SEDIMENTATION

1. Introduction

Sedimentation has been defined as "the separation of a suspension into a supernatant clear fluid and a rather dense slurry containing a higher concentration of solid" (1). This definition is too broad. It does not specify the acceleration, gravitational, centrifugal, magnetic, or electrostatic external field that causes the separation. There is also a possible ambiguity whether the suspension is gaseous or liquid. Herein sedimentation is restricted to the most common definition, ie, the gravitational settling of particles in liquids (see also MINERAL RECOVERY AND PROCESSING; SEPARATION, CENTRIFUGAL SEPARATION; SEPARATION, MAGNETIC SEPARATION). Most of the equipment and methods described are for separation of solid particles but where the same or similar technology is used for separation of immiscible liquids, this is also mentioned.

As buoyancy occurs in sedimentation, the deciding factor for particle settling to take place is the difference between the density of the particles and that of the suspending liquid. The existence of this density difference is therefore a necessary prerequisite.

The uses of sedimentation in industry fall into the following categories: solid–liquid separation (2); solid–solid separation; liquid–liquid (immiscible) separation; particle-size measurement by sedimentation (3); and other operations, such as mass transfer and washing.

In solid–liquid separation, the solids are removed from the liquid either because the solids or the liquid are valuable or because these have to be separated before disposal. If the primary purpose is to produce the solids in a highly concentrated slurry, the process is called thickening. If the purpose is to clarify the liquid, the process is called clarification. Usually, the feed concentration to a thickener is higher than that to a clarifier. Some types of equipment, if correctly designed and operated, can accomplish both clarification and thickening in one stage (see also Extraction, Liquid–Solid). In an overwhelming majority of applications in industry the solids to be separated are denser than the suspending liquid. However, where the solids are lighter and, therefore, settle upword (or float), this has some obvious design consequences.

In solid–solid separation, the solids are separated into fractions according to size, density, shape, or other particle property (see SIZE REDUCTION). Sedimentation is also used for size separation, ie, classification of solids (see SEPARATION, SIZE SEPARATION). One of the simplest ways to remove the coarse or dense solids from a feed suspension is by sedimentation. Successive decantation in a batch system produces closely controlled size fractions of the product. Generally, however, particle classification by sedimentation does not give sharp separation (see SIZE MEASUREMENT OF PARTICLES).

Where emulsions are to be resolved or other mixtures of immiscible liquids of different density are to be separated by sedimentation, the principles are the same as for solids, but some specific challenges present themselves. For example, the dispersed phase must not be emulsified by violent pumping or flow shear because this obviously reduces the particle size and makes the subsequent separation more difficult. Another specific feature of such systems is the possibility of coalescence. Where this is possible it is actively promoted because it can greatly improve the efficiency of the separation by effectively locking out of the continuous phase from the coalescing dispersed phase. If coalescence occurs during sedimentation this leads to increases in particle size and the consequent increase in the settling rate. Applications of liquid–liquid separation by sedimentation can be found in many chemical, pharmaceutical, nuclear, petrochemical, and petroleum industries. An excellent review of such applications, and of electrostatic coalescence, can be found in (4).

In particle-size measurement (3), gravity sedimentation at low solids concentrations ($<0.5\%$ by vol) is used to determine particle-size distributions of equivalent Stokes' diameters in the range from 2 to 80 μm. Particle size is deduced from the height and time of fall using Stokes' law, whereas the corresponding fractions are measured gravimetrically, by light, or by X-rays. Some commercial instruments measure particles coarser than 80 μm by sedimentation when Stokes' law cannot be applied.

Sedimentation is also used for other purposes. For example, relative motion of particles and liquid increases the mass-transfer coefficient. This motion is particularly useful in solvent extraction in immiscible liquid–liquid systems (see Extraction, Liquid–Liquid). An important commercial use of sedimentation is in continuous countercurrent washing, where a series of continuous thickeners is used in a countercurrent mode in conjunction with reslurrying to remove mother liquor or to wash soluble substances from the solids. Most applications of sedimentation are, however, in straight solid–liquid separation.

2. Principles

2.1. Gravity Settling of a Single Particle. If a particle moves relative to the fluid in which it is suspended, the force opposing the motion is known as the drag force. Knowledge of the magnitude of this force is essential if the particle motion is to be studied. Conventionally, the drag force F_{D} is expressed according to Newton:

$$F_{\mathrm{D}} = C_{\mathrm{D}} \cdot A \cdot \frac{\rho v^2}{2} \tag{1}$$

where v is the particle–fluid relative velocity, ρ is the fluid density, A is the area of the particle projected in the direction of the motion, and C_{D} is a coefficient of proportionality known as the drag coefficient. Newton assumed that the drag force results from the inertia of the fluid and that C_{D} is then constant.

Dimensional analysis (qv) shows that C_{D} is generally a function of the particle Reynolds number (Re):

$$Re_{\mathrm{p}} = \frac{v \cdot x \cdot \rho}{\mu} \tag{2}$$

where x is the particle size and μ is the liquid viscosity. The form of the function depends on the regime of the flow. This relationship for rigid spherical particles is shown in Figure 1. At low Re, under laminar flow conditions when viscous forces prevail, C_D can be determined theoretically from Navier–Stokes equations; the solution is known as Stokes' law:

$$F_D = 3\pi\mu v x \tag{3}$$

This is an approximation that gives the best results for $Re_p \rightarrow 0$. The upper limit of its validity depends on the error that can be accepted. The usually quoted limit for the Stokes' region of $Re_p = 0.2$ is based on the error of $\sim 2\%$.

Elimination of F_D between equations 1,2, and 3 gives another form of Stokes' law, as shown in Figure 1 as a straight line.

$$C_D = \frac{24}{Re_p} \qquad Re_p < 0.2 \tag{4}$$

For $Re > 1000$, the flow is fully turbulent. Inertial forces prevail and C_D becomes constant and equal to 0.44, the Newton region. The region in between $Re_p = 0.2$ and 1000 is known as the transition region and C_D is either described in a graph or by one or more empirical equations.

In solid–fluid separation, the fine particles are most difficult to separate, ie, Re_p is low, almost inevitably < 0.2, owing to low values of x and v. Therefore, only the Stokes' region has to be considered.

A single particle settling in a gravity field is subjected primarily to drag force, F_D; gravity force, $m \cdot g$; and buoyancy, $(m \cdot g)\rho/\rho_s$; which have to be in equilibrium with the inertial force:

$$m\frac{dv}{dt} = (m \cdot g) - (m \cdot g)\frac{\rho}{\rho_s} - F_D \tag{5}$$

assuming positive downward forces, where m is particle mass, g is gravity acceleration, ρ_s is particle density, and t is time. Equation 5 can be solved, assuming Stokes' law:

$$v(t) = \frac{gx^2(\rho_s - \rho)}{18\mu}\left[1 - \exp\left(-\frac{t18\mu}{x^2\rho_s}\right)\right] \tag{6}$$

This relationship is exponential with respect to time t and with increasing time quickly approaches equation 7, where v_g is known as the terminal settling velocity under gravity.

$$v_g = \frac{gx^2(\rho_s - \rho)}{18\mu} \tag{7}$$

The terminal velocity in the case of fine particles is approached so quickly that in practical engineering calculations the settling is taken as a constant velocity motion and the acceleration period is neglected. Equation 7 can also be applied to nonspherical particles if the particle size x is the equivalent Stokes' diameter as determined by sedimentation or elutriation methods of particle-size measurement.

