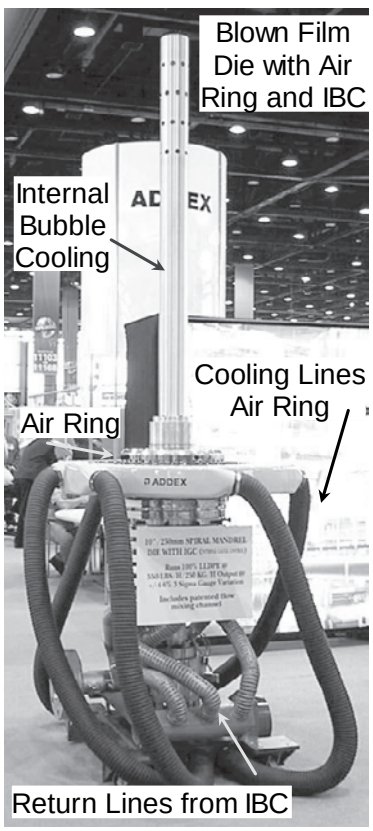

Figure 46.8. Addes, Inc. blown film die with air ring and internal bubble cooling.[3]

Figure 46.9. Brampton Engineering, Inc. pancake IBC.[4]

thick-walled tube is extruded downward and cooled in a water calibrator to produce uniform wall thickness. After pinch-off, the tube is reheated with infrared heaters and reinflated, stretched biaxially, and rapidly cooled to produce BOPP. The American Kuhne system is aimed at a specialty coextrusion market to produce sausage casings. Again, a narrow tube is extruded downward through a water vacuum calibrator before the bubble is blown a second time. The bubble is nipped and reinflated prior to being drawn in an annealing oven.

Black Clawson[7] has a machine to produce machine direction oriented (MDO) film. In the MDO, the film is heated and drawn between slow rolls and fast roll from 2.5 to 10:1 in the machine direction. After drawing, the film is annealed to reduce stress and then cooled. The web enters the MDO at 100 feet/minute and leaves at 250–1000 feet/minute. This reduces the film gauge up to 50% while increasing the machine direction strength. Drawing barrier film increases the barrier properties, allowing less to be used or a cheaper barrier resin to be incorporated into the structure.

Figure 46.10. Winding blown film from Maachi S.r.l. blown film line.[1]

46.4 Winders

Winders for blown film are similar to those used in cast film production. Both surface winders and turret center drive winders are used. The surface winders operate through friction between the roll being wound and a drive roll that runs at constant speed. Rolls up to 5 feet in diameter and 30 feet in width are possible with line speeds up to 1000 feet/minute. Automatic roll changeover requires the operator to remove the completed roll and place a new core in position for the next changeover. Options available with different winders include variable pressure between the driven roll and the wound roll to adjust the roll hardness to customer specifications, higher pressure during changeover to prevent slippage, digital counters, PLC controls, full automatic roll unloading when complete, static eliminator bar, cooling systems to prevent post-winding shrinkage, trimming and slitting within the winder station, and additional nip rolls.

Center winding is done with a variable speed motor that drives the roll core. Automatic torque reduction as the roll diameter increases allows the roll to slow down. Rolls up to 3.5 feet in diameter and 30 feet wide are possible with 1000 feet/minute line speeds. As with surface winders, automatic roll changeover, PLC controllers, and pneumatic-driven knife cutters are standard equipment. Bowed rolls are used to produce wrinkle-free film. Separate nips at the film lay-on roll allow the film to relax before it is wound on the finished roll, reducing film tension.

One issue in winding both blown and cast film is controlling air entrapment during the winding operation. Air entrained in the wound rolls leads to telescoping, lower contact between the film web and the roll, and lower torque capacity. Figure 46.11 shows air captured between the roll layers and also between the film and an idler roll. Air trapped between the idler roll can cause film scratching or wrinkling. Entrained or trapped air reduces friction and can cause processing problems.[8] New winders incorporate mechanisms to minimize air entrapment.

Addex Reinhold[3] blown film winders use vacuum to remove entrained air. A. R. Engineering Machine Inc.[9] uses a winding roll with a diamond-shaped groove in the rubber cover. Brampton Engineering Inc.[4] uses a linear lay-on device that controls air. Different manufacturers have unique features that can make processing easier. Models come in both single and turret winders with surface, surface/center assist, and center driven drive systems.

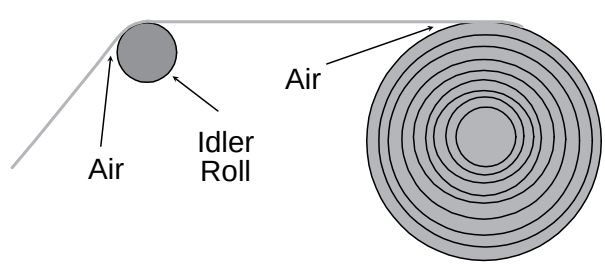

Figure 46.11. Entrained air in winding blown or cast film.

46.5 Gauging

NDC Infrared Engineering[10] has developed a non-contact, blown film, rotary, gamma backscatter gauge to determine the film thickness in blown film processes where the film is not rotated or oscillated. A frame containing an automatic bubble tracking system with a gamma sensor is mounted around the film. The gamma sensor moves toward the bubble to read the film thickness. The sensor rotates around the film at 0.1 to 1 revolution/minute. Gauges are available to handle bubble sizes up to 40 inches (1000 mm) and 70 inches (1800 mm) in diameter. Plast-Control[11] measures the film thickness with a C-head sensor that is based on capacitance and constant contact with the film surface. The technology provides high resolution, whether mounted in the collapsing frame or on a radially adjusted scanner.

46.7 Operation

Operating a blown film line is similar to other extrusion processes in that the polymer exiting the extruder must have uniform melt temperature and pressure while also being at the proper temperature and pressure. Compared to cast film, blown film is usually run at a lower melt temperature with lower melt flow index resins. Factors affecting process stability are

- Bubble cooling
- Bubble stability
- Frost line height
- Air flow
- Cooling air temperature
- Bubble sizing cage
- Bubble collapsing frame
- Tension control
- Winder

Throughput is controlled by

- Tower height
- Resin melt temperature
- Extruder capacity

To string up the machine, the melt exiting the die is connected to film that is strung through the film path and hung over the die. As the extrudate is removed from the die, it is connected to the old film and pulled through the system at a slow rate. Once the new film passes the top of the tower, the nip rolls are closed and the film continues through to the winder. Air is used to blow the bubble to size. Extruder screw, puller, and winder speeds are increased to provide the desired throughput rate and film gauge. Once the bubble is properly inflated, the air entering the bubble replaces air leaving the bubble, as the total pressure within the bubble must remain constant to provide uniform bubble diameter and film gauge. The two blown film processes are conventional and high stalk. In the conventional process, the bubble is blown to its desired diameter almost immediately upon exiting the die, and the frost line is within a couple of feet of the die exit. In the high stalk process, the diameter of the bubble exiting the die remains the same size as the die opening for many feet above the die exit before it is blown out to the desired bubble diameter.

Parameters and terminology associated with blown film are shown in Fig. 46.12 and defined below:

- **Frost line height** is the point where the film changes from the melt to the solid state with semi crystalline polymers. Film changes from a transparent amorphous state to a hazy or translucent crystalline state.
- $\mathbf{D_D}$ and $\mathbf{D_B}$ are the die and the bubble diameters.
- $\mathbf{T_o}$ and $\mathbf{T_f}$ are the initial and final film thickness. The film thickness is dependent on the bubble size and the draw ratio.
- $\mathbf{V_o}$ and $\mathbf{V_f}$ are the die exit and final film velocity. Higher draw ratios result in relatively higher V_f values.
- **MD** and **TD** are the machine and transverse direction orientation. (TD is also called the cross machine direction and is designated as CD.) The

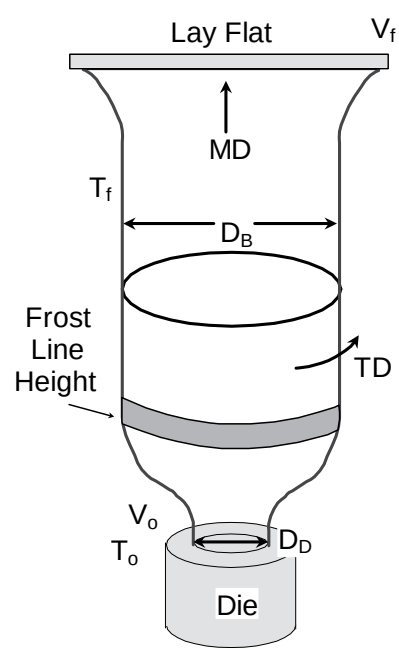

Figure 46.12. Blown film parameters.

property ratios depend on the bubble size and draw ratio. Larger bubble size provides more transverse direction orientation and higher draw ratios provide more machine direction orientation.

Blow-up ratio (BUR) is defined as the bubble diameter divided by the die diameter. It measures how large the bubble has expanded in the transverse or cross-machine direction. Blow-up ratio is given by Eq. (46.1):

$$BUR = \frac{D_B}{D_D} = \frac{2 \times Lay\ Flat\ Width}{\pi \times D_D} \tag{46.1}$$

Lay-flat width (LFW) is the collapsed bubble width before slitting. It is given by Eqs. (46.2) and (46.3):

$$LFW = \frac{\pi \times D_B}{2} = 1.57 D_B \tag{46.2}$$

$$LFW = \frac{\pi \times D_D \times BUR}{2} \tag{46.3}$$

The draw down ratio (DDR) is the drawing taking place in the film tower in the machine direction, and it is related to the blow-up ratio. Equation (46.4) is used to calculate the draw down ratio:

$$DDR = \frac{T_o}{T_f \times BUR} \tag{46.4}$$

The output rate (Q) of a blown film line can be calculated from Eq. (46.5):

$$Q = \pi \times D \times V_f \times T_f \times \rho \tag{46.5}$$

where

ρ = Film density

Specific die output rates are calculated differently in North America and Europe. North America calculates pounds/hour/inch of die circumference, and Europe uses kg/hr/mm of die diameter. To convert the blown film output rate calculated in Eq. (46.5) to specific die output rates, use Eqs. (46.6) and (46.7):

$$Specific\ Rate_{N.Am.} = \frac{Q}{\pi \times D_D} \frac{\text{lbs}}{\text{hr} \times \text{inch}} \tag{46.6}$$

$$Specific\ Rate_{Europe} = \frac{Q}{D_D} \frac{\text{kg}}{\text{hr} \times \text{mm}} \tag{46.7}$$

In all calculations it is important to keep the units consistent within a given calculation to obtain the correct results. Film calculators are available from various equipment suppliers to quickly provide answers to the various equations.

One blown film advantage compared to cast film is the ability to make many film widths with the same die by modifying the blow-up ratios. Table 46.1 presents some blow-up ratios for different die diameters and lay-flat widths with MD draw ratio equal to 1. From Fig. 46.13 one can see the relationships among the blow-up ratio, lay-flat width, and die diameter. The larger the blow up ratio, the higher the transverse direction molecular orientation with corresponding decreases in the draw down ratio. With the increased bubble surface area and decreased film thickness, cooling time is reduced. All the bubble expansion takes place between the die exit and frost line, where the polymer is still molten. The time frame for this expansion is given by Eq. (46.8), which is the time required for the film to travel from the die exit to the frost line:

$$Time\ to\ Frost\ Line = \frac{Height\ of\ Frost\ Line}{V_f - V_o} \cdot Ln \frac{V_f}{V_o} \tag{46.8}$$

The blow up ratio affects properties such as tensile and flexural strength, toughness, and optical properties in crystalline materials. These properties are all related to the molecular orientation and the ability to orient the molecules in more than one direction. As the transverse or cross-machine direction properties increase, the machine direction properties decrease, as some molecules that were initially oriented in the machine direction are now oriented in the transverse direction. Tear properties change substantially as the molecular orientation changes. If all the molecules are aligned in the machine direction, it is easy to tear the film in that direction and quite difficult to tear the film in the cross-machine direction. Balancing the molecular orientation in the machine and transverse directions makes it difficult to tear the film in either direction.

Blown film is sometimes produced using coextrusion, where both extruders use the same resin and same melt flow. This is normally done to improve process stability and line efficiency. If a pinhole should occur in the film during production, allowing the bubble to collapse, considerable time can be lost restringing the line and getting back to equilibrium or steady-state operation. The odds of two pinholes occurring simultaneously at the same location in each of the film layers are highly unlikely. Consequently, a production run is very unlikely to be interrupted by pinholes in a given film layer when coextrusion is being run.

The principal advantages and disadvantages of blown film versus cast film are summarized below:

- Cast film has better optical properties.
- Cast film lines are more expensive to purchase.
- Cast film has better gauge uniformity.

Table 46.1. Lay Flat Film Widths Using Various Die Diameters with Specific Blow Up Ratios

LFW, inches	5	10	15	20	25	30	35	40	45	50	60
Die Diameter, inches											
2	1.6	3.2	4.8								
3	1.1	2.1	3.2	4.2	5.3						
4		1.6	2.4	3.2	4	4.8	5.6				
5		1.3	1.9	2.5	3.2	3.8	4.5	5.1	5.7		
6		1.1	1.6	2.1	2.7	3.2	3.7	4.2	4.8	5.3	5.8
7			1.4	1.8	2.3	2.7	3.2	3.6	4.1	4.6	5.3
8			1.2	1.6	2.0	2.4	2.8	3.2	3.6	3.9	4.4
9			1.1	1.4	1.8	2.1	2.5	2.9	3.2	3.5	3.9
10				1.3	1.6	1.9	2.2	2.5	2.9	3.2	3.5
11				1.2	1.5	1.7	2.0	2.3	2.6	2.9	3.2
12				1.1	1.3	1.6	1.9	2.2	2.4	2.7	2.9

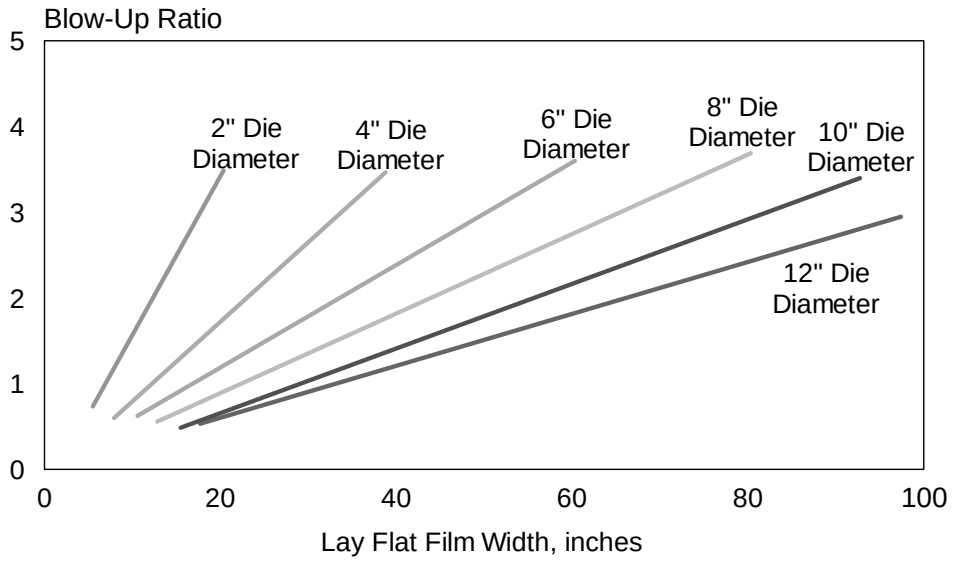

Figure 46.13. Effect of blow-up ratio on lay-flat width with different diameter dies.

- Cast film can be produced at higher production rates
- Blown film can produce a wide range of film sizes from the same die.
- Blown film provides a better balance of properties between the machine and transverse direction.
- Blown film lines are used to make shrink wrap films.
- Blown film lines require lower melt temperatures.
- Blown film towers required large vertical heights.
- Bags produced by blown film operations require no seals on the side of the bag.

Comparison of the effect of polyethylene polymer properties on processing and performance of blown film is shown in Table 46.2.[12] While the property/processing performance was generalized for metallocene catalyzed polyethylene, it is generally true for all polyethylene resins.

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Table 46.2. Effect of Polymer Properties on Blown Film Processing

Polymer Property	Effect	Process Impact
Lower Degree of Branching	Lower Chain Entanglement	Easier Draw Down
Lower Density	Softer	Higher Friction in Collapsing Frame
Lower Density	Tacky	Hard to Wind
Less Low-Molecular-Weight Fraction	Higher Die Pressures	Higher Melt Temperature
Less Low-Molecular-Weight Fraction	Higher Torque	Less Throughput
Less High-Molecular-Weight Fraction	Poorer Melt Strength	Lower Bubble Stability
Less High-Molecular-Weight Fraction	Lower Viscosity	Easier Draw Down

Review Questions

1. What is meant by the following terms and why are they important to blown film processing: blow-up ratio, draw down ratio, and frost line height?
2. What are the principal advantages and disadvantages of blown film versus cast film processing?
3. How does internal bubbling cooling improve blown film process stability?
4. What are the requirements of good blown film die design?
5. During processing, how is the air changed in the center of the bubble with and without IBC?
6. What are the three types of dies used in blown film production? Which die is most common in production today?
7. What causes weld lines in blown film dies and how are weld line effects eliminated?
8. What is the purpose of an air ring?
9. What is the purpose of rotating either the die, collapsing frame, or both?
10. How does a pancake internal bubble cooler work?
11. What is the difference between conventional and high stalk blown film?
12. How is the bubble stabilized in the blown film tower?
13. What is the purpose of the collapsing frame at the top of the tower?
14. What are the effects of air in the winding operation?
15. How is film lay flat width calculated?
16. What is the lay flat width of LDPE film running on a 6-inch diameter die with a blow up ratio of 4:1?

47 Extrusion Coating and Lamination

Extrusion coating extrudes a polymer onto a substrate. Combining two inexpensive materials to make a higher performance product adds value and utility. Materials typically coated with plastic include paper, polyester, metal foils, cellophane, paperboard, cloth, and other plastics. The objective is to combine the best properties of each material into a third product that can perform a function neither of the individual products can do on their own. Potential value-added functionality might be

- Heat sealability for packaging applications
- Improved tear or crease resistance
- Better barrier properties to water or oxygen and other gases
- Improved appearance
- Additional chemical resistance
- Improved printing or decorating ability

Figure 47.1 shows a potato chip bag with a decorated plastic coating laminated over top of foil. The plastic overlay may be coextruded with different layers to improve the surface printing, barrier properties to maintain the freshness of the potato chips, material to properly heat seal the top of the bag in the packaging operation, and a foil liner that maintains content freshness as it acts as a moisture barrier to protect the food while protecting the plastic from chemical attack by any of the food or its additives.

Figure 47.2 shows a possible extrusion coating configuration. The unwind roll contains the material to be coated. It is passed through idler and tensioning rolls to remove wrinkles and flatten the web. Before coating, the substrate may be preheated on a heated roll, bringing it closer to the coating resin melt temperature. Using a film die, a polymer is coated onto the substrate, analogous to a cast film process. Instead of extrusion onto a casting roll for cooling, the substrate and molten film are passed through pressure nip rolls to improve the adhesion between the substrate and the molten polymer. Which surface is exposed in the final applications and the surface requirements determine the finish to be applied in the coating operation. Gloss is specified as standard gloss, mirror gloss, or pocket gloss, depending on the reflective image and the surface brightness or shininess. Mirror gloss gives an outstanding reflected image, which is different from just having a shiny surface. Surface finishes available depend on the press roll surface characteristics and can range from a highly reflective gloss finish to a matte finish.

After being joined, the structure is cooled before secondary operations that may include slitting, surface treatment, printing or decorating, and winding. In the slitting operation, edge trim can be removed or the web can be slit into different widths in-line. Some operations slit into multiple rolls off-line. Surface treatment might be used with either polyethylene or polypropylene films to modi-

Figure 47.1. Foil laminated food package.

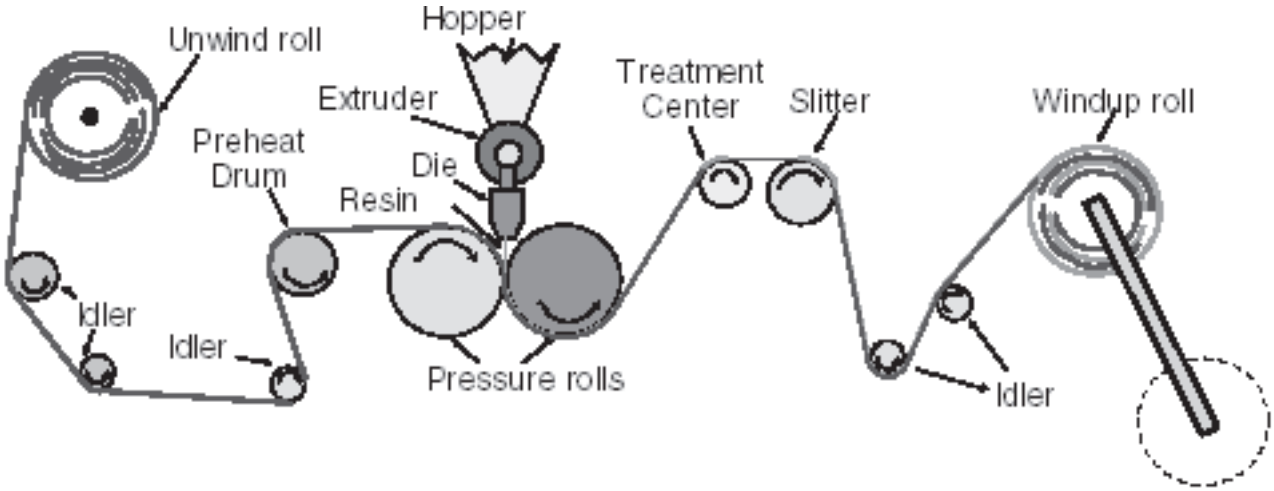

Figure 47.2. Equipment configuration for extrusion coating.

fy the surface chemistry, improving the ink adhesion. Corona discharge or plasma are standard treatment procedures, discussed previously in Part 7, Chapter 45, that add polarity to the polyolefin nonpolar surface. Printing or decorating can be done in-line between surface treatment and windup. Idler rolls are installed between the different secondary operations to provide tension and change web direction as necessary.

Extrusion coating operations use high melt temperatures to lower the melt viscosity. This improves coating thickness uniformity and adhesion. On-line gauging is used in some operations to measure the film thickness. Either an operator makes manual die adjustments or a microprocessor controls an auto die from the on-line gauge data. A second critical performance criterion is film adhesion to the substrate. Adhesion depends on

- Resin melt temperature
- Resin viscosity (the reason high temperatures are used)
- Film/substrate compatibility
- Coating speed
- Coating thickness

Chemical bonding, mechanical interlocking, or a combination of the two provides the film/coating adhesion. In chemical adhesion, the substrate and film are compatible, having similar polar or nonpolar chemistry at the interface or chemical groups that tend to bond together rather than repel each other. With mechanical adhesion, the substrate surface being coated is rough, providing a larger bonding surface area, with micro-undercuts or areas on the surface where the polymer can flow into and become mechanically attached on cooling. A third option is where both chemical and mechanical forces are positively binding the film and the substrate together. If the adhesion is inadequate under optimum processing conditions, it may be necessary to add an adhesive between the substrate and the film to provide the required performance. Typically an adhesive is a material that will chemically bond to both surfaces. The adhesive could be the top layer in coextruded film that is compatible with the substrate, producing good adhesion. A water- or solvent-based adhesive can be applied to the substrate prior to the preheat drum. The substrate can pass through a corona or plasma treatment to modify the surface chemistry and enhance adhesion. Corona or plasma treatment of a substrate being coated with a polyolefin resin is not anticipated to enhance adhesion to the polyolefin. Adding oxygen atoms to make the surface more polar will not improve adhesion with nonpolar polymers such as polyolefins.

Defects in the coating that can render the coated product useless are

- Voids
- Pinholes
- Thick or thin coating in the machine direction
- Orange peel
- Contamination due to gels or foreign material

Voids are caused by poor adhesion between the coating and substrate, where the two materials are not properly bound together. Pinholes are tiny holes in the coating. Voids and/or pinholes may be caused by excessively high coating speeds that are drawing the polymer melt too much. The other defects (gels, oxidized or burnt particles, orange peel) are the result of poor extrusion conditions or raw material supply. The extrusion operation has to have a properly adjusted die with uniform and constant melt temperature and pressure. Processing conditions may introduce too much shear heat that degrades the material, or stagnant places in the die may allow polymer to degrade over time. Polymer gels result in hard particles in the final film.

Blocking is when the laminated structure sticks to itself in the wound roll. Unwinding causes defects, ripping, or tearing. Insufficient cooling prior to winding, electrostatic charge, and sticky polymer surface cause blocking. When the product is not cooled prior to winding, the residual heat takes a long time to dissipate. This heat, plus high windup tension, bonds or seals the polymer film to the substrate below it on the reel. Electrostatic charges by themselves do not normally cause blocking; however, electrostatic build-up can magnify any blocking problems that already exist. Adding an antiblocking material, such as talc, to the formulation may eliminate blocking attributed to tackiness or when the roll is wound hot.

Lamination is similar to extrusion coating with the exception that two substrates are added to each side of the extruded film. Figure 47.3 demonstrates the basic lamination process. This can be considered a three-ply process, with two substrates and a molten film. If the film is produced on a coextrusion line, it can have multiple layers. A structure could be made with substrate A/three-layer film/substrate B, where the three-layer film might be an adhesive layer/film for structural integrity/adhesive layer to bind the entire laminate together. Building on the idea of multiple layers, a five-layer extrusion laminate might consist of substrate A/film 1/substrate B/film 2/substrate C. Figure 47.4 shows an extrusion configuration to produce a five-layer laminate. Adding coextrusion on both extruder 1 and extruder 2, very complex structures can be produced for food, medical, and industrial packaging applications.

Applications for extrusion coating and lamination include the following:

- Film lamination
- Heat seal layer used in general packaging
- Dairy packaging
- Juice and folding cartons
- Cups
- Paper
- Foil
- Carpet coating and backing
- Food pouches
- Cheese bags
- Can linings
- Photographic paper
- Potting soil bags
- Release paper
- Frozen food containers
- Paperboard trays
- Oven-safe paperboard trays

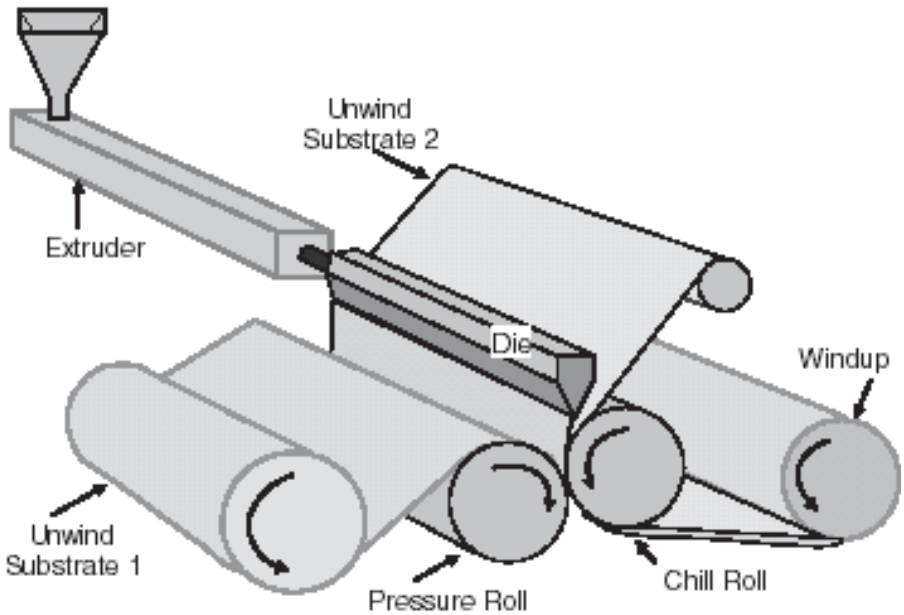

Figure 47.3. Extrusion lamination.

Another lamination process uses a lamination press after materials are combined in a thicker structure than film. As an example, Azdel®, a glass-reinforced thermoplastic composite made with continuous glass mat, is produced using an extrusion lamination operation. Film, continuous random orientation glass mat, and molten resin are combined in a lamination process using a Sandvik double belted press, shown in Fig. 47.5. Either continuous glass mat, made in a random configuration, or chopped glass strand mat, made with 0.5 inch (12.7 mm) or longer glass fibers, is added between layers of resin delivered either as film or from an extruder. The film layers are heated above their melting points in the hot section of the press, where pressure is applied to force the resin through the glass mat, impregnating the individual glass bundles and the glass fibers within each bundle. The top and bottom belts move at the same speed, carrying the product through the heating cycle, impregnating the glass, and making a uniform glass/resin distribution in the product. In the continuous process, the

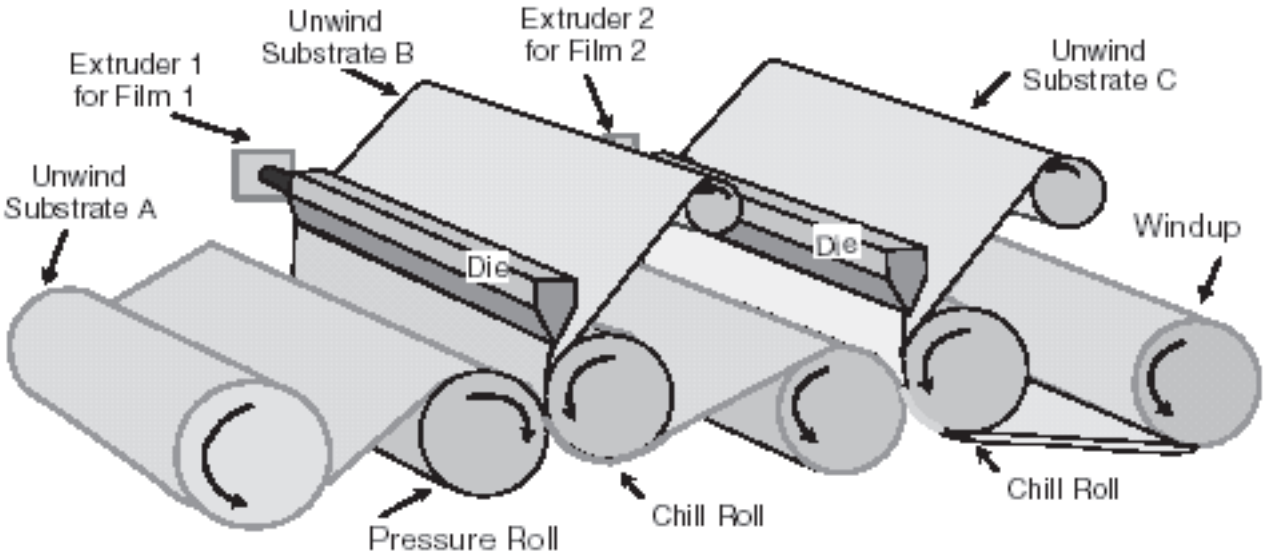

Figure 47.4. Five-layer lamination with three substrates and two films.

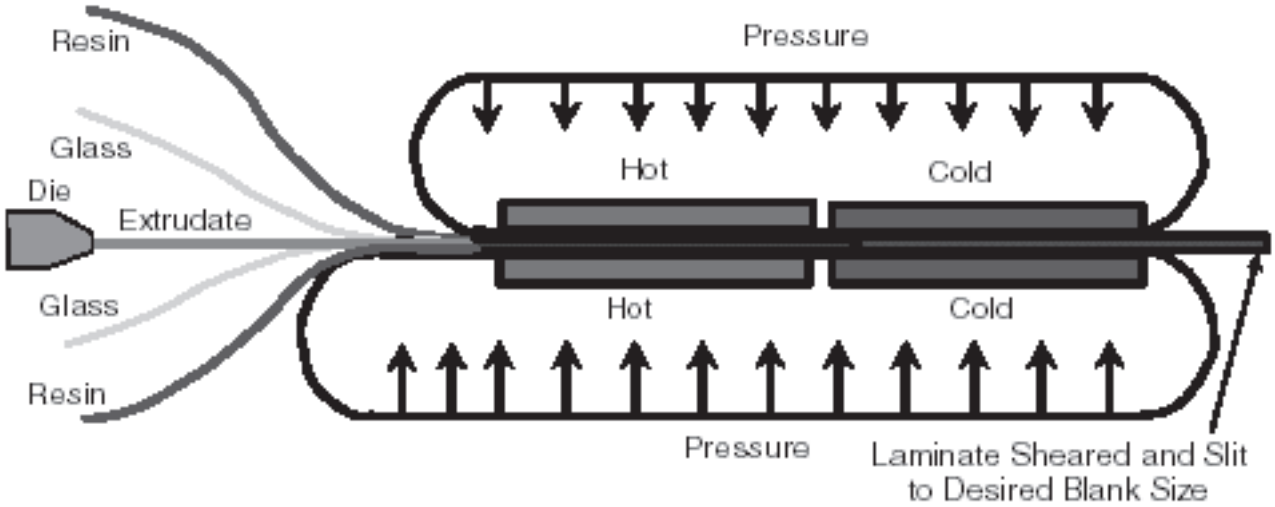

Figure 47.5. Double-belted press laminator used to produce thermoplastic composite laminate sheet and blanks.

product is then cooled below the resin melt temperature. As the product solidifies, it releases from the belt and is drawn out by puller rolls. After the puller rolls, the product is slit to the desired width and cut to the desired length. Blanks are used in compression molding operations to produce high performance composites with continuous glass mat. Blanks are preheated in an oven to 40–100°F (22–56°C) above their melting point if crystalline resins are used, and 150–200°F (83–111°C) above their glass transition temperature if the resin is amorphous. After exiting the preheat oven, blanks are rapidly stacked and transferred to the mold in the compression press. The press is closed very rapidly, and the resin and glass flow in a uniform distribution to fill the mold. After cooling, the part is removed and any necessary secondary operations are performed.

Review Questions

1. What is the difference between extrusion coating and laminating?
2. What materials are used as substrates and films in extrusion coating, and what are some of the applications?
3. What is the principal extrusion process used in extrusion coating or lamination?
4. Why might corona or plasma treatment of the surface be necessary in extrusion coating?
5. What are some possible defects and extrusion problems that might be encountered in extrusion coating operations?
6. Why is the polymer melt temperature normally run on the high side in extrusion coating operations?
7. Why is preheating used in extrusion coating?
8. What are the two types of bonding mechanisms that might be present in extrusion coating, and how might each be optimized to provide higher strength?
9. How does an adhesive layer function?

