Chapter 10
Particle Separations by Filtration and Sedimentation

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Abstract This chapter presents an overview of methods for separating particles (and the like) from fluids by filtration and sedimentation. Focussing primarily on separations from liquids involving recovery of the dispersed 'solids' phase and thickening, features of the types of equipment available are provided together with their advantages, limitations etc., and indicative process models for predicting or characterising their performance. A section on membranes, which are frequently used for the separation of sub-micron (micrometre) particles and nano-sized materials, is also included.

10.1 Introduction

The separation of solids (particles) from fluids by filtration and sedimentation is practised in almost every industry sector including chemical, oil and gas, pharmaceutical, food, water, aeronautical and automotive as well as pollution control. The product may be the particles themselves, the liquid phase, a solute dissolved within the liquid, and sometimes all three components. Cake filtration and sedimentation are common methods for the handling and recovery of particles and this chapter concentrates on descriptions of these two processes.

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Fig. 10.1 A 92 m^2 roller discharge rotary vacuum filter (*left*, Courtesy of Dorr-Oliver Eimco) and a vertical leaf pressure vessel filter (*middle*, Courtesy of LFC). Micrograph of a typical woven filter medium (*right*, Courtesy of Sefar)

10.1.1 Cake Filtration Processes

Industrial filters are available in forms ranging from units capable of handling different filtration applications to those restricted to use with specific fluids and process conditions, from very small to very large scale, and for either continuous or batch operation. Continuous filters are essentially capable of carrying out filter cake formation, deliquoring, washing and discharge, sequentially without interrupting flow of the process feed. Conversely, batch filters can perform the same range of functions but need to be stopped in order to discharge the cake. The surfaces on which separation takes place, the filter media, may be orientated horizontally or vertically, and be either planar or cylindrical. Woven media are frequently used in cake filters, but there are many alternatives and the range of variants is considerable [see 18]. The driving force may be pressure, either a positive pressure or a vacuum, centrifugal or in a few cases just gravity. Photographs of some industrial filters and a typical filter medium are shown in Fig. 10.1.

The above and other factors have led to a bewildering choice in filter design, but even so the underlying principles of all cake filters are similar. In cake filtration a particulate deposit, the cake, accumulates on the surface of the semi-permeable filter medium whilst liquid passes through. Ideally, particles bridge over the pores in the medium and thus limit so called ‘blinding’ where pores can be blocked by particle ingress. Particle bridging is promoted by a raised feed concentration (see Fig. 10.2). After an initial period of deposition the cake itself starts to act as the filter medium whilst further particles are deposited. This accumulation continues until the pressure drop across the cake exceeds the maximum permitted by economic or technical considerations, or until the filtrate (liquid) flow rate falls to an unacceptable level. Provided that cakes are of sufficient thickness to discharge efficiently and any washing/gas deliquoring can be carried out to the required level without cracking, it is generally more productive to operate a filter with thinner cakes.

The most important factor in cake filtration is permeability of the filter cake. Fundamental understanding, as well as the development of useable process models, is conventionally based upon Darcy’s Law which describes the flow of fluid through a porous medium such that

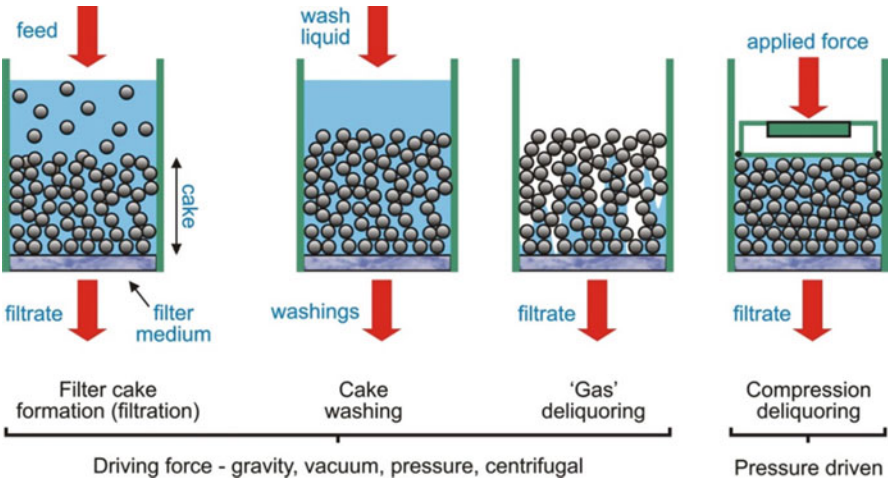


Fig. 10.2 The basic processes involved in deadend cake filters. As water is often the process liquid, the term 'dewatering' is often used interchangeably with deliquoring (Courtesy of Filtration Solutions, UK)

$$u = -\frac{k}{\mu}\frac{dp}{dz} \tag{10.1}$$

where dp is the dynamic (hydraulic) pressure difference across thickness dz of porous medium of permeability k, and u the filtrate flux (superficial velocity) of fluid with a viscosity μ flowing through the bed. Those factors which determine the permeability, the porosity of the filter cake (ε) and size of particles (x) in the cake, together with the particle size distribution and state of aggregation of the particles, are fundamental parameters. They dictate the ease with which any filtration process, and post-treatment process such as compression deliquoring (where pressure is directly applied to the cake to compress it and thereby improve its homogeneity), displacement washing (where clean liquid is passed through the cake to remove solute(s) and, for instance, 'clean' the constituent particles) and gas deliquoring (where gas displaces the liquid in the pores of the cake to 'dry' it), will be accomplished. The combination of sequences shown in Fig. 10.2, together with cake discharge, is known as the filter cycle.

In cake filtration, the cake permeability is most often interpreted via specific resistance (α)

$$k = \frac{1}{\rho_s(1-\varepsilon)\alpha} \tag{10.2}$$

where ρ_s is the true solids density. Important terms such as α, ε and x can be related through, for example, Eq. (10.2) and the Kozeny-Carman equation which is expressed in terms of specific surface (S_0) as

$$k = \frac{\varepsilon^3}{5S_0^2(1-\varepsilon)^2} \quad \Rightarrow \quad \alpha = \frac{180}{\rho_s x^2}\frac{1-\varepsilon}{\varepsilon^3} = \frac{180}{\rho_s x^2}\frac{C}{(1-C)^3} \tag{10.3}$$

where C is the solids concentration in the cake ($=1-\varepsilon$) and spherical particles are assumed ($S_0 = 6/x$). In reality the relationship between variables is complex and made more difficult because of filter cake compressibility which makes a separation more difficult. Here, the cake structure, and hence α and ε, are variant with pressure and generally variant with spatial position in the cake and with time. The consequence is that average values of cake resistance and porosity (or solids concentration) often need to be used in modelling. Despite the inherent complexities, the following general observations can be made:

- *Particle size and size distribution*. Greater rates of filtration can be expected with larger particles in the feed and narrower size distributions are usually favoured. Particles may vary in size from very fine or colloidal matter or molecular aggregates to coarse granular solids.
- *Solution environment and particle surface charge*. Particles less than about 50 μm have a tendency to aggregate and form more compressible filter cakes. Aggregation may arise as a result of the chemical composition of the suspension, or it may be induced/changed by the addition of chemical coagulants (e.g. acids, alkalis, metal salts) or flocculants (e.g. polyelectrolytes) prior to filtration; with chemical addition the particle charge is generally altered. The resulting aggregates are referred to as flocs which can range from being relatively compact and strong assemblies to fragile and 'stringy' collections of loosely bound particles depending on the chosen conditions, particle/liquid interactions etc. The effects of particle charge become more pronounced below 5–10 μm and tend to dominate over hydrodynamic effects below (say) 1 μm. Around the isoelectric point (IEP) of the suspension, i.e. the point of net zero surface charge, more rapid filter cake formation (lower cake resistance) and faster settling rates can be expected.
- *Particle shape*. The majority of particles are of irregular shape, but it is a property that is often difficult to control, and characterise/account for in filtration (and sedimentation) processes. Particle shape affects the volume and surface area of a particle, and hence the specific surface. In general, extremes of particle shape have undesirable effects in filtration.

10.1.2 Sedimentation Processes

Sedimentation is a term that is used to describe particle settling phenomena in suspensions, where particles or aggregates are suspended by hydrodynamic or particle-particle interaction forces with compression being absent. Initially, the particles or aggregates, which are most often more dense than the suspending liquid, settle with return (upward) flow of liquid between the downward moving particles. Eventually the aggregates come into close proximity of one another and

Fig. 10.3 A view of two thickeners showing the central driveheads which drive submerged rotating arms (*left*, Courtesy of Dorr-Oliver Eimco) and a vane decanter centrifuge with the motor drive and gearbox being visible at the right hand end of the unit (*right*, Courtesy of Mitsubishi Kakoki Kaisha)

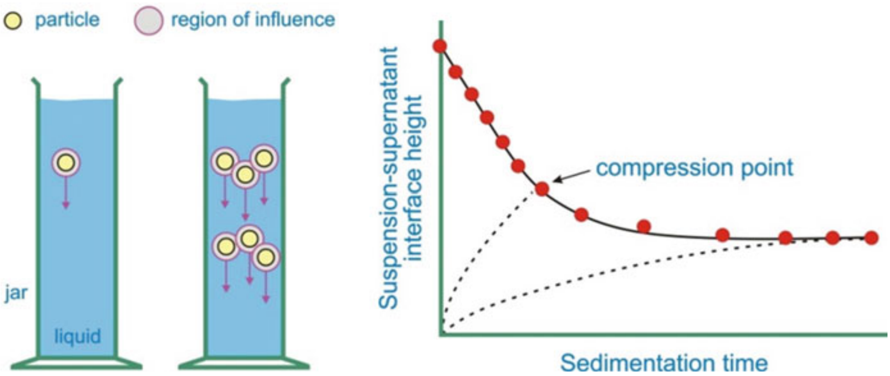


Fig. 10.4 Single particle and hindered settling (*left*) and the typical form of a batch settling curve (*right*) (Courtesy of Filtration Solutions, UK)

those at lower levels feel compressive forces due to the presence of those higher in the mixture; the settled material is known as sediment. Gravity driven sedimentation processes are frequently carried out industrially on a continuous basis at large scale in order to ideally separate the feed suspension into an overflow of clear liquid, the supernatant which is removed toward the top of the unit, and an underflow of more concentrated solids, the particles, that are removed in the form of a sludge/paste toward the bottom (see Figs. 10.3 and 10.4).

In order to increase the settling rate of smaller particles, the force applied to the particles must be increased. One way to do this is to use a centrifugal force. Whereas in a gravity force field particle motion is upwards or downwards, depending on whether the particle is less or more dense than its suspending liquid, in a centrifugal field the motion is radial through the liquid either inwards or outwards, again depending on relative densities. Centrifugal sedimentation can be performed industrially either batchwise or on a continuous basis and units generally occupy a smaller footprint than gravity driven variants.

Particle size, particle density and fluid viscosity are the primary factors to be considered in a sedimentation process, but suspension concentration and particle shape can also have a significant influence. Suspensions with particle diameters of the order of microns settle too slowly for most practical operations. So, to increase their settling rate, the particles are aggregated or flocculated into relatively larger particles known as flocs. Fundamental considerations of sedimentation are normally based upon Stokes Law which relates the drag force on a sphere (F_D) for a low Reynolds number (Re) to liquid properties and particle size such that

$$F_D = 3\pi\mu u x \tag{10.4}$$

where μ is the viscosity of the liquid and u the particle-liquid relative velocity. A single (discrete) spherical particle settling in a gravity field at low concentration is subjected primarily to drag and gravity forces and buoyancy, and a force balance gives:

$$\begin{array}{ccccccc} m\dfrac{du}{dt} & = & mg & - & m\dfrac{\rho}{\rho_s}g & - & F_D \\ \left\{\begin{array}{l}\text{inertial}\\ \text{force}\end{array}\right\} & = & \left\{\begin{array}{l}\text{gravity}\\ \text{force}\end{array}\right\} & - & \{\text{buoyancy}\} & - & \left\{\begin{array}{l}\text{drag}\\ \text{force}\end{array}\right\} \end{array} \tag{10.5}$$

where m is the particle mass, g the acceleration due to gravity and t the time. Incorporating Stokes' law for F_D, integrating and considering longer times gives the terminal (or Stokes) settling velocity (u_t) which is achieved rapidly for small particles:

$$u_t = \frac{x^2(\rho_s - \rho_l)g}{18\mu} \tag{10.6}$$

where ρ_l is the liquid density. If the settling occurs in a centrifugal force field, such as in the spinning basket of a centrifuge, then the particle never reaches a terminal velocity (because the value of r continually increases) and its equation of motion can be written as

$$\frac{dr}{dt} = \frac{x^2(\rho_s - \rho_l)r\omega^2}{18\mu} = u_t\frac{r\omega^2}{g} \tag{10.7}$$

where r is the radius relative to the basket centreline and ω the angular velocity of the basket. That is, the instantaneous velocity of the particle is equal to the terminal velocity in a gravitational field increased by $r\omega^2/g$. The term $r\omega^2/g$ is known as the g-factor and is a basic measure of the separating power in an industrial centrifuge.

When the concentration of the feed suspension increases (say >1 % by volume), the particles are closer together and the motion of any single particle is usually affected by the motion of neighbouring particles. For most practical suspensions the settling rate declines with increasing concentration and the process is referred to as hindered settling. Although there are alternative approaches, for non-flocculated systems Richardson and Zaki [20] showed that the hindered settling velocity (u_h) can be equated to the product of u_t and ε^n, where ε is the voidage or porosity of the suspension and n can take a range of values dependent upon x, D (the diameter of the vessel in which the sedimentation is taking place) and Re_t (the Reynolds number based on u_t); by way of example, for $\mathrm{Re}_t < 0.2$, $n = 4.6 + 20x/D$.

10.1.3 Overview of the Chapter

The many forms of solid/liquid separator developed over the years encompass a wide range of variants which manufacturers claim to give productivity and/or cost benefits. Whilst it is not possible to describe all variants or aspects, Sects. 10.2, 10.3, 10.4, 10.5, 10.6, and 10.7 give a descriptive overview of exemplar equipment types with advantages and limitations highlighted. Typical values for filter sizes, operating parameters etc. are provided, however, the reader should be aware that exceptions are likely to exist. Capital costing equations are given where available and whilst these should be used with some caution, they do provide some basis for comparison; no attempt has been made to correct costings for inflation.

Tables are used to show a range of information, including the solids concentration and particle size found in a typical feed; n/a is used to indicate where information is not available. Process indices give relative values between 0 and 9 for cake dryness (and state), washing performance, liquid product clarity and particle breakage where 9 represents the best performance currently available; a '–' indicates that either a rating is not applicable or the equipment is not capable of performing the operation. For instance, the '1 S, 2, 5, 9' ratings shown in Table 10.6 for the circular basin thickener signify a wet solids discharge in the form of a slurry ('C' designates a cake and 'S' a slurry), poor washing performance, near average liquid product clarity and minimal breakage of particles/aggregates. These indices, and knowledge of typical solids concentration and particle size, can be used as an aid to equipment selection [8].

Figure 10.5 shows the broad classifications of filtration and sedimentation equipment.

