CENTRIFUGAL SEPARATION

1. Introduction

Centrifugal separation is a mechanical means of separating the components of a mixture of liquids that are immicible or of liquids and insoluble solid particles. The material is accelerated in a centrifugal field that acts upon the mixture in the same manner as a gravitational field. The centrifugal field can, however, be varied by changes in rotational speed and equipment dimensions, whereas gravity is essentially constant. Commercial centrifugal equipment can reach an acceleration of 20,000 times gravity (20,000 G); laboratory equipment can reach up to 360,000 G. The ultracentrifuge and gas centrifuge represent special cases that establish separation gradients on a molecular scale. The usual gravitational operations, such as sedimentation (qv) or flotation (qv) of solids in liquids, drainage or squeezing of liquids from solid particles, and stratification of liquids according to density, are accomplished more effectively in a centrifugal field (see SEPARATION, SIZE SEPARATION).

The development of theory for centrifugation equipment has been slow. Flow patterns in the centrifuge bowls are complex and difficult to model mathematically. The concept of a theoretical capacity factor for sedimentation depends only on equipment characteristics and is independent of the system (1,2). The theoretical effect of particle size distribution in single and multistage centrifugation has been demonstrated (3). Extensive application of centrifugation to dewatering (qv) and thickening of relatively soft solids and hydrogels associated with industrial and municipal waste treatment has resulted in changes in centrifuge design (see WASTES, INDUSTRIAL). A sound theoretical basis for centrifugal drainage or squeezing of liquids from solids has been only partially developed for compressible solids. Many aspects are in need of amplification.

Herein centrifugal separation in a liquid medium is discussed. For a brief discussion of centrifugal separation in a gaseous medium see previous editions of this article.

2. Theory

2.1. Separation by Density Difference.

A single solid particle or discrete liquid drop settling under the acceleration of gravity in a continuous liquid phase accelerates until a constant terminal velocity is reached. At this point the force resulting from gravitational acceleration and the opposing force resulting from frictional drag of the surrounding medium are equal in magnitude. The terminal velocity largely determines what is commonly known as the settling velocity of the particle, or drop under free-fall, or unhindered conditions. For a small spherical particle, it is given by Stokes' law:

$$v_g = \frac{\Delta\delta d^2 g}{18\ \mu} \tag{1}$$

where v_g = the settling velocity of a particle or drop in a gravitational field; $\Delta\delta = \delta_S - \delta_L$ = the difference between true mass density of the solid particle or liquid drop, and that of the surrounding liquid medium; d = the diameter of the solid particle or liquid drop; g = the acceleration of gravity; and μ = the absolute viscosity of the surrounding medium.

Stokes' law can be readily extended to a centrifugal field:

$$v_s = \frac{\Delta\delta d^2 \omega^2 r}{18\ \mu} = v_g \left(\frac{\omega^2 r}{g}\right) \tag{2}$$

where v_s = the settling velocity of a particle or drop in a centrifugal field; ω = the angular velocity of the particle in the settling zone; and r = the radius at which settling velocity is determined. Analogous equations describe the terminal velocity of a light particle or drop rising in a heavier continuous medium.

The settling velocity, v_s, is relative to the continuous liquid phase where the particle or drop is suspended. If the liquid medium exhibits a motion other than the rotational velocity, ω, the vector representing the liquid-phase velocity should be combined with the settling velocity (eq. 2) to obtain a complete description of the motion of the particle (or drop).

These concepts are used to analyze separations in the bottle centrifuge, the imperforate bowl centrifuge, and the disk centrifuge. Separation by density difference in other types of centrifuges can be analyzed by analogy.

The Bottle Centrifuge. Analysis of the performance of a bottle centrifuge is based on the model shown in Figure 1. A solid or liquid particle is considered in an initial position, X, at a radius, r, from the axis of rotation. If equation 2 is applied to this specific particle, assuming that $v_s = dr/dt$, then

$$\int_r^{r_C} \frac{dr}{r} = \int_0^t v_g \left(\frac{\omega^2}{g}\right) dt \tag{3}$$

where r_C = the radius of the sedimented cake, and t = the time during which the particle is subjected to centrifugal acceleration. Integration of equation 3 leads to the following:

$$\ln \frac{r_C}{r} = v_g \left(\frac{\omega^2}{g}\right) t \tag{4}$$

A radius, r', that divides the volume of supernatant into two equal parts can be defined as follows:

$$(r' - r_1) = (r_c - r') \quad \text{or} \quad r' = (r_c + r_1)/2 \tag{5}$$

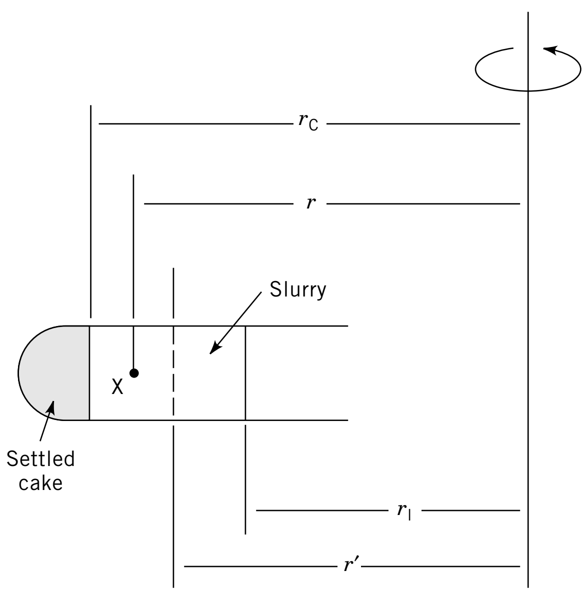


Fig. 1. Separation in a bottle centrifuge, where X is the initial position of a particle (drop). See text for definition of terms.

where r_1 is the radius of the free surface of the liquid. Assuming the presence of more than one particle in the liquid as well as a uniform initial particle distribution, then each of the two volumes defined by r'; initially contains the same number of particles. By making the further assumption that the particles in the suspension are identical, a settling time, t, is chosen so that those particles starting from radius r'; all reach the cake r_C after time t. Under these conditions, one-half of the particles that were in suspension at $t = 0$ are sedimented after the time, t, has elapsed. The other one-half, initially located above the level defined by the radius r', remain in suspension. If r is replaced in equation 4 by r', a sedimentation condition is established that is referred to as 50% cutoff. To determine 100% cutoff (complete solids capture) r should be replaced by r_1 in equation 4. An effective capacity, Q_0, for the bottle centrifuge is determined by the ratio between the volume, V, occupied by the slurry in the bottle, and the spinning time, t, calculated from equation 4:

$$Q_0 = V/t = 2v_g(\omega^2/g)V/(2\ \ln(2r_c/(r_c + r_1))) \tag{6}$$

In equation 6, v_g characterizes the settling behavior of the solid particles or liquid drops in the suspension, whereas the second part of the right-hand side refers to speed and size of the bottle centrifuge and is expressed by the capacity factor Σ_B:

$$\Sigma_B = \omega^2 V/2g\ \ln(2r_c/(r_c + r_1)) \tag{7}$$

This capacity factor has the dimension of an area and represents the area of a static gravity settling tank having a separation performance equal to that of the rotating bottle centrifuge handling the same particles. By combining equations 6 and 7, to eliminate volume, V, equation 8 is obtained:

$$\frac{Q_0}{\Sigma_B} = \frac{2g}{(\omega^2 t)} \cdot \ln\left(\frac{2r_C}{r_C + r_1}\right) \tag{8}$$

Equation 8 provides the basis of comparison of the performance of various bottle centrifuges containing the same material, and also, under certain circumstances, of other types of sedimentation centrifuges, if geometric dissimilarities are also considered.

The capacity factor, Σ_B, defined by equation 7, is derived from a set of assumptions. An additional assumption is specific to the bottle centrifuge. Namely, a particle is considered sedimented when it reaches the surface of the cake without contacting the tube wall.

The Imperforate Bowl Centrifuge. In an imperforate bowl centrifuge the flow of the continuous liquid phase is nominally axial, except for areas immediately adjacent to the feed inlet and effluent outlet. Tubular bowl, imperforate basket and decanter centrifuges satisfy this definition.

The mathematical model chosen for this analysis is that of a cylinder rotating about its axis (Fig. 2). Suitable end caps are assumed. The liquid phase is introduced continuously at one end so that its angular velocity is identical everywhere with that of the cylinder. The flow is assumed to be uniform in the axial

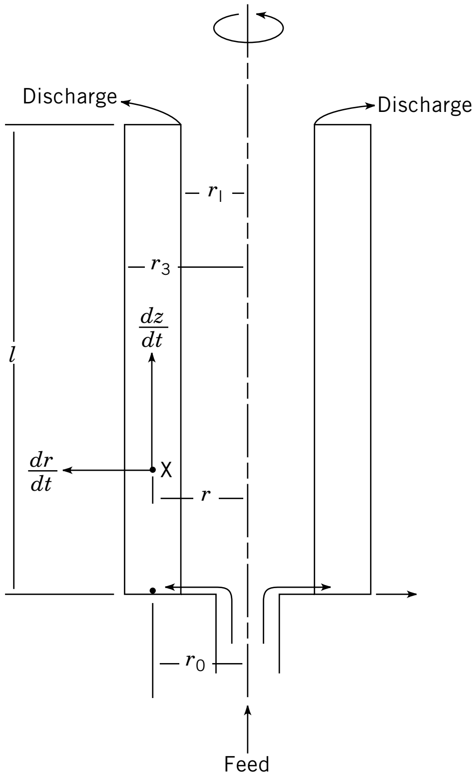


Fig. 2. Separation in a basket or tubular centrifuge. Terms are defined in the text.

direction, forming a layer bound outwardly by the cylinder and inwardly by a free air–liquid surface. Initially the continuous liquid phase contains uniformly distributed spherical particles of a given size. The concentration of these particles is sufficiently low that their interaction during sedimentation is neglected.

Under these circumstances, the settling motion of the particles and the axial motion of the liquid phase are combined to determine the settling trajectory of these particles. The trajectory of particles just reaching the bowl wall near the point of liquid discharge defines a minimum particle size that starts from an initial radial location and is separated in the centrifuge. A radius r' is chosen to divide the liquid annulus in the bowl into two equal volumes initially containing the same number of particles. One-half of the particles of size d present in the suspension are separated; the other one-half escape, which is referred to as a 50% cutoff.

The feed rate corresponding to this condition is related to the bowl geometry r_1, r_3, and l; the bowl angular speed, ω and the Stokes' settling velocity, v_g (eq. 2).

$$Q_0 = V/t = 2\ v_g(\omega^2/g)\pi l(r_3^2 - r_1^2)/(\ln\ (2r_3^2/(r_3^2 + r_1^2)) \tag{9}$$

As an approximation with a maximum error of 4%:

$$Q_0 = 2v_g(\omega^2/g)2\pi l(3/4r_3^2 + 1/4r_1^2) \tag{10}$$

where v_g = the Stokes' settling velocity (see eq. 1); ω = the angular velocity of the centrifuge; g = the gravitational acceleration; 1 = the length of the settling zone; r_3 = the radius of the inside wall of the cylinder; and r_l = the radius of the free surface of the liquid layer in the cylinder. Equation 9 can be rewritten as equations 11 and 12:

$$Q_0 = 2v_g\Sigma_T \tag{11}$$

where

$$\Sigma_T = 2\pi l\left(\frac{\omega^2}{g}\right)\left(\frac{3}{4}\,r_3^2 + \frac{1}{4}r_1^2\right) \tag{12}$$

Equation 11 estimates the flow or throughput rate, above which particles of size d are <50% sedimented, and below which >50% are collected. Equations 11 and 12 are also applicable to the light particles rising in a heavy phase liquid, provided that r_3 and r_1 are interchanged in equation 12.

The theoretical capacity factor, Σ_T, defined by equation 12 has the dimension of an area and can be interpreted as the area of a gravity settling tank where the separation performance is equal to the centrifuge provided that the factor v_g is the same for both. This restriction is required because the particles suspended in the continuous phase can be deaggregated and further dispersed by the vigorous shearing to which the feed is subjected during acceleration in the centrifuge. If this effect is not considered in comparison with that of the settling tank, centrifugal sedimentation might be less favorable in practice than is anticipated by theory.

Equation 12 can be further reduced to facilitate understanding of its use and application. If, instead of the radii of pond surface, r_1, and bowl wall, r_3, a mean radius, r_m, is introduced, equation 12 can be rewritten as follows:

$$\Sigma_T = 2\pi r_m l\left(\frac{\omega^2 r_m}{g}\right) \tag{13}$$

Equation 13 shows that Σ_T can be expressed as the product of a mean sedimentation area ($2\pi r_m l$) and the G level ($\omega^2 r_m/g$), and therefore reflects the increased sedimentation rate expected through a defined area having centrifugal acceleration instead of gravity.

Disk Centrifuge. The separation of particles inside a disk stack is illustrated in Figure 3. The continuous liquid phase, containing solid or liquid particles to be separated, flows from the outside of the disk stack, having radius r_2, to the inside discharge opening, having radius r_1. Assuming that the liquid phase is evenly divided between the spaces formed by the disks, the flow in each disk space is Q_0/n, where Q_0 is the total flow through the entire disk stack and n is the number of spaces (disks). The flow of the continuous liquid phase is also assumed to be in a radial plane and parallel to the surfaces of the disks, and as having the same angular velocity as the stack.

Here again an equation is established (2) to describe the trajectory of a particle under the combined effect of the liquid-transport velocity acting in the x

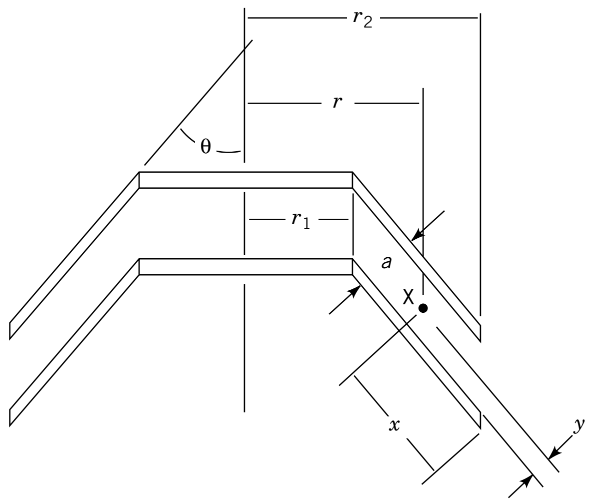


Fig. 3. Separation in a disk centrifuge, where X is the initial position of the particle. Other terms are defined in the text.

direction and the centrifugal settling velocity component in the y direction. Equation 14 determines the minimum particle size that originates from a position on the outer radius, r_2, and the midpoint of the space, a, between two adjacent disks, and just reaches the upper disk at the inner radius, r_1. Particles of this size initially located above the midpoint of space a are all collected on the underside of the upper disk; those particles initially located below the midpoint escape capture. This condition defines the throughput, Q_0, for which a 50% recovery of the entering particles is achieved. That is,

$$Q_0 = 2v_g\left(\frac{2\pi n\omega^2}{3\,g}\ \cot\,\theta(r_2^3 - r_1^3)\right) \tag{14}$$

where v_g = the Stokes' settling velocity (see eq. 1); n = the number of spaces between disks in a stack; ω = the angular velocity of the centrifuge; θ = one-half the included angle of the disks; r_2 = the outer radius of the disks; and r_1 = the inner radius of the disks (see Fig. 3). If v_g, which describes the settling characteristics of the particles, is separated from parameters relating to the geometry and rotational speed of the disk stack, then,

$$Q_0 = 2v_g\Sigma_D \tag{15}$$

where

$$\Sigma_D = \frac{2\pi n\omega^2}{3\,g}\ (\cot\,\theta(r_2^3 - r_1^3)) \tag{16}$$

Again, Σ_D corresponds to the area of a gravity settling tank capable of the same separation performance as the disk stack defined by the parameters included in equation 16.

