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Review of Desliming Methods and Equipment

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UNIT OF MEASURE ABBREVIATIONS USED IN THIS REPORT

°F	degree Fahrenheit	lb/in^2	pound per square inch
ft	foot	min	minute
ft^2	square foot	min/rev	minute per revolution
ft^3/h	cubic foot per hour	pct	percent
gal/h	gallon per hour	qt	quart
gal/min	gallon per minute	rev/min	revolution per minute
h	hour	ton/d	ton per day
in	inch	V/in^2	volt per square inch
kW	kilowatt	wt pct	weight percent
lb	pound	yr	year
$lb/h{\cdot}ft^{-2}$	pound per hour per square foot		

REVIEW OF DESLIMING METHODS AND EQUIPMENT

By Christopher H. Roe[1]

ABSTRACT

This Bureau of Mines report reviews the various methods of removing the liquid from tailings slurries that contain very fine grained solids. Gravitational settlement, centrifugation, filtration, and thermal drying methods are discussed in detail. Chemical additives, electrokinetic dewatering, and the rotary trommel screen are also mentioned. Lists of dewatering equipment and suppliers are given in the appendixes to assist the planner who must choose the most efficient, economic, and practical method of dewatering very fine grained mill tailings.

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INTRODUCTION

The disposal of wet fine-grained wastes from a milling operation can be a difficult challenge for the plant operator. He or she must consider the stability of the material after deposition, environmental consequences, and, above all, economic constraints. Waste material or tailings with a minimum of material smaller than 200 mesh (0.0029 in) can be piled up or used as backfill with few complications because it drains water easily and is relatively stable.

On the other hand, materials with a high percentage of particles smaller than 200 mesh present far more problems for water drainage and stability. The small size allows intermolecular attraction between water molecules and solid particles to influence the hydraulic and structural characteristics of the material. As a result, when these fine tailings are saturated, they have low permeabilities and have little or no shear strength (16).[2]

Current tailings disposal practice is to mix these fine tailings or slimes (35, p. 1026) with large volumes of water and to pipe the resulting slurry to settling ponds. In these ponds, the solid particles settle out of the slurry and the remaining liquid is decanted out of the pond. This disposal method works well; however, it requires large ponds. Also, as a consequence of the small particle size, these slimes may not become completely settled for years and may present a possibility of structural failure.

These problems can be mitigated or eliminated if the water content is substantially reduced prior to disposal. Many different methods have been developed for separating small-sized solids from liquids. A cursory evaluation of these methods was done by the Bureau of Mines. This report presents the results of the study in a format that should assist mill operators who must establish a tailings disposal system or upgrade an existing dewatering circuit for metal or nonmetal mining operations; dewatering coal slimes is reviewed in a report done under contract to the Bureau of mines (14). Many of the conclusions in this report are based on experience from the coal processing, power generation, and sewage treatment industries in the hope that a sharing of knowledge will be beneficial to those interested in dewatering slimes.

The descriptions of each dewatering method, along with the lists of manufacturers and equipment in appendixes B and C, should assist mill operators in making decisions about the most appropriate and economic methods of tailings dewatering to investigate further.

ACKNOWLEDGMENTS

The author thanks Ken Miyoshi, mill superintendent, Western Nuclear Corp., Wellpinit, WA, for reviewing this paper and offering considerable technical advice.

CHEMICAL TREATMENT

This paper discusses the physical methods of separating solids and liquids. It must be acknowledged, however, that any discussion of solid-liquid separation must include some comments on chemical additives used in this process. Other Bureau projects are investigating the various aspects of chemical additives, so only a brief review of these substances is included in this report.

Chemical treatment is often the first step in slurry concentration. Chemical additives (or reagents) are added to a

[2]Underlined numbers in parentheses refer to items in the list of references preceding the appendixes.

slurry to promote formation of more easily separated solid masses (24). For many of the dewatering devices evaluated in this report, chemical pretreatment is frequently used or may even be necessary to remove as much water from the solids as possible prior to mechanical dewatering.

For many years, glue, gums, starch, lime, and similar additives were used as flocculating aids to improve the separation rate of small solids from liquids (10). These substances were reasonably successful but increased the volume of the solids that had to be transported and disposed after dewatering (24).

With the introduction of polymers or polyelectrolytes, great improvements occurred. Settling rates for solids increased by a factor of 10, 20, or more, and solids that could not previously be thickened were responsive to the addition of polymers (10). In addition, lower doses of these additives were needed to produce the desired thickening. As a result, the volume of the solids did not increase and production rates improved (24).

ADVANTAGES AND DISADVANTAGES OF CHEMICAL TREATMENT

The proper application of chemical additives will--

1. Increase separation efficiency.

2. Increase throughput.

3. Require a minimum investment for mixing equipment.

There are, however, several disadvantages to using chemical additives--

1. They can be very expensive.

2. A large concentration of some additives may be needed to produce the required results.

3. Not all slurries are responsive to the chemical additives (1).

Chemical additives are often needed to increase the efficiency or throughput of a dewatering system. The use of chemical additives must be carefully considered because they are expensive to purchase and will increase the processing cost.

GRAVITATIONAL METHODS

BACKGROUND

This section describes equipment known as thickeners because of their capability to concentrate or thicken the solids from a slurry that has a very low solids content. In gravitational methods, the force of gravity causes the solids to settle and separate from the liquid. In the mining industry, flocculants are added to the slurry which cause the solids to form larger masses or "flocs" that settle at an accelerated rate. With respect to dewatering slimes, thickening equipment is often used to concentrate the fine particles for further processing by other dewatering equipment which will then produce the final dewatered product.

Over the years, nomenclature has been adopted that is specific to thickener functions in the mineral processing industries. The solid-liquid mixture that is to be separated by sedimentation is known as the feed. The sedimented material having a solids content higher than that of the feed is the underflow. The clarified liquid from which solids have been removed is the overflow. This terminology is used even for equipment where the overflow does not migrate over a weir or the underflow does not emerge from the bottom of the sedimentation device (17).

Regardless of name or variation in design, all gravitational equipment depends on sedimentation to produce the thickened product. During initial sedimentation, solid particles in a nonturbulent solution move downward under the influence of gravity relative to the liquid. The velocity of this movement increases until the upward force of drag, caused by the viscosity of the liquid, equals the downward force of gravity on the particles. These particles then fall at a constant velocity, called the terminal or free-settling velocity. In addition to gravity, the size of the reacting force is dependent on the particle diameter and density and the solution density and viscosity. The magnitude of the terminal settling velocity can be shown as--

$$v_s = \frac{Gx^2(D_s - D_l)}{J8\mu}, \tag{1}$$

where v_s = the terminal settling velocity, feet per second,

G = the gravitational constant, feet per second per second,

x = the particle diameter, inch,

D_s = the particle density, slugs per cubic foot,

D_l = the liquid density at a specified temperature, slugs per cubic foot,

μ = the viscosity of the liquid at a specified temperature, pound-seconds per square foot,

and J = a correction factor for particle shape, dimensionless (26).[3]

Where nonspherical particles are concerned, v_s alters by factor (J), which is less than 1.0. Where there is a low density of particles, v_s also decreases by a factor that is a function of the particle concentration. The purely mathematical description of sedimentation is imprecise owing to variations in particle diameter, shape, and distribution for any given slurry. Therefore, the above formula will be used for the determination of boundary conditions such as the smallest particle size just completely retained by a given device (26).

As sedimentation continues, the concentration of the solids increases through the process of "hindered settling" and then continues into the "compression phase." In this phase, a further concentration of the sediment occurs and an increase in the concentration of solids takes place; however, the process proceeds at a slower rate. This slowing down is due to the fact that during the exchange of solid matter for water, the water does not reach the top relatively unhindered but has to pass through increasingly narrowing capillaries between the more densely packed particles. Also, the friction between the touching particles slows down the compression process (22).

As the sedimentation process reaches equilibrium in the tank, four zones will be present (fig. 1):

1. Clear solution zone.

2. Feed zone.

3. Transition zone.

4. Compression zone (13, p. 27-71).

The clear solution zone contains the clarified solution of the overflow. The feed zone has the solids concentration of the unsedimented feed. The transition zone has a higher concentration of solids in hindered settling. The compression

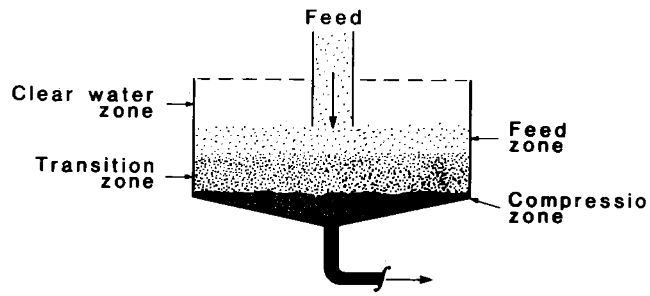


FIGURE 1. - Cross section of gravity thickener showing zones of the sedimentation process (13).

[3] A list of all mathematical symbols used in this report is given in appendix A.

zone has the highest concentration of solids in compression and is the origin of the underflow.

Gravity thickeners consist of open tanks with a feed inlet at the top and a means of collecting the sludge at the bottom by a rake. As the contents are slowly stirred by the rake rotation, the solids settle and are drawn off the bottom in a continuous underflow. Gases escape from the surface, and clear solution is removed by the overflow weir (24). Chemicals are added to the feed to aid settling, and, in the mining industry, many gravitational thickeners produce underflows having over 40 pct solids with recoveries approaching 100 pct.

Three general types of thickeners will be discussed in this report: conventional, high-rate, and multiple-plate thickeners. The following sections discuss each of these in detail.

CONVENTIONAL THICKENERS

Conventional thickeners use the sedimentation principle for liquid-solid separation. They are much larger than high-rate or multiple-plate thickeners and have several characteristic design features such as--

1. Cylindroconical shape.
2. Annular overflow weir.
3. Walkway and feed pipe support.
4. Feed well.
5. Drive mechanism and rake.
6. Underflow cone or trench (17).

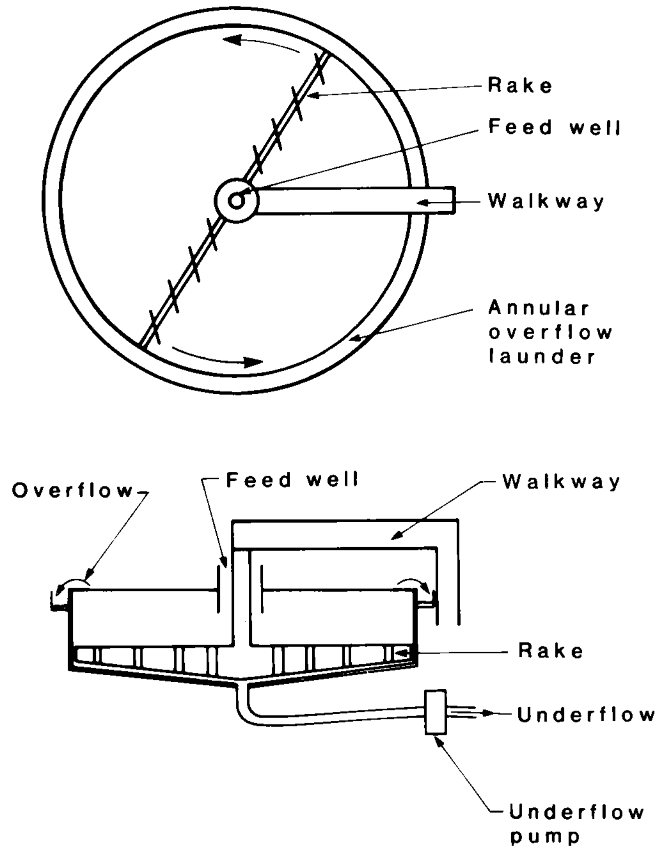


FIGURE 2. - Plan and cross-section views of conventional thickener (17).

Conventional devices are typified by a cylindrical upper portion attached to a shallow conical section having the apex oriented downward (fig. 2). The width or diameter of these thickeners is much larger than the depth. Most conventional thickeners are equipped with an annular overflow weir, which may be located internal or external to the tank and which is equipped with a froth baffle. Usually, overflow is regulated through a notched weir so that adjustments can be made to compensate for uneven tank settlement in the subjacent soil. Most thickeners have a walkway to the center of the thickener. The walkway usually serves as support for the piping that carries the feed to the center of the machine. The feed pipe terminates at a device located at the center of the thickener, which is called the feed well or center well. The function of the feed well is to dissipate the kinetic energy of the incoming feed and to form a zone of quiescence conducive to sedimentation. Feed wells are manufactured in many shapes and sizes because recent information shows that the feed introduction method and the feed well shape greatly affect sedimentation behavior. The function of the rakes in a thickener is to gently move the sedimented solids from the periphery towards the center discharge point. The movement of the rakes is provided by a drive mechanism. After the sedimented solids have been moved toward the center of the tank by the rake, they are removed from the thickener through a cone or trench located near the bottom center (17).

Although they are all based on the same sedimentation process, a variety of conventional thickeners are being manufactured which differ from each other in their design. For instance, the feed pipe support may extend from the edge of the tank to the center or may span completely across the diameter of the tank. The rake arms in some thickeners are rigidly attached to a central vertical shaft or lattice, while in other designs, the arms may be suspended from cables. Drive mechanisms also vary from worm-spur gear combinations to hydraulically operated push-pull arrangements for rotating the rake arms about the central axis. Each of these thickener configurations is designed to assist the process of continuous sedimentation by steadily removing the consolidated solids and the clear liquid to make room for the introduction of more feed material (17).

HIGH-RATE THICKENERS

Recently, several manufacturers have introduced high-capacity or high-rate thickeners, which have much smaller tanks than conventional thickeners. The lateral area for a conventional installation ranges from 5 to 10 ft^2 for each ton of solids thickened per day, but the area of high-rate units may be as low as 0.3 to 0.6 ft^2 for each ton of solids per day (17).

These high-capacity thickeners have smaller tanks because they discharge the feed directly into the bed of settled solids and use chemical additives to hasten the flocculation of the sediments. The efficiency of the flocculation depends on how thoroughly the flocculant and the slurry are mixed. The many different high-capacity thickeners on the market use various mechanisms or designs to mix the flocculant with the slurry.

High-rate thickeners, while smaller than conventional models, can often meet or exceed the performance of the larger units. In one such comparison between a conventional and a high-rate thickener, the latter unit produced underflow solids concentrations equal to or greater than those produced by the conventional unit. This was done even at feed input rates 10 times that of the conventional thickener. These results, though, were dependent on the slurry solids being responsive to the chemical flocculant used during the trials (10).

Figure 3 shows a typical high-capacity thickener. The chemical additives are combined with the incoming feed and previously thickened solids in the mixing chamber. This mixture is thoroughly churned by the blades of the mixing mechanism. The high concentration of the solids causes all chemical reactions to occur quickly and completely so that flocculation is considerably improved. This mixture moves from the reaction chamber to the clarified zone, where the flocculated solids quickly settle, to be used again with the incoming feed. A portion of the solids is collected inward by the rakes and evacuated as underflow. As in conventional thickeners, the clear solution migrates upward through the circulating sludge bed and exits over the overflow weir (23, p. 19-52).

Gravity plays only a part in the solids-liquid separation in this thickener. It would be more precise to refer to this as a filtration unit in which the filter media is a suspended sludge bed. This is true because as the solids settle, the pore spaces between them become more restricted and trap other solids being carried along with the water migrating upward (17).

The horsepower required to operate these thickeners is approximately equivalent to the horsepower for conventional units having a similar capacity. The drive unit employed by high-rate thickeners is usually a hydraulic arrangement (17).

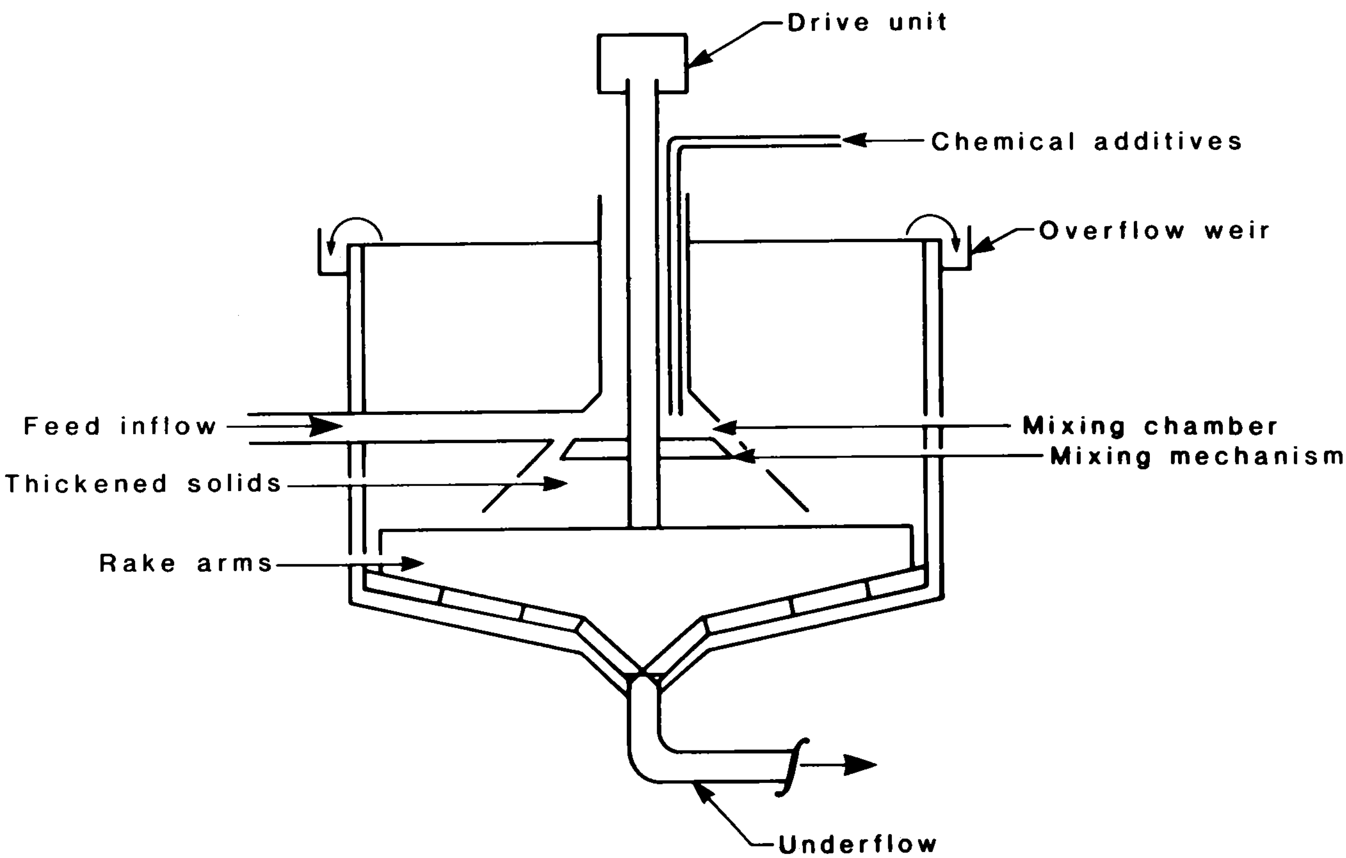


FIGURE 3. - Cross section of high-rate thickener (23).

ADVANTAGES AND DISADVANTAGES OF CONVENTIONAL AND HIGH-RATE THICKENERS

Thickeners have been used extensively by the minerals industry for concentrating slurries, and over the years many design improvements have been made that enable them to be very reliable and efficient for thickening operations. Gravitational thickeners, in general, may be advantageous to an operation because they--

1. Are capable of continuous operation.

2. Are capable of processing slurries with a variety of solids concentrations and size ranges.

3. Have fairly low maintenance and operational costs for the mechanical equipment.

4. Provide a kneading action by the rake mechanism which is beneficial to the compression process (26).

In addition, high-rate thickeners have some advantages over conventional thickeners, such as--

1. Smaller lateral space requirements (i.e., lower installation costs).

2. Greater throughput of solids based on available area for settlement (10).

Gravitational thickeners, though, do have some drawbacks, which include--

1. Requirements for large spaces.

2. Necessity for large watertight basins.

3. Sensitivity to persistent strong winds, especially for very large units.

4. Expensive operation if large amounts of flocculants are needed to obtain the desired concentration (10).

MULTIPLE-PLATE THICKENERS

Innovations have been made to reduce the space needed for gravity thickeners. First among these is the multiple-plate thickener, which uses a series of evenly spaced inclined plates positioned in a settling tank.

Multiple-plate thickeners are in use by the minerals industry around the world for clarifying, classifying, and thickening. In the United States, this equipment is being used in the coal industry for dewatering the waste products of coal cleaning (7).

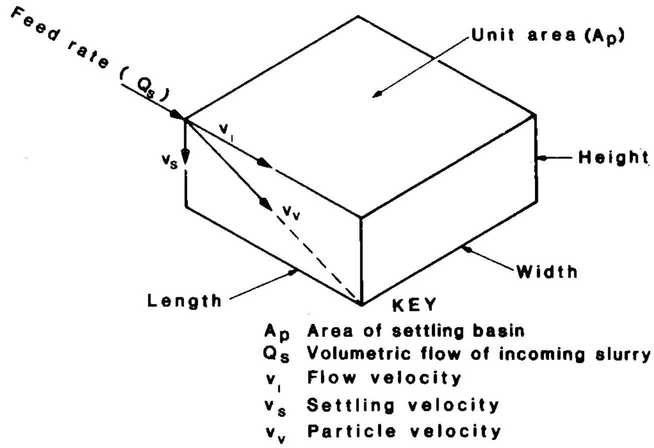


FIGURE 4. - Diagram of particle path in ideal settling tank (7).

