1. EVOLUTION OF OIL EXTRACTION

1.1. Hand Press Extraction

Documented oil extraction dates back to 1650 B.C. when ripened olives were pressed by hand in Egypt using wooden pestles and stone mortars. The extracted olive oil was filtered through goat hair filters and used as a lubricant. Sesame, linseed, and castor oils were recovered in Egypt by hand pressing as far back as 259 B.C. (1).

1.2. Early Mechanized Extraction

By 184 B.C., the Romans developed more sophisticated technology such as edge-runner mills and screw and wedge presses. These technologies combined leverage and the use of animal power to aid in the milling and extraction of the oil. From Roman times until the eighteenth century, similar technology was used for oil extraction (1).

In the eighteenth century, wind and water power largely replaced animal power to assist in oil extraction. Large wind-driven stamper mills became popular in Europe. The wind turned a vane outside the oil mill, and the rotational energy was transmitted into the mill via shafts and gears, eventually rotating a horizontal cam shaft. The horizontal cam shaft had vertical stamper shafts connected to it. The initial stamper shafts were used as mortars to beat the oilseeds into a pulp inside a wooden pestle. The pulp was then transferred into filter bags woven from horse hair.
and placed vertically between opposing wedges. Additional stampers pounded the wedges together, squeezing the oil through the filter bags, where it could be collected. Hundreds of such oil mills sprung up across Europe (2).

1.3. Hydraulic Press Extraction

In 1795, J. Bramah of England invented the hydraulic press for oil extraction (1). Oilseeds were milled, cooked, and wrapped in filters cloths woven from horse-hair. The oilseeds wrapped in filter cloths were manually loaded into perforated, horizontal boxes below the head block and above the ram of the press. The boxes were pressed together using upward hydraulic pressure on the ram. The oil was pressed out through the filter cloths surrounding the oilseeds. The filter cloths and spent cake were manually removed from the hydraulic press. The residual oil in spent cake was approximately 10%.

In 1801, the first cottonseed oil mill was constructed in the United States using hydraulic presses (1). By the 1870s, American technology in hydraulic pressing had outpaced European technology. Large hydraulic presses with up to 16 press boxes and up to 400 tons of force were being used. In 1874, Rose, Down, and Thompson (2) of Hull, England, began marketing the American-designed hydraulic presses with the advantage of preforming the cakes to increase productivity. Facilities using this joint technology were commonly referred to as Anglo-American oil mills. In the late 1800s, German companies were producing hydraulic cage presses, with rams pressing the oilseeds inside of vertical slotted barrels that did not require filter cloths. By the end of the nineteenth century, hydraulic press oil mills were the standard technology for oil extraction.

In 1900, Alfred French founded the French Oil Mill Machinery Company in Piqua, Ohio, for the purpose of advancing hydraulic press technology. He was awarded a patent (3) for the automatic cake-trimming machine for automating the sizing of the cakes prior to pressing. He also developed and patented the “change valve” in 1905, which allowed the hydraulic press to change pressures near the end of the pressing cycle to squeeze additional oil. The continuous stacked cooker was patented in 1907 along with an innovative cake former. Two-pass pressing was another of French’s developments, taking final residual oil in cake below 5%. French hydraulic presses became the industry standard in the United States in the 1920s (3). Figure 1 indicates a typical French hydraulic press of that era.

Hydraulic press oil mills remained in use as late as the 1950s before the last of them were replaced with continuous screw presses and continuous solvent extraction plants, both of which required far less labor and could process at much higher rates.

The olive oil industry is the only oilseed industry still using hydraulic presses today. This is possible because of the price premium paid for natural olive oil, processed without the use of heat or chemicals.
1.4. Screw Press Extraction

In 1900, Valerius D. Anderson invented the mechanical screw press in Cleveland, Ohio. He was awarded a U.S. patent for the apparatus (4). The mechanical screw press was a radical departure and significant technological advancement over the hydraulic presses being used at the time. The mechanical screw press used a vertical feeder and a horizontal screw with increasing body diameter to impart pressure on the oleaginous material as it proceeded along the length of the screw. The barrel surrounding the screw was slotted along its length, allowing the increasing internal pressure to first expel air and then expel the oil through the barrel. The expelled oil was collected in a trough under the screw, and the de-oiled cake was discharged at the end of the screw. The primary advantage of the mechanical screw press was that it allowed continuous oil extraction and could process large quantities of oleaginous materials with minimal labor. Figure 2 illustrates an early Anderson Expeller™ (trademark for Anderson mechanical screw press).

In the early 1900s, a number of European companies developed variations of the V. D. Anderson design concepts and began manufacturing mechanical screw presses. In the 1930s, after the U.S. patent had expired, other American firms did the same. These machines were able to replace large numbers of hydraulic presses.
In 1951, the V.D. Anderson Company was again the pioneer in mechanical screw pressing, being first to patent the process of using a mechanical screw press to continuously prepress oleaginous materials ahead of continuous solvent extraction plants. The company was awarded U.S. Patent 2,551,254 (5).

In the past 100 years, the primary improvement in mechanical screw press design has been developing materials of construction that extend wearing part life. Screw and barrel parts that once lasted three months before requiring replacement may now last up to two years. Additionally, mechanical screw presses have been built to much larger scale, going from initial capacities of 5 tons per day up to present-day capacities of over 100 tons per day for full pressing and over 800 tons per day for prepressing applications.

1.5. Solvent Extraction

In 1855, Deiss of Marseilles, France, was first to employ solvent extraction (1). He used carbon disulfide to dissolve olive oil retained in spent olive cakes. This technology used batch solvent extraction, where the material was held in a common kettle for both the extraction process as well as the subsequent meal desolventizing process. Deiss obtained a patent for batch solvent extraction of olive oil in 1856 (1). Small batch solvent extraction plants were installed in France and Italy, and by 1870, small batch solvent extraction facilities had spread across Europe. Larger scale solvent extraction plants were supplied by Rose, Downs, and Thompson (2) of Hull, England, starting in 1898.
In the early 1920s, with the availability of petroleum-based solvents, German inventor Hildebrandt invented the continuous countercurrent immersion extractor and German inventor Bollman invented the continuous two-stage percolation extractor. The first commercial scale continuous solvent extraction plants were installed in Germany in the late 1920s, with the Hansa-Muhle facility in Hamburg using Hildebrandt extractors being the largest (6). Between 1934 and 1937, the Hansa-Muhle Company of Germany supplied the first continuous solvent extraction plants to oilseed processors in the United States, using Hildebrant-type extractors. Leading German solvent extraction technology was set back due to World War II.

The American company, French Oil Mill Machinery Company of Piqua, Ohio, developed continuous solvent extraction plants starting in 1939, using a multistage Bollman-type percolation extractor (3). Soon thereafter, the Belgian Company, Extraction De Smet, started manufacturing continuous multistage belt-type extractors in 1945. V.D. Anderson of Cleveland, Ohio, joined the solvent extraction equipment supply market in 1948 with continuous Hildebrandt-type immersion extractors (5).

A small-scale chain conveyor type of extractor was developed in the 1940s at Iowa State University with the intent of using trichloroethylene solvent to extract soybeans. Crown Iron Works of Minneapolis, Minnesota, licensed the technology and extractor design, and supplied several continuous solvent extraction plants in 1951 using the new extractor and trichloroethylene solvent (6). The meal from these plants proved detrimental to animals, so the plants were either closed or converted to petroleum-based solvents (7). The chain-type extractor apparatus continued on and is the basis of the modern Crown Iron Works Model III extractor used today.

In the first continuous solvent extraction plants in Germany, the meal was desolventized in Schnecken’s desolventizers. The Schnecken’s desolventizer design consisted of a series of steam-jacketed conveyors, stacked one above the other, which used indirect heat to desolventize the meal. This was the prevalent meal desolventizer design used by all equipment suppliers through the 1940s.

In 1939, Kruse and Soldner (8) of Central Soya Company patented a process of mixing water into desolventized soybean meal and drying it in a vertical stacked cooker to improve meal quality. In 1940, Hayward (9) of Archer Daniels Midland described a similar process, referred to as “toasting” the meal. In 1948, Hutchins of French Oil Mill Machinery Company and Kruse of Central Soya Company worked together to develop an apparatus and process for desolventizing soybean meal and toasting soybean meal in the same vessel. The apparatus was a vertical stacked vessel. Live steam was introduced primarily in the top tray of the vessel to desolventize the meal and increase its moisture. The high moisture meal was then dried using indirect steam conducted through the trays lower in the same vessel. This new apparatus was known as a combination desolventizer-toaster, commonly abbreviated as “DT.” Hutchins was awarded an apparatus patent, and Kruse was awarded a process patent for the DT (10, 11). This new desolventizing technology was considered state-of-the-art until the 1980s.

The Blaw Knox Company of Buffalo, New York, invented a rotary-type continuous solvent extractor in 1958 (12). The rotary-type extractor held the material in
cells that rotated in a horizontal plane around a vertical axis, and it included four countercurrent extraction stages. The Blaw Knox trademark for this apparatus was the Rotocel® Extractor. The Rotocel Extractor was licensed to equipment suppliers around the world and quickly became very popular. Figure 3 illustrates a Rotocel Extractor. A similar apparatus, named the Stationary Basket Extractor, was invented by French Oil Mill Machinery Company in 1962 (3).

As solvent extraction plants were scaled up in size in the 1960s, the drying trays of the DTs were often not scaled up proportionately. The result was higher meal moistures exiting the DT, requiring a subsequent downstream meal drying step. Meal drying was generally accomplished using a rotary steam-tube dryer. In 1972, Schumacher of Germany invented the vertical stacked meal dryer-cooler, often abbreviated as “DC.” This apparatus used heated air to dry the meal and cool air to cool the meal. As the DC was similar in construction to a DT, Schumacher also developed the concept of stacking a desolventizer-toaster above a dryer-cooler to form a combination desolventizer-toaster-dryer-cooler, commonly abbreviated as a “DTDC.”

Desolventizing technology was further advanced in the late 1970s by Schumacher. The original Schumacher DT was a single, deep material depth tray with live steam injected from below. Steam injection from below allowed countercurrent stripping of solvent, and the deep material depth allowed adequate time for toasting. Steam distribution through the deep material bed was uneven, and the original Schumacher DT was difficult to operate. By 1980, Schumacher developed an improved version of the Schumacher DT, which included predesolventizing trays at the top of the vessel, along with perforated countercurrent trays below that supported thinner layers of meal and allowed the steam entering the bottom of the vessel to pass through the meal more evenly. Schumacher patented the hollow stay-bolt type of perforated countercurrent tray (13) used in the improved Schumacher DT in
1981, and he licensed this technology to equipment suppliers around the world. The Schumacher DT was often supplied above a DC to form a Schumacher DTDC.