2.2. Separation by Drainage. The theory covering drainage in a packed bed of particles is incomplete, and requires more development for a centrifugal field. Liquid is held within the bed by various forces. Removal involves several flow mechanisms. In addition, the centrifugal acceleration changes with radius in the bed, causing changes in packing tendencies of particles and accelerating forces on the residual liquid.

There are three types of liquid content in a packed bed: (*1*) in a submerged bed, there is liquid filling the larger channels, pores, and interstitial spaces; (*2*) in a drained bed, there is liquid held by capillary action and surface tension at points of particle contact, or near-contact, as well as a zone saturated with liquid corresponding to a capillary height in the bed at the liquid discharge face of the cake; and (*3*) essentially undrainable liquid exists within the body of each particle or in fine, deep pores without free access to the surface except perhaps by diffusion or compaction.

The last type of liquid can be removed by evaporation (qv) or displacement by another liquid but cannot be removed by simple flow in either a gravitational or centrifugal field. There is no sharp distinction between the first two types. The rate and extent of liquid removal from a submerged bed during drainage depends on the physical characteristics of the components, the force of the centrifugal field, and the time of exposure to the field. The residual liquid content of a drained cake consists largely of the capillary and irremovable types at the time of discharge from the separation equipment.

During cake formation and drainage the liquid moves into and through the bed in three different ways. During cake deposition, a continuous head of liquid ranging in composition from feed to clarified supernate may exist over the deposited cake. After feed is stopped, a layer of essentially clarified liquid may still exist over slow-draining cakes. Wash liquor, if it is used, may also create a liquid layer over the deposited cake. Drainage under these conditions requires continuous flow through the cake. The interstitial spaces are assumed to be full. When the free liquid layer no longer exists above the cake, the free liquid surface moves through the cake to an equilibrium position at the capillary height, leaving behind the larger voids filled with gas or vapor. Then, after bulk drainage of the larger voids, liquid still exists in the cake's upper zone in a film covering the surfaces of the solids and in partially filled voids having very restricted outlets. In time, some of this liquid flows as a film to the continuous liquid layer at the capillary height.

If a cake is sufficiently impermeable to permit the buildup of a feed or wash liquid head, flow through the cake approaches steady-state conditions except for changes in compaction or cake thickness as more feed is added, or in the liquid head if the drainage rate differs from the liquid rate addition. An equation for full-pore flow in a centrifugal field has been developed based on Darcy's equation. The hypotheses set forth and for the most part proved are as follows: flow radiates out from the rotation axis and the effect of the gravitational field is negligible; voids at all points are filled with liquid that moves in laminar flow through the cake; kinetic energy changes of the liquid in the cake may be neglected, ie, the filter medium is sufficiently permeable so that it does not run full of liquid; and ambient pressure exists at the outer face of the cake. Essentially incompressible solids produce very similar cake permeabilities in a vacuum, under pressure, and during centrifugal filtration, although significant local variations in permeability may occur because of irregularities in the feed and its distribution. When pores are filled, the flow rate of the supernatant liquid through the cake (4,5) is as follows:

$$Q = \left(\frac{\pi K \omega^2 h \delta_1}{\mu_l g}\right)\left(\frac{r_m^2 - r_l^2}{\ln\,(r_m/r_c)}\right) \tag{17}$$

where K is the permeability; h is the basket height; r_C and r_m are radii from the axis of rotation to the inner and outer faces of cake, respectively; ω is the angular velocity; and r_l is the radius from the axis to the inner face of the liquid layer. Comparison to the usual term α for cake resistance in pressure filtration shows that $K = \delta_l g/\alpha$. The functions related to cake thickness, (r_m/r_C); liquid layer thickness, $(r_m^2 - r_l^2)$; angular velocity, ω^2; and viscosity, μ_l, have been verified by experiment. The exponent on the angular velocity varies for materials that exhibit cake compression as speed is increased, but compression effects are minor on starch, chalk, and kieselguhr, ie, loose or porous diatomite (qv). In practice, the exponent of ω can probably be assumed to have a value of two and experimental variations in permeability can be absorbed by changes in the permeability coefficient. The real value of equation 17 lies in its use for estimating the effect of changes in operating variables for a material where the characteristics are already known from experimental data or plant operation.

Low permeability cakes draining under the conditions of equation 17 are usually handled in perforate basket centrifuges that have relatively long (20 min to several hours) cycles. Following elimination of the supernatant liquid, unsteady-state drainage of the cake may often be neglected. These slow-draining cakes often support such a large capillary height, eg, 90–99% of void volume for chalk (5), that little additional dewatering (qv) is obtained after completion of free-liquid drainage. Solids of larger particle size, or freer drainage characteristics, are handled in automatic perforate baskets or continuous screen centrifuges. Low final moistures are usually achieved. For cakes of high permeability, the period when a liquid layer exists above the cake is either short or nonexistent. The time cycle depends chiefly on the rate of film drainage at the completion of bulk liquid flow. Under these conditions film drainage and permanent residual moisture are most important.

The quantity of undrainable residual moisture cannot be predicted without the benefit of experimental data. Equation 18 (6) indicates the important parameters where the exponents were determined using limited experimentation. Introducing the approximation that $s\ \delta_s$ is proportional to $1/d$, where s is the specific surface area per weight of solid, the modified equation for undrainable liquid becomes

$$S^{\propto} = k\left(\frac{1-\epsilon}{\epsilon}\right)\left(\frac{1}{\bar{d}^2 G}\right)^{1/4}\left(\frac{\sigma \cos \Psi}{\delta_L}\right)^{1/4} \tag{18}$$

Where S^{∞} is the fraction of void volume occupied by liquid after infinite drainage time, k is an experimental coefficient, ε is the void fraction, d is the mean particle diameter, G is $\omega^2 r/g$, σ is the surface tension, and ψ is the wetting angle. Appreciable internal porosity of the particles can badly distort an experimental value of S.

For cakes of high permeability, the capillary drain height may be an insignificant fraction of cake thickness, and film drainage becomes the controlling factor in a centrifugal field (7). Under unsteady-state conditions, equation 19 represents the drainable liquid left in the cake as a function of the centrifugal filtration parameters:

$$S - S^{\propto} = \frac{3}{\pi}\,\frac{s'}{\epsilon}\,\omega\left(\frac{\mu_2}{\delta_L}\right)\left(\frac{h}{2\,r_m - h}\right)\left(\frac{1}{t - t'}\right)^{1/2} \tag{19}$$

where S is the fraction of void volume occupied by liquid at time t, s'; is surface area/volume of cake, h is cake thickness, t'; is time at which free liquid surface enters the cake, and μ_2 is the viscosity of the surrounding medium.

It is difficult to obtain void volume data for a cake under drainage conditions in a centrifuge. Prediction of these values from filter cake data is uncertain because the compressive force in centrifugation increases with radius throughout the cake depth, and the effective mass of a particle, proportional to ($\delta_S - \delta_L$) when the cake is submerged, becomes essentially δ_s after bulk drainage is completed. For engineering purposes, it is simpler to approximate the ratio of the volume of liquid to the volume of solid. Assuming that $G = \omega^2(2r_m - h)/g$, and $s' \sim (1 - E)/d$ as for spheres, equation 19 becomes

$$q - q^{\propto} = \left(\frac{k'}{d}\right)\left(\frac{\mu_L}{\delta_2}\right)^{1/2}\left(\frac{h}{Gt}\right)^{1/2} 100 \tag{20}$$

where q is the ratio of the volume of liquid to the volume of solid. This measurement is readily obtained from the volume percentage of liquid in cake and is also easily converted by a density ratio to a weight ratio of liquid to solid. The value for q^{∞} the ratio of the liquid volume to the volume of solid at infinite time, may also be applied when known.

Reasonable experimental agreement was obtained (7) using the exponents given in equation 20 for relatively slow-draining chalk and kieselguhr. Figure 4 (8) shows the effect of drain time after the disappearance of a free liquid head above the cake. The sharp break probably indicates completion of bulk drainage and start of drainage by film flow only. Drainage before the break is rapid and proportional to $t^{-1.9}$. After the break in the curve, the value $t^{-0.3}$ indicates that film drainage becomes controlling if low residual moisture is required. This exponent, -0.3, is appreciably lower than the -0.5 in theory but is close to -0.25, the value previously obtained (9). Considerable data indicate the validity of $G^{-1/2}$; limited data corroborate the theoretical exponent of 1/2 for kinematic viscosity.

Experimental exponents for cake thickness vary from 0.5 to as much as 3.0. The theoretical value of 1/2 may be approached only by incompressible cakes of a narrow range of sizes. The proper and characteristic value for the mean particle size, d, is difficult to ascertain. In practice, the most finely divided particles, eg, 10–15 wt% of solids, almost wholly determine the liquid content of a cake, regardless of the rest of the size distribution. It seems reasonable to use a d closely related to liquid content, eg, the 10% point on a cumulative weight-distribution curve.

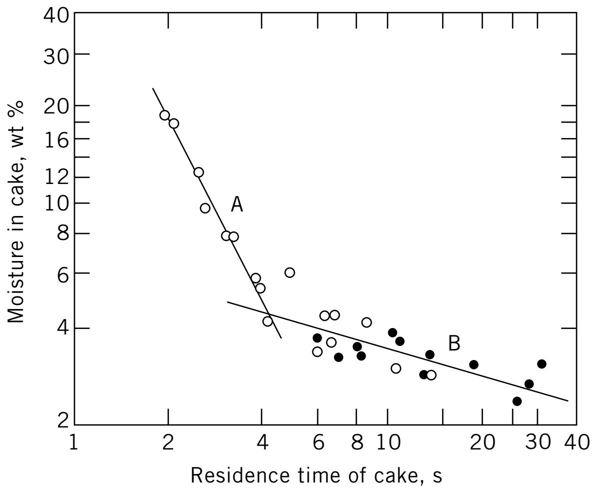


Fig. 4. Drainage of salt crystals in a cylindrical screen pusher-discharge centrifuge (8), where the cake thickness is 3.3 cm, the centrifugal field = 320 G, and the crystals 14 wt% <250 μm. (•) Represents moisture in the discharge cake, and (○) moisture in the cake by material balance with drainage flows; line A has slope = −1.9 and B, −0.3.

2.3. Separation by Compaction. In centrifuges used for dewatering compactable slurries, cake compaction follows the sedimentation of the suspended solids. Both solid bowl and perforated bowl centrifuges are used for dewatering compactable cakes. In a solid bowl centrifuge the expelled less dense liquid flows inwards countercurrent to the compacting cake, whereas it flows outwards through the compacting cake in a perforate bowl centrifuge, provided that the liquid pressure generated by rotation exceeds the capillary pressure. The bulk filtration rate of a filtering centrifuge can be estimated from equation 17 including accounting for the pressure drop over the filter medium and assuming an average cake permeability across the cake thickness.

In a decanter centrifuge the liquid released during compaction flows predominately radially inward and Darcy's equation can be used for estimating this liquid flow, with the following assumptions: (*1*) radial solids velocities are neglected, (*2*) cake motion due to scrolling is disregarded, and (*3*) the time required for full cake consolidation is much longer than the cake retention time in a centrifuge (10,11).

$$Q_0 = ((C_c - C_e)/(C_c - C_o))(K_m/\mu)(\delta_C - \delta_L)\pi l(r_3^2 - r_c^2)/\ln(r_3/r_c) \qquad (21)$$

$$Q_0 = \phi K_p \qquad (22)$$

Where K_p is a process constant and

$$\phi = \pi l(r_3^2 - r_c^2)/\ln(r_3/r_c) \qquad (23)$$

Comparing equation 23 with equation 9 and assuming that $R_c > R_1$, which is the normal operating mode for sludge compacting solid bowl centrifuges, one sees that the centrifuge related terms are identical, considering that equation 23 was developed for 100% solids capture, whereas, equation 9 assumes 50% cutoff. This in turn implies that scaling of solid bowl centrifuges based on sedimentation considerations (Σ concept) is identical to scaling based on cake compaction considerations.

2.4. Σ-Concept and Its Application. The assumptions and conditions for deriving equations 7, 12, and 16 impose limitations on the application of the Σ concept and fall into two groups. The first concerns the particulate material. Particles (or drops) are assumed to be spherical in shape and uniform in size. These should not deaggregate, deflocculate, coalesce, or flocculate during passage through the zone in which separation occurs. Initially, particles are evenly distributed in the continuous liquid phase, where their concentration is low enough for them to settle as individual particles without interaction. The settling velocity, v_s, of the particles is such that the Reynolds number does not exceed 1.0, thus ensuring that the deviation from Stokes' law does not exceed 10%. The settling velocity, v_g, in a gravity field, or v_s in a centrifugal field, is theoretically never reached because the accelerating time required for a particle to reach its terminal velocity is infinite. However, a particle of up to 100 μm approaches 90% of its terminal velocity in milliseconds. The time available for the particles to settle in a centrifuge is enough for each individual particle to be almost at its theoretical settling velocity, despite variation in G.

The second group of assumptions and conditions concerns flow conditions. Flow is assumed to be streamlined. Fresh feed is introduced uniformly into the full space available for its flow. In an imperforate bowl centrifuge, this condition requires that the continuous liquid phase immediately occupy the full liquid layer thickness between the free surface and the inside radius of the cylindrical wall. In a disk centrifuge, the continuous phase is assumed to divide evenly between all the disk spaces axially as well as circumferentially. In any imperforate bowl centrifuge, the continuous phase rotates everywhere at the same angular velocity as the bowl, ie, there is no forward or back swirl. The displacement of the flow pattern of the continuous phase by the layer of deposited material is neglected. Remixing at the interface of the separated material is negligible. Finally, the detrimental effect resulting from heavy separated material crossing the fresh feed stream outside of the disk stack is neglected.

Few of the assumed conditions are fully satisfied in practice. The last three items relate to potential interference between separated phases. Such interference can occur and leads to poor sedimentation performance if an excessive volume of the sedimented phase is retained in the centrifuge.

Excessive volume of solids may be retained in the bowl of conveyor centrifuges if (*1*) the conveyor volumetric displacement is not sufficient to handle the sedimentation rate of solids; (*2*) the sedimented solids cannot be successfully conveyed and discharged over the solids port until a sufficient layer has been built up inside the bowl; and (*3*) solids do not easily slide outwardly on the underside of the disk of a disk centrifuge.

In the case of the nozzle disk centrifuge, the flow of the solids phase through the discharge nozzles may be so restricted that an excessive layer can accumulate inside the bowl shell. When this layer reaches the zone utilized by the fresh feed stream entering the disk stack, reentrainment of the sedimented solids by the fresh feed may lead to poor sedimentation performance.

The sedimentation phenomenon that the Σ concept attempts to describe quantitatively is only part of the total task that the centrifuge has to accomplish. Thus, attempts to predict separation performance solely on the basis of Σ concepts have sometimes given disappointing results.