48 Wire and Cable Coating

Coating wire and cable with a polymer is accomplished via a crosshead extrusion process. The wire or cable to be coated is pulled at a constant rate through a crosshead die, where molten plastic covers it. Most wires and cables used in electrical and telecommunication applications and the electronics industry are coated this way. Two extruders in tandem can be used to apply more than one coating. Consider the production of standard two-wire-plus-ground cable used in home wiring. Initially one copper wire is coated with white insulation, while a second wire is coated with a black insulation. In a secondary process, the white and black insulated wires are combined with a third, paper-wrapped copper wire used for the ground; all three wires are fed through a die, where the final insulating jacket is applied to protect and hold all three wires together in a common plastic sleeve for use in home wiring. An alternative approach to producing this product is to do it in a continuous process where two extruders individually apply the white and back coating to the two conductive wires. These two products are then brought together with the third, copper ground wire, and the three wires are then sent through a third crosshead die, where the exterior jacket is added. With all three extruders running at the same speed, the final product can be produced with minimal handling. A downside of this process is the production loss if any one of the three extruders is not running properly or a problem is encountered with one extruder and not the others.

Wire coating is normally done using single screw extruders. The extruder function is to melt the resin and deliver it to the die at a uniform and constant melt temperature and pressure. Depending on the formulation, the only mixing required may be adding colorant to provide the correct color. Telephone wire contains many very small wires with different color insulation to prevent wires from getting crossed during installation. Telecommunications cables using fiber optics are produced using the same process. Dies in wire and coating operations are normally run at high pressure because the line speed requires high throughput rates coupled with a very small die opening.

Figure 48.1 shows a block diagram of a crosshead extrusion operation with typical equipment in the line. The basic pieces in each line are similar:

- Unwind station or other wire or cable source to feed the line
- Pretensioning station to set the tension through out the process
- Preheat station to prepare the wire for coating
- Crosshead die
- Cooling trough to solidify the polymeric coating
- Test stations to assure the wire is properly coated
- Puller to provide constant tension through the process
- Winder to collect the product

A crosshead extrusion operation has the extruder set at a right angle to the wire reel and the rest of the downstream equipment, as shown in Fig. 48.1. Wire enters the die at a 90° angle to the extruder, with the polymer entering the side of the die and exiting at a 90° angle from the extruder. Figure 48.2 shows a line described in Fig. 48.1 from ground level.

When there are multiple coating operations, where multiple wires are brought together, the extruders can be placed on the side with the extrudate running down the center, as shown in Fig. 48.3. Figure 48.4 shows another potential configuration for applying multiple coatings. Coextrusion with a second small extruder is used to apply a strip down one side of the wire and is referred to as a stripper extruder. The configuration used in a particular operation depends on the coating objective. Line lengths can be very long with sophisticated tensioning devices at the beginning and end to accurately control the tension and speed. Large cables obviously require larger tension devices, pullers, and winders compared to small gauge wire.

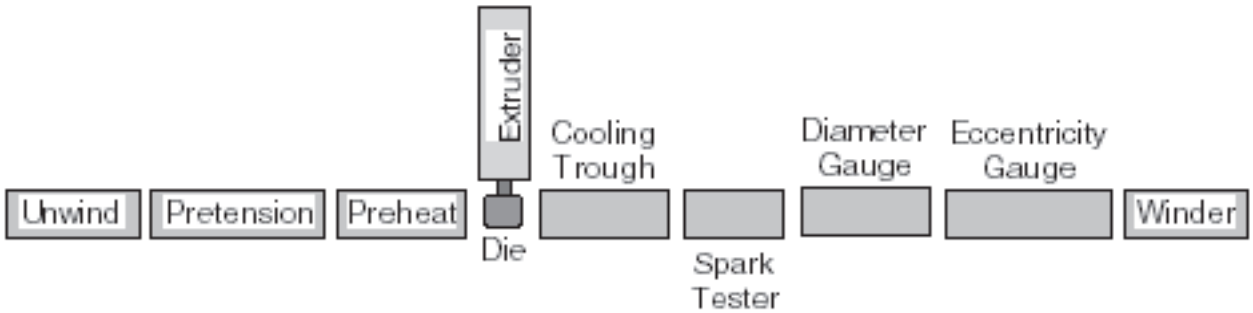

Figure 48.1. Block diagram of a wire coating line.

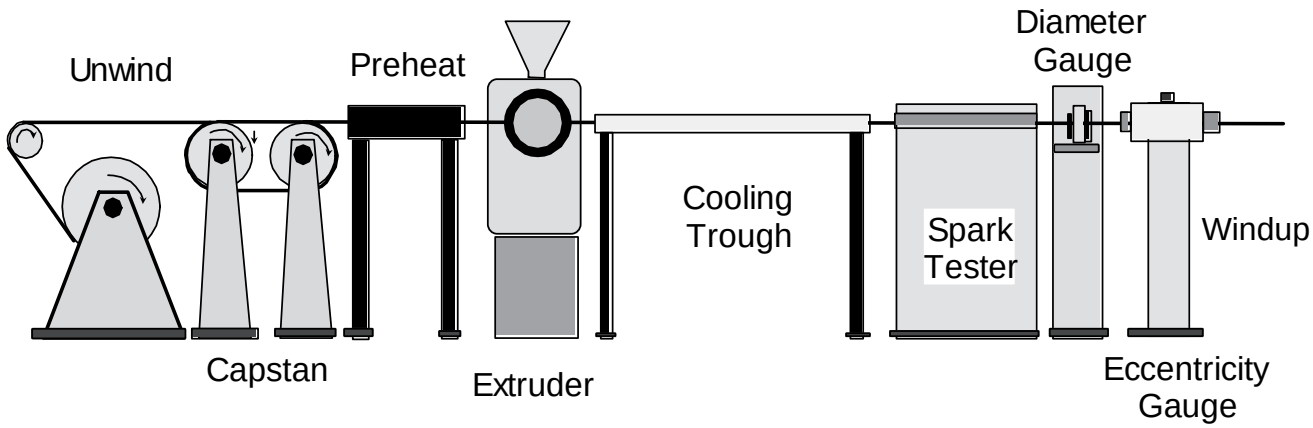

Figure 48.2. Crosshead extrusion process for coating wire and cable.

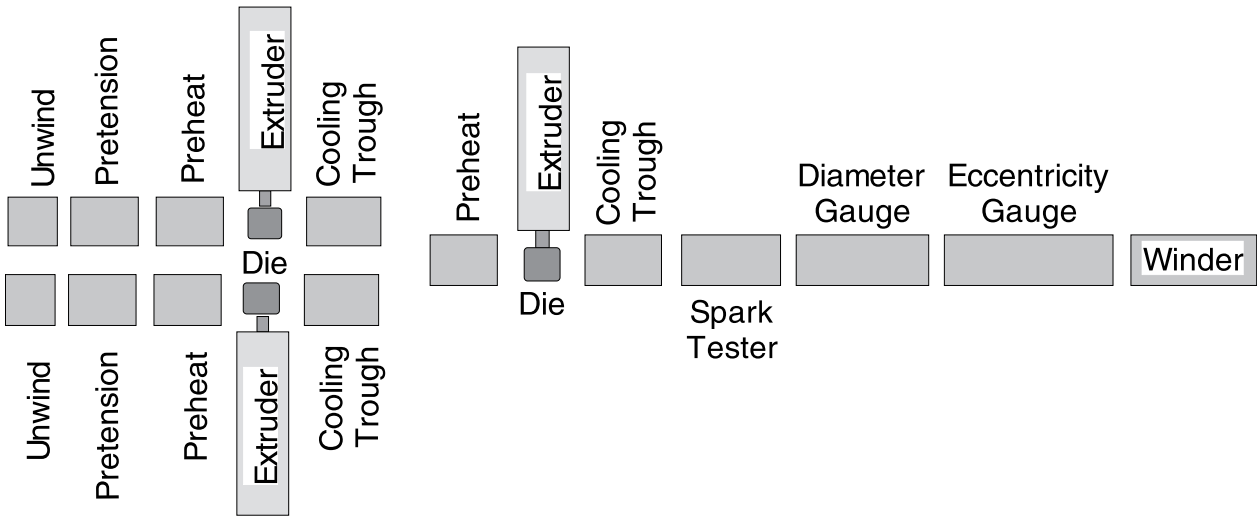

Figure 48.3. Equipment diagram with multiple coating operations in one line.

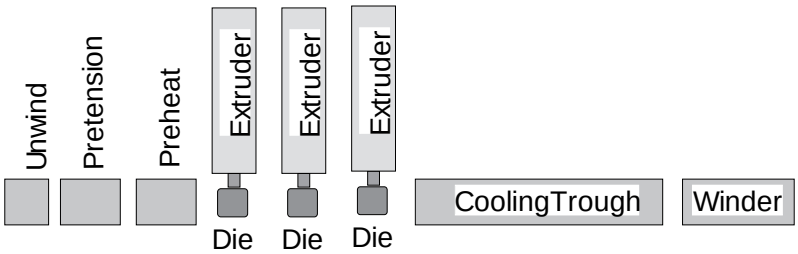

Figure 48.4. Multiple coating operations with additional extruders.

This chapter will now cover the extrusion process and equipment used in wire coating operations from the extruder die through windup.

48.1 Pretreatment

The object to be coated can range from a single wire to multiple wires made of copper, aluminum, glass fiber (fiber optic cable), or other continuous length metal fibrous material requiring an insulative or protective coating. The material to be coated can be wound on spools or reels and placed at the end of the extrusion line, or it can be brought directly from another part of the plant in a continuous operation. If the material to be coated is supplied on a spool, there are two potential ways to remove the wire from the spool. The first option is to use a stationary creel with the wire being pulled continuously off the end of the spool. This has the disadvantage of putting a twist in the wire for each revolution pulled off the spool. The second option, which is normally used, is to have a spool or reel unwind where the reel turns continuously as it pays out the wire. This supplies a wire to the extruder with no twist. Most coating operations require very long sources without any breaks. Fiber optic cable may have to be 5 to 10 km in length without any breaks or other defects in the product. Regardless of the wire or cable source, it has to be continuous in many wire coating applications to produce a specific length that is free of defects.

Once the supply source is established, the wire or cable is wrapped around a capstan or other tensioning device to provide wire at constant speed and tension to the process. The capstan size depends on the wire or cable being coated.

Finally the wire passes through a preheater prior to the die to bring the wire up to the temperature of the plastic used to coat the wire. Heated wire provides better adhesion between the wire and the plastic coating and expands the wire, reducing the shrinkage difference occurring between the wire and the coating on cooling. The plastic coating will always shrink more than the wire, because the plastic's coefficient of thermal expansion is greater than that for metal or glass. However, heating the wire brings the overall wire and coating shrinkage closer together on cooling. Another advantage of heating the wire is to help maintain the die temperature during normal operations. Cold wire passing through a die at high speed is a tremendous heat sink. Finally, preheating can be used to remove any moisture or other contaminants (such as lubricants left on the wire from a wire drawing operation) from the wire surface that might interfere with adhesion to the plastic coating. Preheaters are normally either gas or electrical resistance heat and are designed to heat the wire to the melt temperature of the plastic being applied to the wire or just slightly below the melt temperature.

48.2 Die

Figure 48.5 shows the designs of a pressure die and tubing die used in wire coating operations. The pressure die coats the wire inside the die, while the tubing die coats the wire core outside the die. The core tube, also referred to as the mandrel, is used to introduce the wire into the die while preventing resin from flowing backward where the wire is entering. Mandrel guide tip tolerances in a pressure die are approximately 0.001 inch (0.025 mm). This tight tolerance plus the forward wire movement prevents polymer backflow into the mandrel even at high die pressures. The guide tip is short, allowing contact of the polymer and the wire inside the die. The distance between the guide tip and the die exit is called the gum area. A spider ring with three to five legs holds the mandrel or core tube in place. The polymer melt divides, flows around each leg, and recombines as it flows toward the die exit. This causes a potential coating defect at each spider leg, called a weld line or knit line. Without proper chain reentanglement after the spider leg, the coating has less

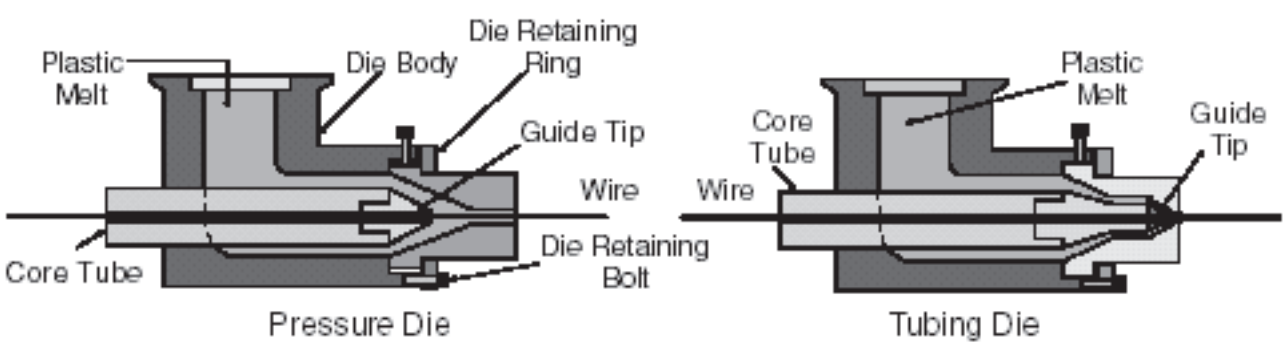

Figure 48.5. Pressure and tubing dies to coat wire and cable.

toughness and durability at the knit line. Process changes to improve knit line strength are

- Higher melt temperature (more molecular motion)
- Higher pressure (forces the polymer chains together)
- Longer residence time after recombining (more time for chain reentanglement)

Lengthening the land area at the die exit increases the residence time. Figure 48.6 shows a flow channel cross section. The open areas are the spider legs, and the annular darker area is the melt flow channel. As the wire is pulled through the die at high speed, it acts as a pump, pulling the polymer out of the die. Typically the die opening is similar to the final wire diameter.

Figure 48.6. Flow around spider supporting mandrel.

The centering ring at the die exit has adjustments to keep the wire centered within both the flow channel and coating.

In the tubing die, the polymer and wire are combined outside the die. As the wire exits the mandrel and guide tip, the polymer coats the wire as it is pulled away from the die. To obtain uniform melt thickness around the wire, the die tip must be properly centered, with the wire running in the center of the opening.

Spiral mandrel dies are common in blown film and pipe applications, and they are being applied to wire coating. Spiral mandrel dies do not have weld or knit lines and have better product concentricity. In 1998 a successful wire coating spiral design die was demonstrated.[1]

One die design parameter to consider is the draw down ratio (DDR) resulting from the changing cross sectional area in the die. Draw down ratio is the melt channel cross sectional area divided by the coated product cross sectional area. It is calculated from Eq. (48.1):

$$DDR = \frac{DD^2 - TD^2}{OD^2 - ID^2} \tag{48.1}$$

where

DD = Die diameter
TD = Tip diameter
OD = Outside wire diameter
ID = Inside wire diameter

as shown in Fig. 48.7.

The draw down ratio measures the polymer melt size reduction. The draw down ratio to use is normally based on past experiences: what has worked previously or what

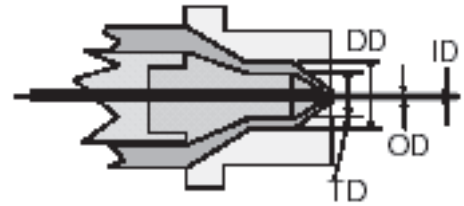

Figure 48.7. Definition of terms for draw down ratio.

the resin manufacturer may recommend. In some cases the draw down ratio is not specified; a multiplier is specified that states how much larger the die channel is relative to the final wire dimensions. Larger draw down ratios allow higher output rates at the expense of overall product dimension control. Lower draw down ratios produce a more stable process.

A second factor in die design is the draw ratio balance, which is the rate at which melt outside the die is drawn down, compared to inside the die at the tip area.[2] The cone shape at the end of the die causes polymer melt near the center to move at a different rate than the polymer at the outside diameter of the cone. Referring to Fig. 48.7, if the ratio of DD/OD is the same as the ratio of TD/ID, the draw ratio balance is 1. However, if the ratio of DD/OD is much greater than TD/ID, the die is out of balance, which leads to processing problems. A slightly positive draw ratio balance (1.03–1.08) provides good processability in wire and cable operations, while a negative draw ratio balance should be avoided.

Dies in wire and coating operations can be quite simple when coating a single wire or group of wires. However, complexity can increase substantially as three wires are coated simultaneously, with one extruder producing three individual jackets that are bound together in a shape similar to that shown in Fig. 48.8. Die complexity increases further as coextrusion is added to stripe the products or make the core with different color resins. In long lines with multiple extruders, braiding, support wires, and metal shielding can be added to the product between extruders to produce a value-added wire or telecommunications cable.

Figure 48.8. Complex wire coating cross section.

48.3 Downstream Equipment

After exiting the die, the polymer coating is cooled in a water trough, where the water is applied uniformly on all sides of the wire coating to prevent differences in resin

shrinkage around the wire. The preheater, die, and water cooling trough must be properly aligned to prevent the wire from moving up or down or sideways during these stages. Proper die adjustment plus alignment promote uniform coating around the wire and uniform shrinkage in cooling. If the wire is pulled in a different direction as it exits the die, it puts one side of the coating in tension and the other side in compression and may affect the product performance later in its life cycle.

Normally there is an air gap between the die and the water trough where initial air cooling occurs. Cooling is then controlled very carefully to prevent vacuum voids from forming or excessive insulation shrinkback. Two water cooling troughs may be used with hot water in the first trough and cold water in the second. Hot water helps control the cooling rate, and vacuum void formation is reduced. In place of a cooling trough, a water tube might be used, where the cable passes through a tube that is filled with water. Tubes are more practical if the environment tends to be dirty and final product contamination is an issue.

Shrinkback occurs when the insulation pulls back excessively from the wire after the wire is cut or exposed to heat. Remember, this is a continuous process and polymers shrink more than the materials that are coated. Consequently, if the polymer is not tight around the substrate, it will want to relieve any tension by sliding back from the wire end as it is cut. The shrinkback can be calculated by heating a piece of wire in an oven at a specific temperature and time and measuring the shrinkage associated with a given length. Factors affecting shrinkback are

- Wire preheat temperature prior to the die—To reduce shrinkback, increase the preheat temperature.
- Water bath is too close to the die or it is too cold—If cooling occurs too rapidly vacuum voids can be formed.
- Polymer melt temperature
- Draw down ratio
- Poor core substrate adhesion from foreign contamination.

After cooling, the wire can be passed through on-line gauges for quality control. Three different gauges are normally used to measure product quality:

- Diameter gauge
- Eccentricity gauge
- Spark gauge

The diameter gauge measures the wire diameter. If the diameter is too large, speed up the puller or slow down the extruder screw. If the diameter is too small, do the opposite. The eccentricity gauge measures the coating uniformity around the wire. It is important to have the same polymer wall thickness around the wire. The concentricity is adjusted by centering the guide tip with the adjusting bolts. Finally, the spark tester checks for pinholes in the coating that can cause electrical shorts or carbon deposits in the polymer that can cause electrical conductivity through the coating. The three gauges may be installed in any order on the line.

A printer can be installed in-line to print the wire codes. If printing is a problem on polyethylene coating, it may be necessary to install a plasma or corona treatment station before the print station to improve wet-out and adhesion.

Finally a capstan, caterpillar-type puller, or other pulling device is installed to provide constant line speed and tension during processing. A capstan is normally used with small diameter wire, where the wire is wound around a large diameter reel run at constant speed numerous times to provide a uniform pulling speed. A caterpillar-type puller with belts is used with large diameter wire. Sufficient pressure has to be applied to prevent the wire from slipping, providing uniform speed to the winder.

Typically, two center winders are required in a continuous operation, with one winding up the product while the second waits in reserve for the first spool to be completed. Once the first spool is complete, the wire is transferred to the second spool as the first one is being emptied and prepared for the next reel.

48.4 Polymers

Typical polymers used in wire coating applications are

- Polyethylene
- Polyvinylchloride
- Polyamide
- Polybutylene terephthalate
- Thermoplastic elastomers
- Ethylene propylene copolymers
- Polypropylene
- Fluoropolymers

These polymers are chosen because of their electrical properties, flexibility, durability, and ability to withstand abuse.

Cross-linked polyethylene is used in wire coating operation. Wire is extrusion coated with polyethylene in the process described in this chapter. After the coating operation is completed, the polyethylene is cross-linked on the wire. There are two ways to cross-link polyethylene:

- Polyethylene containing a crosslinking agent such as peroxide or sulfur is extruded and applied to the wire as a coating. Once outside the die the peroxide initiator is activated with steam, nitrogen, or pressurized liquid where the polymer is heated causing the peroxide to decompose forming free radicals on the polymer backbone. At elevated temperature the polymer free radicals react to form a crosslinked polymer. This process is referred to as vulcanization or curing. In the crosslinked state the polyethylene no longer melts and has high temperature resistance.
- Polyethylene can also be cross-linked via radiation. The coated wire is passed through an electron beam that generates free radicals on the polyethylene backbones. They react at elevated temperature to cross-link the polymer, forming a thermoset.

Regardless of the cross-linking process used, the operation is completed outside the extruder after the wire has been coated. This places another operation in the overall wire coating line for cross-linking.

REFERENCES

1. B & H Tool Company Inc., "Understanding and Communicating Draw Ratio Balance," B & H Extruder, 2(3), San Marcos, CA, 1999.

Review Questions

1. Name two different die designs used to produce wire and cable, and describe how they differ.
2. Why is a crosshead die used in wire and coating operations?
3. What is the purpose of preheating the wire prior to entering the crosshead die?
4. What is the function of capstans at the entrance and exit of the extrusion line?
5. What does an eccentricity gauge measure?
6. What are draw down ratio and draw ratio balance, and why are these important in wire coating?
7. How is cross-linked polyethylene applied as a wire coating, and what are some methods of cross-linking the product?
8. What polymers are used in wire coating operations?
9. How is a coated wire cooled after extrusion, and what are some important parameters in the cooling process?
10. How are multiple extruders set up to apply more than one coating to a given wire or group of wires?
11. What type of winders are used to spool or creel the product?

49 Profile Extrusion

Profile extrusion is one of the most common types of extrusion processes, producing a variety of product shapes in large volume. Any plastic cross section requiring long lengths can be produced in a profile extrusion process and cut to the desired length. The process uses

- Either single or twin screw extruders equipped with a specific die configuration to produce the desired profile
- Cooling or calibration tank
- Puller
- Cutter

Other secondary operations can be included in the process, such as

- Printing
- Decorating
- Combination with another product that requires adhesive application and bonding
- Attachment of fasteners or some other posttreatment

Profiles can be more difficult to run because there are no die adjustments, and the different shapes associated with each product require modifying downstream equipment. Where sheet dies have adjustable lips and choker bar for on-line gauge adjustment, a profile die has a fixed opening requiring a specific draw down ratio, with only one speed ratio between the extruder and puller or haul-off that provides the correct product dimensions. While profiles can be produced with any resin system, the most common resins are

- PVC
- Polyethylene
- Polypropylene
- ABS
- High impact polystyrene

Rigid PVC is used to make parts for building and construction markets such as

- Window profiles
- Gutters
- Vinyl siding
- Fence posts
- Railings
- Plastic lumber

PVC profiles are commonly run on counterrotating twin screw extruders and particularly conical extruders, where PVC powder can be compounded with the required plasticizer concentration, lubricants, thermal stabilizers, fillers, and colorants to produce a commercial profile in one operation. Some extrusion profiles are shown in Fig. 49.1. Potential profile configurations are limited only by one's imagination and appropriate applications.

Figure 49.2 shows a profile line with an extruder, cooling tank, puller, movable saw to cut the product, and product packaging. The cooling tank can be a vacuum calibration tank or a cooling tank without vacuum, depending on the profile being produced. The movable saw can

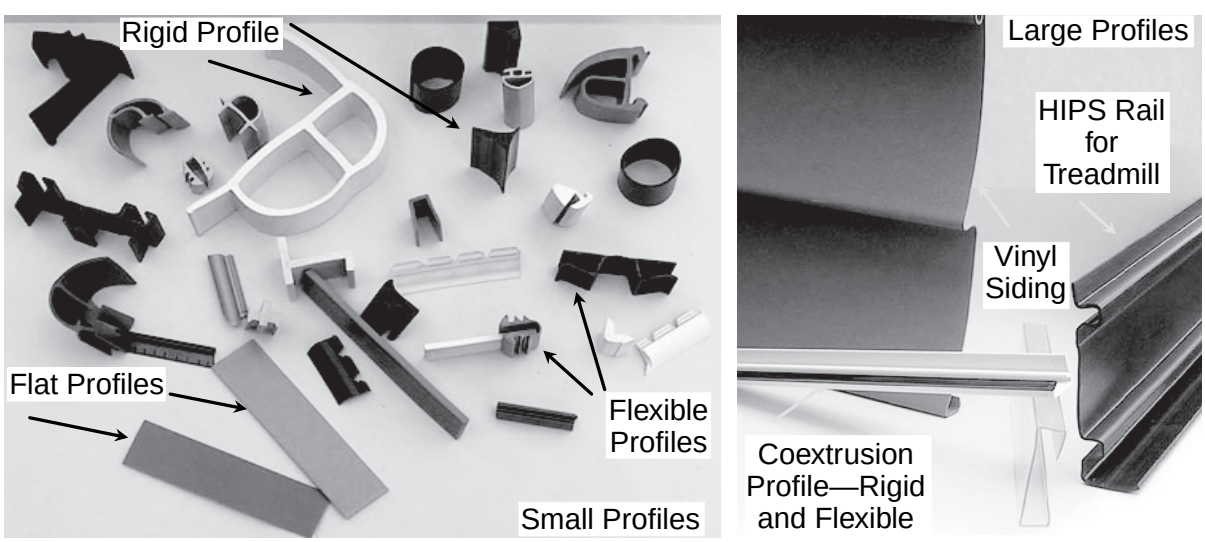

Figure 49.1. Various large and small profile shapes.

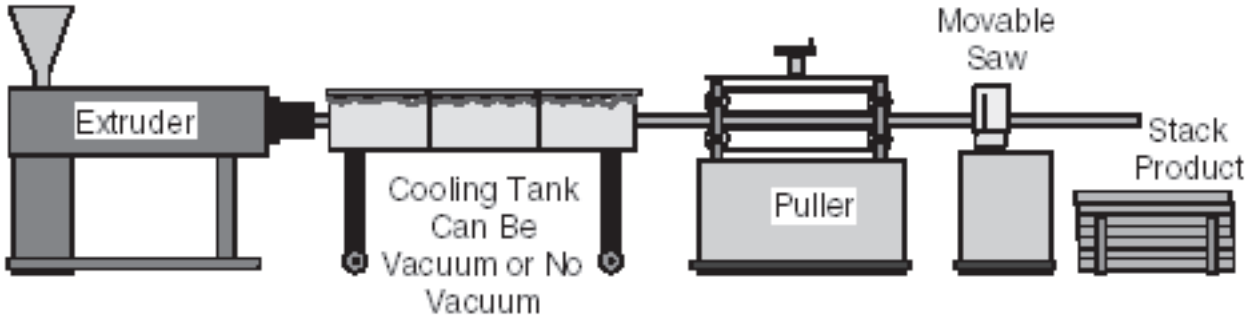

Figure 49.2. Schematic of equipment used in profile extrusion.

be replaced with an on-demand rotary cutter or guillotine cutter; as with the calibration and sizing tank, the cutter required depends on the profile size and the material.

We will now cover the equipment used in profile extrusion and its operation from the die through the end of the line.

49.1 Die

One of the keys to successful profile extrusion is combining good profile cross sectional designs with good die design. Frequently, die design for profile extrusion is done using trial and error methods in-house by the company producing the profile. A good die designer can machine the correct die shape to produce the specified profile in three to four die cuttings or modifications, while an inexperienced die designer may have to machine seven to eight dies to obtain the desired profile. Some companies specialize in computer-aided die designs based on computer modeling. With the desired profile cross section defined, the rheological parameters (viscosity versus shear rate, extension modulus and flow modulus) for the resin and the throughput rate, computer programs can design the die cross section required, while predicting the pressure drops, melt temperature, and flow profiles within the die.

The die has to be properly designed to have uniform flow and pressure drops in all the legs of the profile. Improperly designed dies can lead to severe warpage problems associated with the profiles. Higher polymer flow rates in one section of the die compared to another will result in different molecular orientation. This causes different shrinkage characteristics from one section to the other. Melt temperature in all channels needs to be similar to prevent one profile section from being hotter than another. This can lead to differential cooling outside the die and differential shrinkage and warpage. Another factor that can lead to warpage in the final part is different molecular packing in one part of the die versus another, attributed to higher die pressure in one area compared to another. Part areas with higher molecular packing produce less shrinkage compared to other part sections. These factors are normally grouped under a general heading of internal stresses in the final part. And nonuniform internal stresses can cause warpage in the final profile because the shrinkage is different between one profile section relative to another.

Polymer melt pressure and flow in the various die sections can be kept constant through uniform wall thickness and proper design. Figure 49.3 shows some different cross sections with two different geometries, one being preferred over the other. In all cases the preferred geometry has uniform wall thickness around the

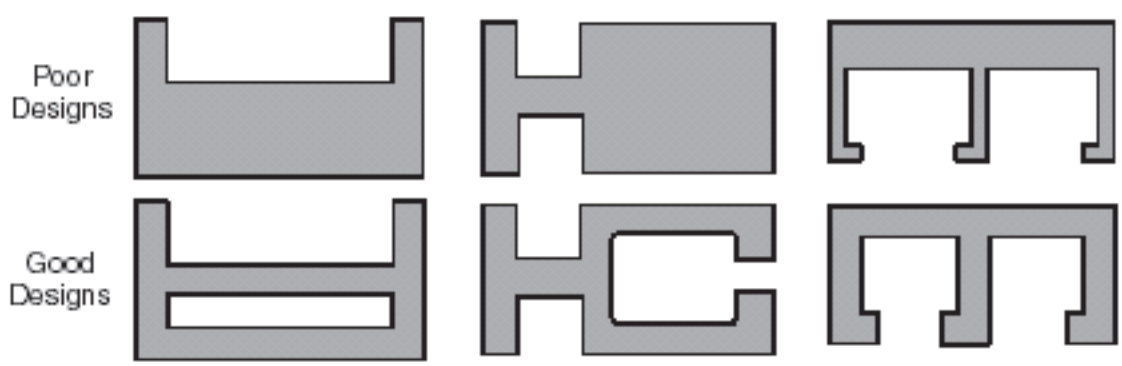

Figure 49.3. Comparison of profiles with uniform versus nonuniform cross sections.

entire part. Depending on the profile, constant wall thickness over the entire part is not always possible. The resistance to flow in a die channel is given by Eq. (49.1):

$$R = \frac{\Delta P}{\eta \times Q} \tag{49.1}$$

where

R	=	Resistance to flow
ΔP	=	Pressure drop
η	=	Melt viscosity
Q	=	Volumetric throughput rate

Since the volumetric throughput of the different channels has to be constant to prevent warpage, $Q_1 = Q_2 = Q_3 = \ldots Q_n$. If the melt temperature is the same and the shear rates are similar, the melt viscosity in the different channels will be the same. This leaves the pressure drop to balance the resistance in each section or leg of the die. The pressure drop is calculated from Eq. (49.2):

$$\Delta P = \frac{2 \times \tau \times l}{h} \tag{49.2}$$

where

τ	=	Shear stress = force/area
l	=	Channel length
h	=	Channel height

If the cross sectional area of each output section or leg is the same, the pressure drop in each leg will be identical, assuming the rest of the die has a uniform melt flow. To change the flow resistance in different legs, the land length is modified, making it longer in thicker profile sections and shorter in thinner profile sections to produce equivalent pressure drops. If it is possible, it is better to maintain uniform wall thickness in all areas of the profile.

Sink marks opposite ribs are eliminated with part design. Sink is caused by shrinkage. Extruded profiles with thicker cross sectional areas shrink more due to the higher mass than adjoining thinner areas. This creates an imperfection opposite a rib that is readily visible. Figure 49.4 shows a T-type profile and the corresponding sink mark. Eliminating the sink mark in the final part is accomplished by making the area opposite the rib slightly thicker, so with shrinkage it becomes the same thickness as the rest of the profile. The profiles shown in Figs. 49.3 and 49.4 should have radii on all the inner angles. In addition the die profile shown in Fig. 49.4 has to be cut different than the part profile to get flat sections on each leg. Stress relaxation as the polymer molecules exit the die land area and are no longer bound by the die cause the extrudate to swell. This extrudate swell phenomenon is commonly and incorrectly called die swell.

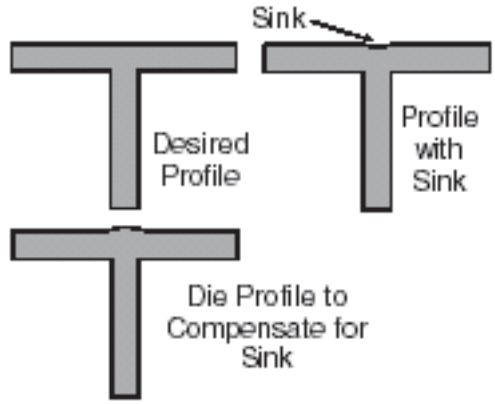

Figure 49.4. Effect of sink and die modification to eliminate sink.

Extrudate swell, discussed previously in Part 1, is attributed to the elastic portion of the viscoelastic behavior that makes the polymer want to spring back after the polymer exits the die. In the die land area, the polymer flow simulates the plug flow in a pipe, where the polymer velocity at the wall is zero. After the polymer exits the die, the velocity profile is the same at the surface as it is in the center. This is shown in Fig. 49.5. Extrudate swell requires the die to be cut differently than the actual profile desired. A die to produce a square rod is shown in Fig. 49.6 versus a die cut with a square hole and the resulting profile.