In Sects. 10.2, 10.3, 10.4, and 10.5 the principal equations for process models are also presented. Whilst these models can facilitate equipment sizing and predictions of equipment performance in terms of the solids, liquid and solute throughputs, it should be recognised that their successful implementation often requires the experimental measurement of characterising parameters. Such empirical relationships are necessary as it is not currently possible to routinely predict the behaviour of suspension, sediment and cake properties from a knowledge of fundamental particle (e.g. size, shape, charge) and liquid (e.g. viscosity, pH) properties. The interested reader is directed to texts such as Tarleton and Wakeman [27] where full

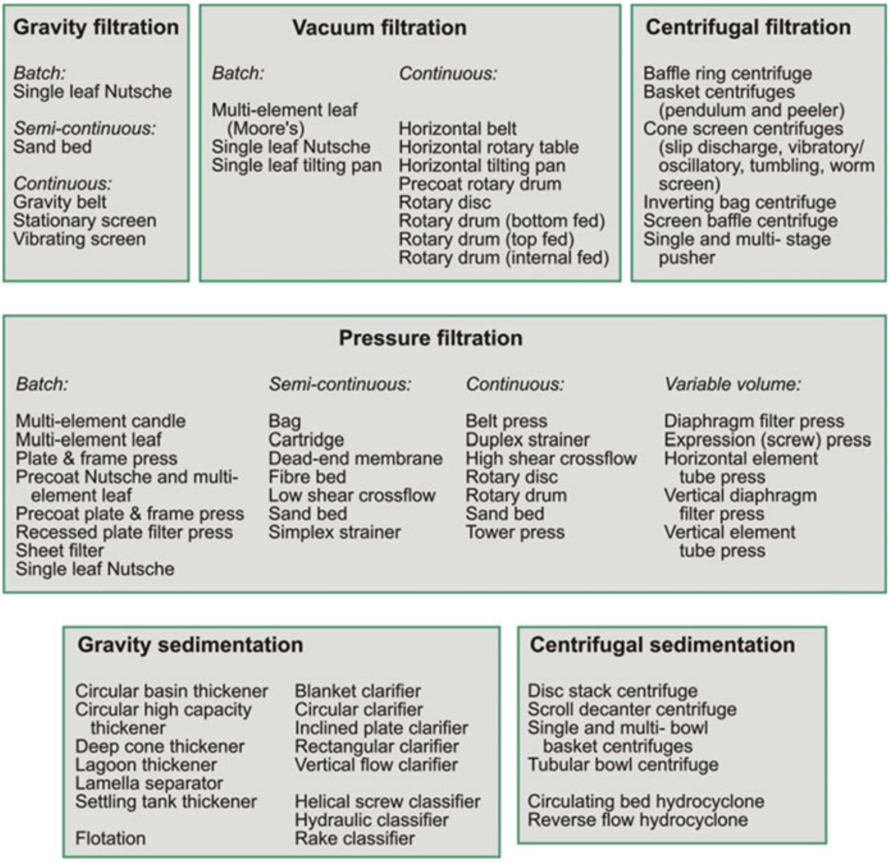


Fig. 10.5 Broad classifications that highlight some of the many forms of solid/liquid separator (not all are described in this chapter)

descriptions and worked examples of model implementations in filters are provided. See also Wakeman and Tarleton [32] for more details about filtration, washing, deliquoring and sedimentation fundamentals, and Wakeman and Tarleton [33] for details of scale-up procedures for filters, centrifuges and membrane separators.

10.2 Pressure Filters and Presses

Pressure filters/presses most often operate in a batchwise manner and use positive pressure above the semi-permeable separating surface(s) to remove liquid and retain solids in the form of cakes. They are used in a wide range of chemical and process industries for the separation of suspensions which contain finer particles that settle slowly and exhibit poor filterability, and/or suspensions that contain higher solids contents. Filtration pressures are typically in the range from 0 to 800 kPa, but can be higher, and usually provided by centrifugal or positive displacement pumps. Smaller units employ compressed gas as the driving medium. Many types can be fully automated to sequence cake formation, washing and deliquoring operations. Some filter processes allow for cake consolidation through the inclusion of flexible diaphragms and several pressure filters have been designed for semi-continuous and continuous operation.

Table 10.1 shows some typical characteristics whilst capital equipment costs for some filters can be estimated over the specified filter area ranges using equations adapted from Couper et al. [6]:

$$\begin{array}{ll} \text{Leaf filter}\,(A = 3 - 230\,\text{m}^2): & 4600A^{0.7}\ \$ \\ \text{Plate and frame filter press}\,(A = 1 - 100\,\text{m}^2): & 2070A^{0.55}\ \$ \end{array} \tag{10.8}$$

10.2.1 Single Leaf (Nutsche)

Typical uses: Fully enclosed batch processing of a wide range of feeds requiring good solids washing.

The Nutsche is a versatile batch filter comprising of a cylindrical pressure vessel with a single planar filter medium (leaf) at the bottom (see Fig. 10.6). The feed suspension, which may be toxic, volatile or flammable, is introduced to the fully enclosed vessel and a constant pressure (typically up to 800 kPa but can be higher) is applied within the vessel to initiate downward filtration. Cake formation may be followed by deliquoring (by gas-blowing) and cake washing (by displacement or reslurry, where the latter involves a repeated cycle of clean wash liquid addition, cake redispersion and subsequent filtration to reform the cake). Smaller Nutsche filters utilise a manual cake discharge and are generally able to operate at higher pressures, whilst larger machines employ mechanical ploughs or rakes to discharge the cake either centrally or through a side port. Fully automated versions of the pressure Nutsche filter are readily available and most include cake smoothing devices to help reduce the problems of cake cracking. A range of filter media can be accommodated with filter areas up to a maximum of 30 m^2, although up to 15 m^2 is more normal.

10.2.2 Multi-element Leaf or Candle

Typical uses: Batch operations with solids forming lower compressibility or incompressible cakes.

Multi-element pressure filters find widespread use and comprise of a cylindrical vessel inside which many horizontal or vertical permeable elements covered by filter cloths are placed (see Figs. 10.1, 10.7, and 10.8). The process suspension, which may be toxic, volatile or flammable is pumped into the vessel at pressures

Table 10.1 Typical characteristics of some industrial pressure filters and filter presses

	Single leaf[a]	Multi- leaf or candle	Filter press	Diaphragm filter press	Tube press	Vertical diaphragm filter press	Belt press	Rotary pressure drum/disc
Operation mode	Batch	Batch	Batch	Batch	Batch	Batch[c]	Continuous	Continuous
Max. pressure (kPa)	800	600	1000	1600	10,000	1600	<500	700
Max. cake thickness (m)	0.5	<0.1	<0.05[b]	0.08	0.03	0.08	Varies	0.18/varies
Filter area (m^2)	0.1–15	2–300	5–2000	5–2000	1–3	2–234	1–75	Up to 40/170
Solids dryness index	6 C	5 C	6 C	8 C	8 C	8 C	8 C	5–6 C
Washing index	8	6–8	8	8	4	8	7	6/–
Liquid clarity index	8	8	8	8	7	8	7	6
Particle breakage index	8	8	8	7	7	7	7	8
Particle size in feed (μm)	1–200	0.5–100	1–100	1–200	1–200	1–200	1–200	1–100
Feed conc. (%w/w)	<1–20+	<1–20	<1–30+	0.3–30+	0.3–30+	0.2–30+	0.2–30+	5–30+
Installed power (kW)[d]	<20	n/a	<25	<35	10–30	20–140	<10	<30

[a]Vacuum and gravity driven variants also exist
[b]Can be up to 0.2 m with additional spacers
[c]Semi-continuous as cake discharge time is relatively short
[d]Can be higher (typically +15 kW) depending on compressed air/gas requirements.

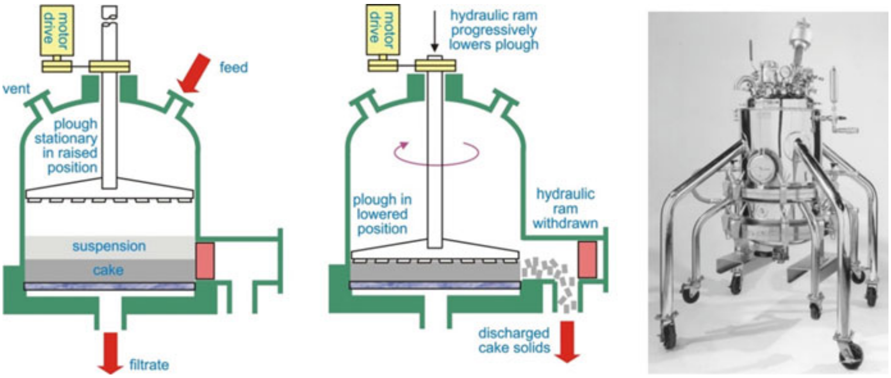


Fig. 10.6 Filtration (*left*) and side port cake discharge (*middle*) in a single leaf (Nutsche) filter (Courtesy of Filtration Solutions, UK). Photograph (*right*) shows a fully assembled 0.15 m^2 Nutsche filter (Courtesy of Pope Scientific). Units can be pressure, vacuum or gravity driven

typically up to 600 kPa, although this can be higher in special units. The positive pressure induces cakes to grow on the outer surfaces of the cloths and filtrate is transported away through the elements via a suitable manifold system. Either flat elements, in the form of square, circular or rectangular leaves, or tubular candles are used. These are spaced sufficiently far apart (up to 20 cm) to avoid the possibility of cakes touching on adjacent elements.

The pressure vessel can be jacketed for operation at elevated (or lowered) temperatures. Up to 150 °C is a typical limit, but this very much depends upon the integrity of the filter medium at the raised temperature. Filter leaves can be automatically extracted for cake discharge if adequate floor/height provisions are made. Relatively frequent cake discharge is normally required and solids are usually removed with the filter leaves in-situ either by vibration, rotating blades, centrifugal force (horizontal elements only) or liquid sluicing to give a wet discharge. Element precoating can be used to, for instance, help prevent particle ingress into the filter medium, and for more extreme duties metallic or ceramic filter elements can be employed.

Vertically mounted vessels: These filters contain either horizontal or vertical leaves with a normal maximum filter area of 120 m^2 (see Fig. 10.7); a specialist filter that is used in alumina processing has a significantly larger filtration area of 440 m^2. They utilise floor area economically but can require excessive height allowance, particularly when leaves need to be withdrawn vertically for cake discharge or cloth cleaning. Horizontal leaves are preferred when either washing is required, rapidly settling feeds are processed or intermittent operation is envisaged, however, the installed cost can be higher as filtration takes place only on the upper surfaces. Multi-element filters having vertical rectangular leaves are best suited to the processing of feeds with a particle settling velocity less than 3 cm s^{-1} but give relatively poor washing performance as cakes tend to prematurely fall off the filter leaves. With tubular candles, cakes form on the outer surfaces of the candle elements and this arrangement is most frequently used when washing is not required.

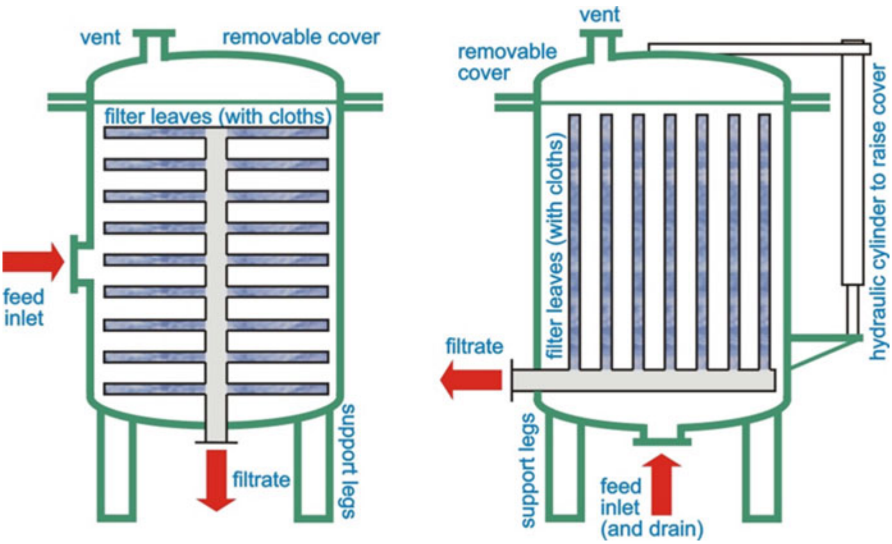


Fig. 10.7 Cross-sectional views of vertical vessel, multi-element pressure filters with horizontal (*left*) and vertical (*right*) leaves (Courtesy of Filtration Solutions, UK)

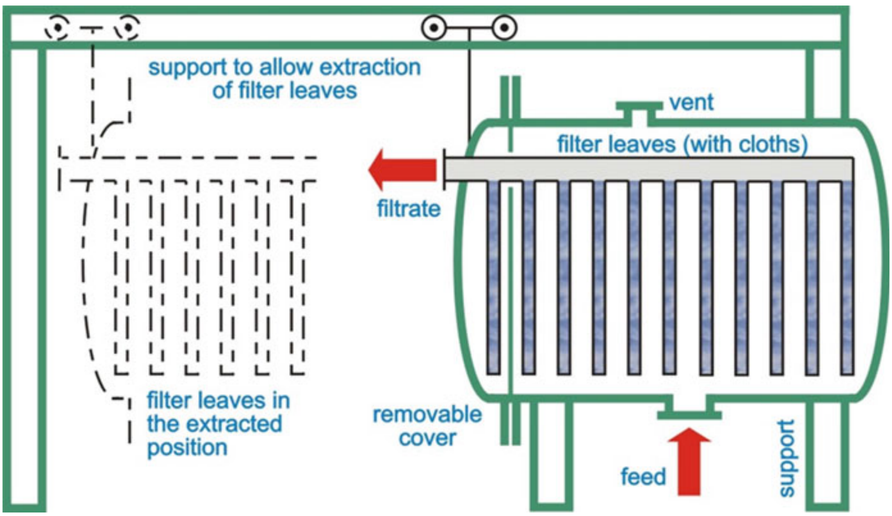


Fig. 10.8 Cross-sectional view of a horizontal vessel, multi-element pressure filter with vertical leaves showing the in-situ and extracted positions of the filter leaves (Courtesy of Filtration Solutions, UK)

Horizontally mounted vessels: These filters contain vertically mounted flat elements with filtration areas up to 300 m^2 (see Figs. 10.1 and 10.8). Whilst needing little height, a large floor space can be required, particularly where filter elements are withdrawn for cake discharge or cloth cleaning. Washing, although possible, can be troublesome if cakes fall off the filter leaves prematurely.

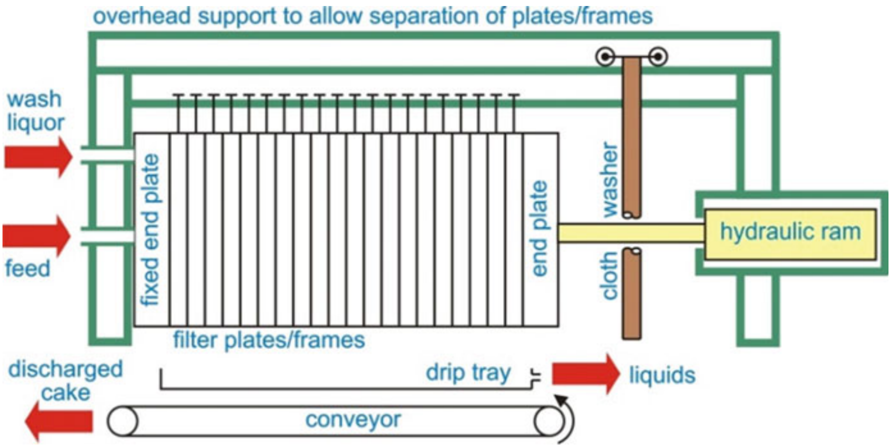


Fig. 10.9 Schematic of a typical overhead bar horizontal filter press showing aspects of filter cycle operations (Courtesy of Filtration Solutions, UK)

10.2.3 Filter Presses

Typical uses: Batch processing of solids forming incompressible and moderately compressible filter cakes.