Nevertheless, the Σ concept is a valuable tool, allowing in theory a comparison between geometrically and hydrodynamically similar centrifuges operating on the same feed material. Equations 7, 12, and 16 show that the sedimentation performance of any two similar centrifuges having the same feed suspension is the same if the quantity Q_0/Σ; is the same for each. In practice, an efficiency factor, e, is often introduced to extend the use of Σ so as to compare dissimilar centrifuges. This factor takes into consideration differences in feeding, discharging, flow, turbulence, and remixing that exist in different types of centrifuges operating on the same feed material. The flow rate, $Q_{0\ 2}$, of a No. 2 centrifuge can thus be compared to the rate, $Q_{0\ 1}$, of a No. 1 centrifuge operating on the same feed. For equal sedimentation performance,

$$\frac{Q_{0_2}}{\Sigma_2 e_2} = \frac{Q_{0_1}}{\Sigma_1 e_1} \quad \text{or} \quad Q_{0_2} = Q_{0_1}\left(\frac{e_2}{e_1}\right)\left(\frac{\Sigma_2}{\Sigma_1}\right) \tag{24}$$

If the two centrifuges are geometrically and hydrodynamically similar, then $e_1 = e_2$ and equation 20 can be simplified to

$$Q_{0_2} = Q_{0_1}\frac{\Sigma_2}{\Sigma_1} \tag{25}$$

The Σ concept permits scale-up between similar centrifuges solely on the basis of sedimentation performance. Other criteria and limitations, however, should also be investigated. Scale-up analysis for a specified solids concentration, eg, requires knowledge of solids residence time, permissible accumulation of solids in the bowl, G level, solids conveyability, flowability, compressibility, limitations of torque, and solids loading. Extrapolation of data from one size centrifuge to another calls for the application of specific scale-up mechanisms for the particular type of centrifuge and performance requirement.

2.5. Other Sedimentation Scale-Up Equations. Some centrifuge suppliers use an area-equivalent, Ae, description instead of Σ; others use KQ or Lf_2 values. All of these are in units of area. For a disk centrifuge,

$$\Sigma_{\mathrm{D}} = \frac{2\pi\eta\omega^2}{3g}\ \cot\theta(r_2^3 - r_1^3)$$

$$KQ = \frac{2\pi n\omega^{1.5}}{3g}\ \cot\theta(r_2^{2.75} - r_1^{2.75}) \tag{26}$$

$$Lf_2 = \frac{2\pi n\omega^2}{3g}\ \cot\theta\left(r_2^3 - \frac{r_1}{r_2} - \left(\frac{r_1}{r_2}\right)^2 - \left(\frac{r_1}{r_2}\right)^3\right) \tag{27}$$

For an imperforate tubular centrifuge,

$$Ae_{3/4} = 2\pi l\omega^2 (0.75 r_3)^2 \tag{28}$$

$$\Sigma = \frac{2\pi l\omega^2}{g}\ (0.75\ r_3^2 + 0.25\ r_1^2) \tag{29}$$

All of these equations work in the scale-up of geometrically similar centrifuges. The KQ reduces the effect of rotational speed from ω^2 to $\omega^{1.5}$ and the disk radius from r^3 to $r^{2.75}$, based on empirical experience. The Lf_2 uses the projected cylinder area of the disks at their mean radius (see equation 13 of Σ_T for imperforate bowls. The parameter $Ae_{3/4}$ is the cylindrical area at three-quarter of the bowl wall radius).

When testing a new material for centrifugal separation, a bottle centrifuge is usually used to obtain the general G range needed, and to choose the centrifuge type and size. To estimate size, the Q_0/Σ_B must be determined for the bottle centrifuge (eq. 7), and then used in equation 25 to determine the Σ value of the centrifuge to be used. Efficiency factors for the various types of centrifuges (12) have been reported:

Disk bowl centrifuges	45–73 %
Scroll centrifuges	54–67 %
Tubular bowls	90–98 %

2.6. Factors Influencing Centrifugal Sedimentation. The sedimentation velocity of a particle is defined by equations 1 and 2. Each of the terms therein effects separation.

Viscosity. Sedimentation rate increases with decreased viscosity, μ, and viscosity is dependent on temperature. Often mineral oils, which are highly viscous at room temperature, have a viscosity that is reduced by a factor of 10 at 70–80°C. Tar, solid at room temperature, is a low viscosity liquid at 150–200°C and can be clarified of inorganic solids at high flow rates. Even the viscosity of water changes significantly when the temperature changes between 10 and 35°C.

Density Difference Between Particle and Liquid. Separation cannot take place if $\Delta\delta = 0$. Some mineral oils have the same density as water at room temperature. When heated to 80°C, the reduction of the density of water is less than that of the mineral oil, resulting in the water becoming heavier. Therefore separation is possible. Dilution of a liquid by a solvent, eg, molasses by water or heavy oil by naphtha, results in lower density and lower viscosity of the liquid. Solvent stripping may be required at a later stage in the process.

Particle Size. Doubling particle size, d, increases sedimentation by a factor of 4. Thus, methods to increase size are important. Additives are commonly used to flocculate many fines into an agglomerate that acts as a single large particle. Many chemical companies offer a wide range of organic and inorganic flocculating products, in dry or liquid form. Bench tests are usually required to

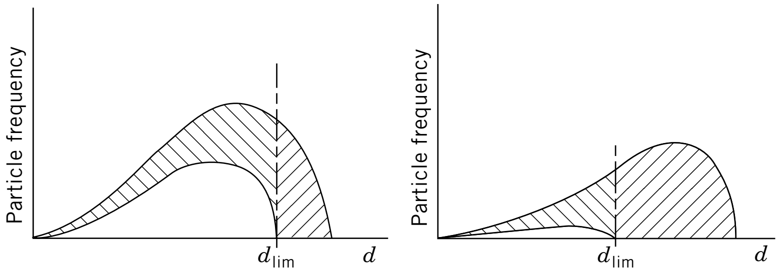


Fig. 5. Particle distribution (upper line) before and (lower line) after action of the separator where the cross-hatched areas represent the particles separated out. By definition, all particles of $d > d_{\mathrm{lim}}$ are separated out. A number of particles having $d < d_{\mathrm{lim}}$ are also separated. (**a**) Fine and (**b**) coarse particle dispersion (13).

determine best type and dose, ie, to optimize the flocculent choice. pH control, electrostatic devices, and mechanical coalascers are used to combine fine liquid drops in emulsions to produce larger particles. Special care must be exercised in pumping, and feeding mixtures of these easily breakable particle agglomerates, thus preventing the large particles from becoming fine before sedimentation.

Particle Shape. Whereas the Stokes' particle is assumed to be a sphere, very few real solids are actually spherical. Flat and elongated particles sediment slower than spheres. Normally an equivalent diameter would be used.

Particle Size Distribution. Almost every feed slurry is a mixture of fine and coarse particles. Performance depends on the frequency of distribution of particle size in the feed. Figure 5 shows that whereas all of the coarse particles having a diameter greater than some d_{lim} are separated, fewer of the very fine particles are, at any given feed rate. The size distribution frequency of particles in feed and centrate for a fine and coarse feed are quite different. More coarse particles separate out than fine ones. Classification of solids by size is often done by centrifugal sedimentation.

3. Liquid–Liquid-Phase Behavior

Liquid drops, suspended in a continuous liquid medium, separate according to the same laws as solid particles. After reaching a boundary, these drops coalesce to form a second continuous phase separated from the medium by an interface that may be well or ill defined. The discharge of these separated layers is controlled by the presence of dams in the flow paths of the phases. The relative radii of these dams can be shown by simple hydrostatic considerations to determine the radius of the interface between the two separated layers. The radius is defined by

$$r_{\mathrm{i}}^2 - r_{\mathrm{h}}^2 = \frac{\delta_{\mathrm{l}}}{\delta_{\mathrm{h}}}\ (r_{\mathrm{i}}^2 - r_{\mathrm{l}}^2) \tag{30}$$

where r_{i}, r_{h}, and r_l are the radii of the interface, the liquid surface at the heavy discharge dam, and the liquid surface at the light discharge dam, respectively, and δ_{h} and δ_{l} are the densities of the heavy and light phases, respectively. Control of the interface radius, achieved by varying r_h or r_l for the desired ratio, is an important factor in liquid–liquid separation, as it determines whether the heavy or light phase is exposed to the greater separating effect (13).

Equation 30 is accurate only when the liquids rotate at the same angular velocity as the bowl. As the liquids move radially inward or outward these must be accelerated or decelerated as needed to maintain solid-body rotation. The radius of the interface, r_{i}, is also affected by the radial height of the liquid crest as it passes over the discharge dams, and these crests must be considered at higher flow rates.

4. Centrifuge Components

4.1. Power, Energy, and Drives. Centrifuges accomplish their function by subjecting fluids and solids to centrifugal fields produced by rotation. Electric motors are the drive device most frequently used; however, hydraulic motors, internal combustion engines, and steam or air turbines are also used. One power equation applies to all types of centrifuges and drive devices.

The total power, P_{T}, needed to run a centrifuge, ie, delivered by the drive device, is equal to sum of all losses:

$$P_{\mathrm{T}} = P_{\mathrm{P}} + P_{\mathrm{S}} + P_{\mathrm{F}} + P_{\mathrm{W}} + P_{\mathrm{BD}} + P_{\mathrm{CP}} \tag{31}$$

where P_{P}, the process power, $= Q\delta\omega^2 r^2$; Q is the flow rate of liquid or solids; $\delta =$ mass density; and $r =$ discharge radius of material being discharged. Power for each liquid and the solid phase must be added to get P_{P}. P_{s}, the solids process power $= k_{\mathrm{p}} T_{\mathrm{C}} \cdot \Delta N$ for scroll decanters, where $T_{\mathrm{C}} =$ conveyor torque k_{p} is a conversion factor and $\Delta N =$ differential speed between bowl and conveyor. The parameter P_{F} is the friction power, ie, loss in bearings, seals, gears, belts, and fluid couplings. P_{W}, the windage power $= k_{\mathrm{s}}\mu^{0.2}\rho^{0.8}N^3D^{4.5}$ and $\mu =$ viscosity of surrounding gas; $\rho =$ density of gas; $D =$ rotor outside diameter; $N =$ bowl speed; and $k_{\mathrm{s}} =$ shape constant. Increased density owing to gas pressure increases the windage power, and this may be very significant for high pressure applications. Also, many hydrocarbon gases are heaver than air, resulting in high windage power. Very high G centrifuges often operate in a vacuum to avoid excessive windage power. For constant G, the scale-up windage power $\sim D^3$; for constant rotor stress, the scale-up of windage power $\sim D^{1.5}$.

Windage power is a very important loss for large machines and must be determined. Whereas windage power can be calculated from drawing dimensions (14), it is preferable to measure the windage power for an actual rotor, and then extrapolate using the formula given for windage power for a geometrically similar (larger or smaller) size. Doubling the size of a rotor while maintaining the g level results in eight times the windage power loss.

P_{CP} is the friction power consumed by the centripetal pump. The centrate kinetic energy is partially recovered by the pump, which delivers the centrate flow at a positive pressure. The added power must be supplied by the centrifuge main drive, but use of a centripetal pump avoids the need for a separate centrate pump. The power required to bring the feed to the centrifuge is supplied by a feed pump at the feed tank, not by the centrifuge drive. This power may be significant where the feed pressure required due to flow rates is high.

For scroll centrifuges having back-drives, P_{BD} is the back-drive power:

$$P_{BD} = k_p (T_C/R) N_{BD} \tag{32}$$

where T_C = conveyor torque; R = gear box ratio; and N_{BD} = back-drive speed. Depending on the type of gearbox and wether the conveyor is leading or lagging the bowl, the backdrive may be a driver or a break. In the case of a driving backdrive, P_{BD} in equation 32 is provided by the backdrive driver, otherwise P_{BD} is provided by the centrifuge drive device. Braking backdrives can be regenerative or non regenerative. In the former case some of the braking power is regenerated using a motor as a generator, otherwise the energy is dissapated as heat. Regenerative back-drives reduce total power consumption. A high gear box ratio results in lower back-drive power. Direct hydraulic motor conveyor drive devices get their power from an external hydraulic power supply, not from the main drive motor, and must meet the direct conveyor torque demands. The equation for direct hydraulic conveyor power is $P_{BD} = k_p T_C \Delta N$; however, hydraulic losses in the power supply, rotary seals, and the hydraulic motor must be added. At least one manufacturer avoids these losses by using a direct drive planetary gear box powered by a variable frequency ac motor.

The choice of the main drive, usually an ac motor, must include starting specifications. Various methods are used to start centrifuges. These include mechanical or fluid couplings, as well as wye-delta electric motor starters or variable frequency drives. Centrifuges are high inertia rotating devices, sometimes taking up to 15 min to accelerate to operating speed. Details of the mass moment of inertia, friction, and windage losses must be considered to specify a drive device. The inertia seen by the drive device, when comparing centrifuges operating at constant G, is proportional to D^4: Small rotors are easy to start, but large rotors must be carefully reviewed.

Disk centrifuges having nozzles to discharge the solids slurry through small backward-pointing nozzles must have this power included in the calculations. Some thickening scroll centrifuges also use such nozzles, usually with an intermittent flow. The parameter P_N is the nozzle power:

$$P_N = Q_N \delta_N \omega r_N (\omega r_N - v_N \cos \phi) \tag{33}$$

where Q_N = the volume rate of the material discharged through the nozzles, δ_N = the mass density of the solids slurry discharged through the nozzles, ω = the angular velocity of the centrifuge, r_N = the radius at which the nozzles discharge, v_N = the linear discharge velocity out of the nozzles, and ϕ = the angle measured between the direction of the nozzle and the tangent to the circle of radius, r.

The discharge velocity out of the nozzles is given by equation 34:

$$v_N = C(2p_N/\delta_N)^{1/2} \tag{34}$$

where $C =$ a discharge coefficient, and $p_N =$ the hydraulic pressure at radius r_N resulting from the rotation of the bowl. The pressure p_N is given by equation 35:

$$p_N = \omega^2 \pi (r_N^2 - r_1^2)/2 \tag{35}$$

where r_1 is the free-surface radius of the liquid phase, δ_p is the weighted average density of the process material in the bowl.

The coefficient C in equation 34 is a function of two phenomena. First, the presence of viscous friction accounts for a small loss of energy. Nozzle orifice contraction results in discharge coefficients which range from 0.5 to 0.85. The coefficient falls in the upper portion of this range when the length of the nozzle is two or three times its diameter, and in the lower end of the range where the nozzle diameter is more than five times its length.

Special vortex nozzle designs that deliver lower flows using a large opening have been used to reduce plugging problems. Also, viscosity-sensitive nozzles, where flow is increased as viscosity is increased, are used to control variation in solids concentrations.

Filtering centrifuges must consider the power needed to bring the solids to a final radius. The radius used to determine the centrate power consuption in equation 31 is the outside radius of the rotor supporting the filtering screen. Many filtering centrifuges discharge solids at a reduced bowl speed to avoid particle breakage. The main drive is often designed to recover the rotor kinetic energy during deceleration.

The energy absorbed by the liquid and solids stream, P_P, is transferred in the feed zone of the rotor or conveyor. One-half of the total energy is converted to the kinetic energy associated with the tangential velocity of the pond surface. This kinetic energy is dissipated as heat when the centrate is discharged to the stationary casing. The other one-half of the total energy is lost as turbulence in or near the feed zone. The intensity of this turbulence can break friable particles, or create tight emulsions of two immiscible liquids, both making the separation that follows more difficult. Reducing the pond radius reduces the total process power and particle degradation and thus improves total separation performance. Reducing the pond inside radius by 25% reduces power consumed by 44%.

4.2. Materials of Construction and Operational Stress. Before a centrifugal separation device is chosen, the corrosive characteristics of the liquid and solids as well as the cleaning and sanitizing solutions must be determined. A wide variety of materials may be used. Most centrifuges are austenitic or duplex stainless steels; however, many are made of ordinary steel, rubber or plastic coated steel, Monel, Hastelloy, titanium, and others. The solvents present and of course the temperature environment must be considered in elastomers and plastics, including composites.

Once the material choice based on corrosion is made, a careful analysis of the stresses produced by rotation for the particular type of centrifuge is required, so that for the given liquid and solids specific gravities a maximum operating speed can be determined. In general, the metals used are ductile and elastic in the operating speed range. The usual limits are the ultimate strength and yield strength (0.2% offset). The stresses of the centrifuge bowl are primarily tangential stress and axial stress. These are the result of the weight of the bowl material itself, and the need to contain the rotating process solids and liquids within the bowl.