In 1985, Mason of French Oil Mill Machinery Company developed a variation to the Schumacher DTDC design. The Mason design used a hollow stay-pipe type of perforated countercurrent tray design with clusters of perforations through a single plate at the top of the pipes. The DC also included the option of steam dryer trays prior to the air dryer trays, with the water vapor from the steam dryer trays being returned to the solvent extraction process for heat recovery. Another variation of the Schumacher DT was developed in 1997 and patented in 1999 by Kemper of the French Oil Mill Machinery Company and the Farmer of Bunge Corporation, which used a perforated countercurrent tray composed of a combination of indirect heating surface and slotted screen elements for more even distribution of countercurrent stripping steam (14).

The latest extractor to be developed for mainstream oleaginous material extraction applications is a rotary-type extractor with fixed slotted floor and bevel gear drive trademarked as the Reflex® Extractor. This extractor was developed by Kemper et al. (15) of the French Oil Mill Machinery Company in 1995 and patented in 1997. The Reflex Extractor is an improvement on the Rotocel Extractor. One of these units installed at the Louis Dreyfus solvent extraction plant in Argentina has processed in excess of 9000 metric tons of soybeans per day, making it the world’s highest capacity extractor.

2. SEED PREPARATION

Upon receipt at the processing plant, oleaginous materials require varying degrees of seed preparation prior to the oil extraction process. Seed cleaning, seed drying, size reduction, hull removal, heating/drying, flaking, and extruding are all potential unit processes involved in seed preparation.

2.1. Seed Cleaning

Good quality oleaginous materials entering a processing plant will contain up to 2% foreign material. Foreign material is generally removed twice, once prior to storage and again as the oleaginous material enters the continuous process. The foreign material to be removed may consist of a combination of weed seeds, sticks, pods, dust, soil, sand, stones, and tramp metal.

Tramp metal is generally the first foreign material to be removed to protect all downstream equipment from damage. Tramp metal is separated using magnetic force to pull the iron-based materials from the oleaginous materials. Magnetic devices commonly used are plate magnets, drum magnets, and magnets suspended over a conveyor. Plate magnets allow the oleaginous materials to pour across their surface and rely on the metallic material sticking to the magnetic surface. Plate magnets are relatively inexpensive but require stopping the process occasionally to clean the magnetic surface and suffer from high wear rates. Drum magnets
rely on a fixed magnet inside a quadrant of a rotating drum. The metal debris sticks to the rotating drum, allows the oleaginous materials stream to fall away, and then the metallic objects fall off the drum later in the rotation. Drum magnets have the advantage of being continuous and self-cleaning, but suffer from high wear rates. The last type of magnet is a magnet suspended over a conveyor. This system generally consists of a pair of electromagnets suspended over the oleaginous materials passing in a thin stream over a belt conveyor. The metallic debris is pulled upward onto the surface of the electromagnets. This type of system has the advantage of no wear and, due to redundancy, allows cleaning during operation. There are numerous suppliers of various devices to remove tramp metal.

Sticks and pods are typically larger than oleaginous materials. In addition, sticks, pods, and dust are typically lighter than oleaginous materials. Therefore, the method of removing this foreign material from the oleaginous materials stream is typically a combination of course screening followed by aspiration. This combination of processes is commonly referred to as scalping. Companies such as Buhler of Switzerland, Carter-Day of the United States, and Kice Metal Products of the United States are well known for supplying scalping equipment.

Weed seeds, sand, and soil are typically smaller in physical size than the oleaginous materials. These foreign materials can be removed by fine screening. The hole sizes in the screen are selected slightly smaller than the oleaginous materials, allowing the soil, sand, and weed seeds to sift through. Buhler of Switzerland and Rotex of the United States are popular suppliers of such screening equipment.

Particularly with ground nuts (peanuts), there may be a significant quantity of stones approximately the same size as the ground nuts. These stones are not removed by scalping or by fine screening. The stones are heavier than the ground nuts and need to be removed by gravity separation. Gravity separators manufactured for stone removal are commonly referred to as destoners. Buhler of Switzerland and Triple-S Dynamics of the United States are suppliers noted for manufacturing destoners.

### 2.2. Seed Drying

The moisture of oleaginous materials often needs to be reduced to minimize degradation in storage and to enhance the effectiveness of downstream unit operations. For example, soybeans are often received at 13% moisture and need to be dried to 10% moisture to facilitate hull removal.

Oleaginous materials are generally dried in large, vertical, open-flame grain dryers. These dryers have multiple columns of oleaginous materials that slowly migrate downward in a plugged flow. The oleaginous materials are dried in the upper portion of the column and cooled in the lower portion of the column. Cool air is pulled laterally through the lower portion of the column to cool the oleaginous materials and warm the air. The warm air is further heated with open flame and pushed laterally through the upper portion of the column to heat and dry the oleaginous materials. Many regional manufactures of cereal grain dryers are used to supply this equipment.
2.3. Size Reduction

Most oleaginous materials require size reduction prior to further processing. Exceptions are canola, rapeseed, and corn germ, which are already sufficiently small in size. For most oleaginous materials, they need to be broken into pieces 2 to 3 mm across to enhance the downstream unit processes of hull removal, heating/drying, and flaking.

Soybeans, ground nut (peanuts), and copra are typically reduced in size using an apparatus known as a cracking mill. These mills generally consist of two sets of cylindrical corrugated rolls in series. The rolls operate at differential speeds to assist in breaking the oleaginous materials apart. The first set of rolls often has a coarser corrugation, and the second set of rolls has a finer corrugation to reduce the particle size in two steps. Modern cracking mills can process up to 1000 tons per day of oleaginous materials each. Figure 4 illustrates a modern, high-capacity cracking mill. Buhler of Switzerland and CPM Roskamp of the United States are the primary providers of cracking mills to the vegetable oil industry.

Sunflower seed and delinted cottonseed have very coarse hulls on the exterior of the seed, requiring more intense breaking. Devices for breaking these seeds into smaller pieces are generally supplied by Buhler of Switzerland and Carver of the United States.

2.4. Hull Removal

Soybeans, sunflower seeds, and delinted cottonseed have a course outer seed coat known as a hull. The hull fraction is high in fiber content and low in oil and protein
content. Soybeans, sunflower seeds, and delinted cottonseed often have a large portion of their hull fraction removed so that the finished meal fraction will be higher in protein by weight, and therefore higher in value.

The process of removing the seed coat of soybeans is commonly referred to as dehulling, and the process of removing the seed coat from sunflower seeds and delinted cottonseed is commonly referred to as decortication. In both dehulling and decortication, the process has two distinct stages. In the first stage, aspiration is used to remove the lighter hull fraction from the heavier meats fraction. A certain amount of small meats particles are also aspirated away with the hulls stream. In the second stage of dehulling, the fine meats are separated from the hulls through various means of hull agitation and screening. The effectiveness of a dehulling system is often measured by the residual fiber content (from hulls) in the meal fraction, and the residual oil content (from meats) in the hull fraction. For soybeans, the industry standards are less than 3.5% fiber content remaining in the meal fraction and less than 1.5% oil content remaining in the hull fraction.

In soybean dehulling, the aspiration can take place with the soybean at elevated temperature, or at ambient temperature. Dehulling at elevated temperature is referred to as hot dehulling, and dehulling at ambient temperature is referred to as cold dehulling. Dehulling at elevated temperature has the advantage of fewer fine meats particles being aspirated away with the hull fraction in the first stage of dehulling and, therefore, requires less separation of meats from the hull fraction in the second stage of dehulling.

Dehulling systems are most commonly supplied by Buhler of Switzerland, Crown Iron Works of the United States, De Smet of Belgium, Kice of the United States, and Rotex of the United States. Decortication systems are most commonly supplied by Buhler of Switzerland and Carver of the United States.

2.5. Heating/Drying

With the exception of cold pressing, all oil extraction processes require that the oleaginous materials be heated and sometimes further dried before oil extraction.

In order to enhance the downstream unit operation of flaking, oleaginous materials are typically heated in the range of 60–75°C temperature. By heating and softening the oleaginous materials, it enables the oleaginous material to stretch and flatten in the flaking operation with a minimum of fragmented particles being created.

For oleaginous materials being prepressed prior to solvent extraction, they are typically heated in the range of 90–110°C temperature to decrease the viscosity of the oil and allow the prepress to expel the oil and make a good quality cake. This heating is often in two stages, initial heating to approximately 65°C prior to flaking and then final heating to approximately 100°C prior to entering the prepress.

Oleaginous materials that are full pressed are often heated to temperatures of 110–150°C temperature, and dried to as low as 3% moisture. The high temperature decreases the viscosity of the oil, making it easier to expel. The high degree of drying largely ruptures the cellular structure of the oleaginous material as the internal
moisture vaporizes and expands. The low final moisture maximizes friction within the full press to maximize internal pressure. These functions all improve de-oiling and allow residual oil in cake to be minimized.

Another aspect of heating soybeans in particular is the impact on the phospholipase enzyme. The phospholipase enzyme is activated at approximately 55°C and remains activated up to approximately 100°C. In this temperature range, and with sufficient exposed surface area and time, the phospholipase enzyme modifies a portion of the phosphatides in the oil fraction by splitting off the non-fatty acid moiety (16). The resultant calcium and magnesium salts of phosphatidic acids that are formed tend to be more oil-soluble than water-soluble, thereby converting phosphatides from a hydratable form to a nonhydratable form (16). This has a resultant impact on the quantities of acid, caustic and silica needed to reduce the phosphorus content of the soybean oil in the downstream degumming and refining unit operations.

There are several types of apparatus used for heating/drying oleaginous materials. The rotary steam-tube conditioner, stacked-tray cooker, plugged flow conditioner, and hot air conditioner are most popular.

Rotary steam-tube conditioners are horizontal, cylindrical, shell, and tube heat exchangers that rotate on large rollers mounted under the shell. High-pressure steam is used on the tube side as the heating medium, and the oleaginous materials fill the lower half of the shell side, between the tubes. The rotating vessel is sloped downhill, with the oleaginous material particles entering at the center of the higher end of the vessel, and exiting at the center of the lower end of the vessel. Inside the unit, the oleaginous materials are mixed and heated by conductive heat transfer with the steam-filled tubes. The distribution of residence time of the oleaginous materials within the rotary steam-tube conditioner is very consistent. Typical residence times for oleaginous material applications are in the range of 10 to 30 minutes. Single units process as much as 5000 tons per day. The disadvantages of the rotary steam-tube conditioner are that it requires significant floor space, and it is difficult to maintain the seals at the product inlet and outlet. There are a number of regional suppliers of rotary steam-tube conditioners that supply the industry.