Although variants exist, the basic unit shown in Fig. 10.9 comprises of many narrow vertical chambers lined on both sides by filter cloths. The chambers, formed between hollow frames and flat filter plates or between adjacent recessed plates, allow for filter cake formations as well as washing and gas-blown deliquoring operations. Suspension is fed to the square, rectangular or circular chambers through a variety of plate porting arrangements and a suitable positive displacement or centrifugal pump. The cakes grow inside each chamber until they meet. Pressures, typically limited to a maximum of 800–1000 kPa, are sufficient to allow centre ported plates to deliver higher solids content feeds whilst bottom and top fed plates generally facilitate more even cake formations and the processing of faster settling suspensions, respectively. Some specially reinforced presses utilise filtration pressures up to 7000 kPa.

Although filter cakes can be removed by reslurrying, they are usually discharged by releasing the mechanical/hydraulic clamping pressure on the press and manually or automatically separating the plates and/or frames. Good cake/cloth release properties are thus preferable, particularly in an automated press. Cloth washing using sprays can also be performed when the filter plates are separated. Modern filter plates are made from either polymers or steel with polymer coatings and formed to provide good drainage surfaces for the covering filter cloths. Plates with dimensions up to 4×5 m are used and filter cloth areas can be as large as 2000 m^2, however, plates up to 2×2 m and cloth areas in the range 50–1000 m^2 are far more typical.

Filter presses are available in two basic forms as shown schematically in Fig. 10.10:

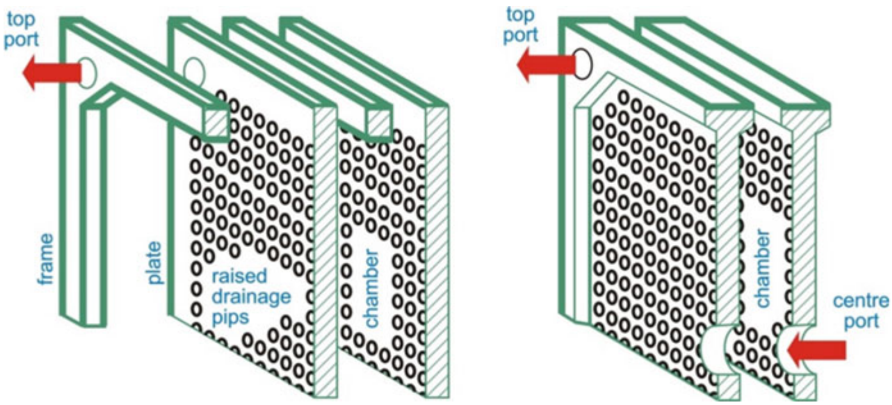


Fig. 10.10 Examples of a top ported plate and frame (*left*) and a centrally ported recessed filter plate (*right*); only one corner of each square/rectangular plate is shown (Courtesy of Filtration Solutions, UK)

Plate and frame press: The basic arrangement comprises an alternating sequence of flat filter plates and frames to facilitate cake formations, normally up to 50 mm thick. The feed suspension and wash liquor enter through the same ports to allow filtration and 'simple' washing, respectively. A more sophisticated arrangement incorporates flat wash plates where suspension and wash liquor enter through separate ports to facilitate improved 'through' washing. In the context of filter presses, plate and frame units offer the advantages of longer cloth life, easily replaced cloths, more uniform cakes and an ability to accommodate alternative filter media such as paper. Their disadvantages include higher capital cost, inlet ports which are prone to blockage at higher feed concentrations and a tendency towards leakage.

Recessed plate press: Here, the functions of the plate and frame are combined such that cake is formed within a recess on each plate. Unlike the plate and frame press, cake thickness is restricted to 32 mm unless additional frames are used as spacers. Feed suspension usually enters through centrally ported plates. The inherent advantages of recessed plate presses include lower initial costs, less tendency towards leakage, an ability to process higher concentration feeds and ease of automation. Their disadvantages include shorter cloth life, longer cloth change times, a tendency to form uneven cakes and an inability to accommodate filter papers.

10.2.4 Variable Volume Filters and Presses

These comprise a family of filters devised to handle suspensions of finer solids which are difficult to pump and/or filter. Typical feeds include suspensions of gelatinous and fibrous materials and those particulates containing occluded liquid within an inherently porous structure.

Fig. 10.11 A sidebar diaphragm press installation (Copyright © Outotec)

10.2.4.1 Horizontal Diaphragm Filter Press

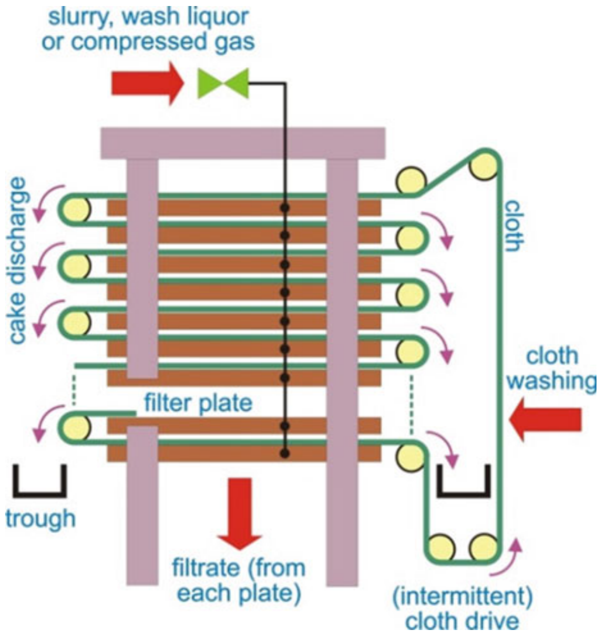
Typical uses: Batch processing of suspensions forming compressible filter cakes where dry cakes and/or efficient post-treatment are required.

These machines are similar in form and general operation to filter presses (see Sect. 10.2.3 and Fig. 10.11). However, the plate surfaces are modified by the addition of flexible diaphragms to form ‘membrane plates’. Although different processing conditions are employed, feed pumping can be stopped after ~80 % of the required volume of filtrate has been produced. In this state the chambers in the press are partially filled with cake and residual unfiltered suspension. The diaphragms on each membrane plate are then inflated from behind to induce pressures up to 1600 kPa which filters the remaining suspension and squeezes the now joined cakes in each chamber. The combination of compression by the diaphragms and subsequent gas deliquoring reduces cake moisture content by up to ~25 % more than that achieved in a conventional filter press. The compression process also tends to produce more uniform cake with improved washing characteristics and release properties; the latter also being aided by the correct choice of filter cloth. Although diaphragm presses are significantly more expensive than conventional filter presses the additional capital and operating costs are often justified by shorter cycle times and the beneficial properties of the final cake.

10.2.4.2 Vertical Diaphragm Filter Press

Typical uses: Semi-continuous processing of solids forming compressible filter cakes that require efficient post-treatment.

Fig. 10.12 Schematic of the vertical diaphragm filter press (Courtesy of Filtration Solutions, UK)



The vertical diaphragm filter press (or tower press, see Fig. 10.12), may be thought of as a conventional horizontal diaphragm press mounted on its end. In place of the fixed filter cloths, a continuous cloth zigzags through the plate pack and is supported on grids within the horizontal chambers. After hydraulically closing and sealing the plate pack, pressure driven cake filtration takes place in the downward direction via a pump. Compression with elastomer diaphragms at up to 1600 kPa, cake washing, and gas-blown deliquoring can then performed in sequence within each chamber. At the end of the cycle the plate pack opens and the cloth is driven forward to discharge the cakes without manual assistance. At the same time, the filter cloth leaving the plate pack is washed by high pressure water sprays to help maintain its permeability.

The largest available units can produce in excess of 200 te h^{-1} (metric tonnes per hour) of dry solids and offer significantly better washing characteristics due to the preferable orientation of the cakes. A filtration area up to 234 m^2 is technically feasible whilst individual plate areas are in the range 0.4–9 m^2. Although filtration usually takes place only on the upward facing part of the filter cloth, this disadvantage is reduced by complete automation and short down-times. Some more complex machines have a filter cloth on both sides of each chamber to form so called ‘double sided presses’, however, particle sizes must be toward the lower end of the allowable range to prevent excessive sedimentation.

10.2.4.3 Tube Press

Typical uses: Batch processing of compressible materials where drier cakes are required.

A tube press comprises of two concentric cylinders where a permeable tube covered with a filter cloth is positioned centrally within a solid outer tube lined by an elastomer diaphragm. The filter cycle is initiated by pumping the feed suspension into the annular space so formed and with sufficient suspension in the press, pressure is applied to induce filtration in the radial direction. This process is most often performed at constant pressure via the diaphragm in two stages where a lower pressure is used initially to promote more even cake formation. When filtration is complete the elastomer diaphragm is further inflated (hydraulically) to compression deliquor the cake.

The most widely used version, the vertical axis tube press, is usually used in parallel groups to give the desired filter area as individual units are limited to ~3 m^2. Squeeze pressures are high, up to 10,000 kPa, and thus very low moisture content cakes can be obtained. Cake discharge is achieved by automatically opening the bottom end cover, lowering the central element by ~0.3 m and applying a reverse back pulse of compressed air to dislodge the cake. Although capital and running costs can be higher, tube presses offer the potential for short cycle times, near optimum cake thickness and reduced thermal drying requirements for the discharged cake.

10.2.4.4 Expression Press

Typical uses: Batch or continuous deliquoring of finer particle suspensions forming compressible filter cakes.

Batch units are characterised by series of cylindrical or square boxes containing semi-permeable cloths at their closed ends. The feed is introduced to the boxes and a moving piston expresses liquid through the cloth at pressures up to 40,000 kPa to leave the deliquored solids. Although many variants exist, the continuous expression press, which may also be classed as a continuous pressure filter, is typified by the most widely used screw press shown schematically in Fig. 10.13. This comprises a variable pitch helical screw rotating at up to 2 rpm inside a perforated cylindrical or conical screen surround. As the (usually flocculated) feed moves through the unit, pressure is progressively increased to continuously express the liquid phase and discharge cake through a variable orifice nozzle. Larger units can have a length and screen diameter in excess of 8 m and 1 m, respectively, and are capable of dry solids throughputs greater than 1 te h^{-1}.

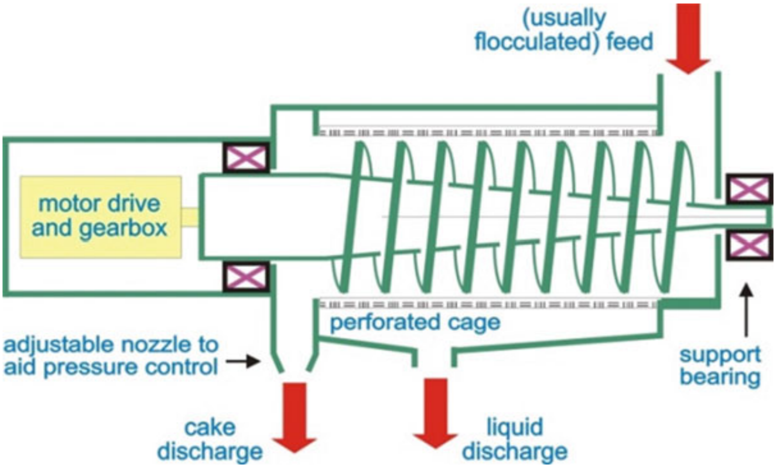


Fig. 10.13 Schematic of a typical expression press (Courtesy of Filtration Solutions, UK)

10.2.5 Continuous Pressure Filters

Continuous pressure filters are often inherently complex and some are based on vacuum driven filters with the addition of an enclosing pressurised shell.

10.2.5.1 Belt Press

Typical uses: Flocculated sludge deliquoring, wastewater treatment.

The belt press shown in Fig. 10.14 was originally conceived to deliquor highly flocculated materials but may be used to process a range of feeds. Although several variants exist, belt presses are characterised by two continuous, tensioned belts. Flocculated material is introduced onto the lower semi-permeable cloth (belt) and then progressively squeezed under pressure as the belts move over a sequence of successively smaller diameter rollers. Liquor is removed through the cloth by a combination of gravity drainage and mechanical squeezing to (ideally) produce a dry, crumbly cake. Filter cloth washing, to help recover initial permeability, may be performed using water sprays at a convenient place after cake discharge. Whilst power consumptions are relatively low, these complex machines have several inherent disadvantages including high flocculant use (dependent on the nature of the feed), relatively low squeeze pressures and a need to use long, strong filter cloths which can be expensive to replace. A guideline (maximum) feed flow rate is 10–15 $m^3\ h^{-1}$ per metre width of filter cloth which typically equates to *ca*. 850 kg $h^{-1}\ m^{-1}$ of dry solids loading.

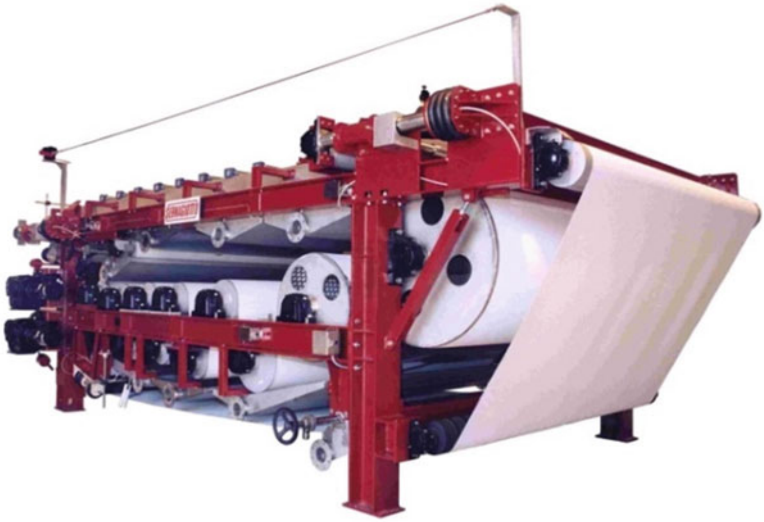


Fig. 10.14 Photograph of a belt filter press (Courtesy of Sernagiotto)

10.2.5.2 Rotary Pressure Drum

Typical uses: Continuous separation of finer particle suspensions where cakes require post treatment.

The rotary pressure drum filter is similar in principle and basic form to the rotary vacuum drum filter (see Sect. 10.3.3). It comprises a rotating, bottom fed drum of area up to 40 m^2 enclosed within a sealed housing (see Fig. 10.15). Rather than applying a vacuum inside the compartments of the drum, the pressure inside the housing is raised by compressed gas up to 700 kPa and this facilitates constant pressure filtration at the outer drum surfaces. Raised temperatures can be accommodated as can volatile and toxic feeds. Both cake deliquoring by gas blowing and displacement washing can be performed reasonably effectively at different pressures through a multi-compartment arrangement within the housing. Cake discharge usually occurs at atmospheric pressure.

10.2.5.3 Rotary Pressure Disc

Typical uses: Continuous, generally larger scale, separation of finer particle suspensions where cake washing is not required.