The geometry of the bowl parts is important, and for intermittently discharging centrifuges, fatigue strength must be considered. In general the tangential stress in a bowl wall is σ_r.

$$\sigma_{\mathrm{T}} = \sigma_{\mathrm{self}} + \sigma_{\mathrm{p}} \tag{36}$$

where for a thin-shell bowl

$$\sigma_{\mathrm{self}} = K'\delta_{\mathrm{M}}\omega^2 r^2 \tag{37}$$

$$\sigma_{\mathrm{p}} = K'\delta_{\mathrm{p}}\omega^2(r^2 - r_{\mathrm{p}}^2) \tag{38}$$

where δ_M = mass density of bowl material; r = bowl radius; δ_p = maximum density of process material; r_p = smallest radius of the process material (assuming that the bowl is full of process material); and K' is a function of bowl geometry and includes stress concentrations owing to changes in section, holes, slots, fillets, etc. The total stress increases with the square of bowl speed, ω^2, and the pond radius, and is directly proportional to the liquid and solid density.

Within the bowl, the pressure, p, developed also exerts axial separation forces. The internal pressure in a bowl is shown in equation 39:

$$p = \pi\omega^2(r^2 - r_{\mathrm{p}}^2)/2 \tag{39}$$

The axial projected area is $A = \pi(r^2 - r^2{}_p)$ and the average pressure is 1/2 p. The axial force $F_A = pA/2$, where

$$F_{\mathrm{A}} = \delta_{\mathrm{p}}\omega^2\pi(r^2 - r_{\mathrm{p}}^2)/4 \tag{40}$$

The stress resulting from the axial forces must be considered in analyzing the parts that resist axial separation, such as bolts, nuts, rings, etc. On scroll centrifuges the axial force owing to the axial component of the conveyor torque must also be considered.

Typically, the total stress on the bowl is 50–65% self-stress, and 35–50% process stress. The axial stress is usually 5–10% of the tangential stress. When the bowl material density is low (such as titanium), or where solids are heavy (as is coal), or when the pond surface is close to the bowl wall (as when using shallow pond scroll centrifuges), these proportions differ. All centrifuges must have a factor of safety against general yeilding and rupture. The ratio of the material strength to the actual stress is the factor of safety. Factor of safety is set by the manufacturer, and sometimes, especially in Europe, by government regulation (15). Every centrifuge supplier sets the limits of bowl speed, temperature, pressure and liquid, and solids density. On some very high G preparative or zonal centrifuges, the number of cycles of bowl use must be recorded and the rotor retired after a given number, because fatigue determines the bowl life.

Most disk and scroll centrifuges are made from forgings, centrifugal castings, or fabrications. Each manufacturer must carefully monitor the mechanical properties of the bowl material to ensure that the required factor of safety is maintained as well as the specified ductility. A rotor failure during operation is very serious, and can not only destroy the centrifuge, but also damage nearby equipment and injure operators.

Many of the centrifugal separation applications are abrasive and erosive to centrifuge parts because of the high relative velocity or high contact pressure between the particles and exposed parts. Areas of wear must be protected using materials that resist both mechanical and chemical attack. In general, the main areas of abrasive wear are the feed zone surfaces, where nonrotating process slurries are accelerated to speed; the tips and faces of the conveyor flights on decanter centrifuges; solids discharge openings; screens, where solids slide across the screen; and stationary collection surfaces that receive the impact of material being discharged from the bowl. Many abrasion resistant materials are used. Examples are sintered tungsten carbide bound with cobalt or nickel; sintered aluminum oxide ceramic; sprayed and fused or weld-deposited hard-surfacing alloys; and elastomeric coatings or inserts. There is great variation in the design and application of these materials that effect the actual working life in service and the cost to rebuild the worn areas or replace the worn insert.

Maintenance and operating costs are lower if replaceable wear protection is used, compared to the requirement to rebuild (and thus rebalance) the worn parts. In very severe applications, such as coal (qv) or coal refuse dewatering that use screens, wedge wire screens made from ceramic or carbide materials are required. The use of carbide or ceramic flight-tip protection on decanter centrifuges has permitted economic use in applications such as dewatering (qv) tar sands as well as coal and coal refuse, and dewatering and thickening mixed primary and secondary sewage sludges. Continuous nozzle discharge disk and decanter centrifuges would not be feasible without replaceable ceramic or carbide nozzle inserts. The use of relatively soft abrasion-resistant elastomers, such as urethanes, has been successfully applied in the solids receiver housing and the feed zone targets of scroll centrifuges. Hard chrome plating has been used on the first few disks of disk centrifuges. Stellite or carbide inserts are often used at the solids discharge opening.

4.3. Noise. Centrifuges, as do any rotating equipment, create noise. When the motion of air or gas entrained by a rotating bowl shell is deflected or otherwise disturbed, its energy is transferred to the environment through the casing or chutes. This mechanism suggests that the noise level created by the bowl is related to the surface linear speed of the rotor. High linear speed is important for maintaining separating capacity, so the noise level should be reduced without reducing the rotational speed, if possible.

In addition to surface speed, rotor imbalance, surface irregularities, clearance between covers and rotor, resonance in the supporting structure, conditions of installation, and particularly the drive motor contribute to centrifuge noise. An inadequate supporting platform can amplify centrifuge vibration. Open piping and venting and discharge connections allow noise generated inside the unit to escape. Discharge connections should be tight, yet flexible enough to prevent transfer of vibrational energy to plant piping. Size, spacing, materials of construction, and other properties of the centrifuge room also affect noise level. Sound-absorbing materials or enclosures should be provided when other means are inadequate.

The electric motors are often the noisiest component of the centrifuge assembly. Most standard motors in the 75–250 kW range develop noise levels of 85 dbA (weighted sound pressure level using filter A, per the ANSI standard). A quiet motor can reduce this level by 5 dbA and should be used whenever noise is of concern.

4.4. Equipment. Centrifugation equipment that separates by density difference is available in a variety of sizes and types (16,17) and can be categorized by capacity range and the theoretical settling velocities of the particles normally handled. Centrifuges that separate by filtration produce drained solids and can be categorized by final moisture, drainage time, G, and physical characteristics of the system, such as particle size and liquid viscosity.

For optimum results, a combination of several types of equipment may be used, eg, a gravity separator for oil recovery from sludge at a petroleum refinery. The sludge, an aqueous suspension of 1–5% oil and 5–30% solids <50 μm, is screened to remove trash, and degritted in a cyclone to eliminate the coarse solids that would cause excessive abrasion. A decanter centrifuge then removes 60–70% of the solids in oil-free condition. The resulting oil-in-water emulsion, stabilized by residual fine silt, is passed through a 0.25-mm (60-mesh) screen, and sent to a disk centrifuge that discharges an oil stream at 0.5–2% bottom sediment and water, and an oil-free peripheral nozzle discharge containing the remaining solids in water.

Equipment Materials and Abrasion Resistance. Stainless steel, especially Type 316, is the construction material of choice and can resist a variety of corrosive conditions and temperatures. Carbon steels are occasionally used. Rusting may, however, cause time-consuming maintenance and can damage mating locating surfaces, which increases the vibration and noise level. Titanium, Hastelloy, or high nickel alloys are used in special instances, at a considerable increase in capital cost.

Abrasion, a serious problem in some applications, requires the addition of hard-surfacing materials to points exposed to abrasive wear (18). The severity of wear depends on the nature, size, hardness, and shape of particles as well as the frequency of contact, the force exerted against the wearing parts, and solids loading as related to feed rate and solids concentration.

A wide range of abrasion-resistant materials is available. Nickel–chrome–boron and cobalt–chrome–tungsten hard-surfacing alloys have been used for many years. Composite coatings of nickel-base alloys containing crushed tungsten carbide particles, applied by flame spraying and fusing, are also used. Solid tungsten carbide, pressed to shape and sintered at high temperatures, provides the best protection. Tungsten carbide plates, previously induction brazed, or bonded are now in light of superior corrosion resistance more suitably vacuum brazed to stainless steel supports, which in turn can be easily welded to portions of a centrifuge such as conveyor flights. Ceramics have been used where minor impact and abrasive particle pressures are involved.

5. Centrifuges

Like other manufacturers, those building sedimentation and filtration centrifuges are subject to acquisition and merger. In order to find updated information it is perhaps best to check on line sources such as those listed in Reference 19.

5.1. Sedimentation Equipment. Centrifugal sedimentation equipment is usually characterized by limiting flow rates and theoretical settling capabilities. Feed rates in industrial applications may be dictated by liquid handling capacities, separating capacities, or physical characteristics of the solids. Sedimentation equipment performance is illustrated in Figure 6 on the basis of nominal clarified effluent flow rates and the applicable Q_0/Σ values. The latter are equivalent to twice the theoretical gravity settling velocities. In liquid–solid separations, the effluent rate represents the clarified stream of the liquid medium and does not include the volume of solids discharged or the volume of medium discharged with the solids. The effluent rate of liquid–liquid separation refers to the clarified, heavy, or light continuous phase that usually occupies the greater volume within the separating equipment. The flow range for a particular piece of equipment does not represent its absolute limitations, but the

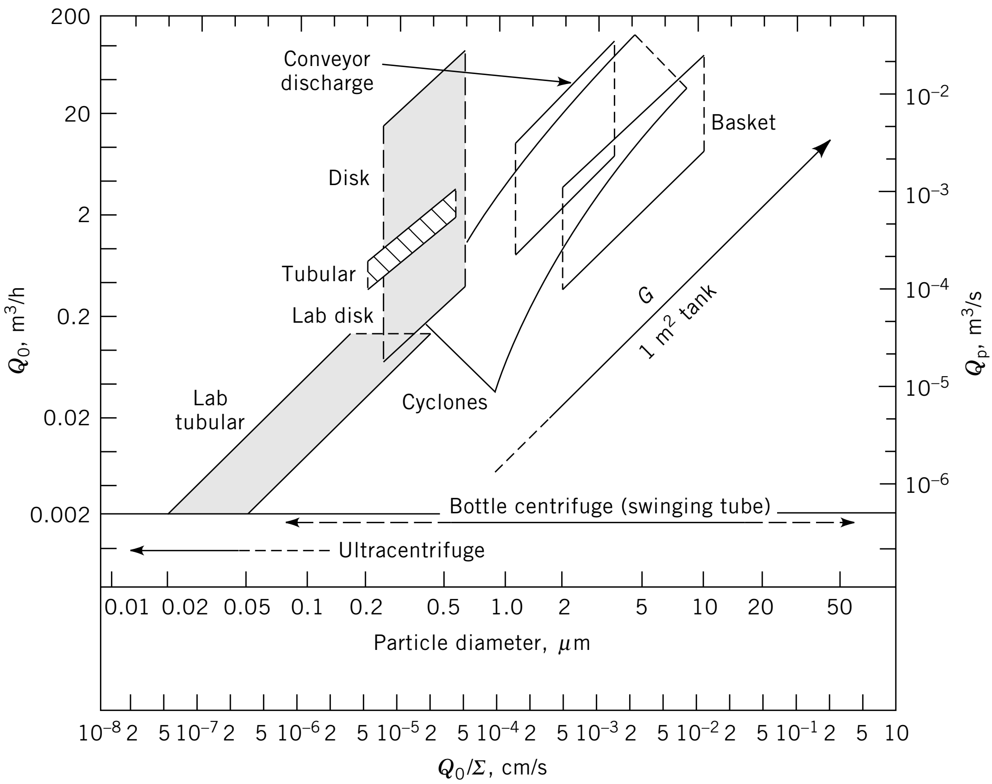


Fig. 6. Sedimentation equipment performance where the particles have a $\Delta\delta$ value of 1.0 g/cm^3 and a viscosity, μ, value of 1 mPa·s (=cP). The value of Q_0/Σ is twice the settling velocity at $G = 1$, and Q = overflow discharge rate in measurements given.

normal flows for good clarification in standard applications. Similarly, large particles can always be sedimented.

As an additional guide, the Q_0/Σ values are correlated with the equivalent spherical particle diameter by Stokes' law, as in equation 1. A density difference $\Delta\delta$ of 1.0 g/cm^3 and a viscosity of 1 mPa · s (=cP) are assumed, thus conversion to other physical characteristics of the system requires that the particle size scale be adjusted to equate a particle of 1.0-μm diameter to its Q_0/Σ in cm/s, according to the relationship $Q_0/\Sigma = 10^{-7} \times 1.09\ \Delta\delta/\mu$, for $\Delta\delta$ in g/cm^3, and viscosity μ in Pa · s. For interpretation of the particle sizes, the scale refers to the 50% cutoff particle size, and under actual centrifugation conditions the value of Σ, determined from Figure 6, must be increased by efficiency factors to give the theoretical value of Σ.

Figure 6 serves as a guide to the types of equipment that can handle a given separation. Other characteristics further narrow selection. For example, for the separation at $Q_0 = 3.5 \times 10^{-3}$ m^3/s (50 gpm) of kaolin clay solids from an aqueous suspension, where the particle density is 2.55 g/cm^3 and the size ranges from 0.25 to 30 μm with 55% >2 μm, the 1.0-μm point on the particle size scale would be equivalent to $Q_0/\Sigma = 1.69 \times 10^{-4}$ cm/s. Assuming that a high recovery is desired, a disk centrifuge is required, and recovery of most particles >0.4 μm is satisfactory, the Q_0/Σ equivalent to 0.4 μm on the adjusted scale is about 2.3×10^{-5} cm/s. Using a disk machine efficiency of 40%, the centrifuge needed would have a Σ value of

$$\Sigma = \frac{3.15 \times 10^{-3} \times 10^6}{0.40 \times 2.3 \times 10^{-5}} = 34.3 \times 10^7\ \text{cm}^2 \tag{41}$$

Because clay tends to pack hard, only the continuous nozzle discharge bowl would be satisfactory. Intermittently discharging disk centrifuges could not be used.

If classification of solids were desired, several other types of centrifuges could be used as well, assuming that only particles over ~2 μm were to be removed from the suspension and that the oversize stream should be highly concentrated for disposal. Figure 6 shows that a decanter or basket centrifuge or standard cyclone could theoretically be used. Because cyclones cannot concentrate the oversize as much as the centrifuges, these are less satisfactory for this example. If the feed concentration were low, eg, less than a few percent solids, a basket centrifuge would be used with intermittent discharge of solids. A decanter centrifuge gives almost as good a concentration of oversize as the basket and is a more efficient classifier. The decanter centrifuge would thus be the better choice and is actually used in the kaolin industry.

In general, solids-retaining batch and batch automatic machines are limited to low feed concentrations to minimize the time required to unload the solids. Continuous disk centrifuges can have higher feed concentration. The limit is the underflow concentration. Conveyor discharge centrifuges can handle high feed concentration and are limited only by the volume of solids displacement, or torque capacity.

Using flocculent to create aggregated solids, varying degrees of deflocculation of the solid particles may occur during acceleration in the centrifugal field.

An additional problem is the removal of the resulting soft, slimy cakes under centrifugal force. Scroll centrifuges have been successfully used to discharge soft, slippery solids continuously. Specially modified, automated basket centrifuges can handle a broad range of soft sludges, often without polymer addition. Disk centrifuges are particularly well/suited for clarification of streams containing solids such as aluminum hydroxide or a secondary waste-activated sludge.

The disk centrifuge having high capacity and G level is normally used for separating a liquid–liquid mixture or for clarification of such a mixture containing fine solids (20). Specially modified conveyor centrifuges are also used for three-phase separations. Settling velocity is a criterion of selection, but the actual separating of emulsified liquid–liquid mixtures may not strictly follow a settling theory. To break an emulsion, a threshold level of centrifugal force may have to be exceeded. In addition, drainage of liquid from the continuous-phase film of the emulsion has a time factor. Centrifugal force and time cannot be calculated interchangeably as Σ theory would indicate. In centrifugation equipment, coalescence of the dispersed liquid occurs coincidentally with its separation because there is neither time nor space for appreciable interfacial retention of unbroken emulsion.