Note that equation 7 essentially applies to settling at low particle concentrations ($< 0.5\%$ by volume) in Newtonian liquids, which have a constant viscosity. In principle, it can be used also in non-Newtonian fluids, where viscosity μ then becomes the apparent viscosity but, depending on the type of non-Newtonian behavior, its determination may require an iterative procedure. Not only is such behavior shear dependent (ie, the apparent viscosity depends on how fast the particle is settling), but it may also be time dependent and the model may contain a zero shear viscosity as a parameter. Reference 2, section 18.4 reviews the state of the art to 2000 with a specific reference to research that is still in progress, eg, on particle settling in the Carreau model fluids (eg, polymeric liquids).

2.2. Settling of Suspensions. As the concentration of the suspension increases, particles get closer together and interfere with each other. If the particles are not distributed uniformly, the overall effect is a net increase in settling velocity because the return flow caused by volume displacement predominates in particle-sparse regions. This is the well-known effect of cluster formation that is significant only in nearly monosized suspensions. For most practical widely dispersed suspensions, clusters do not survive long enough to affect the settling behavior and, as the return flow is more uniformly distributed, the settling rate steadily declines with increasing concentration. This phenomenon is referred to as hindered settling and can be theoretically approached from three premises (5): as a Stokes' law correction by introduction of a multiplying factor; by adopting effective fluid properties for the suspension different from those of the pure fluid; and by determination of bed expansion using a modified Carman-Kozeny equation. These three approaches yield essentially identical results:

$$\frac{v_{\mathrm{p}}}{v_{\mathrm{g}}} = \epsilon^2 f(\epsilon) \tag{8}$$

where v_{p} is the hindered settling velocity of a particle, v_{g} is the terminal settling velocity of a single particle as calculated from Stokes' law (eq. 7), ϵ is volume fraction of the fluid (voidage), and $f(\epsilon)$ is a voidage function, which for Newtonian fluids has different forms, depending on the theoretical approach adopted. The differences between the available expressions for $f(\epsilon)$ are not great and are frequently within experimental accuracy. The most important forms are as follows. From the Carman-Kozeny equation (6),

$$f(\epsilon) = \frac{\epsilon}{10(1-\epsilon)} \tag{9}$$

from Brinkman's theory (7–10), applied to Einstein's viscosity equation (11,12),

$$f(\epsilon) = \epsilon^{2.5} \tag{10}$$

and from the well-known Richardson and Zaki equation (13),

$$f(\epsilon) = \epsilon^{2.65} \tag{11}$$

For irregular or nonrigid particles, eg, flocs, the Einstein constant (2.5) and the Richardson and Zaki exponent (2.65) can be considerably larger than for spheres.

Strictly speaking, these correlations apply only to the cases where flocculation is absent, such as for coarse mineral suspensions. Suspensions of fine particles, because of the very high specific surface of the particles, often flocculate, and therefore show different behavior. With increasing concentration, C, of such suspensions, at a particular concentration, C_1, an interface is observed that becomes sharper at $C > C_1$. The slurry is then said to be in the zone-settling region. The particles below the interface, if the size range is not $> 6{:}1$, settle en masse, ie, all at the same velocity irrespective of size. There are two possible reasons for this: either the flocs become similar in size and settle at the same velocity, or they are joined and fall as a web. Interestingly enough, the settling rates of the interface and of the solids below it, of many practical suspensions, can still be described by the Richardson and Zaki equation (eq. 11), but the value of v_g has to be determined by extrapolation of the experimental log–linear plot for $\epsilon = 1$. The value of this intercept has in fact been used for indirect size measurement of the flocs. The slope of the plot determines the exponent.

The concentration C_1 at which zone settling is first observed depends on the material and its state of flocculation and no guidance can be given. Addition of flocculating agents (qv) or dispersants (qv) drastically changes this concentration and only experimental evaluation can yield its value. If the concentration is increased still higher, a point is reached when the flocs become significantly supported mechanically from underneath, as well as hydraulically, and the suspension is then known to be in compression or compression settling. The solids in compression continue to consolidate. The consolidation rate depends not only on the concentration, but also on the structure of the solids, which in turn depends on the pressure and flow conditions. This is a complex problem closely related to cake filtration and expression (see FILTRATION). At intermediate concentrations between those of zone settling and fully established uniform compression, channeling is sometimes observed, which particularly occurs in slowly raked large-scale thickeners. Under these conditions, a coarser structure of pores becomes interconnected in the form of channels.

Most authors who have studied the consolidation process of solids in compression use the basic model of a porous medium having point contacts that yield a general equation of the mass-and-momentum balances. This must be supplemented by a model describing filtration and deformation properties. Probably the best model to date ($\sim$1996) uses two parameters to define characteristic behavior of suspensions (14). This model can be potentially applied to sedimentation, thickening, cake filtration, and expression.

2.3. Coagulation and Flocculation. Both coagulation and flocculation are classical pretreatment methods used to increase the effective particle size, thereby improving sedimentation settling rates. Although these two terms are often used interchangeably, coagulation is sometimes defined as agglomeration of the primary particles into particles up to 1 mm in diameter. Flocculation, on the other hand, not only agglomerates particles, but also interconnects them by means of long-chain molecules of the flocculating agent into giant loose flocs up to 1 cm in size (see FLOCCULATING AGENTS). The term flocculation is used here to include coagulation as defined. Chemical agents create favorable conditions for flocculation by neutralization of surface charges and thus reduce interparticle repulsion. Mineral coagulants are in the form of electrolytes, such as alum or lime, whereas flocculation agents are mostly synthetic polyelectrolytes of high molecular weight. Development of the latter group since the 1970s has resulted in a remarkable improvement in sedimentation equipment. Such agents are relatively expensive and the correct dosage has to be carefully optimized. Overdosage is not only uneconomical, but may inhibit the flocculation process and cause operating problems. Because surface charges are also affected by pH, control of this variable is essential in pretreatment.

Although reduction or elimination of the repulsion barrier is a necessary prerequisite of successful flocculation, the actual flocculation in such a destabilized suspension is effected by particle–particle collisions. Depending on the mechanism that induces the collisions, the flocculation process may be either perikinetic or orthokinetic.

Perikinetic flocculation is the first stage of flocculation, induced by the Brownian motion. It is a second-order process that quickly diminishes with time, and therefore is largely completed in a few seconds. The higher the initial concentration of the solids, the faster is the flocculation.