The multiple-plate thickener can be used to accomplish both free settling and hindered settling; however, for simplicity, only free settling will be discussed in describing the theory of operation. For an ideal settling basin, the thickener feed enters at one end of the basin, flows uniformly along its length at velocity v_l, and exits at the other end (fig. 4). Any one particle will settle at velocity v_s. The actual trajectory of the particle is indicated by the vector v_v. If the trajectory takes the particle to the bottom of the basin before it reaches the far end, then the particle is assumed to have been removed from the liquid. A particle starting at the top must settle to the bottom at velocity v_s in the same time or less than the velocity of the liquid, v_l, in the basin. Thus, the feed quantity, Q_s, divided by the settling area, A_p, of the basin is known as the overflow rate or surface loading and is often expressed as gallons per minute per square foot. Based on this relationship, all particles are removed that have a settling rate equal to or greater than the overflow rate. It should be noted that the height or detention time of the basin is not one of the main parameters that affect the separation efficiency (7).

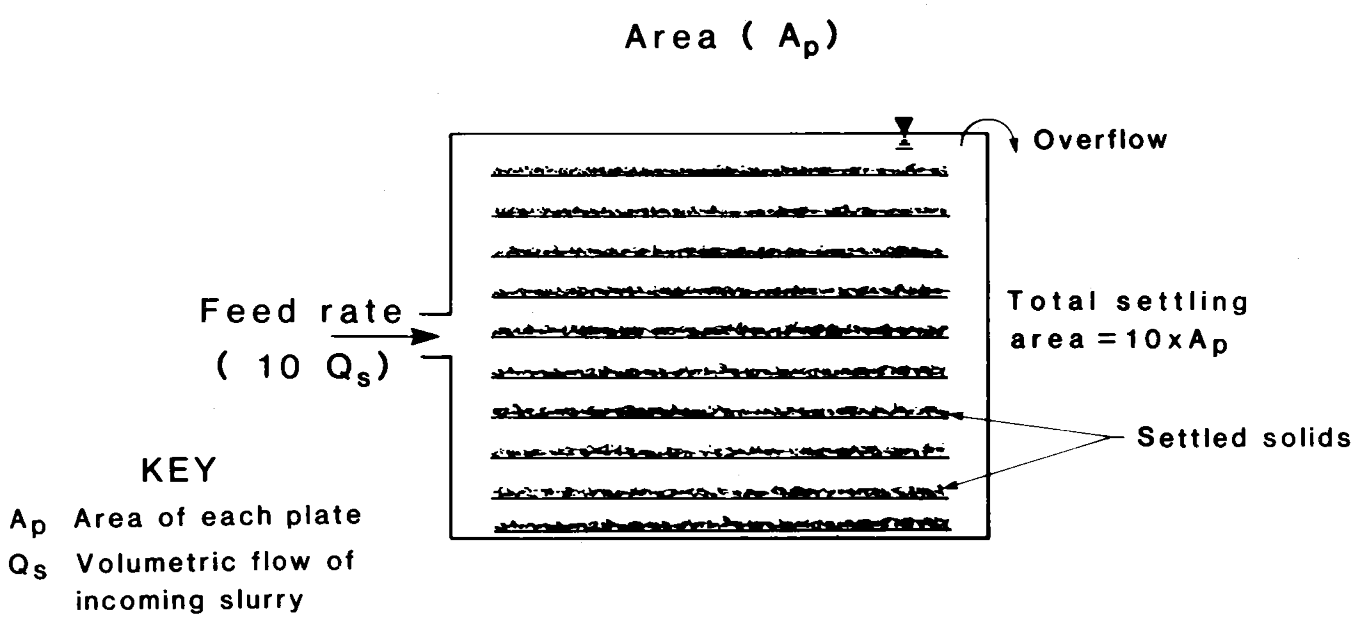


FIGURE 5. - Settling basin containing 10 parallel plates (7).

If the depth of the basin is reduced to a few inches and a number of such basins are stacked on top of each other, the result is a simple multiple-plate sedimentation device. Figure 5 shows such a unit containing 10 plates, which can theoretically handle 10 times the flow rate as could the same basin without any plates. The liquid detention time is one-tenth as long, and the settling area is 10 times as large, so the same separation efficiency is achieved because the overflow rate is unchanged (7).

In practice, the plates are inclined so the settled particles will slide downward and the plates are, in effect, self-cleaning. Figure 6 shows an arrangement of 10 plates set at an angle of 60° above the horizontal. In this case, however, the plate area must be multiplied by the cosine of the angle to correctly determine the capacity and overflow rate, because only the projected area of each plate on a horizontal plane is counted. Thus, the total settling area is 10 times the plate area times cosine 60°, which equals $5A_p$, and the capacity of the unit is $5Q_s$ (7). The following equation is used for determining the terminal settling velocity of the smallest grain size just completely retained:

$$v_s = \frac{S_p\, v_l}{L_p + \dfrac{2\, S_p}{\sin 2\, A_l} \cos A_l}, \qquad (2)$$

where v_s = the terminal settling velocity, feet per second,

S_p = the spacing between plates, feet,

v_l = the feed flow velocity, feet per second,

L_p = the plate length, feet,

and A_l = the inclination angle above horizontal of the plates, degrees (26).

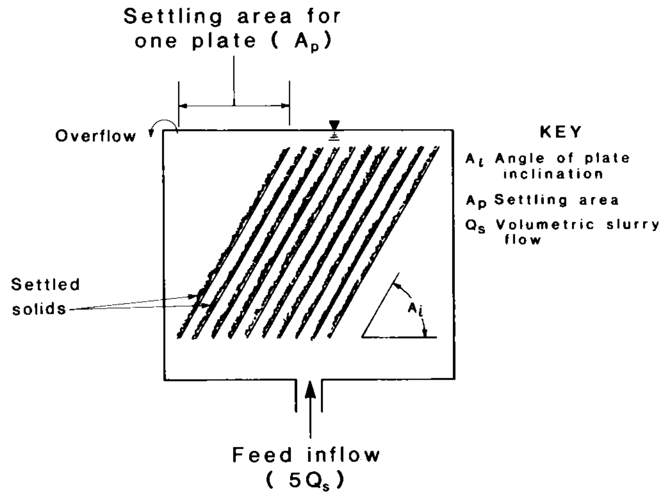


FIGURE 6. - Settling basin containing 10 plates set at angle of 60° above horizontal (7).

From equation 2, it appears that even at very high flow velocities the finest particles can be retained if the plates are long enough. This is not true, though, because the flow becomes turbulent with high flow velocities. At sufficiently high flow velocities, particles that may have been deposited on the plates will be flushed away when the force exerted on them by the flow becomes larger than the force of gravity on them (26).

A typical thickener plate is usually 2 ft wide and 10 ft long. Plate spacing is a critical variable because it must be large enough to prevent the settled solids from being disturbed by the upward flowing liquid but close enough to give the benefit of compactness. A 2-in spacing between plates is generally safe and is often used. The plate angle is another critical variable because the angle must be steep enough for the solids to flow or slide down the plate easily. For mineral-type solids having a high specific gravity, 45° is adequate; however, application experience is the best guide in this area (7).

TYPES OF MULTIPLE-PLATE THICKENERS

Two different types of multiple-plate thickeners are available. Each type differs according to the direction of the feed flow in relation to the inclination of the plates. The first type operates using the counterflow principle, where the feed material rises between the inclined plates against the direction of inclination, as described in the previous section, to develop the principle of multiple-plate sedimentation. Figure 7 shows a cross section of a typical countercurrent thickener and its component parts. Note that the feed inflow line extends to the center of the tank in order to distribute the feed evenly to all plates (26).

The second type of multiple-plate thickener uses the crossflow principle (fig. 8). The incoming slurry flows across the width of the inclined plates. The terminal settling velocity for the solids is determined from the equation--

$$v_s = \frac{S_p\ v_l}{L_p\ \cos\ A_i}, \tag{3}$$

where the variables are the same as those defined for equation 2 (26).

The crossflow multiple-plate thickener is more efficient in small-particle separation and allows higher feed velocities than the counterflow model. This is because the directions of the feed flow and the solids flow down the plate are perpendicular and not opposed to each other as in the first type. There are practical limits to the highest velocities

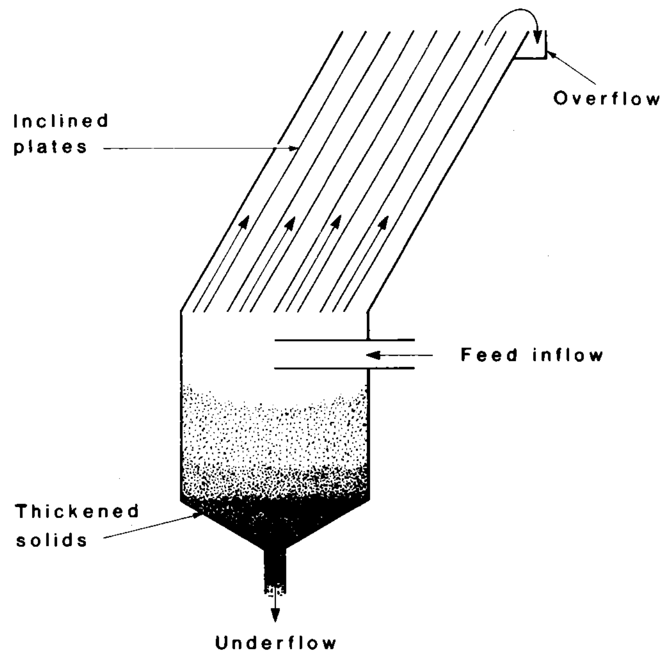


FIGURE 7. - Cross section of countercurrent multiple-plate thickener (26).

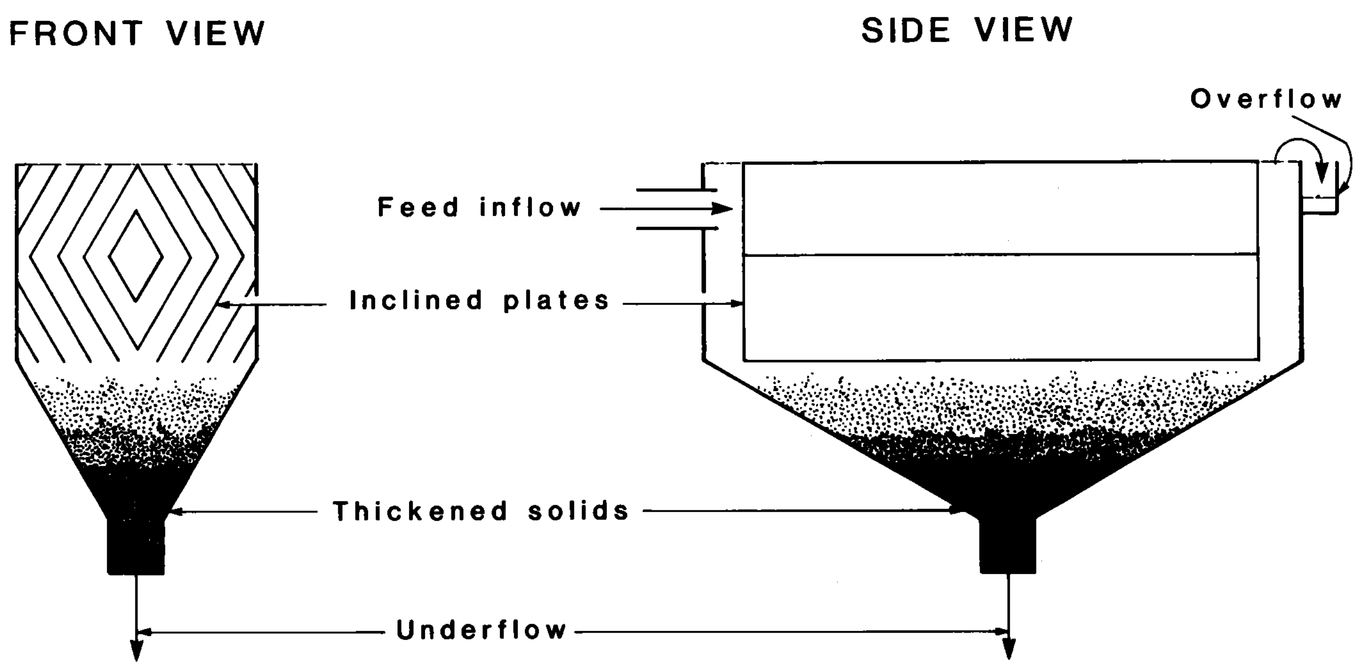


FIGURE 8. - Front and side cross sections of typical crossflow multiple-plate thickener (26).

possible because the very fine solids may not have sufficient time to settle to the plate at very high feed velocities (26).

ADVANTAGES AND DISADVANTAGES OF MULTIPLE-PLATE THICKENERS

The multiple-plate thickener is simple in design and operation. The only item requiring operator attention is the sludge withdrawal rake located at the underflow outlet. This is usually controlled by a variable-speed, positive displacement hopper, which is manually or automatically set by a detector monitoring the sludge level in the hopper (7).

Spacing and inclination of the plates are very critical for an efficient operation; however, once these parameters have been determined, the multiple-plate thickener can reduce the space requirements by as much as 90 pct compared with conventional thickeners (21). Multiple-plate thickeners are simple in construction and can be prefabricated for rapid erection (31). These thickeners are normally constructed of mild steel or 3/16-in stainless steel. For corrosive slurries, polyvinylchloride (PVC) or fiberglass-reinforced plastic plates can be used with rubber-lined tanks (7).

SIZING THICKENERS

Settling Rate Basis

The sizing of thickeners can be done using the method proposed by Coe and Clevenger. When the suspending fluid is water, the area necessary for particle settlement can be estimated by the following formula:

$$A_{cs} = \frac{1.333(W_{rs} - W_{rd})}{S_f}, \qquad (4)$$

where A_{cs} = the cross-sectional area of the basin, square feet per ton of dry solids thickened in 24 h,

W_{rs} = the weight ratio of water to solids in the slurry,

W_{rd} = the weight ratio of water to solids in the discharge,

and S_f = the settling rate for a given W_{rs}, feet per hour (13, p. 27-72).

This equation uses various values of W_{rs} obtained from batch sedimentation tests. The procedure for these tests begins by mixing the solids to be tested in a graduated cylinder of water at a specific water-to-solids ratio, W_{rs}. The cylinder is shaken, and the fines are allowed to settle. The settling rate, S_f, is the subsidence rate for the solids in the bottom of the cylinder. Several values of W_{rs} should be tested, including the final discharge value, W_{rd}. The largest area, A_{cs}, obtained will govern the size of the settling basin or number of settling plates (13, p. 27-72).

The capacity of a thickener can then be estimated by using the equation proposed by Deane.

$$V_{sc} = \frac{1.33\, t_{hs}\, (S_{gs} - 1)}{S_{gs}\, (S_{gsc} - S_{gw})}, \qquad (5)$$

where V_{sc} = the volume of solids in compression, cubic feet per ton of solids per 24 h,

t_{hs} = the holding time for the solids to settle from the entering dilution of compression to the dilution of the discharge, hours,

S_{gs} = the average specific gravity of the solids,

S_{gw} = the specific gravity of water, which is 1.0,

and S_{gsc} = the average specific gravity of the solids in compression (13, p. 27-72).

The foregoing analysis assumes that the settling rate is dependent only on the feed solids concentration. Running settlement tests with different solids concentrations will indicate the settling rates at different depths in a thickener tank. Talmadge and Fitch proposed another method based on the work done by Kynch. This method uses just one settlement test and provides an estimate of the coincident settling rate and solids concentration for various depths (32, p. 90).

In this analysis, a slurry having the anticipated solids concentration of the full-scale operation is put into a tall cylinder. The height of the interface between zones A and B is recorded at frequent intervals. From these data, a graph of the interface height versus time is plotted as shown in figure 9 (18, p. 12-32).

The relationship among solids concentration, particle settling velocity, and interface height is shown by the following equation:

$$C_{se} = \frac{C_{ss}\,H_{abo}}{H_{abe} + v_s\,t_e}\,, \qquad (6)$$

where C_{se} = solids concentration at time t_e, pounds per cubic foot,

C_{ss} = solids concentration of the feed slurry, pounds per cubic foot,

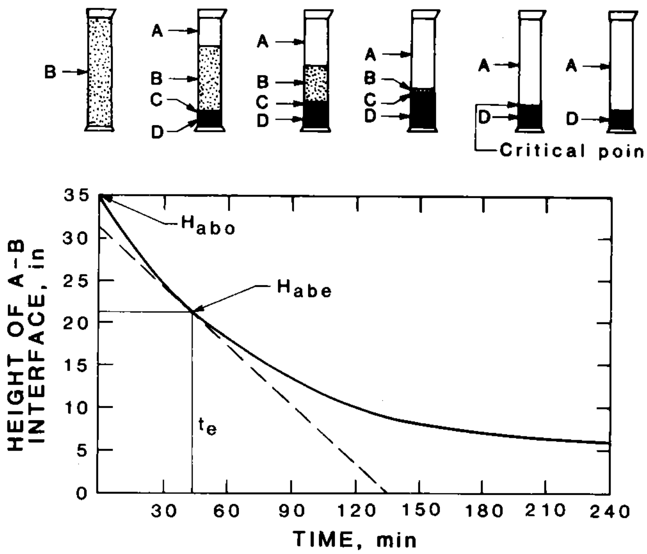
H_{abo} = original height of A-B interface, feet,

H_{abe} = height of A-B interface at time t_e, feet,

v_s = solids terminal settling velocity, feet per hour,

and t_e = an arbitrary time after solids settlement begins, hours (18, p. 12-36).

The solids settling velocity, v_s, is the slope of the tangent to the curve for interface height versus time at time t_e. The quantity $H_{abe} + v_s\,t_e$ is the intercept of the tangent with the ordinate.



KEY

—— Height of A-B interface

– – – Tangent to curve

A Zone A-clear solution

B Zone B-solids at concentration of feed

C Zone C-transition between B and D

D Zone D-solids in compression

H_{abe} Height of A-B interface at time t_e

H_{abo} Original height of A-B interface

t_e Arbitrary time after settlement begins

FIGURE 9. - Graph of A-B interface height versus time for batch settling test (18).

When the feed inflow rate, feed solids concentration, and underflow solids concentration are established, the area of the thickener tank can be determined from the formula--

$$A_{cs} = \left(\frac{Q_s\,C_{ss}}{v_s}\right)\left(\frac{1}{C_{se}} - \frac{1}{C_{su}}\right), \qquad (7)$$

where A_{cs} = tank area, square feet,

Q_s = volumetric feed rate, cubic feet per hour,

and C_{su} = solids concentration of underflow, pounds per cubic foot (18, p. 12-36; 32, p. 93).

A series of solids concentration and settling velocity values can be determined from the interface height versus time curve. These values can then be used to establish and plot the solids concentration at different times. Several thickener areas are calculated using the corresponding values of v_s and C_{se}. A graph of A_{cs} versus v_s can then be plotted. The maximum value for A_{cs} will be used for designing the thickener (18, pp. 12-36).

After the required area has been established, the depth of the compression zone can be determined based on the retention time of the solids in this zone. This time is dependent on the rate of discharge and the concentration of the underflow. These values can be obtained from the circuit required for the underflow solids concentration and from the interface versus time graph (32, p. 93).

The volume and the height of the compression zone can then be computed:

$$V_{sc} = \frac{Q_s\ C_{ss}\ t_{hs}}{D_s} + \left(\frac{Q_s\ C_{ss}}{D_l}\right)\left(\frac{1-C_{ave}}{C_{ave}}\right), \quad (8)$$

and $$H = \frac{V_c}{A} = \frac{4\ V_c}{\pi d^2} \text{ for a circular tank,} \quad (9)$$

where V_{sc} = volume of the compression zone, cubic feet,

t_{hs} = retention time, minutes,

D_s = density of the slurry solids, pounds per cubic foot,

C_{ave} = average slurry solids concentration, pounds per cubic foot,

and D_l = density of the slurry liquid, pounds per cubic foot (32, p. 93).

The total height of the thickener tank can be determined by allowing 2-1/2 to 5 ft above the compression zone for the solution in the clarified and transition zones (32, p. 94).

Of the two methods discussed, the Kynch method is more often used for flocculated slurries. It must be remembered that these equations will provide only approximate values for sizing a conventional thickener. Pilot plant experimentation must be done using the slurries to be thickened in order to determine the optimum design features.

Solids Flux Basis

The Coe-Clevenger and Kynch methods have been used many years for designing settling tanks; however, they have limitations because they are used to design a dynamic process based upon data obtained from a static settling test. A recent innovation for designing gravity settling equipment uses the solids flux concept.