Batch equipment, such as the bottle centrifuge or ultracentrifuge, does not have a real throughput capacity. By increasing the time of operation, according to Figure 6, the smallest particle size of solids usually sedimented in the bottle centrifuge may be 0.1 μm, at which size Brownian movement controls. In the ultracentrifuge, separation can be achieved down to molecular size, perhaps 0.005 μm, where Brownian movement is controlling. There is clearly no limitation to the larger particles that may be settled in bottle centrifuges so that an arbitrary upper limit is indicated for practical minimum conditions of 1000 rpm and 10 s. Similarly, for the ultracentrifuge the upper limit for Q_0/Σ was estimated for minimum conditions of 1 h and 5000 rpm.

Commercial sedimentation centrifuges are characterized principally by how solids are discharged, and the general dryness of these solids. There are batch and automatic batch solid bowl machines which collect the solids at the bowl wall. Solids are removed very dry. Almost any solid is collectable, even those that are very soft and compressible.

Disk-type solid bowl machines are batch, batch automatic, and continuous. The solids are removed in many different ways, but are usually wet. Scroll centrifuges discharge solids continuously and usually drier than disk and imperforate batch types. Generally, disk centrifuges have the highest values of Σ or KQ for a given size and therefore the best ability to collect fine particles at a high rate.

Bottle Centrifuge. A bottle centrifuge is designed to handle small batches of material for laboratory separations, testing, and control. The basic structure is usually a motor-driven vertical spindle supporting various heads or rotors. A surrounding cover reduces windage, facilitates temperature control, and provides a safety shield. Accessories include timer, tachometer, and manual or automatic braking. Bench-top bottle centrifuges operate at 500–5000 rpm, producing centrifugal fields up to 3000 G in the lower speed range, and operate up to 20,000 rpm with 34,000 G in the high speed units. Larger models operate up to 6000 rpm and develop 8000 G, using special attachments that permit 40,000 G.

These models may also be equipped with automatic temperature control down to -10°C and other programmable controls to manage the cycle.

There are three types of rotors: swinging bucket, fixed-angle head, or small perforate or imperforate baskets for larger quantities of material. In the swinging bucket type, the bottles are vertical at rest but swing to a horizontal radial position during acceleration so that solids are deposited in a pellet at the bottom of the tube. Although sedimenting particles must travel up to the full depth of the liquid layer, which requires appreciable time, the long path of travel and the perpendicularity of the sedimenting boundary to the axis of the tube are distinct advantages in effecting fractional sedimentation. Heads carrying fixed tubes at a 35–50° angle reduce centrifuging time because the maximum distance traveled by a particle is the secant of the tube angle times the diameter of the tube. Particles strike the wall and slide down the tube to collect near the bottom, but the angle makes it difficult to measure relative volumes of supernatant liquid and sedimented solids. Rotors carry 2–16 metal containers having tubes and bottles of various sizes and shapes. Containers range in capacity from capillaries for microanalysis to a 1 L maximum, limiting the batch capacity of this type of centrifuge to 4 L. Although glass bottles and tubes are generally used, plastic and metal containers are available for high speed operation or corrosive liquids. Tubes are usually cylindrical, tapered, and graduated; special shapes for analytical work are available, including pear-shaped tubes having capillary tips for measuring small quantities of solids.

The bottle centrifuge is primarily used in the laboratory to separate small quantities of material. It is also used for standard analyses including many ASTM methods and in preliminary testing for scale-up to commercial centrifuges. The Σ_B value for free-swinging tubes is determined by equation 7 and the Q_0/Σ_B value by equation 8; Q_0/Σ_B data can be prepared by bottle centrifuge (21). The bottle centrifuge has been used to study waste treatment sludges for estimation of cake concentrations and feasibility of handling in a conveyor centrifuge (22). Because compaction of solids is largely a function of the centrifugal field force and the exposure time, the bottle centrifuge has been used to study these parameters. The results are not always in the range applicable to full scale industrial equipment which also utilizes the motion of the cake. Similarly, drainage of packed solids can be studied by using tubes with fretted glass or perforated metal bottoms. Closed containers should be used to prevent drying by windage.

Specialty rotors permit ordinary bottle centrifuges to achieve some of the results previously considered possible only in ultracentrifuges. A modified zonal rotor, shown in Figure 7, permits collection of sediment using continuous addition of feed and discharge of centrate.

Preparation Ultracentrifuge. Preparation ultracentrifuges are suitable for a range of applications, such as processing quantities of subcellular particles, viruses, and proteins (qv). Many design variations are available and only the common features are considered here. Preparation ultracentrifuges range in operating speed from 20,000 rpm, generating ~40,000 G, to 75,000 rpm and ~500,000 G. The rotor is surrounded by a high strength cylindrical casing and underdriven by an electric motor. To avoid overheating of the rotor by air friction at these speeds, the pressure in the casing is reduced to ~0.13 Pa (1 mm Hg).

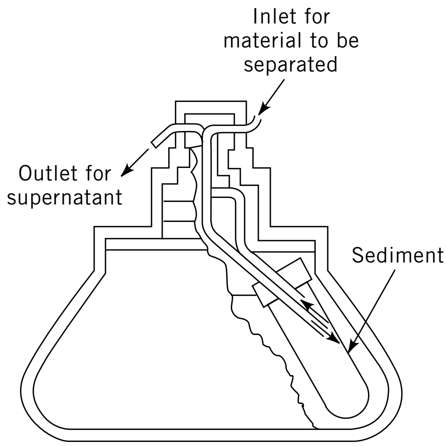


Fig. 7. Tube-type continuous-flow rotor. Courtesy of Sorvall.

Sensors (qv) monitor the temperature and a cooling system controls the temperature in the range of -15 to 30°C within $\pm 1^{\circ}$C. Electronic controls maintain the rotor speed within a required narrow range and may be automatically programmed for sequential changes in speed, including control of the acceleration and deceleration (23).

Preparation ultracentrifuges are guaranteed for several billion revolutions and can be rebuilt using relatively few parts. Among the great number of rotors available are batch rotors and those accepting feed and discharging centrate continuously during rotation. Batch rotors include angle and swinging-bucket types as well as those having vertical tubes parallel to the axis of rotation, which present a very short sedimenting distance and time requirement. Swinging-bucket rotors are also used for density-gradient separations or volume evaluation of the settled cake.

Separation by selective sedimentation on the basis of size and density of the particles may be satisfactory for polydisperse particle systems. However, the cake contains a range of material depending on its starting position in the container. Selectivity of separation can be improved by introducing the sample near the surface after the container is up to speed. Reslurrying and recentrifuging may be necessary to achieve purer fractions. Isopycnic separation improves initial separation efficiency where particles differ in density. If the density of the medium is intermediate to the range of densities of particles, higher density particles settle, whereas others remain suspended or rise regardless of size.

Zonal Centrifuge. The use of density gradients in centrifuge rotors greatly increases the sharpness of separations and the quantities of material that can be handled. In principle, the density gradient is established normal to the axis of rotation of the rotor and the highest density is located at the outer radius of the rotor. Low molecular weight solutes such as cesium chloride, sucrose, or potassium citrate, which are compatible with many systems in solution, are frequently used. A natural gradient may be formed by introducing a homogeneous solution and centrifuging for long periods of time. Continuous or step gradients may also be formed by introducing successive layers of solution,

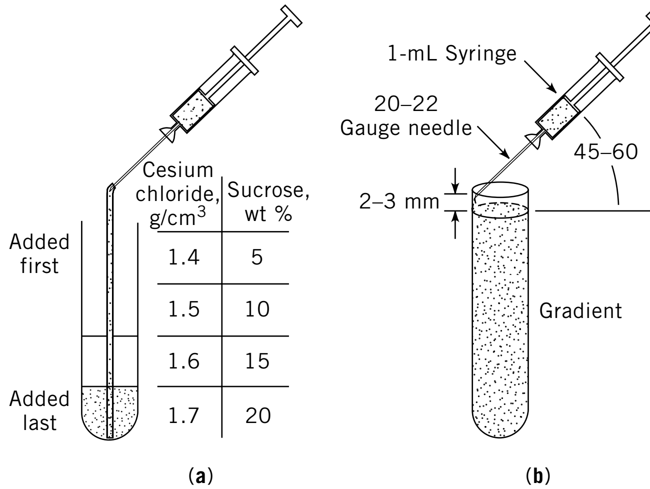


Fig. 8. (**a**) Forming a gradient and (**b**) applying the sample to the gradient before inserting tubes in a centrifuge rotor.

the composition of which varies continuously or stepwise from low to high density, where the latter displaces the former toward the center of the rotor (17,18) (Fig. 8).

In the simpler rotors using batch containers having swinging, angle, or vertical tubes, the gradient is introduced while the rotor is at rest and then accelerated to speed. The gradient shows relatively little mixing. Slowing the rotor gradually at the end of the run allows retention of the gradient and permits collection of the material banded isopycnically (Fig. 9).

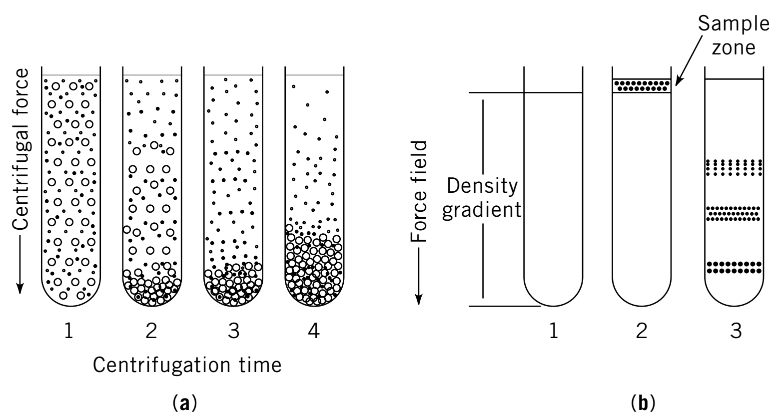


Fig. 9. (**a**) Differential centrifugation (pelleting), where time 1 < time 2 < time 3 < time. (**b**) Rate zonal separation in a swinging-bucket rotor, where tube 1 represents the density gradient solution, tube 2 the sample plus the gradient, and tube 3 the separation of sample particles under a centrifugal force, where the particles move at differing rates, depending on their mass.

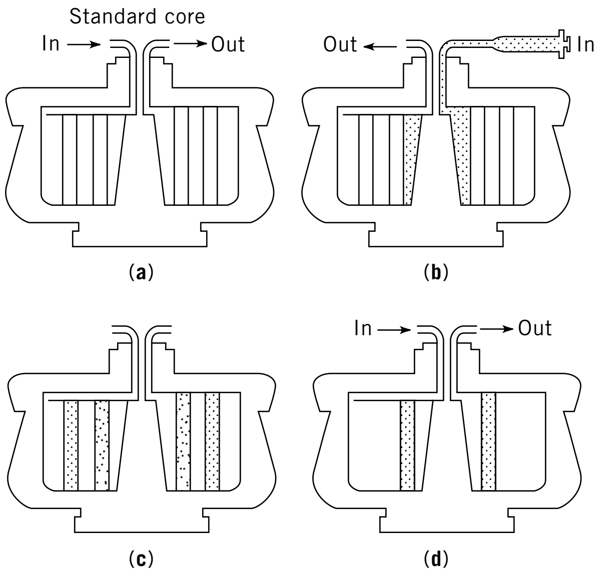


Fig. 10. Dynamic loading and unloading of a zonal rotor. (**a**) Gradient is loaded while rotor is spinning at 2000 rpm; (**b**) a sample is injected at 2000 rpm, followed by injection of overlay; (**c**) particles separated when the rotor is running at speed; and (**d**) contents are unloaded by introducing a dense solution at the rotor edge, displacing fractions at the center.

More sophisticated rotors can be loaded with gradient and sample while rotating. When the batch is finished or the bands are sufficiently loaded with material, the bowl may be stopped slowly and the reoriented layers displaced under static conditions. Rotors may also be designed to establish gradients and isopycnic bands of sample and then be unloaded dynamically by introducing a dense solution near the edge of the rotor as shown in Figure 10.

Particles in the gradient may be separated on the basis of sedimentation rate; a sample introduced at the top of the preformed gradient settles according to density and size of particles, but the run is terminated before the heaviest particles reach the bottom of the tube. If the density of all the particles lies within the range of the density limits of the gradient, and the run is not terminated until all particles have reached an equilibrium position in the density field, equilibrium separation takes place. The steepness of the gradient can be varied to match the breadth of particle densities in the sample.

Rotors are made of titanium or aluminum and may be cylindrical or bowl-shaped (see Fig. 10). Larger bowls reach 100,000 G; smaller units reach 250,000 G. The tubular rotors permit feed rates up to 60 l/h at 150,000 G or 120 l/h in a larger unit at 90,000 G. Such centrifuges may be used to separate relatively large quantities of viral material from larger quantities of cellular and subcellular matter, as, for example, in the production of vaccines (see VACCINE TECHNOLOGY).

Tubular Centrifuges. Tubular centrifuges (Fig. 11) separate liquid–liquid mixtures or clarify liquid–solid mixtures having less than 1% solids content and fine particles. Liquid is discharged continuously, whereas solids are removed

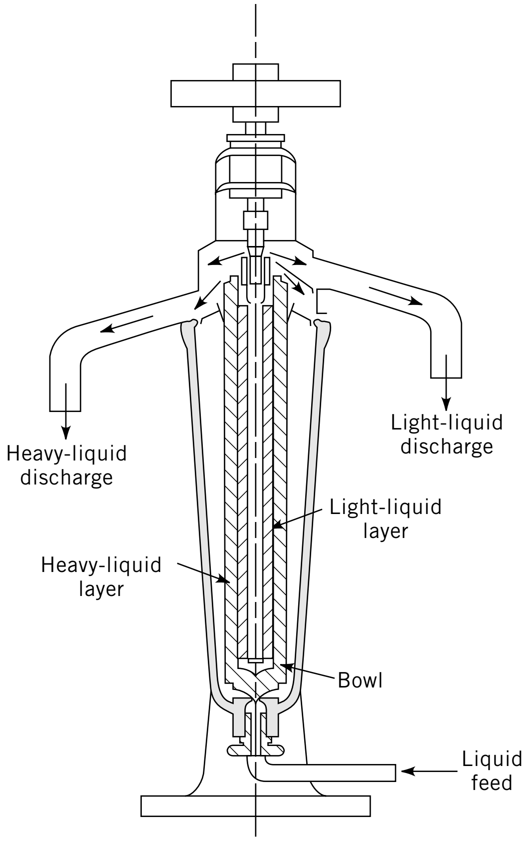


Fig. 11. A tubular centrifuge. Courtesy of Alfa Laval Inc.

manually when sufficient bowl cake has accumulated. For industrial use, the cylindrical bowls are 100–180 mm in diameter with length/diameter ratios ranging from 4 to 8. Bowl speeds up to 17,000 rpm generate centrifugal accelerations up to 20,000 G at the bowl wall (20). Because of the small bowl diameter, however, Σ_T values according to equation 11 result in flow rates in the range of 0.1–4 m^3/h (0.5–16 gpm). The tubular centrifuge handles low to medium flows and theoretical particle settling velocities in the range of 5×10^{-6} to 5×10^{-5}cm/s. Clean in place (CIP) and sterilize in place (SIP) are available. SIP usually is done at 121–124°C, and slightly >1.5 bar. Drive motors of 1.5–7.5 kW are used.

The laboratory tubular centrifuge is similar to the industrial model. It operates with a motor or turbine drive at speeds to 50,000 rpm, generating 65,000 G at the latter speed in the 4.5-cm diameter bowl. The nominal capacity range is 30–2400 cm^3/min. This centrifuge is uniquely capable of separating far finer particles than any other production centrifugation equipment except the bottle centrifuge. It is widely used in the production of flu virus.