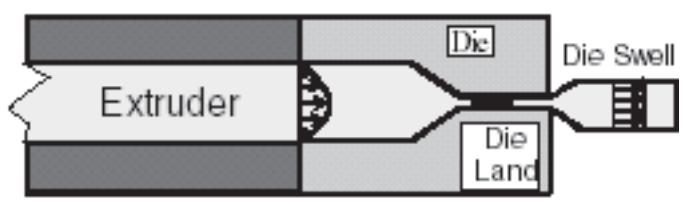

Figure 49.5. Die swell as polymer exits die.

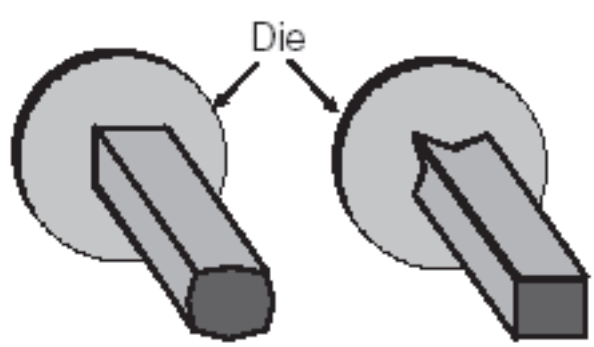

Figure 49.6. Die configuration to produce a square profile.

Proper die design is required to produce uniform parts at high rates without warpage. Design criteria include the following:

- Thicker walls require more time for cooling, leading to lower production rates. Calculate the wall thickness required based on the product performance requirements. Ideal wall thickness is approximately 0.03–0.12 inch (0.7–3.0 mm).
- Wall thickness can be reduced and still provide the same stiffness with the incorporation of ribs on the back side or the use of corrugated structures.
- Uneven wall thicknesses lead to uneven flow in the die and differential profile cooling that can lead to warpage, bow, and twist in the final part. If more than one wall thickness is present in the part, make the transition from one wall thickness to the next gradual.
- Ribs lead to sink marks caused by the extra mass and the polymer shrinkage. Eliminating sink marks can be accomplished by building the profile up opposite the rib or making the rib smaller so there is less material. In any event, generous radii between the rib and the rest of the part are required to provide the necessary strength.
- All inner and outer corners need to be designed with proper radii, eliminating sharp corners. Sharp corners generate high stress areas that can cause premature part failure.
- Eliminate through-holes in the profile wherever possible; these are expensive from both a tooling perspective and the potential need for vacuum sizing. It is better to have through-holes in a profile than uneven wall thicknesses. Uneven walls can lead to more significant processing problems.

A profile die is attached to the extruder with an adapter that funnels the melt to a die cross section that is slightly larger than the profile cross section. Three die designs are used, depending on the production run length and the polymer being processed:

- Flat back
- Semi-streamlined
- Streamlined

Flat back dies are used for low volume or prototype runs four to six hours long. A flat back die is a flat plate with the desired profile cut into its center and mounted to a round tube that conveys the molten polymer from the extruder. The die plate is flat on the side facing the extruder and is bolted directly onto the tube comprising the die body. The negatives associated with a flat plate die are dead spaces in the die body around the front plate, where molten polymer can stagnate and degrade, and poor flow patterns within the die body. A semistreamlined die, shown in Fig. 49.7, has a tapered channel in the die body that compresses the polymer flow as it approaches the die plate. While dead and stagnate places are still present, the area is greatly reduced where polymer degradation can occur. The best die has a streamlined polymer melt channel that leads directly to the land area with no dead spots for polymer stagnation and degradation. There is no dead space within the die where polymer can hang up or stagnate and degrade over time. The die is designed to produce uniform flow velocity in all the channels with less extrudate swell and melt fracture. Streamlined dies are used in long production runs where polymer cannot be allowed to degrade. Obviously, semistreamlined dies are somewhere between the two die extremes. Some dies are made with inserts allowing different shapes or slight modifications of a basic profile to be made more easily.

Key factors in die design are the land length and the draw down ratio. The land length is defined as the uniform flow channel just prior to the die exit. The land lengths within the die determine the flow resistance for the different profile sections or channels. Thinner profile sections require less land length than thicker sections in the profile to produce the same flow resistance. Longer land lengths are used to produce profiles with higher gloss. The proper land length depends on the die gap, the orifice area, flow rates, melt viscosity, and the optimum pressure for the die and extruder. Just selecting the land length as some multiple of the gap (assume 10, 20 or 30 times the gap) ignores the factors that are critical in a given die and profile. Die land lengths are recommended by resin manufacturers, and some guidelines are provided in Table 49.1.[1] These should be considered as guidelines only, as each die is different. The best way to establish the land length is to provide sufficient land to generate 1000 to 1500 psi across the land.

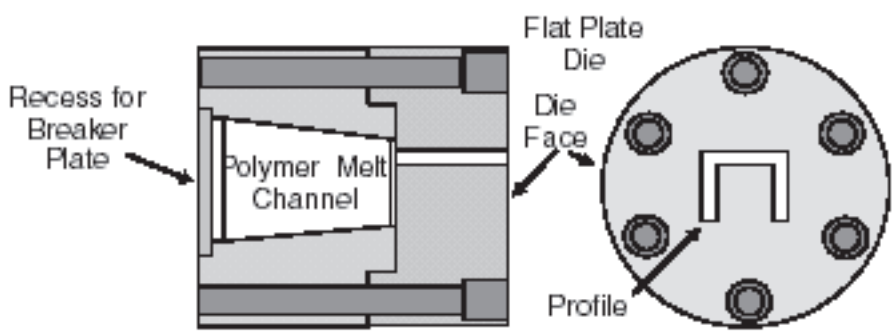

Figure 49.7. Semi-streamlined die.

Table 49.1. Recommended Draw Down Ratios and Land Lengths

Material	Die Land Length*	Draw Down, %
LDPE	15:1	30
MDPE	15:1	25
HDPE	15:1	20
PP	15:1	30
PPVC	15:1	5–25
UPVC	20:1	4–10
GPS	20:1	8–20
ABS	25:1	25
Nylon 6,6	20:1	15–20
Nylon 6	20:1	20
PC	10:1	15–25
Noryl®	20:1	25–40

*Land length to thickness ratio

Draw down ratio is the draw between the die and the final shape. As polymer exits the die, it is drawn by the puller and shaped by a sizing device to produce the desired shape. The draw determines the final molecular orientation in the part and the product performance. Higher draw ratios yield higher properties in the machine direction and lower properties in the transverse direction. With a specific cross section, the throughput rate is matched to one puller speed to produce the desired profile dimensions. Increasing or decreasing the puller speed results in the part being too small or too large. Table 49.1 provides some draw down ratio guidelines. The draw down ratio in a particular operation depends on the profile and the performance criteria. Since most profile dies do not have adjustable lips, the cutting of the die determines the draw. If a different draw is required to obtain the correct profile, a new die generally has to be cut. This is one reason three or four dies may be cut to produce the correct profile dimensions and a final part that meets all the customer specifications.

Die temperature control is another key variable affecting the profile and uniformity. The die has to have uniform heat to provide a uniform melt viscosity across the entire profile. Large profiles may require insulation around the die to maintain uniform die temperature. The die face may be a large area with only a single heater band around the outside edge to maintain the temperature. Changes in die temperature affect the melt temperature, with high temperatures producing lower melt viscosities.

Many profile dies are required due to the large variety of cross sections produced. Compared to other extrusion processes, there are more tricks to the trade and art involved, rather than science, in profile extrusion. In addition to the many profiles and resins used in profile extrusion, the process is further complicated by adding regrind or using regrind for the entire profile. Regrind is used to extend the resin in many applications. Figure 49.8 shows an American Maplan Corporation[2] single screw extruder producing tubing with a secondary extruder adding a stripe to the tube and a calibration tank to properly size the tube.

Die construction is usually made of 4140, 420, or 17-4 PH steel. All three steels have similar Rockwell hardness, but 4140 has poor corrosion resistance, with 420 having excellent corrosion resistance and 17-4 PH falling in the middle. 4140 and 420 can be hardened by heat treating and quenching. 17-4 PH is not practical for applications above 600°F (316°C) because it embrittles. For high gloss finish, the die land area is highly polished or can be chrome or nickel plated. Polishing to a very low micro-roughness or plating with chrome or nickel leads to extra expense.

49.2 Calibration and Cooling Tank

Cooling and sizing or calibration is used to set the part dimensions while preventing the part from warping and twisting. Part fixtures or sizing depends on the part complexity. The simplest setup has adjustable metal fingers shaping the soft polymer as it exits the die, with fans blowing air on the profile to cool it. The profile may be supported by wood forms or a metal rack. While this is the simplest way to cool a profile, it is also the most inaccurate, producing profiles with very low dimensional tolerances. For thermoplastics generally, more intricate sizing fixtures or calibrators are used with complex shapes in calibration tanks. The sizing unit may be a long metal fixture containing the desired profile shape where the molten polymer is pulled though and allowed to solidify, possibly in a water environment. As the part exits the die, it passes through a small air space and into the sizing unit or calibrator where it is cooled with air or water spray, flooded with water, or goes through a metal calibrator with circulating water and a heat exchanger to freeze or solidify the part to its final dimensions.

Air cooling can be used for simple shapes where tolerances are relatively unimportant. Metal fingers and jigs are used to bend or shape the profile to the desired configuration. One advantage of air cooling is it is relatively slow; consequently there is sufficient time to alter the profile with jigs to freeze it in the desired configuration. This is a quick approach to use with small simple profiles that do not have critical dimensions, such as thin flat

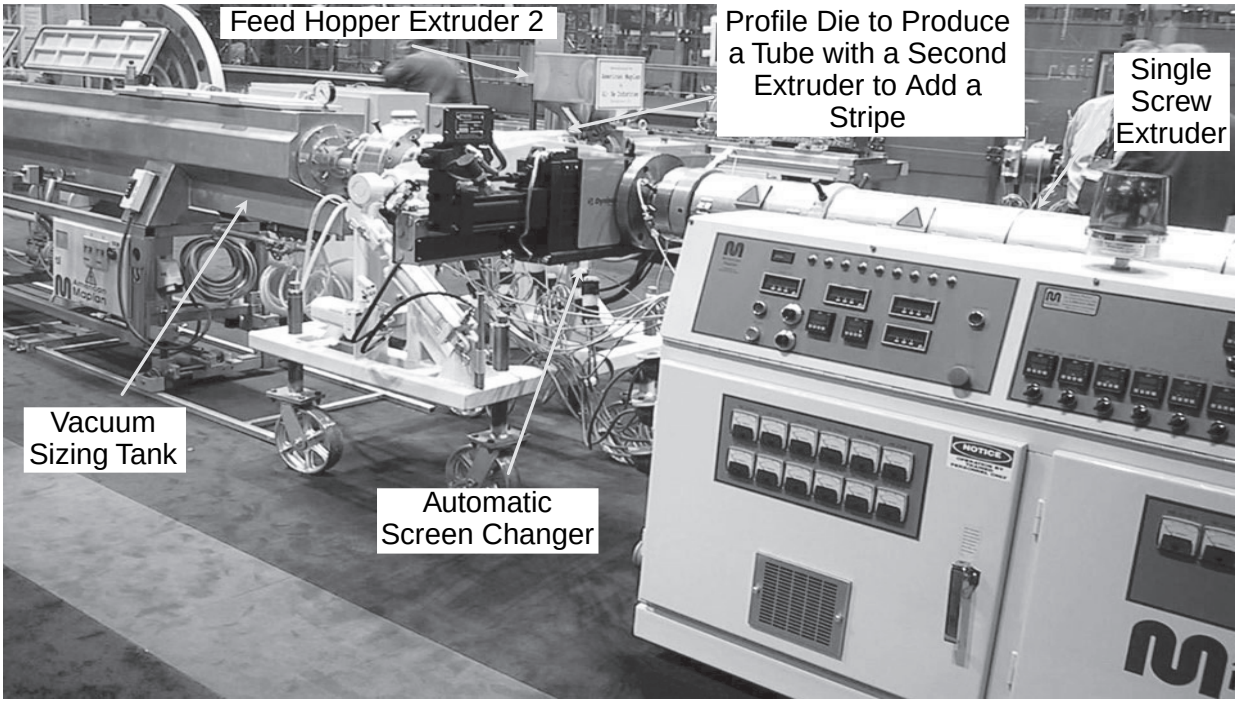

Figure 49.8. American Maplan profile coextrusion line with tubing die, stripper extruder, and vacuum sizing tank.

stock, vertical blinds, U- or T-shaped profile, and so forth. The most important criteria in these applications are a uniform melt temperature, good melt strength, and uniform cooling around the entire part. Air is directed onto the profile with adjustable position nozzles mounted at the beginning of the cooling section as the profile exits the die. Uniform cooling is required to eliminate warpage and twisting associated with differential shrinkage. Assuming the flow from the die is uniform and the profile is not being dragged across a surface in the cooling operation, profile bowing or twisting is frequently caused by nonuniform cooling. The profile will bend or cup toward the hot surface or the side that cooled last. This fact helps identify the causes and the proper corrective actions to take in modifying the cooling operation.

Water cooling is where the profile enters a bath below the water level and enters a fixture to size and hold the profile during cooling. The profile may pass through a ring spraying water on all sides simultaneously or water nozzles prior to entering the water tank. Water cooling is very efficient, is good for short runs, and may suffer in repeatability between short runs, depending on the fixture quality, its location, and the operator skill. Options available are the tank length; the number of chambers in the tank; whether the water is applied in the form of a mist or spray, flood, or a combination of both in different parts of the chamber; temperature of the water (hot or cold); self-contained recirculation systems; and heat exchanger.

A more sophisticated cooling system is vacuum sizing or calibration under water, where a hollow profile, whether round or square, exits the die and enters a sizing or calibration unit in an underwater vacuum tank. The water acts as a lubricant in the calibrator. Lower vacuum pressure enables the air in the hollow profile to expand uniformly outward until it comes in contact with the sizing sleeve or calibrator to produce the desired external dimension. Vacuum sizing units may have several compartments where the vacuum level can be adjusted and the water can either surround the profile or be sprayed on it at different temperatures. Normally the highest vacuum is in the first compartment, where the polymer is being expanded the maximum amount as the surface skin forms. Initially the profile will be larger than the final dimensional requirements to accommodate the polymer shrinkage as the polymer cools completely to room temperature. Figure 49.9 shows a vacuum sizing tank produced by Boston Matthews.[3] All tanks are fitted with lifters and adjustments allowing the tank to be move horizontally toward or away from the profile as it exits the die. The profile enters the tank passing down through the various compartments and exiting out the opposite end after it has been solidified in the correct cross section. Options available on most sizing tanks include heat exchanger, full spray cooling compartments, water circulation system, multiple compartments, vacuum pumps, additional vacuum or water manifolds to attach external calibration units for initial sizing, additional vacuum and

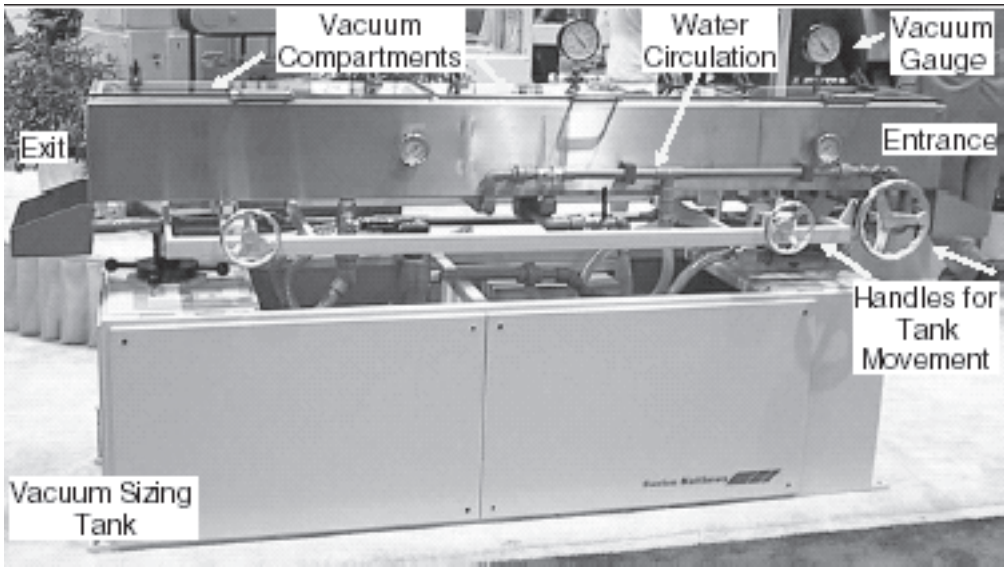

Figure 49.9. Vacuum sizing tank produced by Boston Matthews.[3]

water ports that can be connected to calibrators inside the vacuum tank, and uniform water or spray temperature.

Dry vacuum calibration is used to size complex, rigid shapes requiring a high dimensional precision and accuracy, such as window profiles. One or more calibrators or sizing fixtures are used with a secondary spray cooling tank. Enough heat is dissipated in the calibrator to hold the profile dimensions after exiting the calibrator, even though additional profile shrinkage occurs as the profile cools to room temperature. Calibrators can be made of aluminum, steel, or stainless steel. Cast aluminum calibrators may have steel inserts machined to the desired profile shape to extend the wear life of the calibrator. Aluminum is inexpensive and easy to machine, but the wear surface has to be hardened to extend its life. If a calibrator is going to be used in high-volume production, stainless steel is the material of choice. The calibrator surfaces can be highly polished or chrome or nickel plated to improve the profile surface. One of the most important factors is the calibrator length. It has to be sufficiently long to provide enough cooling to hold the dimensions while running at high throughput rates. Once the profile exits the calibrator, it is cooled further with water spray. For cooling, the calibrator body has water channels to circulate water through the calibrator and a heat exchanger to remove the heat from the circulating water. The calibrator may also contain vacuum connections where a vacuum source can be connected to the calibrator to expand hollow profiles.

The calibrator can be mounted on a calibration table where the water and vacuum lines are connected to the calibrator. After exiting the calibrator, the profile may pass through an additional cooling tank. Alternatively, the calibrator may be placed on a table that is in the first calibration tank section. The hoses and vacuum lines are connected to the calibrator with the lines going to a manifold that can be quickly connected and disconnected through individual couplings. Manifold systems with individual flow control valves and vacuum gauges with quick disconnect couplings allow calibrator tooling to be changed quickly. Most calibration tables have their own self-contained vacuum system and water recirculation system with heat exchanger to provide both uniform vacuum and constant water temperature and flow. Figure 49.10 shows the water and vacuum manifolds on the calibration table prior to the spray unit that is enclosed in a tank. The equipment in

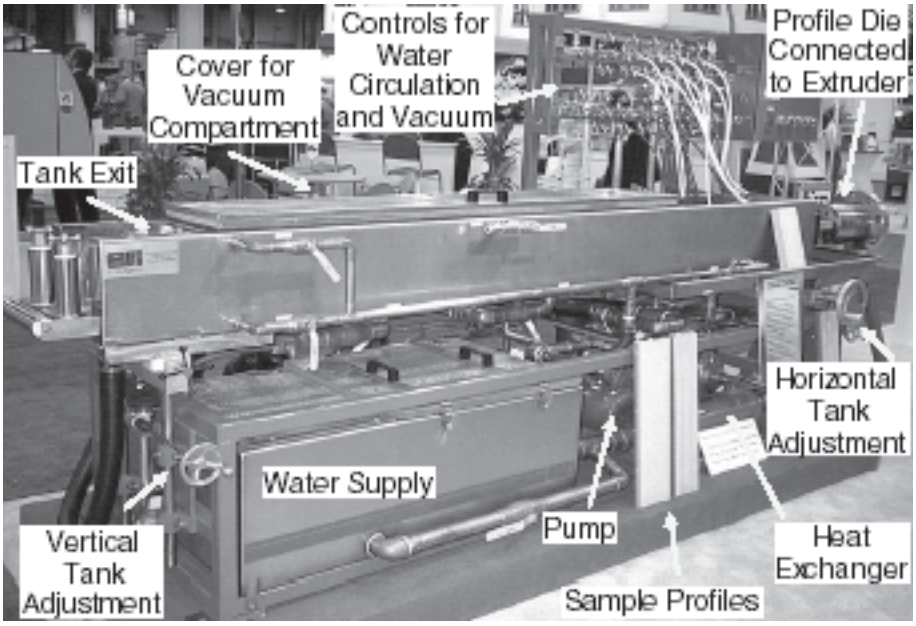

Figure 49.10. ESI vacuum sizing tank.[4]

Fig. 49.10 is supplied by Extrusion Services Incorporated (ESI).[4] The first part of the unit is where the dry calibrator is located, with water and vacuum lines going from the manifold on the calibrator table to the tooling. In the second part of the tank, a water spray or additional calibrators are used to fix the part geometry as the profile is being cooled to room temperature.

The calibration table can be moved up or down, or from side to side, to align the calibrator with the extrusion die and the downstream equipment. A good calibration table is required with high-tolerance parts to guarantee the profile meets the final dimension specifications. The process should not be speed limited by the calibration unit length or its ability to hold dimensions. While the calibration unit is expensive in comparison to the extruder, die, and raw materials, it is a relatively inexpensive part of the production line. Profile rate limitation by either cooling capacity or the ability to hold dimensions established by the calibrator is a problem that is normally easily corrected. At high rates, the calibrator in complex profiles may be five feet long, or multiple calibrators with different cooling along the way may be required to produce a suitable profile at high throughput rates in good yield. Calibrators may operate at slightly lower throughput rates compared to either sizing units or air cooling because of the higher friction between the calibrator surface and the moving profile. This is more than compensated for with improved dimensional precision.

Conair[5] has introduced supercold nitrogen gas to cool parts more rapidly in less space than water. Since cooling is quicker, line speeds can be increased and calibrators and cooling tanks shortened. Liquid nitrogen is converted to a chilled gas and it is circulated around the profile at rates up to 2000 cfm. The liquid nitrogen itself never comes in contact with the hot extruded profile, preventing thermal shock from altering the material performance. Liquid nitrogen is converted to gas when it enters the cooling chamber, expanding rapidly to replace any room air. The chilled nitrogen gas is circulated with a high-velocity blower around the extrudate, producing turbulent air flow, aiding the heat transfer from the extruded profile. Temperatures in the cooling zones can be controlled between -300 and 75°F (-184–24°C). Multiple cooling zones can be used during calibration and cooling to optimize the process efficiency. Nitrogen cooling has been successfully used to cool window, fence, and tubing profiles in less space at higher throughput rates.

Since all thermoplastics shrink on cooling, the die has to be properly cut and the calibrator properly sized to allow for shrinkage as the product cools to room temperature. Additional shrinkage occurs with cooling and post-crystallization if the profile is not completely cooled down to room temperature before it is cut. Shrinkage is balanced with the draw down to ensure the correct dimensions in the final part.

49.3 Puller

The puller or haul-off unit pulls the product from the die through the calibrator or sizing tank at a constant rate with the correct draw down ratio to produce the proper dimensions. Caterpillar-type pullers come with a wide selection of belt materials and styles for different profiles. Pressure applied by the puller has to pull the profile at uniform speed regardless of the tension created by the calibration process, without crushing or distorting the profile in any way. Flat belts are used with flexible profiles, rigid profiles where sophisticated calibrators are not required, and tubing. Various belt materials have different properties:

- Natural gum rubber used for smooth extrudates where a high grip surface is required.
- Cellular neoprene provides a good gripping surface where high abrasion resistance is required.
- Silicone rubber that has good high temperature resistance.
- EPDM is good for high temperature resistance.
- Latex rubber has good frictional characteristics.
- Solid neoprene is hard with good resiliency and is non-marking.
- Cellular polyurethane has very high abrasion resistance.

Other belt materials are available; belts need to be fit to the application. Belts are available with varying hardness, abrasion resistance, and coefficient of friction or gripping abilities. Pullers come in widths and lengths ranging from 4.5–12 inches (114–305 mm) wide and 2.5–12 feet (760–3050 mm) long. The belt size required depends on the profile dimensions, with larger, wider profiles requiring wider and longer caterpillar belts. Longer puller lengths allow the profile to be gripped with less pressure, minimizing profile distortion while still providing enough pulling power to pull the profile uniformly through the calibrator and sizing fixtures. Figure 49.11 shows a puller with a flat belt produced by Conair.[6] Custom profiler's prefer to use belt pullers due to their flexibility and ease of operation.

The motor and drive control are related to cost, the speed range required, and the ability of the drive to hold constant speed. Belt pullers are available with different transmissions and broad speed ranges. If the profile operation produces many profiles, wide speed ranges with

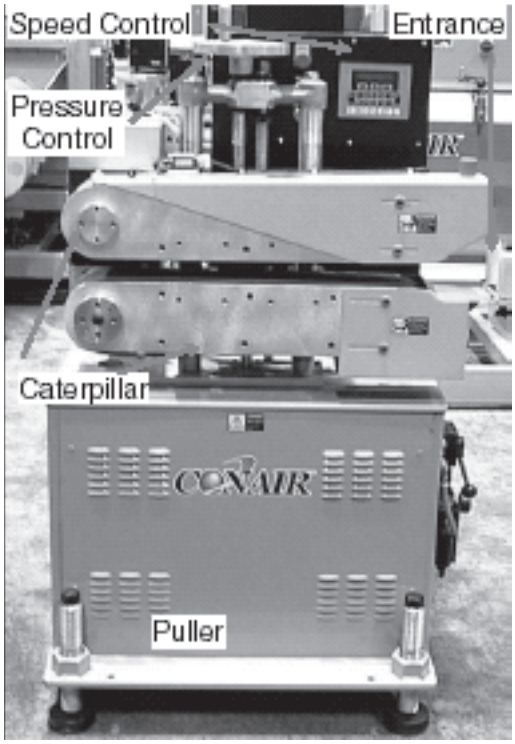

Figure 49.11. Conair puller.[6]

different pulling tensions are required to accommodate the product portfolio. However, if a captive profiler is running the same profile every day on a specific line, it is more important to hold the speed constant over a smaller range with a specific tension. Following are some of the control systems available:

- DC SCR drive, which is relatively low cost with a rather small speed range and adequate but not outstanding ability to hold a constant speed at load.
- DC brushless drive costs more and has a wider speed range with the ability to slave the speed to other equipment in the line such as the extruder screw speed or cutter speed. It does a good job at holding constant speed under load and is more expensive than the DC SCR drive.
- AC inverter drive is a low cost system that can provide higher torque at low motor speed. The speed range is rather small and may vary slightly under load.
- AC vector drives provide a wide speed range with good speed control under load and high torque at low motor speed at the expense of higher cost and more design sophistication.
- AC servo control provides the widest speed range, with good speed control under load. This system is the most complex and most expensive.

Drive motors are available in either AC or DC, using either one or two motors. The controls are normally digital with start and stop buttons. Gear boxes may be connected through either V-belts or chain and sprocket to the puller. Remote speed controls for the puller make the operator's job easier. If the speed control is located near the sizing unit or calibrator table, the operator can easily adjust the puller speed as he or she is stringing up the operation. After the equipment is properly running through the calibration unit, the ability to increase both the extruder speed and the corresponding puller speed from one location facilitates increasing the line speed and getting to the correct operating speed more quickly. Pressure is applied to the profile either manually or through pneumatics by raising or lowering the top caterpillar.

The most significant issue with any puller is slippage, whether it occurs in the drive system or within the caterpillar belts. Slippage anywhere within the puller leads to the wrong profile dimensions. Some haul-off producers recommend specialty belts with their equipment, such as cleated belts (Metaplast[7]) or slit or laterally slotted, profiled belts, and tangent contact belts (Boston Matthews[3]).

49.4 Cutters

In-line cutters are either flying knife cutters or traveling saws. On-demand cutters make a single pass through the profile, based on a predefined signal. The blade sits at rest until the signal is given, at which time it moves very rapidly through the profile, providing a clean cut. An on-demand or flying knife cutter cuts at high speed and is used with flexible, semirigid, or thin-wall profiles at speeds not attainable by traveling saws. The cutting units can be packaged with a puller, providing an integral unit that both pulls and cuts the profile. Rates up to 750 cuts per minute are possible with the latest vacuum clutch brake technology. After cutting, the profile is ejected. Cutter heads with one to four blades are available. The cutter blade slices through the profile at high speed, providing an accurate clean cut. Different suppliers have several models to choose from, based on speed, cuts per minute, and control. Options include a preheating station to preheat the plastic, making it easier to cut while also providing a cleaner cut in brittle polymers.

In-line saws, also referred to as automatic or cutoff saws, clamp down on the profile and move with it as they cut through the profile. After the cut is complete, the saw returns to its initial position, where it again clamps the profile and moves with it through the cutting operations. There are chop saws, up-cut saws, and cross-cut saws. As the words describe, the up-cut saw moves from the profile bottom to the top and can cut up to 10 inches,

depending on the saw blade size. Chop saws come down from the top, cutting through the profile from top to bottom, and crosscut saws cut the profile from side to side. The cuts per minute depend on the saw blade size used and the distance the blade has to travel to cut the profile. Smaller blades have the capability of cutting more parts per minute. On-demand cutters have the capability of cutting 500-plus parts per minute. Saws generally cut in the range of 20 to 40 parts per minute. Saws are obviously used in larger profiles running at slower speeds. Carbide tipped blades are used to prevent the plastic from melting due to friction generated during the cutting operation. Several signals are used to indicate the time to clamp down on the profile and activate the sawing operation; these include a timer, length control based on an electronic signal, or a limit switch. All saws come with safety lockouts that stop the saw blade immediately in the event the E-stop is activated or the saw guard is opened.

49.5 Thermoplastic Materials

Almost any type of thermoplastic polymer can be used in profile extrusion. Obvious choices are most grades of PE, rigid and flexible PVC, HIPS, PS, and ABS. These resins offer a wide range of properties and are relatively low cost in comparison to the engineering resin grades. Rigid PVC has good color stability, is low cost, has good toughness and durability, and is used in many outdoor applications in building and construction. Flexible PVC is clear, tough, and flexible and is used in tubing, gaskets, and seals. Polyethylene is low cost, flexible, and can be coiled and cut to length in filled applications. The attributes of each polymeric grade determine which material to use in different profile applications. Engineering thermoplastics are more expensive and are used in high-performance applications. Using the proper extrusion processing conditions with the correct resin viscosity allows virtually all resins to be used in profile extrusion.

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5. The Conair Group and Material Enhancement, Inc. (MEI), “How Nitrogen Gas Cooling Can Increase Production and Profits,” One Conair Drive, Pittsburgh PA. and 1400 W. Niccum Ave., Effingham, IL.
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Review Questions

1. What are the different dies used in profile extrusion?
2. Why is it critical to have uniform polymer flow out of all sections or legs in a profile?
3. What causes sink in a profile and how can it be eliminated or minimized?
4. What is the effect of nonuniform melt temperature or variations in cooling temperature on the final profile?
5. What controls profile dimensions and final profile size?
6. Name five important profile design factors and explain why each one is important.
7. What is the importance of die land in a profile die?
8. What are three profile cooling designs and how do they differ in profile tolerances, complexity, effect on run size and expense?
9. What is meant by draw down, and how does it affect the final profile?

Review Questions (continued)

10. What is the difference between a calibration table and a vacuum sizing tank?
11. Where is vacuum sizing used and how is it accomplished?
12. Why is it important to be able to move the vacuum tank or calibration table up and down, left and right, and forward and backward?
13. How can supercold nitrogen gas be used to cool extruded profiles?
14. What types of polymeric materials are used for the puller belts?
15. What are two different types of profile cutters? How do they differ? In what profile applications are they used?
16. What are the different types of drives available to use with pullers?
17. What are some common polymeric materials used in profile extrusion?

50 Pipe and Tubing Extrusion

Pipe and tubing is profile extrusion with dies designed to produce round cross sections. A pipe and tubing extrusion line is very similar to a profile line with a vacuum sizing cooling unit, puller, and saw or on-demand cutter. Products can be rigid or flexible and vary from something very small, such as a catheter tube used in medical applications, to large-diameter pipe used to transport water or other fluid. Pipe or tubing can be wound up as a continuous product or cut to length. Pipe and tubing can be coextruded for added value and to meet specific end-use requirements. Figure 50.1 shows some monolayer and multiple layer pipe and tubing products. The clear flexible tubing is a three-layer, coextruded structure containing PE/tie layer/PVC; it is used in medical applications. The corrugated pipes are two-layer coextruded products, while the larger pipe is monolayer ABS. Pipes and tubing are used in many consumer and industrial applications, including water and sewage pipes, drinking straws, fluid and vacuum lines in automotive products, medical tubing, plastic pipets, circular plastic posts, refrigeration hose, vacuum cleaner hose, cable protection, drip irrigation, and so forth. Large volumes of pipe are made from PE, rigid PVC, ABS, and nylon. Tubing is produced from most thermoplastic resins. Flexible tubing is made from elastomers, cross-linked PE, flexible PVC, and polyurethane, while rigid tubing is made from commodity resins to reinforced engineering resins.

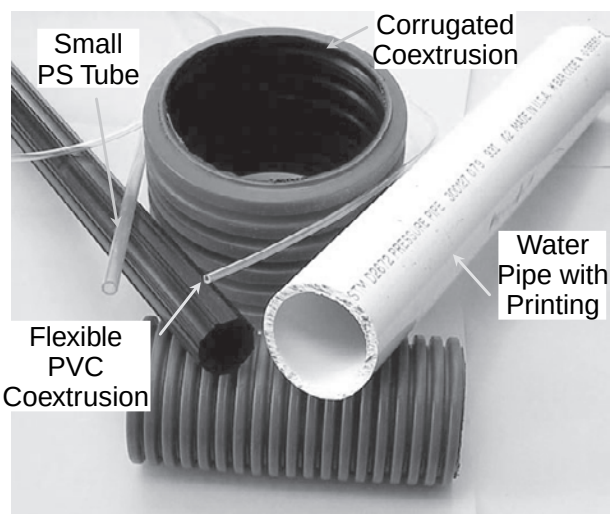

Figure 50.1. Examples of pipe and tubing.

The key components in a pipe and tubing line are similar to a profile line. The cooling section may use a vacuum sizing tank or mold blocks attached to a caterpillar belt for corrugated pipe. Figure 50.2 shows the components in a pipe and tubing line. The extruder used depends on the polymer. For small tubing, 0.5–3.5 inch single screw extruders are quite common. As the pipe diameter requirements become greater, the extruder diameter increases substantially, to where a 48 inch diameter pipe may be produced on an 8 or 10 inch diameter single screw extruder. Counterrotating twin screw extruders and conical extruders are common in processing rigid PVC pipe and tubing. Conical twin screws have positive pumping action that provides outstanding mixing and high head pressure for in-line compounding PVC resin powder with the appropriate additives and plasticizers. A laser gauge monitors the pipe outside diameter (OD) and ovality, and a printer is used to identify the pipe and its specifications.