The well-known DLVO theory of colloid stability (15) attributes the state of flocculation to the balance between the van der Waals attractive forces and the repulsive electric double-layer forces at the liquid–solid interface. The potential at the double layer, called the zeta potential, is measured indirectly by electrophoretic mobility or streaming potential. The bridging flocculation by which polymer molecules are adsorbed on more than one particle results from charge effects, van der Waals forces, or hydrogen bonding (see COLLOIDS).

The flocculation rate is determined from the Smoluchowski rate law, which states that the rate is proportional to the square of the particle concentration by number; inversely proportional to the fluid viscosity, and independent of particle size.

Orthokinetic flocculation is induced by the motion of the liquid obtained, eg, by paddle stirring or any other means that produces shear within the suspension. Orthokinetic flocculation leads to exponential growth, which is a function of shear rate and particle concentration. Large-scale one-pass clarifiers used in water installations employ orthokinetic flocculators before introducing the suspension into the settling tank (see WATER, MUNICIPAL WATER TREATMENT). Scale-up of orthokinetic flocculators, generally in the form of paddle devices, are based on the product of mean velocity gradient and time, for a constant volume concentration of the flocculating particles.

Another type of flocculation results from particle–particle collisions caused by differential settlement. This effect is quite pronounced in full-size plants where large rapidly falling particles capture small particles that settle more slowly.

A third type of flocculation, mechanical syneresis, has been defined (16,17). This is the process of shrinkage and densification of loose and bulky flocs through uneven application of local fluctuating mechanical forces leading to exudation of the liquid from the floc. It is achieved by slowly stirring the blanket zones with rotating paddles in sludge-blanket clarifiers. Pellet-like flocs can be produced by this process allowing higher overflow rates than those obtained using conventional blanket clarifiers. So far, the process has been applied successfully to only a few suspensions in water treatment.

3. Settling Tests

For the simplest case of particulate clarification, where no flocculation takes place during settling, ie, either the flocculation process is completed before entering the settling tank or the suspension is entirely nonflocculant, the basic test is the so-called short-tube procedure (18). A sample settles in and is decanted from a large measuring cylinder in order to evaluate the settling rate, ie, the specific overflow rate that produces a satisfactory clarity of the overflow. The long-tube procedure is designed for systems where flocculation (or deflocculation) takes place during settling, thus the settling tank's performance depends not only on the specific overflow rate, but also on the residence time in the tank. Tests are conducted in a vertical tube that is as long as the expected depth of the clarifier, under the ideal assumption that a vertical element of a suspension that has been clarified maintains its shape as it moves across the tank.

When the overflow clarity is independent of overflow rate and depends only on detention time, as in the case for high solids removal from a flocculating suspension, the required time is determined by simple laboratory testing of residual solid concentrations in the supernatant versus detention time under the conditions of mild shear. This determination is sometimes called the second-order test procedure because the flocculation process follows a second-order reaction rate.

The design of the sludge-blanket clarifiers used primarily in the water industry is based on the jar test and a simple measurement of the blanket expansion and settling rate (18). Different versions of the jar test exist, but essentially it consists of a bank of stirred beakers used as a series flocculator to optimize the flocculant addition that produces the maximum floc-settling rate. Visual floc-size evaluation is usually included.

The critical settling flux essential for evaluating the requirement of the zone-settling layer area of a gravity thickener is measured either using the Coe and Clevenger method (19) or the simpler Talmage and Fitch procedure (18). The former consists of a series of settling tests in a measuring cylinder where the initial settling rates of a visible interface within the settling suspension are measured for different initial solids concentrations. The Talmage and Fitch procedure is simpler because it requires only one test at any concentration, providing it is in the zone-settling regime. Theoretically, the two methods should give an identical critical settling flux, and therefore identical pool areas, but this is not so in practice. Usually, the Coe and Clevenger method leads to underdesign of the thickener area, whereas the Talmage and Fitch procedure leads to overdesign. For highly flocculant slurries, the area requirement of the compression layer may exceed that of the zone-settling layer; the compression zone also has a depth requirement. A laboratory test for the latter, employing a multiple batch upflow test for compression-zone evaluation, has been described (14). This test is by no means generally accepted and its reliability remains to be demonstrated.

4. Design Methods

The simplest case of a gravity settling tank without coagulation or flocculation in clarification applications, ie, when removing small amounts of solids, is based on the identical principle as laminar settling chambers for cleaning gases (20). The grade-efficiency curve $G(x)$ is given by equation 12:

$$G(x) = \frac{v_g A}{Q} \tag{12}$$

where v_g is the terminal settling velocity of a particle of size x (see eq. 7), Q is the feed flow rate that is equal to the overflow rate, and A is the plan area of the tank. Equation 12 was derived assuming laminar flow in the tank and no end effects. The tank area is the only design parameter affecting the theoretical separational performance irrespective of the shape or depth of the pool. Equation 12 can be expressed in terms of the more conventional dimensionless groups:

$$Stk.\ Fr = G(x) \tag{13}$$

if Stokes' law is assumed for the particle-settling velocity, and therefore the Stokes' number, Stk, is defined as follows:

$$Stk = \frac{x^2 \Delta\rho}{18\mu} \cdot \frac{Q}{A \cdot H} \tag{14}$$

and the Froude number, Fr, as in equation 15:

$$Fr = \frac{HgA^2}{Q^2} \tag{15}$$

where H is a characteristic dimension of the pool, eg, the height, which in equation 13 cancels; $\Delta\rho$ is the particle–fluid density difference; μ is the liquid viscosity; and g is the acceleration of gravity.

Scale-up can be calculated with the help of equation 12, based on a simple specific overflow rate model: for the same performance, the flow rate is proportional to the area, or vice-versa. Because of the long residence time involved during which the feed solids must remain constant, it is difficult to measure the grade-efficiency curve of a settling tank, particularly on a large scale. It is therefore more practical to measure the specific overflow rate (or overflow volume flux) Q/A that gives satisfactory overflow clarity from simple settling tests (18). If the clarification is not completely particulate, ie, when flocculation takes place and has not been completed before the suspension enters the settling tank, the overflow clarity depends not only on the overflow rate, but also on the detention time. The scale-up under such conditions is based on the long-tube procedure. Sometimes the effect of the detention time is so strong that the overflow rate can be ignored and the scale-up is based on the detention procedure. For the whole range of coagulation clarifiers used predominantly in the water industry, the scale-up is usually based on the overflow rate determined from jar tests primarily designed to select the best flocculant and determine the settling rates that give a clear supernatant.

Probably more relevant to the chemical industry is the scale-up of thickeners. Thickeners are basically gravity settling tanks that, apart from producing a clear overflow, are designed to have a thick underflow with as little water content as possible. The feed into a thickener is generally more concentrated than the feed into a clarifier, and quite often exhibits zone-settling behavior because of the application of flocculants.