A long, hollow, cylindrical bowl is suspended by a flexible spindle and driven from the top as shown in Figure 11. Axial ribs in the bowl ensure full acceleration of the liquid during its short time in the bowl. Feed is jetted into the bottom of the bowl and clarified liquid overflows at the top, leaving deposited solids as compacted cake on the bowl wall. The clarifying performance of the bowl is reduced as the deposited cake decreases the effective outer radius of the bowl in accordance with equation 12. Consequently, cake capacity of the industrial model is limited to 0.1–10 l. For liquid–liquid separation, the interface position (eq. 30) is determined by selection of ring-dam diameter or by the length of a hollow nozzle-type screw dam.

The tubular centrifuge was long used for the purification of contaminated lubricating oils because of the high centrifugal force developed and the simplicity of its operation. Colloidal carbon and moisture are removed from transformer oils to maintain dielectric strength; carbon and acid sludges are removed from diesel engine lubricating oils; and water and solid contaminants are removed from steam turbine lubricating oil. Polishing operations include the removal of small quantities of solids in the clarification of varnish, cider, fruit juices (qv), and even highly viscous chicle. In vegetable oil refining, oil losses in the semisolid soap stock are kept low by compaction of the soap phase under high centrifugal force. Automatic disk centrifuges which do not require manual solids unloading have largely replaced the batch-operating tubular. The laboratory tubular centrifuge is used to recover fine solids in batch preparations too large for bottle centrifuge separation, to estimate scale-up rates in larger centrifuges, and to analyze particle size distributions involving settling rates too low for feasible gravity sedimentation (24). Modern units having variable-frequency drives are available (20).

Disk Centrifuge. Centrifuges that channel feed through a large number of conical disks to facilitate separation combine high flow rates with high theoretical capacity factors (see Fig. 6). For industrial units flow rates up to 250 m^3/h (1100 gpm) can be obtained on easy separations, and theoretical settling velocities may range from 8×10^{-6} to $\sim 5 \times 10^{-5}$ cm/s. Both liquid–liquid and liquid–solid separations are performed using feed solids concentration <15% and small particle sizes. As seen from equation 15, the theoretical capacity factor depends on the number of disks, which is limited by the height of the disk stack. The performance is proportional to the cube of the disk diameter.

Several of the assumptions in the development of Σ_D do not apply in practice (25), and mathematical representation of the actual flow pattern has been difficult to achieve. Computer studies of the flow to and between the disks, as well as experimental analysis of flow patterns within the disk stack, have improved the effectiveness of disks. Not all disk machine designs can be compared using Σ_D alone. Details such as type and size of disk spacer, number of spacers, and location of the feed holes with respect to the spacers and solids discharging ports are very important. Also the method of feed acceleration is especially important in liquid–liquid separations and in the presence of fragile solids. If the disk centrifuge design is geometrically similar, scaling up by Σ_D from one speed and size of stack to another is reasonably accurate.

The outstanding feature of the disk bowl design is a stack of thin cones, commonly referred to as disks, which are separated by thin spacers. These are so arranged that the mixture to be clarified must pass through the disk stack before discharge. The resulting stratification of the liquid medium greatly reduces the sedimenting distance required before a particle reaches a solid surface and can be considered removed from the process stream. The angle of the cones to the axis of rotation is great enough to ensure that solid particles deposited on the surfaces slide, either individually, or as a concentrated phase according to the difference between their density and that of the medium.

The general flow patterns for a liquid–liquid separator and a recycle clarifier, respectively, are illustrated in Figure 12. Feed enters near the center of the bowl from either the top or the bottom, depending on the support, and is accelerated by vanes or disks (26) to the radius at which it enters the disk stack. When the disk stack is used for one phase, as in clarification or classification, the feed is distributed to the stock through the zone between the outer edges of the disks and the bowl wall. The clarified medium is discharged at a relatively small radius, generally at the top of the bowl. For a liquid–liquid separation, with or without solids, feed is distributed by a number of feed channels. The interface of

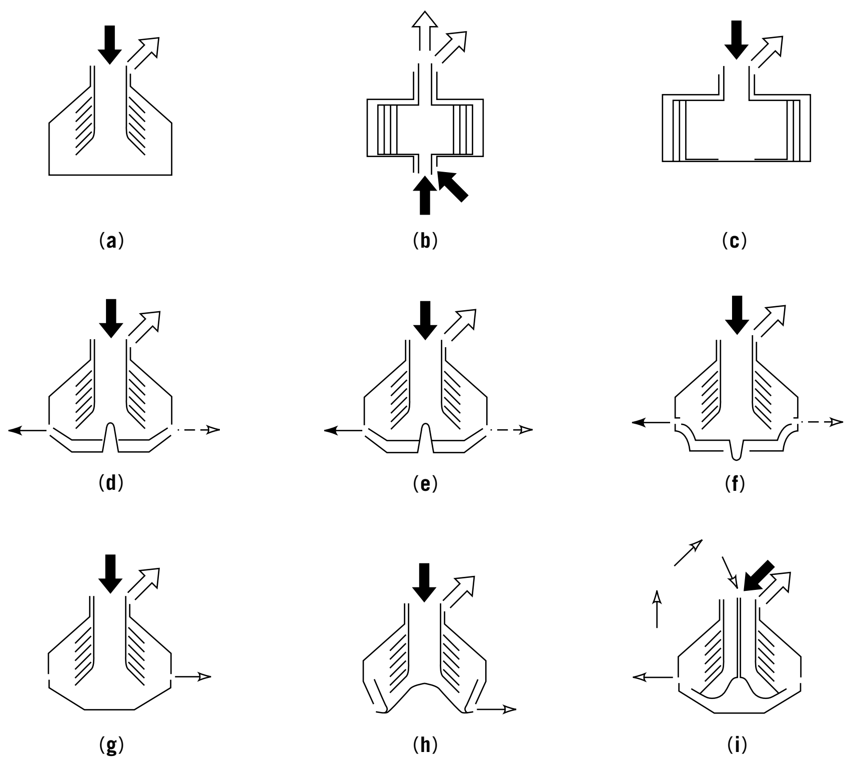


Fig. 12. Disk centrifuge bowls, where bowl diameters range from 10 to 90 cm; feed flow rates from 0.06×10^{-3} to 38×10^{-3} m^3/s (1–600 gpm); and operating temperatures from 10 to 90°C, and (➡) represents the main inlet, (⇨) the main outlet, (←) continuous flow solids, (· ⟶) intermittent flow solids, and (→) auxiliary liquid. (**a–c**) Solid wall bowl: base design, spiral cylinder inserts, and cylinder inserts, respectively; (**d–f**) bowl for intermittent solids discharges: base design, radial peripheral parts, and shoe; peripheral parts and shoe with nozzles for both continual and intermittent solids discharge; and axial peripheral parts and shoe, respectively; and (**g–i**) bowl with nozzles: base design and peripheral nozzles, nozzles at reduced diameter, and peripheral nozzles and solids circulation, respectively.

the two coalesced liquid phases is located at the disk feed holes, by appropriate selection (eq. 26) of the heavy- and light-phase discharge radii. During handling of the two liquids, the heavier moves toward the edge of the disks and the lighter moves inward. Separate channels in the bowl and separate cover compartments segregate the discharges.

Solids in either phase are sedimented to the underside of the disks and slide outward along the surfaces because of their higher density. The aggregated solids must move from the outer edges of the disks to the bowl wall; some may be reentrained into new feed material, and carried into the disk stack, which accounts in part for actual performance falling short of theoretical prediction.

Commonly used with disk centrifuges are centripetal pumps (Fig. 13) that discharge the clarified liquid phases under pressures up to 0.7 MPa (100 psi) at reduced aeration, and scoop the rotating liquid out by using a stationary impeller. Interface location can be altered by varying the centripetal pump-back pressure, allowing interface control without shutdown to change a discharge weir. Centripetal pumps are capable of discharging at rates to 250 m^3/h (1100 gpm), and often eliminate the need for a tank and conventional pump (see PUMPS).

To maximize cake capacity, the simplest disk centrifuge bowl (Fig. 13**a**) is designed having a nonperforate bowl wall parallel to the axis. Feed solids should not exceed 0.5%. Bowl diameters of industrial units range from 180 to 600 mm with operating speeds from 8000 to 4500 rpm; the disks, between 30 and 200, are stacked at spacings of 0.3–6 mm; the half-angle is frequently 35–40° because solids handling is not critical. This type of centrifuge was originally employed for the separation of cream from milk and is still used widely in this field. Other uses include purification of fuel and lubricating oils having a low percen-

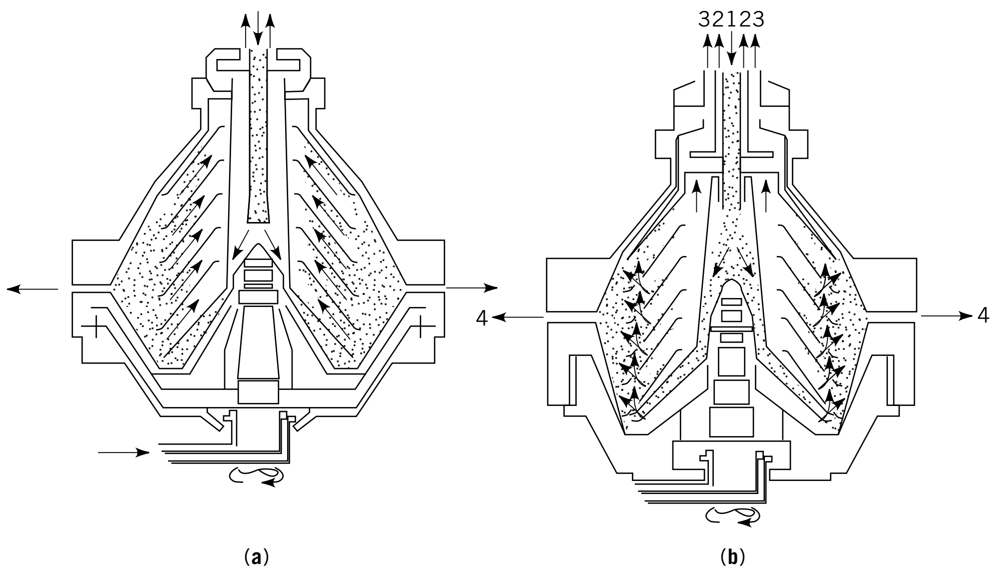


Fig. 13. (**a**) A clarifier and (**b**) a purifier of the paring disk-type design, where intermittent discharges of solids are designated by →, and 1 represents feed; 2, light phase; 3, heavy phase; and 4, solids.

tage of solids, separation of wash water from fats and vegetable or fish oils, and removal of moisture and solids from jet fuel. Solids that move readily in plastic flow can be continuously discharged as the heavy phase, eg, in the separation of soap stock from oil in vegetable oil refining.

Continuous discharge of solids, as a slurry, is achieved by sloping the inner walls of the bowl toward a peripheral zone containing between 8 and 24 discharge points containing nozzles, as shown in Figure 12**g–i**. The nozzles must be spaced closely enough so that the natural angle of repose of the solids deposited between the nozzles does not cause a buildup of cake to reach into the disk stack and interfere with clarification. The size of the nozzles is limited because the fluid pressure at the wall, which can be 6.9–13.8 MPa (1000–2000 psi), produces high nozzle velocities (see eq. 30). On the other hand, the nozzles must be large enough to prevent obstruction by individual particles; nozzle diameters at least four times the size of the largest particle are satisfactory. Replaceable, wear-resistant bushings in the nozzles provide orifices in the range of 0.8–2.5 mm. From 5 to 50% of the feed may be discharged with the solids through the nozzles. The upper limit of solids concentration depends on the particle packing characteristics but seldom exceeds 20 times that in the feed. The nozzle flow is directed backward with respect to rotation to recover much of the pressure energy and to reduce centrifuge power (see eq. 29). Bowl diameters range from 100 mm in laboratory units to 900 mm in industrial units and speeds from 12,000 to 3000 rpm, respectively; power requirements are between 10 and 2000 kW. Centrifugal accelerations to 12,000 G are obtained at the wall in the smaller bowls. Feed rates range from 1 to 136 m^3/h (4–600 gpm).

Applications include kaolin clay dewatering, separation of fish oils from press liquor, starch and gluten concentration, clarification of wet-process phosphoric acid, tar sands, and concentrations of yeast, bacteria, and fungi from growth media in protein synthesis (20).

A variant of the continuous-discharge disk centrifuge provides for introduction of a recycle stream. Restrictions on number and size of nozzles sometimes prevent adequate concentration of the discharged solids to satisfy further process requirements. To increase this concentration, a portion of the discharged slurry is returned to the feed, thereby increasing the overall loading of solids in the bowl. A more efficient method, shown in Figure 12**i**, is to return some discharged slurry through a recycle system to the region of the nozzles where the higher density recycle stream preferentially joins the nozzle flow.

A further modification of the disk centrifuge provides peripheral ports that are opened only intermittently to discharge sludge (see Fig. 12**d–f**, and 13). This type is employed for medium quantities of solids (1–4%) for which neither continuous discharge nor batch operation is suitable, and for solids that break down under the shear forces of nozzle discharge and are therefore not suitable for nozzle recycle. Intermittent discharge provides longer holding time and better concentration of solids, often at the expense of decreased disk size and reduced bowl throughput. The frequency and duration of opening can be controlled to discharge very high concentrations of sludge by partial emptying of the bowl.

These centrifuges are available in 180-mm bowl laboratory and 460–600-mm bowl industrial units. Disks are mostly spaced at 0.6–1.0 mm and the half-angle is 40–45°. In the industrial units, bowls with 60–200 or more disks operate at maximum speeds of 4400–6500 rpm and require 10–50 kW. Feed flow rates range from 2.5–114 m^3/h (10–500 gpm); temperatures from 30–90°C. Theoretical capacity factors are generally lower than in continuous discharge bowls of the same size. The disk outside diameter is smaller in order to provide solids-holding space of 4–20 L. Applications are limited to free-flowing solids that do not pack, and include recovery of wool grease from wool scouring liquor, orange juice clarification, recovery of soya protein, clarification of animal fats and food extracts, and purification of marine and jet engine fuels and lube oils as well as dairy applications.

Other modifications have special but more limited applications. A centrifugal bowl may contain, instead of disks, several annular baffles that take the liquid through a labyrinth path before discharge. The multiple cylinders increase cake capacity to as much as 70 L for easily sedimented solids. This centrifuge is used for clarification of food syrups and antibiotics (qv), and for recovery of heavy metallic salts and catalysts (see Fig. 12**c**).

Decanter Centrifuges. A comprehesive discussion of the decanter centrifuge is availible (27). Decanter centrifuges collect solids by sedimentation and continuously discharge both liquid and solid material. These centrifuges have bowl diameters of 150–1400 mm and are essentially tubular shells with a length/diameter ratio of 1.5–5.2, as shown in Figure 14. Deposited solids are moved by a helical screw conveyor operating at a differential speed of 0.5–100 rpm with respect to the bowl. Centrifugal fields are lower than in disk or tubular centrifuges because of the conveyor and its associated mechanism. Maximum speeds range from 300–9000 rpm. Figure 8 shows that particles of intermediate settling velocities, such as 1.5 to $\sim 15 \times 10^{-4}$ cm/s, are handled at medium to large flow rates. For clarification, this type of centrifuge recovers medium and coarse particles from feeds at high or low solids concentration. Particle sizes less than ~2 μm are normally not collected without the addition of flocculating agents. For classification of solids, the flow rates are higher than for clarification and the overflow usually contains most of the finer solids. Feed flow rates range from 1 to 136 m^3/h (4–600 gpm).