The dimensions of plastic pipe and tubing are characterized by Eqs. (50.1) and (50.2) for the wall diameter ratio and the wall cross sectional area.

$$DR = \frac{OD}{ID} \tag{50.1}$$

$$A_w = \frac{\pi \times (OD^2 - ID^2)}{4} \tag{50.2}$$

where

DR	=	Diameter ratio
A_w	=	Wall cross sectional area
OD	=	Outside diameter
ID	=	Inside diameter

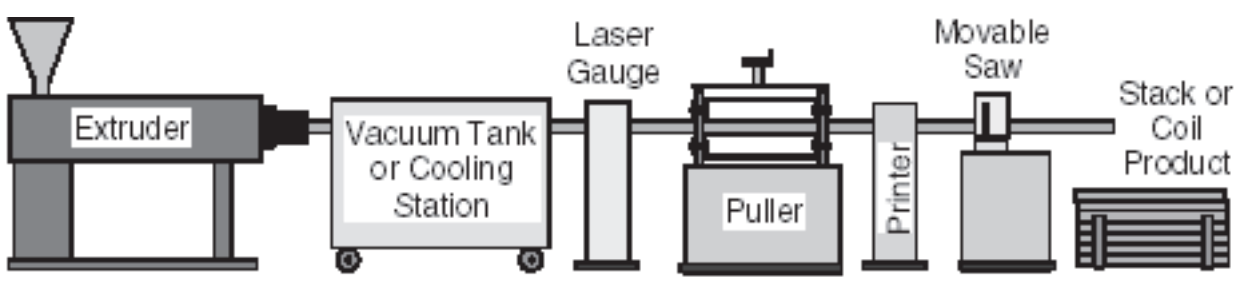

Figure 50.2. Equipment schematic of pipe and tubing extrusion line.

Parameters defined by Eqs. (50.1) and (50.2) are controlled by varying the puller speed or the extruder output. Either modification changes the draw down ratio. The only way to vary parameters defined by Eqs. (50.1) and (50.2) independently is to change the die cross sectional area. To change the parameters defined by the equations, a series of dies or at least die parts are required to change the cross section.

We will now cover the equipment from the extruder die through product packaging used in pipe and tubing extrusion.

50.1 Die

The extrusion head or die is determined by the diameter and wall thickness required in the final product. A single-layer pipe or tubing die is shown in Fig. 50.3. (Coextrusion dies used for coextruded pipe or tubing applications are covered in Part 6.) The extrudate enters directly from the extruder through the breaker plate into the die. Large dies require a die stand or cart to support the die weight and prevent distorting the extruder barrel. An adapter may be present between the extruder and the pipe or tubing die, versus the direct connection shown in Fig. 50.3. The entrance cone distributes the melt uniformly around the mandrel. The mandrel or center section is held in place by spokes radiating out from the mandrel, called a spider ring with the individual spokes referred to as spider legs. The spider legs support the mandrel weight and the shearing forces from the high upstream pressure. The number of spider legs is determined by the die size and the individual spider leg size. The load on each spider leg is calculated similar to the extruder thrust bearing load, where the head pressure is multiplied by the cross sectional area, divided by the number of spider legs. Equation (50.3) can be used to calculate the force on each spider leg:

$$L_{sl} = \frac{P_d \times \pi \times R^2}{N} \tag{50.3}$$

where

L_{sl} = Spider leg load
P_d = Die pressure
R = Die radius
N = Number of spider legs

Calculate the force on each spider leg in a tubing die, assuming a 4-inch diameter mandrel is supported with three spider legs and the die pressure is 3000 psi. Equation (50.4) shows the calculation:

$$L_{sl} = \frac{3000 \times 3.142 \times 2^2}{3} = 12{,}566\,\text{lbs} \tag{50.4}$$

As the number of spider legs is increased, their size can be decreased. Each spider leg divides the polymer flow, causing a knit or weld line that is a potential weakness in the final pipe or tube. Spider legs are shaped to minimize the melt stream disturbance. Sufficient die length has to be present after the spider ring for the polymer chains to reentangle and provide the proper strength at the weld line. Molecular chain reentanglement is encouraged by

- Higher melt temperature (more molecular motion)
- Higher die pressure (forces more molecules to pack into a given area)
- Longer die and land length (provides more time)

A larger melt channel area after the spider ring followed by a flow channel restriction (increased mandrel diameter or decreased bushing inside diameter) builds pressure to encourage molecular reentanglement.

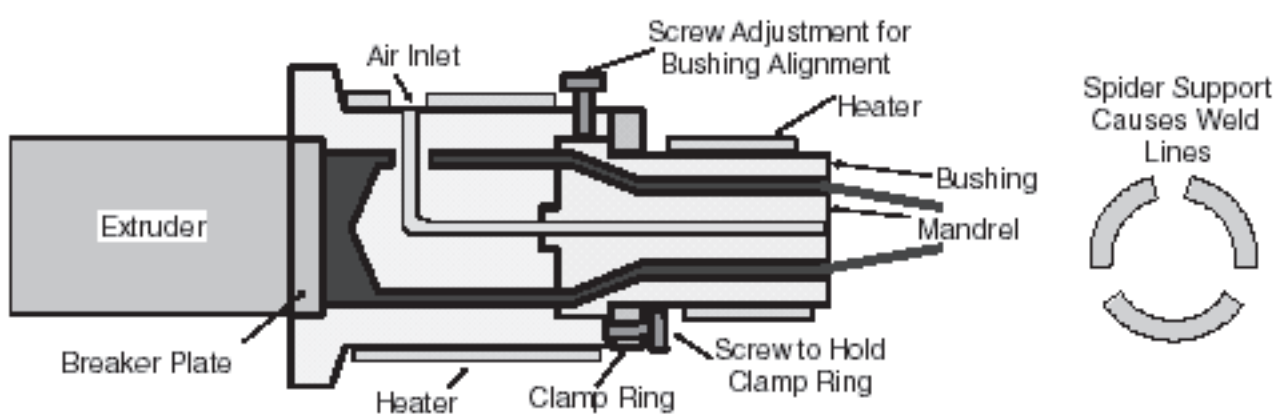

Figure 50.3. In-line pipe or tubing die.

The mandrel is centered in the die and the bushing is centered around the mandrel. To make adjustments, the bushing clamping ring is loosened and the die adjustment bolts are used. If the bushing is improperly centered, the pipe or tube concentricity will be incorrect, with one side having a thicker wall than the other side. Through one spider leg there is an air inlet tube that supplies air to the center of the pipe. Air is critical in obtaining the correct pipe or tube diameter in the calibration tank. The air supply is normally at atmospheric pressure. The die area where the melt channel is a constant diameter is called the die land area and is used to shape the molten polymer into a pipe or tube before it exits the die. Die land lengths need to be proper for the polymer being processed. Longer land lengths may generate too much shear, while shorter land lengths may not provide sufficient molecular orientation and lead to extrudate swell.

Die development over the years has focused on minimizing weld lines caused by the spider ring. Two approaches have been successful in eliminating weld lines. First, spiral dies similar to those used in blown film are used in large-diameter pipe applications. Second, a melt pool is generated directly after the spider ring by expanding the flow channel followed by a restriction. Expanding and contracting the flow channel one or two times forces the polymer chains to reentangle. Regardless of the die used, the polymer melt channel has to be properly streamlined to prevent stagnation, which can generate degraded or cross-linked polymer. Degraded polymer in the melt channel can cause a weld line with no mechanical mechanism to encourage the polymer molecules to recombine. This causes a weak area in the pipe.

Larger extruders can be used to produce multiple pipes by splitting the melt stream as it exits the extruder and feeding two dies. Instead of using end-fed dies, the dies are side-fed. As the polymer melt converges on the back side of the mandrel, it has to mix to allow the polymer molecules to properly entangle. Melt impinging on the mandrel cannot deform or move the mandrel in the die. Multiple dies are used when small-diameter pipe or tubing is produced on large extruders with excessive capacity for one small-diameter product.

To save money on dies for different size pipes or tubes, the end that is connected either to the extruder or adapter can be reused with different diameter mandrels and bushings. This allows the parameters defined in Eqs. (50.1) and (50.2) to be changed by modifying the tips, mandrel, and/or bushing.

Rotating the mandrel tip, as shown in Fig. 50.4, allows modifying Eqs. (50.1) and (50.2) independently, without changing the tip, mandrel, or bushing. By varying the rotation speed with the puller speed or extruder screw speed, the ID and OD of the tube can be changed independently, using the same tubing die without changing any of its components.[1] Advantages of this die are

- A single die can be used to manufacture different OD and ID.
- Tubing is weld line free.
- Tubing with different OD and ID can be produced by changing the rotation speed during tube production.
- Tubing is more uniform.

50.2 Cooling and Sizing

Pipe or tubing exiting the die travels a short distance in air, allowing a thin skin to form before it enters the vacuum sizing tank, where more rapid cooling occurs in a sizing fixture. Proper pipe or tube sizing can be done several ways. One way is to use a vacuum sizing tank equipped with sizing rings or sizing tube, shown in Fig. 50.5. Sizing rings are an assembly of thin metal plates with a slightly larger hole bored through the center than the final pipe or tube diameter. This accounts for shrink-

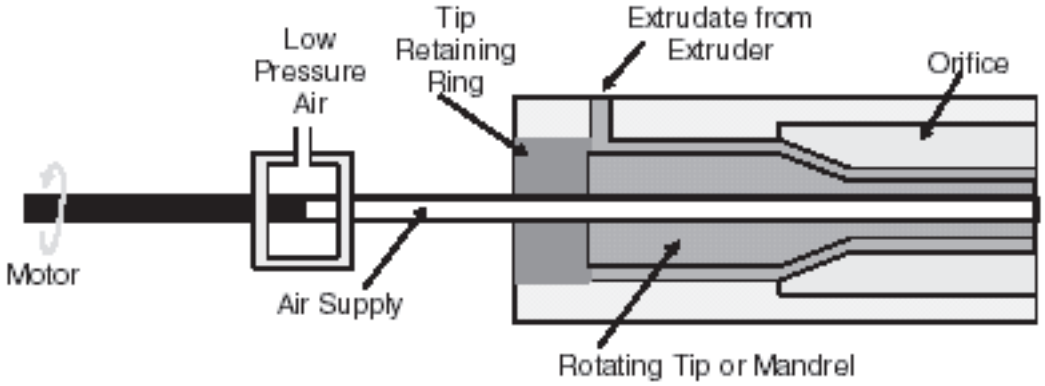

Figure 50.4. Rotating tip die.[1]

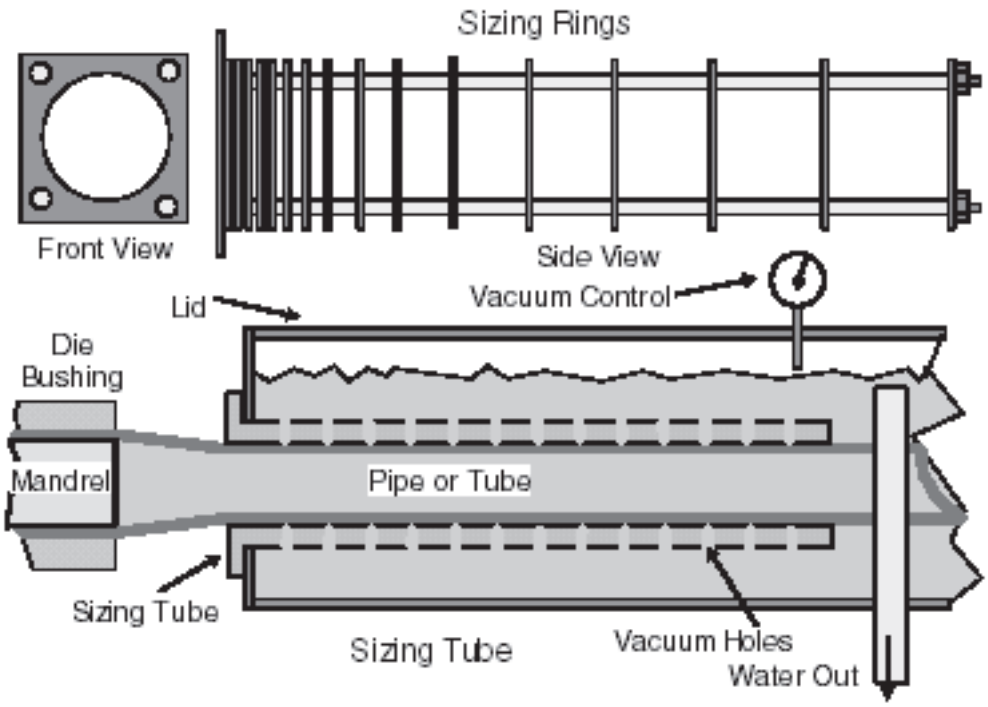

Figure 50.5. Sizing rings and sizing tube.

age that occurs due to cooling to room temperature and crystallization in semicrystalline polymers. The submerged plates in water in a vacuum tank cool the product as it passes through the sizing holes. The vacuum above the water reduces the pressure, allowing the pipe or tube to expand to the sizing rings. Vacuum sizing is preferred in most small-diameter pipes and tubes for convenience and versatility. An alternative way is to use a vacuum sizing tube where a tube with holes for the cooling water to contact the pipe is used in place of the sizing rings. The tube is either submerged in the vacuum tank or water is sprayed around the tube, keeping it cool, with small holes through the tube to allow the vacuum to expand the pipe or tube. Normally the pipe or tube diameter exiting the die is larger than the vacuum tube to create a good vacuum seal. If the pipe or tube is slightly smaller than the sizing tube or the shape is somewhat irregular, the vacuum seal will be broken. Vacuum tanks may have three or four compartments where the water temperature can be controlled independently. Controlling the water temperature helps control the cooling rate, which controls crystal development in semicrystalline polymers. Rapid quenching with chilled water generates more and smaller crystals than slow cooling in hot water. In amorphous polymers, rapid cooling can lead to higher internal stresses due to shrinkage differences. In addition to altering the water temperature in the various compartments, the vacuum levels can be adjusted to improve the final dimensions. Both the water temperature and the vacuum level are directly related to the polymer morphology and consequently the physical properties.

How much the pipe diameter is drawn down is referred to as the diameter draw down ratio. High-diameter draw down ratios lead to high molecular orientation, which may reduce the product hoop strength. A draw down ratio guideline is 1.5–2:1 for small tubes. In other words, the diameter of the die is approximately 1.5 times the sizing tube diameter, with the product being drawn down in the space between the die and the sizing rings or tube. Since the polymer in the gap between the die and the sizing tube must support its own weight, large-diameter pipes or tubes have very short distances between the die and the sizing tube. Small-diameter tubes may be as much as 12 inches away from the die. Figure 50.6 shows a coextrusion setup with a stripper die and an appropriate distance between the die and the vacuum sizing tank. The second extruder providing the stripe is behind the die, with the melt entering the backside. The distance between the die and the vacuum tank can be increased or decreased as the vacuum tank is moved toward or away from the die. While diameter draw down occurs in this area, pipe or tube sag can be a major issue and is addressed with internal pipe cooling.[3]

Draw down ratio also refers to the wall thickness draw down. The wall thickness draw down ratio is defined by Eq. (50.5):

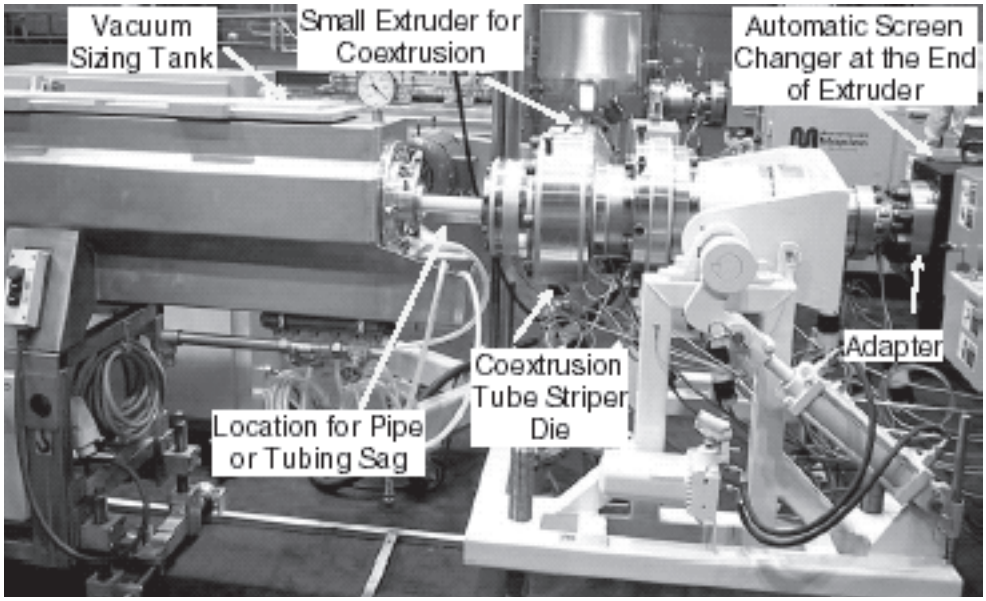

Figure 50.6. American Maplan coextrusion striper die and vacuum sizing tank for tubing.[2]

$$WDD = \sqrt{\frac{ID_B^{\,2} - OD_M^{\,2}}{OD_T^{\,2} - ID_T^{\,2}}} \qquad (50.5)$$

where

WDD	=	Wall draw down ratio
ID_B	=	Bushing inside diameter
OD_M	=	Mandrel outside diameter
OD_T	=	Tube outside diameter
ID_T	=	Tube inside diameter

As the pipe is pulled through the sizing tube or rings, the inside surface is dragged along by the hard outside surface, causing the wall thickness to become progressively thinner as it solidifies. The diameter draw down ratio occurs between the die and the vacuum sizing tank, while the wall thickness draw down ratio occurs in the sizing rings or sizing tube. The final pipe diameter and wall thickness exiting the sizing tube or rings has to account for shrinkage as the part cools to room temperature.

Another way to size pipe and tube is to use air pressure to expand the pipe against the sizing tube. Vacuum may still be used in the cooling tank later in the process. This is referred to as external pipe sizing because a sizing tube encases the pipe as it exits the die, preventing the air pressure in the pipe from blowing a hole through the molten pipe. The pipe slides over a plug that is fastened to the die and remains inside the pipe during operation. The plug acts as an airtight seal, preventing the air from escaping along the pipe while expanding the pipe to the sizing tube dimensions. Air inside the pipe is introduced through the die mandrel with the pressure controlled very accurately. Water is circulated through the sizing tube to remove heat from the molten plastic pipe as it solidifies. This process overcomes the sag problem between the extruder and the vacuum tank. The plug has to be far enough away from the die to allow the pipe inside diameter to solidify before the pipe comes in contact with the plug. Figure 50.7 shows the external sizing equipment. If a sizing tube is being used, the dimensions are critical to obtain the pipe specifica-

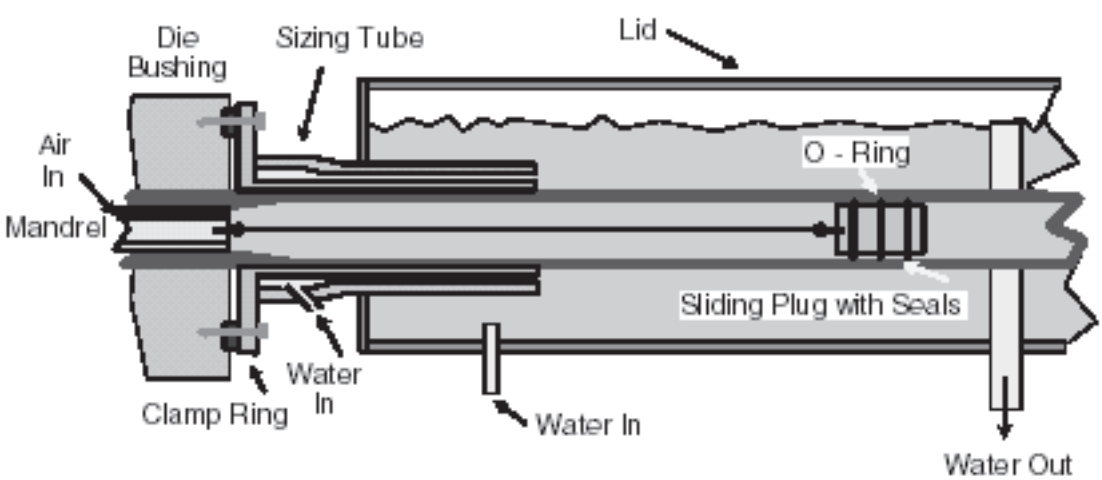

Figure 50.7. External sizing tube with sliding plug.

tions. Initially the sizing tube may be designed with slightly thicker walls and a diameter that assumes approximately 2% shrinkage. If the shrinkage is greater than 2%, the sizing tube can be drilled out, making the pipe diameter larger, allowing for additional shrinkage.

For very large pipe diameters it may not be practical to use a vacuum tank or water cooling tank with an external sizing tube. The external sizing tube has to be water jacketed to remove heat from the process. Water can be sprayed or cascaded over the product once it exits the sizing tube to provide the additional cooling needed to get the product to room temperature. To prevent warpage and bowing, the pipe has to be uniformly sprayed with water to provide a uniform cooling rate. If the pipe bottom is allowed to sit in water or one side is sprayed and not the other, the pipe will bow or warp toward the hot side or away from the cooling due to differential shrinkage.

Aluminum, brass, or beryllium copper is used to fabricate most sizing tubes. The inside diameter has to be free of any nicks, scratches, or burrs that can cause continuous defects on the pipe surface. Harder metals than aluminum are preferred for long runs. The sizing tube inside surface needs to have some micro-roughness to allow air and water to circulate between the pipe and the wall. Holes are drilled through the sizing tube to allow the vacuum to pull the pipe against the tube. In Fig. 50.5 the sizing tube vacuum holes are exaggerated to demonstrate how it works. The holes need to be designed for uniform contact and vacuum around the sizing tube.

The pipe or tube outside is cooled most efficiently with water. Normally there is enough movement in the water to dissipate the heat. To create a reproducible, constant process, it is critical to control the water temperature, which controls the polymer cooling rate and the crystal growth in semicrystalline polymers. In addition to the water temperature, the tank water level needs to be controlled. Depending on the vacuum tank, the end has a flexible gasket through which the pipe passes, with a catch basin underneath to trap any water that may leak. This gasket has to fit tight enough to prevent water from leaking at low friction but with no pipe distortion. Once the pipe exits the sizing tube or sizing ring and enters other chambers in the vacuum tank, it is normally held underwater by rollers on its top side. Due to the air in the pipe and in some cases the material (density less than water), the pipe wants to float.

Regardless of the cooling methods used, the cooling time and corresponding cooling distance has to be long enough to cool the product at the production rate. As with other limiting steps in extrusion, the extruder is more expensive than the sizing tank. Don't let the sizing tank length control the production rate, because the product cannot be cooled sufficiently to package or perform secondary operations.

Internal pipe cooling as it exits the die minimizes sag between the die and cooling tank as the diameter is drawn down. Sag in thick-wall pipes can be one factor limiting the production rate. Krauss-Maffei has developed a system where a cooling mandrel is connected to the die. Air, blown from the die head cools the inside pipe surface.[3] Other equipment manufacturers have developed their own systems to cool the pipe inside surface to minimize sag.

As the pipe or tube exits the sizing tank, it passes through an air wipe to remove water. Drying is necessary before the pipe or tube passes through a laser micrometer that measures diameter, ovality, and wall thickness.

A corrugator replaces the sizing tank when producing corrugated pipe, as shown in Fig. 50.1, or thin-walled drainage pipe. Corrugation provides extra stiffness in the hoop direction in both thin-walled pipe and large-diameter pipe. Corrugators are large mechanical devices designed to corrugate pipe. Two mold blocks containing the negative pipe corrugations are clamped around the pipe as it exits the die and remain in contact with the pipe through the cooling stage. As the mold clamps around the pipe, the pipe conforms to the mold shape, with the pipe flowing into the mold. Several pipe configurations are available. To change the pipe size or configuration, the mold blocks are changed. Mold blocks may be part of a horizontal conveyor, as shown in Fig. 50.8, that moves like a caterpillar. Regardless of the mold block path from the unclamping back to the initial position, the moving pipe is cooled in a continuous operation in contact with a closed mold. As the mold blocks travel back to clamp onto the pipe, they cool so they can absorb heat from the molten polymer. In some operations, the mold blocks are sprayed with a light water mist both when they are in contact with the pipe and when they are being transferred back to their starting position to assist the heat removal. In large pipe, the mold block may be filled with chilled water before it comes in contact with the pipe; after disengaging from the pipe as it returns to the original position, the hot water is removed and the mold block is recharged with cold water. Some mold blocks use vacuum to draw the pipe into the mold block.

Many mold block systems are available. In a closed system, the mold blocks are shuttled through the system in a specific channel on a moving platform, as shown in configuration A of Fig. 50.8. A closed configuration is normally used on smaller pipes run at high rates. Configuration B is an open system, where the mold block halves are connected together with a chain and ride on moving plates. Some open mold block configurations are set horizontally, while others are set vertically to save floor space. In a vertical configuration, the mold blocks look like caterpillar threads on a puller with the mold

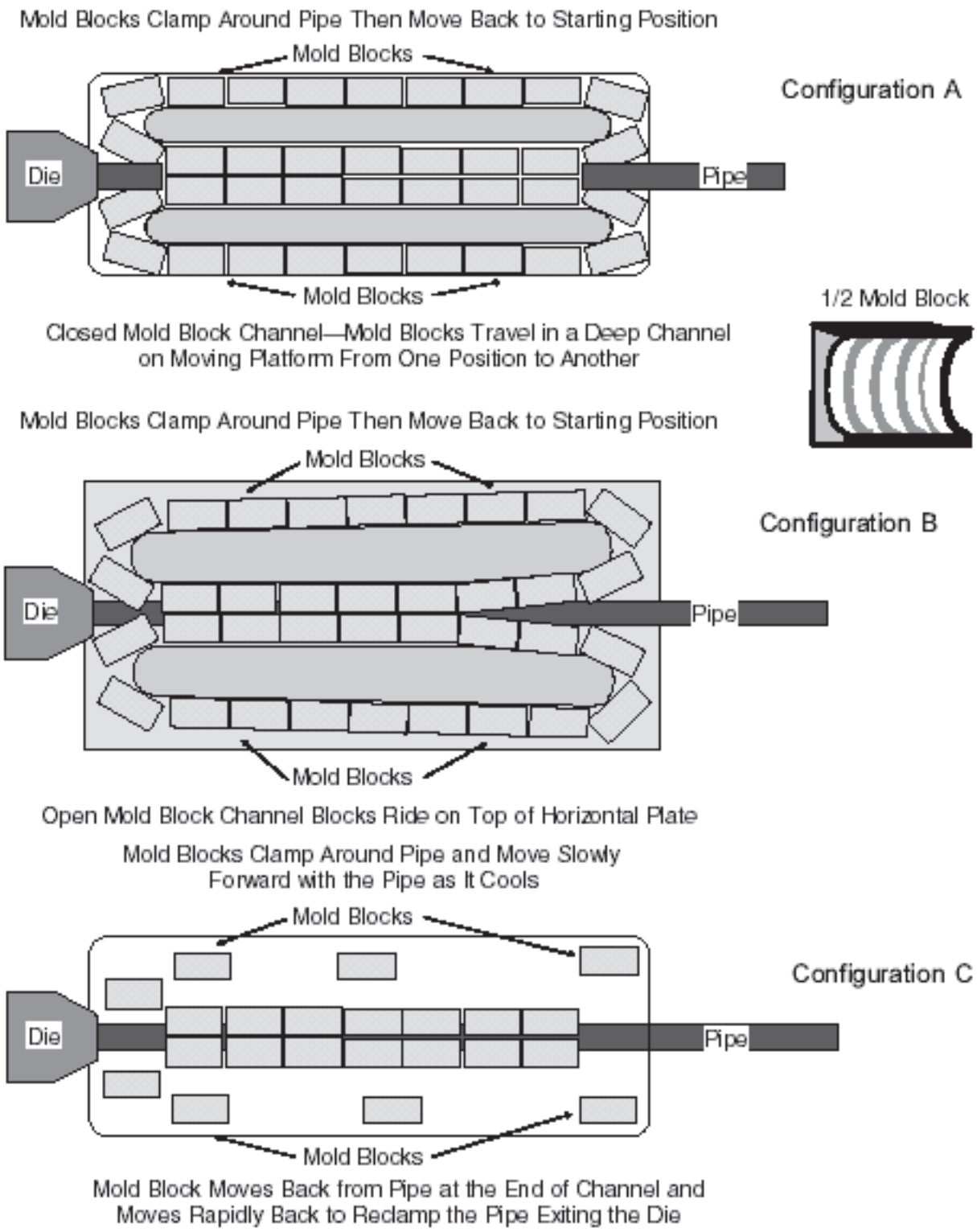

Figure 50.8. Schematic of various equipment configurations to produce corrugated pipe.

blocks attached. Very large mold blocks are set on a shuttle table where the mold blocks move with the pipe at slow speed. They disengage and return to the die at a higher speed, where they reengage the pipe. This system, shown in configuration C of Fig. 50.8, requires fewer mold blocks. Mold blocks can be used to provide different pipe geometries, excellent tolerances, high production speeds, and uniform wall thicknesses. The equipment is more expensive and complicated than vacuum sizing tanks used on smooth-surface pipe; in addition, mold blocks can typically be used on only one pipe configuration.

Advances in corrugated pipe are in size and production speed. The largest dual-wall pipe is 9.8 feet (2.98 meters) in diameter,[4] made by Corma Inc. with a pulsating corrugator. Pipe speeds are constantly increasing with dual-layer corrugated pipe. This is accomplished with bigger molds and improved techniques for both air and water cooling. Corma's "Super Cooling" technology blows chilled air between the mold and the pipe as the pipe is shrinking away from the mold after solidification. Unicor uses two different ways to cool the pipe rapidly: direct cooling on the mold surface by hollowing out the mold block, and a closed-circuit form of direct cooling. Chilled water is circulated below the ground plate holding the corrugator rails. As the mold travels over the rails, the cold water is sucked into the fast moving mold blocks. Increasing the length of the mold blocks is another method of increasing the cooling capacity and increasing the line speed.

50.3 Puller

A caterpillar-type puller or capstan for tubing is used to pull the product from the die through the sizing rings or tube and the cooling tank. Between the cooling tank and the puller may be an on-line laser micrometer to measure the pipe diameter, wall thickness, and ovality, and a length counter to monitor the length produced, which sends a signal to a moving saw to cut the product. The caterpillar puller is very similar to those discussed in Chapter 49 for profile extrusion. Figure 50.9 shows downstream equipment from American Maplan Corporation[2] producing pipe. The pipe is seen exiting the vacuum sizing chamber and entering a combination puller/cutter; in the first station the pipe is pulled by a caterpillar-type puller to a cutting station, seen in the second window. After cutting, the tube is ejected from the cutting station.

The laser micrometer can be used to feed a signal to a microprocessor that can adjust the puller speed or extruder screw to change the pipe or tubing dimensions if they are out of specification. If the diameter is too large, the extruder screw speed has to decrease or the puller speed increase. However, if the pipe diameter is too small, the opposite corrective action is taken; either the extruder screw speed is increased or the puller speed decreased. If the wall thickness is too thin, the puller speed is decreased, while the opposite corrective action is used if the pipe is too thick.

Printing, as shown on the pipe in Fig. 50.1, can be done before or after the puller/cutter operation. If the pipe or tube is polyolefin, it may be necessary to flame or corona treat the pipe surface to make it more polar prior to printing to improve ink wet-out and adhesion. Plasma and corona surface treatment of polyolefin surfaces were covered in Chapter 45. The simplest way to apply continuous printing is an offset wheel containing the type set, which transfers the ink to the flexible wheel, which applies it to the pipe. The type set wheel is inked by a roller in contact with the ink reservoir.

Other secondary operations may include drilling or punching uniformly spaced holes in the sides of drain pipes, attaching a flared end to one end of the pipe so var-

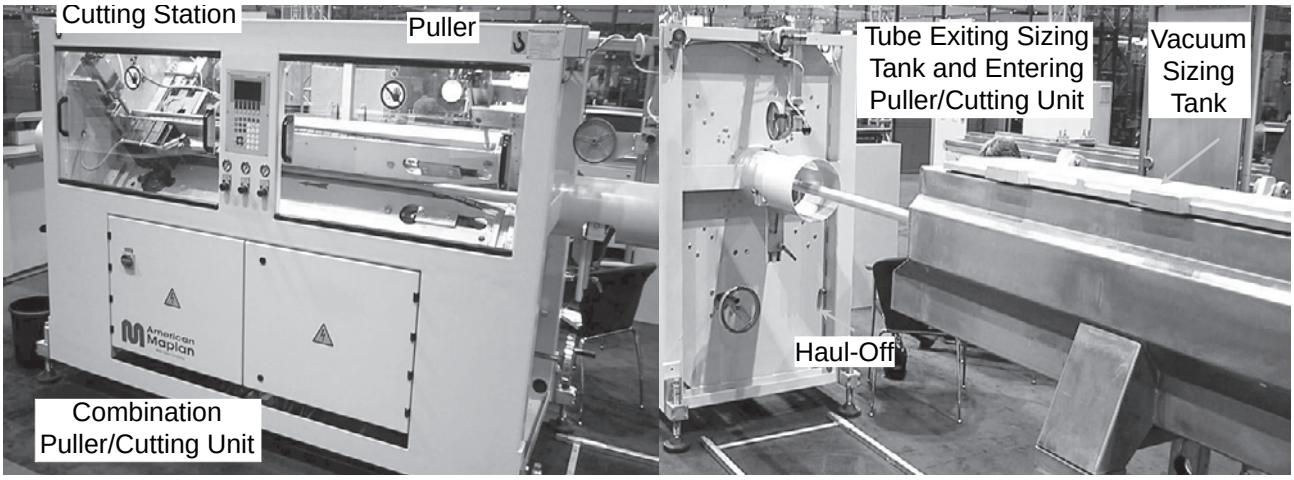

Figure 50.9. Tube exiting vacuum sizing tank entering puller/cutter (equipment by American Maplan Corporation).[2]

ious sections can be easily connected together in the field, slitting one side so it can be wrapped around something of small diameter (for example, a guide wire), slots cut in the surface, and so forth.