Solids flux is defined as the mass rate of solid flow through a unit area and portrays the conditions in a modern continuous process gravitational thickener. The solids flux in a batch settling test is dependent only upon the velocity of the settling particles. In a continuous thickener, however, the solids flux is dependent not only on the settling velocity but also on the bulk transport of solids in the underflow. This can be represented as--

$$F_{ts} = F_{bt} + F_s = C_{ss}\ (v_{bt}) + C_{ss}\ (v_s)$$
$$= C_{ss}\ (v_{bt} + v_s), \quad (10)$$

where F_{ts} = the total solids flux, pounds per square foot per hour,

F_{bt} = the bulk transport flux component based on the underflow, pounds per square foot per h,

F_s = the settling flux component, pounds per square foot per hour,

v_{bt} = the bulk transport velocity of the solids, feet per hour,

v_s = the settling velocity of the solids, feet per hour,

and C_{ss} = the solids concentration of the slurry, pounds per cubic foot (32, p. 95).

Figure 10 shows a graphic relationship among the various components of the solids flux for a continuous gravity thickener. This graph can be constructed for any gravity thickener, once the underflow rate and the bulk transport velocities have been determined (32, p. 96).

The figure indicates that for low solids concentrations, most of the total flux is composed of the settling component. As the solids concentration is increased, the bulk transport component represents a correspondingly greater portion of the total flux. When the total flux is less than the solids handling capacity, there is only one associated value of solids concentration, and there will be a zone of constant concentration for any particular underflow rate. If the total flux is greater than the solids handling capacity, there will be two zones of constant concentration for any particular underflow rate, but the solids handling capacity represents the maximum practical value for stable thickener operation (29).

Mathematical manipulation of the variables indicates that

$$F_s = F_{ts}\left(1 - \frac{C_{ss}}{C_{su}}\right), \quad (11)$$

where C_{su} = the solids concentration of the underflow, pounds per cubic foot.

When designing a thickener, the total solids flux, F_{ts}, and the underflow solids concentration, C_{su}, will be known or fixed, but the settling flux component, F_s, and solids concentration,

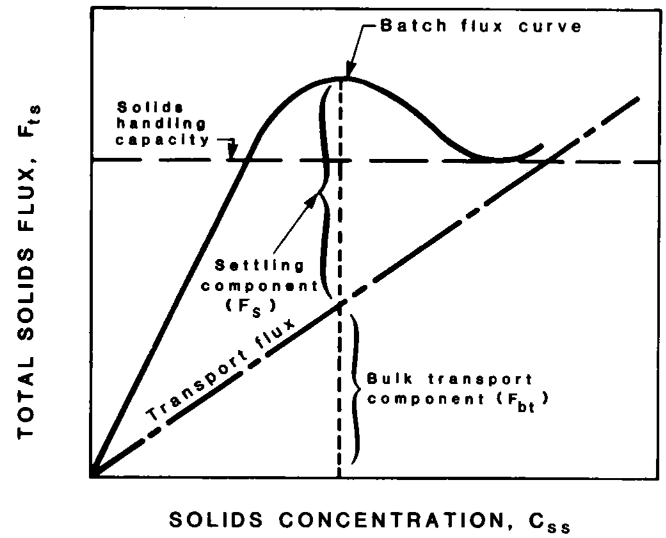


FIGURE 10. - Graph of total solids flux versus solids concentration showing relationships of the various components (32).

C_{ss}, will be unknown. Using values for F_{ts} and C_{ss} and picking values for C_{su}, equation 11 can be used to obtain corresponding values for F_s. A graph of the settling flux, F_s, versus concentration, C_{ss}, can then be constructed, as shown in figure 10 (32, p. 97).

If values of C_{ss} and the corresponding F_s are plotted as shown in figure 11, the result is a straight line with a slope of $-F_s/C_{su}$ and an intercept of F_{sI}. This is referred to as the operating line and represents the thickening characteristics for a particular thickener at any specified underflow rate. The batch flux curve can also be plotted, and the intersections of the operating line and the batch flux line indicate those solids concentrations that satisfy both relationships (32, p. 97).

In figure 11, an operating line has been plotted for a thickener that is functioning within its limits. In this example, only one zone of concentration exists at C_{ss1} to produce an underflow concentration, C_{su1}. If, however, the slurry concentration and the underflow concentration are increased to C_{ss2} and C_{su2}, respectively, the operating line intercepts the tangent to the batch flux curve and a second zone of concentration

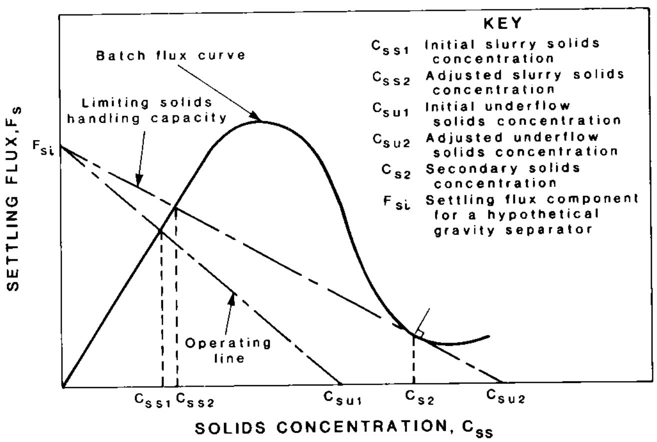


FIGURE 11. - Graph of settling flux versus solids concentration showing operating line and limiting solids handling capacity (32).

begins to appear with a concentration of C_{s2}. This condition represents the limiting solids handling capacity of a thickener at a particular underflow rate (32, p. 97).

In practice, the necessary data are obtained by running tests using a thickener and the slurry to be thickened. Altering the applied flux, F_{ts}, and the underflow concentration, C_{su}, will provide an indication of the limiting flux value. After the maximum value of C_{su} has been determined, a tangent to the batch flux curve can be drawn. The area of the tank can then be determined by dividing the maximum rate of solids loading by the limiting flux. Experience has shown, though, that the optimal thickener throughput will be about 90 pct of the values calculated (32, p. 98).

As with conventional thickeners, the use of a pilot plant is also necessary for determining the best size for a high-capacity thickener and for establishing the amount of flocculant needed for proper thickening. The properties of the slurry and the many available flocculants will vary considerably, so the best flow rates and flocculant injections should be determined experimentally.

Multiple-plate thickeners should also be sized according to results of laboratory and pilot plant testing. The results will aid the designer in determining the settling rate of the solids so that the overflow rate and effluent quality can be established. The sludge volume can also be determined in order to establish the underflow solids concentration. Finally, the need for chemical pretreatment can be evaluated (7).

CENTRIFUGAL SEDIMENTATION

BACKGROUND

Centrifugal sedimentation depends on the density difference between solids and liquids where the particles are subjected to centrifugal forces that make them move radially outwards or inwards through the liquid, depending on whether they are heavier or lighter than the liquid (32, p. 125). Centrifuges are compact machines and are capable of producing high liquid clarification and solids concentration. Most of the units available today are designed for continuous operation (24).

Centrifuges have relatively modest capital costs, but they may be expensive to operate because of the need for chemical conditioning in most applications, high power consumption, and extensive maintenance requirements. Lower speed units consume less power and, as a result, have fewer wear problems. Maintenance difficulties can be greatly reduced if the construction materials are specified to match the abrasive or corrosive characteristics of the slurry to be handled (24).

A large variety of centrifuges are available on the market; however, several types are used primarily for clarifying and not dewatering, so they are not included in this report. Solid-bowl, screen-bowl and disk centrifuges are of interest in dewatering slimes and are discussed in the following sections.

SOLID-BOWL CENTRIFUGES

Solid-bowl, scroll, or decanter centrifuges consist of a horizontally rotating chamber that has one end tapered into a cone. The slurry is admitted through axial feed tubes and removed radially out of the bowl. A screw or scroll mechanism rotates in the same direction as the bowl but 5 to 100 rev/min faster or slower than the bowl and thus can push the solids along the length of the chamber. The speed of the bowl rotation can vary from 1,600 to 6,000 rev/min. The solids are collected by the scroll towards the tapered end of the bowl, while the solution overflows a weir at the other end. The tapered section serves as a drying zone or beach area prior to the cake discharge (24, 32, p. 139). Figure 12 shows a simplified cross section of a typical solid-bowl centrifuge.

With regard to the machine design, a number of variations are available in the contour of the centrifuge shell, scroll flight angle and pitch, beach angle and length, conveyor speed, and feed position. An alternative to the liquid overflow outlet is an internally mounted tube for skimming off the liquid (32, p. 140).

Specially designed models of this centrifuge have been used to dewater very fine slurries in hydrocyclone circuits for recovering and dewatering deslimed coal. Other solid-bowl centrifuges have been used to dewater froth flotation tailings after thickening in a static thickener (18). Another major area of application for this type of centrifuge is in the classification of solids such as kaolin clay and titanium oxide (32, p. 140).

Polyelectrolytes are widely used for the flocculation of the solids to be dewatered in solid-bowl centrifuges. The point of addition varies, depending on the type of the polyelectrolyte and slurry. Anionic polyelectrolytes are usually introduced upstream from the centrifuge, while cationics are added within the centrifuge because they react very quickly with the slurry (32, p. 140).

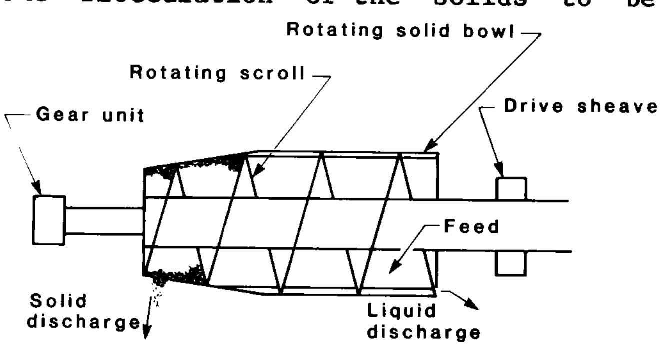


FIGURE 12. - Cross section of solid-bowl centrifuge (3).

Wilson and Miller (36) compared the effectiveness of solid-bowl centrifuges to that of disk vacuum filters and commented on the parameters that affect their use in coal dewatering. The solid-bowl centrifuge and the disk vacuum filter produced similar results on the same slurry. The solids from each of the two dewatering machines contained 20 pct minus 325 mesh with 20 pct surface moisture and a centrate or filtrate of about 0.1 to 0.2 pct solids. The products from these units are very similar in nature, so the choice between the devices should be made on the basis of handling characteristics, throughput, floor space, cost, and overall circuit or layout considerations. If the centrifuge is chosen, in-plant adjustments to feed rates, speed, and pool depth can be made to obtain the proper balance of throughput, moisture, maintenance cost, and effluent clarity (36).

Wilson and Miller also found that steam-heating the feed to 110° F resulted in a 4-pct reduction in product moisture; however, higher temperatures did not further improve results. The use of heat may be economically feasible as a means of reducing the product moisture if an inexpensive source of heat is available (36).

One of the principal advantages of this machine is that it can dewater dilute slurries. In plants that use shaking tables, no dewatering screens are required between the table and the centrifuge as would be needed when a screen-type machine is used. The solid-bowl unit will require more horsepower, though, because it must accelerate the water as well as the solids during the dewatering operation (18, p. 12-20).

SCREEN-BOWL CENTRIFUGES

Screen bowls, also called basket or perforate units, are a second type of centrifuge. Positive-discharge machines are screen-bowl centrifuges with transport devices and are the most common type of centrifuge found in the minerals industry today. These units have two elements that rotate about a vertical or horizontal axis. These elements consist of an outside conical screen frame and an inside solid cone that carries spiral hindrance flights. A gear arrangement produces a differential speed in the two rotating elements so that they both rotate in the same direction, but the screen element moves slightly faster than the cone carrying the spirals. The operation is similar to that of the solid-bowl centrifuge. The slurry enters the machine at the top and falls on the apex of the cone. The centrifugal force developed by the rotating cone throws the solid-liquid mixture against the screen. The water passes through the perforations and is collected in an effluent chamber. Meanwhile, the flights spiral downward and the solid material is gradually transported to the bottom of the screen. The conical shape of the basket causes the solids and water to be subjected to zones of increasing centrifugal force (18, p. 12-16). Figure 13 shows a cross section of a typical screen-bowl centrifuge.

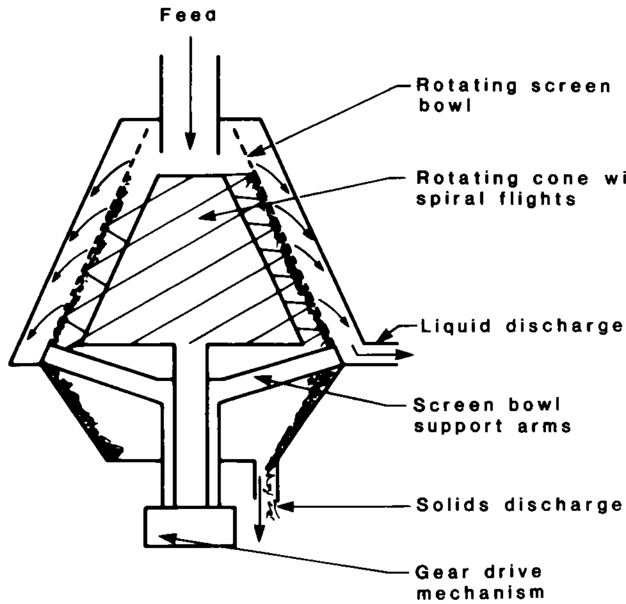


FIGURE 13. - Cross section of screen-bowl centrifuge (5).

Another type of screen-bowl centrifuge is the vibrating-basket type that is frequently being installed in new plants. This centrifuge has either a vertical or horizontal basket that vibrates in such a manner as to cause the solids to move through the machine. This vibration tends to loosen the bed of particles so that they are free-draining and only moderate force is required to effect thorough dewatering. Because of the low speed generally used in these centrifuges, the moisture content of the solids is usually higher than that produced by the transport-type unit; however, wear and horsepower are low and solids degradation is minimal. The principal difference between the horizontal and vertical screen-bowl types is that the horizontal axis machine requires less headroom than the vertical device (18, pp. 12-16 to 12-19).

Wilson and Miller also conducted tests on screen-bowl units. They found that the positive-discharge screen-bowl centrifuge provided 4 to 6 pct lower moisture in the solids product than the solid-bowl centrifuge. Thus, for similar sized centrifuges, the screen bowl is preferable to the solid bowl for this particular instance. As with the solid-bowl centrifuge, they found that the screen-bowl centrate must be bled out of the plant to a pond or backed up by a secondary recovery system. Because of the recovery requirement, this type of centrifuge should operate for maximum moisture reduction as opposed to maximum effluent clarity to fully benefit from this costly system (36).

Further test results indicated that only minor moisture reduction was obtained by steam heating the feed, so heating was not recommended for this type of centrifuge. Surface-tension-reducing chemicals and a flocculant were also tested; however, both approaches were ineffective in reducing moisture (36).

DISK CENTRIFUGES

The last type of centrifuge to be discussed is the disk centrifuge. Its construction is similar to that of the vertical screen-bowl unit; however, instead of using just one cone, it uses multiple cones for dewatering (fig. 14). The basic idea of increasing the settling capacity by using a number of disks in parallel is the same as the multiple-plate principle in gravity sedimentation (32, p. 141).

The slurry enters the unit at the top, and the solids are forced to the outer circumference by centrifugal action, then removed through the outer rim. The clarified water is channeled upward through passages between the disks. These centrifuges can handle inflows of up to 3,300 gal/min containing low-density particles up to 0.1 in in diameter and concentrations to about 1 pct solids. The output can be up to 6 pct solids, or even up to 10 pct solids if chemical additives are used (24).

Disk centrifuges are operated at speeds up to 12,000 rev/min, depending on the bowl diameter. The bowls usually have

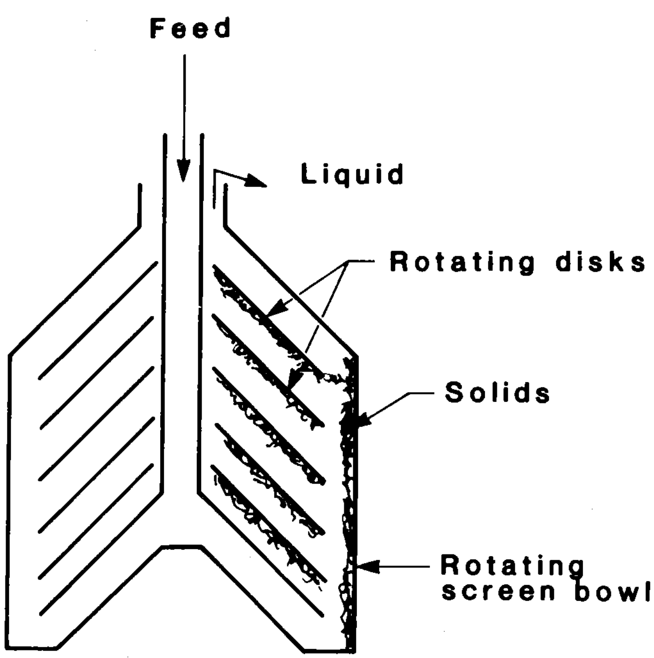


FIGURE 14. - Cross section of disk centrifuge (32).

equal dimensions of height and diameter for optimum capacities, and the angle of the cones is usually between 35° and 50°, which is large enough to facilitate the sliding of the particles on the disk surfaces (32, p. 143).

There are several variations in this design, which include recirculating the solids discharge, a facility for washing before discharge, and a paring tube for pressurized solids discharge (32, pp. 143-145).

Disk centrifuges are very effective for dewatering fine-grained solids and are often used for dewatering kaolin clay (32, p. 144).

CENTRIFUGE PERFORMANCE

The performance and efficiency of a centrifuge depend on a number of factors. The more important factors are--

1. Centrifuge rotation speed.
2. Diameter.
3. Length.
4. Beach angle and length (for horizontal centrifuges).
5. Feed point of slurry.
6. Feed point of flocculants.
7. Scroll rotation speed differential and pitch (for positive-discharge units).
8. Pool depth (for solid-bowl units) (11, p. 4-3).

Increasing the bowl rotation speed usually increases the solids recovery. There may be an increase in the solids cake concentration; however, the increase of the fines in the cake tends to increase the cake moisture. Higher speeds also increase both maintenance and operating costs of the centrifuge (11, p. 4-3).

Centrifuge-bowl diameters generally range from 6 to 50 in, and bowl lengths are generally from two to four times the bowl diameter. Bowl speed is normally a function of the bowl diameter because the effects of speed and diameter determine the resulting centrifugal force acting on the slurry. Typical values for centrifugal force range from 1,000 to 4,000 times the force of gravity; the higher centrifugal forces are associated with the smaller bowl diameters (11, p. 4-3).

The bowl length will affect the centrate clarity. A longer bowl increases the residence time of the slurry as it travels from one end of the bowl to the other. This increased time allows the finer particles sufficient time to separate from the liquid (2, p. 11-35).

In solid-bowl units, the beach angle and beach length will affect both the percent solids in the final cake and the torque needed to move the solids out of the centrifuge at a constant scroll differential speed. The longer the cake is allowed to dewater on the beach, the higher the cake solid concentration, and the higher the torque requirements for discharge (11, p. 4-4).

The feed entry point into the bowl will influence the percent solids in the cake and the solids recovery. For solid-bowl centrifuges, recovery will be improved if the feed entry point is near the beach because the slurry particles have a longer distance to migrate to the end of the machine where the liquid exits (11, p. 4-4).

Chemicals may be added to the slurry to accelerate the flocculation of fine particles that do not immediately separate from the liquid. The high degree of agitation and mixing within the centrifuge generally necessitates large polymer doses to effect an increase in the solids recovery. In one instance, a dosage range of 0.1 to 0.3 lb polymer per ton of solids increased the capacity of a centrifuge 5 to 10 pct at the same solids recovery (11, p. 4-9).

In positive-discharge units, changes in the scroll rotation speed and pitch will affect the solids recovery and cake dryness. Small residence times allow only the heavier and larger solids to be removed, while the finer solids remain suspended in the liquid. This factor will be particularly important when dewatering slimes.

Most solid-bowl centrifuges are provided with an adjustable pool depth setting so that the liquid level in the bowl can be changed after installation. An increase in the pool depth will improve the percentage of the solids that are separated from the liquid. This improvement in solids recovery is due to an increased residence time and a reduction in the agitation within the centrifuge at deeper pool depths; however, deeper ponds will increase the amount of moisture in the cake because of the reduced dewatering time on the beach (11, p. 4-4).

ADVANTAGES AND DISADVANTAGES OF CENTRIFUGES

A main advantage of the centrifuge is its operational flexibility. Within its design limits, a centrifuge can be fed a slurry at various rates and still provide a consistent solids product. If the feed rate exceeds the design limits of the unit, the excess solids appear in the liquid; however, the quality of the discharged cake does not deteriorate and the percent solids remains relatively constant (11, p. 4-4).

A disadvantage of the centrifuge is abrasive wear on the scroll and other interior parts, which results in high maintenance costs. In recent years, scrolls have been manufactured with improved materials such as tungsten carbide on the wearing surfaces. Operational results indicate an order of magnitude improvement in the life of these components (3; 11, p. 4-4). Another drawback of the centrifuge is that the feed slurry may need to be prethickened. Although centrifuges can dewater a wide range of slurries, a very low feed solids content means that the centrifuge must process large volumes of slurry input. The number of centrifuges needed in a dewatering circuit is directly proportional to the volume of feed slurry to be processed. Consequently, the slurry should be prethickened by sedimentation, hydrocycloning, or other means prior to centrifugation. The final circuit configuration will be the result of compromising the performance and economy of the prethickening and centrifuge equipment (11, p. 4-7).