The rotary pressure disc filter is again similar in form and general operation to its vacuum driven counterpart (see Sect. 10.3.4) with the addition of an enclosing housing. Rotating cloth covered discs, having a total filtration area of between 2 and 170 m^2, are pressurised externally up to 700 kPa to promote cake formations and

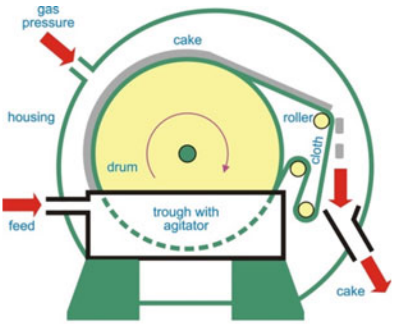


Fig. 10.15 Schematic of a continuous rotary pressure drum filter with belt discharge (*left*, Courtesy of Filtration Solutions, UK) and photograph of a multi-cell pressure drum filter (*right*, Courtesy of Andritz)

generally higher throughputs than equivalent size vacuum units. Although able to handle volatile liquids more readily, the pressure disc filter's inherent advantages are offset by increased costs and the difficulties which can be experienced in cake discharge; the latter is usually achieved by either reslurrying or the use of helical screw conveyors.

10.2.6 Modelling of Filtration

The most widely used process design models for pressure (and vacuum) cake filtration are based on the general filtration equation which is derived from Darcy's Law, Eq. (10.1). The general filtration equation relates the cumulative volume of filtrate (V_f), i.e. the amount of liquid removed, to the filtration time (t_f) and is usually stated as:

$$\frac{dV_f}{dt_f} = \frac{A_f^2 \Delta p_f}{\mu_l \left(\alpha_{av} c V_f + A_f R\right)} \tag{10.9}$$

where A_f is the filter medium area devoted to filtration, Δp_f the filtration pressure, μ_l the viscosity of liquid, c the effective feed concentration and R the filter medium resistance. As most filter cakes are compressible the cake properties, which are characterised by average values of specific cake resistance (α_{av}) and cake solids concentration (C_{av}), are invariably taken to be functions of filtration pressure alone such that:

$$\alpha_{av} = \alpha_0 (1 - n) \Delta p_f^n \tag{10.10}$$

$$C_{av} = C_0 \Delta p_f^{\beta} \tag{10.11}$$

$$c = \frac{s\rho_l}{1 - m_{av}s} = \frac{s\rho_l}{1 - \left(1 + \frac{\rho_l}{\rho_s}\left(\frac{1 - C_{av}}{C_{av}}\right)\right)s} \tag{10.12}$$

where m_{av} is the ratio of mass wet/dry cake and s the mass fraction of solids in the feed. The empirical scale-up constants α_0, n, C_0 and β are most often derived from sequences of constant pressure or vacuum experiments. A value of $\alpha_{av} \leq 10^9$ m kg^{-1} is representative of a very easy to filter material whereas $\alpha_{av} \geq 10^{13}$ m kg^{-1} represents a very difficult to filter material. The value of n is often taken to be the characteristic measure of cake compressibility, i.e. the extent to which a filter cake will compress when it is subjected to a compressive force. When $n \approx 1$ a cake is regarded as very compressible and $n < 0.2$ indicates essentially incompressible.

The cake thickness (L_f) can be related to the cumulative volume of filtrate and cake/particle properties through a mass balance. For example, in the single leaf (Nutsche) filter and filter presses

$$L_f = \frac{V_f}{A_f} \frac{s(\rho_s(m_{av} - 1) + \rho_l)}{\rho_s(1 - m_{av}s)} \tag{10.13}$$

A general feature of pressure driven batch filters is that cake formation is not generally limited by time, but rather by the dimensions and capacity of the filter itself.

10.2.6.1 Cake Formation at Constant Pressure

For a single leaf (Nutsche) filter the filtration pressure is fixed such that α_{av}, C_{av}, m_{av} and c remain constant throughout cake formation as given by Eqs. (10.10), (10.11), and (10.12). Equation (10.9) is then integrated at constant pressure to give:

$$t_f = \frac{\alpha_{av} c \mu_l}{2A_f^2 \Delta p_f} V_f^2 + \frac{\mu_l R}{A_f \Delta p_f} V_f \tag{10.14}$$

With curved elements, such as in the candle filter and the tube press, a modified version of Eq. (10.9) gives:

$$t_f = V_f \left(\frac{\alpha_{av} c \mu_l}{2A_f^2 \Delta p_f} V_f + \frac{\mu_l R}{A_f \Delta p_f} + \frac{4c}{d\rho_c A_f} \left(\frac{\alpha_{av} c \mu_l}{3A_f^2 \Delta p_f} V_f^2 + \frac{\mu_l R}{2A_f \Delta p_f} V_f \right) \right) \tag{10.15}$$

where d is the element diameter and ρ_c the bulk density of the filter cake. In both cases the filtrate flux reduces with time as the cake builds. Cake formation on rotary pressure disc and drum filters is modelled in a similar manner to that described for the corresponding vacuum filter in Sect. 10.3.5.

10.2.6.2 Cake Formation at Variable Pressure

In the case of, for instance, filter and diaphragm presses and multi-element leaf filters, the pressure generally increases throughout cake formation according to a pump curve. That is, the Δp_f vs. q relationship for a positive displacement pump (i.e. constant flow filtration where the flow rate of filtrate is constant with time) or a centrifugal pump (i.e. variable pressure-variable flow filtration where both pressure and flow rate are variant). Substituting $q = dV_f/dt_f$ into Eq. (10.9) and rearranging gives an expression for the volume of filtrate:

$$V_f = \frac{A_f}{\alpha_{av}\mu_l c}\left(A_f\frac{\Delta p_f}{q} - \mu_l R\right) \tag{10.16}$$

As the pressure varies several parameters are time variant whereby α_{av}, C_{av}, m_{av} and c change throughout filtration to an extent dictated by Eqs. (10.10), (10.11), and (10.12). The pressure is specified by a given position on the pump curve and the filtration time is obtained from:

$$t_f = \int_0^{V_f} (1/q)dV_f \tag{10.17}$$

It is generally necessary to solve Eqs. (10.16) and (10.17) using numerical techniques in conjunction with a knowledge of the pump characteristics in order to find sequential values of V_f and t_f, and consequently the amount of solids processed.

10.2.7 Modelling of Compression Deliquoring

In the tube press and all variants of the diaphragm press, flexible diaphragms can be inflated at a constant pressure to facilitate compression deliquoring of the filter cake (s). This consolidation is most frequently performed immediately after cake formation to remove unwanted liquor and/or improve the distribution of cake solids to aid any subsequent cake washing and gas deliquoring. As a rule-of-thumb, the duration of cake compression equals the total filtration time.

The cake properties during compression deliquoring are related to the compression pressure (Δp_c) by:

$$C_e = C_{e0}\Delta p_c^{\gamma} \tag{10.18}$$

$$(C_{av})_\infty = C_0\Delta p_c^{\beta} \tag{10.19}$$

where C_e is the modified consolidation coefficient which characterises the process, $(C_{av})_\infty$ the equilibrium cake solids volume fraction at infinite consolidation time, and C_{e0}, γ, C_0 and β are empirical scale-up constants derived from sequences of constant pressure consolidation experiments. Process design calculations for compression deliquoring are primarily based on the theories of Shirato et al. [22, 23] where the cake thickness (L_c) at a given time (t_c) is related to a dimensionless consolidation time (T_c) by:

$$L_c = L_0 - (L_0 - L_\infty)\frac{\sqrt{\frac{4(j_{II})^2 T_c}{\pi}}}{\left(1 + \left(\sqrt{\frac{4(j_{II})^2 T_c}{\pi}}\right)^{2v}\right)^{1/2v}} \tag{10.20}$$

$$T_c = \frac{i^2 C_e t_c}{\omega_0^2} = i^2 C_e t_c \left(\frac{A_c \rho_s}{M_s}\right)^2 \tag{10.21}$$

$$L_\infty = \frac{M_s}{A_c \rho_s (C_{av})_\infty} \tag{10.22}$$

where L_∞ is the cake height at infinite consolidation time, j_{II} is a factor that accounts for any filter element curvature, i the number of drainage surfaces (normally = 1), M_s the mass of solids in the cake, ω_0 the volume of solids per unit filter area, A_c the active filter area during consolidation and v is an empirical scale-up constant. Values for i, A_c and j_{II} vary according to the type of press, but use of Eq. (10.20) allows the change in cake thickness with time to be evaluated.

10.2.8 Modelling of Cake Washing

Design equations for the most frequently used displacement washing process are usually based on the dispersion model [see 30]. The model requires the determination of a dispersion number (D_n) that characterises the washing process and use of a design chart (Fig. 10.16) which allows the amount of wash liquid used (specified as the number of wash ratios, W) and the amount of solute removed from the cake (e.g. fractional solute recovery, F) to be found. The pressure (Δp_w) is fixed throughout washing and cake properties such as specific resistance, solids concentration and thickness (L_w) are normally assumed to remain constant.

In the dispersion model the superficial velocity (u) and pore velocity (v) of wash liquor are related to the intrinsic properties of the cake through a version of Darcy's Law

$$u = \frac{\Delta p_w}{\mu_w(\alpha_{av}\rho_s L_w C_{av} + R)} \tag{10.23}$$

$$v = \frac{u}{\varepsilon_{av}} = \frac{u}{1 - C_{av}} \tag{10.24}$$

where μ_w is the viscosity of the wash liquor. For the solute, the ratio of the molecular diffusion coefficient (D) to the axial dispersion coefficient (D_L) is

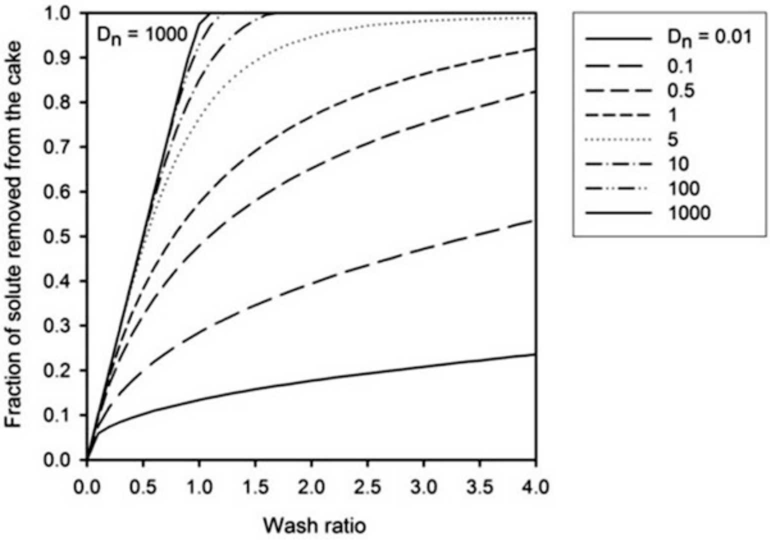


Fig. 10.16 Variation of the fraction of solute removed (F) from a saturated filter cake with wash ratio (W) and dispersion number (D_n)

dependent on the product of the Reynolds number (Re) and Schmidt number (Sc) as well as the cake thickness. For instance, with ReSc ($= vx/D$) >1 and L_w <10 cm

$$\frac{D_L}{D} = 0.707 + 55.5(\mathrm{ReSc})^{0.96} \tag{10.25}$$

and hence

$$D_n = \mathrm{ReSc}\frac{L_w}{x}\frac{D}{D_L} \tag{10.26}$$

Depending on whether t_w, W or F are specified the other values can be found with Eqs. (10.23,) (10.24), (10.25), and (10.26) and use of Fig. 10.16.

10.2.9 Modelling of Cake Gas Deliquoring

Process design calculations for the gas deliquoring of filter cakes require the specification of a threshold pressure/vacuum (p_b), which is the pressure or vacuum required to start deliquoring, and an irreducible saturation (S_∞), which is the cake saturation where deliquoring due to liquid displacement stops [see 31]. These can be either measured in a capillary pressure experiment or the former can be calculated with some confidence using:

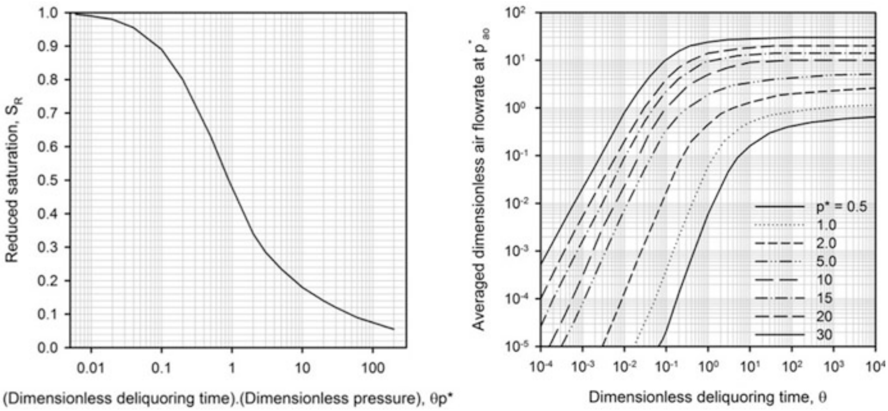


Fig. 10.17 Reduced saturation of a filter cake as a function of dimensionless deliquoring time (*left*) and the dimensionless air/gas flow rate through a filter cake as a function of dimensionless pressure p^* (*right*). In both cases deliquoring is by or pressure (or vacuum) applied in the gas phase

$$p_b = \frac{4.6 C_{av} \sigma}{x(1 - C_{av})} \tag{10.27}$$

where σ is the cake liquid surface tension. Knowledge of p_b and S_∞, together with the intrinsic properties of the cake and two design charts (Fig. 10.17), allows the evaluation of cake moisture content (M) or deliquoring time (t_d) as well as the flux of gas (or gas/air rate) required to deliquor the cake. The applied gas pressure (Δp_d) is fixed and properties of the cake such as thickness (L_d), solids concentration and specific resistance are assumed to remain constant.

The basis of the approach requires the calculation of a dimensionless time (θ) or reduced saturation (S_R) depending on whether time (t_d) or the required saturation (S) is known, respectively, whereby

$$\theta p^* = \left(\frac{t_d p_b}{\alpha_{av} \rho_s C_{av} (1 - C_{av}) \mu_l (L_d)^2 (1 - S_\infty)} \right) \left(\frac{\Delta p_d}{p_b} \right) \tag{10.28}$$

$$S_R = \frac{S - S_\infty}{1 - S_\infty} \tag{10.29}$$

In this way the unknown value (either t_d or S) can be determined with the aid of Fig. 10.17 (*right*) and the cake moisture content (M) can be calculated using

$$M = \frac{100}{1 + \frac{\rho_s}{S \rho_l} \left(\frac{C_{av}}{1 - C_{av}} \right)} \tag{10.30}$$

Calculations for the amount of gas/air required to drive the gas deliquoring are more complex and best described by worked example [see 27]. In general, calculations for air rate tend to over-estimate requirements which leads to an over-specification of compressor (or vacuum pump) requirements.

10.3 Vacuum Filters

A category of filter that uses vacuum induced driving forces and semi-permeable media to facilitate a separation. Whilst pressure differences across a filter are limited to less than 85 kPa (usually <75 kPa), most units are capable of processing a wide range of (coarser particle size) feed materials in a continuous manner. Many types employ a rotary valve arrangement to set different vacuum levels over sequential phases in a filter cycle thus facilitating more control over cake formation, deliquoring and washing; any given phase is operated at a constant vacuum level. Several vacuum filters have counter-current washing capability where the washings from downstream are used to wash filter cakes further upstream in order to reduce overall wash liquor consumption. Although it is possible to enclose some types to conserve heat and/or vapours the processing of more volatile constituents at higher altitudes can cause significant problems. Most variants have a minimum cake thickness requirement to help ensure adequate cake discharge. Woven filter cloths or specially developed coated media are used almost exclusively.