An operating thickener has basically three layers: the topmost clarification layer, the zone-settling layer, and the compression layer at the bottom. Each of the three layers requires a certain area. Ideally, the largest of these governs the design of the thickener. In most cases, it is the function of the clarification layer to prevent those particles that have escaped from the zone-settling layer or the feed from leaving with the overflow. This function is frequently less important than the thickening function and thus the thickener area is usually chosen on the basis of the zone-settling or compression layer requirement.

The conventional design and scale-up of thickeners operated with mineral or certain metallurgical slurries is based on the area requirement of the zone-settling layer and assumes that the compression zone only imposes a solids-detention (hence depth) demand and has no independent demand on area, with the exception of the empirical 3-ft (1 m) rule, which if applicable, introduces an area demand (18). It is the basic principle of this method that the solids on their way downward from the feed layer to the underflow continuously increase in concentration from that of the feed to that of the underflow, usually determined by a time-retention test. The total solids flux (mass flow rate of solids per unit area), which different layers in the thickener are capable of accommodating, usually goes through a minimum between the feed zone and the underflow. This critical solids flux, G_c, determines the minimum design area of the thickener. The zone-settling layer does not form under these conditions and makes no depth demand. The thickener is then in no danger of overflowing solids. The area of the thickener, A, is calculated from the critical solids flux, G_c; the feed flow rate, Q; and concentration, C_f, using a simple mass balance, which assumes a complete separation of all feed solids.

$$G_c A = Q \cdot C_f \tag{16}$$

Thus the design and scale-up of the thickeners in this category centers around the determination of the critical solids flux, G_c. This value cannot be estimated from the primary properties of the particulate system because of the unpredictable effect of flocculation. It must be obtained experimentally from large- or pilot-scale thickeners. However, with a few exceptions, this method is impracticable because of the scale and cost of such experiments. If the settling velocity of the solids is assumed to be a function of concentration only, ie, $v = v(c)$, then this function should be unique for a given suspension and should be the same in batch-settling and continuous operations. This is the basis of the conventional Coe and Clevenger, and Talmage and Fitch methods that only differ in the way in which G_c is determined from experimental tests. The latter method requires simpler tests. According to Coe and Clevenger, if the function of $v = v(c)$ is known, the critical flux G_c corresponds to the minimum on the total flux G curve, which is as follows:

$$G = v \Big/ \left(\frac{1}{C} - \frac{1}{C_v}\right) \tag{17}$$

where the underflow concentration, C_v, is determined by time-detention tests. The thickness of the compression zone is also determined by time-detention tests, assuming that the solids concentration reached in the compression layer in a batch test after a given time is the same as in the compression layer of a thickener.

5. Equipment

Sedimentation equipment can be divided into batch-operated settling tanks and continuously operated thickeners or clarifiers. The operation of the former is simple. Whereas use has diminished, the batch settling tanks are employed when small quantities of liquids are to be treated, eg, in the cleaning and reclamation of lubricating oil (see RECYCLING, OIL). Furthermore, batch settling as repeated decantation is an effective and well-predictable particle classification process for small quantities of products. Most sedimentation processes, however, are operated in continuous units as reviewed in the following.

5.1. Clarifiers. The largest user of clarifiers is probably the water-treatment industry. The conventional one-pass clarifier uses horizontal flow in circular or rectangular vessels (Fig. 2) with feed at one end and overflow at the other. The feed is preflocculated in an orthokinetic (paddle) flocculator which often forms an integral part of the clarifier. Settled solids are pushed to a discharge trench by paddles or blades on a chain mechanism or suspended from a traveling bridge (see Water, Industrial Water Treatment). Rectangular settling tanks are also used in liquid–liquid (such as oil-water) separation in the petroleum industry where they are known as the API (American Petroleum Institute) separators. The design calculations for these are set out in API 734–53.

Where circular basin clarifiers are used, these are most commonly fed through a centrally located feed well. The overflow is led into a trough around the periphery of the basin. The bottom gently slopes to the center and the settled solids are pushed down the slope by a number of motor-driven scraper blades that revolve slowly around a vertical center shaft. This design closely resembles a conventional thickener. Like thickeners, circular clarifiers can be stacked in multitray arrangements to save space. Some juice clarifiers are also arranged in this way.

Circular raking mechanisms are sometimes also used in square basins with horizontal flow across the basin. Such designs have to incorporate supplementary rake arms that reach into the corners of the square vessel (Fig. 3).

The conventional one-pass clarifier is designed for the lowest specific overflow rate (flow per unit area of liquid surface), which is usually 1–3 m/h depending on the degree of flocculation. These clarifiers can be started and stopped without difficulty.

Newer designs incorporate some vertical flow and combine flocculation, gravity and inertial clarification, and solids recirculation. Because such units achieve higher overflow rates, they are referred to as rapid settling or high rate clarifiers. Figure 4 gives an example in this category.

Flocculation is accelerated and higher overflow rates are achieved by external or internal recirculation of settled solids into the feed that leads to the collection of fine particles by interception. Addition of conditioned fine sand to the feed induces separation by differential sedimentation, and sometimes increases overflow rates to 6–8 m/h.

Design and operation of recirculation systems can be complicated. Problems are avoided by using a sludge-blanket clarifier, in which feed enters below a blanket of accumulated and flocculated solids that become fluidized in the zone-settling regime by the upflowing feed. Feed solids are trapped in the blanket. The solids content of the blanket continuously increases and part must be bled off in order to maintain the mass balance.

Sludge-blanket clarifiers are available in flat-, trough-, and hopper-bottom types. The hopper-bottom vertical-flow clarifier shown in Figure 5 achieves rise rates of 1–6 m/h in wastewater applications. It is a 3–4 m deep, 60° triangular or circular hopper tank, with feed introduced through a downward-pointing inlet at the bottom of the tank. The flocculated feed suspension is clarified by passage through the blanket and overflows into decanting troughs that are usually 1–1.5 m above the blanket, to allow for blanket-level variations with feed flow rate. The blanket can be continuously bled off through a submerged weir-type regulator and then thickened in a conical concentrator, or the clarifier can be periodically shut down to allow settlement bleeding.

Sludge-blanket clarifiers are difficult to start up because the first blanket must be established, and large-scale units require extensive excavation. Sizes range from 600×600 mm to 50×50 m. Precipitation and crystallization can be carried out in similar hopper-designed units, having overflow rates of 80 m/h or higher.

A combination of gravity settling and slow cyclonic action, first patented in 1983 (21), is available commercially. Known as the hydrodynamic or vortex-induced separator, it is used for excess storm water treatment or for the separation of grits and sands from raw sewage and other liquids. The principle of this separator (Fig. 6) is gravity settling in a circular vessel from a slow vortex flow induced by a tangential feed. The bottom of the vessel is conical and has a large included angle from which the settled solids are swept toward the central underflow discharge. The radially inward flow at the bottom, which is responsible for the sweeping action, is often referred to as the “tea cup” effect. This is caused by rotating eddies that form in cyclonic flows in vessels having wide included angles or flat bottoms.