SIZING CENTRIFUGES

When centrifuging a slurry, it is not necessarily true that an increased force will decrease the moisture content of the product. Materials that deform, break, or degrade will not be dewatered proportionally to the applied forces. It should also be realized that horsepower, wear, maintenance, and degradation will accelerate with increased forces applied in the machine. While a large centrifugal force developed by a machine may be an indication that it is sturdily built, this force should not be the only criterion used in selecting a centrifuge to dewater any material (18, p. 12-15).

For any particle traveling in a circular motion about a point, the centrifugal acceleration is

$$A_c = \frac{v_p{}^2}{R_c}, \tag{12}$$

where A_c = centrifugal acceleration, feet per second per second.

v_p = the linear peripheral velocity, feet per second,

and R_c = the radius of curvature, feet (18, p. 12-14).

Centrifugal acceleration is then expressed as multiples of the gravitational acceleration:

$$\frac{A_c}{G} = \frac{v_p{}^2}{GR_c}, \tag{13}$$

and

$$F_c = \frac{v_p{}^2}{GR_c} = \frac{(2\pi R_c N)^2}{GR_c} = \frac{4\pi^2 R_c N^2}{G}, \tag{14}$$

where $v_p = 2\ R_c N$, feet per second,

N = the number of revolutions per second,

$\pi = 3.1416$,

F_c = the centrifugal acceleration expressed as multiples of gravitational acceleration, dimensionless,

and G = the gravitational acceleration, feet per second per second (18, p. 12-14).

SIGMA CONCEPT

The "sigma concept" has been widely used in the field of centrifugal sedimentation for the last 30 yr. It is a simplified representation of machine performance in terms of the particle size, the total volumetric rate, and an index of the centrifuge size. The sigma concept characterizes a centrifuge's ability to separate solids from liquids and is widely used in industry (32, p. 130).

The volumetric throughput of a centrifuge can be expressed as--

$$Q_c = 2\ v_s\left[2\pi L_{sz}\frac{\omega^2}{G}\left(\frac{3}{4}\ R_{cb}{}^2 + \frac{1}{4}\ R_{sz}{}^2\right)\right], \tag{15}$$

where Q_c = the volumetric throughput, cubic feet per second,

v_s = the terminal settling velocity of the solids, feet per second,

ω = the angular velocity of the solids, radians per second,

G = the acceleration of gravity, feet per second per second,

L_{sz} = the length of the settling zone, feet,

R_{sz} = the radial length from the rotational axis to the settling zone surface, feet,

and R_{cb} = the radial length from the rotational axis to the centrifuge bowl surface, feet (11, p. 4-5).

This equation is composed of two parts. The first component, v_s, describes the settling characteristics of the solids. The remainder of the equation represents the machine variables that effect separation efficiency, such that--

$$\Sigma = 2\pi L_{sz}\frac{\omega^2}{G}\left(\frac{3}{4} R_{cb}^2 + \frac{1}{4} R_{sz}^2\right), \quad (16)$$

where Σ = the sigma value for a particular centrifuge configuration (11, p. 4-5).

Equation 16 is the basic expression of the sigma concept, which gives an estimate of the maximum flow rate that will allow solids of a particular size to separate from the liquid. Sigma is a constant containing factors pertaining only to the centrifuge and can be thought of as the theoretical capacity factor. It is expressed in terms of area and facilitates comparison between the performances of geometrically and hydrodynamically similar centrifuges processing the same slurry (32, p. 131).

A shortcoming of the sigma concept is that the cut size or smallest particle separated from the liquid is not suitable as a criterion for separation efficiency because different total efficiencies can be obtained for a given cut size, depending on the size distribution of the solids. The best method of describing the performance of a centrifuge is by using the grade efficiency curve, which is briefly discussed in appendix D. This requires many tests, together with deeper theoretical considerations, but the results will provide more meaningful and reliable predictions of total efficiencies for different slurries (32, p. 133).

PILOT-PLANT TESTING

The centrifuge evaluation procedure, generally used in pilot testing, relates the percent of solids recovered and cake solids concentration to the feed rate at various operating conditions. The recovery of feed solids in the centrifuge cake is determined from the measurements of flow rates and concentrations entering and leaving the centrifuge. By combining the material balance equations and the definition of recovery, the following simplified equation relates the recovery to various solids concentrations:

$$R_{sc} = \frac{W_{sc}\,(W_{ss} - W_{sl})}{W_{ss}\,(W_{sc} - W_{sl})}, \quad (17)$$

where R_{sc} = the percent recovery of solids in the cake,

W_{sc} = the weight percent of solids in the cake,

W_{sl} = the weight percent of solids in the centrate,

and W_{ss} = the weight percent of solids in the fed slurry (11, pp. 4-6 to 4-7).

This equation simplifies the testing procedure so that the solids recovery can be calculated from concentration measurements made on the three process streams entering or leaving the centrifuge. Thus, flow rate measurements need not be made to calculate the recovery of solids in the cake (11, p. 4-7).

FILTRATION EQUIPMENT

BACKGROUND

Filtration is a process where solids are separated from liquids by passing a slurry through a permeable filter medium that retains the solids. To cause the fluid to flow through the filter medium, a pressure drop has to be applied. The pressure drop can be achieved by various means, including gravity, vacuum, or direct pressure (32, p. 171). If the liquid is induced to flow through the medium by hydrostatic head, it is called gravity filtration. If higher than atmospheric pressure is applied upstream from the filter, the process is called pressure filtration; if lower than atmospheric pressure is applied downstream from the filter, it is referred to as vacuum filtration (20, p. 1473).

There are two types of filtration systems available: surface filters and depth filters. Surface filters are used for cake filtration, where the solids are deposited in the form of a cake on the surface of a relatively thin filter medium. Depth filters are used for deep-bed filtration, where particle deposition takes place within the medium (32, p. 173). Surface filtration is the process in common use by the minerals industry and is the only one discussed in this report.

At the commencement of surface filtration, the slurry particles that are the same size as or larger than the openings of the filter medium are held at these openings and create smaller passages, which remove even smaller particles from the slurry. A filter cake is formed which, in turn, functions as a more efficient filter for the subsequent filtration (32, p. 172).

In practice, solids are deposited on the filter medium as a cake, which is removed from the filter medium by a mechanical method such as scraping. After the cake is removed, the filter is cleaned by spray washing, then put into position to receive more solids. The performance of filter equipment is usually expressed as the yield in pounds of dry solids per square foot of filter area per hour (24).

The filter medium is probably the single most important component of a filter that affects equipment performance. Many different materials are now available as filter media. Filter cloth can be composed of duck, chain, and twill weaves, as well as felt made of cotton, nylon, polyester, polypropylene, and other natural or synthetic materials. Filter cloths and screens made of steel, stainless steel, and other alloys are also on the market (29).

Filter paper is often required as the filter medium to obtain high solids retention. Filter paper requires the support of filter cloth, screens, or perforated metal sheets to prevent its breaking. The filter paper can be easily removed from the support filter medium (29).

Covering the surface of the filter cloth or filter paper with a filter aid is known as a precoat and is often required to prevent blinding or clogging the filter medium. Materials such as diatomaceous earth, paper pulp, or perlite can be very effective. A filter aid can also be added to the slurry as a body feed to produce a filter cake that remains relatively permeable during the filtration cycle and gives a good overall filtration rate (29). Various combinations of filter cloth, filter paper, and precoating are capable of filtering out particles as small as 0.0002 in in diameter (31).

The following sections discuss the filtration equipment commonly used today.

FILTER PRESSES

Filter presses are batch units that use higher than atmospheric pressures and a series of filter plates to separate the solids from the liquid. They are also called pressure, plate-and-frame, or press filters. Although this type of device is well known, it has had little attention or application in the U.S. minerals industry because of its batch process nature, high capital cost, and associated high labor requirements. In spite of these drawbacks, the filter press should receive close attention because of its separation efficiency (24).

There are two types of filter press systems available--the high-pressure system and the low-pressure system. The first type uses a pressure of up to 225 lb/in^2, while the second type uses a maximum pressure of 100 lb/in^2. Research has shown that the higher filter pressure does not provide a significant benefit over the lower pressure unit with respect to dryer filter cakes or shorter cycles on chemically conditioned slurries (22).

Pressure filtration is a fairly simple process. Initially, the slurry is chemically treated and then pumped into the press (fig. 15). The filtrate flow from a large press can be 10,000 to 15,000 gal/h, so that this part of the cycle is often called the fast fill. During this phase, the cake chambers of the filter press collect the major amount of slurry solids. As the chambers become progressively filled with solids, the pressure inside the press rises and the filtrate flow rapidly decreases. This portion of the cycle is the cake consolidation

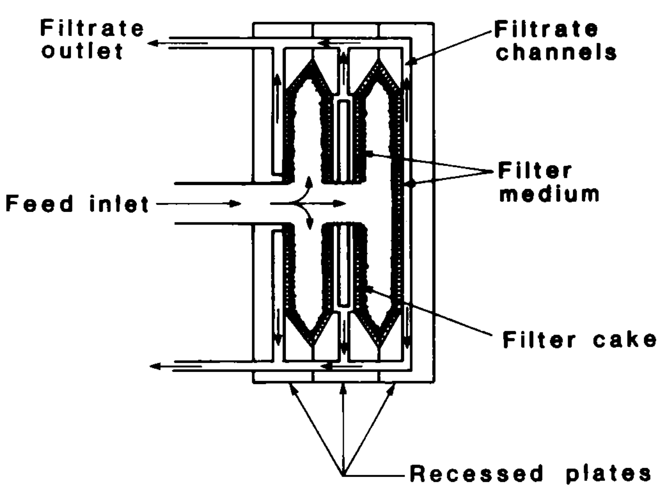


FIGURE 15. - Cross section of recessed-plate filter press (9, 22).

phase, where slurry solids are being forced under pressure into the cake chambers. This compacting action displaces more water from the loosely formed slurry cake and enables the press to produce harder and drier cakes than other means of dewatering (6, 22, 24).

At a predetermined low-flow condition, pumping is stopped and the feed holes that carried the unconsolidated sludge into the press are blown clear. The stack of filter plates is unclamped, and the individual plates are mechanically separated, allowing the filter cake between adjacent plates to drop out of the chambers. The cakes then fall against breaker bars that reduce them to a convenient size before they are transported for final disposal (6, 9).

Although pressure filtration is a batch process, a series of presses can be organized to work in a semicontinuous operation. This is done by having one press being filled, while a second is being airblown and a third is being unloaded or waiting to be filled again (25).

Filter presses find wide application in the process industries for the separation of slow-settling solids from liquids when the solids content is 1 to 10 pct and the filterability is poor. The use of these presses depends mainly on the particle size and on the quantity of solids in the slurry feed (32, p. 211). The capacity of the press is dependent on such factors as the characteristics of the material being filtered, operating temperature, pressure, filter medium being used, and type and size of filter press. There is no mathematical formula or scientific method for determining the exact filtration capabilities or rates. The most reliable method is to use past experience or records and to make controlled tests on laboratory-sized equipment (29).

ADVANTAGES AND DISADVANTAGES OF FILTER PRESSES

Filter presses are efficient at dewatering very fine grained solids such as slimes and consume only moderate amounts of power during their operation. Labor requirements are high, however, and the batch operation may be a limitation in some facilities. Chemical pretreatment is frequently needed to obtain the necessary solids capture. Also, large space requirements and the possible need to treat the decant liquors may restrict use (24). Despite these problems, a well-designed system of multiple filter presses might be the answer to difficult slime dewatering problems.

CONTINUOUS PRESSURE FILTERS

Although they are not well known, some pressure filters are available that work continuously. These units operate similarly to a drum or disk vacuum filter, except that the entire mechanism is enclosed in a pressurized tank. The cake is removed from the filter medium by a blade under superatmospheric pressure. The dislodged cake is throttled out of the tank to atmospheric pressure by a self-sealing screw conveyor or a series of receivers and pressure locks (23, p. 19-75).

The operating pressure in these machines is normally 100 lb/in^2 with a pressure drop of 40 lb/in^2 across the filter medium. Filter areas from 4 to 700 ft^2 are available. The cost of continuous pressure filters can be two to four times that of a vacuum filtration unit having the same filter area (23, p. 19-75).

Continuous pressure filters are advantageous because they work continuously and do not require close operator supervision. These units are also capable of much higher filtration rates for low-compressibility solids than are vacuum filters. There are drawbacks, too, because these machines are mechanically complex, do not allow access for any maintenance during operation, and have problems with lubrication. As mentioned before, these units are more expensive than similarly sized vacuum filters (23, p. 19-75).

BELT FILTER PRESSES

Belt filter presses have been manufactured in Europe since the midsixties and have been used mostly in pulp or paper plants and for sewage sludge dewatering. Only in the last several years have these machines been used by the minerals industry. Experimental work with the belt press indicates that this machine produces a drier cake with lower polymer consumption than centrifuges or similar dewatering devices. The belt press cake also has a relatively high shear strength, which is important if the cake is to be transported to a waste dump (28).

The belt filtration process is composed of three operational stages: chemical conditioning of the feed slurry, gravity drainage to a nonfluid consistency, and compaction of the dewatered solids. Figure 16 depicts a simple belt press and shows the location of the three stages. Although present-day belt presses may be very complex pieces of equipment, they follow the same concepts indicated in this figure. Good chemical conditioning is of primary importance to successful and consistent dewatering by the belt filter press. The recent developments in high-grade polymeric reagents have encouraged the development of belt presses to their present performance levels (19).

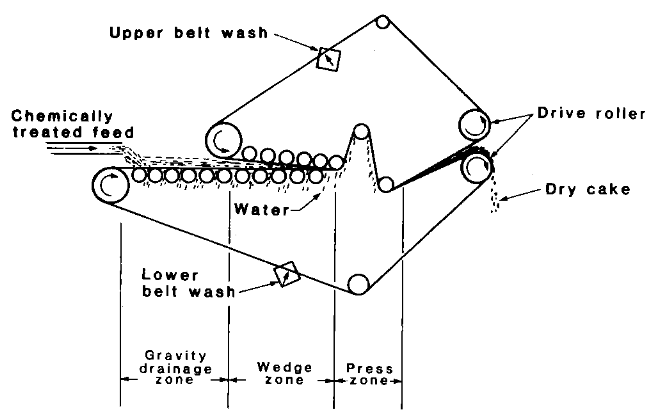


FIGURE 16. - Cross section of belt filter press (28).

The first stage in belt press dewatering is to add a flocculating agent to the slurry in a conditioning zone (12). The resultant mixture must be allowed to set under low shear conditions to allow flocculation to take place. If this mixture is disturbed during flocculation, more chemical agents may be needed to obtain the necessary degree of flocculation and any economic advantage will be lost. The conditioning zone should have an adjustable residence time which can be increased or decreased to suit the slurry being treated. Correct conditioning liberates the free water originally contained in the slurry and also facilitates the capture of the finest particles present by causing them to flocculate with the larger particles. This is necessary since water removal will be carried out on belts whose openings are larger than the smallest slurry particles (19).

The next stage after chemical conditioning is gravity dewatering, which allows the free water to drain from the solid particles. The slope of the belt will normally be set at 5° to 15° from horizontal to assure proper distribution of the solids across the belt as the water is separated from the slurry. At the end of the drainage section, the slurry will lose about 50 pct of its free water. At this time, the formation of an even surface cake is essential for the successful completion of the following stage in the cycle. The even surface prevents uneven belt tension or distortion, and the rigidity of the semicake enhances distribution through the process (19).

Following the gravity dewatering stage, the semicake is sandwiched between the carrying belt below and the covering belt above. The semicake is squeezed between the two belts and subjected to flexing in opposite directions as it passes around various rollers. This action increases water release and allows greater compaction of the cake (19). At least one manufacturer has developed a high-performance belt press that has a low and a high belt pressure stage. This machine is basically the same as other belt presses except that in the high-pressure stage, an additional belt is added to increase the pressure on the solids cake and remove more water (12).

Finally, the dried cake passes from between the two belts and is carried away from the machine by truck or conveyor belt for ultimate disposal (19).

The material and construction of the filter belts themselves are very important. Belt filter presses do not utilize the belt fabric as a filter medium but as a support for the self-filtering solid-liquid mass. The mesh size of the belt fabric is not critical because very large flocs will be produced by the polyelectrolytes in the first stage of operation. Texture, however, is a significant feature of the belts because it influences the adhesion of the cake to the belt and the physical strength of the belt. Currently, the available belts are those used in the papermaking industry, although a wider variety of special-purpose woven belts are becoming available. Most belts are woven from polyester monofilament, and many weaving patterns are available (19).

A major problem with the belt press is that an uneven layer of sludge causes the belt to deviate from the desired tracking line. This occurs because the filter belts usually have short lengths and widths. Direct side pressure should not be used because guide rollers would cause serious wear on the edges of the belt. One system that has been devised uses a delicately balanced sensor flap which rests lightly on the edge of the belt. This flap is attached to the spindle of a rotary directional fluid-flow control valve. Displacement of this valve causes hydraulic fluid to flow to either one or the other end of a cylinder which supports one end of the belt roller. The cylinder's movement causes the roller to slew across the line of travel of the belt, and belt tracking is induced rather than forced (19).

The openings in the belt are easily plugged by the solids particles so the belt must be washed continuously. The contamination of these perforations depends on the belt tension, pressure on the cake, complexity of belt weave, type of sludge being treated, and quantity and type of polyelectrolyte being used. Washing is done using water; however, the volume and pressure used vary from manufacturer to manufacturer. The current trend is to recycle the filtrate as belt washing water. This is satisfactory if the spray system is well designed to allow cleaning in the event of a blockage. At low solids concentrations the volume of filtrate produced will be adequate for belt washing.

ADVANTAGES AND DISADVANTAGES OF BELT FILTER PRESSES

Belt filter presses have several advantages, which include--

1. Continuous operation.

2. Low power consumption, as low as 3.7 kW according to one manufacturer (12).

3. Low consumption of chemical additives; one manufacturer claims his unit uses only 25 to 50 pct that used by centrifuges for some applications (12).

The disadvantages of belt filter presses include--

1. Problems with proper belt tracking (19).

2. Shortened belt life if very coarse solids are dewatered.

3. Difficulty in dewatering slurries having a very low solids content (28).

VACUUM FILTRATION EQUIPMENT

Vacuum filtration equipment has been in use in this country for over 100 yr, and continuous vacuum machines represent the most prevalent type of filtration equipment seen in the mineral industry today. Their simple operation and need for a minimum of auxiliary equipment has made them popular for solid-liquid separation in a variety of industrial applications.

Continuous vacuum filters separate the solids and liquid in a slurry by placing the filter in the slurry and applying suction behind the filter so that the water and solids are drawn to it. The solids are collected on the filtration surface, while the water is drawn through the filter and separated from the solids. The particles accumulate to form a cake on the filtering surface that is gradually lifted from the slurry. The cake is removed from the filter by various means, and the filter is returned to the slurry to repeat the process (18, p. 12-51).

Vacuum filters must have several items of auxiliary equipment in order to function properly. An agitator is needed to keep the slurry solids in suspension until they are drawn to the filter medium (32, p. 226). A vacuum pump provides the subatmospheric pressure to draw the liquid through the filter. A receiver with a barometric leg is located between the filter and the vacuum pump and separates the liquid from air drawn through the filter. Last, a filtrate pump removes the liquid from the receiver and forces it to the disposal point or back into the circuit for reuse (19).

The following sections describe the filtration equipment found in the minerals industry.

DRUM VACUUM FILTERS

Drum vacuum filters are probably the most common type of continuous vacuum filters in use. They consist of a horizontal cylinder or drum whose circumference holds the filter medium. The drum rotates in a slurry tank, and a vacuum inside the drum draws the liquid through the unit, leaving the solids captured on the filter surface. Scrapers or other devices then remove the cake from the drum, and the medium reenters the tank, enabling operation to be continuous (24). Figure 17 shows the front and side views of a typical drum vacuum filter.

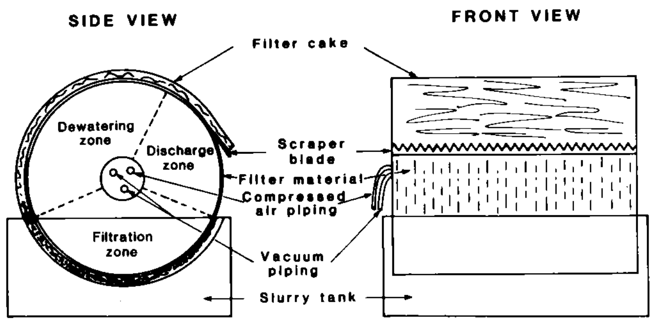


FIGURE 17. - Front and side views of drum vacuum filter (20).

Vacuum drums can handle throughputs up to 250 gal/min, particles as small as 0.0020 in, and solids loadings between 8 and 10 pct. The average yield is between 2 and 10 lb of dry solids per square foot per hour (24).

Drum vacuum filters have been widely used in industry because of their adaptability with respect to filter media and cake discharge methods. A variety of filter media can be easily used with a drum filter because the material itself is attached around the circumference of the cylinder. This configuration simplifies installation, inspection, maintenance, and removal of the filter medium. (24).