The stacked-tray cooker is a vertical, cylindrical vessel with a multitude of horizontal trays. The oleaginous materials enter at the top and are supported by the tray. The material is mixed above each tray and conveyed downward from tray to tray by agitating sweeps anchored to a center rotating shaft. The heat for increasing particle temperature and evaporating moisture is conducted into the oleaginous materials from the upper surface of the trays, filled with high-pressure steam. The stacked-tray cooker has the advantage of limited floor space. However, the stacked-tray cooker has the disadvantage of a wide distribution of residence time among particles, and it is limited to about 2000 tons per day capacity in a single vessel. Major suppliers of stacked-tray conditioners are Crown Iron Works of the United States, De Smet Group of Belgium, and Krupp Elastomertechnik of Germany.

Plugged flow conditioners are vertical, square columns with a multitude of horizontal oval-shaped tubes inside the column. The oleaginous materials have a
plugged flow and totally surround the steam-heated tubes. The oleaginous materials discharge the conditioner by variable-speed live bottom screws. The advantages of the plug flow conditioner are that they require no electrical mixing power, they have very tight distribution of residence time among particles, and they require little floor space. Plugged flow conditioners are supplied by Buhler of Switzerland and Crown Iron Works of the United States.

In soybean plants using hot dehulling, the cracked soybeans are shocked by hot air to aid in the separation of the hull fraction from the meats fraction. The hot air conditioner uses either a hot air fluidized bed to heat and convey or hot air spouting to heat and a belt to convey. The hot air is partially recirculated for heat recovery. Hot air conditioners for soybean plants are provided by Buhler of Switzerland, Crown Iron Works of the United States, and Escher Wyss of Germany.

### 2.6. Flaking

Most oleaginous materials are flaked prior to solvent extraction. Flaking distorts the cellular structure of the oleaginous material and reduces the distance that solvent needs to penetrate to reach the oil in the oleaginous material cells. The apparatus used for flaking is commonly referred to as a flaking mill. The flaking mill has two large diameter rolls in parallel, turning in opposing direction at approximately 250 to 300 r.p.m., and forced together by hydraulic cylinders. The softened oleaginous materials are uniformly fed into the nip of the two rolls. As the oleaginous materials are pulled through, they are stretched and flattened. Typical flakes are in the range of 0.3 to 0.4 mm thick and 8 to 18 mm in diameter.

Good solvent extraction performance is highly dependent on good flaking performance. It is critical that no particles pass through the flaking mill without being flaked, and it is critical that flake thickness is kept uniform across the entire width of each flaking mill. To accomplish good flaking mill performance, the flaking mills must be fed very uniformly at all times, and the roll surfaces must be kept parallel. Common practice is to grind the surfaces of the rolls on an as-required basis to maintain the parallel condition. It is also important to maintain good top-down aspiration of the flaking mill to remove dust from the feeding apparatus, and to remove surface moisture from the flakes.

Modern flaking mills are manufactured in capacities of 300 to 500 tons per day. Figure 5 illustrates a modern, high-capacity flaking mill. Buhler of Switzerland and CPM Roskamp of the United States are the major suppliers of flaking mills to the vegetable oil industry.

### 2.7. Extruding

A wet extruder can be used to enhance the performance of solvent extraction on soybeans and cottonseed, and a dry extruder can be used to enhance the performance of a full press.

A wet extruder is a high-strength cut-flight screw conveyor equipped with steam injection along the housing and a restricted orifice discharge. These units are
commonly referred to as Expanders® (Anderson International trademark). The oleaginous flakes are heated in the range of 90–110°C temperature using a combination of electrically generated internal friction and live steam heat. The pressure developed prior to the restricted orifice discharge increases the product density. As the high-temperature, high-moisture product exits the restricted orifice, the pressure drop allows the product to abruptly flash off moisture. The result is a hot, fragile, pellet-like product, with greater bulk density than the incoming flakes, greater porosity than the incoming flakes, and significant cell rupture. To enable good solvent extraction performance, the extrudate must be dried to 10% moisture and cooled to 60°C temperature to firm up the soft extrudate and prevent evaporative cooling and recondensing of moisture enroute to solvent extraction.

Wet extruders are used extensively prior to solvent extraction in cottonseed processing. Cottonseed flakes with 30% oil content are very fragile and have poor miscella flux rates within the extractor, causing inconsistent results and requiring very long extraction times. By extruding the flakes, consistently good results can be achieved with moderate extractor sizing.

Wet extruders are also used ahead of a number of soybean solvent extraction plants. In plants with undersized extractors, the energy used by the extruder can be compensated by additional oil yield from the extractor. Steam consumption is reduced in the desolventizer toaster. Additionally, the wet extruder in a soybean plant increases temperature sufficiently to stop the phospholipase enzyme from converting additional hydratable phosphatides into nonhydratable phosphatides. This reduces acid, caustic, and silica consumption in the downstream refinery. However, in soybean plants with adequately sized extractors where potential oil yield
improvements are minimal, the additional operating costs of wet extruders are difficult to justify.

Modern wet extruders can process up to 1500 tons per day of flakes per unit. They are equipped with variable orifice apertures to aid cleaning after power stoppages. Figure 6a illustrates a modern, high-capacity Expander. Anderson International of the United States and Technal of Brazil are the two major suppliers of wet extruders used in the vegetable oil industry.

Dry extruders are occasionally used ahead of full presses. Dry extruders use electrical power to generate internal friction to heat the product as high as 150°C temperature. When the extruded product exits the dry extruder, it is liquid-like in consistency with thorough cell rupture. When this product is fed to a full press, the

Figure 6. (a) Expander. Courtesy of Anderson International. (b) Dry-type extruder. Courtesy of Insta-Pro International.
Press throughput can be increased significantly. Because of the short residence time of less than 30 seconds in the dry extruder (17), the protein solubility of the meal produced is higher than with traditional processing, whereas trypsin inhibitors are sufficiently deactivated. The resultant high-protein solubility meal is best suited for poultry and swine consumption.

The principle advantage of dry extruder preparation is that no expensive stacked-tray cookers or steam boiler are required, and the total capital investment for facilities under 100 tons per day in size is significantly less than for traditional full-press or solvent extraction processes. Therefore, this means of preparation is commonly applied in very small-scale plants commonly referred to as “mini-mills.” The primary disadvantage of dry extruder preparation is that the meal produced is higher in fat content than many poultry producers desire, and lower in bypass protein than many cattle producers desire. Thus, meal produced from a full-press mill using dry extrusion preparation generally obtains a lower price than meal produced from a traditional full-press operation.

Dry extruders available today are limited in capacity, typically processing less than 50 tons per day each. Figure 6b illustrates a typical dry-type extruder. Major suppliers of dry extruders are Anderson International of the United States and Insta-Pro International of the United States.

3. MECHANICAL EXTRACTION

3.1. Palm Fruit Extraction

Palm fruit is different than most oleaginous materials and requires a unique preparation consisting of sterilization, stripping, and digestion. Sterilization is a heating process under pressure to stop the development of lipolytic enzymes in palm fruit bunches. Stripping is the mechanical process used to separate the stalks and leaves from the palm fruits. Digestion is a process of heating the palm fruits to approximately 90–100°C temperature for 20 minutes prior to pressing (18).

Mechanical extraction of the palm oil from the digested palm fruit is a delicate process. The screw press must squeeze as much oil as possible from the fruit without breaking the palm kernels. Palm-kernel oil will mix with the palm fruit oil if the kernels are crushed. International trading rules allow a maximum of 5% palm-kernel oil in palm oil.

The presses used to extract the palm oil are low-pressure screw presses. A low-pressure screw press is a mechanical device that uses a horizontal screw with increasing body diameter to impart pressure on the palm fruit as it proceeds along the length of the screw. Some designs use parallel twin-screw technology. The barrel surrounding the screw(s) is perforated along its length, allowing the increasing internal pressure to first expel air and then expel the oil through the barrel. The expelled oil is collected in a trough under the screw(s), and the de-oiled material discharges at the end of the screw. Low-pressure screw presses can reduce residual oil in palm fruit fibers to 5% while keeping palm-kernel breakage rates below 15% (18).
The palm oil exiting the low-pressure screw press has approximately 66% oil, 24% moisture, and 10% solids (18). The solid particles are typically separated from the oil using the traditional method of pumping the oil into a tank with approximately 2 hours of residence time to allow the heavier solid particles to settle and be continuously dredged from the base of the tank. After gravity separation, the oil is then pumped through a liquid cyclone to remove residual solids. Solid particles separated from the clean oil stream are saturated with oil and recycled back into the process. After the palm oil is cleaned of solid particles, it is heated and pumped through a vacuum oil-dryer to remove moisture.

3.2. Full Press Extraction

Because of comparatively poor oil yields, sole use of mechanical extraction to separate the oil and meal fraction is not as commonly used as solvent extraction. The mechanical extraction process can reduce the oil in meal to 5% to 10% by weight, whereas the solvent extraction process reduces the oil in meal to less than 1% by weight. As the value of the oil fraction is typically two to three times the value of the meal fraction by weight, the loss of yield is very costly. The mechanical extraction process also has comparatively higher energy and maintenance costs per ton of oleaginous materials processed.

There are four primary reasons why the mechanical extraction process is still selectively used. First, the mechanical extraction process can be furnished in very small scale, as low as 10 tons per day. The capital cost for small mechanical extraction facilities is considerably less than small solvent extraction facilities. In remote locations, freight differential can compensate for higher operating costs and lower yields. Second, there is a niche, high-value market for natural oils that have not been in contact with solvents or chemicals, requiring the use of mechanical extraction. Third, mechanical extraction can create a high bypass protein meal for ruminant animals that sells at a price premium over solvent extracted meal. Finally, mechanical extraction is often considered more reliable than solvent extraction when processing difficult materials (copra and palm kernel) in hot, tropical climates.