Table 10.2 shows typical characteristics of the most common continuous vacuum filters; some batch variants are listed in Fig. 10.5. Capital costs for continuous vacuum filters can be estimated over the specified filter area ranges using equations adapted from Couper et al. [6]:

$$\begin{array}{ll}
\text{Horizontal belt}(A = 1 - 110\,\text{m}^2): & 1.1 \times 10^5 A^{0.5}\ \$ \\
\text{Rotary disc}(A = 10 - 300\,\text{m}^2): & 3.2 \times 10^4 A^{0.43}\ \$ \\
\text{Rotary drum, knife discharge}(A = 1 - 140\,\text{m}^2): & -2.9A^2 + 2470A + 65100\ \$ \\
\text{Rotary drum, belt discharge}(A = 1 - 75\,\text{m}^2): & -14A^2 + 3710A + 72800\ \$
\end{array} \tag{10.31}$$

10.3.1 Horizontal Belt

Typical uses: Separation of relatively free filtering solids where good post treatment is required.

The horizontal belt is a continuous filter with an endless cloth supported on a perforated belt (see Fig. 10.18). The belt and cloth are driven around two rollers and across a sequence of evacuated suction boxes at linear speeds up to 0.5 m s^{-1}. The feed suspension is introduced at one end and filtered to produce a cake. The length of the filter, which can be in excess of 60 m, is arranged to allow adequate cake formation as well as the chosen number of deliquoring and washing operations. Due to the ease with which wash liquors can be segregated, it is relatively simple to perform counter-current washing to exacting requirements provided wash liquor

Table 10.2 Typical characteristics of some industrial continuous vacuum filters

	Rotary drum	Horizontal belt	Rotary table	Rotary tilting pan	Rotary disc
Max. submergence (% of cycle)	30[a]	As req'd	As req'd	As req'd	28[a]
Area under active vacuum (% of cycle)	75–80[e,b]	As req'd	80	75	75[b]
Max. for washing (% of cycle)	29[c]	As req'd	As req'd	As req'd	None
Max. for deliquoring only (% of cycle)[(d)]	45–60[e]	As req'd	As req'd	As req'd	45–50
Cake discharge (% of cycle)	10–25[e]	0	20	25	25
Min. cake discharge thickness (mm)	1–6[e]	3–5	20	20–25	10–13
Filter area (m^2)	0.1–180	3–170	5–320	5–200	1–300
Solids dryness index	6 C	7 C	7 C	7 C	4 C
Washing index	7	9	8	9	–
Liquid clarity index	7	7	7	7	6
Particle breakage index	8	8	8	8	8
Particle size in feed (μm)	1–200[e]	20–80,000	20–80,000	20–80,000	1–700
Feed conc. (%w/w)	1–20[e]	5–30+	10–30+	5–30+	5–20
Motor drive power (kW)	0.5–20[f]	1–90	n/a	3–45	5–60[f]
Power to generate vacuum (kW)	$\approx$ (filter area in m^2)$^{1.25}$				

[a]Higher submergence (cake formation period) is available, consult manufacturer
[b]Assumes no trunnion stuffing boxes
[c]Washing starts at the horizontal centreline on the rising side and extends up to 15° past top dead centre
[d]Deliquoring means drainage of liquor from cake formed during submergence
[e]Depends on variant
[f]Including agitator

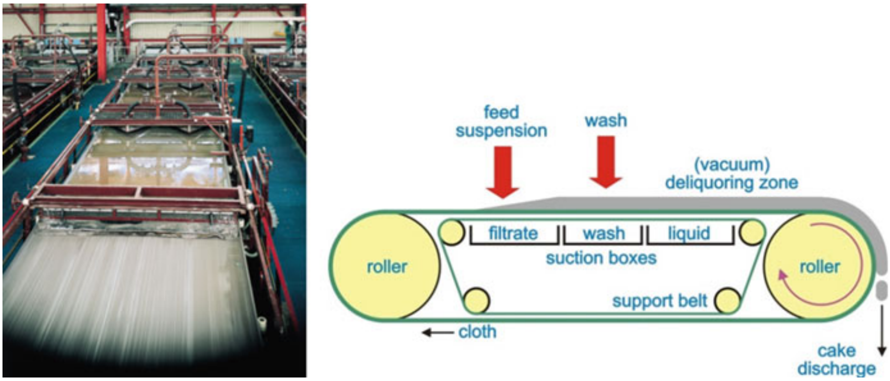


Fig. 10.18 Photograph of a vacuum driven horizontal belt filter (*left*, Courtesy of Clear Edge) and a typical filter cycle (*right*, Courtesy of Filtration Solutions, UK)

carry-over into the next suction box is avoided. The final cake is naturally discharged as it passes over the second roller and separation of the belt and cloth beneath the filter allows the cloth to be cleaned by sprays as it returns.

Horizontal belt filters are best suited to the larger scale filtration of medium and faster settling slurries. Although they occupy a large floor space and the cost of installation per unit filter area is relatively high, these disadvantages are generally offset by full automation, flexibility, high capacity and relatively high speeds of operation. Filters may be sealed to prevent the escape of heat and/or vapours, however, should the belt or cloth be damaged then replacement of either component can be expensive. Some units are programmed to operate semi-continuously via intermittent motion of the belt.

10.3.2 Horizontal Rotary Filters

Typical uses: Processing of fast settling slurries where good washing is required.

The two forms of horizontal rotary filter differ primarily in the manner in which the filter cloth is arranged around the periphery of the circular separation surface (see Fig. 10.19).

Table: A filter comprising of a rotating horizontal table with an annular filter cloth. Vacuum is applied over individual segments of the table to initiate filtration and the formed cake is subsequently deliquored and/or washed by sprays according to requirements. The final cake is continuously discharged via a screw conveyor to typically leave a 3–4 mm residual heel of cake on the cloth. Dependent on the properties of the solids, the presence of the heel can have undesirable consequences for future cycles and necessitates the use of more open filter media with the potential for cloudier filtrates. As the cloth surface is not physically divided into individual sectors some short-circuiting of the feed may occur as well as

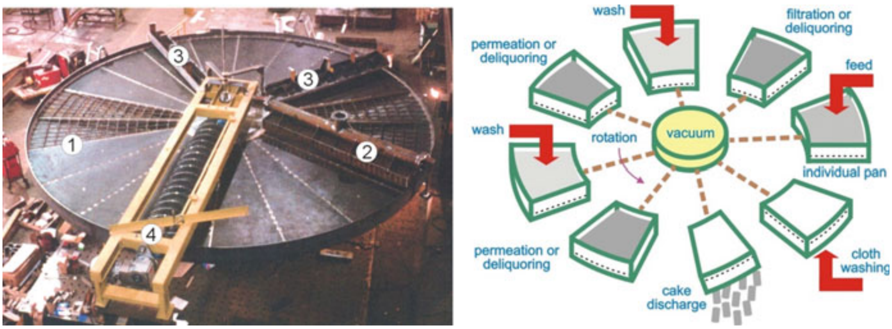


Fig. 10.19 Photograph of a table filter undergoing refurbishment (*left*, Courtesy of Dorr-Oliver Eimco). *1* individual segments on which the cloth is mounted; *2* feed trough; *3* wash liquor delivery or additional feed points; *4* screw conveyor for cake discharge. The schematic (*right*) shows a tilting pan filter (Courtesy of Filtration Solutions, UK)

unwanted mixing of the wash and mother liquors. Although cloth area can be up to 320 m^2, alignment difficulties usually restrict machine size and the available area to below 100 m^2.

Tilting pan: A filter that is similar in general form to the table filter, except the cloth is replaced by a series of annular sectors or pans, each one of which is lined along its perforated bottom by an individual filter cloth. After suspension is introduced to a pan, filtration, deliquoring and washing can take place under the applied vacuum before cake discharge is achieved by a relatively complex tilting mechanism that inverts the pan. The discharge procedure, which may be assisted by air blow-back, leaves no heel of cake and thus in-situ cloth medium cleaning is readily performed using sprays. As all liquors are kept separate, there is little unwanted mixing of mother and wash liquors and counter-current washing can be very good. Tilting pan filters are available with total filter cloth areas up to 200 m^2, however, many of their inherent advantages are offset by their mechanical complexity and higher capital cost.

10.3.3 Rotary Drum Filters

Typical uses: Separation of relatively easy to filter suspensions, the efficiency of cake post treatment depends on the type of drum.

The versatile rotary drum filter (or rotary vacuum filter, RVF) is perhaps the most widely used of the continuous vacuum filters (see Figs. 10.1 and 10.20). The generic type is characterised by a rotating, multi-compartment drum covered externally by a fixed filter cloth of total surface area between 0.1 and 180 m^2; most commercial units are in the range 1–80 m^2. The drum on bottom fed units rotates about a horizontal shaft at speeds up to 5 rpm and is partially submerged in a mechanically agitated tank of constantly replenished suspension. A vacuum is applied inside the drum via a control valve to initiate upward filtration over the submerged region. As the drum rotates so the filter cake formed on the cloth is exposed and a limited number of deliquoring and washing procedures can then be performed at the appropriate level of vacuum. Although washing efficiency is reasonable, the restricted horizontal filter area near the top of the drum prevents further efficiency gains. Cake discharge normally occurs at a point where the final cake is almost vertically oriented and variants differ primarily in the manner in which cake is discharged:

Knife/scraper discharge: The most widely used method when avoidance of cloth blinding can be more or less guaranteed, but requires a minimum 6 mm cake formation. The knife is arranged to leave a heel of cake on the cloth and thus avoid potential damaging contact between the knife and drum. If the cake is thinner then air blow-back can be employed to break the vacuum and assist discharge, though with some designs this tends to cause filtrate to re-enter the cake.

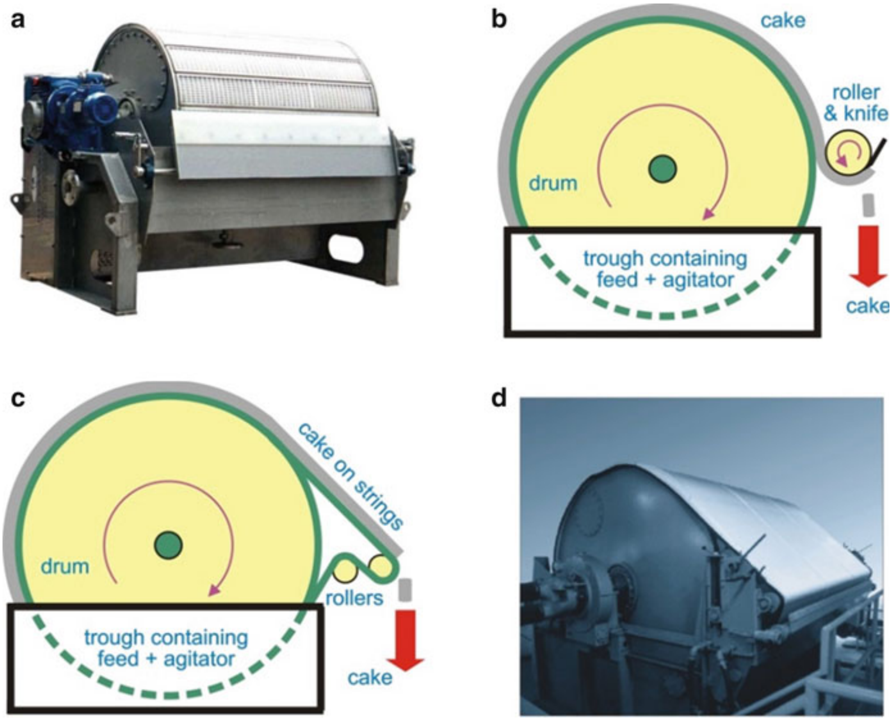


Fig. 10.20 Representations of bottom fed rotary vacuum drum filters. (**a**) Knife/scraper discharge (Courtesy of Filtration Services); (**b**) roller discharge (Courtesy of Filtration Solutions, UK); (**c**) string discharge (Courtesy of Filtration Solutions, UK); (**d**) belt discharge (Courtesy of Dorr-Oliver Eimco)

Roller discharge: Generally used for the complete removal of finer particle, sticky cakes that don't crumble. The 0.5–3 mm cake must preferentially stick to the roller placed adjacent to the drum, a process that is aided by the shearing action of the faster rotating roller. A knife scraper continuously removes the cake from the roller.

String discharge: Suited to the discharge of fairly thick cakes of a fibrous nature that don't crumble. A number of endless strings pass over a series of external rollers and the surface of the filter cloth. In the discharge zone the strings lift away from the cloth to remove the cake completely. The strings may be replaced by endless wires, coils or chains as appropriate.

Belt discharge: Mostly used for the discharge of sticky, thin cakes (<3 mm) whose constituent particles may tend to blind a filter cloth. In this case the endless cloth is not fixed to the drum, instead it passes around its outer periphery and a series of external rollers. At the discharge point the cloth lifts away from the drum and movement over the rollers causes all the cake to be released. The exposed cloth is then cleaned by sprays before returning to the drum. Although relatively expensive to install, belt discharge systems are claimed to raise throughput by up to 30 %.

Drum filters can be enclosed to prevent the escape of heat and/or vapours and typically operate with a submergence equivalent to 30–40 % of the available filter cloth area. Where cake formations are more difficult this may be raised to 60–75 % to produce a 'submerged axis' filter. However, the option is rarely preferred as the trunnion mounts for the drum must be wholly or partly flooded with slurry which necessitates the use of stuffing boxes (sealing units) with their attendant capital and maintenance costs. For faster settling solids the mechanical agitation of suspension in the tank is more problematical and top fed or internally fed drums are options, but rarely favoured as, for instance, horizontal rotary filters are generally better suited.

10.3.4 Rotary Disc Filters

Typical uses: Continuous larger scale separation of relatively free filtering suspensions where washing is not required.

Units comprise up to 12 flat, circular discs mounted vertically on a central horizontal shaft (see Fig. 10.21). The discs, which are themselves permeable, are usually covered externally by sectored filter cloths. Rotation causes them to pass through individual agitated tanks containing the feed suspension(s) and the vacuum is applied inside the discs to promote cake filtration. After this is complete, deliquoring by air-suction can be performed, however, cake washing is essentially impossible due to the vertical cake formation. The final cakes are discharged by blade or wire scrapers on either side of the discs. As an air blow-back system is often employed to aid cake removal, wetter cakes tend to be discharged (in comparison to drum filters) and the discharge of thin cakes can be particularly troublesome. The need to place the discharge scrapers close to the cloth surface frequently leads to cloth damage, though cloth sectoring means only portions need be replaced.

Rotary disc filters are available at a relatively low capital cost with total cloth areas between 1 and 300 m^2. They have an inherently large filter area to floor space

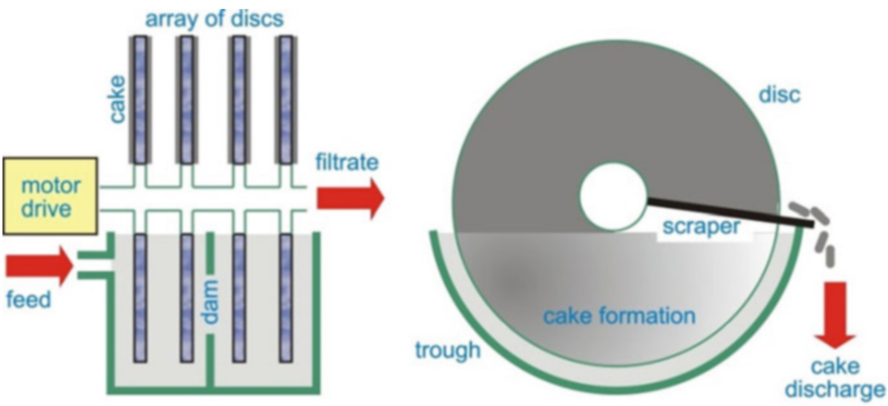


Fig. 10.21 Side and end views of a rotary disc filter (Courtesy of Filtration Solutions, UK)

ratio and their flexibility is enhanced by an ability to process multiple feedstocks at the same time within a single unit.