CAKE DISCHARGE METHODS

In addition to the variety of filter media available, there are many methods of removing the cake from the filter surface of a drum vacuum filter. The scraper-blade method of discharge is commonly used where the cake is friable and has poor mechanical qualities. A cake that is fairly thick and is not strongly held to the filter cloth could be easily dislodged by a blade. This blade could be steel, rubber coated, or made of other materials compatible with the particular corrosive or abrasive qualities of the filtrate or the solids. In many cases, a pressure reversal or blow is sufficient to loosen the cake from the filter medium, and the blade itself only guides the loose cake to the discharge chute (20, p. 1476).

When a precoat method is used, an advancing-knife mechanism must be precisely positioned to dislodge the cake and a minimum amount of the precoat material from the drum surface. This is an exacting procedure because the knife-advance system must not move more than several thousandths of an inch. The knife system has to be constructed so that its linear integrity, with respect to the drum face, is absolute and any deviations from the correct distance must be extremely small (20, p. 1477).

A roller discharge device may be used where a thin cake of sticky or thixotropic material is formed on the drum. This system uses a roller positioned close to the drum filter at the point of cake discharge. The cake is transferred from the filter to the roller and then removed from the roller by a cutting blade (20, p. 1476).

The belt discharge arrangement uses a filter cloth that winds around the drum to form the cake but leaves the drum to carry the solids to the discharged area. The cloth or belt passes over a small-diameter roller which causes the cake to separate from the filter. This system can be used for a cake having mechanical strength, for a thin cake, or where intensive washing of the filter is necessary to maintain the cloth openings. After the belt has passed over the discharge roll, it returns to the drum through a belt-washing system (20, p. 1476). Some of the various discharge methods are illustrated in figure 18.

EQUIPMENT MODIFICATIONS

Drum vacuum filters have been modified to separate solids and liquids under adverse conditions. For example, these machines can be fitted with simple hoods which limit the escape of poisonous or foul-smelling vapors. They can be adapted for complete sealing and for operation in a nitrogen environment; however, this complicates access to the internal parts,

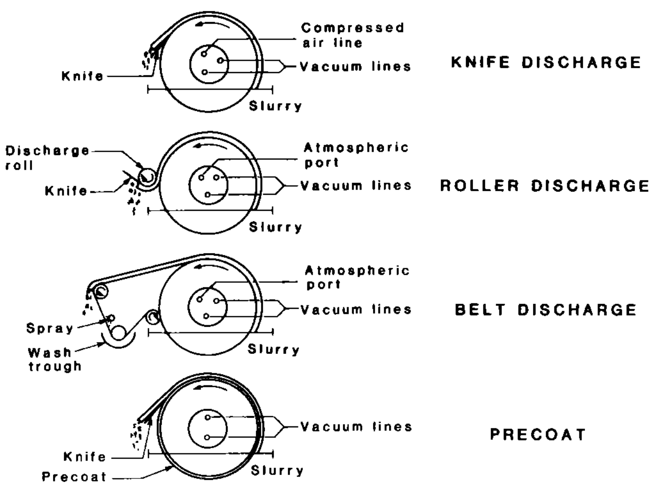


FIGURE 18. - Schematic drawings of various discharge methods for drum vacuum filters (20).

and the drum design must be such that the vacuum system and the cake receiving system are isolated to prevent vapor loss (20, p. 1481).

The drum vacuum filter can also be modified with accessories that improve the quality of the cake with respect to its washing and drying characteristics. This is possible because the cake moves through the washing and drying zones in the form of a continuous sheet and because the cake and filter medium are adequately supported on the drum shell. The filters can be fitted with simple rollers which extend the full width of the filter drum and can be arranged to eliminate irregularities or cracks in the cake prior to washing and drying. The wash waters and air are therefore applied to a uniform surface and will not short circuit or channel the deposited solids. The cake compression system may also have a "wash blanket" draped over the cake, which further limits any tendency for the air or wash water to channel the solids. These blankets allow the wash water to be used near the point where the cake emerges from the slurry (20, p. 1481).

Drum vacuum filters can also be designed so that the trough and hood are thermally insulated. If needed, the unit can be equipped with heat exchange equipment for heating or cooling the cake (20, p. 1481).

The basic design of the drum vacuum filter has undergone little change; however, with the development of new and improved construction materials, many options have become available. Drums can now be fabricated in a variety of metals, plastics, and rubber for handling corrosive materials. Internal piping and valves have also been improved by the use of corrosion- and abrasion-resistant materials, which greatly reduce shutdowns for repair of leaks or loss of vacuum. Tanks and agitators can also be fabricated of special materials, and tank linings have greatly reduced wear (20, p. 1475).

PERFORMANCE

For any given feed condition, the performance of drum vacuum filters can be optimized by the drum speed, the vacuum, and the percentage of drum surface submerged in the feed slurry. Most drum filters have controls for the manual adjustment of these variables, and some models have automatic adjustments which are actuated by changes in the quality and quantity of the feed or cake (20, p. 1480).

A drum filter should be operated with the greatest degree of submergence and at the highest drum speed in order to maximize the throughput of solids. It should be remembered that any increase in the submergence reduces the proportion of the drum area available for washing and drying. Drum submergence above 40 pct necessitates the use of seals on the drum shaft where it passes through the trough. No matter what combinations of drum speed, vacuum, and submergence are used, the sum effect must produce a cake that can be completely and easily removed from the drum. If this cannot be achieved, the filter medium will quickly deteriorate and its life will be severely shortened (20, pp. 1480-1481).

ADVANTAGES AND DISADVANTAGES OF DRUM VACUUM FILTERS

The advantages of the rotary drum vacuum filter are--

1. Continuous operation, which results in low operating labor costs.

2. Many design and operational variations available for a wide range of suspensions of divergent nature.

3. Clean operation.

4. Low maintenance costs.

5. Effective washing and dewatering (32, p. 226).

6. Provides a filtrate with a low suspended solids concentration.

7. Does not require skilled personnel.

8. Has low maintenance requirements for continuously operating equipment (19).

The disadvantages of the rotary drum vacuum filter are--

1. High capital cost.

2. Limitations imposed by vapor pressure of hot or volatile liquids.

3. Incapable of handling products that form explosive or inflammable gases under vacuum.

4. Unsuitable for quick-settling slurries.

5. Tendency for cloth blinding due to thin cakes and short cycles, although this may be alleviated by the application of a belt or string discharge (32, p. 227).

6. Auxiliary equipment such as vacuum pumps is very loud.

7. Consumes the largest amount of energy per unit of slurry dewatered, in most applications.

8. Produces wetter cakes if blowback is used, and greater filter medium wear if blowback is used in conjunction with a scraper knife (32, p. 227).

ROTARY DISK VACUUM FILTERS

The principle of operation for the rotary disk vacuum filter is the same as that for the rotary drum filter. The disks are oriented in a vertical plane and are composed of several pie-shaped sectors which fit into a central pipe for support of the disks and for transport of the filtrate. These sectors can be removed without disturbing the others in the same disk, and at slow speeds, it is possible to change a sector while the filter is still in operation. The filter medium cloth can be slipped over each sector and fastened at the innermost end of each sector (20). As the disks rotate, they go through pickup and dewatering operations similar to those carried out on the drum filters. At the discharge point, the cake is removed by means of wires or knives. Figure 19 shows two simplified views of a disk filter. Disk vacuum filters are available with areas from 0.5 ft^2 to approximately 3,300 ft^2; for large areas, as many as 12 disks are used in a single unit (32, p. 229).

Owing to the vertical disk orientation, the wide variety of discharge methods used on drum filters cannot be applied to disk filters. Thus, their application is somewhat more limited than that of the drum filters, but a disk filter will occupy only one-third the floorspace of a drum filter having the same filter medium area and is less expensive (20, p. 1477).

Disk filter provide more efficient agitation than drum filter agitators. In some applications a disk filter may be more cost efficient (18, p. 12-63).

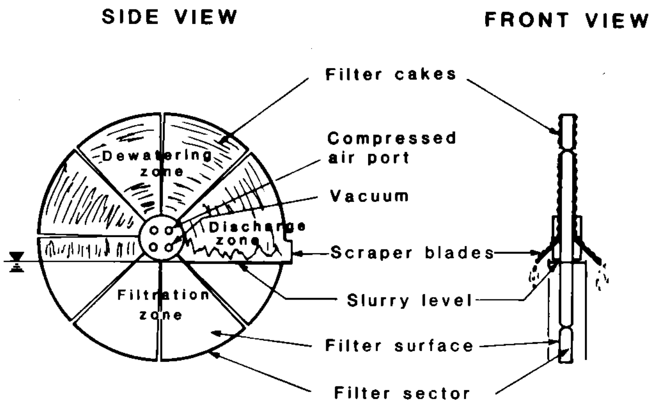


FIGURE 19. - Front and side views of disk vacuum filter (13).

ADVANTAGES AND DISADVANTAGES OF ROTARY DISK VACUUM FILTERS

The advantages of using a rotary disk vacuum filter are--

1. Low capital costs per unit area.

2. Large filter areas with minimum floorspace.

3. Rapid medium replacement (32, p. 229).

The disadvantages of the rotary disk vacuum filtration system are--

1. Difficulties in washing the cake.

2. Difficulties in discharging very thin cakes.

3. Inflexible operation.

4. High rate of medium wear with a scraper discharge.

5. Unsuitability for noncoherent cakes (32, p. 229).

HORIZONTAL CONTINUOUS VACUUM FILTERS

These vacuum filters use a horizontal filter surface in the form of a table, a belt, or multiple pans in a circular arrangement. The operating principle of the horizontal filter is the same as that of the rotary drum or disk filter, where the liquid in the slurry feed passes through a filter, leaving the solids deposited on top of the media. These units can be characterized as flat-bed, high-capacity filters which lend themselves to granular, fast-filtering materials and high-specific-gravity concentrates (20, p. 1478).

ROTARY TABLE VACUUM FILTERS

The rotary table units have a circular shape, the filter medium and supports rotate about a central axis. The feed slurry is deposited along the radius of the unit and rotates while it is being subjected to vacuum dewatering, washing, and drying, finally being removed from the filter medium by mechanical means (20, p. 1478). Figure 20 shows a cross section of a typical horizontal table filter.

The rotary table machines permit a choice of cake thickness, washing time, and drying cycle where 4- to 5-in-thick cakes can often be handled. Sharp separations between countercurrent wash waters are also possible because of the horizontal drainage configuration. Rotary table filters, though used in industry, are better suited for dewatering free-draining solids and not sticky or thixotropic slimes because of the difficulties in removing the solids from the filter medium (20, p. 1478).

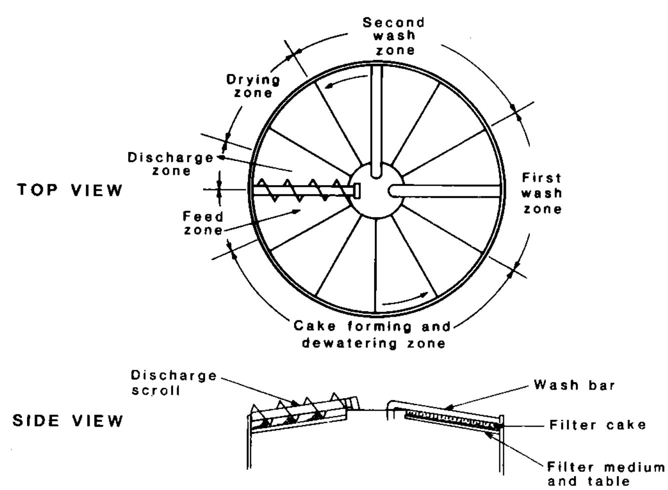


FIGURE 20. - Plan and cross-section views of horizontal rotary vacuum filter (20).

HORIZONTAL BELT VACUUM FILTERS

The horizontal belt vacuum filter uses an endless belt of filter fabric supported by a slotted or perforated endless belt. Both belts travel over one or more vacuum zones. The slurry is deposited onto the filter at one end, wash water is applied at one or more points along the path of belt travel, and the cake is dumped at the other end. The support belt and the filter are parted and directed along separate lines of pulleys. The filter is washed and rejoins the support belt just ahead of the slurry deposition point. Figure 21 shows a cross section of a typical horizontal belt vacuum filter (15; 20, p. 1478).

The horizontal belt machines have a high capacity per square foot of area under vacuum, similar to the horizontal rotaries. They are well adapted for a countercurrent discharge circuit and enable the cake to be flooded with wash solvent so that it can be steeped in the wash liquid. They are suited for countercurrent leaching or washing (20, p. 1478).

Horizontal belt filters have been manufactured for more than 30 yr, but recently have they been used for large-tonnage applications that require filter cake washing. For example, the largest unit available in 1950 had only 40 ft^2 of active filter area, while modern machines have over 900 ft^2 of filter area. Many mechanical improvements have also been made on these machines, which have improved their reliability. The biggest single improvement, though, was the ability to manufacture a continuous drainage belt to very close tolerances, which spurred the development of large-capacity units (20, p. 1479).

The horizontal filter shows several advantages because the slurry deposition on the horizontal filter belt eliminates the constant slurry agitation necessary for rotary types. In the horizontal filter, the cake travels as a ribbon of unwashed cake which is gradually washed to the required purity, either concurrently or countercurrently (19).

Another advantage of the horizontal machine is its ability to resist filter medium blinding. Most slurries contain, in addition to some medium-sized material, a fraction of fines. No matter how efficient the agitation with the rotary drum, these fines will always be concentrated near the surface within the trough. Consequently, the fine material is sucked against the filter cloth of the rotary drum filter and causes blinding before the large particles reach the cloth. This does not happen on horizontal belt filters because the biggest particles reach the filter cloth first,

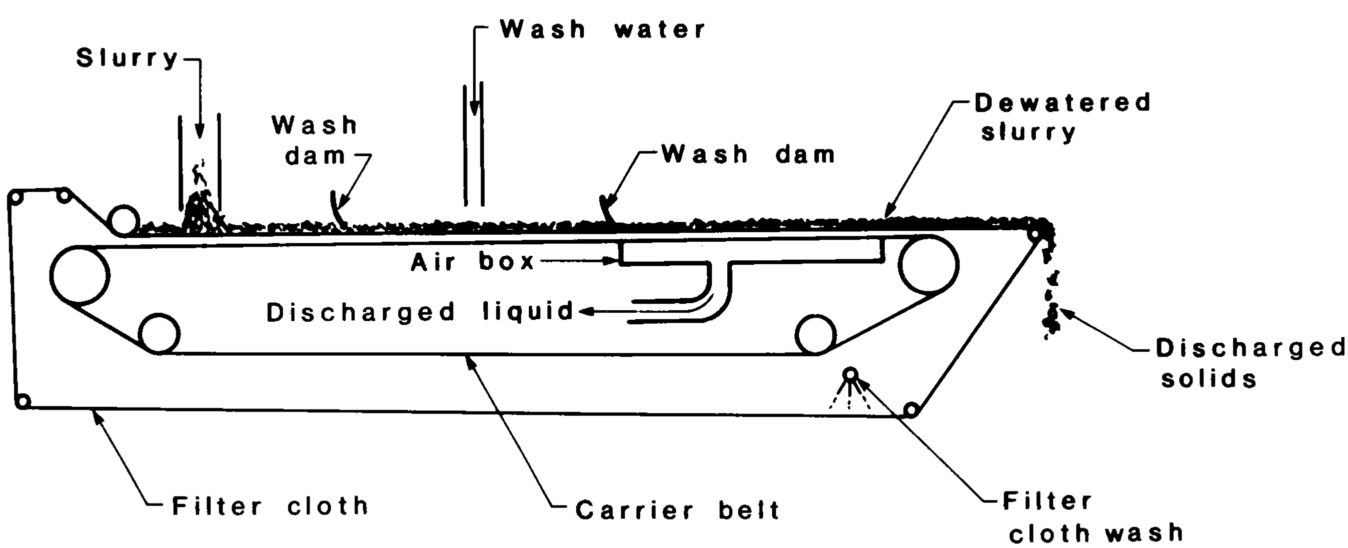


FIGURE 21. - Cross section of typical horizontal belt vacuum filter (15, 20).

followed by the finer particles. As a result, the larger particles act as a precoat for the finer solids (19).

Another concern with the rotary drum and belt filters involves maintaining a vacuum. Many cakes, after being dewatered, shrink and crack. This allows air to pass freely through the cake, which reduces the vacuum and the dewatering effect on the cake. On the rotary drum filter, this is remedied by using expensive blankets, rakes, or squeeze rollers. On the belt filter, a simple sheet of impervious material such as polyethylene trailed over the cake is usually adequate to maintain the maximum vacuum (19).

ADVANTAGES AND DISADVANTAGES OF HORIZONTAL VACUUM FILTERS

The advantages of the horizontal vacuum filters are--

1. Excellent wash capability.

2. Flexible operation.

3. High-volume operation for fast-settling solids (32, p. 230).

The disadvantages of the horizontal vacuum filters are--

1. Requirements for large floorspace.

2. High initial costs.

3. Unsuitability for slow settling solids (32, p. 230).

SELECTING AND SIZING FILTRATION EQUIPMENT

In the selection of filtration equipment, the job requirements must be compared to those associated with the equipment characteristics. Job-related factors include slurry character, production magnitude, process conditions, performance requirements, and permissible materials of construction. The equipment-related factors are the type of cycle (batch or continuous), driving force (gravity, pressure, or vacuum), production rates of largest and smallest units, separation sharpness, washing capability, dependability, feasible materials of construction, and costs. This last item must include depreciation (installed cost plus expected equipment life), maintenance, operating cost (labor, services, and filter media), and penalty of product loss (if any). In addition, consideration must be given to preconditioning and the use of filter aids (20, p. 1485). The suitability of the most common types of filters for various classes of slurries is summarized in table 1.

Continuous filters are the most desirable when the process to which they contribute is a steady-level, continuous one; however, the rate at which the cake forms and the magnitude of production rate will probably be the critical factors. For example, the use of a rotary vacuum filter is not practical if a 0.1-in-thick cake will not form under normal vacuum in less than 5 min and if more than 50 ft^3/h of wet cake is to be produced. The production use of batch filters is harder to define, although they have been used in some processes that turn out 200 ton/d of dry solids. Occasionally, equipment flexibility and high filtering pressures will become more important than other factors that would otherwise dictate continuous equipment. Small-scale tests are essential for estimating the filtration rate, the washing characteristics, and other important features. Filtration is essentially an art rather than a science, and experience with the various aspects of vacuum filtration will help in better approaching the selection of equipment and evaluating test results (20, p. 1485).

TABLE 1. - Classification of selected vacuum filters (20, p. 1481)

(Performance index: 1 = very poor or negligible; 9 = highest possible performance)

Type	Maximum area, ft²	Slurry classification[1]					Relative performance		
		A	B	C	D	E	Cake dryness	Cake washing	Filtrate clarity
Horizontal belt[2]	900	X	X	-	-	-	5-8	7-8	6
Horizontal rotary table[2]	160	X	X	-	-	-	4-7	8-9	7
Rotary drum:									
Knife discharge[3]	860	-	X	X	-	-	5-8	6	8
Roller discharge[3]	860	-	-	X	X	-	5-6	5	8
Belt discharge[3]	860	-	-	X	X	X	5-8	6	7
Precoat[4]	860	-	-	-	X	X	NA	6	9
Rotary disk[5]	3,230	-	X	X	-	-	2-3	1	6

NA Not available.

[1]A--High solids concentration (>20 pct), free draining, fast settling, high filtration rates.

B--Rapid cake formation, reasonably fast settling solids.

C--Lower solids concentration, slow thin cake formation, difficult to discharge.

D--Low solids concentration, slow cake formation, very poor strength properties.

E--Very low solids concentration, solids usually blind normal filter media.

[2]For free-draining materials where good washing is necessary.

[3]Wide range of types and sizes. Generally suitable for most slurries of types B and/or C. Can be fitted with various devices to improve washing and cake drying.

[4]Suitable for slurries that blind most filter media.

[5]Large throughput for small floorspace.