The apparatus used to mechanically extract the oil fraction from the meal fraction is a full press or Expeller® press (Anderson International trademark). A full press is a mechanical device that uses a horizontal screw with increasing body diameter to impart pressure on the oleaginous material as it proceeds along the length of the screw. The barrel surrounding the screw is slotted along its length, allowing the increasing internal pressure to first expel air and then expel the oil through the barrel. The expelled oil is collected in a trough under the screw, and the de-oiled oleaginous material cake discharges at the end of the screw. The full press has one function, to de-oil the oleaginous material as low as possible. The key to full-press performance is to apply maximum pressure to a thin cross section of the oleaginous material to squeeze out as much of the oil as possible. As a result, full presses create tremendous heat. It is common for full presses to use water-cooled shafts and water-cooled or oil-cooled barrels to dissipate the heat to
maintain adequate internal friction and pressure. Most full presses are capable of processing 10 to 100 tons of oleaginous materials per day. Figure 7 illustrates a 100 ton per day, modern full press. Smaller capacity full presses can typically obtain lower residual oil in meal, at the expense of consuming more power and maintenance cost per ton of seed processed. Today there are numerous suppliers of full presses, with Anderson International of the United States, De Smet Rosedowns of the United Kingdom, French Oil Mill Machinery Company of the United States, Insta-Pro International of the United States, and Krupp Elastomertechnik of Germany being the most notable international suppliers.

The fraction of oil removed in the full press is laden with fine meal particles, typically in the range of 5–15% by weight. These meal particles are typically separated from the oil using the traditional method of pumping the oil into a tank with 30 to 60 minutes of residence time to allow the heavier meal particles to settle and be continuously dredged from the base of the tank. After gravity separation, the oil is then pumped across a vibratory screener or through a pressure-leaf filter for final meal particle separation. Meal particles separated from the clean oil stream are saturated with oil and recycled back into the process ahead of the full press.

3.3. Pre-Press Extraction

The solvent extraction process relies on the prepared oleaginous material maintaining its structural integrity while in the extractor. Oleaginous materials with greater than 30% oil by weight tend to break down in the extractor after a large portion of the oil is extracted, causing poor final extraction and high solvent retention. A solvent extraction plant operating with an oleaginous material containing approximately 20% oil is in thermodynamic balance. The waste heat from desolventizing
the meal fraction is sufficient to serve as the principle heat source for evaporating the solvent from the oil fraction. If a high oil content material is sent to the solvent extraction plant, the thermodynamic balance is lost and the operational energy required in the solvent extraction process increases significantly. As a result, oleaginous materials with greater than 30% oil content typically have their oil content reduced to 20% prior to solvent extraction. Rapeseed/canola, sunflower, ground nut, wet process corn germ, palm kernel, and copra are common oleaginous materials with an excess of 30% oil content, which require a reduction in oil content prior to solvent extraction.

The apparatus used to reduce the oil content of oleaginous materials prior to solvent extraction is commonly referred to as a prepress. A prepress is a mechanical device that uses a horizontal screw with increasing body diameter to impart pressure on the oleaginous material as it proceeds along the length of the screw. The barrel surrounding the screw is slotted along its length, allowing the increasing internal pressure to first expel air and then expel a portion of the oil through the barrel. The expelled oil is collected in a trough under the screw, and the partially de-oiled oleaginous material cake discharges the end of the screw. The prepress has two important functions, partially de-oiling the oleaginous material to 20% oil content and producing a porous cake with adequate structural integrity to allow high downstream solvent extraction efficiency. The key to prepress performance is to convert electrical energy into pressure rather than heat, by configuring the segmental screw elements to provide sufficient forward conveyance with an optimum combination of intermittent pressurization and mixing along the length of the barrel. Modern prepresses are capable of processing 500 to 1000 tons of oleaginous materials per day. Figure 8 illustrates an 800 ton per day, modern full press. These large machines are typically driven with 200–400-kw motors. Today there are two major suppliers of prepresses, De Smet Rosedowns of the United Kingdom and Krupp Elastomertechnik of Germany.

![Figure 8. Prepress. Courtesy of De Smet Rosedowns.](image-url)
The fraction of oil removed in the prepress is laden with fine meal particles, typically in the range of 5–10% by weight. These meal particles are separated from the oil using one of two methods. The traditional method is to pump the oil into a tank with 30 to 60 minutes of residence time to allow the heavier meal particles to settle and be continuously dredged from the base of the tank. After gravity separation, the oil is then pumped through a pressure-leaf filter for final meal particle separation. The alternative method of separating the meal particles from the oil is through use of a high-speed centrifugal decanter. In both methods, meal particles separated from the clean oil stream are saturated with oil. These oil-rich meal particles are either recycled back into the process ahead of the prepress or are mixed with the partially de-oiled cake and sent forward to the solvent extraction process.

4. SOLVENT EXTRACTION

4.1. Introduction

The overwhelming majority of all vegetable oil is extracted using solvent extraction. The solvent extraction process has the benefit of significantly higher oil yields than mechanical extraction, along with lower unit operating costs. The major drawback of solvent extraction is the high initial capital cost to construct a facility. Solvent extraction facilities constructed today are commonly in the size range of 1000 to 5000 tons per day, costing $15 million to $75 million to construct.

The solvent used in the majority of oilseed solvent extraction plants around the world is commercial hexane, a mixture of hydrocarbons generally boiling in the temperature range of 65–69°C. Most commercial hexane available contains approximately 65% normal hexane, with the remaining 35% of the composition consisting of cyclopentane and hexane isomers. As hexane vapor is three times heavier than air and slight amounts of hexane mixed in air can create an explosive mixture, special care must be taken in constructing and operating solvent extraction plants. The National Fire Protection Agency bulletin NFPA-36 Solvent Extraction Plants is the recognized guide for safe construction and operation of such facilities.

Because of special safety considerations, the solvent extraction process is constructed in a separate facility from the seed preparation process. The solvent extraction process consists of five closely interrelated unit processes: solvent extraction, meal desolventizing, meal drying and cooling, miscella distillation, and solvent recovery.

4.2. Solvent Extractor

The extractor is the apparatus in the solvent extraction process where the vegetable oil fraction of the oleaginous material is separated from the meal fraction of the oleaginous material by dissolving the oil fraction in a solvent.

The prepared oleaginous material is conveyed from the seed preparation process to the solvent extraction process and enters the solvent extractor. The solvent
extractor conveys the prepared material from its inlet to its exit, providing the prepared material approximately 30 to 120 minutes of residence time. While the material is being conveyed forward, miscella (solvent and oil solution) is washed down through the bed of material to extract out the vegetable oil. Each miscella wash is of a decreasing concentration of vegetable oil. After four to eight miscella washes, the material is washed once more by fresh solvent, ending the extraction process. Before the material exits the extractor, it is allowed to gravity drain to reduce its solvent retention. The extracted, spent material then falls into the extractor discharge and exits the apparatus. The miscella with the highest concentration of vegetable oil also exits the apparatus to a full miscella tank.

To understand the extraction process on a macroscale, it is helpful to understand the extraction process on a microscale. Figure 9a indicates the microstructure of a soybean cotyledon parenchyma (meats fraction) cell. This transmission electron micrograph is at 9000:1 magnification and represents a 0.020-mm tall by 0.023-mm wide cross-sectional view of a soybean flake, the approximate size of a single cell. The cell wall (CW), protein storage vacuoles (PSV), oil bodies (OB), nucleus (N), nucleolus (Nu), and intercellular spaces (*) are all indicated on the micrograph. As clearly seen, the oil within the cell exists as thousands of spherical oil bodies clinging to the inside surface of the cell walls and to the exterior surface of the protein storage vacuoles.

Figure 9. (a) Soybean cellular structure. Courtesy of USDA-ARS. Special thanks to Dr. Robert Yaklich and Dr. Charles Murphy at the Soybean Genomics and Improvement Laboratory in Beltsville, MD, for creating this electron transmission micrograph especially for this chapter. (b) Crown extractor.Courtesy of Crown Iron Works. (c) Reflex extractor. Courtesy of De Smet Group. (d) LM extractor. Courtesy of De Smet Group.
Figure 9  (Continued)
In the solvent extraction process, the miscella at the surface of the oleaginous material diffuses through the cell walls to the oil bodies located within the cells. The miscella quickly goes into solution with the oil bodies. As miscella continues to enter and go into solution, internal pressure builds within the cell and concentrated miscella diffuses back out of the cell. This concentrated miscella diffuses through the adjacent cell walls and eventually reaches the surface of the particle. Once the more concentrated miscella reaches the bath of miscella outside the oleaginous material, it quickly goes into solution with the miscella, incrementally increasing its concentration. This process continues until the concentration of the miscella inside the cells of the oleaginous material comes into equilibrium with the concentration of miscella outside of the oleaginous material.

All oleaginous materials have a somewhat different cell structure and, therefore, a different time required for the miscella in the cells of the oleaginous material to come into equilibrium with the miscella outside of the oleaginous material. Soybean flakes have a cellular structure (16) that will allow equilibrium to occur within approximately 5 minutes at each extractor wash, whereas sunflower cake requires approximately 9 minutes and rapeseed cake requires approximately 12 minutes.

There are six parameters that affect the performance of the solvent extractor apparatus. These six parameters are outlined below.

4.2.1. Contact Time  The total time that the oleaginous material spends in the extractor is residence time. Residence time can be subdivided into wash time and drain time. Wash time is the time the oleaginous material spends under the washing nozzles of the extractor, and drain time is the time the oleaginous material spends draining prior to discharge.

Wash time can be further subdivided into contact time and dormant time. Contact time is the time a particle of oleaginous material spends in the washing zone of the extractor where the particle is in contact with miscella. Extraction only takes place during contact time. Dormant time is the time a particle of oleaginous material spends in the washing zone of the extractor where it is not
in contact with miscella. The ratio of contact time to dormant time varies with extractor design.

Extractors with deep material bed depth and small bed surface area are generally operated by immersing the oleaginous materials in miscella. This is accomplished by providing a sufficient miscella flow rate per unit of material bed surface area to ensure that miscella fills all of the voids around the oleaginous material particles as it passes down through the material bed. This type of extractor operation provides a very high ratio of contact time to dormant time in the washing zone.

Extractors with shallow bed depth and large bed surface area are generally operated by percolating the miscella down through the oleaginous materials. With most oleaginous materials, the oleaginous material particles occupy 40% to 50% of the material bed and 50% to 60% of the material bed is composed of void spaces between the particles. In percolation, the oleaginous material particles are surrounded by both solvent vapor and miscella as the miscella rains downward. This type of extractor operation provides a lower ratio of contact time to dormant time in the washing zone.

In comparing two different extractor designs, the contact time can be similar, despite very different configurations:

**Deep Bed Extractor Example**

50 min. residence time = 30 min. wash time + 20 min. drain time.
30 min. wash time – 5 min. dormant time = 25 min. contact time.