In a variant, the filter cloths and discs are replaced by sintered alumina membrane discs with near uniform micropores. Although a relatively small vacuum pump is needed to promote cake formation, cake deliquoring proceeds via capillary action with little or no air consumption which significantly reduces the cost of vacuum generation (by *ca.* 90 + %). Whilst the use of a ceramic membrane material allows a clear filtrate to be readily obtained, disc replacement can be expensive. Moreover, backflushing with filtrate and periodic in-situ ultrasonic cleaning must be performed in order to maintain the original permeability of the filtering discs. Ceramic disc filters are available with filter areas of 15–240 m^2 and have been used in both metal and mineral concentrate processing at claimed solids throughputs up to 400 te h^{-1}.

10.3.5 Modelling of Filtration

Cake formation on a vacuum filter takes place at a constant level of vacuum. Thus, process modelling is usually based on a version of the quadratic Eq. (10.14), which is derived from Darcy's Law Eq. (10.1), and the fixed values of cake resistance and solids concentration given by Eqs. (10.10), (10.11), and (10.12). Noting that the configuration of vacuum filters limits the cake formation to a fixed, and generally limited, time period, the volume of filtrate is given by:

$$V_f = \frac{A_f}{c}\left(-\frac{R}{\alpha_{av}} + \sqrt{\left(\frac{R}{\alpha_{av}}\right)^2 + \frac{2ct_f\Delta p_f}{\mu_l\alpha_{av}}}\right) \tag{10.32}$$

and the cake thickness by:

$$L_f = \frac{1}{\rho_s C_{av}}\left(-\frac{R}{\alpha_{av}} + \sqrt{\left(\frac{R}{\alpha_{av}}\right)^2 + \frac{2ct_f\Delta p_f}{\mu_l\alpha_{av}}}\right) \tag{10.33}$$

In these equations the area of filter medium (A_f) devoted to filtration is given by:

$$A_f = \begin{cases} z_f h_B & \text{horizontal belt} \\ \varphi_f \pi D h_D & \text{drum} \\ 0.25\varphi_f\pi\left(d_o^2 - d_i^2\right) & \text{table} \\ \varphi_f n_p A_p & \text{tilting pan} \\ 0.5\varphi_f n_d \pi\left(d_o^2 - d_i^2\right) & \text{disc} \end{cases} \tag{10.34}$$

where z_f is the belt length devoted to filtration, h_B the belt width, φ_f the fraction of filter medium area devoted to filtration, D the drum diameter, h_D the drum width, n_p

the number of pans, A_p the filter medium area in a single pan, n_d the number of discs, and d_i and d_o are respectively the inner and outer diameters of the filter medium on a table or disc. With rotary type filters, such as the rotary drum and tilting pan, the duration of the filtration phase is given by $t_f = 2\pi\varphi_f/\omega$ (ω is the angular velocity), whereas for the horizontal belt $t_f = z_f/v_B$ (v_B is the linear belt velocity).

10.3.6 Modelling of Cake Washing and Gas Deliquoring

Cake washing and gas deliquoring on vacuum filters are modelled in a similar manner to that described for pressure filters in Sects. 10.2.8 and 10.2.9, respectively. However, vacuum filters such as the rotary drum do not always adequately constrain the wash liquor flow in terms of directing it through the filter cake and washing is generally not as efficient as it could be. In such cases it is necessary to correct the calculated dispersion number according to an empirical correlation such as:

$$(D_n)_{corr} = 3.22\log(D_n) + 0.395 \tag{10.35}$$

The value of $(D_n)_{corr}$ is used in place of D_n in design charts.

10.4 Filtering Centrifuges

Filtering centrifuges use centrifugal forces to perform batch and continuous cake filtration on either cylindrical, or conical, semi-permeable surfaces. Displacement washing operations can be accommodated by most centrifuges in addition to efficient cake deliquoring. Several machines are capable of operating in both vertical and horizontal orientations whilst some rely on the favourable sliding and conveying properties of the formed cake for successful operation. The typical characteristics of filtering centrifuges are shown in Table 10.3 and exemplar capital equipment costs can be estimated using equations adapted from Leung [15].

$$\begin{array}{ll}\text{Inverting bag}\,(D = 0.3 - 1\,\text{m}): & 2.3\times10^5\exp(1.1D)\ \$\\ \text{Peeler}\,(D = 0.6 - 1.2\,\text{m}): & 1.8\times10^5\exp(0.86D)\,\$\\ \text{Pusher}\,(D = 0.3 - 1\,\text{m}): & 2.6\times10^5 D + 52200\,\$\end{array} \tag{10.36}$$

10.4.1 Basket

Typical uses: Deliquoring of suspensions with reasonable drainage characteristics.

Table 10.3 Typical characteristics of some industrial filtering centrifuges

	Basket	Pusher	Inverting bag	Slip discharge	Vibratory, oscillatory	Worm screen
Centrifugal force or g-factor (g)	700–2200[c]	800–1700	700–1500	Up to 2900[a]	30–150	500–2600
	200–1200[d]					
Operation	Batch	Continuous	Batch	Continuous	Continuous	Continuous
Max. throughput (te h^{-1})	12–15[e]	5–80[e]	10–450 kg[e]	<150	350	<150
Cake condition	Dry → pasty, granular	Dry, granular	Dry, granular	Dry, granular	Dry, granular	Dry, granular
Bowl diameter (m)	0.4–1.8	0.3–1.3	0.3–1.3	0.5–1.6	0.7–1.5	0.2–0.9
Solids dryness index	9 C	9 C	9 C	7 C	8 C	9 C
Washing index	6	7–8	6	5	5	5
Liquid clarity index	5	4	5	4	4	4
Particle breakage index	5–6	4	6	4	3	4
Particle size in feed (μm)	2–1000	40–7000	2–1000	80–10000	100–10000	60–5000
Feed conc. (%w/w)	4–30	10–40	5–30	10–40	10–40	10–40
Motor drive power (kW)	5–90	3–200[b]	3–90	6–70	8–75	2–90

[a]At the lip (i.e. maximum diameter) of the cone
[b]Including pusher drive
[c]Peeler
[d]Pendulum
[e]Of solids

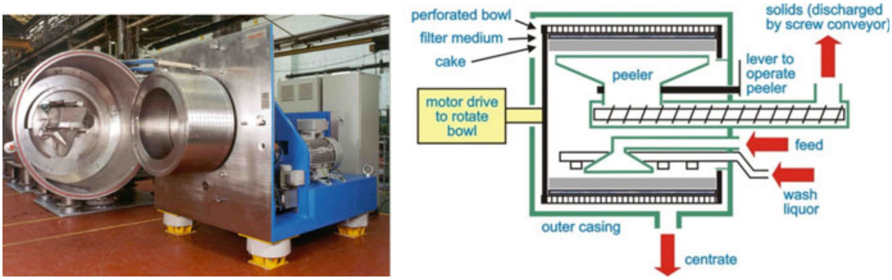


Fig. 10.22 Photograph of a horizontal peeler centrifuge with the front end casing open to facilitate safe operator inspection of the interior (*left*, Courtesy of Thomas Broadbent & Sons) and schematic showing the principal features (Courtesy of Filtration Solutions, UK). In a centrifuge the filtrate is known as centrate

These centrifuges are essentially batch operated and comprise a vertically or horizontally mounted basket with one closed end and one partially open end. The basket, which is perforated and covered by a combination of metal screen(s) and filter cloth, is rotated to give solids throughputs up to 15 te h^{-1}. The induced centrifugal forces allow centrate (filtrate) to pass through the cloth/screen whilst particles accumulate in the form of a filter cake. The cake may subsequently be washed by sprays and/or allowed to deliquor prior to discharge. Variants of the basket filtering centrifuge differ primarily in the process limitations imposed by the axis of rotation.

Horizontal axis: The fully automated horizontal axis basket, or peeler, centrifuge operates for most of its cycle at constant rotational speed to give a g-factor up to 2200 g (see Fig. 10.22). The operating cycle is generally shorter than for vertical axis machines and, with less time lost for acceleration and deceleration, higher throughputs can be achieved. Solids are discharged at moderate speed at the end of the cycle by a rigidly constructed peeler, sometimes with the aid of a compressed gas jet or reciprocating knife. The relatively high speed discharge can induce glazing of the cake heel and hence reduced centrate flow rates in subsequent cycles. Horizontal axis centrifuges tend to be more expensive than equivalent capacity vertical axis machines.

Vertical axis: The vertical axis basket centrifuge, which is also known as the three-column or pendulum centrifuge, allows the feed suspension to be introduced when the basket is either stationary or rotating at a moderate speed. The rotational speed is often varied through a cycle with cake washing and deliquoring being performed at high speed (~1500 rpm) and cake discharge at a lower speed (~60 rpm). On bottom driven machines the basket is usually lifted out manually to allow for cake discharge. Generally more expensive top driven machines are employed for heavier duties and faster filtering feeds. These units are discharged automatically by plough or with the assistance of a gas jet and/or compressed gas blow-back when a residual heel of cake is unacceptable. For cakes that are inherently hard a partial length plough that moves up and down the axis of the basket can be used. Due to uneven cake formation, washing performance can be variable.

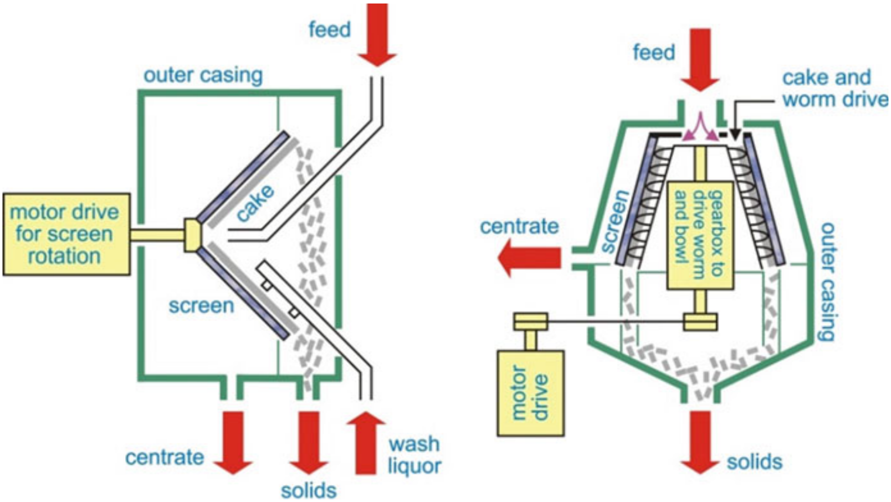


Fig. 10.23 Schematics of two variants of cone screen centrifuge, slip discharge/wide angle cone (*left*) and single stage worm screen (*right*) (Courtesy of Filtration Solutions, UK)

10.4.2 Cone Screen

Typical uses: Continuous deliquoring of suspensions containing relatively free filtering solids.

A cone screen centrifuge comprises of a conical perforated metal screen across which wet solids slide after filtration from relatively high concentration suspension (Fig. 10.23). During their passage, the solids, in the form of a cake, can be washed by sprays and/or deliquored prior to discharge at the wider end of the cone. The four variants of cone screen centrifuge differ primarily in the manner in which the solids are caused to translate along the screen:

Slip discharge/wide angle cone: In these vertical or horizontal axis machines the cake is caused to move by providing a cone with a half-vertex angle in excess of the angle of friction between the cake and the screen. The cone angle is critical for good operation and is typically in the range 25–35°, though the lubrication provided by the liquid in the cake can greatly assist the sliding operation and particularly toward the start of the translation process. Whilst good deliquoring is generally achieved with centrifugal forces up to 2900 g, the rapid transit of solids through the centrifuge means there is a limited time available for washing on the angled surfaces of the cone. Slip discharge centrifuges are best suited to the processing of fairly coarse, fast filtering, granular solids.

Vibratory/oscillatory: These centrifuges work on a similar principle to the slip discharge centrifuge, however, the addition of an eccentric vibratory drive facilitates use of a cone angle lower than the angle of friction between the cake and the screen. Cone angles of 13–18° are common and vibrations in the region of 1700 min^{-1} induce partial fluidisation of the cake which enhance its translation across the screen to give throughputs as high as 350 te h^{-1}. As relatively low centrifugal forces are generated (typically below 120 g), deliquoring and centrate clarity can sometimes be poor and this may in turn lead to lower quality particle and liquid products. Both horizontal and vertical axis machines are available and these are best suited to the processing of relatively coarse, fast filtering solids.

Tumbling: A gyratory motion of the screen bowl about a vertical axis causes the inclination of the cone walls to alter about the angle of friction between the cake and the screen. The tumbling of the cone induces intermittent cake movements and throughputs up to 120 te h^{-1}. Despite relatively modest centrifugal forces, deliquoring can be very good. However, tumbling centrifuges are usually only employed when washing is not required and coarser, faster filtering solids are present in the feed. Variations of cone angle and the speeds of rotation and gyration dictate the compromise between throughput and final moisture content of the cake.

Worm screen: Also known as the conveyor discharge or screen scroll centrifuge, the worm screen centrifuge causes solids to move along the cone via an internal screw conveyor. The conveyor rotates at a differential speed to the cone screen and centrifugal forces below 2600 g facilitate reasonable throughputs. The presence of the conveyor can sometimes lead to particle breakage and abrasion problems, as well as relatively poor washing. There is a compromise between throughput and final cake moisture and this is dictated by the conveyor speed. Typical cake residence times on the screen lie in the range 4–15 s. Worm screen centrifuges are available in either vertical or horizontal orientation and are most frequently used for the processing of fibrous solids/particles. More sophisticated versions employ cones with up to four stages that allow cake formation, two periods of displacement washing (with the potential for segregation of the wash liquors) and final deliquoring to take place.

10.4.3 Pusher

Typical uses: Deliquoring of relatively coarse particulate suspensions where good cake dryness at discharge is required.

The horizontal axis pusher centrifuge is probably the most commonly used design employing a continuous feed of suspension. A single-stage machine comprises of a rotating cylindrical screen bowl into which suspension is introduced and filtered to form a cake. A plate positioned at the closed end of the bowl reciprocates with a 20–80 mm stroke at up to 100 strokes min^{-1} to push the forming cake toward the open end of the bowl and discharge. During transition across the screen, the cake may be washed by sprays and efficiently deliquored as a result of the 500–1700 g centrifugal forces generated and 6–20 s cake residence times. In order for

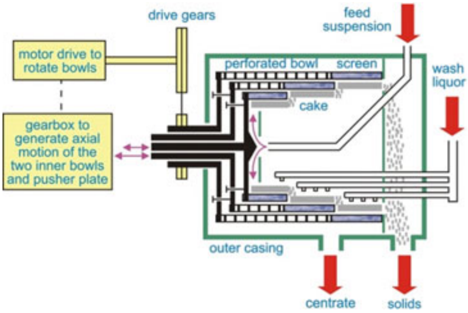


Fig. 10.24 Schematic of a three-stage pusher centrifuge (*left*, Courtesy of Filtration Solutions, UK) and photograph of a pusher centrifuge (*right*, Courtesy of Andritz)

the pusher centrifuge to work correctly the formed cake must have sufficient strength to withstand buckling and slide efficiently across the screen.

For weaker, more friable cakes, or cakes containing finer particles of greater frictional resistance, a multi-stage unit may be needed (see Fig. 10.24). These more expensive machines include a sequence of up to four, relatively short, concentric bowls with progressively increasing diameter. The solids are pushed more readily along the shorter bowls by the reciprocating piston and an ability to separate wash liquors allows for improved washing. The transfer between the bowls tends to lead to cake break-up which can enhance the deliquoring process.