The vessel design features a Chinese hat-like conical core stopper above the underflow sump, which is there to prevent the vortex from reaching the latter and reentraining the settled solids. The core stopper is also believed to stabilize and locate the vortex flow in the vessel. Overflow from the vessel is through a wide cylindrical insert through the lid, similar to a vortex finder in a hydrocyclone

(22), and an optional provision can be made for collecting any floatables in a float trap.

Units are available in stainless steel or protected mild steel, often prefabricated, up to 12.5 m in diameter, capable of processing $\geq 5\,m^3/s$ depending on the separation efficiency required. When the separator is used for classification of granular solids, smaller diameter (≤ 4 m) units are used, separating nearly all particles coarser than $\sim 150\,\mu m$.

Another development of the gravity settling principle reported recently is the spiral channel separator (23). This is an attempt to counter some adverse hydraulic conditions encountered in both rectangular and cylindrical settling vessels. The efficiency of these conventional units is often reduced by short circuiting, recirculation, backflows, turbulence, scouring flows, and stagnant zones. These can be reduced by using a rectangular tank with a large length/width ratio (30:1) and wrapping it into a spiral-wound unit with a vertical axis. This might be simply a spiral baffle inserted into an open tank, with the baffle walls rising above the water surface and open at the bottom where the settled sludge provides the seal. This patented and tested idea has not yet found wider acceptance, but it has potential especially for sewage and water treatment works as well as for smaller installations in stand alone or retrofitted units.

Finally, as pointed out in the Design Methods section, when particulate clarification predominates, the most important design parameter is the pool area. The capacity per unit floor area can be substantially increased (and thus the footprint area of the settler reduced) by using the Swedish invention called the Lamella clarifier/thickener where a number of inclined plates (flat or corrugated) or tubes is stacked vertically, closely together. As the design features of these are essentially the same whether for clarification or thickening duties, the reader is referred to the following section on Thickeners for a fuller discussion and the schematic diagram of the Lamella principle.

5.2. Thickeners. The most common thickener is the circular basin type shown in Figure 7. After treatment with flocculant, the feed stream enters the central feed well that dissipates the stream's kinetic energy and disperses it gently into the thickener. The feed finds its height in the basin, where its density matches the density of the inside suspension and spreads out at that level. Solids concentration increases downward in an operating thickener giving stability to the process. The settling solids and some liquid move downward. The amount of the latter depends on the underflow withdrawal rate. Most of the liquid moves upward and into the overflow that is collected in a trough around the periphery of the basin.

A typical thickener has three operating layers: clarification, zone-settling, and compression. Frequently, the feed is contained in the zone-settling layer that theoretically eliminates the need for the clarification zone because the particles would not escape through the interface. In practice, however, the clarification zone provides a buffer for fluctuations in the feed and the sludge levels.

The most important design dimensions of a thickener are pool area and depth. The pool area is chosen to be the largest of the three layer requirements. In most cases, only the zone-settling and compression layer requirements need to be considered. However, if the clarity of the overflow is critical, the clarification zone may need the largest area. As to the pool depth, only the compression layer has a depth requirement because the concentration of the solids in the underflow is largely determined by the time detention and sometimes by the static pressure. Thickness of the other two layers is governed only by practical considerations.

Thickeners are widely used, particularly in the mineral processing industry and in wastewater treatment. Typical applications include thickening of alumina red mud, alumina hydrate, coal tailings, copper middlings and concentrate, magnesium hydroxide, china clay (kaolin), phosphate slimes, potash slimes, pulp-mill wastes, and gas-washing effluents. In hydrometallurgical installations, thickeners are employed for the separation of dissolved components from leached residues in countercurrent washing configurations, eg, in copper, uranium, alumina, and precious metals production (see also Flotation).

The most widely used conventional thickeners in metallurgical and mineral processing applications give solids fluxes (mass flow rate of solids per unit area) in the range of 0.011–0.022 kg/(m^2·s) (24). The conventional thickeners are constructed of steel ($\sim$25-m diameter) or concrete ($<$200-m diameter). The floor is usually sloped toward the underflow discharge in the center. Large thickeners have earth bottoms. Raking mechanisms turning slowly around the center column consolidate the solids in the compression layer and facilitate the discharge of solids. Smaller basins can be covered for conservation of heat or to prevent freezing.

The center-drive mechanism and feed launder are usually supported by a walkway that extends across one-half or the whole diameter of the basin. Devices having drive mechanisms and rakes supported by a truss across the diameter of the thickener are referred to as bridge machines. The bridge thickeners usually do not exceed 25–45 m in diameter. In thickeners with larger diameters, the drive mechanism is supported by a central column or pier and the rates are driven and supported by a drive cage. The sediment is discharged into an annular trench around the bottom of the column.

In even larger caisson thickeners, the central column is sufficiently large to accommodate a discharge pump at the bottom. The discharge passes through the column and along the access walkway above the basin. Caisson thickeners can be built having diameters of up to 200 m. These often have an earth bottom.

The rake arms are driven by fixed connections or dragged by cables or chains suspended from a drive arm that is rigidly connected to the drive mechanism. The rake arms are connected to the bottom of the central column by a special arm hinge that allows both horizontal and vertical movements. This arrangement lifts the rakes automatically if the torque becomes excessive. The drive arm can be attached below the suspension level or, if scaling is a problem, above the basin.

The traction thickener includes a traction mechanism where the movement of the rake is supplied by a single long arm pivoted around the center column and driven by a trolley that moves on a peripheral rail around the basin. Such units have diameters of 60–130 m.

Improved application of flocculating agents has resulted in several other high capacity thickeners capable of handling fluxes up to 0.19–0.38 kg/(m^2·s). A good example is the deep-cone thickener developed by the National Coal Board (NCB, U.K.) (25,26). It is based on the deep-cone vessel used for the processing of coal and metallurgical ores since the turn of this century (see COAL). As can be seen in Figure 8, the vessel is equipped with a slow-turning stirring mechanism (2 rpm) that enhances flocculation in the upper part and acts as a rake in the lower section. The unit is used for densification of froth-flotation tailings at overflow rates from 6.5 to 10 m/h; the final discharge contains 25–35 wt % moisture. Other commercial deep-cone thickeners are of particular advantage, as is the NCB unit, where the final underflow density is increased by the large weight of solids above the discharge point, eg, with flocculated clays.

An unraked version of the deep-cone thickener was developed for the thickening of red mud, the insoluble residue from caustic digestion of bauxite in the Bayer process, in a countercurrent decantation system (27). The deep-cone units are the last two in a nine-stage system. Whereas these are smaller in diameter for the same solids throughput, they are able to produce thicker underflows than the remaining seven conventional, raked thickeners. This greater thickening translates into fewer stages needed for the same washing, and thus represents a further space saving.