**Shallow Bed Extractor Example**

50 min. residence time = 45 min. wash time + 5 min. drain time.
45 min. wash time – 20 min. dormant time = 25 min. contact time.

Regardless of extractor design, adequate contact time is critical for maximizing extraction efficiency and minimizing the amount of residual oil remaining in the oleaginous material. Increased contact time requires a larger extractor. Even though this represents a higher initial investment, the long-term benefits of surplus contact time are significant and justify oversizing the extractor.

4.2.2. **Particle Thickness** Various oleaginous materials are prepared for extraction using different process steps, but with virtually all oleaginous materials, one process step is flaking. The principle purpose of flaking is to reduce the thickness of the oleaginous material to reduce the distance and number of cell walls that miscella needs to diffuse through to reach the oil bodies. Note that for soybeans, a typical 0.38-mm-thick flake is approximately 20 cells thick. By reducing the particle thickness, the time required for miscella within the cellular structure of the oleaginous material to reach equilibrium with the miscella surrounding the oleaginous material is reduced. By reducing particle thickness, desired results
can be achieved with less contact time. If all other extraction parameters remain constant, reduced particle thickness will allow a smaller extractor to be used.

Reducing the particle thickness represents additional cost. For example, on soybeans, reducing particle thickness from 0.38 mm down to 0.30 mm will increase flaking mill electricity requirements by 1 to 2 kwh/ton of soybeans processed. As this is a significant ongoing operating expense, it is not economically feasible to undersize an extractor and reduce particle thickness. Conversely, by increasing particle thickness, desired results will demand more contact time. An extractor can be oversized to obtain desired results with increased particle thickness, thereby reducing ongoing operating cost.

For all oleaginous materials, the economic balance between the initial cost of the extractor and the ongoing electricity costs required for flaking can be analyzed and the optimum particle thickness can be determined.

**4.2.3. Extractor Temperature**  As the temperature of the miscella increases, its rate of diffusivity through the cell walls of the oleaginous material increases. As the prepared oleaginous material enters the extractor at approximately 60°C, and both the oil and meal fractions are heated in excess of 100°C in subsequent process steps, there is no extra energy required for operating the extractor at a warm temperature. As a result, optimizing extraction results requires operating the extractor as warm as possible.

There is an upper limit for the extractor operating temperature. The solvent must remain safely in a liquid state. As the boiling range of commercial hexane is typically 64–69°C at sea level, the maximum possible temperature to prevent boiling is 63°C. Operating on the edge of the boiling range could cause rapid evaporation during an upset condition. Rapid evaporation can cause pressurization of the extractor, leading to excessive solvent loss, a safety hazard. Therefore, most processors operate the extractor at 60°C to provide several degrees of safety margin below the lower end of the solvent boiling range.

If the prepared material temperature is too low, or heat loss in the extractor is too high, then it may not be possible to achieve an extractor temperature of 60°C. Additional contact time will be required to achieve desired results. Insulating the conveying system and extractor to prevent heat loss to enable operation at 60°C is typically less expensive than oversizing the extractor to compensate for low operating temperature.

**4.2.4. Miscella Flux Rate**  The miscella flux rate is the maximum volumetric flow rate of miscella that can flow down through the bed of material per unit of material bed surface area. In SI units, it is commonly expressed as m³/h per m², and in Imperial units, it is commonly expressed as gpm per ft². By simplification of units, the miscella flux rate can also be expressed as the maximum downward velocity of the miscella through the material bed (in m/min or ft/min). Miscella flux rates for various prepared oleaginous materials vary widely (see Table 1).

The miscella flux rate is determined by the screen below the bed of material. As stated earlier, the material bed is approximately 40% to 50% solids and 50% to 60%
void space. Therefore, as the miscella is moving downward, it has 50–60% open area. The screen under the material bed has less open area, and therefore, the material interface with the screen creates the greatest restriction to flow. Most deep bed extractor screens have approximately 30% open area, whereas most shallow bed extractor screens have less than 10% open area. In both cases, the screens have less open area than the material bed, and the material interface with the screen provides the greatest flow restriction.

As downward miscella flow reaches the miscella flux rate, the material/screen interface reaches its maximum flow rate and begins restricting the flow of miscella. All void spaces between the oleaginous material particles fill with miscella as the solvent vapors are pushed out of the top of the material bed. Eventually, the entire material bed becomes immersed in miscella with no void spaces. At this point, miscella breaks through the top of the miscella bed and forms a pool. This phenomenon is often referred to as “flooding” the material bed. Once the material bed is flooded, no additional rate of flow will pass down through the material bed.

At each washing stage of the extractor, miscella needs to have an opportunity to wash the material bed, pour through the screen, and then enter the proper miscella collection receptacle underneath the material bed. For a given extractor and prepared oleaginous material, each miscella collection receptacle is carefully calculated to be located a specific distance after its washing nozzle to maintain separation between washing stages.

For example, if there is a deep bed extractor operating on 0.38-mm-thick soybean flakes with a 3.0-m bed depth and a forward velocity of 0.3 m/min, the distance that the miscella collection receptacle needs to follow the washing nozzle can be calculated as follows:

Downward miscella velocity for 0.38-mm-thick soybean flakes = 0.42 m/min.
Time for miscella to pass through material bed = 3.0 m/0.42 m/min = 7.1 min.
Wash nozzle to miscella receptacle distance = 7.1 min* 0.3 m/min = 2.1 m.

As another example, if there is a shallow bed extractor operating on 0.38-mm-thick soybean flakes with a 1.0-m bed depth and a forward velocity of 1.0 m/min,
the distance that the miscella collection receptacle needs to follow the washing nozzle can be calculated as follows:

Downward miscella velocity for 0.38-mm-thick soybean flakes = 0.42 m/min.
Time for miscella to pass through material bed = 1.0 m/0.42 m/min = 2.4 min.
Wash nozzle to miscella receptacle distance = 2.4 min * 1.0 m/min = 2.4 m.

If the miscella flux rate is significantly reduced for a given prepared oleaginous material, the miscella can partially discharge into a later, undesired miscella collection receptacle. This will cause concentration contamination caused by mixing of the extractor washes and reduce the efficiency of the extractor.

Miscella flux rates can reduce as a result of thinner than normal flakes, surface moisture, or an abundance of fine particles. Flake thickness is a normal operator controlled parameter, and surface moisture and an abundance of fine particles are more difficult to control.

As hexane solvent is not soluble in water, the liquids repel each other. If surface moisture exists on the material in the extractor, the solvent has difficulty penetrating the particle surface. Also, the moisture can collect on the screen at the bottom of the material bed, building a protein layer and narrowing the screen slots, thus further reducing the miscella flux rate. To prevent this from occurring, it is very important to adequately aspirate flakes from flaking mills, cake from screw presses, or extruded pellets from extruders to remove all water vapor created when the material evaporatively cools from preparation temperature to extraction temperature. As a secondary precaution, it is ideal to have an extractor designed such that the oleaginous material moves with respect to the screen surface to constantly keep the screen surface brushed clean so that the impact of surface moisture on miscella flux rate is minimized.

The material bed is approximately 40–50% material particles and 50–60% void space. If there is an abundance of fine material particles, these particles can sift down through the material bed and settle in the void spaces just above the screen. This causes an additional flow restriction, and the miscella flux rate will be reduced. An abundance of fine material coming to the extractor is generally caused by over-drying the material at some point in the seed preparation stage, or by rough handling of the friable material during a seed preparation step. Both should be avoided to ensure uniform material shape to the extractor and uniform miscella flux rates.

4.2.5. Number of Miscella Stages  In most extraction applications, the prepared material has approximately 20% oil by weight and the goal is to reduce the oil content to approximately 0.5% oil by weight. If an extractor had one miscella stage, then the miscella concentration exiting the extractor (1.15% oil) would be in equilibrium with the miscella concentration remaining in the material cells (1.15% oil). A mass balance for a single-stage extractor can be calculated (see Table 2). An extractor that has only one miscella stage would require 17.2 parts of solvent per 1.0 part of material to be extracted. The energy required to evaporate the solvent in
the miscella would be tremendous. As a result, countercurrent, multistage extractors are required.

Through iteration of the mass balance, the minimum number of miscella stages can be calculated for a given solvent to material ratio. For an energy-competitive distillation system, the solvent to material ratio should be maintained below 1 to 1. To achieve a ratio below 1 to 1, the minimum number of stages required can be calculated and determined to be four stages.

A four-stage extractor with sufficient contact time to allow the miscella concentration in the material cells to come into equilibrium with the miscella concentration in the surrounding miscella bath at each miscella stage can be designed to extract to less than 0.5% residual oil using a solvent to material ratio of 1 to 1. The resultant outgoing miscella concentration will be in the range of 27% oil. A four-stage extractor is sufficient in a theoretical sense, but it leaves no contingency for lack of achieving equilibrium at each miscella stage.

The more miscella stages, the greater the theoretical extraction efficiency will be. In practice, however, if too many miscella stages are designed into an extractor, causing the individual stages to have insufficient contact time to reach equilibrium, residual oil will not be further reduced by adding stages. In this case, more miscella stages simply lead to more pumps, and therefore more pumping energy and more potential points for solvent loss. The number of miscella stages is generally determined by the total washing zone time and the number of stages that can theoretically come to equilibrium within the washing zone time. Commercially, most extractors have in the range of five to nine miscella stages.

### 4.2.6. Solvent Retention

After the washing zone of the extractor, the extracted material is left to gravity drain. This gravity drain time is generally in the range of 5 to 20 minutes. Extractors designed with shallow material bed depths will generally have a drain time closer to 5 minutes, and extractors designed with deep material

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<tbody>
<tr>
<td>Prepared Material Entering</td>
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<tr>
<td>Solids: 800 units</td>
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<tr>
<td>Oil (20%): 200 units</td>
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<tr>
<td>Total: 1000 units</td>
</tr>
<tr>
<td>Solids: 800 units</td>
</tr>
<tr>
<td>Oil (0.5% residual): 4 units</td>
</tr>
<tr>
<td>Solvent (30% retention): 344 units</td>
</tr>
<tr>
<td>Total: 1148 units</td>
</tr>
<tr>
<td>Miscella Concentration</td>
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<tr>
<td>$\frac{4}{4 + 344} = 1.15%$ oil</td>
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bed depths generally have a drain time closer to 20 minutes. After gravity drainage, the solvent retained with the extracted material will be in the range of 25% (fast-draining extrudate) to 35% (slow-draining flakes).