LABORATORY TESTING

Extensive research has been done to develop laboratory procedures for determining the filterability of slurries. The filter-leaf test is the commonly used method of estimating the necessary filter area for a particular slurry. In this method, a test leaf is used which is covered with a filter medium identical to that intended for the full-scale filter. Figure 22 shows the typical apparatus

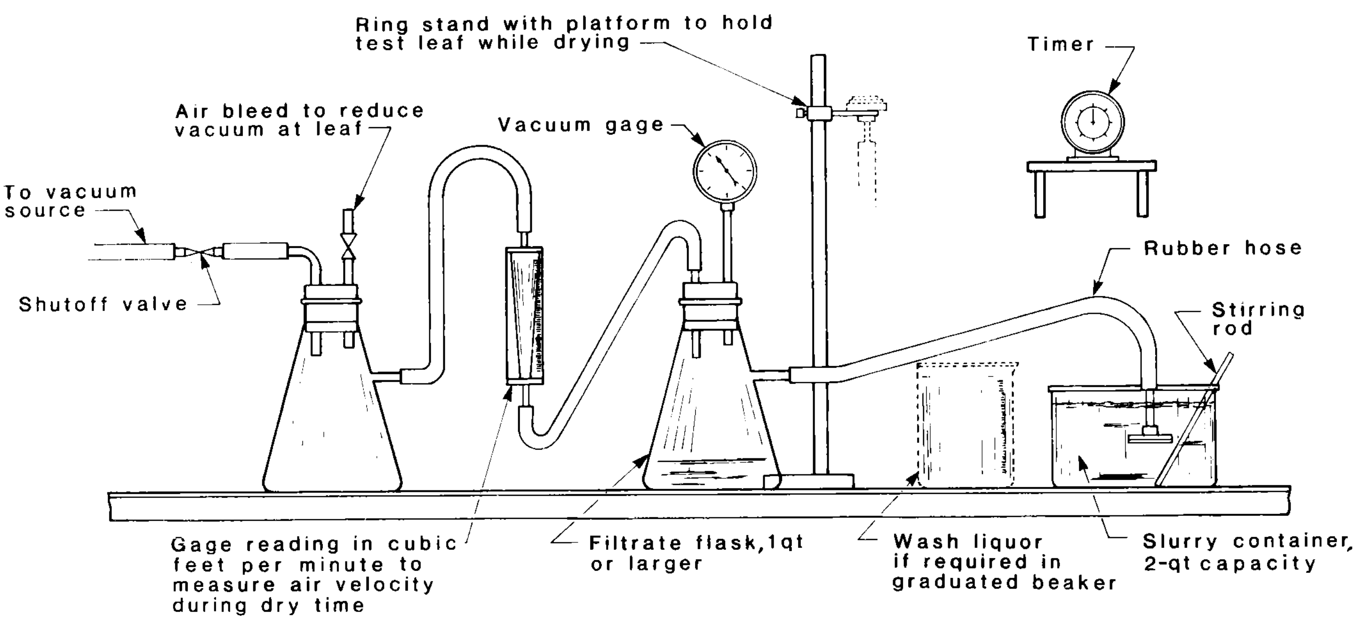


FIGURE 22. - Typical laboratory installation for vacuum leaf tests (20).

needed for a leaf test. The procedure for the test follows:

1. Condition 2 qt of sludge for filtration. The sludge should be thickened to the same concentration as the production slurry.

2. Apply the desired vacuum to the filter leaf and immerse in the sample for 1-1/2 min while maintaining sample agitation. The test leaf is usually inserted upside-down in the slurry to simulate the cake formation zone of a drum filter.

3. Bring the leaf to a vertical position and allow it to dry under vacuum for 3 min. This simulates the cake draining and drying part of the cycle.

4. Blow off the cake for 1-1/2 min, which gives a total drum cycle of 6 min. To discharge the cake, the leaf is disconnected from the vacuum, and air pressure of not more than 2 lb/in^2 is applied.

5. Dry and weigh cake to determine percentage moisture, which can be computed from the equation--

$$f_{cv} = \frac{W_{ds}\ t_{cv}}{A_{tl}}, \tag{18}$$

where f_{cv} = the filter cake formation rate, pounds per square foot per hour,

W_{ds} = the dry weight of the solids cake, pounds,

t_{cv} = the cycle time, hours,

and A_{tl} = the filter test leaf area, square feet (2, p. 11-29).

The test can be easily modified for other cycle times or discharge mechanisms. Filter leaves and testing instructions are available from most filter manufacturers. It may be necessary to adjust the results obtained by a factor to compensate for partial medium blinding and for scaling over a long period of operation. The test results will provide filtration parameters for the cake formation, drying, and washing portions of the filtration cycle. The filter-leaf test is easy to perform; however, several precautions should be observed to assure accurate results:

1. Representative slurry samples should be used.

2. The test should be repeated 5 to 10 times to observe any filter medium blinding.

3. The sample must be continually agitated to assure that it is homogeneous.

4. The vacuum must be regulated so that it does not vary during the test. The vacuum should be the same as that intended for use in the full-scale operation (2, p. 11-29).

After the tests have been completed, the results can be analyzed. A graph of the moisture content of the filter-leaf test cakes versus a correlating factor should be constructed. The correlating factor is calculated from--

$$F_v = V_{av}\ t_{dv}\left(\frac{P_{dc}}{W_{dst}}\right), \tag{19}$$

where F_v = the filter cake correlating factor, dimensionless,

V_{av} = the volume of airflow through the cake per unit area of filtering surface, cubic feet per minute per square foot,

t_{dv} = the drying time, minutes,

P_{dc} = the pressure differential, pounds per square inch, gage,

and W_{dst} = the weight of the dry cake solids for a given cake thickness, pounds per square foot (2; 18, p. 11-31).

A decreasing moisture correlation indicates that the moisture content decreases; and as the air rate through the cake per unit of filtering area is increased, the vacuum differential, or the length of the drying time, is increased. On the other hand, if the cake thickness and the cake weight are increased, the moisture content increases. Knowing the percentage of available drying time of the filter cycle and using the design formation, such as the proper cake thickness for a given type of filter, the vacuum level, and the airflow rate through the cake, it is possible to predict for each cycle time the discharged filter cake moisture content expected from the full-scale filter. The filter area provided in the design should be for the maximum solids removal rate plus a 5- to 15-pct safety factor (2, p. 11-31).

FACTORS AFFECTING FILTRATION

Efficient vacuum filtration is influenced by many variables, of which some can be controlled and others cannot. The following items represent many of the factors that affect the final moisture content of a filter product:

1. Cake thickness.

2. Pressure drop across cake.

3. Drying time.

4. Volume of air or gas per minute per square foot of filtering area.

5. Viscosity of filtrate.

6. Surface tension of filtrate.

7. Filter medium.

8. Size distribution of solids.

9. Permeability of cake.

10. Specific gravity of dry solids.

11. Inherent moisture of dry solids.

12. Surface properties and other characteristics of solids.

13. Type of filter and construction.

14. Homogeneity of cake formation.

15. Temperature of solids and gas (18, p. 12-53).

Two other conditions--feed solids concentration and cycle time--are important in vacuum filtration because they influence many of the above factors.

The feed solids concentration is very important in the filtration process; consequently, a thickening device often precedes the filter to ensure that the feed solids concentration is consistent with economic and efficient operation. A general plot of dry cake output versus feed solids concentration reveals a curve, as shown in figure 23 (18, p. 12-54).

Each slurry has its own characteristic filtration curve, which must be experimentally determined. In this example, the slurry exhibits a sharp incremental rate increase above 35 pct solids. Controlling the solids concentration between the limits of 45 and 55 pct solids

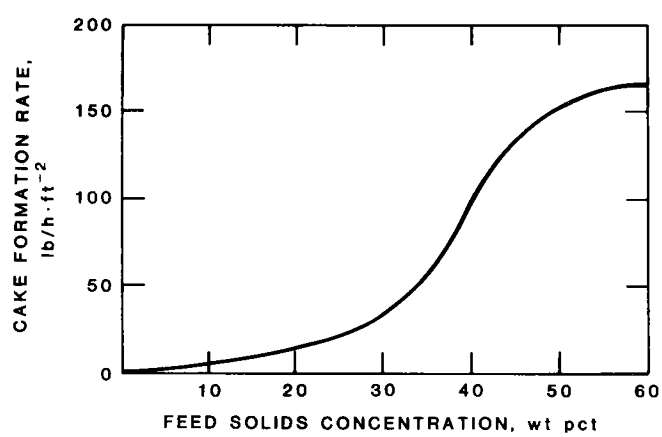


FIGURE 23. - Representative curve for cake formation rate versus feed solids concentration (18).

will require less filtration area and the resulting filter operating costs will be reduced. Above 58 pct solids, this slurry becomes relatively viscous and transportation to the filter will be difficult. There is a point of inflection at about 55 pct solids where the curve becomes asymptotic. This indicates that further slurry thickening above 55 pct solids is impractical and uneconomical because it produces only a slight increase in the filtration rate (18, pp. 12-54 to 12-55).

The other important consideration for filtration operations is the cycle time. In this discussion, cycle time will be concerned with rotary vacuum filters, although the same principles apply to other types of filters.

The cycle time of a continuous vacuum filter is the amount of time the filter takes to make one complete revolution and is given in terms of minutes per revolution. During each cycle, there are three phases of filter operation: cake formation, cake dewatering, and cake discharge. At the end of each cycle, the filter discharges a certain amount of cake per given filter area. With these data, the dry cake formation rate can be expressed in pounds per hour per square foot of filtering area. A log-log plot of dry cake formation rate versus cycle time for an arbitrary feed solids concentration is shown in figure 24 (18, p. 12-55).

For this example, the slope of the curve is -0.5, based on an assumption that solids concentration and cake compressibility remain constant. This relationship can be mathematically expressed as--

$$f_{cvn} = f_{cvo}\left(\frac{t_{cvo}}{t_{cvn}}\right), \qquad (20)$$

where f_{cvo} = the old cake formation rate, pounds per square foot per minute,

f_{cvn} = the new cake formation rate, pounds per square foot per minute,

t_{cvo} = the old cycle time, minutes,

and t_{cvn} = the new cycle time, minutes (18, p. 12-55).

Vacuum filters are normally equipped with variable-speed drives operating within a range of 1.5 to 9.0 min/rev. Thus, for any filter area, cake output can be doubled, tripled, or halved, as the situation requires (18, p. 12-55).

Cycle time also affects the filter cake moisture content and dischargeability. As a general rule, the filters should be sized for a cycle time of at least 3 min/rev and preferably 4 min/rev. Cycle times less than 3 min/rev will increase the cake moisture and produce thin cakes, which are difficult to remove from the filter medium. Difficult cake discharge can mean sizable increases in filter maintenance costs (18, pp. 12-55 to 12-56).

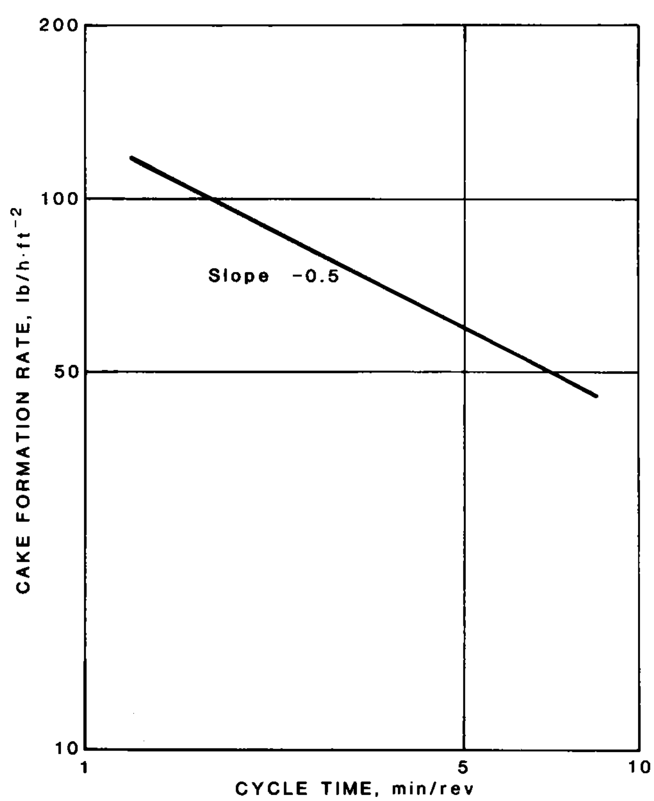


FIGURE 24. - Representative curve for cake formation rate versus cycle time (18).

HYDROCYCLONES

BACKGROUND

Cyclones have found wide use in industry for solids classification and concentration. The cyclones designed for liquids are called hydrocyclones, hydraulic cyclones, or hydroclones. The basic separation principle used in cyclones is centrifugal sedimentation, where the suspended particles are subjected to a centrifugal force which causes them to separate from the fluid. Cyclones have no moving parts, and the necessary vortex motion is provided by the fluid itself (32, p. 101).

Hydrocyclones, by themselves, are not capable of producing a dry solids product because they use liquids for their operation. Cyclones are useful, however, in concentrating the solids content of a slurry ahead of another dewatering machine, such as a filter or centrifuge.

The cyclone consists of a short cylindrical section attached to an inverted, truncated, conical section. The apex or bottom of the conical section is called the underflow orifice. A central overflow orifice or vortex finder is fitted to the base of the cone, and a feed orifice is attached tangentially to the cylindrical body section. Figure 25 shows the cross section of a typical hydrocyclone. The slurry enters at high pressure through the tangential feed orifice into the cylindrical section, where a rotating force field is established. The solids in the slurry are settled to the side wall by this force, slide down the inclined wall to the apex of the cone, and exit through the underflow orifice. The liquid portion of the feed travels to the center of the cone with some of the finest solids and exits through the overflow orifice. A vortex of air extends throughout the length of the cyclone (18, p. 12-26).

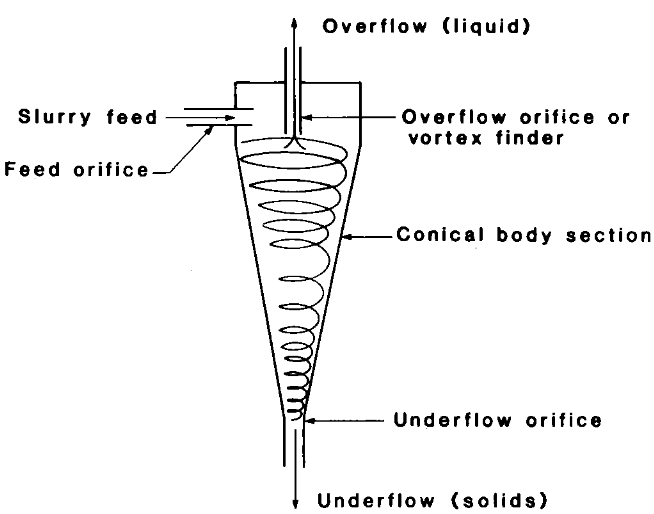


FIGURE 25. - Cross section of hydrocyclone (18).

A cyclone is a simply constructed device; however, its principles of operation are complex and there are many variables that must be carefully evaluated to produce the desired separations. The most significant variables are--

1. Cyclone diameter.

2. Cyclone cone angle.

3. Feed, overflow, and underflow orifice sizes.

4. Length of cylindrical section.

5. Feed pressure.

6. Feed concentration.

7. Particle size (18, pp. 12-26 to 12-27).

The hydrocyclone diameter is the most important factor influencing the application and efficiency of a cyclone because the smaller the particle to be dewatered, the smaller the cyclone diameter that must be used. For most applications, the cyclone manufacturer will determine the size and cone angle of the cyclone to be used (18, p. 12-27).

The orifice size is another important factor that influences cyclone performance. The underflow orifice will determine the concentration and flow of the thickened solids from the cyclone. Enlarging the underflow orifice increases the flow rate and the percentage of fines in the underflow. The larger underflow orifice also allows more liquid to pass through it and decreases the concentration of the product. A larger overflow orifice increases the total flow, the concentration of solids, and the maximum particle size contained in the overflow of the cone. A secondary effect is that the underflow of the cone increases in solids concentration and contains a larger percentage of coarse sizes (18, p. 12-27).

Changes to the feed orifice affect the volume processed by the cone because as the area of the feed orifice is increased, there is an accompanying proportional increase in the flow to the cyclone. The additional flow reduces the retention time of the slurry within the cone and causes the cyclone to reject coarser materials to the overflow. As a result, when a cyclone feed is increased, the cyclone underflow orifice should also be increased to accommodate the higher tonnage of material fed to the cyclone (18, p. 12-28).

The feed inlet can be either rectangular or circular; however, a rectangular inlet with its long side parallel with the axis of the cyclone produces better results. The top of the feed orifice should be flush with the top of the cyclone to eliminate a dead space which would cause short circuiting of the feed (32, p. 113).

In the mineral industry it is common practice to have several vortex finders with different diameters or nozzles which can be put into the exit pipe. This enables the operator to change the length of the vortex finder, when needed. An increase in the length of the vortex finder improves the efficiency of removal of the coarse particles but decreases the efficiency for the finer particles (32, p. 114).

The feed pressure affects the volume processed and the relative efficiency of the cyclone. The total flow to a cyclone will vary proportionally to the square root of the pressure. Increasing the feed pressure causes the underflow concentration to increase and become finer in size, while the overflow discharge increases and the overflow solids also become finer (18).

The interior surface of the cyclone should be as smooth as possible to promote good material flow. Abrasion resistance should be built into a cyclone if it is to be operated with abrasive solids. A wide range of construction materials, such as steel, nylon, ceramics, polyurethane, and rubber, are available (32, p. 114).

ADVANTAGES AND DISADVANTAGES OF HYDROCYCLONES

Hydrocyclones have been used for many solid-liquid separation applications because of the following advantages:

1. High capacity.

2. Simple operation.

3. Compact design, which uses a minimum of floorspace.

4. Relatively low capital costs.

5. Low maintenance and processing costs (8).

Hydrocyclones do have a drawback, however, because the smallest particles in the slurry will be carried away with the overflow. Proper cyclone design can minimize this loss, but clarification of this liquid may be necessary, and this possibility should be recognized by the mill operator.

SIZING HYDROCYCLONES

The designer of a hydrocyclone installation will be concerned with the size and number of cyclones needed. This will be based on the desired separation efficiency and flow rate. Several small-diameter cyclones working in parallel are more efficient than one large cyclone handling the same capacity (32, p. 114).

A simple, straightforward method of sizing hydrocyclones is to first determine the requirements of the installation in terms of particle size, amount of throughput, or pressure differential between the inlet and overflow orifices (32, p. 114). Knowing this information, manufacturers' bulletins can be consulted. Operational data for hydrocyclones are often presented in graphic form; figure 26 shows a plot of pressure drop versus throughput capacity for two representative hydrocyclones. As can be seen, these two cyclones are designed to handle very small solids.

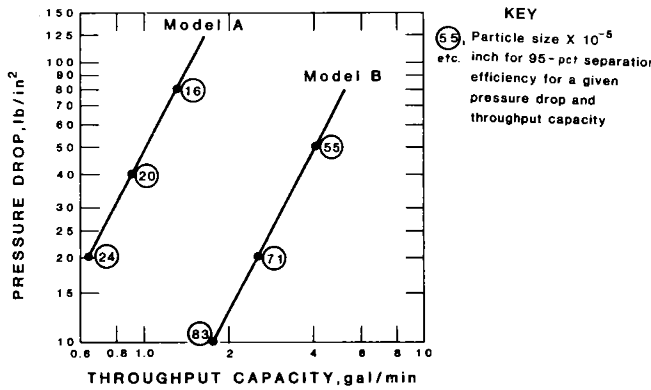


FIGURE 26. - Graph of pressure drop versus throughput capacity for two hydrocyclones (8).

More sophisticated methods can be used to determine the diameter, cone angle, and other dimensions for a particular situation. Their application, however, is very complex and a reference such as Svarovsky (32, pp. 106-118) should be consulted for a complete theoretical discussion.

THERMAL DEWATERING

BACKGROUND

The use of heat is another method of separating liquids from solids. Thermal processes can be used to dry a slurry or just the thickened solids; however, fuel consumption will increase proportionally with the increase in the moisture present. It is, therefore, cost effective to use thermal drying after a substantial amount of moisture has already been removed by other methods such as centrifugation or filtration.

There are basically three methods of heat transfer for drying:

1. Convection--the direct contact of particles with warm air.

2. Conduction--the direct contact of particles with a heated shell of a dryer or other heated particles.

3. Radiation--heat radiating from a hot surface to the particles (13, p. 27-75).

THERMAL DRYER OPERATION

Dryers used commercially for drying minerals commonly utilize heat transfer by convection or direct contact between the wet solids and hot air. Over the years, many different configurations of convection dryers have been available for the minerals industry. Now, six basic types exist, as follows:

1. Drum.

2. Suspension or flash.

3. Multilouvre.

4. Vertical tray.

5. Continuous carrier.

6. Fluidized bed (18, p. 13-7).

In each of these dryers, the wet solids are separated from each other and subjected to a flow of hot gases or air. In the drum-type dryer, the solids enter one end of a rotating cylinder through which the hot air is blowing. The solids are tumbled the length of the cylinder and exit the other end (13, p. 27-76).

In the suspension or flash dryer, the wet solids are carried upward through a vertical duct by a blast of superheated air for a very brief time. The air is usually heated to 1,200° F, but the solids are in contact with the hot air for only a fraction of a second and are not changed chemically (18, p. 13-39).

The louvered arrangement has a series of specially designed flights which carry the solids upward through a flow of hot air. The vertical tray, on the other hand, has a series of shelves placed in a terrace arrangement. The solids are fed in at the top, and a vibrating action causes them to fall from one shelf to the next through heated air. The continuous carrier type uses a vibrating inclined screen to support the solids as they tumble through the hot airstream (18, pp. 13-39 to 13-43).

The fluidized-bed dryer uses a slightly different approach and has seen wider use in the past two decades. In this type of dryer, the solids are fed into a heater box and subjected to a high-velocity flow of hot air. The violent bubbling action of the solids is similar to that of boiling water, leading to the description of the dryer as a fluidized bed (4). Figures 27 and 28 show simplified cross sections of the various thermal dryers.