There are two U.S. thickeners based on feeding the slurry under the settling–solids interface (sludge blanket) in a way similar to the sludge-blanket clarifiers described. These also offer other features designed to accelerate flocculation and thus increase the capacity. In the Eimco Hi-Capacity thickener, the feed is introduced from the top through a hollow shaft that incorporates flocculant addition and a mixing device. The feed is then directed into the established sludge blanket under the sludge line that partially submerges a set of inclined settling plates. The settled solids are moved by a conventional raking mechanism at the bottom of the basin.

The principle of another high capacity unit, the Enviro-Clear thickener, is shown schematically in Figure 9. Here the flocculated feed enters vertically from the bottom and is directed horizontally at a controlled velocity into the sludge blanket by an impingement plate. A number of other arrangements are also available. The feed can, eg, enter from the top through a center well surrounding the rake driveshaft. A glass window allows visual observation of the sludge line. Available unit sizes range from 4 to 18 m with typical overflow rates from 2.4 to 14.4 m/h. The applications include sugar (qv) and paper (qv) production, and mineral processing.

Stacking of sedimentation units in vertical arrangements increases the capacity per unit area. Multiple compartment or tray thickeners consist of two or more conventional thickener compartments up to 35 m in diameter stacked vertically. Each compartment has a set of raking arms operating from a rotating central shaft common to the whole stack. The compartments are used either in series or parallel.

Another development in this category is the Swedish Lamella thickener (Fig. 10), which consists of a number of inclined plates (flat or corrugated) stacked closely together. The flocculated feed enters the stack from the side feed box. The flow moves upward between the plates while the solids settle onto the plate surfaces and slide down into the sludge hopper underneath, where they are further consolidated by vibration or raking. Even distribution of the flow through the plates is assisted by flow-distribution orifices placed in the overflow exit. In theory, the effective settling area is the sum of the horizontal projected areas of all plates. In practice, only $\sim 50\%$ of the area is utilized. When treating sticky sludges, the whole Lamella pack can be vibrated intermittently or continuously to assist the sliding motion of the solids down the plates. In some instances, the plates are corrugated instead of flat or they are replaced by tube bundles in the tube settler; these tubes are square and have a cross-section of $\sim 50 \times 50\,\mathrm{mm}^2$ and are 950 mm long. The Lamella thickener and the tube settler are used in the treatment of coal, gas-scrubber effluents, fly ash, leach solutions, and many other applications. In coal concentration, the typical overflow rates range from 1.7 to 2.9 m/h.

The Lamella principle has also been extensively used in immiscible liquid–liquid separation, where such units are better known as tilted plate separators. The plates, which can again be either flat or corrugated (with the corrugations running down the sloping surface), also provide surfaces where coalescence of the dispersed phase can take place. The actual configuration of such units has to be somewhat different from those for solid–liquid separation, in order to accommodate the necessary collection–skimming of the light component, such as in oil from water applications. Figure 11 gives an example of such a configuration similar to that marketed, eg, by CJB Develpments Ltd. for oily storm water treatment (Tilted Plate Separator) or by Stetfield Ltd. (Stetpak separator) for oil–water separation in wash and coolant systems, power stations, or marine bilge water applications. The main flow may be down the plates (countercurrently to the rising film of oil), across the plates (cross-flow) or a combination of the two.

An interesting variation on the theme of the Lamella principle in oil–water separation is the Vertical Gravity Separator VGS from Axsia Mozley company of the NATCO Group. Instead of a pack of flat parallel plates, it uses a conical spiral plate pack in a vertical cylindrical vessel. In parallel plate units for solid–liquid separation, another development reported (28) is in making the plates into electrodes so that the settlement rates are enhanced by electrophoretic mobility of the particles. Finally, the overflow rates of the Lamella and tilted plate settlers can also be increased by installing, within the same unit, a deep bed filter in series with the plate pack. However, that brings it into the realm of series connections of separators and that is outside the scope of this article.

Nomenclature

Symbol	Definition
A	area of particle projected in direction of motion; plan area of settling tank or thickener
C	solids concentration
C_D	drag coefficient
C_f	feed solids concentration
C_v	underflow concentration
$F(\epsilon)$	voidage function
F_D	drag force
Fr	Froude number
G	total flux
g	gravity acceleration
$G(x)$	grade efficiency
G_c	critical solids flux
H	height
m	particle mass
Re_p	particle Reynolds number
Stk	Stokes' number
t	time
Q	flow rate
v	particle–fluid relative velocity
$v(c)$	velocity as a function of concentration
v_g	terminal settling velocity
v_p	hindered settling velocity of a particle
$v(t)$	velocity as a function of time
x	particle size
$\Delta\rho$	particle–fluid density difference
ϵ	voidage
μ	liquid viscosity
ρ	liquid (fluid) density
ρ_s	solids (particle) density

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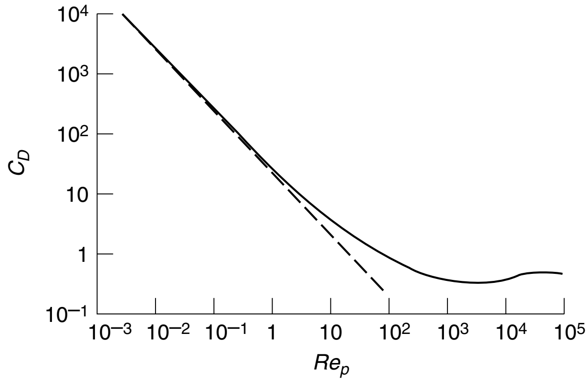

Fig. 1. Drag coefficient versus particle *Re* for spherical particles where (– – –) corresponds to the theoretical value of $C_D = 24/Re_p$ (eq. 4).

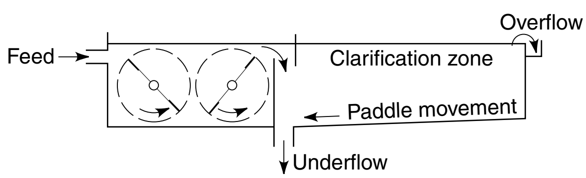

Fig. 2. Schematic diagram of a rectangular basin clarifier having an orthokinetic flocculator where the feed is mixed with a flocculant.