The “solvent retention” of the drained material is a bit of a misnomer, and it would be more accurately defined as the “weak miscella retention.” The weak miscella retained in the drained material contains approximately 1% oil. In the meal desolventizer, the solvent is evaporated leaving behind the traces of oil, often referred to as the residual oil. In order to minimize the residual oil left in meal, it is important to minimize the amount of weak miscella carried forward to the meal desolventizer.

Adequate extractor drain time is the most economic manner in which to minimize weak miscella retention. Maintaining the desired miscella flux rate is also important. Thus, once again, there is a need to maintain proper flake thickness, minimize surface moisture, and minimize fine material particles in the material bed to minimize weak miscella retention.

Today there are two major suppliers of solvent extractors, Crown Iron Works, headquartered in the United States, and the De Smet Group, headquartered in Belgium. The Crown Model III Extractor is a shallow material bed depth extractor, using a chain conveyor design to convey the material over fixed screens in a loop pattern (see Figure 9b). De Smet supplies two major types of extractors, a Reflex Extractor and an LM Extractor. The De Smet Reflex Extractor is a deep material bed depth extractor, using a cylindrical rotating set of baskets to convey the material over a fixed circular screen (see Figure 9c). The De Smet LM Extractor is provided in both shallow and deep material bed depths, using a belt conveyor made of screens to convey the material along a linear path (see Figure 9d). Extractors of these three types process in the range of 25 tons per day of specialty oilseeds to 10,500 tons per day of soybeans through a single unit.

4.3. Meal Desolventizer Toaster

After the prepared material has had its oil extracted in the solvent extractor, it is conveyed to the meal desolventizer toaster, commonly referred to as the DT. The material entering the DT is typically at the extractor temperature of 60°C, and it contains 25–35%(w/w) of solvent. The primary purpose of the DT is to remove the solvent from the meal fraction so that the solvent can be recovered.

DTs are vertical, cylindrical vessels with a multitude of horizontal trays. The extracted material enters at the top and is supported by the tray. The material is mixed above each tray, and it is conveyed downward from tray to tray, by agitating sweeps anchored to a central rotating shaft. The heat for increasing meal temperature and evaporating the solvent is supplied by steam, introduced directly and indirectly into the meal via the trays. Figure 10a illustrates a typical Schumacher DT.

The trays of the DT are designed with an upper plate, lower plate, and structural members between designed-to-hold pressurized steam. The DT has three different types of trays: predesolventizing trays, countercurrent trays, and a sparge tray.
4.3.1. Predesolventizing Trays

The predesolventizing trays have the sole purpose of providing conductive heat transfer through their upper surface to the solvent-laden material supported above. The steam is typically held at 10.5-kg/cm² pressure within the predesolventizing trays, providing a surface temperature of 185°C. Steam condenses within the tray, providing its latent heat to maintain the 185°C tray surface temperature and allow heat to be conducted into the solvent-laden meal layer above. Heat is transferred into the meal at a rate of 500–600 kJ/m²-C-h with a temperature differential of approximately 120°C.

A DT may have as many as seven predesolventizing trays, or as few as one. The predesolventizing trays are located in the upper portion of the DT and must allow ascending vapors from below to pass around them to the vapor exit at the top of the DT. Some manufacturers design disk-shaped trays providing space for the ascending vapors to pass between the outside perimeter of the tray and shell wall, and others design donut-shaped trays providing space for ascending vapors to pass.
between the inside perimeter of the tray and the central shaft. As an alternative to a large number of predesolventizing trays, the upper section of the DT is often expanded in diameter to enable fewer, larger diameter trays.

4.3.2. Countercurrent Trays The countercurrent trays have three purposes, providing conductive heat transfer through their upper surface to heat wet material supported above, providing convective heat transfer through their lower surface to superheat direct steam swirling below, and providing apertures for direct steam to evenly ascend up through the tray and into the meal supported above. The steam is typically held at 10.5-kg/cm² pressure within the countercurrent trays, providing a surface temperature of 185°C. Steam condenses within the tray, providing its latent heat to maintain the 185°C tray surface temperature and allow heat to be conducted into the meal layer above, and transferred into the direct steam swirling below. Heat is transferred into the meal above at a rate of 450–550 kJ/m²-C-h with a temperature differential of approximately 85°C. Heat is also transferred into the direct steam swirling below at a rate of approximately 100 kJ/m²-C-h with a temperature differential of approximately 75°C.

A DT will have from one to four countercurrent trays. The countercurrent trays are located directly under the predesolventizing trays in the center of the DT. The apertures in the trays must allow the direct steam swirling below to pass through the
tray and into the meal supported above. There are three designs of countercurrent trays with differing apertures for allowing the ascending direct steam to pass through. Initial modern DTs (Schumacher, circa 1982) used hollow stay-bolts as apertures and generally had 1% to 2% open area for the vapors to ascend upward. Later designs (Mason, circa 1985) used hollow stay-pipes capped by a perforated plate, with generally 2% to 4% open area for vapors to ascend upward. The latest design (Kemper, circa 1997) uses pie-shaped enclosures capped with stainless steel slotted screens, with approximately 10% open area for the vapors to ascend upward. The choice of countercurrent tray design is generally determined by the amount of open area required to maintain stable operation for the quantity of direct steam passing through.

4.3.3. Sparge Tray The sparge tray has the dual purpose of providing a uniform means of introducing direct steam into the meal layer, and providing conductive heat transfer through its upper surface to the wet material supported above. The direct steam introduced through the sparge tray provides approximately 75% of the total heat required for desolventizing and heating the meal in the DT. The sparge tray is typically designed with a plurality of apertures across its entire upper surface to evenly introduce direct steam into the meal. The size and quantity of apertures is calculated based on the anticipated direct steam flow rate to provide a pressure drop of 0.35 to 0.70 kg/cm². The direct steam supply is 10.5-kg/cm² pressure saturated steam (185°C), and after passing through a flow control valve, its quality changes to 0.35–0.70-kg/cm² pressure superheated steam (150–160°C). Therefore, the upper surface of the sparge tray is maintained at approximately 155°C. Heat is transferred into the wet meal above at a rate of 450–550 kJ/m²-C-h with a temperature differential of approximately 50°C. The sparge tray is the last tray in a modern DT.

Solvent-laden meal enters the DT at a temperature of 60°C, and it contains 25% to 35% (w/w) of solvent. The solvent-laden meal is stirred across the surface of the predesolventizing trays by the rotating sweeps. As the heat is transferred into the meal layer by conduction, a shallow layer of 150-to-300-mm meal depth is held above each tray. The solvent-laden meal temperature is increased to approximately 68°C and approximately 10% to 25% of the solvent is evaporated on the predesolventizing trays.

The material exits the predesolventizing trays of the DT and falls onto the top countercurrent tray. This is perhaps the most critical tray of the DT. As most of the heat is transferred into the meal layer by condensation of direct steam, a deep layer of 700-to 1000-mm meal depth is held above the tray. The solvent-laden meal is stirred above the top countercurrent tray by the rotating sweeps. The direct steam passes from below up through apertures in the countercurrent tray. As the direct steam penetrates the upper meal layer, it reaches the solvent-laden meal and condenses, providing direct latent heat to evaporate solvent. The solvent evaporates and exits the meal layer as vapor. The condensation of steam causes the meal exiting the tray to be wet, typically in the range of 17–21% moisture. After the majority of the solvent evaporates, the meal temperature increases by direct and indirect steam
heat, surpassing 100°C before the material exits the tray. The protein solubility of soybean meal is reduced from approximately 90 PDI down to 45 PDI as a result of the elevated moisture and temperature conditions.

After the wet meal exits the top countercurrent tray, it has had over 99% of its solvent removed. On the remaining countercurrent trays and the sparge tray, the meal is held in a 700-to 1000-mm deep layer on each tray to provide residence time for stripping solvent and toasting. The wet meal is stirred above each tray by rotating sweeps. The final desolventizing takes place as the ascending steam passing through the meal slowly strips out final traces of residual solvent down to 0.01–0.05% (w/w). The meal temperature increases from 100°C up to 105–110°C, and the meal moisture decreases approximately 1% before the meal discharges from the sparge tray. The meal color darkens slightly and provides the meal with a toasted color. For soybeans, antinutritional factors such as trypsin inhibitors and urease are reduced on these trays by maintaining the meal moisture and temperature elevated for a period of time. The protein solubility drops approximately 1% PDI for every minute the meal spends in the remaining countercurrent trays and the sparge tray. Ideal feed for monogastric animals (poultry and swine) is high in protein solubility, and ideal feed for ruminant animals (cattle) is low in protein solubility (high in rumen bypass protein). Meal residence time on the remaining countercurrent trays and the sparge tray is dictated by both the degree of solvent recovery required as well as by meal quality parameters.

An important parameter in the energy efficiency of the DT is the exit vapor temperature. The condensing sparge steam provides a plentiful supply of surface moisture, allowing the solvent and water to evaporate as an azeotropic mixture. Figure 10b indicates the solvent/water equilibrium boiling curve. As the chart indicates, a mixture of 94% solvent with 6% water can boil as low as 62°C. Therefore, the lowest possible DT exit vapor temperature would be 62°C. In practice, to maintain low solvent loss and maintain a safety margin, modern DTs are operated with exit vapor temperatures ranging from 66–78°C, with the most typical temperature being 71°C. As the DT vapor temperature increases, the ratio of water vapor to solvent vapor increases. Therefore, to minimize total DT energy, it is very important to maintain a vapor temperature as low as safely possible.

Determining the optimum DT configuration for a given process application is complex. It requires determining all input parameters and calculating the mass and heat balance of both the DT and the follow-on meal dryer cooler (DC). The mass and heat balance of the DC will determine the maximum allowable DT exit moisture that will minimize meal drying energy. This moisture is generally in the range of 18–20%. With the DT exit moisture determined, the amount of direct steam introduced into the meal can be calculated. The DT diameter is generally determined by the direct steam flow rate per unit area. It is important to have a sufficiently high direct steam flow rate per unit area for adequate solvent stripping. The number of countercurrent trays is determined by the residence time needed to balance meal quality with residual solvent objectives. By calculating the total DT heat demand and subtracting the heat supplied by live steam, the total heat supplied by indirect steam can be determined. The total heat supplied by indirect steam less
the heat supplied by countercurrent tray indirect steam will provide the amount of indirect steam heat needed to be supplied by the predesolventizing trays. With this data in hand, the diameter and quantity of predesolventizing trays can be selected. Major manufacturers of DTs use process simulation tools to assist processors in optimizing the DT configuration for a given application.