50.4 Packaging

At the end of the line is a packaging station to create a uniform package acceptable to the customer. Small-diameter pipe, less than 3 inches (7.6 mm) in diameter, may be coiled rather than cut; this provides the advantage in application of supplying longer lengths with fewer joints. Coilers similar to winders used in sheet and film come with two winding stations. At a fixed length, the reels automatically stop and transfer to the second reel while the first is being removed. Coilers may wind either vertically or horizontally, with sides sufficiently open to band or strap the product prior to removing it from the coiler. Some winders have an index guide that allows the pipe to be laid down next to the previous wrap of pipe in the package. If all the pipe is placed next to each other in an orderly fashion on the reel, more pipe can be packaged in a smaller space. As with any winder, the tension has to be controlled and the winding speed has to remain constant, regardless of the package size. Consequently, as the reel gets larger, the reel rpm slows down to maintain a constant speed at the outside circumference of the package. This speed has to be matched to the puller speed delivering the pipe or tube.

50.5 Scale-Up

Issues involved with scaling up a pipe extrusion line to a larger extruder were presented in a paper by John Colbert.[5] The first part of the paper covers scaling up from one size extruder to another. Similar topics were covered in Part 1, Chapter 7, for single screw extruders and Part 2, Chapter 16, for twin screw extrusion. Issues involved with scaling up the extruder die, sizing die, cooling tube, and pitfalls in scale-up are presented by J. Colbert.

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2. American Maplan Corporation, 823 South ByPass, McPherson, KS.
3. Mapleston, P., “Novel Internal Pipe-Cooling System Boosts Output and Efficiency of Extrusion Lines,” *Modern Plastics*, September 1999, pp. 26–27.
4. Schut, J. H., “Corrugated Pipe—Multi-Wall Takes Off,” *Plast. Tech*., June 1999, p. 60.
5. Colbert, J. A., “The Inter-Related Aspects of Scaling up a Complete Tube Extrusion Line,” 56th ANTEC, Atlanta, 1998, p. 242.

Review Questions

1. What other extrusion process does pipe and tubing most closely resemble and why? What are the common elements between the two processes?
2. How do you calculate the cross sectional area of a pipe? What is the cross sectional area of a 3-inch diameter pipe with a 1/8-inch wall thickness?
3. What is the spider in a tube or pipe die, and what effect doe this part have on the melt and the pipe strength?
4. What is the load on each spider leg in pipe die with a mandrel that is 2.5 inches in diameter, running at 4000 psi, with four legs in the spider?
5. If the die bushing is not properly centered on the die mandrel, what effect does this have on the geometry of the final product?
6. What are some other pipe or tubing dies that do not create weld lines?

Review Questions (continued)

7. What is the difference between sizing rings and a sizing tube? How does the sizing device affect the operation of the pipe or tubing line?
8. What is the purpose of using vacuum in a sizing tank?
9. What are the critical parameters in determining the diameter of a sizing tube and sizing rings? Relative to the pipe, are the sizing rings and tube larger, smaller, or the same diameter as the final pipe specifications and why?
10. What are the types of draw down ratios and where are they controlled in the process?
11. What is the process difference between external sizing and using a sizing tube?
12. Why might internal pipe cooling be important, and what size pipes might use internal cooling?
13. How is a pipe cooled internally?
14. Using sizing rings, where does the diameter draw down occur in the process?
15. What is the purpose of an air wipe at the end of the cooling tank?
16. Why might it be necessary to flame or corona treat a pipe surface?
17. What parameters are determined by using a laser micrometer in-line, and how might this information be used to automatically adjust the line speed?
18. What are three ways to produce corrugated pipe?
19. What are some possible methods of cooling mold blocks in corrugated pipe lines?
20. What haul-off units are used to process pipe and tubing?
21. What secondary operations are performed in-line on pipe and tubing?

51 Monofilament

Monofilament fiber extrusion is used to produce products used in all walks of life. In sports, monofilament is used for fishing line and strings for tennis, badminton, squash, and racketball rackets. Monofilament is used to make synthetic monoyarn used in weaving process and synthetic rope that is used in greenhouses, orchards, fences, construction, fastening, and so forth. Monofilaments are combined to produce

- Decorative tapes
- Ribbons
- Packaging
- Webbing in lawn chairs
- Bristles for toothbrushes
- Grass trimmer lines
- Tire reinforcements
- Sewing thread for home furnishings, industrial textiles, apparel, and floor coverings

Monofilaments are a part of our everyday lives, making both our work easier and our leisure more fun. Typical polymers used to produce monofilaments include

- Polyamide (nylon 6, nylon 6,6, and nylon 6,10)
- Polyester
- Copolyester
- Polypropylene
- Polyethylene
- Polyphenylene sulfide
- Polyvinyilidene fluoride

Monofilament extrusion uses rather long lines and smaller extruders. In other extrusion processes discussed earlier in Part 7, the line length is directly related to the extruder throughput and the extrudate size. Larger extruders producing pipe are used to make large-diameter pipe at high throughput rates that require long cooling areas, pullers, saws, and packaging stations. However in monofilament, the product diameter is very small with a large number of ends. Due to the small product diameter, cooling is accomplished very quickly in a small space. Once the filaments are cooled, they have to be drawn and wound onto packages. The drawing uses relatively large ovens with godet rolls on each side to draw the filaments to the correct denier (grams/9000 meters). After the filaments are drawn, the product is wound onto bobbins, where they may be used directly or combined to produce another product such as rope. Figure 51.1 shows a monofilament line. Most monofilament processes use a 2.5–3.5 inch single screw extruder. Between the extruder and the die there is a gear pump to provide uniform polymer flow and pressure. The circular dies extrude filaments vertically into the cooling bath, where they are separated so that each filament is run separately through the line, where it is wrapped on a bobbin or spool. Exiting the water bath, the filaments are dried and passed through godet rolls that control the speed and draw from the die. Between the two godet roll stands is a large oven that heats filaments to facilitate filament drawing. The second godet roll stand runs at a higher speed than the first set and determines the draw in the oven. The schematic in Fig. 51.1 shows two sets of godet rolls and one draw oven. Depending on the draw required and line sophistication, there may be multiple godet rolls and ovens to incrementally change the draw and to anneal the monofilament before winding it onto a bobbin. Exiting the last godet rolls, the filaments may pass through a laser micrometer to monitor individual filament diameter. Finally, each monofilament is wrapped on an individual package for

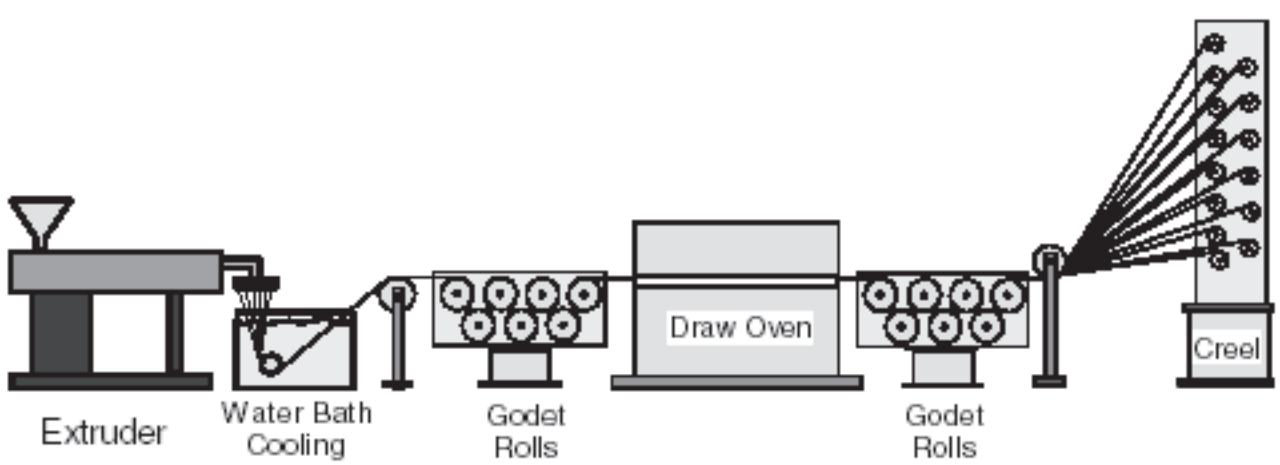

Figure 51.1. Equipment schematic of a monofilament extrusion line.

subsequent operation. The bobbin or spool size used to wrap the product depends on the end-use application. Bobbins are frequently used for textile such as sewing thread, while spools are used for fishing line and grass trimming line. Basic product differentiation is based on polymer type, denier, and tensile properties.

In monofilament production, the fibers are drawn to elongate and orient the polymer chains. Maximum fiber strength is obtained when all the polymer chains are elongated in one direction. A typical tensile stress-strain curve is shown in Fig. 51.2, with the different stages in the drawing process. Figure 51.3 shows how a fiber necks down if it is drawn from both ends. Once all the polymer molecules are aligned, the necking stops and the filament attains its maximum strength. Additional elongation causes the filament to rupture. Applying heat during the drawing process increases polymer chain mobility, and the fibers are easier to draw. A speed difference between consecutive godet rolls draws the fiber, with the oven heat used to facilitate the drawing process. The drawing process is critical to obtain the proper tensile strength and diameter.

The throughput rate in pounds/hour of a monofilament line is given by Eq. (51.1):

$$Q = 25.99 \times \pi \times r^2 \times N \times V \times \rho \tag{51.1}$$

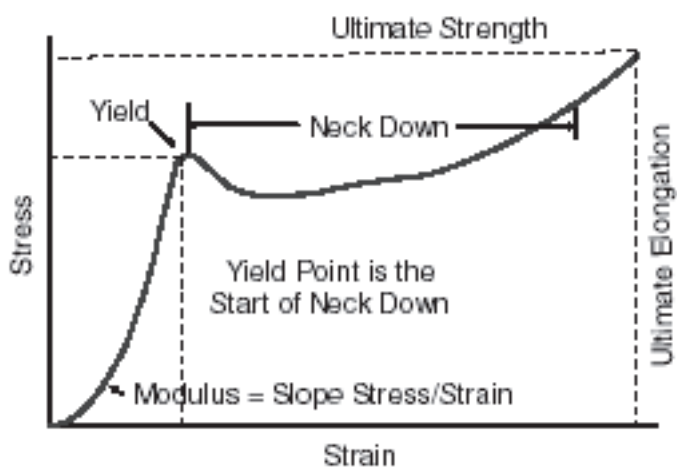

Figure 51.2. Tensile stress-strain curve showing area of neck down and maximum strength.

Figure 51.3. Simulated drawing of a monofilament.

where

Q	=	Throughput rate (lbs/hr)
25.99	=	Conversion factor to obtain pounds/hour
r	=	Monofilament radius (inches)
N	=	Number of monofilaments
V	=	Monofilament speed (feet/minute)
ρ	=	Density (g/cc)

To do the calculation, the monofilament rate and radius have to be taken at the same place. This is important because the monofilament speed changes as it is drawn down.

This chapter explains the equipment and operation of a monofilament extrusion line from the extruder through the creel.

51.1 Die

It is a common practice to install a gear pump between the extruder and die to provide uniform pressure and flow through the die. Monofilaments are small-diameter strands exiting the die, and any pressure or throughput variation changes the filament diameter, which affects both processing and product quality. Gear pumps are slaved to the extruder screw to maintain a minimum pump suction pressure. Any surging or output variation from the extruder is greatly reduced by the gear pump.

Circular dies are used in monofilament extrusion to distribute the melt uniformly to the individual monofilament channels. The die is positioned directly over the cooling tank, with the filaments flowing down into the tank. Figure 51.4 shows the die configuration relative to the extruder and the gear pump, plus a potential die hole pattern. Inside the die is a screen pack for filtering the melt and mixer plates to uniformly distribute the melt in the die. Screen packs use several layers of different mesh screens to filter contamination that might block the die hole or reduce the polymer flow through a particular die hole. The circular die hole pattern gives an identical polymer flow path from the die inlet to all holes. In addition, all holes are machined to exactly the same size. Normally the die holes are in a single circle around the die; however, in some situations two and three circles are present. Die land lengths tend to be two to five times the filament diameter. Uniform heating around the entire die, transfer pipes, and adapters is critical to guarantee uniform melt temperature and viscosity, resulting in consistent polymer flow out of each die hole. If fans or other cooling devices are blowing on one side of the die, this can affect the melt temperature distribution and affect the polymer flow. Fiberglass or other insulation can be wrapped around all

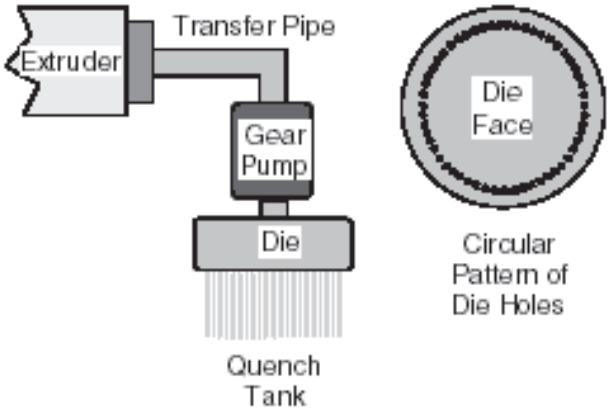

Figure 51.4. Die configuration in monofilament extrusion.

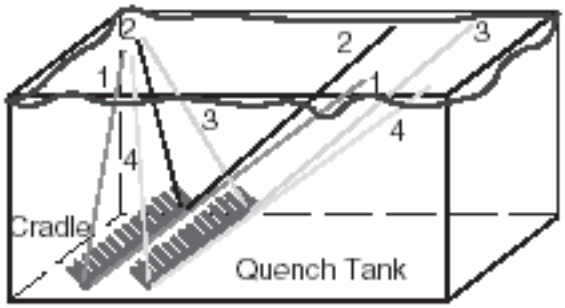

Figure 51.5. Simulated quench tank.

transfer pipes and the die to keep the melt temperature uniform. The die or adapter between the gear pump and die will contain both melt temperature and pressure transducers.

A pressure controller is programmed to control the gear pump suction pressure by adjusting the extruder rpm. The gear pump rpm sets the extrusion rate.

One large extruder can feed two or more monofilament dies. If one gear pump is used, then the pressure drops to each die have to be equal. Otherwise, each die requires its own gear pump. Other equipment modifications required are to ensure that the quench tank, godet rolls, oven, and winders have suitable capacity for the multiple dies. If different diameter products are being produced, separate godet rolls, ovens, and winders are required for each die.

51.2 Quench Tank

The quench tank water level is approximately one inch from the die exit, with essentially no draw between the die and water other than to reduce the die swell. Filaments fall vertically to the bottom of the quench tank, where a cradle separates them. Each filament passes through the line as a discrete unit, never touching its neighbor. This requires each strand to have sufficient melt strength and strand integrity to be processed. The quench tank is normally three to four feet deep with a cradle or deflection cage on the bottom. Each strand has its own unique position in the cradle, with all filaments properly separated so they don't cross or touch one another. The cradle consists of two bars with pins to separate the monofilaments. Half the strands exiting the die farthest from the creel are separated by the pins on the back bar, while the other filaments closest to the creel are separated in the front of the cradle. Figure 51.5 shows the cradle in the quench tank. Only the end filaments are shown passing around the cradle and out of the quench tank. The monofilaments are placed across the cradle in the same order they exit the die, with no monofilament ever crossing the path of another monofilament. The cradle in Fig. 51.5 is a bar with pins to separate the filaments. This bar can be replaced by a grooved roller or a grooved bar to accomplish the same objective. The cradles can also be designed so that the distance traveled by each strand is the same. Normally the quench tank has a light in the bottom of the tank to assist the operator in properly positioning each strand in the cradle.

The quench tank water can be heated or chilled, depending on the process. Hot water provides a slower quench, leading to larger and more crystal development. Regardless of the water temperature used, it is controlled very accurately in the range of 45° to 175°F (5° to 80°C). Hot water can also be used to draw the filaments, although this is not the prevalent method.

Stainless steel and brass are normally used in the tanks to resist rust or other corrosion. The tanks are very rigid to prevent vibration that might affect filament diameters. Other options are to mount the quench tank on wheels for positioning, vertical lifting to raise or lower the tank (this option can be particularly useful in stringing up the line and adjusting the distance from the die), and motor-driven horizontal movement to correctly position the quench tank and cradle under the die.

As the filaments exit the quench tank, they are pulled through a comb guide and across one or more vacuum strippers to remove surface water from the strand. Sponges and/or cloth might be used near the vacuum to help remove water. The filaments must be dry before they enter the drawing unit, as moisture interferes with uniform filament heating. Uniform temperature is necessary to have uniform and consistent filament drawing. At this point, there is a wide band of individual filaments, as shown in Fig. 51.6, traveling through the line. This fila-

ment configuration remains constant through the process until the filaments are wound on bobbins or spools.

Figure 51.6. Monofilament web running through the process.

51.3 Drawing Unit with Ovens

The godet rolls pull the monofilaments, similar to pullers in other extrusion processes. The multiple S wraps around the rolls prevent filament slippage. Figure 51.7 is a block diagram of a typical monofilament extrusion line with multiple ovens and godet rolls. Drawing happens in the ovens between godet rolls. Some lines have two sets of godet rolls with one oven, while other lines have multiple sets of godet rolls and ovens. As the ovens and godet rolls increase, the line flexibility increases, providing capabilities of running more complex products under more stringent conditions with better control.

The first set of godet rolls, or Godet 1 in Fig. 51.7, pulls the product away from the die, maintaining the strand tension through the quench tank and across the vacuum wipe prior to the rolls. Monofilaments are not normally drawn down as they exit the die. They are pulled just enough to compensate for the extrudate swell. Figure 51.8 shows godet rolls with monofilaments. The wrap angle around each roll is approximately 270 degrees. The second set of godet rolls in Fig. 51.7 runs at a higher speed than the first set, with the fibers stretching in Oven 1 as they are heated. The draw ratio in this oven is defined as the product length exiting the oven divided by the product length entering the oven. In the drawing operation, the polymer molecules are aligned in the machine direction to provide maximum tensile strength. Monofilaments exiting the die and passing over the first godet rolls are the largest diameter in the process. Continuous drawing through the process reduces the diameter. Obviously the die is designed to provide the correct final filament diameter based on the draw ratio required to provide the desired filament performance.

Two- or three-stage drawing is possible with multiple sets of godet rolls and ovens, as shown in Figure 51.7. Assume the fiber draw ratio is to be 4:1 with two ovens. A 3X draw ratio or 3:1 could be accomplished in the first oven, with the second oven completing the 4X ratio. Two-stage drawing provides for higher draw ratios with better process control. Godet rolls 3 and 4 in Fig. 51.7 are run at the same speed, with the oven used to anneal the monofilaments. This relaxes the molecular chains and reduces the internal stresses from the drawing operation. Annealing will also reduce the shrinkage that can occur over time. Some postprocessing tests run on monofilaments are high temperature shrinkage and boiling water shrinkage (particularly for polyamide monofilaments). These tests are particularly sensitive to stresses in the monofilament exiting the line and are controlled through the annealing process.

Godet rolls, shown in Fig. 51.8, are also referred to as draw stands. The rolls are cantilevered for easy string-up and operation. A draw stand may have three, five, seven, or nine rolls that are approximately 10 inches (254 mm) in diameter. Rolls can be heated with steam, oil, and water or chilled with water. Rolls are normally steel or rubber coated if required. Heated roll stands can be used in the annealing process to allow the filaments to relax at high temperature.

The ovens use hot air to heat the filaments. Due to the speeds of the lines, the ovens have to be relatively long to provide sufficient time to heat the monofilaments. If three ovens are used to produce a 4X draw ratio, similar

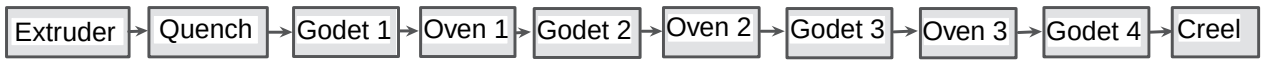

Figure 51.7. Equipment components in a monofilament line.

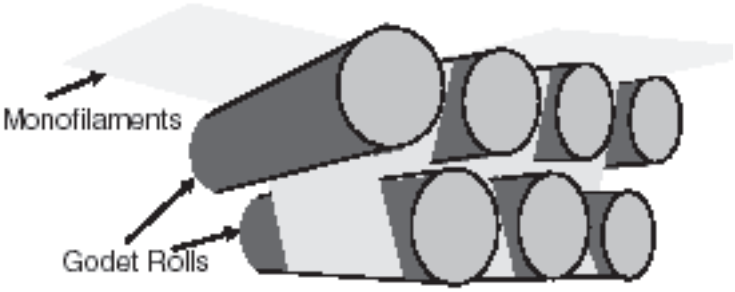

Figure 51.8. Godet rolls with monofilaments.

to the example above, the filament length increases 3:1 in the first oven, 1.33:1 in the second, and 1:1 in the third. Assume the monofilaments exiting the die are running at 100 feet/minute and each oven is 25 feet long (ovens are typically shorter than 25 feet); the time spent in the first oven is approximately six to seven seconds. In the second oven, the strand has about a four second residence time. The residence time in the annealing oven is 3.75 seconds. Considering that most ovens are shorter and the overall draw ratio maybe as high as 10:1 rather than 4:1, the heating time in an oven is relatively short. Oven temperatures are set to attain a specific monofilament temperature to assist the drawing operation or to produce the proper annealing. Oven setpoints are based on the polymer being processed to produce the monofilament and the oven residence time, which is based on the line speed and the draw ratio.

Oven doors are normally pneumatically operated to make it easier to open and close the oven during string-up. Each monofilament has its own path through the oven. Filaments are not allowed to cross or touch each other in the oven. If a filament enters in one location and exits in a different location, the draw ratio increases, resulting in a smaller filament diameter. In addition, filaments crossing or touching each other in the oven can cause abrasion. Ovens have two plenums, one above and one below the monofilaments, with variable speed blowers, fast reacting heaters, and thermocouples to accurately control the temperature. Proper oven operation is as critical as the die, gear pump, and godet rolls.

At the end of each set of godet rolls, there may be a vacuum stand to collect any broken filaments. These vacuum positions can be used to collect filaments during start-up through the individual stages. The vacuum ports act as collecting points until the filaments are properly separated and pulled through the next stage. Initially the strands exiting the quench bath may be pulled to a waste aspirator or vacuum where they are collected, blown to another plant location, or wound on a large bobbin for scrap. As the filaments are strung up through different process sections, the filaments are separated into their individual paths through the process. Additional vacuum collection points are available along the line to collect the filaments until they are ready to be run through the entire line to their bobbin or spool.

In some processes, ovens are replaced with hot water baths containing water below 212°F (100°C), where the filaments are passed through hot water. The strands are pulled out of the bath almost vertically and passed over a vacuum water stripper to remove any water before entering the next draw unit. Grooved rolls entering and exiting the bath are used to change the filament direction and line them up for the next godet rolls.

51.4 On-Line Gauging

On-line diameter monitoring is done with a laser micrometer designed to measure multiple strands. A beta scanner designed by Beta LaserMike[1] has the ability to scan multiple strands, determine the filament diameter, and print an SPC report for the customer. The laser micrometer scans the filaments, similarly to that shown in Chapter 45 for scanning sheet and film thickness. This allows continuous on-line denier measurement and SPC control chart generation. The LaserMike is placed between the last set of godet rolls and the creel.

51.5 Creel

Prior to creeling, a finish or lubricant can be applied to the filaments by contacting or kissing a rubber-coated roll that rotates through fluid in a tray below the roller. The liquid applied to the filaments may be a lubricant, antistat, or some other material to increase the value or utility of the filaments.

The creel is a long rack with individual winding stations and bobbins for each monofilament. Bobbin or spool size depends on the monofilament diameter and the end-use application. Product diameters go from 0.001 inch (0.025 mm) diameter for soft sewing thread, 0.16 inch (4.06 mm) for grass trimmer line, and larger. Heavy gauge monofilaments use larger bobbins than small-diameter filaments produced for sewing thread. Each monofilament is separated and run to individual packages. Figure 51.1 shows a vertical creel. Figure 51.9 shows a horizontal creel. Each winding position has its own transverse guiding system to make a package that is uniform and flat across the wind surface from one spool flange to the other. Filament tension is critical to good wind quality.

51.6 Testing

Monofilament products are tested for tensile strength, water shrinkage, hot air shrinkage, denier, filament smoothness, diameter uniformity, and color. Tests that are unique to monofilament are hot air and water shrinkage, denier, filament smoothness, diameter uniformity, and color. Shrinkage in hot air or water is affected by the process annealing conditions and the residual stresses in the filaments. High stresses can lead to high shrinkage, which may not be acceptable for the application. Denier measures the weight in grams of 9000 meters of filament. Monofilament is classified and sold based on its denier and tensile strength at a

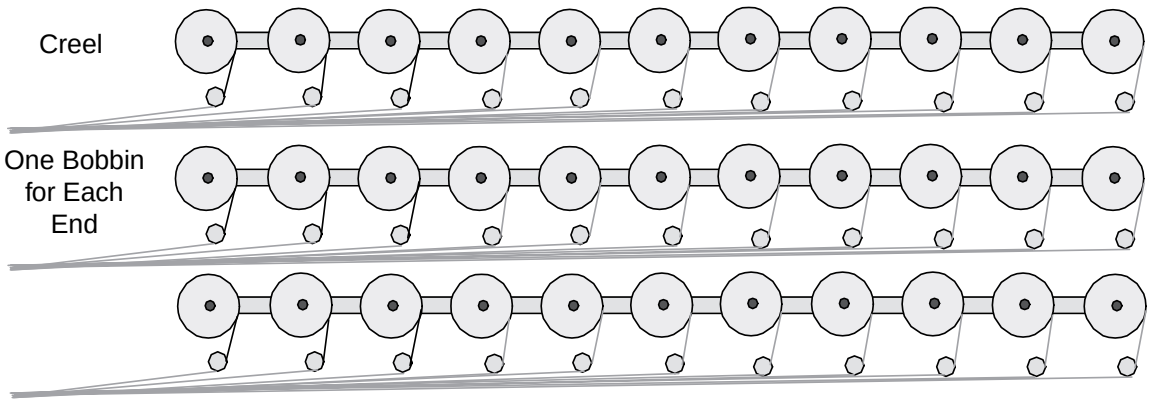

Figure 51.9. Creel to wind individual spools or bobbins.

given denier. Filament smoothness or diameter uniformity is critical in applications such as sewing thread, where diameter variation may hang up in a needle eye and break the thread in a sewing operation. Finally, color is important because monofilaments are sold in many colors for different applications, and the color must be uniform between lots such that combining filaments in a final product such as a rope will have the same color.

SUPPLIER

1. Beta LaserMike Corporation, 8001 Technology Blvd., Dayton, OH.

Review Questions

1. What is the purpose of the godet rolls in a monofilament line?
2. What is the definition of denier?
3. What are draw ratio and neck down?
4. Why do monofilaments have higher tensile strength after drawing?
5. What is unique about the die used to produce monofilaments?
6. Why is a gear pump normally used in a monofilament process?
7. How are the filaments separated in the quench tank?
8. Why is it necessary to thoroughly dry the filaments before entering the draw unit?
9. If there are three ovens in a draw unit, what does the third oven do?
10. How does drawing occur in the draw unit?
11. Why is on-line filament diameter gauging important, and what methods are used for on-line gauging?
12. In a typical tensile stress-strain curve, where does drawing occur?
13. What polymers are used in monofilament processes?
14. What are some typical monofilament applications?

Review Questions (continued)

15. What unique tests are performed on monofilament products?
16. Why is it critical for monofilaments to have their own unique path through the process such that they don't touch or cross the paths of other monofilaments?
17. Why are different solutions applied to monofilaments at the end of the line?

52 Extrusion Blow Molding

Large-part blow molding, also referred to as industrial blow molding or extrusion blow molding, is similar to injection blow molding with the exception that the accumulator, parison, and parts are much larger. Blow molding is used to produce large, irregular, hollow parts in a single operation. The alternative is to injection mold two halves and bond the parts together in a secondary operation. Unlike other extrusion processes, there is no puller to pull the product away from the extruder. Instead the plasticated formulation, delivered by the extruder, enters an accumulator, where the material is delivered periodically to a mold in the shape of a long, hollow, molten tube called a parison. The mold closes around the parison and air is injected into the center of the parison, inflating it to duplicate the mold shape. After the polymer solidifies and cools in the tool, the pressure is released, the mold opens, and the part is ejected and removed from the mold. The process is repeated with a new parison from the accumulator. In comparison to other extrusion processes, extrusion blow molding requires more vertical space and less floor space to produce finished parts. The machine height depends on the final dimensions of the parts being produced. The accumulator and die sit above the tool, allowing the parison to pass between the two mold halves with a blow pin inserted into the bottom to inject air into the parison, expanding it to the mold surface. Figure 52.1 shows an extrusion blow molding process. As the mold closes, the parison is pinched at the top and the bottom to prevent the air from escaping. Both pinch points produce weld lines as the molten polymer is brought together under high pressure and forced to bond. After the part is removed from the mold, both the top and bottom are trimmed to remove the excess beyond the pinch-off points. The scrap is granulated, fed back into the extruder, and reprocessed through the equipment into other blow molded parts. One issue with this process is surface quality. Blow molding is a relatively low pressure process, as air is used to expand the polymer into the tool shape. As the parison exits the die, the surface starts to cool. Even though the time the parison is exposed to air is relatively short, this cooling time plus the low pressure are not conducive to the highly polished surfaces obtained by injection molding. Hotter tool surfaces can lead to improved surface quality at the expense of longer cycle times. Variations of this process are principally in the method of forming the parison and filling the tool.

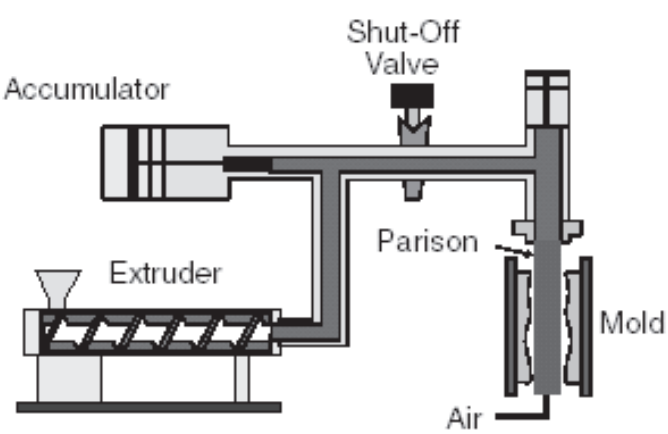

Figure 52.1. Large-part blow molding.

Blow molding is used to produce large hollow parts for

- Automotive uses
- Lawn and garden
- Toys
- Cabinets
- Pallets
- Drums
- Tool cases
- Furniture, containers, and so forth

In automotive applications, parts include

- Bumpers
- Instrument panels
- Knee bolsters
- Distribution system for heating
- Ventilation
- Air conditioning (HVAC)
- Spoilers
- Side view mirrors
- Seating
- Glove box doors
- Door handles
- Fuel tanks
- Fuel filler tubes
- Underhood applications for a variety of tubing and air intake manifolds

Advantages of large-part blow molding include the following:

- Tooling costs are low compared to injection or compression molding as the tools can be made from aluminum and still provide long life.
- The low-pressure process can use composite tools for prototyping or short production runs.

- Very large parts are possible, with accumulator heads being quoted with up to 1000-pound shot size.
- The process can add functionality to molded parts, with carpet or other surface covering attached during the molding operation.
- Thin-wall parts as well as parts with varying wall thicknesses can be made by changing the parison profile.
- Hollow parts are produced on-line without the need for assembly.
- Parts are tougher from the low melt flow resins or high-molecular-weight resins used to provide better melt strength for holding the parison shape.
- Rubber particles added to improve the parison melt strength translate to higher impact strength.

The disadvantages of large part blow molding are

- Sharp corners are difficult to obtain.
- Deflashing is required on all parts, with the flash having to be granulated and reused.
- The high-gloss surface obtainable with injection molding cannot be duplicated.
- Mold parting lines may be present.
- Weld lines may appear at the bottom and the top.
- Parison wall thickness profiling is required to maintain constant wall thickness.

Extrusion blow molding is normally used to make parts that weigh more than 12 ounces (340 grams). This chapter covers the different aspects of large-part blow molding as it relates to extrusion.

52.1 Parison Delivery System

There are two different extrusion blow molding processes: continuous or intermittent parison delivery. In both operations, the extruder runs continuously, but the parison is delivered in a different manner to the cyclic process below the die. In the continuous process, the extruder feeds polymer to the die, where the parison is formed and transferred to rotating or shuttle molds. The second process (shown in Fig. 52.1) uses an accumulator. During the molding process, the accumulator is refilled by the extruder in preparation for the next shot. To increase productivity, two accumulators can be run from one extruder. The extruder is mounted on a platform above the mold, with the parison pushed out of the accumulator and die head directly into the mold. In most operations a single screw extruder is used to supply the melt to the accumulator. Depending on the resin, the extruder may be equipped with a grooved feed throat to increase throughput. In coextrusion blow molding, two or more extruders are mounted on the platform above the mold, providing easy access of the melt to the accumulator and die. Depending on the part size, a large tool is mounted directly below the die in a vertical position between two horizontal platens. During the forming operation, the tool is closed around the parison, with approximately 100 psi pressure air injected into the parison, blowing the polymer to take the mold shape. The extrusion part of a blow molding process is continuous, with the rest of the operation being cyclic.

In the continuous operation, several methods are used to shuttle molds to the parison die. As the parison is exiting the die, it is positioned between the two halves of an open tool. As soon as the parison reaches the proper length, the mold closes, cuts the parison, and, while injecting air into the tool to expand the parison, the next tool is moved into the line with the falling parison, as the process repeats itself.

The parison blow-up ratio is the mold cavity divided by the parison dimensions. Blow-up ratios on the order of 5–6:1 are common. Large parts with sharp corners (large flat panels) may have lower blow-up ratios to prevent the corners from becoming too thin.