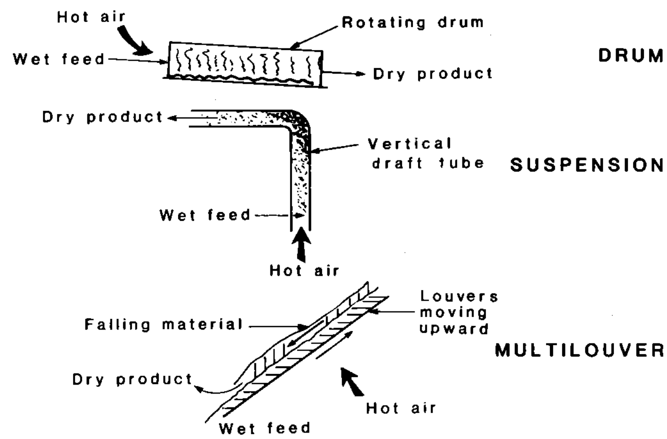


FIGURE 27. - Simplified cross sections showing operation of drum, suspension, and multilouver thermal dryers (18).

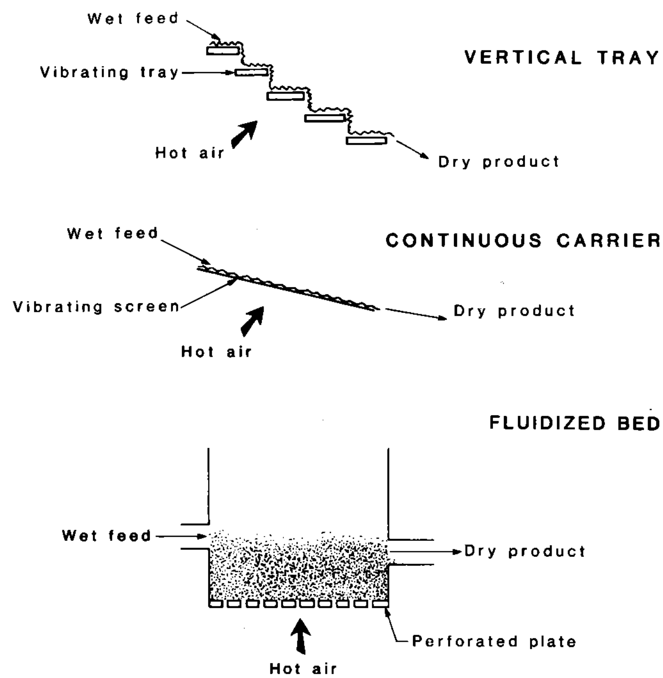


FIGURE 28. - Simplified cross sections showing operation of vertical tray, continuous carrier, and fluidized-bed thermal dryers (18).

ADVANTAGES AND DISADVANTAGES OF THERMAL DRYERS

Thermal dewatering equipment has several advantages, which include--

1. Ability to reduce the moisture content of slimes to 6 pct or less.

2. Minimal labor requirements.

3. Capability to operate continuously as long as feed material is available.

4. Low maintenance costs.

There are, however, several disadvantages, such as--

1. The moisture content of the product must be carefully monitored. If it is too low, the powdery product may be difficult to handle or transport (1).

2. The fuel demands can become very high (4).

3. A mechanical dewatering device is usually needed ahead of a thermal dryer.

SIZING THERMAL DRYERS

As wet solids are dried, they go through three stages with respect to moisture loss (13, p. 27-75):

1. Warming period.

2. Constant-rate period.

3. Falling-rate period.

The constant-rate and falling-rate periods have the greatest impact on the drying time required for a particular material. The warming period will be significant if radiant heat from surrounding surfaces is negligible (13, p. 27-75). The rate of liquid evaporation during the constant rate period can be estimated using mass transfer or heat transfer equations as follows:

$$E_c = M_a K_y (H_{ai} - H_a) A_d$$

(mass transfer), (21)

$$E_c = \frac{h_y (T_a - T_i) A_d}{\lambda_i}$$

(heat transfer), (22)

where E_c = the rate of evaporation, pounds per hour,

M_a = the molecular weight of air, pounds per pound-mole.

K_y = a mass transfer coefficient, pound-moles per square foot per hour,

H_a = the humidity of the ambient air, pounds of water per pound of dry air,

H_{ai} = the humidity of the air at the solid-air interface, pounds of water per pound of dry air,

A_d = the drying area, square feet,

and h_y = a heat transfer coefficient, Btu per square foot per hour per °F.

For airflow parallel with surface of solid, $h_y = (0.0128\ V_{am})^{0.8}$.

For airflow perpendicular to surface of solid, $h_y = (0.37\ V_{am})^{0.37}$.

V_{am} = the air mass velocity, pounds per square foot per hour,

T_a = the temperature of the ambient air, °F,

T_i = the temperature of the air at the solid-air interface, °F,

and λ_i = the latent heat of water at the temperature, T_i, of the solid-air interface, Btu per pound (13, p. 27-75).

Equations 21 and 22 are valid while the surface of each particle is saturated. When this condition is no longer the case, thermal drying enters the falling rate period (33, p. 17-02).

If the total average moisture content (X_{tav}) of any particle is composed of the average free moisture (X_{fav}) and chemically bound or equilibrium moisture (X_{eq}), the rate of liquid evaporation can then be estimated by the liquid diffusion equation, which is--

$$E_f = \frac{X_{tav} - X_{eq}}{X_{ti} - X_{eq}} = \frac{X_{fav}}{X_{fi}} = \frac{8}{\pi^2}\left(e^{-\alpha\beta} + \frac{1}{9} e^{-9\alpha\beta} + \ldots\right), \quad (23)$$

where $\alpha = \left(\frac{\pi}{2}\right)^2$,

$\beta = \frac{D_m t_n}{S^2}$,

E_f = the rate of liquid evaporation during the falling rate period, pounds of water per pound of dry solid,

X_{ti} = the initial total moisture content of the solids, pounds of water per pound of dry solid,

X_{tav} = the average total moisture content during the falling rate period, pounds of water per pound of dry solid,

X_{eq} = the equilibrium moisture content of the solids, pounds of water per pound of dry solid,

X_{fi} = the initial free moisture content at the beginning of the falling rate period, pounds of water per pound of dry solid,

X_{fav} = the average free moisture content during the falling rate period, pounds of water per pound of dry solid,

e = Euler's number - 2.71828,

D_m = the diffusivity value of moisture through a solid, square feet per hour,

t_n = any arbitrary time after the beginning of the falling rate period, hours,

and S = one-half the thickness of the layer of solids in the dryer, feet (13, p. 27-75).

These equations are helpful in estimating the rate of evaporation of moisture and will enable the designer to determine the total drying time, feed rate, air-volume requirements, and other important factors concerning thermal dryers.

When a fluidized-bed dryer is being considered, the pressure drop across the orifice plate must be evaluated and the plate size designed accordingly. For most applications, a pressure drop equal to 10 or 12 in of water is adequate. Pressure differences slightly higher than this should be used if the fluidizing gases are as hot as 1,200° F (18, p. 13-18).

The pressure drop will increase as the percent open area in the plate is decreased. To design for the percent open area, a factor called the discharge coefficient must be determined with the aid of a single-orifice test plate that has the same cross-sectional characteristics as the proposed full-scale dryer. Information obtained from the test plate can be used in the following equation to determine this coefficient:

$$C_d = \frac{q_a\, M_a\, P_{op}}{4825\, A_o\, P_{do}\, T_k}, \qquad (24)$$

where C_d = the discharge coefficient, dimensionless,

q_a = the airflow through the orifice plate for the air temperature and pressure used during the test, cubic feet per minute,

M_a = the molecular weight of air, 29,

P_{op} = the mean absolute pressure in the orifice, pounds per square inch,

A_o = orifice area, square feet,

P_{do} = the pressure differential across the orifice, inches of water,

and T_k = the temperature, kelvins (18, p. 13-17).

After the discharge coefficient has been determined for a particular orifice configuration, the volume of air needed to pass through the orifice to provide the necessary pressure drop can be calculated. The total number of orifices for the full-scale dryer can then be found if the designer knows the volume of air needed to maintain the pressure drop for one orifice and the total volume of air needed for drying the solids (18, p. 13-19).

These equations will provide approximate values for sizing thermal dryers, but as with other equipment, it is recommended that pilot-plant testing be done.

CURRENT BUREAU OF MINES RESEARCH ON DESLIMING METHODS

The bulk of this review has been concerned with well established methods of dewatering slimes; however, at least two Bureau of Mines innovations for dewatering slimes should also be brought to the attention of the mining industry: Electrokinetic methods and the rotary trommel.

ELECTROKINETIC METHODS

Background

The Bureau is currently developing a method of dewatering slimes using an electrokinetic potential. This dewatering technique is intended to be used at the place of the slimes disposal rather than in the mill or other processing location. This method takes advantage of the electrical surface charge on the solid particles in a water suspension. Bureau of Mines research has shown this technique to be generally successful for treating siliceous mine tailings from several north Idaho metal mines, thickener underflow from two Appalachian coal preparation plants, and materials from numerous other coal and metal mine sites (30).

Not all slimes or sludges can be dewatered by this method, however, because the sludge from an acid mine drainage treatment plant and scrubber sludge from a large coal-fired powerplant were not responsive to this technique. The physical properties of the slurries that affect their response to treatment include electrical conductivity, particle-size distribution, water content, and surface charge density. Chemical properties also influence behavior, but their importance is specific to each sample and difficult to characterize (30).

Application

The electrokinetic phenomena of electrophoresis and electro-osmosis are principally responsible for the effects observed. Electrophoresis is the migration of small electrically charged particles through a stationary liquid due to an external electrical potential. Electro-osmosis, on the other hand, is the migration of liquid through a stationary porous solid as a result of an external electrical potential.

Particle sedimentation can be accelerated by imposing a properly oriented electrical field on a slurry. For example, in a slurry composed of negatively charged solids, the anode or positive electrode is located at the bottom of the slurry, and the cathode or negative electrode is positioned at the surface (fig. 29).

Practically speaking, the anode can be a section of abandoned steel track or wire mesh placed on the floor of the fill area prior to the slurry deposition. The cathode can be wire mesh positioned at the slurry surface and suspended from wood floats or cables attached to the roof of the mine opening. A relatively large direct voltage of 2 to 6 V/in^2 of tailings surface area is applied to the electrodes, and the negatively charged solids begin migrating downward to the positively charged anode. The electrophoretic migration will effectively accelerate the settling of these solids (30).

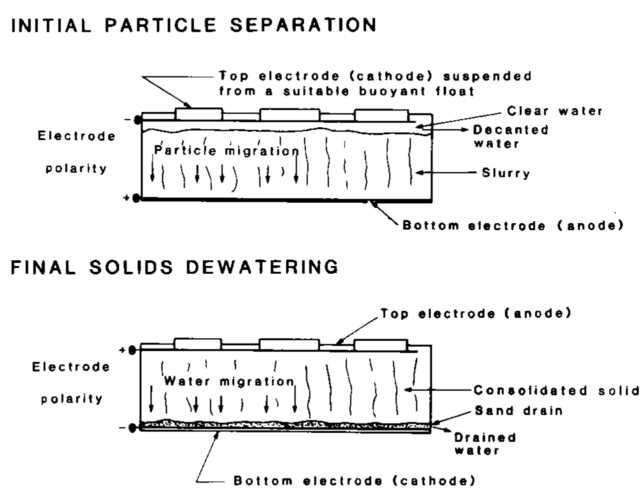


FIGURE 29. - Two configurations for electrokinetically dewatering slimes (30).

When sedimentation is complete, further drainage can be stimulated by reversing the polarity so that the bottom electrode becomes the negatively charged cathode. The water immediately adjacent to the negatively charged solid particles will contain excess dissolved positive ions that, in effect, give this water a positive charge. In an electrical field established between two separated electrodes buried in a slurry, the solid particles will not move appreciably because of their relatively dense packing, but the water will be carried toward the negatively charged cathode by the viscous drag of the migrating positive ions. This movement of water by direct current potential is electro-osmosis. As the water migrates to the cathode, the liquid can be removed by using slotted pipes or gravel drains.

Electrophoretic flow is relatively independent of pore size and is particularly attractive for dewatering dense slurries of fine particles where hydraulic flow of water through the sediment is negligible because of small pore size. Feasibility of electrokinetic dewatering for a particular slurry is best determined by direct testing in the laboratory. The complex interaction of factors affecting the efficiency of the process has prevented the use of physical properties alone as reliable predictors of performance. Change in application methods, such as current density, current reversal, electrode configuration, or settlement time, can also have unique and important effects on the response of a given slurry (30).

Current Research and Use

Field tests conducted in two Idaho mines demonstrated that the electrokinetic process can effectively dewater and densify unclassified mill tailings or slimes for use as backfill, with moderate power consumption. One mine is preparing to use the process to dewater slimes underground as a regular operating procedure, and another mine plans to use it when space occupied by old slime deposits needs to be recovered for other purposes (30).

TROMMEL SCREEN

Description of the Method and Equipment

The Bureau is also doing extensive research on the trommel screen for dewatering slurries. In this method, the slurry is first mixed with a flocculating agent to agglomerate the small particles into much larger masses. Next, the treated slurry passes over a hydrosieve to remove some of the liquid. The remaining wet solids then go into the upper end of a long inclined cylinder or trommel made of steel screen. The trommel rotates about its long axis and allows any remaining liquid to pass through the screen while the dewatered solids move through the cylinder and exit through the lower end (fig. 30) (27, 37).

The most important aspect of this method is adding the proper flocculant to the slurry. It must cause the solids to form sufficiently large masses that will not pass through the screen with the liquid but will remain on top of the screen surface. At the same time, the flocculating agent must be potent enough so that a minimum amount of the chemical will produce the required flocculation. Extensive testing has indicated that polyethylene oxide (PEO) is suitable for use with many different slurry types in the trommel screen. This flocculating agent is a water-soluble polymer having a nominal molecular weight of 8 million. This agent causes slurry solids to flocculate within minutes and works well with the hydrosieve and trommel screen (27, 37).

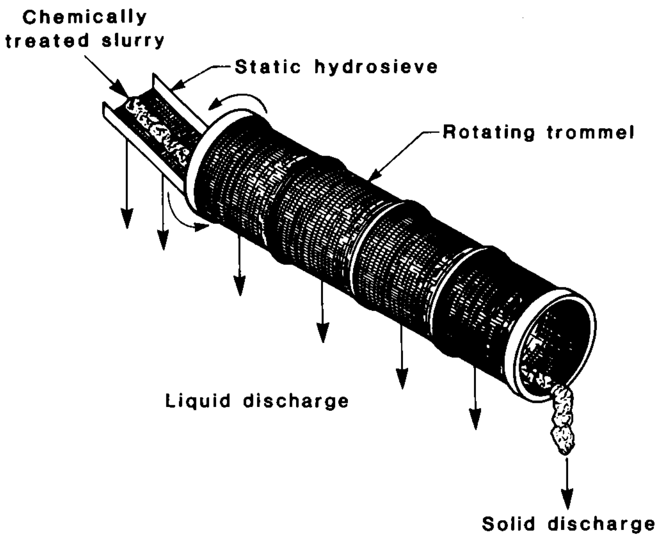


FIGURE 30. - Diagram showing operation of rotary trommel (27).

The hydrosieve used for this research was constructed of stainless steel and was 8 ft long. The first 4 ft had screen openings of 0.04 in, and the last 4 ft had openings of 0.02 in. The hydrosieve was inclined at an angle of 58° from horizontal (37).

The trommel screen was also composed of stainless steel but had 10-mesh openings. The trommel had a length of 36 in and a diameter of 6 in. The angle of inclination from horizontal was between 3° and 9°. The slurry moved over the hydrosieve and through the trommel by gravity (27, 37).

Test Results

Test results using this method indicate that phosphatic and coal-clay slurries can be successfully dewatered. In one particular test, a coal-clay slurry had 23.9 wt pct solids, of which over 70 pct were smaller than 325 mesh. A solution containing 0.125 wt pct PEO was added to the slurry at a dose of 0.78 lb PEO per ton of dry solids. The dewatered product had 60.1 wt pct solids (37). This impressive value is representative of the results obtained so far.

Conclusions

Research on the trommel screen is ongoing, so no information is available on production costs per ton of dried solids. The equipment is fairly simple in design and would not represent a substantial capital investment; however, this method does require chemical pretreatment that would significantly affect the cost of dewatering slimes. Nonetheless, this method shows promise for mineral industry use, and research is continuing on the refinement of the trommel screen.

DISCUSSION

In recent years, economic pressures have caused milling operations to maximize the separation of economic minerals from the waste material. This has been done by grinding the ore to much smaller particle sizes. This fine grinding, while increasing the mineral extraction, has also posed a serious disposal problem for the waste material. As practically all mineral beneficiation involves the use of water, the resulting solid-liquid mixture or slurry must be properly disposed of in accordance with the current environmental regulations.

There is an abundance of dewatering equipment available that can separate the solids from the water with varying degrees of efficiency. Physical separation methods, such as gravity thickeners, centrifuges, filters, thermal dryers, and cyclones, all reduce the water content or increase the solids content of the slurry. Although these items were discussed separately, they can be used in conjunction with each other to produce a solids mass with an acceptably low moisture content. Figure 31 shows the relative capabilities of common pieces of dewatering equipment in terms of the slurry particle size.

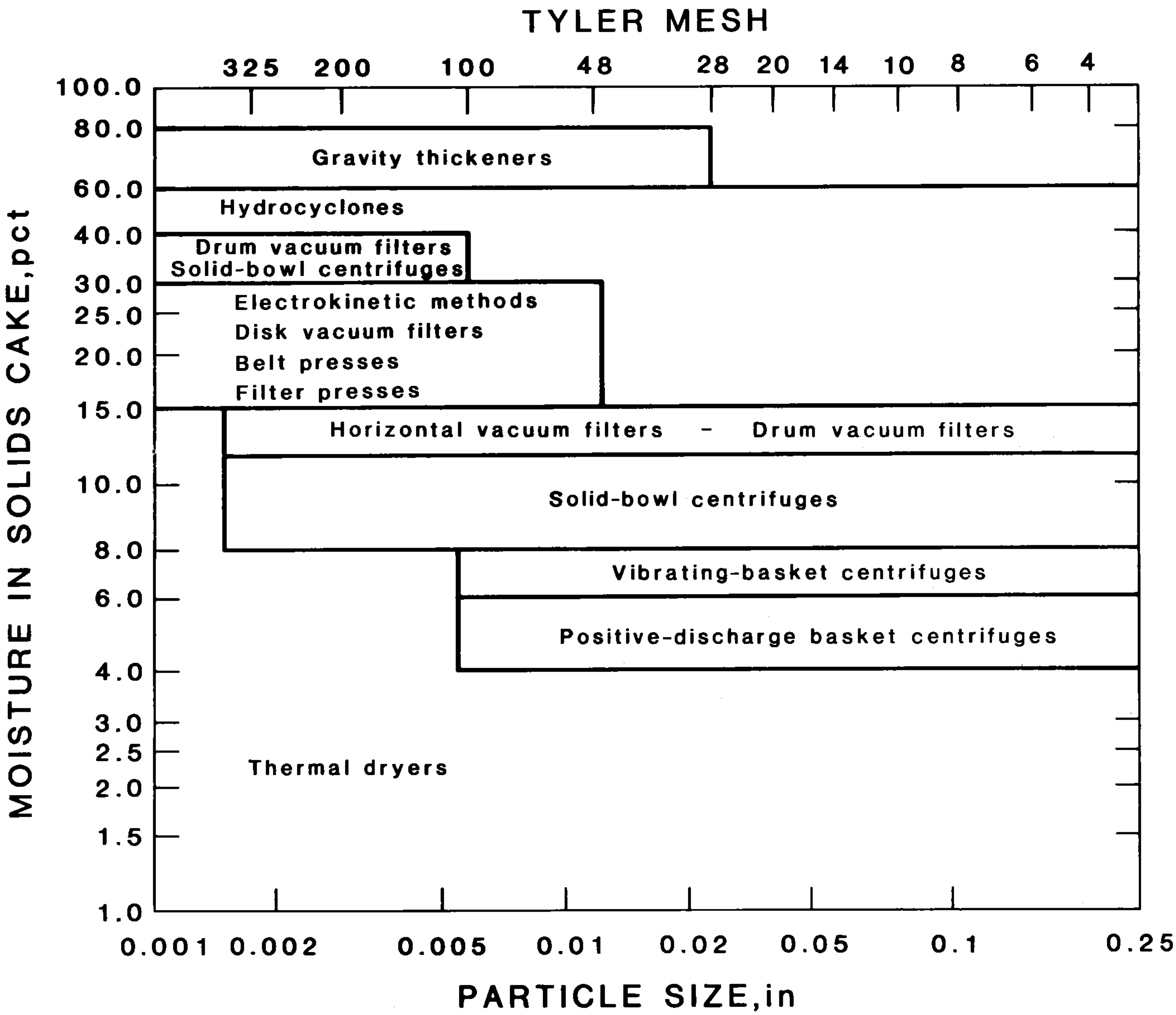


FIGURE 31. - Chart showing generalized capabilities of commonly used dewatering equipment with respect to solids cake, moisture content, and particle size (6, 18, 25, 28, 30-31).

The Bureau is continuously doing research that will benefit the mining industry. Classical dewatering methods are being improved, and new methods are being devised. Electrokinetics and the rotary trommel are but two ongoing Bureau projects that should help the mining industry reduce costs and increase efficiency.