Incompressible solids discharged from the decanter, may not be as dry as those obtained by centrifugal filtration. Coarse crystals may discharge at 2–10% moisture, ground limestone at 15–20% moisture, and kaolin clay in the filler range (1–10 μm) at 30–35% moisture. It has been found that the decanter centrifuge is particularly well-suited for compacting low permeability compressible

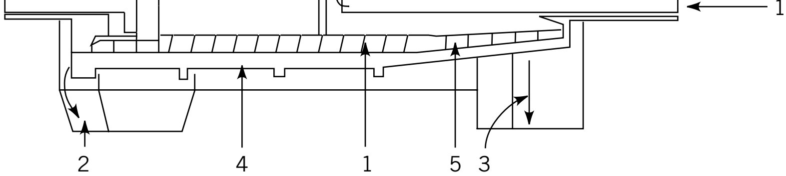


Fig. 14. A decanter centrifuge, where 1 corresponds to feed suspension; 2, to liquid phase; 3, to solid phase; 4, to liquid pool; and 5, to dry beach.

cakes, due to the beneficial motion and shearing of the compaction cake caused by the relative rotation of the conveyor.

Cake dryness's achieved with decanter centrifuges is much higher than what can be achieved with a bottle centrifuge operating at the same g-level, pond depth and with the same cake retention time (28). One manufacturer has recently launched a new series of municipal and industrial waste sludge decanters that optimises the ratio between sludge transportation and cake shearing thus further enhancing the cake dryness or feed rate of the centrifuges. Compressible, amorphous, and fibrous materials, such as sewage sludges, can be dewatered to 60–75% moisture, and meat rendering solids at 60–70% moisture plus 6–8% liquid fat. Operating pressures up to 1.03 MPa (150 psi) and temperatures up to 200°C are standard and 300°C is available.

Feed is introduced through an axial tube. Solids sedimenting to the bowl wall are conveyed along the cylindrical section and up a sloping beach. They are usually discharged at a radius smaller than that of the liquid discharge. Fine and flocculent solids compact under the liquid and show relatively little drainage on the beach. Coarse crystals and fibers do drain on the beach to a low residual moisture. The liquid level in the bowl is maintained by ports adjustable to the desired overflow radius. Considerable variation in design is available for the bowl shell, flight angle and pitch, beach angle and length, conveyor speed, feed position and type, and patterns of liquid and solids movement through the bowl. Bowl shapes having a high l/d exhibit high clarification capacity but may have wetter solids owing to the greater amount of fine particles recovered for discharge in the solids phase.

Bowl designs include countercurrent or cocurrent movement of the phases. In countercurrent flow, feed enters near the conical–cylindrical intersection; liquid flows toward the cylindrical end of the bowl to discharge over dams; while deposited solids are moved up the conical beach by the conveyor. In cocurrent flow, feed enters at the end away from the conical end. Liquid and settled solids move in the same axial direction toward the conical beach end. Axial conduits or a skimming device remove centrate from the pond surface, and solids move up the beach. The actual liquid flow between the helical flights is more tangential than axial so the cocurrent/countercurrent description is not particularly significant in regard to the actual movement of solids vs liquids within the bowl. In either type, solutions of flocculating agents may be introduced in the piping before entering the centrifuge, either in the feed tube, the feed acceleration zone, or in the pond after feed acceleration. Owing to the complexity of the chemistry of polyelectrolyte flocculation rate and efficiency, and floc damage owing to turbulence, the best solution for optimum polyelectrolyte use must be determined experimentally. A wash can be applied to the solids on the beach, but efficient rinsing depends on careful design to direct the rinse liquid to the solids. The wash is not collected separately but is discharged with the mother liquor.

A vertical decanter design using a bowl elastically suspended from a spindle with no bottom bearing is often used for high pressure and high temperature applications instead of horizontal axis decanters. Vertical units are easier to seal for pressure operation, and are well suited to accommodate bowl expansion at high temperatures. A vertical, elastically supported rotor as compared to a horizontal design does not offer separation advantages. There are mechanical advantages, however: the vertical design uses only one seal to three in the horizontal design. Moreover, the process connections are rigid (vertical) rather than flexible (horizontal), thermal expansion is not critical, and noise is lower.

Although Σ_T (eq. 12) indicates a reduced sedimentation performance level at increased liquid depth, this occurs only for coarse solids. The optimum pond depth varies with feed zone design, tendency of deposited solids to redisperse, conveyor differential speed, and particularly the depth of the cake layer required to produce a given solids concentration. Deep ponds are generally more effective for soft, slimy solids because conveying problems may reduce performance level and prevent complete clarification, even at low flow rates. The difficulty with which a decanter centrifuge is able to discharge soft, slippery solids has limited use in the past. To move the solids up the beach the solids must remain at bowl speed, with the helical flights moving the solids inward, against the high G. Often the soft solids slide back into the pond, building up in the bowl and ruining sedimentation performance. Setting the pond level inward of the solids discharge radius and separating the feed portion of the pond from the conical (beach) end using a deeply immersed baffle permits the centrifugal pressure head developed to assist the conveyor in discharging the solids. This method is used to thicken secondary sewage sludge from feed, ie, solids concentrations of 0.5–1% to solids concentrations of 5–10%. This thickening reduces total flow to the next stage of treatment by an order of magnitude. Flow rates up to 100 m^3/h (440 gpm) have been achieved without the use of polyelectrolytes. A small amount of polyelectrolytes, however, can double this rate.

Dewatering sewage sludge to high solids levels has been achieved by the use of higher polyelectrolyte dosage (5–15 kg/t), increased G level (2000–4000 G), and longer solids residence time in the bowl. Longer solids residence has been achieved by mechanically restricting the solids near or at the beach and reducing the differential speed to 0.5–4 rpm. The torque between the bowl and conveyor is a good indication of solids dryness, and use of this factor has permitted automatic differential speed control to maintain a constant solid dryness at varying feed concentrations and feed rates. Differential speed control can be obtained by driving the conveyor by use of a planetary gearbox mounted on the bowl, the gear box receiving controlled torque, and speed from a variable speed motor or electrical brake or motor. Occasionally a hydraulic motor is mounted on the bowl and an external hydraulic power supply is used to deliver variable rates of high pressure oil to the rotor by means of a rotary union. All methods are increasing being controlled by programmable microprocessors. Higher torque differential speed controllers have permitted centrifuge capacity and solids dryness to achieve levels of 100 m^3/h (440 gpm) at 35–40 wt% solids.

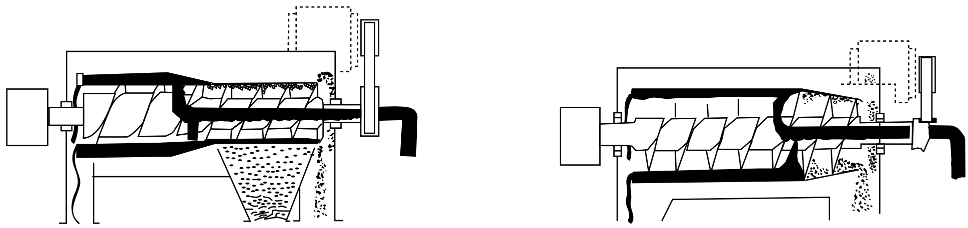
Final dewatering of sewage results from the solids almost filling the interior space between the conveyor hub and bowl shell interior. The use of polymers as flocculating agents results in easy separation in spite of a greatly reduced volume allowed to the centrate. The polymer also conditions the solids so that the pressure resulting from the deep layer of solids and the force needed to move the solids axially and inwardly toward discharge compresses and squeezes additional water from the space between and within the flocs. A drier discharged cake results. Compatibility studies (29) can be made to characterize particular solids. Improvements in dryness is proportional to $G^{1.5}$t, where G is the level at the mean solids radius and t is the solids residence time within the centrifuge. On a decanter having an $1/d = 4.2$, the solids have been found to occupy 75–80% of the bowl. The centrate clarification occurs just before discharging. Theoretical studies (30) comparing the ability to dewater compressible solids by sedimenting and filtering centrifuges to pressure filters, have shown that at high G levels, scroll decanters produce drier cakes than pressure filtration.

The capacity of decanters can be limited by any one of several factors (31): centrate clarity, usually a function of Σ; solids dryness, imposed by requirements of the next process step; conveyor torque, limited by rating of gear box, hydraulic motor, or back-drive capacity; chatter, by torsional instability resulting from stick-slip action of conveyor flights on fusible solids; swallowing capacity, by the ability of conveyor to accept feed without rejection; power, by the rating of drive motor; solids purity, where purity is obtained by rinsing; erosion rate, where abrasive particles are present; size purity, for particle classifications; solids volume, where solids feed concentration is high; fused solids, rate at which pasties are formed; and control, the maximum rate at which process control can be maintained.

Special designs have been offered which include one or more of the following features in vertical or horizontal construction: vapor-tight and pressure-tight enclosures; clean in place (CIP) and sterilize in place (SIP) rotors; three-phase separations having two immiscible liquids and solids; sanitary construction; a wide range of helices, pitches and leads, beach angles, compound beach angles, l/d ratios, pond depth/d ratios, and ribbed or grooved beaches or bowls; centrate centripetal pump devices; and abrasion protection systems, including complex flight tip geometries of ceramic, carbide, and conventional hard surfacing.

In clarifying operations, the conveyor discharge centrifuge recovers many types of crystals, meal from fish press liquor, and polymers, such as poly(vinyl chloride) and polyolefins. It is also used to dewater coal (32) and to concentrate solids from flue gas desulfurization sludges. Vertical designs, vapor-tight or under pressure, are applied to terephthalic acid, polypropylene, and catalyst recovery. Classification includes separation of particles ~2 μm from kaolin coating clay, and of particles over ~5 μm in the mill discharge of ground TiO_2; selective recovery of calcium carbonate from lime-treated waste sludges permits calcining and recycling of the lime without an overwhelming recycle load of inert material. The decanter centrifuge is frequently used to rough out medium and coarse solids before a second separation stage such as a disk centrifuge handling refinery sludges.

A varient of the decanter centrifuge, the screen bowl decanter uses both sedimentation and filtration primarily for the purposes of either drying or rinsing cake solids. The screen bowl and conventional decanter are shown in Figure 15-**a** and **b** respectively. Feed slurry is introduced conventionally into these imperforate scroll centrifuges. The dewatered solids are deposited onto the bowl wall and conveyed up a beach. The screen bowl decanter has a cylindrical screen with an inside diameter equal to the solids discharge diameter. The conveyor moves the solids axially over the bar screen area where further dewatering, fines loss and potentially rinsing takes place. Unlike the conventional decanter in which rinsing on the solid beach ultimately sends the rinse liquor to the centrate, the screen section allows rinse to be segragated from the



(a) (b)

Fig. 15. Comparison of screen bowl (**a**) and conventional (**b**) decanters. Screen section can be used for purposes that include removal of additional fines (thru the screen) from cake solids, futher drying of cake solids or introducing rinse liquor that can be isolated from the centrate.

centrate. Screen bowls are used to dewater coarser solids to a very high degree. Typical screen bowl decanter would be used to dewater fine coal, while the conventional decanter whould more likely be used to capture the solids from a coal refuse stream. Screen bowls are also used in production of purified parazylene (18).

5.2. Centrifugal Filtration Equipment. The important parameters of centrifugal filtration equipment (4,32) are screen area, level of centrifugal acceleration in the final drainage zone, and cake thickness. The latter affects both residence time and volumetric throughput rate. As indicated by equation 20, the particle size of the solids and the kinematic viscosity of the mother liquor also strongly affect the final moisture content. A limited correlation has been developed for the performance of perforate basket and conveyor discharge conical screen bowls, but the range of materials handled and the complexity of the drainage and washing operations do not lend themselves to broad correlations. The variables of correlation may be useful in a particular study, especially if more than one type of centrifuge is involved. An example of centrifugal filtration is the recovery of salt crystals from a mother liquor of $\sim 3 \times 10^{-6} m^2/s$ (3 cSt) viscosity; the crystal size is 170 μm at the 15% level and drained cake bulk density $\sim 1.5 \times 10^3$ kg/m^3 (95 lb/ft^3). The cone screen is 25.4 cm at the larger diameter and the automatic cycle basket centrifuge is 68.6 cm in diameter; other parameters for final values of $q = 7.6\%$ in both centrifuges are given in Table 1.

A conveniently expressed coordinate for plotting filtration performance is the drainage number, $d(G)^{1/2}/v$, where d is the mean particle diameter in micrometers, v is the kinematic viscosity of the mother liquor in m^2/s (Stokes $\times 10^{-4}$) at the drainage temperature, and G is $\omega^2 r/g$; r is the largest screen radius in a conical bowl. Because the final moisture content of a cake is closely related to the finest 10–15% fraction of the solids and is almost independent of the coarser material, it is suggested that d be used at the 15% cumulative weight level of the particle size distribution instead of the usual 50% point.

The other coordinate is $qt^{1/2}$, where q is the percentage of final moisture on the discharged cake, as the volume of mother liquor per unit volume of solids, and t is the drainage time in seconds. A weight ratio may be used for q, but the volume ratio makes the function more universal by eliminating densities.

Table 1. **Centrifuge Parameter Values for $q = 7.6\%$**

	Centrifuge	
Properties	Cone screen	Automatic basket
G	1440	865
bowl		
diameter, cm	25.4	68.6
speed, rpm	3180	1500
$\bar{d}(G)^{1/2}/\nu$	3.750×10^6	2.880×10^6
time, s		
feed and spin		12
spin after rinse		27
drain time	0.6[a]	39
rinse[b]		10
unload and screen rinse		4
cycle time		53
$qt^{1/2}$, $\% \cdot s^{1/2}$	5.89	47.46
h, cm	0.51	3.81
screen area, cm^2	930	7675
approximate cake rate, kg/s	1.19	0.84

[a]Differential speed = 60 rpm, 5/8 turn per helical flight.
[b]Rinse is shown only in the basket centrifuge because rinse efficiency in the cone screen is fairly low compared to the basket. The latter may be selected for the application if good rinsing is needed, and the conveyor-discharge conical screen centrifuge selected if no rinsing is necessary.

For a conveyor discharge conical screen bowl, the helical conveyor is assumed to control the residence time of the solids, so that time becomes the number of turns of helical flight around the conveyor hub divided by the differential speed between the conveyor and the bowl. For a pusher centrifuge, t is the retention time on the screen as controlled by length and diameter of screen, thickness of cake, and frequency and length of stroke of the cake. In a basket centrifuge, dead and unload times should not be included in the calculation of drain time. Bulk drainage is completed so quickly that film drainage is usually controlling. Thus, drain time t is approximated by the sum of feeding time, spin time prior to rinsing, and spin time after rinsing but prior to unloading according to the theory of cyclical centrifuges (34). Filtration correlations generally show a spread as a function of cake thickness. Conical screen bowls characteristically have short residence times and achieve good drainage by maintaining thin layers of cake. Smaller perforate baskets operate having cakes 5–10 cm in thickness, whereas larger baskets may carry cakes up to 15-cm thick. Pusher centrifuges and high speed peeler baskets may handle cakes ranging in thickness from 5 to 20 cm.

Perforate Basket Centrifuges. The simplest and most common form of centrifugal filter is a perforate-wall basket centrifuge, consisting of a cylindrical bowl having a diameter ranging from ∼100–2400 mm and a diameter/height ratio ranging from 1 to 3. The wall is perforated with a large number of holes, more than adequate for the drainage of most liquid loads, and is lined with a filter medium. In the simplest case, the medium is a single layer of fabric or metal cloth or screen. In high speed basket centrifuges, one or more backup screens of relatively large mesh support a finer mesh filter surface. The method of discharging accumulated solids distinguishes three types of basket centrifuge: those that are stopped for discharge, those that are decelerated to a very low speed for discharge, and those that discharge at full speed (35).