Polymer selection for blow molding requires very good melt strength, uniform swell characteristics, and stiffness. The polymer is plasticated and pushed through the die by the accumulator ram into a long, molten tube. The melt strength must support the tube weight as it flows from the die until it is captured by the tool. Two concerns are extrudate swell as the parison exits the die and parison sag from its weight prior to being clamped in the tool. Wall thickness, which is critical to the blow molded part performance, is determined by the parison wall thickness just prior to the mold closing. Sag can result in insufficient wall thickness in the final part. Typically, extrudate swell increases the parison thickness at the die, while sag decreases the parison thickness in the middle or near the bottom. Figure 52.2 shows the die gap, extrudate (die) swell, and parison sag. In principle, sag is overcome through better melt strength and more rapid extrusion through the die.

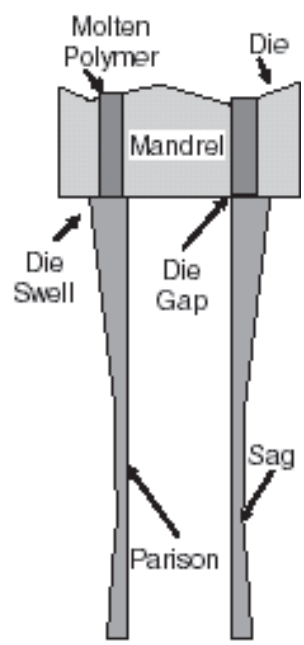

Figure 52.2. Parison swell and sag.

In large-part blow molding, the parison flow

is usually programmed to increase the material pushed out of the die corresponding to thick sections in the part, and to reduce the material in thin sections. This is accomplished by changing the die gap over time as the parison flows from the accumulator. Die adjustments are made by a movable lower lip that can be opened or closed, depending on the profile required. One way to program the parison flow is to cut open the molded part and measure the wall thickness. In thin sections, the die gap is opened as the parison is forced out of the accumulator, and closed in thick sections. In older machines, slides are used to program the parison thickness profile. In newer blow molding equipment, computer touch screens are used to set up the parison profile and save it for later use when the part is run in production again. On-line measurement systems can be used to measure the parison wall thickness without cutting the part, making it easier to determine the correct parison profile more rapidly.[1] Software is commercially available to model the part and the parison. The software predicts the parison thickness, final part thickness, and deformations in the part.

The accumulator and die are required to produce a uniform weight, length, diameter, and wall thickness from shot to shot to produce reproducible parts. The annular dies are either converging or diverging, with converging heads used for parison diameters less than 6 inches (152 mm), and diverging tooling for diameters greater than 6 inches (152 mm). Dies are usually mandrel-type (similar to a pipe die), with a uniform annular gap. A ram above the die in the accumulator, shown in Fig. 52.1, rapidly pushes the molten polymer out of the accumulator to form the parison.

Single or dual accumulator heads are used with large-part intermittent blow molding. The extrudate is pumped into the accumulator through multiple inlets around a spiral mandrel to prevent weld lines in the final parison. Once the accumulator is filled, the parison is pushed out in a repeatable manner, based on a preset profile. The parison push-out speed is not affected by the melt viscosity. This provides uniform parison profiles, regardless of the resin being used. Push-out speed can influence the surface quality of the parison. In the event two accumulators are used, the refill and push-out is computer controlled to coordinate the filling and emptying of each accumulator head in succession.

52.2 Tooling

The mold is the third major component in a blow molding machine. The mold normally is made of two halves that are split vertically down the middle of the part. As the mold closes, the parison is pinched together at the bottom and the top. Then air is injected to expand the part to fill the tool. The 100 psi (6.8 bar) air can be added through a blow pin entering either from the top or bottom. Both halves of the tool have cooling/heating lines that are used to heat the tool prior to production and maintain the tool temperature when it is not in use. During production, the cooling lines remove heat from the plastic while maintaining the mold at constant temperature. The lines must be of sufficient size with turbulent flow to properly remove and control the tool temperature. Water or oil can be used as the heat transfer fluid. Water is a more efficient heat transfer medium than oil and is normally used unless tool temperatures greater than 212°F (100°C) are required. Cooling and heating lines are connected with a manifold to the mold heating unit to provide uniform temperature across the tool face. Serpentine connections allow for the changing heat transfer medium temperature as it is circulating through the tool. Differences in part cooling rate and solidification can lead to warpage, due to differential shrinkage. Depending on the polymer resin being processed, a chiller may be required to remove heat and quench the part very rapidly. If mold temperatures are in the range of 40–70°F (4–21°C), a chiller is connected directly to the tool to remove the heat. If mold temperatures are 70–120°F (21–49°C), a chiller may have to be connected to a mold temperature control unit to remove the heat from the heat exchanger; otherwise, the mold temperature control unit may not be able to control the temperature at the required setpoint.

Molds typically have four guide pins, one at each corner to guide the two mold halves during the mold closing operation. To save money, mold inserts can be used to change the mold shape without changing the complete mold. This uses less steel and can make mold changes faster with small parts.

Similar to injection or compression molding, the mold closes at high speed and then slows down as the two halves approach each other. Finally, higher pressure is applied to lock the tool in the closed position. The slow close operates at low pressure to protect the tool. If a part or something else is caught, the tool will not close due to the low pressure used in that part of the mold closing cycle.

Molds are made out of steel, aluminum, beryllium copper, and kirksite. Injection and compression molding tools are made of tool steels such as at P-20, 4140, and H-13. The low-pressure process used in blow molding allows aluminum and beryllium copper to replace steel for long production runs. Steel is typically used with production runs producing hundreds of thousands to a million parts. Soft aluminum tooling is subject to damage in operation. Advantages of aluminum tooling typically outweigh the disadvantages. Aluminum tools are lighter,

easier to handle, easier to machine, and have better heat transfer properties compared to steel, making them easier to heat and cool.

The pinch-off at the top and bottom of the parison requires higher pressure, and a steel insert is normally installed in an aluminum tool at the pinch-off point. The pinch-off is critical to provide strong weld strength at the top and bottom of the part. Different pinch-off designs are available that influence the weld line strength.

Air injection can be incorporated several ways, depending on the part and the surface appearance. In bottles, containers, and drums, air is introduced through a core pin entering the top of the tool through the die. In these parts, it is necessary to inflate the part from above, because the rest of the part has to be hollow. In a flat panel, the air is normally injected into one of the edges from a blow pin in the bottom of the tool or through a blow needle inserted into the side wall through the tool. The air injection location depends on part geometry. Regardless of where the air is injected to blow up the part, air injection is coordinated with the mold close cycle. The pressure has to properly controlled, along with its duration. The blow pin or needle opening for air injection has to be large enough to handle the air volume required to expand the part rapidly before the resin cools. Clamping forces have to be sufficient to hold the mold closed at the blowing air pressure over the projected tool area. Assume a flat hollow panel is being produced that is 3 feet wide by 6 feet high and 3 inches thick (0.92 m × 1.84 m × 72.2 mm). The projected area of the part is given by Eq. (52.1):

$$3\text{ft} \times 6\text{ft} \times 144 \frac{\text{inch}^2}{\text{ft}^2} = 2592\,\text{inch}^2 \tag{52.1}$$

The clamp tonnage required at 100 psi to mold this part is given by Eq. (52.2):

$$2592\,\text{inch}^2 \times 100 \frac{\text{lbs}}{\text{inch}^2} \times \frac{1\,\text{ton}}{2000\,\text{lbs}} = 130\,\text{tons} \tag{52.2}$$

If this part is being molded with a 100-ton clamp, it is likely the air pressure will blow open the tool and produce a defective part.

Part ejection is accomplished by an ejection system with pins that push the part out of the tool. In place of pins, air poppets can be used on smaller parts. Alternatively, depending on the draw depth and part complexity, parts can be removed with a robot or manually by an operator.

Part surface quality depends on the mold temperature, the parison surface temperature, and the tool surface quality. High-quality parts cannot be produced on molds with bad surfaces. Higher mold temperatures yield improved gloss and surface characteristics such as smoothness. The disadvantage of high mold temperatures is the cycle time required to cool the part. Using multiple molds, optimum part surface can be obtained by cycling the tool temperature. Molds heated to the parison temperature and then quickly cooled once the pressure is applied will produce excellent surface finish (assuming the mold is highly polished). Unfortunately, the energy required to heat and cool the tool in a timely manner is substantial; the cycle tends to be longer and requires more tooling.

In a continuous blow molding operation producing smaller parts, multiple molds are shuttled in and out under the parison delivery system or a manifold die is used to deliver parisons to different locations. Using a fixed table design, molds are placed in a fixed location on a stationary table, with the die having a manifold that delivers parisons sequentially to different locations around the table, as shown in Fig. 52.3. In the top picture, three molds are fed simultaneously, while the other three mold cavities are cooling. In the bottom picture, a parison is being fed to one mold, while two molds in the cycle contain parts that are cooling, and the forth mold is ejecting the part, awaiting parison delivery in the sequential process.

Alternative technologies exist to cycle tools into the parison path with a single die delivering a parison to the same location each time. Dies can be placed on a rotating wheel similar to a Ferris wheel, where each cage is a mold that accepts the parison as it exits the die. The wheel can have a solid vertical base that holds the molds in place. Alternatively, the molds can be on a flat, constantly rotating table that moves the dies underneath the parison. Regardless of which technology is used, after the mold closes around the parison, the parison is cut and the mold moves out of the parison path as the other process steps (blowing, cooling, and ejection) are accomplished. Depending on the mold and extruder size, multiple molds (lined up) can be filled simultaneously by using a die similar to that shown in the top of Fig. 52.3, where three molds are being filled simultaneously with three parisons.

While some possibilities to deliver parisons continuously to multiple molds are explained above, many other possibilities exist, including moving the mold up vertically to capture the parison, dropping the mold back down and moving it out of the way as the next mold is centered under the parison, opened, and moved vertically upward to capture the next parison. The objective is to produce as many parts as possible, based on the extruder throughput. The process limitations have to be the extruder screw speed and producing a quality parison. The limiting step should not be based on the number of molds and the cooling time required before the part can

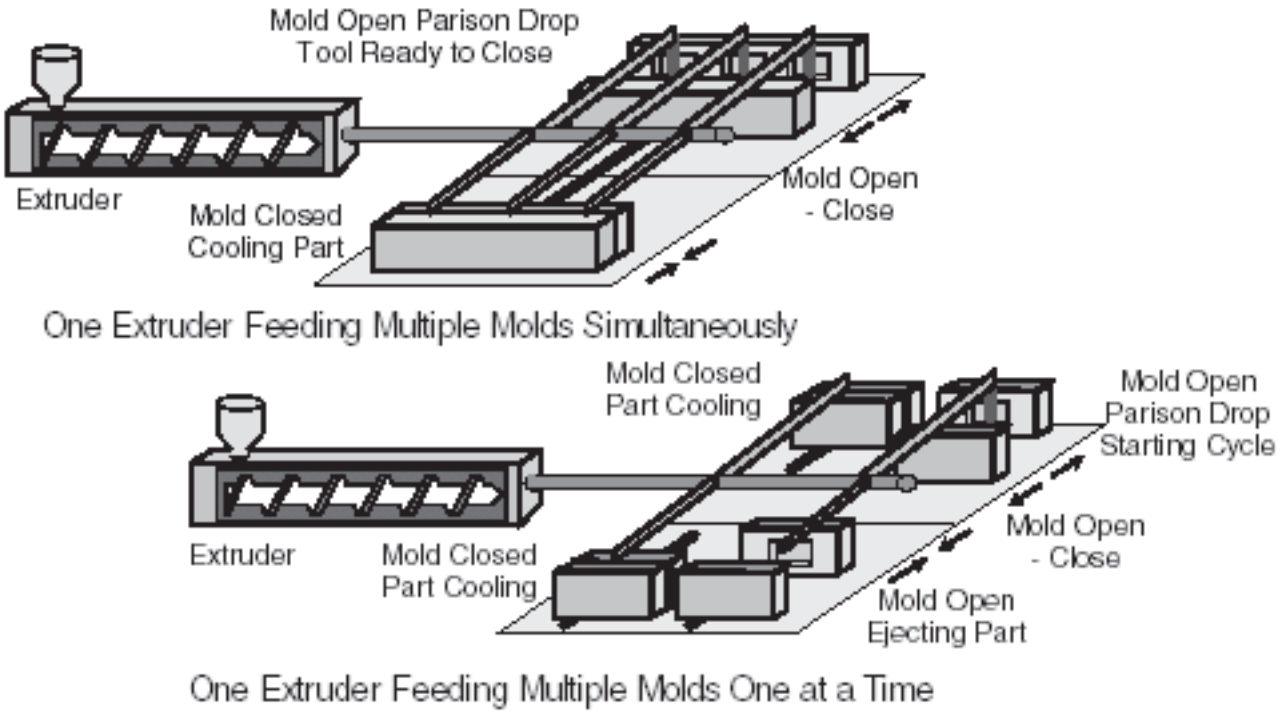

Figure 52.3. Molds on a stationary table with a manifold die.

be ejected. Continuous processes work very well for small molds that are easily moved, require low tonnage or clamping force, and cool rapidly because of part thickness and polymer properties.

In large-part blow molding, the accumulator process, described previously, is required because of the mold size, the mold weight, and clamping forces.

3D blow molding was introduced to

- Reduce flash
- Lower clamping forces
- Improve wall thickness distribution
- Reduce finishing work on the outside surface

3D blow molded parisons are made by moving the tool as the parison is pushed out of the die or accumulator. This lays the parison into the 3D mold. Alternatively, a robot arm can be used to position the parison. This process has gained its name from the three dimensional parison manipulation. Figure 52.4 shows how a part is molded with the mold moving to position the parison properly in the tool. In the first step, the parison is laid across the bottom of the tool; in the subsequent two steps, the bottom half of the tool moves downward as the parison is laid across its surface. After the parison is completely across the bottom mold, the top half closes to clamp the parison just prior to inflation.

To produce a blow molded bent pipe, a multiple-piece mold can be used with a parison diameter smaller than the final pipe diameter. This allows the parison to be laid down in the middle of the tool without pinching off the parison anywhere, creating extra flash that has to be removed in a secondary operation. The pinch lines around the part are removed, providing extra strength. Figure 52.5 shows a simulation of a four-piece tool used to produce a hollow pipe.[2]

Some different systems available to accomplish 3D molding include[3]

- Placo system, similar to that shown in Fig. 52.4, has an inclined mold that moves with the parison to place the parison in the tool in the optimum locations.
- Parison manipulation uses a robot to place the parison in the tool in the desired location.
- Horizontal blow molding places a tool horizontally under the parison, and a robot lays the parison in the proper location.
- Suction blow molding feeds a small diameter parison into a closed tool, where it is sucked through the tool and then sealed off and inflated to match the tool geometry.

Coextrusion, covered in Part 6, is usually used to produce multilayer structures in different shapes, such as

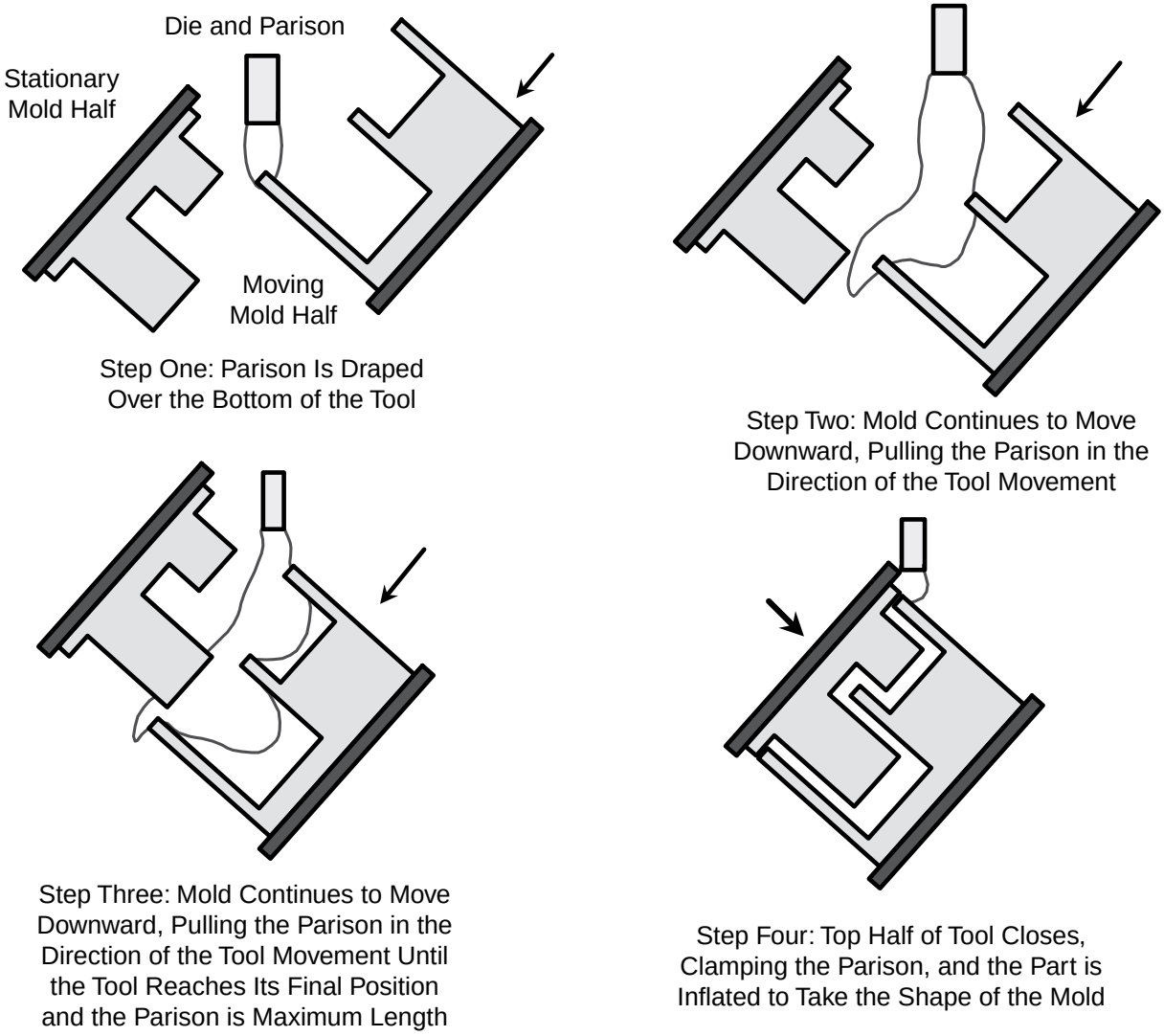

Figure 52.4. Blow molding cycle simulating 3D molding where the mold is moved to produce a complex part.

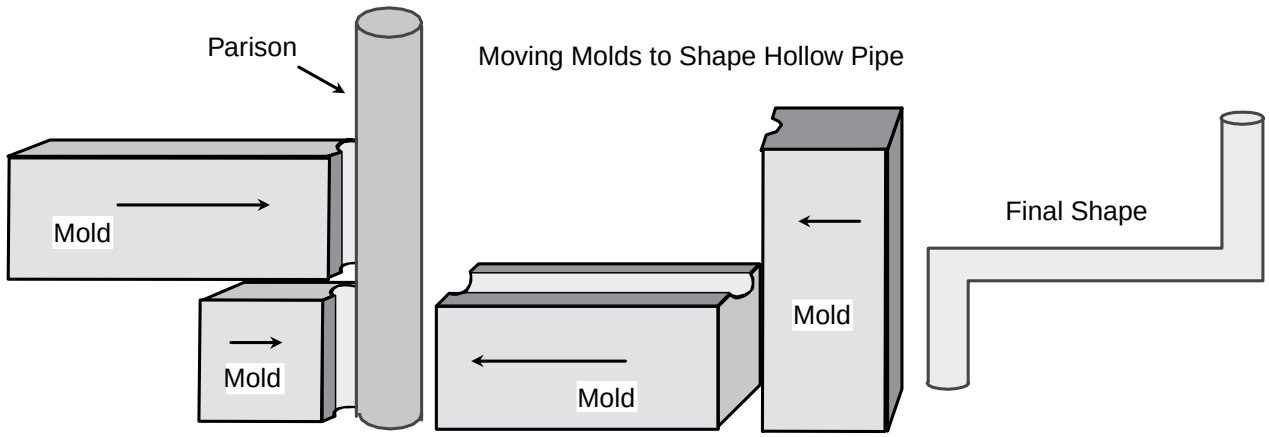

Figure 52.5. Moving tool parts to form final shape in the mold.[2]

sheet, film, pipe, and so forth. One use of coextrusion in blow molding is to make parts with different polymers in different sections of the part. Figure 52.6 shows a blow molded air inlet pipe used in underhood automotive applications. Two extruders are used to supply the different resins. Each accumulator contains the molten polymer supplied by its extruder. The resins are pumped from the accumulator to the head, which is programmed to deliver each polymer to the proper parison location. Figure 52.6 shows the air handling pipe with a hard polypropylene end, followed by a soft rubbery section to provide flexibility, followed by another hard polypropylene section on the other end of the pipe.

52.3 Materials

Most polymers can be used for blow molding, if the melt strength is sufficient to support the parison weight without sag for the time period required to push the parison out of the die and close the clamp. To provide the necessary melt strength, polymers with high molecular weight (low melt flow) or polymers modified with an elastomer are normally used. For containers, the polymer chemical resistance to the liquid the container is going to hold is absolutely critical. Resin selection is based on application requirements such as

- Chemical resistance
- Impact strength
- Temperature resistance
- Tensile and flexural properties
- Clarity
- Durability
- Parison strength
- Processability

To accommodate the application requirements, glass-filled resins are used for stiffness, impact-modified resins are used for parison strength and impact, and coextrusion is used to combine different materials to produce a product that is better than any of the individual components.[4] HDPE is commonly used in blow molding because its high molecular weight (fractional melt flow resin) provides good parison strength, with good chemical resistance, low temperature impact, low cost, and a good balance of properties.

Blow molding foam technology is a way to reduce part weight, increase stiffness, and improve thermal insulative properties. Using foam as a core layer in coextruded structures provides some unique structural possibilities.

52.4 Equipment Selection

Choosing the correct industrial or large-part blow molding equipment depends on balancing the extruder size with the accumulator heads, press tonnage required, hydraulic unit controlling the press, air injection system, part size and weight, and the process controls.[5] Cycle time determines productivity, and the cycle time is controlled by the extruder throughput rate, the accumulator ram speed, and/or the mold cooling required for part ejec-

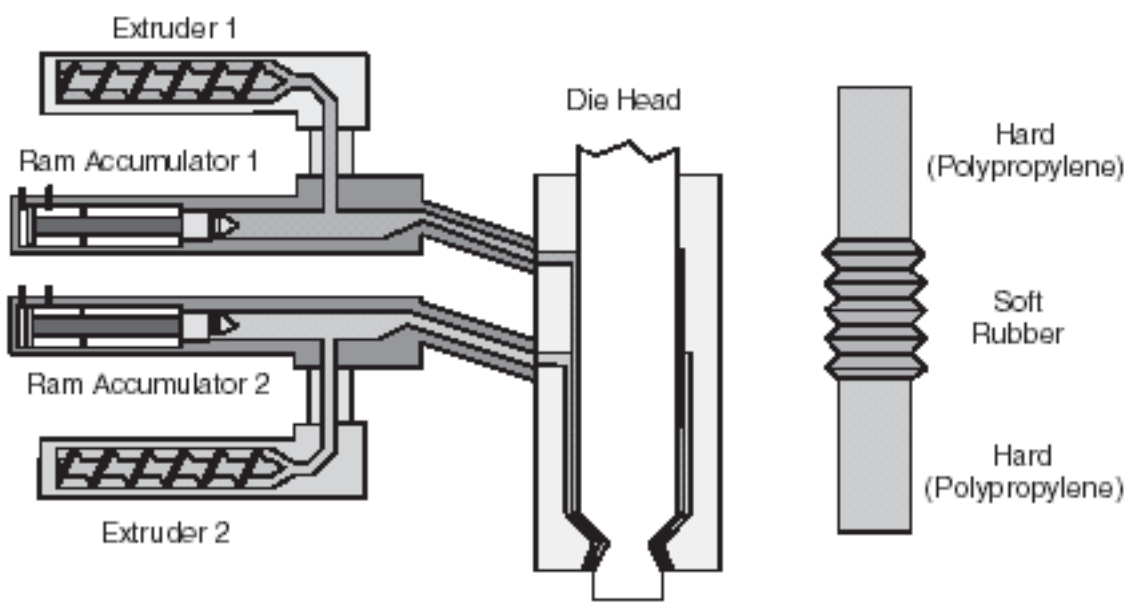

Figure 52.6. Coextruded blow molded pipe with hard and soft sections.

tion. A microprocessor can control the various steps in the process and collect SPC data. The overall control system is complex, especially with 3D molding. Verify that any new machine purchase will meet all processing requirements, both currently and for the next few years.

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Review Questions

1. Describe a large-part blow molding process.
2. What is the difference between conventional large-part blow molding and 3D blow molding?
3. What are the two types of processes used in extrusion blow molding?
4. What critical material parameters are required for blow molding resins and why?
5. Calculate the clamp pressure required to blow mold a barrel with a diameter of 2.5 feet and a height of 4 feet, assuming an inflation pressure of 100 psi.
6. What is the function of an accumulator, and how does it work?
7. What is the purpose of cooling lines in molds, and why are they critical to the process?
8. What are some critical parameters to improve the surface quality of blow molded parts?
9. Why are extrudate swell and sag important in blow molding?
10. How is the wall thickness uniformity controlled in molded parts?
11. What are some advantages of using the blow molding process?
12. How are multiple molds used in a continuous extrusion blow molding process?
13. What are some methods of filling multiple molds sequentially?
14. List some disadvantages of the blow molding process.
15. What materials are used to make blow molding molds, and why are they used?

53 Foam Extrusion

Foam extrusion can be used with most of the processes discussed previously in Part 7, using either single screw or twin screw extrusion, by incorporating a blowing agent into the polymer. Downstream equipment depends on the extrusion process. Foamed products are

- Profiles
- Sheet
- Pipe
- Blow molded items
- Stock shapes

As with other extrusion processes, some polymers are easier to foam than others, based on their melt viscosities and processing temperatures. Foamed structures are produced by incorporating a gas into the polymer melt under pressure in the extruder. As the molten polymer exits the die, the pressure is removed and the gas bubbles expand, creating a polymer structure with many small controlled voids. Two general foam structures are currently produced with extrusion equipment. Standard foam structures with cell sizes >100μm and a cell population density of 10^6 cells/cm^3 are used in thermal insulation, foamed protective coating, tools, pipe insulation, toys, and so forth. Microcellular foam with a cell structure of approximately 10μm and a cell population density of approximately 10^9 cells/cm^3 are used in packaging, aircraft parts, automotive parts, insulation, and so forth. The foam cell number and size determines the density relative to the base resin, the part size expansion, and the product physical properties. One advantage foamed structures offer is their ability to provide higher tensile modulus or stiffness compared to a solid plastic part of the same polymer at the same weight.

Foamed structures are made by incorporating chemical blowing agents, gases, or liquids into the polymer melt. The gas is formed in the polymer melt in a single screw extruder metering section or downstream of a good melt seal in a twin screw extruder, where the pressure from the extrusion process can keep the gas compressed in the melt. Dies are designed to have a high pressure gradient that keeps the gas compressed in the die, preventing premature foaming and handling difficulties at the die exit. As the molten polymer exits the die, the reduced melt pressure outside the die allows the gas to expand, creating a porous foam structure in the extruded shape. Density reduction depends on the blowing agent used, blowing agent concentration, and melt temperature. Chemical blowing agents can be premixed with the resin and fed to the extruder as a blend. Using this procedure, one has to be sure the chemical blowing agent does not decompose prematurely in the melt, releasing gas that can escape back out of the feed throat.

Alternatively, gas or liquid blowing agents are incorporated directly into the melt using either single or twin screw extrusion. The gas or liquid is injected into the vent port, approximately two-thirds of the way between the feed throat and the die in single screw extruders, using a two-stage screw. The gas has to be soluble in the polymer and is mixed into the resin using distributive mixing elements. Nucleating agents are added to the polymer to provide sites for bubble growth. Gas bubble size and solubility determine the foam uniformity in the final product. After gas or liquid injection, the polymer melt temperatures are lowered to increase the melt viscosity. Under typical polymer processing conditions, the melt viscosity may be insufficient to prevent the foam bubbles from rapidly expanding and rupturing, whereupon the cell structure can collapse as the extrudate exits the die. To overcome this problem, melt temperatures are lowered, increasing the polymer viscosity prior to exiting the extruder. In twin screw extrusion, the blowing agent is injected into the melt following a good melt seal, where dispersive mixing elements and a left-handed conveying element work to provide uniform melt mixing. The gas or liquid is injected into the extruder downstream of the melt seal where distributive mixing dissolves and distributes the gas in the polymer matrix.[1]

The standard process used to produce conventional foam products employs two single screw extruders in tandem. In the first extruder, the polymer is plasticated and mixed with a nucleating agent (used to provide uniform cell structure), with the second extruder cooling the product. Pentane or other liquid blowing agent or gas is injected into the first extruder after the polymer is melted or plasticated to provide the medium for bubble formation and cell size. After the blowing agent is dissolved in the melt, the melt is fed through a transfer tube under high pressure into a larger, second extruder. The second extruder reduces the work put into the molten plastic as it prepares the polymer to exit the extruder. The melt is cooled to produce uniform cell growth and stability. As an example, assume tandem extruders are being used to produce a polystyrene foam product. In the first extruder, polystyrene is compounded with approximately 0.5% of fine talc or some other suitable nucleating agent with pentane injection after plastication is complete. The melt temperature in the first extruder is 400–450°F (201–232°C), which is typical for polystyrene processing. The melt is transferred through a melt pipe at high pressure to a second extruder, where the melt temperature is reduced to 275–300°F (135–149°C) before exiting the die in the desired shape. The second extruder is a heat sink to reduce the polymer melt temperature while keeping the gas incorporated in the polymer matrix.[2]

Density reduction depends on the process used. Chemical blowing agents are used to reduce the density up to 50% of the plastic density, while gas and liquid blowing agents added downstream in extruders can reduce the final part density by 80% or more. In HDPE the part density can be reduced from approximately 0.95 to 0.10 g/cm^3 or more with gas or liquid addition downstream of the extruder.

In theory, any thermoplastic resin can be used to produce cellular foam structures. In practice, the most common resins extruded into foam structures are

- Polystyrene
- Polyethylene
- Polyvinyl chloride
- Polypropylene
- ABS

Polystyrene and polyvinyl chloride have the largest processing windows because they are amorphous resins and can be cooled more prior to exiting the extruder. Polyethylene has a smaller processing window, with polypropylene's being even smaller. Foam cores are coextruded with other resins to produce parts with solid skins encapsulating the foam core.

Microcellular foams versus standard foam products provide better

- Impact strength
- Toughness
- Product uniformity

One microcellular foam disadvantage is the reduced part expansion versus conventional foam structures. While conventional part expansion may be 50% greater than the comparable solid part, microcellular foam parts may have only a 1–10% size increase. Microcellular foams have a very uniform cell size, while conventional foams may produce many cell sizes within the same final structure.

53.1 Blowing Agents

Blowing agents come in three forms: solid chemical, liquid, or gas. Solid chemical blowing agents are premixed with the base resin and extruded as a blend under slightly different conditions than are normally used with the base resin. At a specific temperature, the chemical blowing agents decompose, generating gas that forms the cell structure. The correct blowing agent for each polymer is selected based on its decomposition temperature and the gas generated. As an example, some blowing agents generate ammonia gas (NH_3); if this is used as the foaming agent in thermoplastic polyesters such as polycarbonate, polybutylene terephthalate, polyethylene terephthalate, or other copolyesters, the gas reacts to hydrolyze the polymer. Uniroyal Chemical Corporation produces chemical blowing agents under the product labels of Celogen® and Expandex®, with several grades that generate different gases at different decomposition temperatures.[3] Decomposition generates nitrogen (N_2), carbon monoxide (CO), carbon dioxide (CO_2), water (H_2O), and NH_3 in different proportions, depending on the chemical blowing agent used. Decomposition temperature ranges are given in Table 53.1. Reedy International Corporation[4] supplies chemical blowing agents under the name of Safoam® that generate CO_2. Their decomposition temperatures also are given in Table 53.1.

Table 53.1. Blowing Agents

Material	Decomposition Temperature, °F (°C)
Celogen TSH	230–250 (110–120)
Celogen OT	316–320 (158–160)
Celogen AZ	401–419 (205–215)
Celogen RA	442–456 (228–235)
Safoam RIC	248–330 (140–165)
Safoam FP	316–360 (158–183)
Safoam RPC	360–600 (182–316)

Chemical blowing agents are blended with resins between 0.5 and 1.0% by weight. Resin pellets can be precoated with mineral oil or other light paraffin oil to provide a tacky surface for the chemical blowing agent to stick to. Powdered resins can be blended directly with the chemical blowing agent powder. The polymer to be foamed has to be melted or plasticized in the extruder at a temperature below the blowing agent decomposition temperature given in Table 53.1. After plastication is complete, the melt temperature is increased with higher barrel temperatures. This causes decomposition of the chemical blowing agent and generates the gas to form the cell structure. Decomposition is a function of time and temperature. Lower temperatures require longer time for complete decomposition, while higher melt temperatures reduce the decomposition time.

Decomposition temperatures can be reduced by adding an activator that reacts with the chemical blowing agent. The activator is blended in with the resin and the chemical blowing agent during the blending step. An

activator is normally added at approximately the same concentration as the chemical blowing agent, up to 1% loading. Potential activators include

- Zinc stearate
- Zinc oxide
- Dibasic lead phthalate
- Calcium stearate

The optimum activator is based on the chemical blowing agent used. If activators are incorporated into the blend, additional process vigilance is needed to keep the polymer melt temperature below the decomposition temperature until the melt is completely melted.

In a twin screw extruder, the chemical blowing agent can be fed in the feed throat with the resin or downstream after the resin is plasticated. If the chemical blowing agent is fed downstream, the polymer melt temperature where the chemical blowing agent is added has to be lower than the decomposition temperature. The chemical blowing agent has to be distributed uniformly throughout the matrix. A good melt seal in the screw after mixing the chemical blowing agent is used to prevent any gas from escaping back out the side feed port where the chemical blowing agent was added to the extruder. Sufficient additional extruder length is required to provide the correct time and temperature for complete decomposition and generation of the gas. The gas has to be uniformly dispersed in the final melt at high pressure.

Chemical blowing agents are supplied in different particle sizes. Smaller particle size undergoes faster decomposition than larger particle size blowing agents with the same composition. If an activator is used, smaller particle sizes are prone to decompose rapidly.