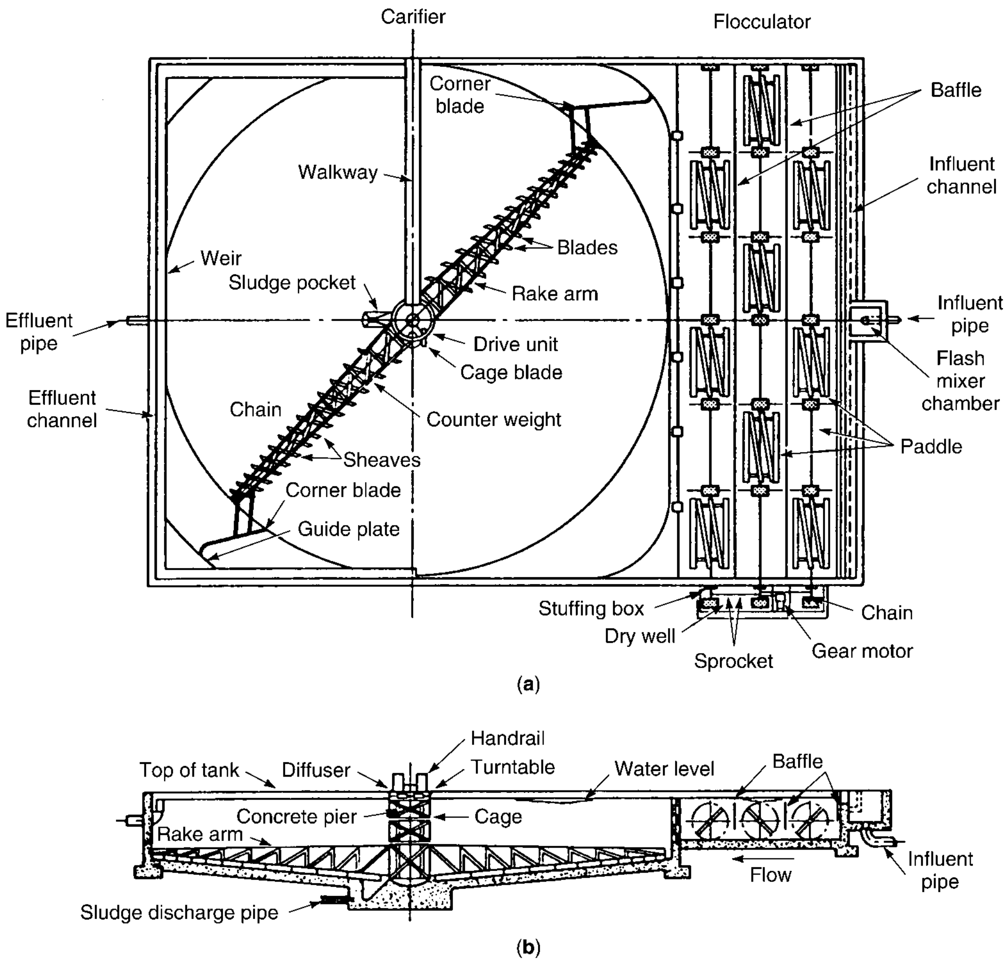

Fig. 3. The Dorrco Flocculator-Squarex clarifier combination (cross-flow arrangement): (**a**) plan and (**b**) sectional elevation. (Courtesy of Dorr-Oliver Inc.)

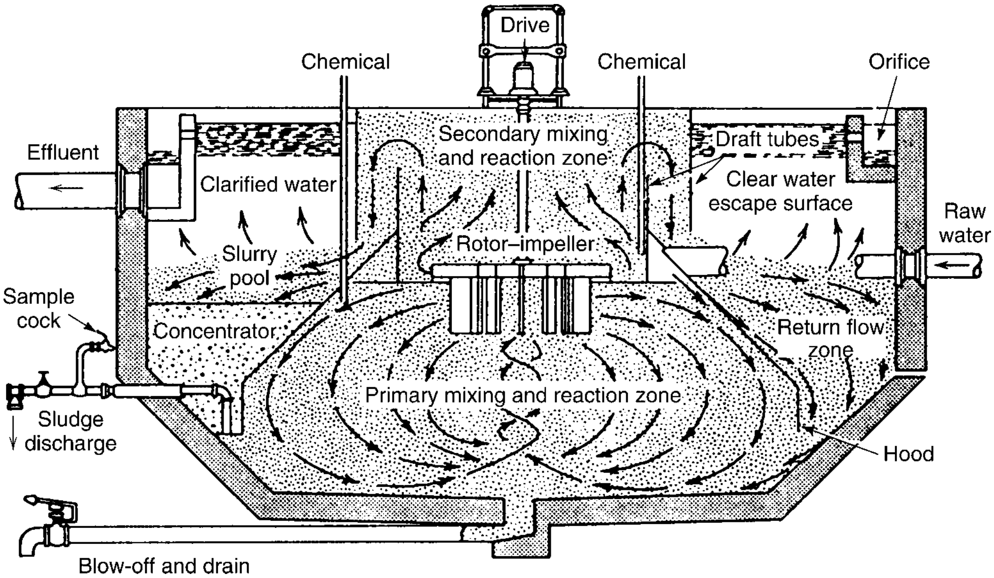

Fig. 4. Functional diagram of an accelerator where the slurry pool is indicated by shaded areas. (Courtesy of Infilco Fuller Co./General American Transportation Corp.)

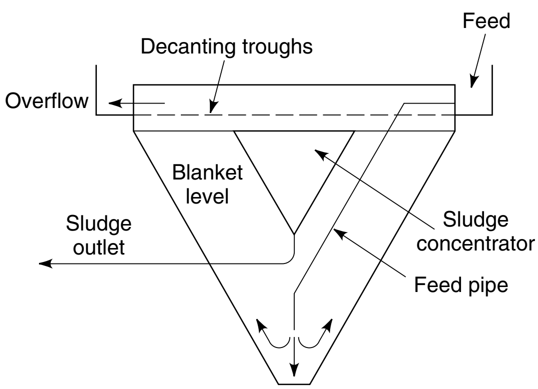

Fig. 5. Hopper-tank sludge-blanket clarifier.

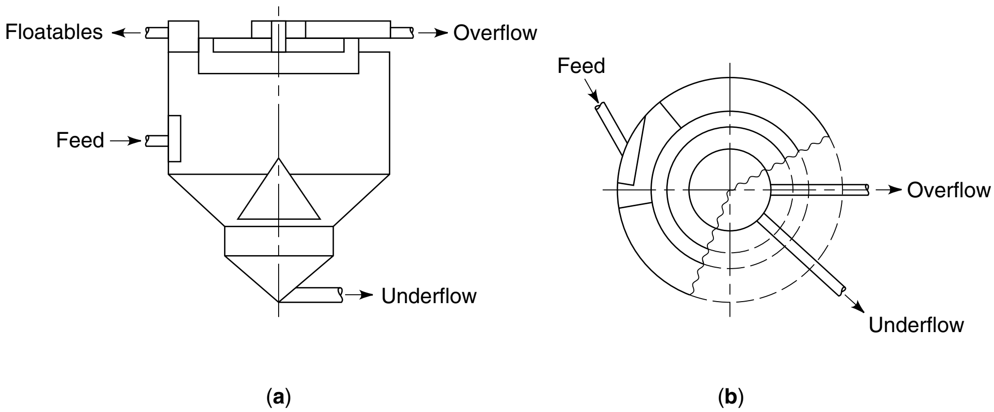

Fig. 6. Schematic diagram of the vortex-induced separator: (**a**) section and (**b**) plan. (Courtesy of Hydro Research & Development Ltd.)