As the open end of a bowl in a pusher centrifuge does not have a retaining lip there is an 'overflow limit' which generally limits solids throughputs to 80 te h^{-1} for single-stage machines and ~45 te h^{-1} for multi-stage machines; some manufacturers claim maximum throughputs of 100 te h^{-1}. The operation of both single and multi-stage centrifuges with solids below 100 μm can be problematical due to blockage of the filtering screen.

10.4.4 Baffle Centrifuge

Typical uses: Deliquoring coarse particulate (e.g. polymer pellet) suspensions where good cake dryness is required.

The family of baffle centrifuges are representative of specialist continuous filtering centrifuges. Both the baffle ring and screen baffle centrifuge can achieve very low residual moistures in granular type materials by causing particles to bounce against (baffle like) obstructions inside the rotating bowl to release additional surface and occluded liquids. Although baffle centrifuges are relatively expensive and restricted to operations with certain types of materials, their use can prove advantageous when other alternatives are unsuitable.

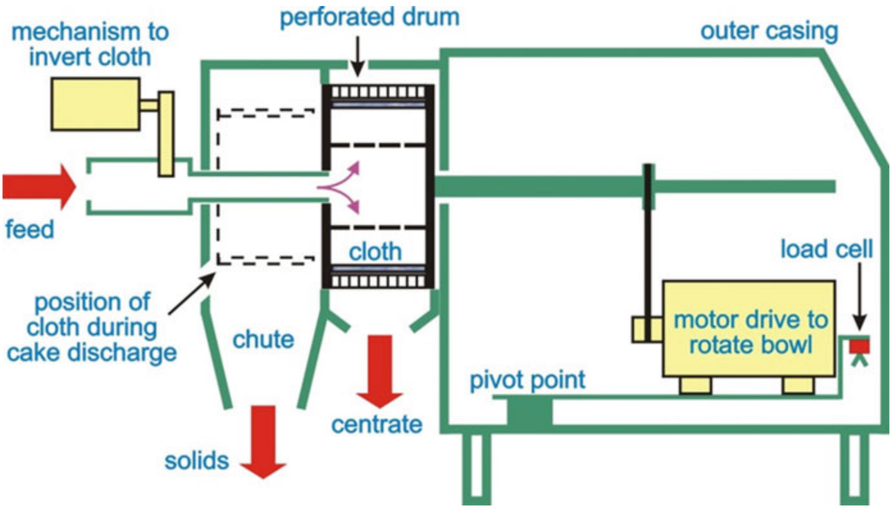


Fig. 10.25 Schematic of the inverting bag centrifuge which can meet clean room requirements, be totally enclosed and gas tight to maintain an inert atmosphere (Courtesy of Filtration Solutions, UK)

10.4.5 Inverting Bag

Typical uses: Semi-continuous deliquoring of suspensions where complete cake discharge and high purity need to be maintained.

The inverting centrifuge (Fig. 10.25) operates semi-continuously via automatic control and features a horizontally mounted, cylindrical drum between 0.3 and 1.3 m diameter which restricts filtration area to ~2 m^2. Suspension is introduced to the drum through gravity by means of a rigid filling pipe that projects through the solids discharge chute. The amount of material delivered is continuously monitored with a non-contact, load cell system resembling a beam type balance.

A typical cycle involves initial cake formation, intermediate deliquoring, rinsing/washing followed by final deliquoring. For discharge, a unique mechanism allows the end of the drum to open through a translational movement and the cake solids are removed completely under rotation as the filter bag inverts through the discharge chute. In this way cloth blinding is avoided and the entire cycle can be performed under high purity conditions. To improve deliquoring the pressure in the filling pipe and internal chamber of the drum can be raised, although any process advantages are offset by the increased mechanical complexity.

10.4.6 Modelling of Filtration

Fundamental theory for fluid flow through porous media has not been used generally for design calculations in filtering centrifuges, instead reliance is often placed

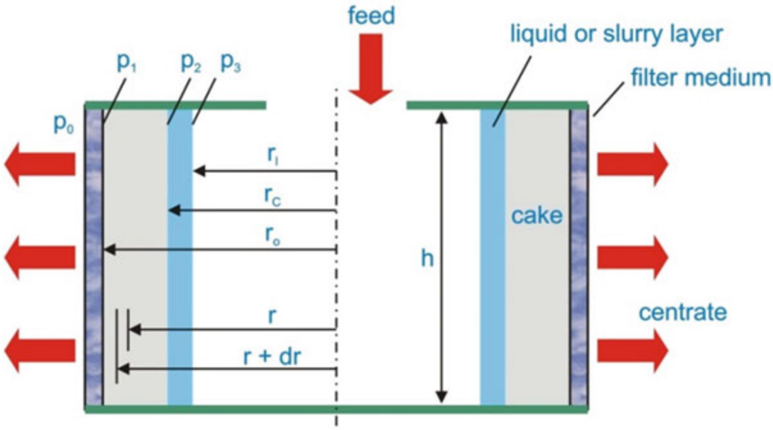


Fig. 10.26 Schematic cross-section through a centrifuge basket showing the filter medium, cake and liquid layers, and defining the notation used for the analysis of centrifugal filtration (Courtesy of Filtration Solutions, UK)

on practical experience and empirical approaches. Modelling difficulties arise from, for instance, the sometimes short processing times, uneven cake formation, variable feed properties, movement of the filter cake through the centrifuge and premature settling of particles. However, the general principles involved can be illustrated with a simple geometry like that associated with a basket centrifuge (see Fig. 10.26); the axis of rotation may be either horizontal or vertical.

By considering a differential cylindrical element of filter cake between radii r and $r + dr$, and assuming that Darcy's Law is valid for a basket radius r_0, height h and angular velocity ω, the volumetric flow rate of centrate can be expressed as:

$$q = \frac{\frac{\rho_l \omega^2}{2}\left(r_0^2 - r_l^2\right)}{\frac{\mu \rho_s \alpha_{av} C_{av}}{2\pi h} \ln\left(\frac{r_0}{r_c}\right) + \frac{\mu R}{2\pi r_0 h}} = \frac{\frac{\rho_l \omega^2}{2}\left(r_0^2 - r_l^2\right)}{\alpha_{av} \mu M_s \frac{\ln(r_0/r_c)}{2\pi h\left(\pi h\left(r_0^2 - r_c^2\right)\right)} + \frac{\mu R}{2\pi r_0 h}} \tag{10.37}$$

where r_0 is the inner radius of the basket, r_l the radius of the liquid layer, r_c the time variant radius of the cake and M_s the mass of solids in the cake. Values for r_c can be obtained from a volumetric balance that equates the solids in the centrifuge feed to the increase of cake volume in the basket

$$r_c = \left(r_0^2 - \frac{V_s}{\pi h C_{av}(1 - V_s)} qt\right)^{0.5} \tag{10.38}$$

where V_s is the solids volume fraction in the feed suspension. Use of Eqs. (10.37) and (10.38) allows the amount of liquid removed and solids accumulation in the bowl to be determined as a function of time. Similar theoretical analyses are available for pusher centrifuges [10, 35].

10.4.7 Modelling of Cake Washing and Deliquoring

Cake washing and deliquoring on filtering centrifuges is modelled in a similar manner to that described in Sects. 10.2.8 and 10.2.9. For deliquoring the principal differences relate to a design chart where the plot of S_R vs. θp^* involves a family of curves with each curve corresponding to a specified value of r_c/r_o, and θ and p^* are defined slightly differently to account for the centrifugal driving force. There is also no corresponding air/gas flow design chart, again because of the different manner in which the process is driven.

10.5 Centrifugal Sedimenters

Sedimenting centrifuges, which typify this category of equipment, employ centrifugal forces to accelerate the settling of particles (or liquid droplets) within rotating, solid walled equipment. With no filtration occurring, a density difference must exist such that the denser material in the feed preferentially settles to the wall where it is removed in a concentrated form. The clarified phase is also discharged, often at the opposite end to the feed position. Both batch and continuous types are available and typical duties range from the clarification of dilute suspensions to the thickening of fast settling slurries. When fast acting flocculants are used to aid separation, the shear resistance of flocs is an important factor as centrifugal forces can be high depending on the type of machine. Some typical characteristics of sedimenting centrifuges are shown in Table 10.4 and exemplar capital equipment costs for variants made from stainless steel can be estimated using equations adapted from Leung [15]:

$$\begin{array}{ll} \text{Tubular bowl}(D = 4 - 15\,\text{cm}): & < \$140,000 \\ \text{Disk}(D = 0.15 - 0.9\,\text{m}): & \$60,000 \; - \; \$350,000 \\ \text{Decanter}\,(D = 0.3 - 1.3\,\text{m}): & -1.2 \times 10^5 D^2 + 5.6 \times 10^5 D - 36000\ \$ \end{array} \tag{10.39}$$

10.5.1 Tubular Bowl Centrifuge

Typical uses: Batch clarification and (occasionally) particle classification.

The tubular bowl centrifuge shown in Fig. 10.27 is usually regarded as the most efficient of the industrial sedimenting centrifuges with separating forces in the range 14,000–65,000 g. The circular bowl, which typically has a diameter between 4 and 15 cm and an aspect ratio between 4:1 and 8:1, is vertically mounted and rotates at up to 50,000 rpm. The relatively dilute feed suspension is introduced at the bottom of the bowl via a distributor and centrate overflows from the top. The normally denser solids accumulate at the wall of the bowl from where they are

Table 10.4 Typical characteristics of some industrial sedimenting centrifuges

	Tubular bowl	Basket	Disk stack	Scroll decanter
Centrifugal force or g-factor (g)	14,000–65,000	Up to 1600	Up to 14,000	2000–6000
Rotational speed (rpm)	50,000 (max)	450–3500	3000–10,000	1600–6000
Mode of operation	Batch	Batch	Batch or continuous	Continuous
Throughput (m^3 h^{-1})	<5	6–10	1–100	<150
Sediment condition	Pasty, firm	Firm	Pasty, flowable → firm	Pasty, granular
Bowl diameter (m)	0.04–0.15	0.3–1.8	0.15–0.9	0.15–1.4
Bowl height or length (m)	0.2–0.75	0.2–0.8	Depends on no. disks	0.35–4.5
Solids dryness index	3 S	2 S	2 S	4 C
Washing index	–	–	–	3
Liquid clarity index	6	5	–	4
Particle breakage index	5	5	6	3
Particle size in feed (μm)	0.1–100	0.1–100	0.1–100	1–5000
Feed conc. (%w/w)	<5	<5	0.05–10[a]	4–40
Power requirement (kW)	<4	10–40	1–90	4–450

[a]Depends on variant

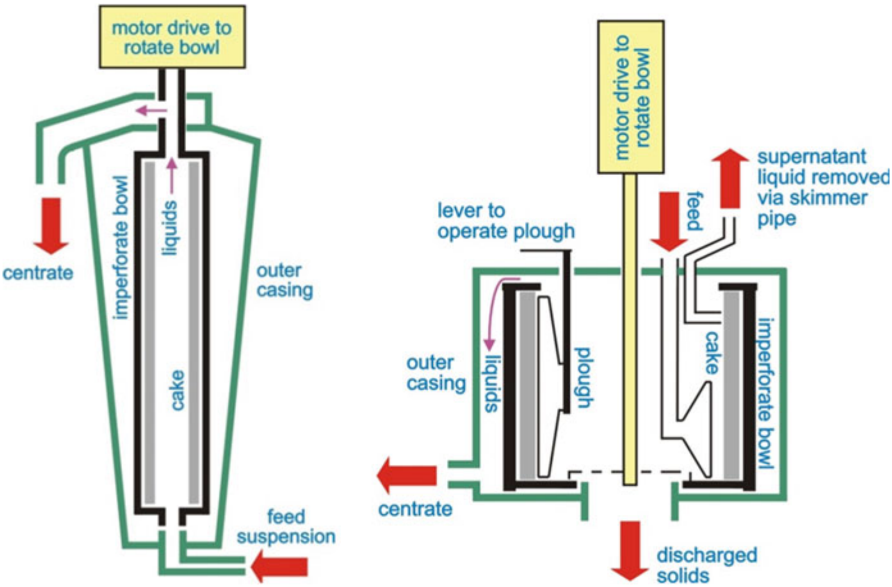


Fig. 10.27 Operating principle and general form of the tubular bowl centrifuge (*left*) and basket sedimenting centrifuge (*right*) (Courtesy of Filtration Solutions, UK)

manually discharged at the end of a batch cycle; the discharge process is sometimes aided by the inclusion of paper liners. Due to the narrowness of the bowl the efficiency of separation is significantly influenced by solids accumulation at the wall and throughputs are restricted to $<5\ m^3\ h^{-1}$ with dry solid yields of up to 4 kg per batch.

10.5.2 Basket Centrifuge

Typical uses: Recovery and concentration of solid sludges.

The basket bowl centrifuge (Fig. 10.27) works on similar principles to the tubular bowl centrifuge. Semi-continuous operation is achieved, however, by using a bowl diameter between 25 and 180 cm and a much lower aspect ratio of ~0.6:1. The solids accumulate on the wall of the imperforate bowl and liquid overflows via a weir at the top. Cake discharge is performed manually on small machines with the bowl stationary. With larger machines the supernatant liquid remaining in the bowl is siphoned-off using a skimmer pipe and the solids on the wall are removed automatically with a plough, sometimes at a reduced bowl speed.

Basket bowl centrifuges operate at rotational speeds of 450–3500 rpm to allow throughputs in the region 6–10 $m^3\ h^{-1}$. The use of relatively low g-forces in the single bowl centrifuge (<1600 g) has led to the development of multi-bowl basket centrifuges which comprise a series of concentric bowls mounted on a common vertical shaft to facilitate higher g-factors of *ca.* 5000–9000 g. During rotation the feed moves progressively from the inner to the outer bowl with ever finer particles being removed each time. Multi-bowl basket centrifuges offer greater efficiency for a given speed of rotation with the disadvantages of increased capital and operating costs.

10.5.3 Disc Stack Centrifuge

Typical uses: Clarification and thickening to produce a solids sludge, useful for bioprocessing.

The disc stack centrifuge is a versatile device which may be used for separating mixtures in continuous, semi-continuous and batch configurations (see Fig. 10.28). All except some batch operated machines are able to handle toxic, flammable and volatile feeds at typical throughputs of up to 100 $m^3\ h^{-1}$ (up to 200 $m^3\ h^{-1}$ is claimed in some cases). Particle-liquid and liquid-liquid mixtures can be separated, and with more sophisticated units a three phase separation (two liquid and one solid) is achievable. In all cases a sufficient density difference must exist between the phases present in the feed.