Despite the amount of knowledge we have concerning dewatering, its practice is still an art. Each mine generates waste material which, in one way or another, is different from the waste of any other mine. Variations in mineral content and physical properties must be evaluated carefully so that the right combination of equipment and methods will produce a sufficiently dry product. This paper was prepared in order to give the mill operator or owner an overview of what dewatering equipment is available. An awareness of the different dewatering practices will enable such operators to evaluate alternatives that perhaps would not have been considered otherwise. This paper should also serve as a basis for further detailed research into the many ways of meeting the desliming challenge.

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APPENDIX A.--MATHEMATICAL TERMS

A_c = The centrifugal acceleration of a particle in a centrifuge.

A_{cs} = The cross-sectional area of a gravitational settling basin.

A_d = The drying area of a thermal dryer.

A_i = The inclination angle above horizontal of the plates in a multiple-plate thickener.

A_o = The area of the orifices in the constriction plate of a fluidized-bed dryer.

A_p = The top surface area of each plate in a multiple-plate thickener.

A_{tl} = The area of a vacuum filter test leaf.

C_{ave} = The average solids concentration in the compression zone of a gravitational thickener.

C_d = The discharge coefficient that characterizes the airflow through an orifice or constriction plate of a fluidized-bed dryer.

C_{s2} = The secondary solids concentration produced when a hypothetical gravitational thickener processes a slurry having the maximum solids concentration permissible for that particular thickener.

C_{se} = The solids concentration in a settling test after an arbitrary time, t_e.

C_{ss} = The solids concentration of the slurry prior to dewatering or thickening.

C_{ss1} = The solids concentration of a hypothetical slurry as shown on a batch flux versus solids concentration graph.

C_{ss2} = The solids concentration of a hypothetical slurry that is greater than C_{ss1}.

C_{su} = The solids concentration of the underflow or thickened solids.

C_{su1} = The solids concentration of the underflow produced by a hypothetical gravitational thickener processing an arbitrary slurry having a solids concentration C_{ss1}.

C_{su2} = The solids concentration of a hypothetical underflow that is greater than C_{su1}.

D_l = The density of the slurry liquid at a specified temperature.

D_m = The diffusivity value of moisture through a solid during thermal drying.

D_s = The density of the solid particles in a slurry.

E_c = The rate of evaporation during the constant-rate drying period during thermal drying.

E_f = The evaporation rate during the falling rate period during thermal drying.

E_t = The total efficiency for a particular piece of solid-liquid separation equipment.

e = Euler's number, which has a value of 2.71828...

F_{bt} = The bulk transport flux component based on the underflow from a particular gravitational thickener.

F_c = Centrifugal force expressed as multiples of gravitational force.

F_s = The settling flux component for a particular gravitational thickener.

F_{si} = An intercept point on the batch flux axis of a hypothetical batch flux versus solids concentration graph for a gravitational thickener.

F_{ts} = The total solids flux for a particular gravitational thickener.

F_v = The filter cake correlating factor for vacuum filters.

f_{cv} = The filter cake formation rate for a vacuum filter.

f_{cvn} = The new cake formation rate for a vacuum filter.

f_{cvo} = The old cake formation rate for a vacuum filter.

G = The gravitational acceleration at the earth's surface.

G(x) = A gravimetric separation function that describes the separation efficiency of a piece of solid-liquid separation equipment.

H_a = The humidity of the ambient air around a thermal dryer.

H_{abe} = The height of the interface between zones A and B in a slurry settling test after an arbitrary time, t_e.

H_{abo} = The original height of the interface between zones A and B in a slurry settling test.

H_{ai} = The humidity of the air at the solid-air interface in a thermal dryer.

h_y = A heat transfer coefficient for thermal drying.

J = A correction factor for particle shape.

K_y = A mass transfer coefficient for thermal drying.

L_p = The length of each plate in a multiple-plate thickener.

L_{sz} = The length of the settling zone in the direction of slurry transport in a centrifuge.

M_a = The molecular weight of air.

M_s = The mass of all solids that have been separated from the liquid of a slurry.

M_t = The total mass of all solids in a slurry before solid-liquid separation.

N = The number of revolutions per second for a centrifuge bowl.

P_{dc} = The pressure differential across the cake of a vacuum filter.

P_{do} = The air pressure differential across the orifice plate of fluidized-bed dryers.

P_{op} = The mean absolute air pressure in the orifice of a constriction plate in a fluidized-bed dryer.

Q_c = The volumetric flow of slurry through a centrifuge.

Q_s = The volumetric flow of a slurry through a settling zone.

q_a = The airflow through the orifice plate under controlled air pressure and temperature conditions in a fluidized-bed dryer.

R_c = The radius of curvature of a particle in a centrifuge.

R_{cb} = The radial distance from the rotational axis to the inside surface of the bowl in a centrifuge.

R_{sc} = The percent recovery of solids in the cake produced by a centrifuge.

R_{sz} = The radial distance from the rotational axis to the surface of the settling zone in a centrifuge.

S = Equal to one-half the thickness of the layer of solids in a thermal dryer.

S_f = The subsidence rate for solids in the bottom of a settling tank.

S_{gs} = The average specific gravity of the solids in a slurry.

S_{gsc} = The average specific gravity of the solids in compression.

S_{gw} = The specific gravity of water, which is 1.0.

S_p = The spacing between plates of a multiple-plate thickener.

T_a = The ambient air temperature around a thermal dryer.

T_i = The air temperature at the solid-air interface in a thermal dryer.

T_k = The temperature in Kelvins.

t_{cv} = The cycle time for vacuum filters.

t_{cvn} = The new cycle time for a vacuum filter.

t_{cvo} = The old cycle time for vacuum filter.

t_{dv} = The drying time for a vacuum filter cake.

t_e = An arbitrary amount of time after the beginning of a settling test for gravitational thickeners.

t_{hs} = The holding time necessary for the solids to settle from the entering feed solids concentration to the underflow concentration.

t_n = An arbitrary time after the falling-rate drying has begun in a thermal dryer.

V_{am} = The air mass velocity through a thermal dryer.

V_{av} = The volume of air passing through a vacuum filter cake per unit of filter surface area.

V_{sc} = The volume of solids in compression in a gravitational thickener.

v_{bt} = The bulk transport velocity of the solids through a gravitational thickener.

v_l = The linear velocity of a slurry particle moving through a settling zone.

v_p = The linear peripheral velocity of a particle in a centrifuge.

v_s = The terminal settling velocity of solids through a liquid.

v_v = The resultant velocity vector of a settling slurry particle as a consequence of its forward motion and the downward pull of gravity.

W_{ds} = The dry weight of the solids cake obtained during a vacuum filter leaf test.

W_{dst} = The dry weight of solids for a given cake thickness on a vacuum filter.

W_{rd} = The weight ratio of water to solids in the discharge of a gravitational thickener.

W_{rs} = The weight ratio of water to solids in the slurry.

W_{sc} = The weight percent of solids in the cake.

W_{sl} = The weight percent of solids in the liquid after liquid-solid separation.

W_{ss} = The weight percent of solids in a slurry prior to liquid-solid separation.

X_{eq} = The equilibrium moisture content of the solids during the falling-rate period of thermal drying.

X_{fav} = The average free moisture content during the falling-rate period of thermal drying.

X_{fi} = The initial free moisture content of the solids at the beginning of the falling-rate period during thermal drying.

X_{tav} = The average total solids moisture content during the falling-rate period of thermal drying.

X_{ti} = The initial total moisture content of the solids at the beginning of the falling-rate period of thermal drying.

x = The diameter of solid particles in a slurry.

x_{max} = The largest particle size on a grade efficiency curve that is capable of being separated from the liquid of a slurry by a particular piece of dewatering equipment.

x_{50} = The particle size on a grade efficiency curve representing a 50-pct probability of being separated from the liquid of a slurry.

x_{98} = The particle size on a grade efficiency curve that represents 98-pct separation efficiency.

α = A factor used in the liquid diffusion equation for the falling-rate drying period of thermal drying.

β = A factor used in the liquid diffusion equation for the falling-rate drying period of thermal drying.

λ_i = The latent heat of water at the temperature of the solid-air interface in a thermal dryer.

μ = The viscosity of the liquid in a slurry at a specified temperature.

π = 3.1416.

Σ = A characteristic value that describes the machine variables for a particular centrifuge.

ω = The angular velocity of a particle undergoing centrifugal acceleration.

APPENDIX B.--MANUFACTURERS OF DEWATERING EQUIPMENT AS OF OCTOBER 1982 (24, 34)[1,2]

AFL Industries, Inc.
3661 West Blue Heron Blvd.
Riviera Beach, FL 33404

Alfa-Laval, Inc.
2115 Linwood Ave.
Fort Lee, NJ 07024

Baker-Perkins, Inc.
1000 Hess St.
Saginaw, MI 48601

The Leon J. Barrett Co.
Box 551
Worcester, MA 01613

Bird Machine Co., Inc.
100-T Neponset St.
South Walpole, MA 02071

Calgon Corp.
Box 1346-C
Pittsburgh, PA 15230

Carus Chemical Co., Inc.
1500 Eighth St.
LaSalle, IL 61301

C-E Bauer
Box 968
Springfield, OH 45501

C-E Raymond
200 West Monroe St.
Chicago, IL 60606

Centrifugal and Mechanical Industries, Inc.
146 President St.
St. Louis, MO 63118

Clow Corp.
1211 W. 22d St.
Oak Brook, IL 60521

Dedert Corp.
20000-T Governor's Dr.
Olympia Field, IL 60461

Denver Equipment Div.
Joy Manufacturing Co.
Box 340
Colorado Springs, CO 80901

Donaldson Co.
Liquid Systems Div.
1400 West 94th St.
Minneapolis, MN 55431

Dorr-Oliver, Inc.
79 Havemeyer Lane
Stamford, CT 06904

Duriron Co., Inc.
Filtration Systems Div.
9542 Hardpan Rd.
Angola, NY 14006

Environmental Elements Corp.
Box 1318
Baltimore, MD 21203

Envirotech Corp.
3000 Sand Hill Rd.
Menlo Park, CA 94025

ERC/Lancy Div., Dart & Kraft Co.
525 West New Castle St.
Zelienople, PA 16063

Filpaco Industries, Inc.
3837 West Lake St.
Chicago, IL 60624

FMC Corp., Materials Handling Systems Div.
3400 Walnut
Colmar, PA 18915

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[1]Reference to specific equipment suppliers does not imply endorsement by the Bureau of Mines.

[2]This list is as complete as possible based on the information available at the time this paper was written. No responsibility can be taken for omissions or changes in listings.

Heyl and Patterson
Dept. 10
Box 36
Pittsburgh, PA 15230

Industrial Filter & Pump
Manufacturing
5900 West Ogden Ave.
Cicero, IL 60650

Infilco Degremont, Inc.
Dept. T-R
Box 29599
Richmond, VA 23288

IU Conversion Systems, Inc.
Dept. T-R
115 Gibraltar Rd.
Horsham, PA 19044

JWI, Inc.
Box 9A
Holland, MI 49423

Keene Corp. Filtration Div.
1571 Forrest Ave.
LaGrange, GA 37743

Komline-Sanderson Engineering
Corp.
100 Holland Ave.
Peapack, NJ 07977

Krebs Engineers
1205 Chrysler Dr.
Menlo Park, CA 94025

Krofta Engineering Corp.
101-T Yokun Ave.
Lenox, MA 01240

Lakos Separators
1911 North Helm
Box 6119
Fresno, CA 93703

Larox OY
Box 29
SF-53101 Lappeenranta 10
Finland

Lavin/Guinard International,
Inc.
500 Davisville Rd.
Hatboro, PA 19040

McNally-Pittsburg
Manufacturing Corp.
Third at Walnut St.
Pittsburg, KS 66762

Nalco Chemical Co.
2901 Butterfield Rd.
Oak Brook, IL 60521

National-Standard Co.
Perforated Metals Div.
Drawer 507
Carbondale, PA 18407

Netzsch, Inc.
119 Pickering Way
Pickering Creek Industrial
Park
Exton, PA 19341

Parkson Corp.
2727-T NW 62d St.
Fort Lauderdale, FL 33309

Passavant Corp.
Carson Rd.
Box 2503
Birmingham, AL 35201

William R. Perrin Co., Ltd.
432 Monarch Ave.
Ajax, Ont.
Canada L1S 2G7

Serfilco Div.
Service Filtration Corporation
1234 Depot St.
Glenview, IL 60025

SFS Div., BINAB USA, Inc.
15271 NW 60th Ave.
Miami Lakes, FL 33014

D. R. Sperry and Co.
112-T North Grant St.
North Aurora, IL 60542

Star Systems Filtration Div.
101 Kershaw St.
Box 815
Timmonsville, SC 29161

Transamerica Delaval, Inc.
Condenser and Filter Div.
Front St.
Florence, NJ 08518

Tretolite Div., Tretolite Corp.
369 Marshall Ave.
St. Louis, MO 63119

Vara International, Inc.
1201-T 19th Pl.
Vara International Plaza
Vero Beach, FL 32960

WEMCO Div., Envirotech Corp.
1796 Tribute Rd.
Box 15619
Sacramento, CA 95813

Western States Machine Co.
1716 Fairgrove Ave.
Hamilton, OH 45012

Zimpro, Inc.
Dept. MZ
Military Road
Rothschild, WI 54474

APPENDIX C.--AVAILABLE DEWATERING EQUIPMENT LISTED BY MANUFACTURER AS OF OCTOBER 1982 (24)

Equipment and manufacturer[1,2]	Type
Gravitational thickeners:	
AFL	Multiple plate.
Denver	Conventional.
Dorr-Oliver	Conventional and high rate.
Envirotech	Do.
Industrial Filter & Pump	Multiple plate.
Larox	Conventional.
Parkson	Multiple plate.
SFS	Do.
Centrifuges:	
Alfa-Laval	Basket, disk, and solid bowl.
Baker-Perkins	Pusher.
Barrett	Basket and solid bowl.
Bird Machine	Do.
Centrifugal and Mechanical	Basket.
Dedert	Do.
Donaldson	Solid bowl.
Dorr-Oliver	Basket, disk, and solid bowl.
Envirotech	Horizontal.
IU Conversion	Solid bowl.
Lavin	Basket and solid bowl.
MET Pro	Solid bowl.
National Standard	Basket.
WEMCO	Do.
Western States	Do.
Hydrocyclones:	
C-E Bauer	Undifferentiated.
Dorr-Oliver	Do.
Krebs Engineers	Do.
Lakos	Do.
Larox	Do.
WEMCO	Do.
Filters:	
AFL	Belt press and gravity.
Bird Machine	Horizontal belt, belt, and drum.
C-E Bauer	Gravity and rotary.
Clow	Filter press and horizontal belt.
Denver	Horizontal belt.
Dorr-Oliver	Belt, disk, and drum.
Duriron	Filter press.
Environmental Elements	Do.
Envirotech	Drum, filter press, gravity, and horizontal belt.
ERC/Lancy	Filter press.
Filpaco	Do.
Industrial Filter & Pump	Do.
Infilco	Horizontal belt.
IU Conversion	Drum.
JWI	Filter press.
Komline-Sanderson	Vacuum belt, belt press, drum, and gravity.
Krofta	Belt and gravity.

See footnotes at end of table.

APPENDIX C.--AVAILABLE DEWATERING EQUIPMENT LISTED BY MANUFACTURER AS OF OCTOBER 1982 (24)--Continued

Equipment and manufacturer[1,2]	Type
Filters--Continued:	
Larox	Disk, filter press, and drum.
Netzsch Inc	Filter press.
Parkson	Belt press.
Passavant	Belt press and filter press.
Perrin	Horizontal belt and filter press.
Serfilco	Gravity and vacuum belt.
Sperry	Filter press.
Star Systems	Do.
Transamerica-Delaval	Gravity and belt press.
Vara	Gravity.
Zimpro	Filter press.
Thermal dryer or incinerator:	
C-E Raymond	Rotary, fluidized bed, and suspension.
Dedert	Drum and suspension.
Dorr-Oliver	Fluidized bed.
FMC	Drum, fluidized bed, and louvre.
Heyl and Patterson	Fluidized bed.
Komline-Sanderson	Suspension.
McNally-Pittsburg	Fluidized bed and vertical tray.
Chemical treatment aids:	
Calgon	Polyelectrolytes.
Carus	Potassium permanganate.
ERC/Lancy	Lime and polyelectrolytes.
Industrial Filter & Pump	Lime.
IU Conversion	Lime and polyelectrolytes.
Keene	Do.
Nalco	Polyelectrolytes.
Passavant	Lime and polyelectrolytes.
Tretolite	Polyelectrolytes.

[1]See appendix B for complete name and address.

[2]This list is as complete as possible based on the information available at the time this paper was written. No responsibility can be taken for omissions or changes in listings.

APPENDIX D.--EQUIPMENT EFFICIENCY

Less than perfect performance of separation equipment can be characterized by the separation efficiency. The grade efficiency concept can be applied to solid-liquid separation equipment whose performance does not change with time if all operational variables are kept constant. Hydrocyclones, centrifuges, and gravitational thickeners are examples of such equipment. This concept is not widely used in filtration because the efficiency changes with the amount of solids collected on the face of the filter medium. For filtration, though, it is helpful to determine the grade efficiency of the clean medium, which influences the initial retention characteristics of the filter and can be used for filter rating (32, p. 31).[1]

TOTAL EFFICIENCY

The total efficiency for dewatering equipment can be determined by the equation

$$E_t = \frac{M_s}{M_t}, \tag{D-1}$$

where E_t = the total equipment efficiency,

M_s = the mass of all solids separated from a slurry liquid,

and M_t = the mass of all slurry solids prior to solid-liquid separation (32, p. 33).

The performance of most available separational equipment is predominantly size dependent, so the total efficiency depends on the size distribution of the feed solids and is not suitable for the general criterion of efficiency. Consequently values of total efficiency stated by equipment manufacturers may not be entirely accurate concerning the separational capability of their equipment. Additional information must be known about the particle-size distribution of the feed solids, the density of the solids, and such operational data as flow rate, temperature, type of fluid, and solids input concentration. A single value for the total efficiency cannot be used to represent the separation capability of the equipment for any materials other than those actually tested (32, pp. 34-35).

GRADE EFFICIENCY

The efficiency of separation equipment, however, can be characterized by a gravimetric grade efficiency function, G(x). This is a probablistic mathematical expression, based on mass efficiency, which describes the particle trajectories during the separation process. A grade efficiency function can be developed for each type of separation equipment that describes the efficiency of separation for a range of particle sizes. This information can be graphed as an S-shaped curve, such as the one shown in figure D-1. This graph is often referred to

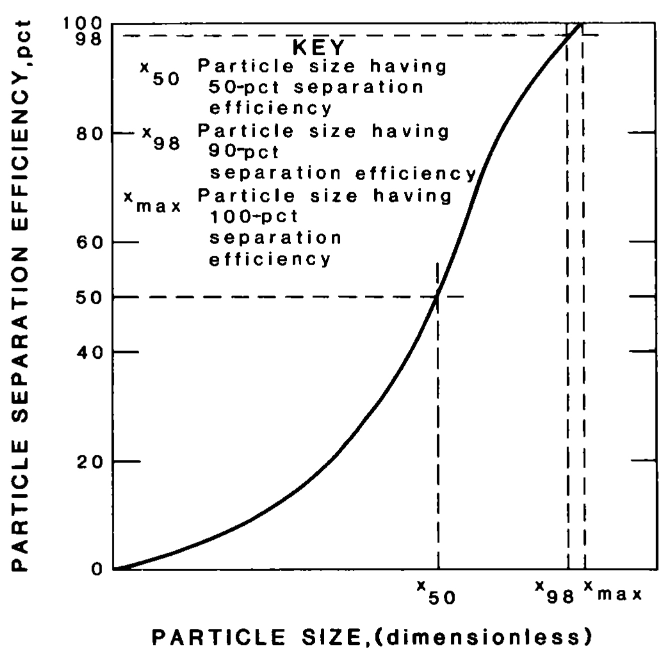

FIGURE D-1. - Example of a grade efficiency curve showing the relationship of x_{50}, x_{98}, and x_{max} (32).

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[1]Underlined numbers in parentheses refer to items in the list of references preceding the appendixes.

as the partition probability curve because it shows the probability for each particle size of either being separated or remaining with the fluid (32, p. 35).

PARTICLE SIZE PARAMETERS

The grade efficiency curve can be used for determining several operational parameters for a particular piece of equipment. The particle corresponding to the 50-pct probability is called the equiprobable size, x_{50}, and is used as the minimum cutoff size or cut size of the particular type of equipment. This cut size is independent of the feed material, and its determination requires a knowledge of the entire grade efficiency curve (32, p. 35).

In any separation operation, there will be a particle size larger than the grade efficiency, which is 100 pct. This is the largest particle remaining in the overflow after separation of the maximum particle size that would have a chance to escape and is called x_{max} (32, p. 38).

If the particle trajectories in the separator can be approximated, the most unfavorable conditions of separation are taken for determining this limit of separation. It is difficult to determine the limit of separation accurately, so the size corresponding to 98-pct efficiency, x_{98}, is used, which gives an easily defined point. This size is called the approximate limit of separation and is widely used in filter rating (32, p. 38). The relationship of x_{50}, x_{98}, and x_{max} is shown in figure D-1.

The concept of grade efficiency is helpful in determining the application of a particular piece of equipment for a particular dewatering or desliming operation. For a more detailed discussion of grade efficiency, consult Svarovsky (32, pp. 31-57).

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