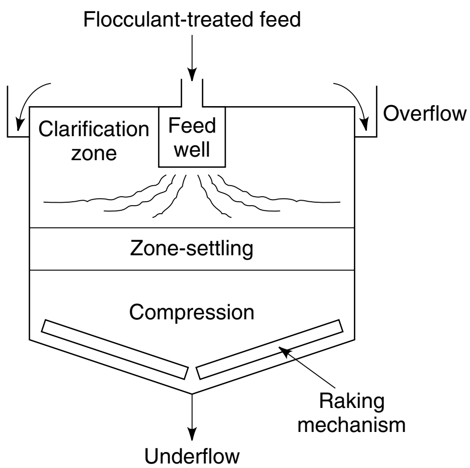

Fig. 7. The circular basin continuous thickener.

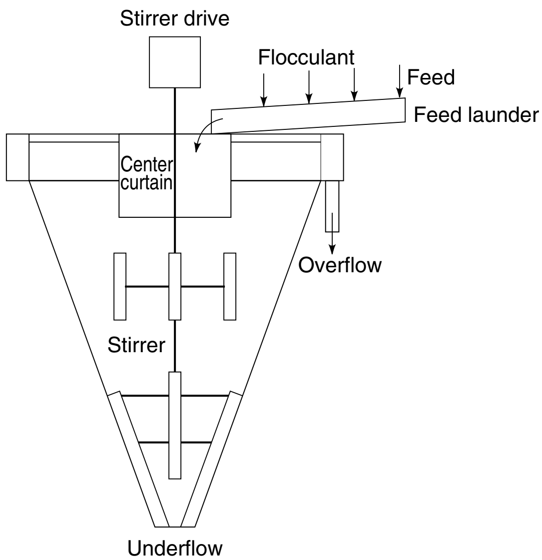

Fig. 8. The NCB deep-cone thickener.

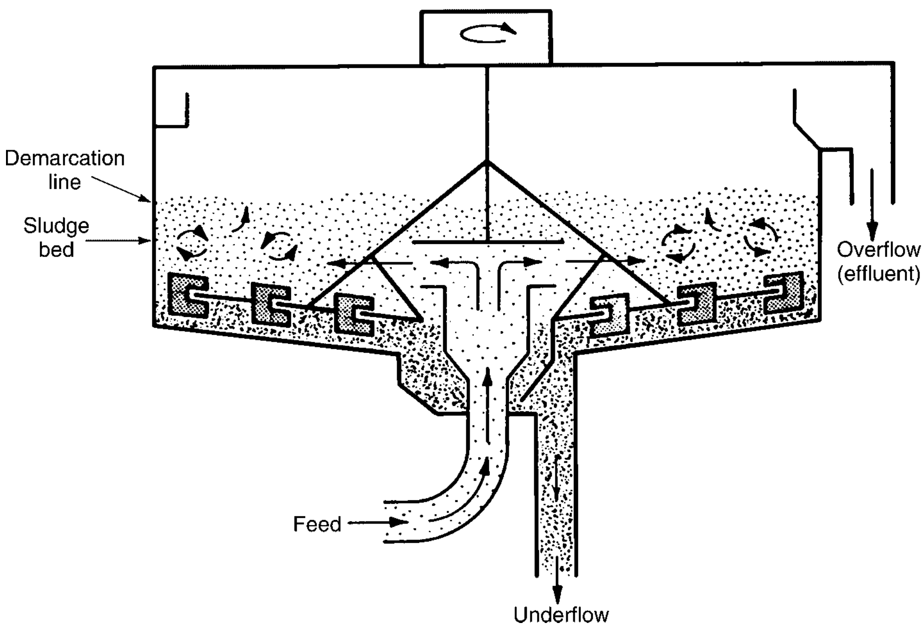

Fig. 9. The Enviro-Clear thickener. (Courtesy of Amstar Corp.)

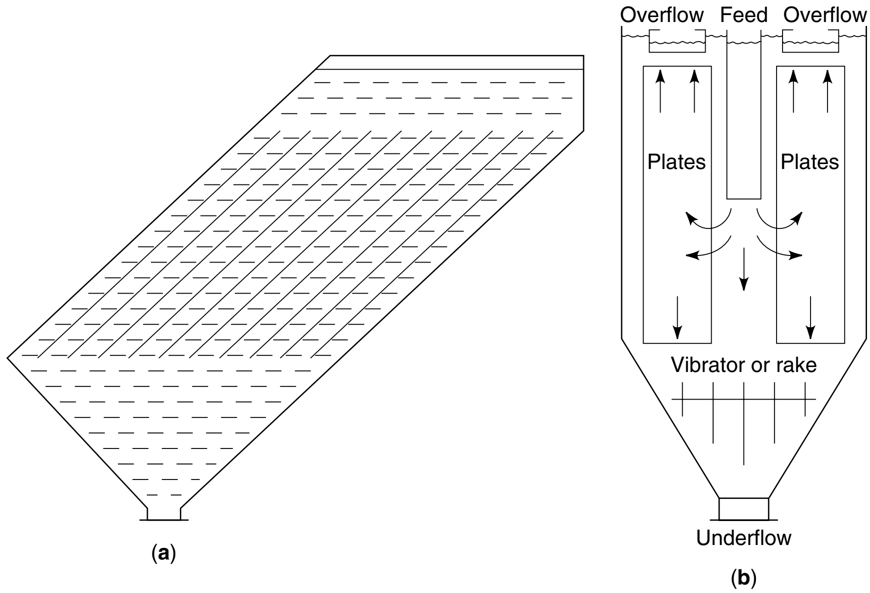

Fig. 10. The Lamella thickener: (**a**) side elevation and (**b**) end elevation.

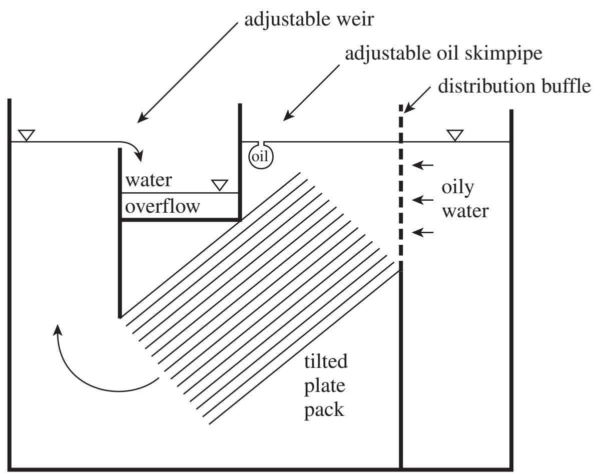

Fig. 11. Schematic diagram of a tilted plate separator for oil–water separation.