4.4. Meal Dryer Cooler

After the solvent-laden material is desolventized, it is conveyed to the meal DC. The material entering the DC is typically at the DT exit temperature of 108°C and contains 18–20% moisture. The primary purposes of the DC are to reduce the moisture in the meal to within trading rule limits and to lower the meal temperature prior to storage.

DCs are vertical, cylindrical vessels with a multitude of horizontal trays. The desolventized material enters at the top and is supported by the tray. The material is mixed above each tray and conveyed downward from tray to tray by agitating sweeps anchored to a central rotating shaft. The DC has three different types of trays: steam drying trays, air drying trays, and air cooling trays.

4.4.1. Steam Drying Trays

The steam drying trays of the DC are designed with an upper plate, lower plate, and structural members between designed-to-hold pressurized steam. The steam-drying trays have the purpose of providing conductive heat transfer through their upper surface to wet meal supported above. The steam is typically held at 10.5-kg/cm² pressure within the steam-drying trays, providing a surface temperature of 185°C. Steam condenses within the tray, providing its latent heat to maintain the 185°C tray surface temperature and to allow heat to be conducted into the wet meal layer above. Heat is transferred into the meal at a rate of 450–550 kJ/m²-C-h with a temperature differential of approximately 77°C.

A DC may have as many as five steam-drying trays or as few as none. The water vapor evaporated from the meal can be compressed in an ejector and have its heat recovered within the solvent extraction plant.

4.4.2. Air Drying Trays

The air-drying trays of the DC are designed with an upper plate, lower plate, and structural members between designed-to-hold low-pressure air. The air-drying trays are designed with a plurality of apertures across their entire upper surface to evenly introduce air into the meal. The size and quantity of apertures is calculated based on the design air flow rate to provide a pressure drop of 0.02 to 0.03 kg/cm².

The air supplied to each air dryer tray is first filtered to remove dust and then pressurized using a spark-proof centrifugal blower. The air for the dryer trays is passed through a steam-heated coil between the blower and the entrance to the dryer trays. After the air enters the trays, it flows upward through the meal at a nominal velocity of 14–21 m/minute, partially fluidizing the meal. The meal evaporatively cools, and the released moisture is transferred to the ascending air. The warm, damp air exits the top of the meal layer and then exits the sidewall of the DC to a
cyclone collector to remove dust prior to discharge to atmosphere. Figure 11 illustrates a meal DC with two air trays and related air-handling equipment.

The major source of heat for evaporating the moisture in the meal is the high temperature of the meal exiting the DT or the DC steam-drying trays. When the meal drops in temperature from 108°C to 38°C, the heat provided is adequate to reduce the meal moisture by 6.5%. For soybean meal, the trading rule moisture limit is 12.5%; therefore, if the incoming moisture from the DT, or DC steam-drying trays, does not exceed 19.0%, the DC will typically require no additional evaporative heat source to dry the meal. If additional heat is required for evaporating moisture from the meal, the air entering the meal dryer trays can be heated to temperatures up to 150°C prior to entering the air drying trays. The heat source can be recovered flash steam, hot glycol-water solution (oil cooler), or fresh steam.

The air must have adequate capacity to carry out the moisture released from the meal without becoming saturated. Cold air can hold less moisture than warm air, so winter conditions may limit the moisture-carrying capacity of the air. If additional heat is required for increasing the dew point of the incoming air, the air entering the air dryer trays can be heated to temperatures up to 150°C.

The energy required to heat the incoming drying air is largely dictated by the incoming meal moisture to the DC. Adequate predesolventizing heat transfer area in the DT or steam-drying tray heat transfer area in the DC is the key to minimizing meal moisture to the DC air drying trays and therefore minimizing DC heater coil steam consumption.

4.4.3. **Air cooling trays** The air cooling trays of the DC are designed with an upper plate, lower plate, and structural members between designed-to-hold low-pressure air. The air cooling trays are designed with a plurality of apertures across

![Figure 11. Meal dryer cooler. Courtesy of De Smet Group.](image)
their entire upper surface to evenly introduce air into the meal. The size and quantity of apertures is calculated based on the design air flow rate to provide a pressure drop of 0.02–0.03 kg/cm².

The air supplied to each air cooler tray is first filtered to remove dust and then pressurized using a spark-proof centrifugal blower. After the cool air enters the trays, it flows upward through the meal at a nominal velocity of 14 to 21 m/minute, partially fluidizing the meal. The meal continues to evaporatively cool and convectively cools. The cool, damp air exits the top of the meal layer and then exits the sidewall of the DC to a cyclone collector to remove dust prior to discharge to atmosphere.

Ambient air is heated approximately 5°C in temperature by the energy of the blower, before it blows into the air cooler trays of the DC. The meal cools down to within approximately 5°C of the air temperature passing through the meal. Therefore, the temperature of the meal exiting the DC is typically cooled down to within 10°C of ambient air.

The dry cool meal is conveyed from the DC to outside the solvent extraction plant for size reduction and then on to meal storage. It is important to properly dry and cool the meal to prevent continued evaporative cooling in storage or transport, which will cause reduced flowability, solidification, and bridging of the meal inside storage and transport vessels.

4.5. Miscella Distillation System

The miscella exiting the extractor contains particles of meal, up to 1% (w/w). Therefore, the first step in miscella distillation is meal particle separation. Meal particles can be separated from the miscella via filtration or centrifugal separation. The large meal particles (+80 mesh) need to be removed to prevent them from settling out in the distillation equipment. The fine meal particles (–80 mesh) are removed from the oil after the solvent extraction plant, typically in the oil degumming process, or oil-refining process. For those plants producing food grade lecithin, the fine meal particles must be completely filtered out prior to degumming.

Some extractor designs include an internal miscella filter (100 mesh) enabling the miscella exiting the extractor to be sufficiently free of meal particles such that it can go directly to distillation. Other extractor designs require external meal particle separation, accomplished by pumping the miscella through a liquid cyclone. The liquid cyclone spins the miscella at high velocity and uses centrifugal force to separate the larger meal particles (over 80 mesh) from the miscella. The larger meal particles, along with 5–10% of the miscella flow, exit the liquid cyclone underflow orifice and return to the extractor. Separation of meal particles by external filtration of the miscella is not recommended because of the safety hazards associated with opening the filter to remove the solvent-laden meal fines.

Once the large meal particles are separated, the clean miscella is stored in a surge tank, generally referred to as a full miscella tank. The full miscella tank has several purposes: It separates the continuous extraction process from the continuous distillation process, it provides miscella storage capacity during a power
outage, and it provides surge capacity so that fluctuations in miscella flow rate from the extractor can be absorbed prior to distillation. The miscella in the full miscella tank is generally 25–30% oil and 70–75% solvent by weight, and at the typical extractor temperature of 60°C.

The miscella is pumped from the full miscella tank to the first of two rising film evaporators. The first-stage evaporator, often referred to as the economizer, uses the waste heat from the DT as its heating source. Miscella enters through tubes at the base of the evaporator, at a temperature of approximately 60°C. As the tube side of the evaporator is held at approximately 300–400-mm Hg absolute pressure, the miscella temperature will drop to approximately 43–48°C temperature upon entry into the tubes. Solvent will begin evaporating and solvent vapor bubbles will rise up through the center of the tubes. Additional DT vapor heat is transferred through the tubes into the miscella, and additional evaporation takes place. When sufficient solvent vapor is formed, the vapor velocity through the center of the tubes will become sufficiently high to drag a thin film of miscella up the inner walls of the tubes, creating high heat transfer rates. At the top of the tubes, the high-velocity solvent vapor and remaining miscella contact an impingement plate to break foam, and they are then centrifugally separated in the evaporator dome. Solvent vapors exit the top of the dome, and concentrated miscella exits the base of the dome. The concentrated miscella exiting the first-stage evaporator is generally 75–85% oil and 15–25% solvent, at a temperature of approximately 48°C.

As the temperature of the miscella exiting the first stage evaporator is low, it is a good heat sink for heat recovery. In various facilities, heat from hot finished oil, heat from steam ejector exhausts, or recovered flash steam is used to preheat the miscella to approximately 75°C in temperature. The preheated, concentrated miscella is then typically heated to 110°C in a steam-heated exchanger prior to entering the second rising film evaporator.

The preheated, concentrated miscella is pumped into the second-stage evaporator, which uses low-pressure steam as its heating source. As the tube side of the evaporator is held at approximately 300–400-mm Hg absolute pressure, the solvent temperature will drop to approximately 43–48°C temperature upon entry into the tubes, providing sufficient heat to immediately begin vigorous evaporation. The vapor velocity through the center of the tubes will be sufficiently high to drag a thin film of miscella up the inner walls of the tubes at relatively high velocity. This is very important to prevent baking phosphatides and fine solid particles in the miscella to the lower, inner surface of the tubes. Additional low pressure steam heat is transferred through the tubes into the miscella and additional evaporation and heating takes place. At the top of the tubes, the solvent vapor and remaining miscella contact an impingement plate to break foam, and they are then centrifugally separated in the evaporator dome. As velocities exiting the tubes are insufficient to break all foam, the dome must be sufficiently large in diameter to allow remaining foam to collapse and not discharge with exiting vapors. Solvent vapors exit the top of the dome, and concentrated miscella exits the base of the dome. The concentrated miscella exiting the second-stage evaporator is generally 95–98% oil and 2–5% solvent, at a temperature of approximately 110°C.
Miscella from the second-stage evaporator is pumped or gravity fed into the oil stripper. The oil stripper is a tall, thin cylindrical vessel that is commonly operated at 150–300-mm Hg absolute pressure. The hot oil passes downward through the vessel across trays. Simple, robust disk-donut trays, or grid-bar trays, are typically used because fouling caused by baking of gums and fine particles onto stripper trays is common. Live steam is introduced into the oil at the top of the vessel to initiate evaporation and again at the base of the vessel to provide countercurrent stripping. The steam and solvent vapors exit the top of the oil stripper through an enlarged diameter dome to prevent entraining oil mist. The oil typically exits the base of the oil stripper with 0.1–0.3% moisture and 0.005–0.020% solvent, at a temperature of approximately 110°C.

The oil leaving the oil stripper, particularly in soybean plants, is often water degummed. In these facilities, the oil temperature is reduced to 70–80°C, and 1–2% soft water is injected and mixed into the oil inline. The oil is then held for approximately 30 to 60 minutes in an agitated tank to allow gums to hydrate. The gums (water, phosphatides, fine meal particles, and some neutral oil) are centrifugally separated from the oil using a high-speed centrifugal separator. The gums are often pumped back into the DT and mixed into the meal fraction. Alternatively, the gums may be dried for food grade lecithin, or feed grade lecithin for animal feed applications. The degummed oil, at a moisture level of approximately 0.5% and a temperature of approximately 70°C, is then heated in a heat exchanger to approximately 110°C.