Basket centrifuges that must be stopped for discharge are available in many sizes. The bowls are usually supported on a vertical spindle. Designs vary from a 300-mm diameter basket of 30 L of cake capacity (0.4-kW motor, 2100 rpm, and ~800 G) to a 1500-mm diameter basket having 500 L of cake capacity (14-kW motor, 600 rpm, and 300 G). Basket cake volumes are always nominal and must be modified according to cake density. Construction materials include carbon or stainless steel, Monel, Inconel, titanium, and a variety of rubber and plastic coatings. Normal operation includes pressures up to 35 kPa (5 psi) and temperatures up to 180°C. This type of centrifuge is used if a variety of materials must be filtered in small batches, if equipment must be sterilized between batches, or if the production rate is too low to warrant more automation. These centrifuges are also used in removing liquid from crystalline materials, in drying bulk materials such as raw leafy vegetables, and in clarifying process liquids and waste streams. Cycle times are >10 min. Large (dia = 1300–1500 mm) baskets operating at 700–800 rpm may use an inner-perforated container that mounts in the bowl but is removable for bulk loading outside the centrifuge. Such units are well suited to laundry and dyeing purposes and for dewatering of textiles, yarn, raw stock, feathers, and hair. Particle sizes range from very fine to 500 μm. Feed flow rates vary from 1 to 20 m^3/h (4.5–90 gpm). A syphon to control the rate of filtrate removal improves dewatering (36).

Improved control systems and rising labor costs have led to manually or automatically controlled cycles with mechanical unloading at reduced speed. Baskets typically load at low to medium speed, accelerate to 900–1800 rpm for drainage and washing, decelerate to 35–75 rpm for mechanical unloading, and then start the cycle again. Cycle times range from 2 to 6 min. Cycles of 30 min for slow drainage or multiple rinses are common. Both a top-driven suspended bowl and underdriven bowl with three-point casing suspension designs are available. The cake is discharged in 20–120 s by means of a single or multiple plow that leaves a heel of cake on the filter medium. The heel can be completely removed with the help of a plastic-tipped plow and a perforated protecting plate. Filter media vary from perforated plates having 3-mm holes to 37-μm (400-mesh) Dutch twill. Vertical automatic dischare baskets always unload through a bottom discharge; a valve mechanism may be used to seal the bottom if the basket is fed with the whole charge at one time, so that an appreciable liquid layer develops.

The most fully automated basket is the horizontal basket or peeler centrifuge that feeds and discharges at spin speed, and normally operates at cycle times of <3 min at pressures up to 1.03 MPa (150 psi) and temperatures from −70 to 120°C, and in special cases to 350°C. It is primarily used for materials draining freely in a high centrifugal field, for medium tonnages, and for multiple rinses where nearly complete segregation of the rinses and motor liquor is desired. Ideally, the feed should have a constant composition and high concentration. For this purpose, a gravity slurry concentrating tank or a cyclone is often installed ahead of the centrifuge. The bowl rotates on a horizontal axis with a

Table 2. **Basket Centrifuge Operation on Free- and Slow-Draining Particles**

	Type of crystal	
Operation	Free draining[a]	Slow draining[b]
solids handling rate, t/h	20–24	1.5–2
conditioning rinse, s	1	0
feeding time, s	7	25
wash time, s	5	25
drain time, s	12	30
unloading time, s	2	6
total cycle, s	27	86

[a]For example, ammonium sulfate.
[b]Requiring several rinses, eg, polyolefins.

metal screen as the filter medium and discharges solids by cutting them out with a hydraulically operated knife. Bowls are 300–1200 mm in diameter and handle loads of 28–170 L (1–6 ft^3) of cake at bowl speeds of 1000–2500 rpm, producing maximum centrifugal fields of ~1250 G. The solids discharge through a front chute, which simplifies sealing against pressure. A distributor riding on the cake during feeding maintains uniform cake thickness, gives better bowl balance, and improves washing efficiency. Power requirements range from 15–150 kW. Materials of construction include carbon or stainless steel and Monel. The high speed during feeding and discharge may cause deformation of particles and breakage of crystals, whereas the heel of cake left on the screen may lose permeability through glazing or plugging with fractured fines. The heel can be conditioned by suitable washes after one or more cycles. Table 2 lists two process cycles.

This type of centrifuge is also used on borax and boric acid, *p*-xylene, sodium bicarbonate, and sodium chloride from glycerol or electrolytic caustic, in addition to dewatering of various products, eg, potato starch, or dewatering and washing slurries, eg, calcium hypochlorite.

Inverting Filter Centrifuge. Another batch automatic horizontal perforated bowl centrifuge inverts the flexible filter to discharge the solids. Feed slurry may be deposited on the inside surface of a cloth filter, with the bucket end completely closed. When the interior is full of dewatered solids, the bowl is decelerated to a slow speed and piston and closure plates move axially, inverting the filter cloth so that the solids reside on the outside diameter of the cloth. Very little residual material remains on the filter cloth surface for the next cycle. Centrifuges are available with diameters of 300–1300 mm, and operate at 720–1920 G. Filter cake rates of 75–250 kg/h are achieved at excellent rinsing efficiencies and low crystal breakage. Drives are electric or hydraulic, with the axial movement hydraulically actuated via a rotary seal. Power demands are low, from 3 kW for the smallest to 100 kW for the largest.

Continuous Cylindrical Screen Centrifuges. Continuous filtering centrifuges are used for very fast draining that do not require extremely dry final products. Rinsing efficiency varies considerably; power requirements are usually low; initial slurry concentrations can be somewhat more variable and not as high as for the high speed basket. Continuous centrifugal filters are equipped with either a cylindrical or a conical screen. Both types are made without a retaining lip on the solids discharge end of the bowl and employ various methods to move the solids through the bowl.

The cylindrical screen centrifuge deposits solids at one end of the bowl in a layer 6–80 mm thick and pushes the annular ring of cake axially through the bowl by means of a reciprocating piston (37,38). Washes are collected in separate sections of the casing but are not as distinctly separated as with basket centrifuge sequential operation. Drained solids at the end of the bowl are thrown into a casing that is separated by baffles from the liquid discharge zones. Bowls rotate on a horizontal axis, range in diameter from 200–1200 mm, and have capacities of 1–25 t of solids/h. Centrifugal fields of 300–600 G are common. To reduce the fines loss and to facilitate movement of the cake on the screen, maximum speeds are rarely used. Power requirements range from 4–60 kW. The cylindrical screens are generally of the bar screen type, with 0.1–0.5-mm spacing. The reciprocating piston operates at 20–100 strokes/min, with stroke lengths up to 80 mm. The thickness of the cake (max ~80 mm) depends on the packing and draining characteristics, the bulking tendency of the layer in front of the pusher, and the frictional resistance of the cake on the screen. Friction also depends on the construction of the bar screen.

The feed slurry is introduced by gravity flow of 20–200 t/h or screw conveyor to an imperforate distributing cone that deposits the slurry at its original concentration immediately in front of the pusher, as shown in Figure 17. The cake must not buckle at this point, so slurries of 40% concentration are generally necessary. Because fast drainage is required, feed particle size should exceed 150 μm; medium and coarse crystals and granules or fibrous solids can be handled in this type of centrifuge. Crystal breakage is low within the basket but some breakage does occur during discharge.

To handle materials that form a soft or plastic cake or have a high frictional resistance, the cylindrical screen may consist of two to six steps with successively larger diameters, as shown in Figure 16. Alternate steps reciprocate with the piston. Thus, the cake is pushed across only a short length of screen before redistribution on the next step at a slightly larger diameter. Drainage and washing efficiency are increased by redistribution of the cake under these conditions. This type of centrifuge is used on sugar (qv), where the high viscosity of the mother liquor causes slow drainage, a high degree of plasticity in the partially drained cake, and poor penetration of wash liquor.

Another type of continuous, cylindrical screen centrifuge discharges the cake by moving it axially through the bowl with a helical conveyor. Crystal breakage through conveyor action is greater than with the pusher-type mechanism. Applications include the handling of copper as trisodium phosphate and purification of paraxylene.

Conical Screen Centrifuge. In conical screen centrifuges the angle of the bowl causes or assists the cake to move axially and redistributes it in an increasingly thin layer which improves drainage characteristics. The feed slurry is deposited at the small end of the cone, where most drainage occurs. The drained solids are discharged from the large end, which has no retaining ring. Screens

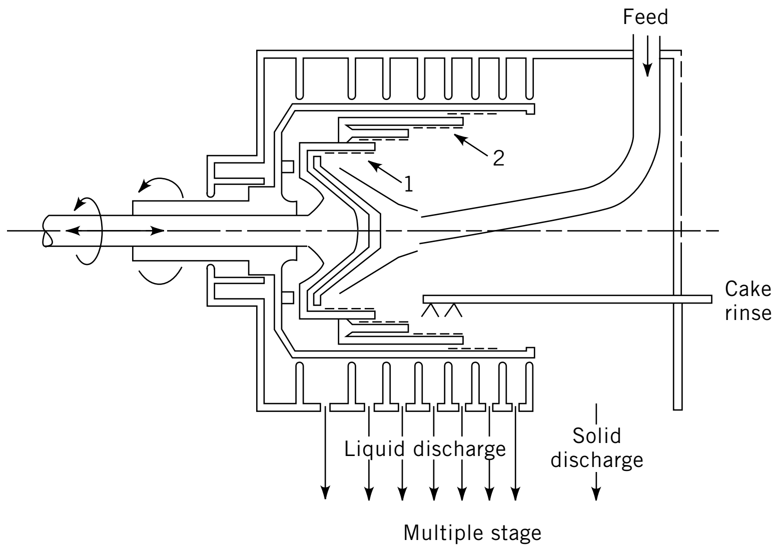


Fig. 16. Multiple-stage pusher centrifuge. Screens 1 and 2 reciprocate with pusher.

generally have 0.08–1.5-mm slots or perforated plate holes. Screens less than ~0.25-mm thick require a backup screen to extend screen life. There are three types of conical screen centrifuges: those that are self-discharging, those that discharge by means of a helical conveyor (Fig. 17), and those that apply an axial vibration or oscillation to the bowl or the bowl and casing.

In its simplest form, the cone angle is slightly larger than the angles of repose of the solids at any stage in their drying cycle. Some bowls are made with two angles, such that the more shallow is at the small end, and the steeper, where solids concentrate, is higher and at the large end. Horizontal or vertical bowls of 500–1000-mm large-end diameter operate at speeds up to 2600 rpm, producing up to 2500 G; cone angles vary from 20–35° and selection of the proper cone angle and suitable screen surface is critical for each application. Temperature control of viscous feeds is necessary to maintain proper distribution and drainage. Feed slurry, usually under gravity flow, is introduced at the small end of the basket, where it is accelerated and spread evenly over the periphery of the screen. Viscous sugar massecuites are successfully handled at capacities of 2–7 t/h on a 750-mm diameter basket. Loss of fines is greater than on the automatic basket centrifuges, but improved rinse performance increases sugar purity. Rinsing efficiency is not generally high on this type of cone and segregation of rinse liquor is incomplete. This centrifuge is also used for the drying of crystalline materials such as ammonium sulfate and separation or dewatering of fine fiber in wet corn milling.

Conical screens with a helical conveyor turning slightly faster than the bowl can handle a variety of materials. The vertical baskets are underdriven and the larger diameter is downward, as shown in Figure 17. Horizontal axis designs are also available. Feed is introduced at the small end and the

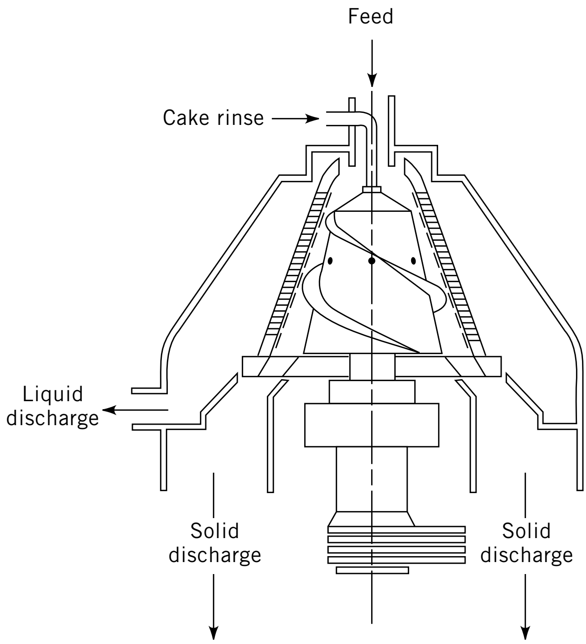


Fig. 17. Conical screen conveyor discharge centrifuge.

rate at which solids move through the drainage zone is controlled to some degree by the differential speed of the conveyor. Bowls have large-end diameters of 250–500 mm, and lengths about two-thirds of the large diameter; bowl angles range from 10 to 20°, and operating speeds are 2500–3800 rpm, giving up to 3500 G. Solids capacities range from 1 to 30 t/h, with feed capacities of 1–15 m^3/h (260–4000 gph). Centrifuge casings may be vapor-tight but are not intended for pressure operation; maximum temperature is about 150°C. Power requirements, low for the tonnages handled, range from 7 to 30 kW. Applications include dewatering of medium and coarse crystals, deoiling of proteinaceous solids, and removal of solids from fruit and vegetable pulps and other food slurries.

The third type of cone screen centrifuge operates with bowl angles of 13–18° and assists solids discharge by a vibratory motion of the bowl or bowl and casing. These units usually have underdriven bowls having 500–1100-mm larger diameter at the upper end; horizontal axis designs are also available where diameter/length ratios range from 1 to 2. Operating speeds are normally 300–500 rpm, and solids capacities range from 25 to 150 t/h. Pressurized units are not available, and operating temperatures range to 100°C. Power requirements are 15–35 kW. Bar screens are frequently used, and applications are largely on coal dewatering; particle sizes from 30 mm down to 0.25 mm (60 mesh) are easily handled. These centrifuges are also used in dewatering of potash and other crystalline solids. Where SW = separative work in kg/year.

6. Nomenclature

Symbol	Definition
$Ae_{3/4}$	the cylindrical area at 3/4 of the bowl radius
a	distance; space between adjacent disks
C	discharge coefficient
c	concentration
d	diameter of particle, particle size
$\bar{d}$	mean particle diameter
e	efficiency factor
E	energy
E_{rel}	relative performance
G	ratio of centrifugal acceleration to the acceleration of earth's gravity
g	acceleration of gravity
h	cake thickness, height
K	permeability
k	experimental coefficient
l	length, particularly from feed zone to centrate discharge
m	torque
N	bowl speed
n	number of disks
P	power
p	pressure
Q	Volumetric flow rate
q	volume of undrained liquid(mother liquir)/unit volume of solids, %
r	radius
S	Fraction of void volune occupied by liquid
s	external surface area/weight of solid
s'	surface area/volume of cake
T	torque
t	time
V	volume
v	velocity
α	cake resistance constance in pressure filtration, length/volume
Δ	difference, particularly density
δ	mass density, mass/unit volume
ε	voil fraction
μ	viscosity
ω	angular velosity
Ψ	wetting angle
Φ	angle between direction of nozzle and tangent to circle intersecting nozzle axis at discharge
Σ	theoretical capacity factor, length^2
σ	surface tension or material stress
θ	half angle of disks
ν	kinematic viscosity
Subscripts	
B	bottle centrifuge
BD	backdrive
C	cake or conveyor torque
F	friction factor
f	film flow
g	settling velocity of particle in gravity field
h	heavy phase

Symbol	Definition
i	interface
L	liquid medium
l	free surface of liquid or light phase
M	filter medium, or mass
m	mean value
N	nozzle
P	process power
p	process
S	solid medium
s	settling velocity of particle in centrifugal field or shape constant
W	windage pored

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