Typical extrusion temperature profiles have the feed zone low to prevent premature blowing agent decomposition, which would allow gas to escape out the extruder feed throat. The second and third zones, where plastication occurs, is run at lower temperature than used in normal resin extrusion to prevent premature blowing agent decomposition. The temperatures in barrel zones 4 and 5 are set at temperatures above the blowing agent decomposition temperature to complete the reaction. Adapter and die temperatures are reduced to minimize the gas expansion in the extruder. Head pressure and die pressures need to be kept as high as possible to minimize expansion inside the die. Immediately upon exiting the die, the gas expands, foaming the extrudate. The foam formed is controlled by the blowing agent added to the formulation. Figure 53.1 shows the resin expansion due to foaming as the extrudate leaves the die.

Pentane and isopropanol are liquid blowing agents that are injected into the extruder, where they are converted to a gas. Pentane has a boiling point of 97°F (36°C) and isopropanol 180°F (82°C). As the pentane is fed to the extruder, it is rapidly converted from a liquid to a gas. Pentane has to be injected into the extruder downstream, where it is incorporated into the melt. In a single screw extruder, a two-stage screw is used, and the pentane is injected through the vacuum vent in a closed system. The same location is used if two extruders are being used in tandem. In a twin screw extruder, the liquid is injected downstream after a good melt seal on the screw typically formed by a rearward conveying element. At the end of the extruder, good ventilation is required to remove pentane gas that escapes from the foamed structure as it expands. Pentane gas is flammable, and appropriate safety precautions and proper handling procedures are required in transporting the liquid to the extruder, injecting the liquid into the melt, and capturing any gas from the expanded structure. The use of a liquid blowing agent or gas (discussed below) injected into the extruder provides the opportunity to produce a lower density part.

In addition to ventilation at the end of the extruder, foamed structures produced with pentane continue to give off pentane for a few days after the extrusion is completed. Consequently, the product is isolated in a well-ventilated, flameproof area for a few days to allow the gas to escape. In large foam sheet rolls used in thermoforming, the rolls are equilibrated to allow the bubbles to spread evenly throughout the roll or the thermoforming process is not repeatable and the cell structure can vary from part to part.

The third class of foaming agents are gases. For years, fluorinated hydrocarbons were used as blowing agents; however, due to their environmental impact, they have been replaced with other blowing agents that are

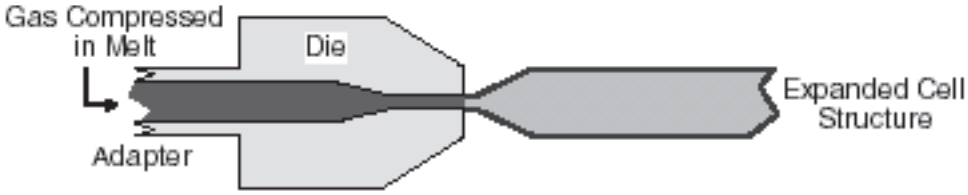

Figure 53.1. Foaming process—cell expansion outside the die.

more environmentally friendly. Chlorofluorocarbons (CFC), N_2, CO_2, Ar, and hydrocarbons (butane) have been used as gas blowing agents. Certain chlorofluorocarbons have been tried that are environmentally friendly. These include 1,1,1,2-tetrafluoroethane (HFC 134a) and pentafluoroethane (HFC125).[5] Selection of the optimum blowing agent depends on the gas solubility in the polymer being processed, the cost, and safety considerations. Regardless of the gas used, it must be capable of producing a consistent cell structure with similar surface finish. An option for balancing cell structure and appearance is to blend gases.

53.2 Downstream Equipment

Depending on the extrusion process, the downstream equipment is similar to that presented previously in earlier chapters of Part 7. Foamed sheet is a unique process, with different methods used to produce thin, flexible sheet used in packaging applications compared to thick, stiff insulative board. This foamed sheet is made with a mandrel-type die that extrudes a large continuous tubular shape that is slit in either one or two locations to form either one or two sheets. The sheets pass through a roll stack and onto the winder. Production of foam board uses a calibration table. A sheet die similar to those presented in Chapter 45 shapes the extrudate. Upon exiting the die, the foamed board passes through calibrators to control the thickness and flatness, as required by the final product. After the calibrator, a puller removes the board and feeds it to a milling and cutting station for trimming to the final product dimensions.

In addition to thin sheet and board lines, foam is extruded into a tube shape, where it is slit on one side for use in pipe insulation and as a protective cover installed on sharp edges or corners. Foam can be extruded as a core layer in coextruded profiles, blow molding, and sheet. Other uses for foam extrusion are in wire coating operations, safety equipment, toys, protective covering on sports workout equipment, and so forth.

53.3 Microcellular Foam

Many articles have appeared in recent years on the development of microcellular foam extrusion.[6][7][8][9] Most microcellular foam extrusion is done with CO_2 gas injected into the system. Key processing parameters include CO_2 concentration, pressure, temperature, gas solubility in the matrix, melt viscosity, and surface tension. High nucleation density is required to create the cell density for a microcellular structure. After plastication in the extruder, the gas is injected into the extruder to form a gas–polymer melt solution, followed by foam cell nucleation, cell growth, and finally stabilization of the cell structure. The critical step in the process is cell nucleation, which is accomplished by creating a thermodynamic instability in the die and reducing the gas solubility in the matrix. A higher pressure drop in the die accompanied with high melt temperature generally generates more cells or a greater cell density.

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Review Questions

1. What are the different types of chemical blowing agents, and how do they work?
2. How does the downstream equipment used in foam processing differ from downstream equipment used in solid part extrusion?
3. Why is it necessary to inject liquid or gas blowing agents downstream in the extruder?
4. What is the difference between conventional foam and microcellular foam?
5. What is the critical element necessary in the screw design used in twin screw extruders to produce foam?
6. How do chemical blowing agents work, and how are they added to the extruder?
7. What products are made using a continuous foam extrusion process?
8. What is the advantage of foam structures?
9. Where does the foaming operation occur?
10. Why is nucleation important?
11. What factors must be considered when designing a die to produce foam?
12. Why is a tandem extrusion process used with single screw extruders to produce foam?
13. What are the dangers associated with pentane as a blowing agent?
14. In using pentane, why must the foamed products be stored in a well-ventilated room for a few days?

54 Solid Stock Shapes

Solid shapes are somewhat different from other extruded shapes in that the extrudate is not pulled away from the die. Typical solid shapes are

- Rods
- Slabs
- Half round rods
- Rectangular strips
- Triangular rods
- Square rods
- Tubes
- Square tubes

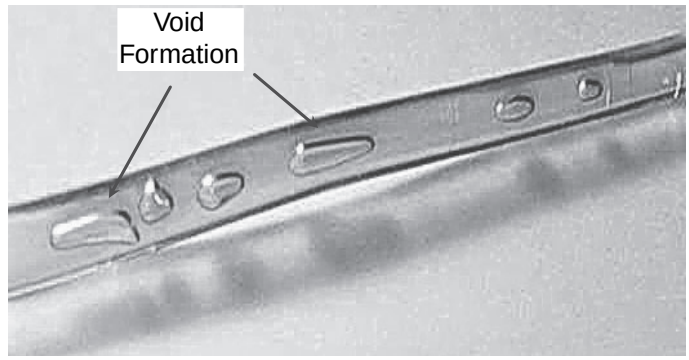

Figure 54.2. Vacuum void formation in extrudate.

These shapes and any other thick, solid profiles are a challenge to produce, due to the potential for void formation in the center of the part. In other extrusion operations, as the polymer exits the die, it is rapidly quenched, forming a solid skin on the surface with molten polymer in the center. Extruded parts cool from the surface toward the center. If quench conditions are much colder than the resin melt temperature, the surface solidifies first, and solidification progresses inward until the entire part is solid. In cooling, the entire part becomes smaller as it shrinks. Consider producing one-inch or larger diameter rod stock; as the skin cools and becomes solid, the center is still molten, as shown in Fig. 54.1. After the skin becomes relatively thick, it stabilizes and, while still shrinking until it gets to room temperature, it does not shrink as much as the core. With the molten core shrinking more than the outside surface, the core pulls away from itself, forming a vacuum void that appears to be a bubble. The greater the cross sectional area of the part, the more difficult it is to prevent void formation in the core. Figure 54.2 shows a relatively small strand with vacuum voids in the center. Void formation weakens that section and can cause premature part failure in use.

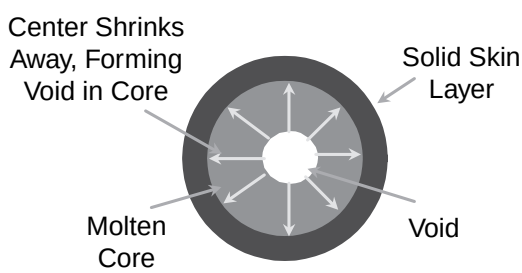

Figure 54.1. Cross section of a solid rod showing void formation in the center due to shrinkage.

To produce solid parts without voids, special dies, cooling, and part removal are required. Figure 54.3 shows an extrusion line to produce solid stock parts. Molten polymer exits the die into a forming box, where the outside skin is formed. As the extruded shape (rod, slab, solid shape, and so forth) passes through the forming box, the skin becomes thicker and thicker. Rather than pulling the shape away from the extruder, causing the extrudate to neck down, the molten polymer is pressured into the profile center, pushing the shape away from the die. This allows the center to be continuously packed with polymer as it cools and shrinks. The process is similar to injection molding, where the part is filled and additional polymer is pumped into the part as the outside cools. In the injection molding cycle, this is accomplished with the hold pressure and time. Polymer is continuously pumped into the center of the injection molded part until the runner system or gate freezes off, preventing additional polymer from being injected into the profile. Similarly in the stock shape extrusion, a long forming box can be used in conjunction with the puller to prevent the part from being pulled away from the extruder too rapidly. As the puller prevents the part from being pulled away from the die, more polymer is packed into the center during the cooling operation, filling the center as it cools and preventing void formation. Another piece of equipment other than a puller to control the haul-off speed so that the extrudate is pushed away from the die, allowing the center to be packed, is called a brake machine.

The greater the stock shape cross sectional area, the longer the required forming box. In Fig. 54.3 the forming box is shown with cooling water circulating around the box to remove heat from the extrudate. The cooling rate determines the potential for void formation. As shown in Fig. 54.3, the wall thickness continues to increase as the part moves away from the extruder, while the center area into which the molten extrudate can be

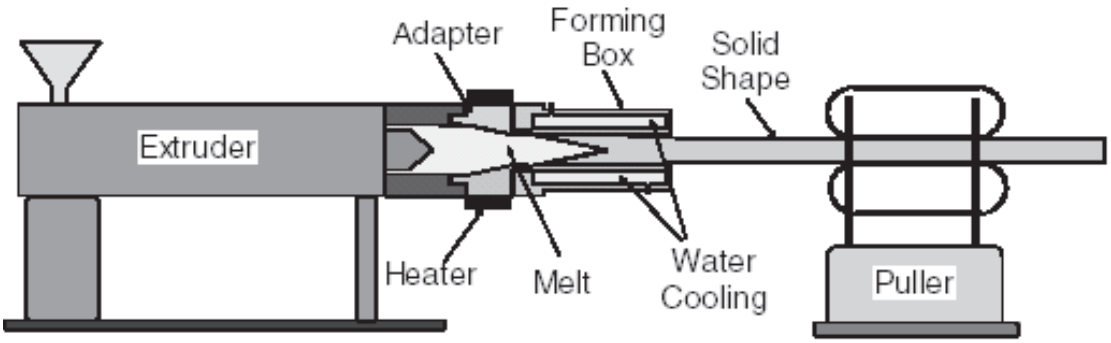

Figure 54.3. Schematic of an extrusion line to produce solid stock parts.

pumped decreases. Cooling water in the forming box, as in cooling of other extruded shapes, is a relative term. Sheet can be cooled on rolls that are 200°F (93°C) for good polish or at 120°F (49°C), depending on the cooling desired and product finish. If slow cooling is required, hot water or hot oil can be used as the heat transfer medium in the forming box. The forming box length is dependent on the stock cross sectional area, the polymer being processed, and the line speed. Higher throughput rates and larger cross sectional areas require longer forming boxes.

After the shape leaves the forming box, it is cooled to room temperature with a cooling tank, spraying water on the surface, or other appropriate cooling method. At or near room temperature, the stock shape is passed through a brake machine or puller to remove the product from the extrusion line. Secondary operations such as printing, drilling, punching, and so forth, are performed after the brake machine. Somewhere between the brake machine and the end of the line, the product is cut to length for packaging. This can be done before or after any secondary operations. The final step in the extrusion line is packaging.

Packing forces required to make void-free parts are on the order of 400 psi. Assume a 2.5 inch diameter rod is being produced; what is the force required inside the forming box to minimize void formation? The cross sectional area of the 2.5 inch diameter rod is given in Eq. (54.1):

$$A = \pi \times r^2 = 3.1416 \times 1.25^2 = 4.91 \text{inch}^2 \quad (54.1)$$

The load in the forming box (assume 400 psi) is given by Eq. (54.2):

$$L_{Forming\ Box} = A \times P = 4.91 \times 400 = 1964 \text{lbs} \quad (54.2)$$

Amorphous polymers are easier to process in solid profiles than crystalline polymers for two reasons:

- Amorphous materials have less shrinkage.
- Amorphous materials have a broad softening range rather than a well-defined melting point.

Lower amorphous material shrinkage suggests the shrinkage in the center of the part after the surface has cooled will be less, reducing the likelihood of voids. The broad softening point range suggests amorphous materials can continue to be pumped into the center of the product over a broader temperature range. The lower the temperature at which material can still be pumped into the center; the less shrinkage between the melt temperature and room temperature. To optimize processing and reduce the shrinkage, solid profiles are run at low melt temperatures. Crystalline melting temperatures can only be reduced to a certain point, as the product state changes rapidly from melt to solid. Amorphous materials on the other hand can be run at cooler temperatures, as the plasticated state does not have a sharp melting point. Instead the viscosity gradually increases as the temperature is reduced until the pressure or work to move the melt exceeds the energy input to the process.

Typically, the process limiting step in solid profiles is the cooling operation. Longer forming boxes allow the process to be run at higher speeds. Alternatively, the second approach to producing higher throughput is to increase the number of ends with multiopening dies and forming boxes. One brake station can be used to process multiple ends, assuming it is properly aligned with each output position and the brake machine is large enough to handle the pressure involved.

After the brake machine or cutting, parts can be annealed to remove any internal stresses that may lead to warpage or premature failure when in service.

Review Questions

1. What is the biggest challenge in producing solid stock profile shapes?
2. What is the difference between a solid profile or stock process and normal profile extrusion?
3. How does a forming box work?
4. Why is it easier to run amorphous resins than crystalline resins in thick cross sectional parts?
5. What is the purpose of the brake equipment?
6. Assuming a 400 psi internal or forming pressure is required to prevent voids, what is the pressure required on a rectangular cross section that is 2×3 inches?
7. Explain void formation in the center of a thick cross section. Why is it a void and not an air bubble or gas?
8. What is the optimum fluid temperature to run in the forming box?
9. What determines the forming box length?

55 Reprocessing

Extrusion reprocessing is utilized in most operations to minimize the scrap generated and to recapture the material value. Regardless of the extrusion process, it is quite common to grind up scrap materials using a granulator or other appropriate equipment to make granules that can be fed back into the extruder. In compounding operations, the start-up material is granulated and fed back into the extruder in some predetermined level to augment the raw materials while reducing the scrap that has to be disposed. Experimental work is required to determine the percent regrind that can be added into the product without affecting the final properties. Another compounding option is to take products that do not meet customer specifications because of color or some other appearance problem, grind up the product, add black concentrate, and reextrude into a black product. Many colors can be added to the extruder with black concentrate to produce a black product that meets all the customer specifications. Regrind and off-specification materials (for color) can be mixed with black concentrate in a ribbon blender, continuous mixer, tumble blender, and so forth, to produce a uniform mixture that is extruded into a black product.

Recycled or reprocessed material can come from edge trim in sheet or film operations and flash from blow molding, off-spec products, and start-up material as the extrusion process is being brought to equilibrium. Scrap can be generated as postconsumer waste (waste generated by the consumer) and postindustrial waste (waste generated by plastic companies). Regardless of the source, scrap material, if properly used, can be a cheaper raw materials source. Reusing scrap generated in an extrusion facility lowers the overall extruded product raw materials costs. However, one must realize that an operation that generates no scrap or reclaim will have the lowest raw materials cost. It costs money to recover and reprocess scrap.

The size of the part or the material to be shredded or granulated determines the size and type of equipment required to recycle or reprocess material. In sheet and film extrusion, the edge trim can be removed pneumatically, if the sheet is both thin and flexible enough to be drawn away from the product. After removal, the sheet or film is either blown by air or sucked by vacuum to a shredder or chipper that cuts the film into small pieces, approximately 0.25×0.25 inch (6.4×6.4 mm). The chips are conveyed to the extruder feed throat, where they are mixed back into the product with the virgin material. Typical regrind concentrations range from 10–30%, with the correct ratio for a particular application determined experimentally. Regrind can affect the property profile and the processing due to viscosity and thermal stability changes. Experimentation determines what concentration can be added to the virgin resin without losing properties while maintaining the same processing characteristics. In a coextrusion operation, the edge trim may contain 7–11 layers with 5–7 different materials. After grinding, the edge trim may be fed back as a center layer to obtain the desired product thickness. Thick sheet is fed back to a web or sheet granulator that is specifically designed to granulate the product into a chip or granular form that can be fed back into the extruder. The feed throat may be equipped with feed rollers that pull the product into the granulator at a controlled speed in a continuous process. The residual granules are pneumatically or mechanically conveyed with augers to the desired location for reprocessing.

Depending on the product, the granulation may occur in a separate building where scrap is segregated and classified to be combined in a specific recipe later for feedstock into a particular product at a predefined rate. Multiple granulators or shredders may be dedicated to a particular product to prevent cross-contamination.

Unwind stations for off-grade film (blown or cast), foamed sheet, or thin-gauge sheet are capable of unwinding rolls 10–20 feet wide. Once the ends of the rolls are fed into a continuous unwind operation, the film is pulled at a constant rate into a shredder, where the product is cut into small pieces (0.25 to 0.50 inch squares [6.4 to 12.8 mm]). Once again the product is either packed directly in boxes and taken to some other part of the plant for processing, or it is mechanically or pneumatically conveyed to the extruder where it is used as 100% regrind or mixed with virgin resin. The end-use requirements determine whether the product can be used as is or has to be mixed with other resin.

In pipe, profile, and large-part blow, the start-up material and out-of-specification products are transported to another location in the plant where the profiles are fed manually or automatically into a large granulator to produce granules that can be fed back to the extruder. Normally the granulating is done in another building or another room to segregate the product, to minimize the noise associated with the granulation procedure on the rest of the plant, and to minimize dust and cross-contamination. Granulators are available with large feed throats and high horsepower to devour parts. If flexible tubing is being ground, a shredder may be necessary to properly cut up the part.

Wire and cable provides specific challenges due to the wire or cable and the plastic. Similar to other extrusion processes, start-up resin that does not include any wire or cable can be ground up. Once the plastic is applied to the wire or cable, special equipment is necessary to strip back the covering and convey it to a shred-

ding process to provide the correct particle size to feed back into the extruder. In this operation, the wire or cable is more valuable than the plastic, and special efforts are used to save the wire or cable so it can be reprocessed.

Laminated products present special challenges in using regrind or off-specification product. If all the components in the laminate are compatible, the laminate can be granulated and fed back into the process to produce its own special layer. Alternatively, the ground-up laminate can be fed to a compounding extruder, where the product is melted and homogenized into a useful plastic that can be sold as pellets or extruded directly into another application that meets the physical property requirements of the new compound. If the components in the laminate are not compatible, the challenge of finding a home for the product becomes greater. Compounding inhomogeneous resins or materials does not normally produce a product with a good physical property profile that can be used in other applications. One alternative is to incorporate the material in a center layer in a coextruded structure. Another alternative is to see if the product can be sold as is to a customer willing to take off-specification or B-grade material. Reextrusion or compounding of recycle to make a new product may require adding heat stabilizers, UV stabilizers, flame retardants, impact modifiers, or other additives to improve the product performance while producing a value-added product.

Munchy Ltd.[1] has a patented process to reprocess most thermoplastic products into pellets. The system can handle most waste streams, such as edge trim, flakes, film, bags, and so forth. The reprocessed plastics can be fed back into the main extrusion process at low percentage if the product meets all the customer specifications. Other applications include high melt flow resins used in nonwoven applications. The system can process nonwoven, melt blown, and spun bonded PP, feeding the products directly into the system and converting the product to pellets for reuse in the nonwoven process feed stream.

SUPPLIER

1. Munchy Ltd, Avtech House, Hithercroft Road, Wallingford, England OX10 9DA.

Review Questions

1. What is the advantage of reprocessing plastic in your facility?
2. What are some ways to recycle film?
3. Why might it be necessary to have the granulating operation in a separate room in the plant?
4. What are the different internal wastes that may be generated in an extrusion operation?
5. What is the difference between postindustrial and postconsumer waste?
6. What is the best way to recycle pipe, profile, blow molded parts, and thick sheet?
7. What is the best way to recycle coextruded parts?
8. What is the difficulty in recycling wire coating?

A

Abrasive Wear: Wear caused by the continual contact, under pressure, of hard particles in the resin against the barrel lining, screw, and valve components. The abrasive particles may be fillers or reinforcements, such as fiberglass, calcium carbonate, powdered metals, and others.

Accelerator: A chemical added to speed up a reaction; usually associated with crosslinking.

Activator: A chemical added to activate or "kick" a reaction; usually associated with crosslinking where they are used with an accelerator.

Adapter: The transfer tube that abuts the barrel, moving molten plastic from the barrel to the die. May also change the direction of melt flow: can have angled, crosshead and offset adapters.

Additive: A substance compounded into a resin to modify its characteristics (i.e., stabilizers, plasticizer, flame retardants, etc.).

Adhesive Wear: Wear resulting from two metals rubbing against each other, such as the screw flight lands and valve rings coming into contact with the barrel lining during operation.

Adiabatic Extrusion: The screw turning inside the barrel moving resin/polymer forward generates significant mechanical/friction heat. Sometimes on small extruders, after the extruder reaches equilibrium the external heaters can be turned off. This is adiabatic extrusion; no heat is entering or leaving the system. External barrel heaters must be turned on if the extruder speed is slowed down or stopped.

Alloy: Polymer blend having a modified interface and/or morphology.

Amorphous: Describing polymers with no crystalline structure; randomly arranged molecular chains.

Amorphous Phase: Devoid of crystallinity, no definite order. (Above the melting point, plastic is in an amorphous state.)

Anneal: To heat an article to a predetermined temperature and slowly cool it to relieve stresses.

Antiaging: Chemicals added to hinder degradation, for example, by oxidation.

Antiblock: An inorganic (i.e., fine ground calcium carbonate) material added to a polymer in very small quantities leading to a surface roughness inhibiting the ability of films to stick together.

Antioxidant: An additive to polymers (i.e., amines, alkylated phenols, phosphates, ester, etc.) that inhibits the oxidation of the plastic material. Used in plastics which are subject to oxidative degradation, i.e., polyolefins, ABS, polystyrene, acetal, and polyphylene oxide.

Antiozonant: A chemical added to hinder the chemical reaction known as oxidation (oxygen attack); usually associated with materials such as polyethylene which readily degrade in this way.

Antistat: Chemical additive (i.e., quaternary ammonium compounds, anionics, amines) that imparts a slight to moderate degree of electrical conductivity to the surface of plastic films to reduce the accumulation of electrostatic charges in finished products.

Apparent Shear Rate: The shear rate determined in capillary rheometer without correcting for shear thinning. The apparent shear rate is equal to $4Q/\pi R^3$ where Q is the volumetric flow rate (m^3/s) and R is the capillary radius.

Apparent Viscosity: The viscosity determined in capillary rheometry without correcting for shear thinning. Apparent viscosity is equal to the shear stress divided by the apparent shear rate.

Applesauce: Rough, wavy appearance of the extrudate; also referred to as orange peel, shark skin, and/or flow patterns.

ASR: Automatic Scrap Recovery System, designed to grind up edge trim and/or defective rolls of product and feed regrind directly back to the extruder hopper.

ASTM: American Society for Testing and Materials.

B

Back Pressure: The resistance to the forward flow of molten plastic in an extruder.

Barrel (Cylinder): A cylindrical housing where the screw rotates, including any special inner surface material, or replaceable liner. Sometimes referred to as a cylinder.

Barrel Shell (Backing Material): Outer thick wall of the barrel made from metal to provide strength and to hold the lining.

Barrier Flight: A secondary flight with reduced outside diameter designed to either separate melted polymer from solid polymer or enhance melting by having the polymer pass over it or to improve dispersive mixing in the metering section.

Barrier Screws: Dual flighted screws where one flight separates the solid bed from melt. The melting capacity of barrier screws is significantly higher than that of conventional screws.

Bayonet Adapter: A cylindrical shaped part with holding pins that threads into a thermocouple hole and is used to mount a spring loaded thermocouple.

Bearing (HUB): Portion of the screw immediately behind the flighted length that prevents the escape of material and provides a seal between the screw and barrel.

Bell End: A flange at the discharge end of the barrel that provides added strength to the barrel.

Biaxial Orientation: Process of aligning polymer molecules in two directions relative to each other. This process increases the force between adjacent molecules causing the end product to have modified properties, such as increased rigidity, tensile strength and barrier properties.

Bimetallic: A term used to indicate that a barrel is composed of two or more metals, commonly used to refer to barrels which have a centrifugally cast lining.

Bindicator: Warning system, attached to the extruder hopper indicating by means of an alarm bell or light that the hopper is "too full" or "too empty."

Blend: An intimate combination of two or more polymer chains not bound to each other with different characteristics.

Blender: A mechanical/electrical device to blend various resins into formulations for extrusion. This can be accomplished prior to extrusion directly above the extruder or in some remote location.

Bleed: Refers to the movement or migration of materials in plastic to the surface or into an adjacent material.

Blister Ring: A raised portion of the root between flights of sufficient height and thickness to effect a shearing action of the polymers as it flows between the blister ring surface and the barrel wall.

Blowing Agents (Foaming Agents): Substances generating gas at the extruder exit to produce a cellular structure in the extrudate. Materials include compressed gases that expand when pressure is released, soluble solids that leave pores when leached out, liquids that change to gases, and chemical agents that react with heat to form a gas.

Blow Molding: A method of fabrication where a warm plastic parison (hollow tube) placed in a mold cavity is blown up with air pressure to the shape of the mold cavity. Pressurized air is introduced to the parison through a blow pin.

Blown Film Extrusion: Process for making plastic film by extruding a tube and blowing it up to several times the die diameter. Also called tubular film or lay flat film.

Blow Up Ratio: The ratio of the bubble diameter to the blown film die diameter.

Bore: The inside diameter of the barrel that holds the screw.

Branched Polymers: Polymers with side chains attached to the molecular chain backbone.

Breaker Plate: A metal plate installed in the flow channel between the end of the extruder screw and the die adapter with holes or slots to allow polymer flow. It commonly holds screens for filtration and changes the motion of the melt before it reaches the die.

Breaker Plate Recess (Extrusion): Internal counterbore at the discharge end of an extruder barrel to accept the breaker plate and provide the seal and alignment of the die adapter.

Bridge: A result of feed material forming a bridge in the hopper or feed throat as the result of the feed material (size, shape, compressibility, or bulk density) or overheating causing the resin to stick to the feed throat of just forward of it.

C

"C" Clamp Flange: Circular tapered flange at the discharge end of the extruder barrel to hold the die adapter.

CAD/CAM: Computer Aided Design and Computer Aided Management.

Capillary Rheometer: Instrument used to measure polymer melt viscosity at different shear rates. Uses a ram extruder to push molten polymer through a round die.

Casting: Flat film process where extrudate is cast into a film in water or against a water cooled roll (chill roll).

Channel: With the screw in the barrel, the space between the flights bounded by the screw root and the barrel wall.

Channel Area (Axial): The channel cross-sectional area measured in a plane through and containing the screw axis.

Channel Depth: Distance in a radial direction from the barrel inside diameter to the screw root.

Channel Volume: Volume obtained from the "axial screw channel area" in one revolution about the screw axis. The measurement location is specific.

Channel Volume (Enclosed): Volume of the screw channel starting at the forward edge of the feed opening to the discharge end of the screw channel. Sometimes called screw inventory.

Channel Width: Distance across the screw channel in a direction perpendicular to the flight measured at the periphery of the flight.

Choke Plate: A single hole unit installed between the extruder barrel and the die holder to produce a controlled pressure drop in the melt.

CIM: Computer Integrated Manufacturing generally consists of a host computer that integrates all aspects of manufacturing. These include order entry, scheduling, material review and planning, inventory control, shipping and billing, as well as process engineering, quality control, production control, maintenance, and management of floor personnel communications.

Clearance (Screw/Barrel): Difference between the barrel inside diameter and the screw outside diameter. The radial clearance is one-half the diameter clearance.

Coefficient of Thermal Expansion: The fractional change in dimension (sometimes volume) specific to a material (plastic, metal or other) per unit change in temperature.

Coextrusion: Extrusion process whereby two or more melt streams are combined in the die to make an extrusion with two or more layers of plastic.

Coextrusion Capping: Extrusion process used to produce a plastic product that is topped or capped with another plastic.

Colorants: Dyes or pigments that impart color to plastics. Dyes are compounds that are soluble in most common solvents, yielding transparent colors. Pigments are organic and inorganic substances that are normally insoluble in common solvents and add color to the final product.

Color Concentrate: A compound with high die or pigment concentration in a carrier resin. Concentrate is added to resin formulations to color the finished plastic product.

Compounding: Combination of polymers with other materials either by means of mechanical (dry) blending or melt blending.

Compression Ratio (Channel Depth): The factor obtained by dividing the feed section channel depth of the screw by the meter section channel depth (or the depth of the last complete flight).

Compression Ratio (Volume): The factor obtained by dividing the channel volume of the screw at the feed opening by the volume of the last full flight prior to the discharge. This is not the commonly used definition of compression ratio.

Compression Section: This is the same as transition section of the screw.

Computer Control: A mode of operation where the extruder and line are controlled by a process computer.

Concentricity: Term used to describe 2 circles of cylindrical shapes having a common center and common axis, such as the inside and outside diameter of a barrel, pipe, or tube. Deviation from concentricity is called runout.

Cone and Plate Rheometer: A viscosity measuring device that uses the torque necessary to rotate a cone over a flat plate with molten polymer in between.

Conical Transition: Transition where the depth change is accomplished with a conical shape root surface (interrupted by the screw flights).

Consistency Index: In the power-law viscosity model, $\rho = m\gamma^{n-1}$, m is the consistency index, which is a function of temperature.

Controller (Heat/Cool): An electrical device in the extruder control panel to control the heating/cooling of the barrel, screen changer, adapter, and die zones to preset temperatures by the extruder operator. Controllers can be individual units for each zone or all zones combined into a single touch CRT screen.

Controller (Process): A computerized system which controls the extruder operation within preset parameters. Functions controlled include zone temperatures, extruder speed, die (gauge) adjustments, and resin blends. Since information is computerized, all extrusion lines in an operation can be monitored and controlled from a central location.

Converter: Company or individual engaged in changing a raw material into another form. For example, the process of converting resin pellets into film, converting films into laminated and/or coated/printed structures or converting the structures/film into bags.

Copolymer: A polymer, or resin, made by polymerizing two monomers together, for example, EVA: The second monomer is added to improve properties such as adhesion.

Core: Internal hole extending longitudinally from the screw shank through a portion of the screw to circulate a heat transfer medium to cool or heat the screw.

Core Pin: The center unit of a tubing die, used to form the extrudate inner wall, also known as a torpedo.

Corrosive Wear: Wear caused by various acids attacking the screws, barrels, valves, and other processing components. Acid formed in the plastic processing can erode and pit metal surfaces. Acids are generated by the polymers themselves or from flame retardants, foaming, and coupling agents.

Crammer Feed: A hopper unit that enhances feeding into the extruder feed throat.

Creeping Flow: Plastic flow over time due to application of force or weight. Flow occurs at very low Reynolds number, i.e., Re<<1, where the dimensionless Reynolds number is defined as

$$\mathrm{Re} = \frac{D(Diameter) \times \rho(Density) \times v(Velocity)}{\mu(Viscosity)}$$

Crosslinking: Mechanism whereby plastic molecules are linked or tied together, increasing the molecular weight and the viscosity. Gels may be crosslinked plastic.

Cross Model: Mathematical expression describing the shear thinning flow behavior. It is more realistic than the power-law model because it fits the data at both high and low shear rates.

Crystallinity: An ordered molecular structure that denotes uniform arrangement, alignment, and compactness of the molecular chains. Most polymers are semicrystalline and contain both crystalline and amorphous regions.

D

Deckle Rods: Metal insert used to close off the ends of a flat film or sheet die to produce a narrower product.

Degradants: Substances added to promote the breakdown of a product, for example, starch may be added to film to assist degradation in a landfill.

Degradation: Damaging change in the chemical structure, physical properties, or appearance of a plastic.

Density: Weight per unit volume of a plastic expressed in grams per cubic centimeter or pounds per cubic foot.

Desiccation: Removal of moisture from polymers.

Diameter (Barrel): The nominal inside dimension of an extruder or injection barrel.

Diameter (Screw): The nominal outside dimension of a screw.

Die: Metal orifice mounted to the discharge end of an extruder barrel to form a desired extrudate shape.

Die Adapter: Part of extruder die that holds the die block.

Die Block: Part of an extruder die that contains the orifice through which the melt flows to form an extruded product shape.

Die Bolts: Fine threaded bolts located around or across most dies to adjust the width of the die gap (distance between the die lips) to control the final product's thickness (gauge).

Die Lip Buildup (Die Drool): Gradual build up of deposits on the edge of the die exit that may partially obstruct the die and/or result in a defective extrudate surface.

Die Swell: Polymer melt exiting the die increases in diameter or thickness compared to the die diameter (or gap). Pulling the extrudate from the die reduces the swell and products can be drawn to sizes smaller than the die opening. Die swell is a misnomer and is properly called extrudate swell. "Dies don't swell, extrudate swells" quote by Bryce Maxwell.