Whether or not the oil is degummed, it is typically pumped to an oil dryer. The oil dryer is a vertical cylindrical vessel that is commonly operated at 50–80-mm Hg absolute pressure. The hot oil is sprayed downward into the vessel, with or without trays. The solvent and water vapors exit the top of the oil dryer and the oil exits the bottom. The oil typically exits the base of the oil dryer with 0.05–0.10% moisture and 0.002–0.010% solvent, at a temperature of approximately 105°C.

The dried oil must have its temperature reduced from 105°C to 50°C to prevent degradation in storage and transport. The hot oil is commonly cooled in two stages. First, the oil is cooled from 105°C to approximately 70°C temperature in a heat exchanger using concentrated miscella or a water-glycol solution (for preheating DC air) as the cooling medium. Second, the oil is cooled from 70°C to 50°C temperature in a heat exchanger using cooling water as the cooling medium. The cool oil is then pumped to storage.

4.6. Solvent Recovery System

Modern solvent extraction plants recover over 99.9% of the solvent pumped to the extractor. The solvent recovery system includes solvent and water vapor condensation, solvent-water separation, stripping solvent from water and air effluent streams, as well as heating the solvent prior to reuse in the extractor.

The solvent vapors from the first-and second-stage evaporators are typically condensed in a common medium-vacuum condenser. The medium-vacuum condenser is a shell and tube vessel with the vapors typically on the shell side and the cooling
water on the tube side. The noncondensable vapors are removed from the condenser by a steam ejector to maintain the 300–400-mm Hg absolute pressure on the shell side and are typically discharged into the first-stage evaporator for heat recovery.

The water and solvent vapors from the oil stripper and mineral oil stripper are typically condensed in a high vacuum condenser. The high vacuum condenser is a shell and tube vessel with the vapors typically on the shell side and the cooling water on the tube side. The noncondensable vapors are removed from the condenser by a steam ejector to maintain the 150–300-mm Hg absolute pressure on the shell side and are typically discharged into the first-stage evaporator for heat recovery.

The water and solvent vapors are evacuated from the oil dryer by a steam ejector to maintain the 50–80-mm Hg absolute pressure on the oil dryer and are typically discharged into the base of the oil stripper. These vapors along with the ejector motive steam serve as the oil stripper’s source of countercurrent stripping steam for heat recovery.

The solvent and water vapors from the DT typically pass through a vapor scrubber to remove meal particles. Two types of vapor scrubbers are commonly used: water-spray scrubbers and centrifugal scrubbers. Water-spray scrubbers use a heavy spray of hot water droplets being sprayed through the vapors in the duct exiting the DT to entrap meal particles. The meal particles and hot water are collected in a tank and recirculated, with a small stream of dirty water being returned to the DT. Centrifugal scrubbers use centrifugal force to separate meal particles from the vapors, allowing the particles to fall out of the bottom of the scrubber directly into the DT. Some centrifugal separators use a solvent wash to keep the walls of the centrifugal scrubber clean.

The clean vapors exiting the DT vapor scrubber, along with the much smaller vapor streams from the steam ejectors and the waste water stripper, are partially condensed in the shell side of the first-stage evaporator. The vapors enter the top of the evaporator shell side at approximately 70°C temperature and quickly condense water vapor until the temperature drops to 62°C, the minimum hexane/water equilibrium temperature. The vapors continue condensing at the mixture of 94% solvent and 6% steam and temperature of 62°C as they progress downward through the shell of the first-stage evaporator. The remaining solvent vapor and water vapor exiting the first stage evaporator shell (approximately 20% to 30% of what entered) are further condensed in either a vapor contactor or a solvent preheater. In a vapor contactor, cool condensate from condensers is pumped and sprayed into the top of a vertical tank. Remaining solvent and water vapors from the first-stage evaporator pass through the cool liquid stream and condense on the droplets, causing the liquid spray temperature to rise. In a solvent preheater, cool fresh solvent en route to the extractor is pumped through the tubes of a shell and tube heat exchanger and the solvent and water vapors from the first-stage evaporator pass through the shell side, condensing on the tubes while warming the fresh solvent passing through the tubes. Whether using a vapor contactor or a solvent preheater after the first-stage evaporator, the remaining DT vapors are reduced to less than 10% of what exited the DT, recovering 90% of the waste heat from the DT. The remaining vapors pass on to the vent condenser.
Approximately 1 m$^3$ of air enters the extractor with each 1 m$^3$ of material. To maintain the extractor under a slight vacuum, the air must be continuously vented from the extractor. At the extractor temperature of 60°C, an equilibrium condition will occur where the vent gas exiting the extractor will contain approximately 10 parts of solvent vapor for every one part of air (see Figure 12). The solvent vapor and air exit the extractor and are typically condensed in an extractor condenser. The extractor condenser is a shell and tube vessel with the vapors typically on the shell side and the cooling water on the tube side. The noncondensable vapors exiting the extractor condenser pass on to the vent condenser.

The vent vapor streams from the extractor condenser, vapor contactor or solvent preheater, and atmospheric tanks pass on to the vent condenser. The normal heat load on the vent condenser is low. However, in the case in which miscella is not passing through the first-stage evaporator, or solvent is not passing through the vapor contactor or solvent preheater, the load on the vent condenser can be very high. Therefore, the vent condenser is generally sized for these contingency conditions and has an excess of heat transfer area for normal operating conditions. The vent condenser is a shell and tube vessel with the vapors typically on the shell side and the cooling water on the tube side. The noncondensable vapors exiting the vent condenser pass on to the mineral oil absorption system.

**Figure 12. Equilibrium vapor composition.**
The composition of solvent vapor with the air exiting the vent condenser is a result of an equilibrium condition determined by the vent gas temperature. Figure 12 illustrates the equilibrium vapor compositions at various temperatures. For example, at a vent gas temperature of 30°C, the vent gas composition will be slightly greater than one part solvent for every one part air. However, if the vent gas temperature increases to 40°C, then the vent gas composition will be slightly less than two parts solvent for every one part air. Therefore, to minimize the load on the mineral oil system, it is important to minimize the vent gas temperature exiting the vent condenser. Some facilities, particularly in hot and humid climates, install a refrigerated vent gas cooler after the vent condenser to minimize the solvent vapor load on the mineral oil system.

The vent gas from the process enters the bottom of the mineral oil absorber. The mineral oil absorber is a tall, small-diameter packed column. Cold mineral oil cascades down through the column, absorbing solvent vapor from the vent gas as the vent gas rises up through the packing. When the vent gas exits the mineral absorber, it generally contains less than 10-g solvent per cubed meter of air. The vent gas is pulled from the mineral oil absorber via a spark-proof suction fan and is discharged to the atmosphere through a flame arrestor.

The mineral oil enters the mineral oil absorber at approximately 30°C temperature and contains 0.1–0.4% (w/w) moisture plus solvent. When the mineral oil exits the mineral oil absorber, its temperature rises slightly because of the heat of absorption, and the mineral oil contains 3–5% (w/w) moisture plus solvent. The cool solvent-rich mineral oil is then heated to 65°C by hot/cool mineral oil heat recovery and then further heated to over 100°C using a steam-heated heat exchanger. The hot, solvent-rich mineral oil enters the mineral oil stripper, a packed column vessel similar in construction to the mineral oil absorber. As the mineral oil stripper is typically maintained under 150–300-mm Hg absolute pressure, much of the solvent evaporates as soon as the mineral oil enters the top of the mineral oil stripper. Most of the remaining solvent is removed as the mineral oil cascades down across the packing countercurrent to rising stripping steam. The water and solvent vapor exit the dome of the mineral oil stripper. The hot, solvent-lean mineral oil exits the mineral oil stripper and has its temperature reduced to 65°C by hot/cool mineral oil heat recovery and is then further cooled to 30°C temperature using a heat exchanger with water as the cooling medium. The cool, solvent-lean mineral oil is then recirculated to the top of the mineral oil absorber.

All water and solvent that drain from the various condensers in the solvent extraction plant enter a decanting tank. As the solvent is immiscible with water, the lighter solvent (0.66 specific gravity) floats above the water. The key to the decanting tank performance is to minimize turbulence within the tank to allow sufficient time for the gravity decanting to take place. The elevation of the interface between solvent and water is established by the highest elevation of the water drain pipe. Water, typically containing 0.01% solvent, exits the decanting tank to the waste water stripper. Solvent, typically containing 0.05% water, exits the decanting tank to the solvent work tank.
The waste water stripper is a small tank used to increase the temperature of the waste water to approximately 95°C to evaporate residual solvent prior to discharge to the plant sump. The water exiting the waste water stripper is typically less than 0.001% (w/w) solvent. The hot water exiting the waste water stripper is often interchanged with the cool water entering the waste water stripper for heat recovery.

Solvent enters the work tank from the decanting tank. The work tank is a surge tank for holding solvent prior to the extractor to ensure that there is always ample solvent available to be pumped to the extractor. If a large surge of solvent flows into the work tank, the work tank will automatically overflow to solvent storage. If the level in the work tank becomes low, additional solvent is pumped from solvent storage to the work tank. The temperature of the solvent in the work tank is typically in the range of 52–57°C in plants equipped with a vapor contactor, and 43–49°C in plants without a vapor contactor.

In plants with a vapor contactor, solvent is pumped from the work tank at 55°C temperature to a steam-heated solvent heater to increase its temperature to 60°C prior to entering the extractor. In plants with a solvent preheater rather than a vapor contactor, solvent is pumped from the work tank at 45°C temperature to a solvent preheater where its temperature is increased to 55°C. The solvent is then further heated in a steam-heated solvent heater from 55°C to 60°C in temperature prior to returning to the extractor.

5. SUMMARY

The solvent extraction process consists of the unit operations of solvent extraction, meal desolventizing, meal drying and cooling, miscella distillation, and solvent recovery. These unit operations are highly interrelated, primarily because of various heat recovery methods that link the operations together. A process upset in any one of the unit operations will typically cause abnormal operation in the others. The key to efficient solvent extraction operation is process consistency. Consistency of seed preparation, consistency of rate, and consistency of solvent extraction operating parameters are all important factors in operating a safe, environmentally friendly, and cost-effective solvent extraction process for oil extraction.

REFERENCES


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