Dispersive Mixing (also called Intensive Mixing): An operation that reduces the size of agglomerates or liquid drops of a minor component within a major fluid matrix. (1)

Distributive Mixing (also called Laminaror Extensive mixing): An operation that increases the minor component randomness or spatial distribution within the major fluid matrix. (1)

Drawdown: Maximum rate that film or sheet can be extruded through a die opening without breaking the film. The ratio between die opening and the thickness (gauge) of the finished product.

Drag Flow (also called Couette Flow): Flow between two surfaces caused by the movement of one surface relative to the other. Fluid is literally dragged by the moving wall. For parallel flat surfaces, the velocity profile is linear, varying from zero at the stationary wall to the velocity of the moving surface. In an extruder, the output is equal to the drag flow minus the pressure flow and the leakage flow over the flights. (1)

Dry Blend: A dry compound prepared without heat fluxing or the addition of a solvent. Done with PVC where powder is blended with plasticizer, stabilizer, lubricant and color in a high speed mixer.

Dynamic Mechanical Analysis (DMA): Response of material to an oscillatory load is measured through a temperature cycle. Provides information on modulus of elasticity which is related to impact strength, tensile strength, toughness, and creep rate.

E

Elastomer: A polymer which can be deformed under load and slowly recovers once the load is removed in a manner similar to rubber compounds.

Elasticity: The ability of plastic material to stretch (elongate) rather than crack or break under tension. (A value that can be recorded on a laboratory test device known as a tensile tester.)

Embossing: Techniques used to create depressions or a specific pattern in plastic film and sheeting.

Extensional Viscosity: Resistance to extensional flow. Common viscosity flow is the resistance to shearing. In polymer melts the elongational viscosity is normally between 3 and 100 times the viscosity. Melt strength is a rough determination of the elongational viscosity.

Extrudate: The stock or melt emerging from the die in a desired product form, such as pipe, coating on wire and others.

Extruder: A device capable of continuously transforming material (by means of heat and pressure) from a solid to a molten state and providing enough pressure so that it flows through the die, where it is transformed into a shape.

Extruder Barrel: Housing for the screw—designed to withstand high pressure of the melt process.

Extruder Drive: Driving mechanism that turns the gear reducer box which rotates the extruder screw. Usually a DC (direct current) variable speed motor with horsepower ranging from 50–1000 hp, depending on extruder size.

Extruder Gear Reducer: Speed reducing device driven by the extruder drive to convert the motor speed to a lower screw speed for melting and extruding plastic. Ratios usually run from 12:1 to 18:1.

Extruder Screw: Device used to convey, melt, mix, and develop the pressure necessary during melt processing.

Extrusion: Process for converting a polymeric material into desired shape which is defined by the die.

Extrusion Valve: An adjustable restriction in the melt stream used to control extruder back pressure.

F

Feed Housing or Section: A separate component of an extruder barrel assembly containing the feed opening, water cooling channels, and possibly a grooved interior lining. It is capable of withstanding high pressure (15,000 psi or more), especially if it is grooved, and may incorporate a thermal barrier between it and the barrel to which it is attached. Also referred to as the feed throat.

Feed Opening or Throat: A hole through the feed section of the barrel to introduce feed material into the barrel; also called a feed port.

Feed Section: The portion of the screw that receives the material to be processed and conveys it to the transition section. The feed section normally has a constant channel depth and constant root diameter.

Filler: An inert substance added to plastics for the purpose of improving the physical properties, processability, or to reduce the material cost.

Film: Thin plastic sheeting, usually taken to apply to thicknesses of 0.010" (250 microns) or below.

Fines: Resin in a fine powder form.

Fish Eyes: Defects in extrudate (usually film) that look like the eyes of a fish and are the result of small unmelted resin particles.

Flame Retardants: Additives that retard but do not self extinguish or completely stop the tendency of a product to burn. They are used in certain applications to conform to flammability safety regulations or codes.

Flaming: The action of applying flame treatment to a plastic surface to make that surface more receptive to printing inks. Also improves the print quality.

Flange: A short external section of the barrel, with a larger diameter, through which bolt holes are placed to either mount the barrel to the machine, or an end cap, die adapter, die, or other member to the barrel. It may also add strength to that section of the barrel.

Flash: Excess material around the edge of extrusion or molding.

Flight: Helical metal thread or raised portion of the screw.

Flight Cutback: The portion of the screw at the discharge end that is not flighted. This is normally included for calculations as part of the flight length.

Flight Depth: Distance in a radial direction from the screw flight land to the root.

Flighted Length: Overall axial length of the flighted portion of the screw, from the start of the feed pocket to the front end of the register or (in the case of extruder screws) the point where the root diameter begins to decrease. Flighted length does not include valves.

Flight Face: Face of the flight extending from the root of the screw to the flight land. The rear face is the side toward the feed section and the front face is the side toward the meter end of the screw.

Flight Hard Surface: A screw flight having its periphery harder (to increase wear resistance) than the base screw metal achieved by flame hardening, induction hardening, heat treating, depositing of hard facing metal, or other means.

Flight Land: The surface at the radial extremity of the flight constituting the periphery or outside diameter of the screw.

Flight Lead: The distance in an axial direction from the center of a flight land to the center of the same flight land after one complete turn. This is not flight pitch in a multi-flighted screw.

Flight Pitch: Distance in an axial direction from the center of a flight land to the corresponding point on an adjacent flight land.

Flood Feeding: System of feeding the extruder by keeping the hopper filled to the same level.

Flow Line (Streak): Flow lines or streaks in the machine direction are visual defects in the form of continuous lines or streaks which occur in the same axial location. They may appear and be very persistent after a change in material, screw or die.

Flow Line (Weld Line): A mark on the surface of a finished plastic extruded or molded piece, caused by the separation and recombining of the plastic melt inside dies and molds.

Form and Fill: A packaging process in which the container is formed and the contents inserted on line. Used in blow molding to obtain high speed production by the rapidly cooling the parison when cold liquid is injected into the bottle. In the packaging industry where a roll of film is formed, filled and sealed into a package in line.

Fourier Transform Infrared Spectroscopy (FTIR): Infrared spectroscopy where radiation is passed through a sample and the amount absorbed is measured as a function of wavelength. FTIR is used for polymer identification by comparing spectra with known materials.

Freeze Line: Place on a blown film bubble where semicrystalline thermoplastic material starts to crystallize on cooling.

Frictional Heating: Highly viscous polymer melts flowing through a channel generate considerable heat from internal friction. Approximately 80% of the heat required in melting a polymer in an extruder is supplied by the screw rotation through generation of frictional heat. Flow in dies can cause significant local temperature increase (hot spots) and thermal degradation from frictional heat resulting in viscosity problems, product discoloration, black specks, etc.

Front Radius: Radius at the intersection of the front or pushing side of the flight and the screw root. This radius is normally smaller than the rear radius and may vary from one section of the screw to another.

G

Gate: The slide at the base of the hopper used to stop and start the flow of material, or part of the die. A hinged gate may be used to attach the adapter to the extruder.

Gauge: Thickness of finished product, as measured by micrometers or beta gauges. A measure of thickness, for example, 100 gauge is 1 mil or 0.001 inch.

Gaylord: A very heavy-duty corrugated box, approximately 4' × 4' × 3' high with its own pallet. It is filled with resin or concentrate, holding up to 1500 pounds, for shipment, storage, or direct application by the end user.

Gear Pump: A pressure control device fitted to the discharge end of a conventional extruder to produce a more stable output.

Gel: Lumps or hard specks seen in film resulting from resin degradation, crosslinking, or high molecular weight particles.

Gel Permeation Chromatography (GPC): Measurement technique to separate polymer fractions by flowing a pulse of a polymer solution through a packed bed of porous particles. Measuring polymer concentration in the effluent stream, and comparing to calibration standards, the molecular weight distribution is determined.

Glass Transition (Tg): Temperature where polymer goes from a glassy brittle state to soft rubber state. Temperature coincides with the on set of molecular motion in the polymer chain occurring in the amorphous region of the polymer. For HDPE and LDPE it is –100°C and for PS +100°C.

Granules: Plastics material supplied in the form of small pellets approximately 0.125 inch in diameter and length.

Grooved Liner: A liner whose bore is provided with longitudinal grooves, usually for several diameters in the feed section only.

H

Haul Off: A series of rolls or caterpillar that pulls the extrudate away from the die.

Heater Bands: Electrical heating elements mounted on or around the barrel, adapter, dies, nozzles, etc.

Heat of Fusion: Heat required to mobilize the molecules of a solid polymer to produce a fluid melt, i.e., the heat required to destroy the solid crystalline structure without increasing the temperature. For amorphous polymers like polystyrene (PS) the heat of fusion is zero.

Heat Stabilizers: Additives used in plastics to protect them from the harmful effects of heat during the extrusion process.

Heating Zone: A portion of the barrel length having independent temperature control of the heater bands.

Helix Angle: The angle of the screw flight at its periphery relative to a plane perpendicular to the screw axis.

Homopolymer: A polymer consisting of only one molecular material.

Hopper (Feed): A funnel-shaped container mounted directly on the barrel over the feed opening to hold material feeding the hopper.

Hopper Car: Railroad car loaded at the resin manufacturer to transport resin to the end user. Divided into four compartments and filled with up to 220,000 pounds of resin.

Hopper Loader: Device that moves resin to the hopper from silos, Gaylords, or other bulk storage by means of a vacuum system.

Hygroscopic: Term applied to polymers having a tendency to absorb moisture from the air.

I

Inlay: Hard surface portion of a flight land not extending across the full flight width.

Interfacial Instability in Coextrusion: Highly irregular or sometimes regular waviness that appears in coextruded structures at the polymer/polymer interface.

Involute Transition: Transition where the change in depth is accomplished with a root surface that remains parallel to the screw axis but is a different depth.

ISO: International Standards Organization, a body like ASTM, which issues test standards or procedures.

J

JIT: Just In Time Production—means that components are not produced until they are required.

K

K: Either temperature measurement in Kelvin (for example 250 K) or it refers to the k value of PVC, which is a measure of molecular weight.

Kelvin: A temperature scale based on absolute 0. 0 °C corresponds to 273 K.

Key: Part of the screw which permits the turning motion of the drive to be transmitted to the screw.

L

Lace-Cut: Normally refers to granules which have been produced by chopping strands of circular cross section.

Laminates: Layers of different film, sheet, or other materials plied together to form one structure.

Lay-Flat Width: Measure of the size of tubular film that is the width in double thickness. A tube with diameter, d, gives a lay-flat width of $\pi\frac{d}{2}$.

L/D Ratio: Ratio of the flighted length (distance from the front edge of the feed opening to the end of the screw flight) to the screw outside diameter. The ratio calculation is simplified to dividing the flighted length of the screw by its nominal diameter.

Lead (Screw): Distance measured parallel to the screw axis from the top of one edge of the screw flight to the same edge of the same flight after one complete turn. It is not the distance measured to the next screw flight in a multi-flighted screw (See Pitch).

Linear Molecule: A long chain molecule with no or minimal side chains or branches.

Liner: Wear resistant removable sleeve in the barrel.

Lining: Internal wear resistant portion of an extrusion or injection barrel. The lining may be metallurgically bonded to the barrel shell inside diameter or be a removable liner that is pressed or shrunk fit into the barrel shell bore.

Load Cell: Instrument placed in a gravimetric blending system to electronically weigh and record the weight of each batch of resin.

Loss Modulus (denoted as G"): Indirect measure of polymer viscosity using a cone-and-plate instrument subjected to dynamic (sinusoidal) deformation (see also Storage Modulus). (1)

M

Machine Direction: The lengthwise or longitudinal direction of an extrudate.

Masterbatch: A concentrated blend of pigments, additives, fillers, etc., in a base polymer. Masterbatch is added in small amounts to a large volume of material (the same as or compatible with the base polymer) to achieve desired properties. (1)

Mandrel: A term used to describe a sizing element either in a die or in a sizer which controls one dimension of the extrudate; this dimension is usually the inside diameter.

Melt: Plastic material in a molten state.

Melt Channel: Channel in a barrier screw designed to collect and convey forward the melted polymer.

Melt Flow Index (also called Melt Index or Melt Flow Rate): The number of grams of polymer that can be pushed out of a capillary die of standard dimensions (diameter: 2.095 mm, length: 8.0 mm) under the action of standard weight (2.16 kg for PE, at 190°C) in 10 minutes (ASTM Standard 1238). The usual melt index range is from less than 1.0 (called fractional) to more than 25 (up to 100 for injection molding). For PP it is usually called MELT FLOW RATE and the standard temperature is 230°C. (1)

Melt Fracture: At higher throughput rates, extrudates can become highly distorted with significant fluctuations in head pressure. This phenomenon is known as gross Melt Fracture. It is possible in some situations to obtain grossly melt fractured extrudates without sharkskin, i.e., the surface remains smooth and glossy but overall the extrudate is distorted (see also Sharkskin). (1)

Melt Plug or Melt Blockage: Sudden drop in output rate of an extruder resulting from insufficient forward conveying of the solid packed bed in the feed section of the extruder.

Melt Strength: Measure of polymer melt extensional viscosity. It represents the maximum tension that can be applied to the melt without rupture or tearing. Usually a capillary rheometer is used to extrude a polymer strand that is pulled by a pair of rollers till rupture.

Melting Point: The temperature at which the structure of a crystalline polymer is destroyed yielding a liquid. HDPE is about 135°C, and LDPE is about 110°C. Amorphous polymers do not have sharp well defined melting points; instead they soften over a range of temperatures. However, in extrusion the glass transition temperature plus 50°C is a temperature above which many amorphous polymers can be processed.

Metallizing: Technique used to deposit a metal (usually aluminum or chrome) on a plastic surface.

Metering Section: Portion of the extruder screw at the discharge end with a constant flight depth and having a length of at least one turn of the flight.

Mixing Section: Area of the extruder screw with a special geometry designed to enhance distributive and/or dispersive mixing in the polymer melt.

Molecular Weight Distribution: The percentage of different length molecular chains within a resin. Defined as narrow when chains lengths are close to average length. Defined as wide when chains have a large standard deviation from the average length.

Molecular Weight: The weight of polymer chains based on the sum of the elemental atomic weights of the atoms comprising the individual polymer chain. The weight of related molecules in a polymer chain. Relates to melt index, i.e., as molecular weight increases the melt index decreases and vice versa.

Monomer: A single chemical compound when combined with many other monomers constitutes a polymer.

MRP—Management Requirements Planning: A part of "CIM" dedicated to reviewing inventories of such things as raw materials, finished goods, packaging, other miscellaneous parts that is based on actual and projected historical order entry and current orders for those items.

N

Newtonian Flow: An isothermal, laminar flow characterized by a viscosity that is independent of the shear rate. The shear rate at all points in the flowing liquid is directly proportional to the shear stress and vice versa. Viscosity does not change with shear rate. Simple liquids such as water and mineral oil exhibit Newtonian flow.

Nitriding (Gas): Surface hardening technique used with certain alloy steels by heating the steel in an atmosphere of nitrogen (ammonia gas) to approximately 950°F. A very hard (70+Rc) case depth of 0.007" to 0.015" results that is wear resistant. Process commonly used for barrel inside diameters, screw, and valve components.

Nitriding (Ion): Surface hardening technique used with certain alloy steels by heating the steel to approximately 600°F in an atmosphere of hydrogen gas, adding an electrical charge to the steel and nitrogen gas, allowing a bombardment of positively charged steel by hydrogen and nitrogen gas ions. This creates a hard (70+Rc) wear resistant surface. Case hardness is slightly greater and more uniform in depth than gas nitriding with less distortion or contamination of the work piece. Process is used on screws and valve components.

Non-Newtonian: Liquids whose viscosities or flow are dependent on the shear rate as well as temperature and pressure. Plastics are non-Newtonian.

Normal Stresses: Polymer melts when sheared (i.e., when subjected to tangential forces) give rise to perpendicular stresses. A fluid flowing in a tube is less compressed in the axial direction than in the radial direction. These stresses lead to extrudate (die) swell. Polymers with high molecular weight tails are more elastic giving larger normal stresses.

Nose Cone: Conical surface at the discharge end of an extrusion screw.

O

Olefins: A group of unsaturated hydrocarbons with the general formula C_nH_{2n}, and named after the corresponding paraffins by the addition of "ene" or "ylene" to the stem in place of "ane." Examples are ethylene and propylene.

Orientation: Alignment of molecular chains or crystalline structure in polymeric materials to produce a highly uniform structure and higher strength in the molecular direction. Can be accomplished by stretching during fabrication.

Oxidation: Degradation of resin by oxygen. Caused by exposure to oxygen at high temperatures. Example is an empty extruder screw sitting at high barrel heat, causing any remaining resin in the screw or barrel to degrade.

Overall Length (Screw): Total length of the screw parallel to the screw axis. It includes the flighted length, the shank, and any nose cone.

P

Parison: Hollow tube of molten polymer that is pinched closed at one or both ends and inflated in a mold to make a hollow part.

Permeability: Passage of gas, vapor, liquid, or solid through a barrier.

Pilot (Barrels): A cylindrical portion at the rear end of an extruder barrel used to locate the barrel to the feed throat.

Pitch: Distance measured parallel to the screw axis from the edge of the top of the screw flight to the same edge of an adjacent flight. In the case of a multi-flighted screw the pitch is less than the lead.

Plasticizer: A chemical added to a plastic or resin to make it softer and more flexible.

Plate-Out: Additive that migrates through the plastic during processing and builds up on the die or the take off equipment.

Pocket: Place where the screw flight is initiated, normally starting from a cylindrical area or another flight. A feed pocket exists on most screws and is located at the intersection of the bearing and the beginning of the flight.

Polydispersity Index (PDI): Ratio of the weight average to number average molecular weight (M_w/M_m).

Polymer: A high-molecular weight organic compound, natural or synthetic, whose structure can be represented by small repeat units (monomers) forming chemical bonds with other monomers. If two or more monomers are present a copolymer is obtained.

Polymerization: A chemical reaction where the molecules of monomer are linked together to form large molecules whose molecular weight is a multiple of the original substance. When two or more monomers are present, the process is called copolymerization.

Power-Law Model: Simple mathematical expression describing the shear thinning behavior of polymers. $\eta = m \cdot \dot{\gamma}^{n-1}$ (1)

Press Fit: An interference fit, characterized by constant bore pressure throughout, achieved by mechanically forcing the entering piece into the receiving piece using a press or similar machine. Barrel liners may be press fit into the prepared barrel shell.

Pressure Plug: A threaded brass plug, located at the end of the barrel prior to the breaker plate. Safety device designed to rupture at high pressures to prevent damage to the extruder or injury to employees.

Pressure Tap: Hole designed to accept a pressure transducer or pressure gauge.

Pressure Transducer: A mechanical/electrical device placed in the melt stream to monitor the back pressure and transfers the information to a readout device.

Pseudoplastic Flow: This term is synonymous to shear thinning flow, i.e., viscosity decreases as the shear rate increases. (1)

Pump Ratio: The ratio obtained by dividing the depth of the second metering section by the depth of the first metering section of a two-stage screw.

Purging: Action of cleaning an extrusion cylinder or barrel by running a special compound through it. Generally done between color changes.

Pushing Side: Flight face of the screw flight facing the discharge end of the extruder and runs from the front radius to the top of the flight land. This surface is commonly close to being perpendicular to the screw axis.

Pyrometer: Instrument used to measure temperature.

R

Rear Radius: Radius at the intersection of the rear or trailing side of the flight and the screw root. Normally this radius is larger than the front radius and may change from one location of the screw to another.

Recycled Plastic: Plastic prepared from discarded articles that have been cleaned and reground.

Regrind: Scrap material left over from previous operations which is pulverized and added to fresh resin as a regular production. Reclaimed material through a process of grinding.

Relief: An area of the screw shank of lesser diameter than the outside diameter and located between the bearing and the spline or keyway.

Relaxation: Polymer melts subject to mechanical work develop stresses that do not immediately go to zero when the mechanical influence is removed. The time required for the stresses to relax to zero is called the relaxation time. Higher molecular weight polymers have longer relaxation times. Polymer solidifying with stresses have frozen-in stresses which are released when the plastic part is reheated.

Resin: Class of solid or semisolid organic products that are either natural or synthetic with high molecular weight and generally no definite melting point. Broadest terms, any polymer that is a basic material for plastics. Most resins are polymers.

Reynolds Number: A dimensionless quantity defined as:

$$\mathrm{Re} = \frac{D(Diameter) \times \rho(Density) \times v(Velocity)}{\mu(Viscosity)}$$

which is equivalent to the ratio of INERTIA forces to VISCOUS forces. The flow is turbulent when the Reynolds number is more than 2100 for tubes. Below 2100 the flow is laminar (i.e., streamlines without disturbances). For molten polymer flow, the Reynolds Number is usually in the range 10^{-4} to 10^{-2} (see also CREEPING FLOW). (1)

Rheology: Study of the flow of polymeric liquids under heat and pressure.

Rheometer: Device used to measure the melt properties of plastic materials.

Rockwell Hardness: Common method of expressing a material's hardness by testing its resistance to indentation. Results are expressed with various scales; the most common is the "C" scale, identified as Rc, used to compare the hardness of barrel linings, screw surfaces, and valve components. A similar test, Brinnell hardness, expresses results on a somewhat different scale and is identified as HB.

Root: Surface of the screw between the flights, usually cylindrical or conical shape, and a smaller diameter than the outside flight diameter.

S

Screen Changer: Mechanical device attached to the end of the extruder, equipped with a sliding plate, having two locations for breaker plates. One plate is in the melt stream while the other is being cleaned or replaced. Some changers incorporate continuous changing based on back pressure.

Screen Pack: Woven metal screen or equivalent device installed across the melt flow between the extruder screw and the die and supported by the breaker plate. Used to filter contaminants in the melt or to increase the back pressure or both.

Screw: A helically flighted shaft that rotates within the barrel to mechanically work and advance the materials being processed.

Screw Axis: Reference line of finite length drawn through the center of the screw at the rear of the shank and the center at the discharge end.

Screw Channel: The open section between screw flights.

Screw Diameter: Dimension defined by the cross-section of the screw bounded by the flight lands. Normally, expressed as a nominal diameter (such as 3.5"), its actually diameter could be 3.490" to 3.492".

Screw Speed: Number of screw revolutions per minute (RPM).

Sealing Ring: Round ring used in place of a breaker plate to seal between the extruder barrel and the die adapter.

Shank: The non-flighted portion of the screw at the drive end.

Sharkskin (Surface Mattness): Failure of extrudate to exhibit smooth and glossy appearance. The surface exhibits a repetitious wavy or ridged surface pattern.

Shear Flow: The sliding of imaginary fluid creates slices that slide parallel to each other, like a deck of cards. Shearing occurs whenever fluids flow through tubes and channels. The velocity is zero at the wall surface and maximum at the center. So the fluid is being sheared as it flows through a tube or channel. (1)

Shear Rate: The velocity gradient, i.e., velocity divided by the gap measured in reciprocal seconds, s^{-1}. In extruder screw channels, the shear rate can usually reach 100 s^{-1} or more. In flow through extrusion dies, it might reach 500 s^{-1} or more, and in injection molding more than 5000 s^{-1}. (1)

Shear Ring: (See Blister Ring).

Shear Stress: A tangential force divided by the area (force/area) over which it is applied. Shear stress is equal to the viscosity multiplied by the shear rate (measured in units of pressure, i.e., MPa or psi). At the die lips under normal processing conditions the shear stress may reach values of 0.2 MPa (29.0 psi) or more. The normally accepted value for the onset of sharkskin in a capillary is 0.14 MPa (20.3 psi), although higher values are reported in industrial production. (1) With additives, critical shear stress value might reach 0.5 MPa (72.5 psi).

Shear Strain: Movement of one layer of polymer relative to an adjacent layer divided by the layer thickness.

Shear Thinning: The reduction of the viscosity as the shear rate increases, which is exhibited by polymeric liquids. Shear thinning is due to molecular chain alignments in the direction of flow and disentanglements. (1)

Shear Viscosity: Ordinary viscosity is the ratio of shear stress to the shear rate. (1)

Sheeting: Blown or cast process producing a continuous product of more than 0.010" thick.

Shrinkage: Volume change associated with the cooling of melted resins. Amorphous resins shrink less than crystalline. Note: Film can continue to shrink for 4–6 weeks after manufacture.

Shrink Fit: Interference fit, characterized by a constant bore pressure throughout, achieved by heating the larger receiving piece to allow the smaller mating piece to enter. In some cases, barrel liners are shrunk fit into the prepared barrel shell.

Slip, Slippage: When fluids flow it is assumed that the velocity at a surface is zero (or equal to the surface velocity if the surface moves). Virtually all polymer melts exhibit some slippage on the surface, especially when the shear stress levels are high, e.g., over 0.1 MPa (14.5 psi). Stick-slip phenomena are responsible for the onset of sharkskin whenever polymers are extruded at shear stresses higher than 0.14 MPa (20.3 psi). Some additives and processing aids promote slippage. Slippage is beneficial for delaying the appearance of sharkskin at higher throughput rates. (1)

Silo: Tall, cylindrical device, approximately 65' tall by 12' diameter. Designed to hold the contents of hopper cars up to 200,000 pounds.

Sleeve: See Liner.

Slip Agent: An organic material (usually purified animal fat) added to resin in small amounts that migrates to the surface and causes the finished plastic to be slippery.

Soak/Soak-In: The time it takes to heat a cold extruder, adapter, screw, die and polymer to the desired temperature. It is a common practice to increase the set points in 100°F (55°C) increments and wait until the zone reaches the set point, and then allow the extruder to stay at that temperature (soak-in) for at least 30 minutes before increasing another 100°F (55°C), etc. As thermoplastics are poor heat conductors they require significant time for external heat to “soak” through the plastic mass and turn it to liquid.

Solids Channel: Channel in a barrier screw that is connected to the feed channel and is designed to contain and convey forward unmelted polymers.

Solids Conveying Zone: The single-screw extrusion process consists of three functional zones: They are:

- The SOLIDS-CONVEYING ZONE, where the polymer pellets or powder are compacted and transported forward;
- The MELTING ZONE, where the polymer melts mainly under the action of shear on the barrel wall;
- The METERING ZONE (PUMPING ZONE), where the polymer is transported forward by DRAG FLOW caused by the rotating action of the screw. (1)

SPC (Statistical Process Control): Method of monitoring the process limits and providing statistical quality control data using Xbar-R charts. Function is to identify when the system is out of the control limits and correction to the process is required.

SPE: The Society of Plastic Engineers.

Specific Gravity: Ratio of the density of a liquid or solid material to the density of water at 4°C, or other specified temperature. Specific gravity is often misused as a synonym for density. Specific gravity is a unitless number.

SQC: Statistical Quality Control.

Square Pitch: Screw flight formed by a helix angle of 17.66°. The pitch is equal to the nominal screw diameter, i.e., 3.5 inch diameter screw with a 3.5 inch lead.

Starve Feeding: An extruder feed system that controls the throughput by using an auxiliary feeder to control the production rate.

Standard Deviation, Sample: This is a statistic that measures the amount of variability within a sample. It is computed using the following equation:

$$s = \sqrt{\frac{1}{n-1}\sum_{i=1}^{n}(x_i - \bar{X})^2}$$

It measures the spread or dispersion in data. It is calculated by:

- taking the difference of the individual observations from the average
- squaring the differences and summing them
- dividing the sum by the number of observations minus one
- taking the square root

Storage Modulus (denoted as G'): The ratio of shear stress to strain (deformation) when dynamic (sinusoidal) deformation is applied in a cone-and-plate rheometer. It relates to the elasticity of the polymer melt. G'', the loss modulus is also determined in dynamic (sinusoidal) measurements and relates to the viscous behavior. So, G' and G'' together give an idea of the dual nature of the polymer melt (partly elastic solid and partly viscous fluid). Measurements of G' and G'' provide information on polymer structure and might be related to molecular weight distribution, crosslinking, etc. (1)

Strain: A technical term synonymous to deformation. (1)

Surging: Pronounced fluctuation in extruder output over a short period of time without a corresponding change in operating conditions.

Swing Gate: Hinged device mounted at the extruder barrel discharge and used to support the die adapter.

Swing Gate Flange: Flange at the discharge end of an extruder barrel using bolts to attach the die adapter to the barrel.

T

Thermal Mechanical Analysis (TMA): In this technique, a sample is deformed under a static load as its temperature is being changed. Glass transition and softening points can be measured. The amount of orientation can also be measured by TMA. (1)

Thermogravimetric Analysis (TGA): This technique is used to measure a variety of polymeric phenomena involving weight changes, such as sorption of gases, desorption of contaminants (monomers, solvents, and additives) and degradation. In TGA, a sample is placed on a balance beam in an oven. From the weight versus temperature curves, kinetic and other studies are carried out. (1)

Thermocouple: Device made of two dissimilar metallic conductors in contact that produces an electrical voltage that depends upon the temperature difference between the sensing point (the bimetallic junction) and the measurement point (voltmeter). The resulting temperature measurement is used to monitor and control the barrel or nozzle temperature. Commercial thermocouples are housed in a steel tubing adapter.

Thermocouple Hole: A machined hole in a barrel, nozzle, or adapter designed to accept the thermocouple adapter.

Thermoform: Process of forming a sheet of plastic around a preformed mold.

Thermoplastic: A material that repeatedly softens when heated and hardens when cooled. Typical thermoplastics are polystyrene, copolymers, acrylic, polyethylene, polypropylene, vinyl, nylon, and engineering resins.

Thermoset: A material that changes as a result of a chemical reaction in the presence of heat and/or pressure, catalysts, or ultraviolet light, from a soluble fusible stage to an insoluble infusible state. Typical thermoset materials are unsaturated polyesters (BMC), amines (melamine and urea), alkyds, epoxies, and phenolics.

Thrust Bearing: Support bearing at the rear of the screw. Designed to withstand the high head pressures developed during extrusion processing.

T.I.R.: An abbreviation used to identify tolerances with respect to concentricity. Total Indicator Reading is standard terminology for drafting and machinist work to describe the deviation in concentricity of a measured surface from a selected surface as shown on a dial indicator. Also known as F.I.M. (Full Indicator Movement).

Tolerance: Predetermined and specified allowance for deviations with a range. Generally used to control weight, gauge, measurements, etc.

Torpedo: See Nose Cone.

Trailing Side: Flight face of the screw that faces the extruder feed end and runs from the rear radius to the top of the flight land.

Transducer: Device for transferring power generated in one system to another system. Various types of electronic transducers are used in connection with molding and extruding equipment to measure, monitor and/or control linear position, position, temperature, speed, and other factors.

Transition Section: Portion of the screw between the feed section and the metering section where the flight depth decreases toward the discharge end. Section is sometimes referred to as the compression section.

Transparent: Capable of a high degree of light transmission, i.e., glass. Transparent materials include polycarbonate, acrylics, styrenics.

Transverse Direction: The direction across the extrudate, or 90 degrees to the machine direction in the horizontal plane. Sometimes referred to as Cross Direction.

Twin Barrel: A dual cylindrical housing in which two extruder screws rotate side-by-side, including a nitrided inner surface or a centrifugally cast bimetallic lining for resistance to wear. Two twin barrel designs are one where two extruder screws intermesh and a second where the two extruder screws do not intermesh.

Two-Piece Barrel: A cylindrical housing where the screw rotates and is constructed of two sections. One section extending the full length of the barrel, consists of an outer shell (typically constructed of annealed alloy steel) with a full outside barrel diameter over the majority of its length that tapers abruptly to a much smaller diameter over the final one-fourth (approximately) of its length toward the discharge end of the barrel. This section commonly has a centrifugally cast, bimetallic lining for resistance to wear. The second section, is an outer shell with full outside barrel diameter (construction of heat treated alloy steel) that is shrunk fit over the smaller diameter of the first section and welded to the larger diameter of that section at the taper area. The second section provides added yield strength to the barrel at its discharge end.

Two-Stage Screw: A screw for a vented barrel consisting of a feed, transition, and meter sections followed by a decompression section (or vent section) located under the vent allowing volatiles to escape without vent bleed. The decompression section is a deep channel and is usually several diameters in length, followed by a second transition and metering section.

U

Ultraviolet: Zone of invisible radiation beyond the violet end of the spectrum of visible radiation. UV has enough energy to initiate some degrading chemical reactions with most plastics.

Ultraviolet Stabilizers (UV Light Absorbers): Chemical agents which absorb or screen out radiation beyond the violet end of the spectrum of visible radiation. Often combined with other additives, like heat stabilizers, and antioxidants.

Under Water Pelletizing: Another name for die face pelletizing. A process used to produce pellets from strands by cutting them underwater at the die face.

V

Vent: The opening in an extruder barrel that allows gases to escape when used in conjunction with an extruder screw with a matching decompression zone.

Vent Bleed: The unplanned escape of melt through the vent during the operation of vented barrel processing.

Vented Extruder: An extruder containing a hole or port in the barrel where material can be removed, or introduced into the plastic material. Often used to extract volatiles from a plastic material.

Vent Port: An opening through the barrel wall to permit the removal of air, water vapor and volatile matter from the material being processed.

Vent Stack: A device surrounding a major portion of the vent deflector designed to prevent any material that escapes through the vent port during start up from collecting on the barrel, heater bands, or wiring.

Venturi Cooling Ring: A unit primarily used to cool film by using a primary air stream to draw in additional air by the Venturi effect and thus improve cooling. It is also used to stabilize, or size, the film while it is cooling.

Viscoelasticity: The dual nature of polymers is that they are partly viscous fluids and partly elastic solids, i.e., viscoelastic. In flowing polymers, viscoelasticity is responsible for time-dependent properties, such as stress relaxation, normal stresses, very large elongational viscosities, and numerous unusual phenomena such as extrudate swell, entry flow vortices and some flow instabilities. (1)

Viscosity: Internal friction or resistance to flow of a liquid.

Volatiles: A general term for material that is a gas at processing temperatures.

W

Weldlines: Weldlines are formed when the polymer flow is interrupted by obstructions in the melt flow field, i.e., in the die. The high viscosity melt must recombine and the polymers reentangle to have high strength in this area. After separation, recombination of polymer molecules is very slow due to the high viscosity and viscoelastic nature of polymer melts. Weld lines tend to be lower in mechanical properties unless the polymers have time to recombine. Such defects are common in extruded pipes, bottles, and film.

Wire Covering: An extrusion operation for the application of a plastic coating onto a wire. Usually done with crosshead extrusion.

Y

Yield: Quantity of finished product produced from a fixed amount of starting material.

Z

ZD (Zero Defects): Process with the goal of producing perfect products all the time.

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