Although variants exist, the generic type is characterised by an imperforate bowl surrounding an inverted stack of 30–200 thin conical discs separated by 0.3–3 mm

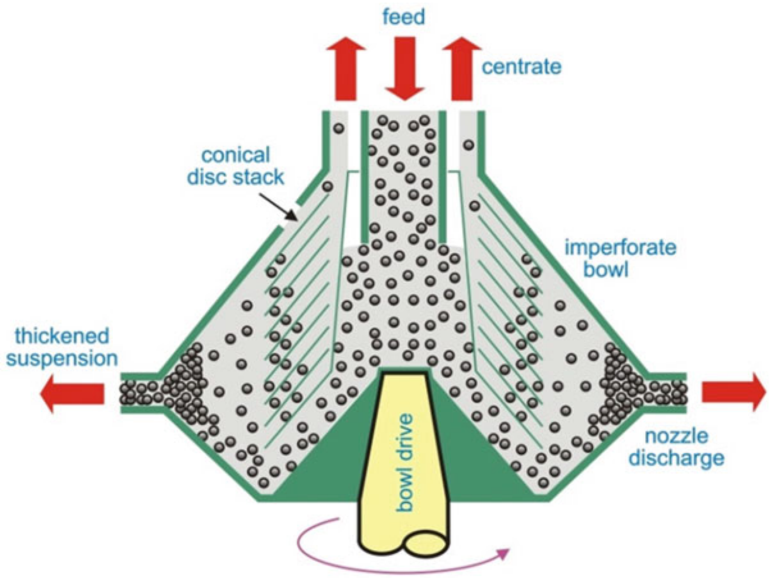


Fig. 10.28 Schematic of a nozzle discharge disc stack sedimenting centrifuge (Courtesy of Filtration Solutions, UK)

spacers. The disc spacing is dependent on the viscosity and solids content in the feed and needs to be fixed accordingly, lower viscosities and solids concentrations favour spacings below 1 mm. As the discs are spun on a common vertical axis the process suspension, which is fed centrally from the top and assumed here to comprise of particles dispersed in a single liquid, travels through the annular spaces between the discs. Centrifugal forces up to 14,000 g cause the particles to accumulate on the underside of the discs from where they slide down toward the outer periphery of the centrifuge bowl to form the sediment.

In batch units the sediment remains in the bowl until the so-termed ‘solids handling capacity’ of the centrifuge is reached. At this point rotation stops and the basket containing the sediment is manually replaced or a discharge valve on the periphery of the bowl is manually operated to facilitate sediment removal. In continuous units the sediment, which must be flowable, is automatically discharged, sometimes intermittently, through nozzles positioned on the outer periphery of the bowl; a typical centrifuge has between 12 and 24 nozzles of 0.5–3 mm diameter. For sediments that exhibit poor flow characteristics, the ‘self-ejecting’ design variant allows the bottom portion of the centrifuge to automatically separate at periodic intervals and discharge the accumulated material.

Whilst disc centrifuges are able to accept a wide range of feeds, they are mechanically complex and often expensive. Moreover, the close stacking of conical discs means that mechanical cleaning is a challenge, and periodic chemical cleaning (with its subsequent disposal issues) may be a necessity.

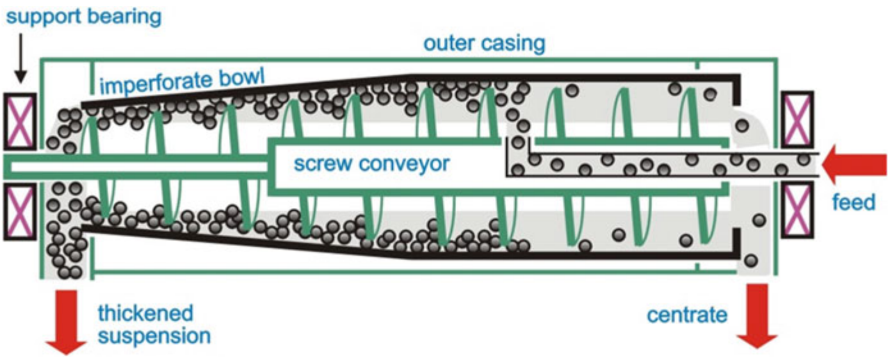


Fig. 10.29 Schematic of a horizontal axis scroll decanter centrifuge. The motor drive for the bowl and the gearbox required to produce the differential rotation speed between the bowl and screw conveyor are omitted for clarity (Courtesy of Filtration Solutions, UK)

10.5.4 Scroll Decanter Centrifuge

Typical uses: Relatively coarse deliquoring and clarification of suspensions.

The scroll decanter is a horizontally or vertically mounted centrifuge which is best suited to the processing of free draining solids from higher concentration feeds (see Figs. 10.3 and 10.29). In extreme cases throughputs of solids can be as large as 100 te h^{-1} whilst liquid throughputs are normally less than 60 m^3 h^{-1}. In a typical unit a cylindrical bowl with a tapered, conical end (the beach) is caused to rotate at speeds between 1600 and 6000 rpm. Inside the bowl a helical screw rotates at a differential speed of up to $\pm$100 rpm. The feed enters through the central axis of the centrifuge where inertial forces of less than 6000 g cause the denser solids to sediment towards the imperforate wall of the bowl. The sediment is conveyed co-currently along the walls of the bowl by the helical screw and moves through the narrower conical end of the centrifuge to discharge. The liquid phase, which may not always be clear due to the presence of fines, leaves the centrifuge via a weir or ports at the broader end of the bowl.

In the alternative screen bowl design the conical section is shortened and a supplementary cylindrical section which is perforated is attached in an effort to promote enhanced deliquoring. Thus, both sedimentation and filtration can be combined within a single unit. Other design variants rely on the addition of baffles, helical discs, vanes, conical disc stacks or fins, all of which alter the flow and/or residence time distributions within the centrifuge (e.g. Fig. 10.3).

When finer particles are being processed the flow properties of the thickened solids can be poor and this leads to high helical screw torques and associated mechanical difficulties. Wear problems on the screw are also caused by more abrasive particles. Scroll decanters can be adapted for use with toxic, flammable and volatile substances and there is some scope to perform (relatively poor) washing.

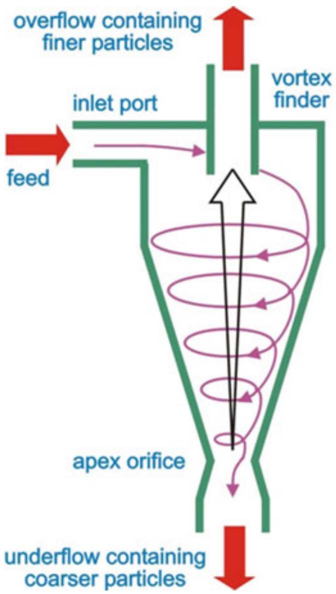


Fig. 10.30 Cross-section through a reverse flow hydrocyclone showing the typical flow patterns (*left*, Courtesy of Filtration Solutions, UK). The *inset* photograph (*right*, Courtesy of Axsia-Mozley) shows a bank of six cyclones connected to a common feed manifold system

10.5.5 Hydrocyclone

Typical uses: Suspension thickening, clarification and particle classification.

Another widely used form of centrifugal sedimenter is the reverse flow hydrocyclone (Fig. 10.30). Either concentration or classification of solids can be performed and the device is particularly attractive because it is relatively cheap, compact, versatile and has no moving parts.

The basic unit comprises an inverted conical bottom section attached to a cylinder containing a tangential inlet port. Feed is pumped through this port at a mean velocity between 10 and 30 m s^{-1} whence geometry induced motion causes the (usually denser) suspended particles to experience centrifugal forces of 70–18,000 g. The combination of these forces and a swirling motion causes the larger particles to exit as a suspension in the underflow stream at the bottom of the hydrocyclone and the finer fractions to leave through the cylindrical vortex finder at the top. With short residence times the particles and liquid move at relatively high speeds and abrasion/particle breakage can sometimes be a problem which necessitates the use of hard, and replaceable, internal linings.

Many standard sizes of hydrocyclone are available with cylinder diameters of 1–30 cm and cone angles of 25–50°. The particle cut size, which is the size equally likely to find its way into the underflow or overflow, is limited to about 5 μm and dependent on several factors including the size and geometry of the hydrocyclone, the inlet flow rate and the pressure drop across the unit. Separation is often more effective (in terms of a lower cut size) with smaller diameter hydrocyclones as higher tangential velocities can be achieved; in such cases overall throughput can be maintained by using several units in parallel (see also Chap. 12).

10.5.6 Modelling

Noting that the modelling of sedimenting centrifuges is inherently difficult and that all published models have their limitations, the most widely used and accessible model for characterising separations is Sigma (Σ) theory which is originally attributable to Ambler [1]. Sigma theory calculates the surface area of a static settling tank that gives the same theoretical performance as the centrifuge. Sigma is defined as

$$\frac{Q}{nu_t} = \sum \tag{10.40}$$

where Q is the volumetric feed flow rate, u_t the Stokes settling velocity given by Eq. (10.6) and n is a constant which is frequently taken as 2. The terms on the left hand side of Eq. (10.40) are solely functions of the process material whilst Σ is related solely to the characteristics of the centrifuge in accordance with the sample formulae given in Table 10.5; alternative formulae and more specific ranges for Σ are given in, for example, Leung [15] and Records and Sutherland [19]. Sigma theory can be used for individual calculations, scaling from one geometrically similar centrifuge to another and assessing relative performance.

Specific aspects for the design of hydrocyclones are provided by Svarovsky [25], but see also Hoffmann and Stein [12].

Table 10.5 Sample formulae and guideline values for Σ

Centrifuge type	Equation for Σ	Σ values (m^2)
Disk stack	$\dfrac{2\pi\omega^2(N-1)\left(R_2^3-R_1^3\right)}{3g\tan\theta}$	<150,000
Tubular bowl	$\dfrac{\pi\omega^2 L}{g}\dfrac{r_o^2-r_l^2}{\ln\left(\dfrac{2r_o^2}{r_o^2-r_l^2}\right)} \approx \dfrac{\pi\omega^2 L}{2g}\left(3r_o^2+r_l^2\right)$	<5000
Decanter	$\dfrac{\pi\omega^2}{2g}\left(L_1\left(3r_o^2+r_l^2\right)+\dfrac{L_2}{2}\left(r_0^2+3r_o r_l+4r_l^2\right)\right)$	<25,000

In the table N is the number of disks, R_1 the inner disk radius, R_2 the outer disk radius, θ the conical half angle, L_1 the length of the cylindrical section and L_2 the length of the conical section

10.6 Gravity Thickeners

A class of solid-walled separator where gravitational forces are used to raise the concentration of a suspension through sedimentation to produce a thickened sludge underflow and a clear liquid. The rate of sedimentation should be as high as reasonably possible to both increase throughput and reduce floor plan area. Sedimentation rates are often artificially increased by the addition of (relatively expensive) coagulants or flocculants. The cross-sectional area of a thickener controls the time available for sedimentation and is important in determining clarification capacity. The physical depth of a separator controls sludge thickening time and is an important parameter in determining thickening capacity. Thickeners and clarifiers can be designed to operate in either batch or continuous modes, although most commercial operations utilise the latter (see Table 10.6). An estimate for the capital cost of gravity thickeners is based on the underflow volumetric flow rate (Q_s) and adapted from ASCE and AWWA [2]:

$$\text{Gravity thickener}\left(Q_s = 0.002 - 0.5\,\text{m}^3\,\text{h}^{-1}\right) : -3.1\times10^5 Q_s^2 + 3.6\times10^5 Q_s + 39000\,\$ \tag{10.41}$$

Table 10.6 Typical characteristics of continuous thickeners and clarifiers

	Circular basin	Circular high capacity	Settling tank or lagoon	Deep cone	Lamella separator	Clarifier[a]
Diameter (m)	3–200	4–18	As required	3–25	As noted[b]	4–12[d]
Diameter/depth ratio	2:1–>10:1	3:1	1:1–>10:1	1:1–1:2	N/a	Varies
Solids dryness index	1 S	1 S	1 S	1 S	1 S	1 S
Washing index	2	–	–	–	–	–
Liquid clarity index	5	5	5	5	5	6
Particle breakage index	9	9	9	9	8	9
Particle size in feed (μm)	0.1–500	0.1–300	0.1–500	0.1–500	1–150	1–50
Feed conc. (% w/w)	<20	<15	<20	<20	<15	<15
Rake motor torque (Nm)	3000–8×10^6	n/a	–	$<14\times10^6$	–	n/a
Power requirement (kW)	1–35	n/a	–	As noted[c]	–	n/a

[a]Primarily used to recover clear liquor from dilute suspension, but operate on a similar principle
[b]Typically 1/5–1/10th the footprint of a circular basin thickener with an equivalent settling area
[c]Dependent on the rheological properties of the underflow, but tends to be higher for a given diameter compared to a conventional thickener
[d]Rectangular versions are also common

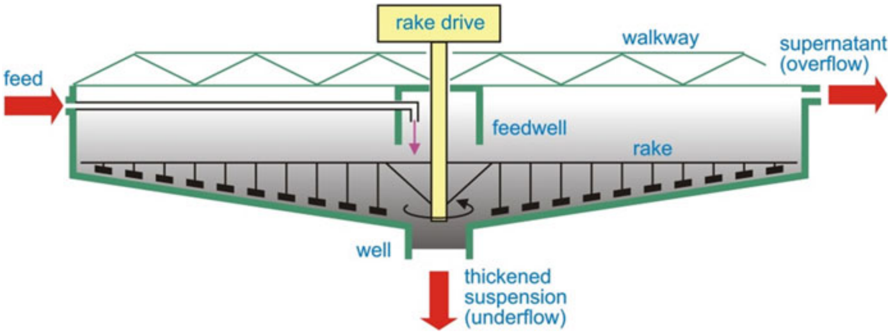


Fig. 10.31 Schematic of a circular basin thickener showing rakes, drive head and walkway (Courtesy of Filtration Solutions, UK)

10.6.1 Circular Basin Thickener

Typical uses: Larger scale thickening and deliquoring of solids from relatively dilute suspension.

The circular thickener comprises a relatively shallow, open top cylindrical tank with either a flat bottom or a bottom shaped in the form of an inverted cone (see Figs. 10.3 and 10.31). The feed mixture is gently and continuously introduced to the feedwell in which exists a pool of settling suspension along with any additional coagulant or flocculant. With settling and thickening proceeding, clear liquid is removed via an annular weir at the top of the unit and solids sludge (sediment) is removed from a 'well' at the bottom. Slowly rotating arms (or rakes) mounted on a central drive head aid the thickening process by directing the sediment towards the well for subsequent discharge, and by creating channels for the release of further liquid from the sediment. The construction and form of the rake are important design parameters as is the rating of the motor in the central drive head which must be capable of moving the rake through the sediment, both during normal operation and during start-up of the rake after a stoppage. Lifting devices are often employed to position the rakes at the optimum height within the basin and thus prevent damage due to excessive torque requirements.

Tanks with a diameter smaller than 25 m are usually formed from steel and have flat bottoms with rake arms at an angle less than 10°. Larger tanks between 25 and 200 m diameter are made from a combination of concrete and steel and employ rakes designed to match the angle of the conical bottom. Circular thickeners are frequently constructed to large scales and can be used to raise suspension concentration prior to another solid/liquid separation process.

Continuous discharge of solids from a gravity settling tank can be achieved without mechanical aid if the tank is shaped so that the sludge flows naturally towards the discharge port. This requires relatively steep sided conical vessels; the angle of the cone is generally 40–60° and thus the diameter of a settling tank is invariably rather less than a thickener. A diaphragm baffle is located near the base to prevent arching of solids across the outlet port.

Particularly large volumes of slowly settling slurries (which are also of low value) may be thickened in lagoons if land is not at a premium. Lagoons usually need to be lined to prevent seepage and are rarely an environmentally friendly option which limits their use.

10.6.2 High Capacity Thickeners

Typical uses: Separation of rapidly settling solids where available space is at a premium.

High capacity thickeners work on a broadly similar principle to conventional thickeners. However, by the correct use of high molecular weight, fast acting flocculants, large flocs that sediment very quickly can be generated to provide thickeners with high solids handling capacities and relatively small floor plan areas. A raised underflow concentration tends to be promoted by an increased thickener depth. Although high capacity thickeners have found many uses, particularly when a significant amount of fines are present, they do not represent a replacement for conventional thickeners as flocculant usage is notably higher and thus more costly.

10.6.2.1 Circular

The circular high capacity thickener is similar in general form to a conventional circular basin thickener (see Fig. 10.31), however, cylinder diameters are generally limited to between 4 and 18 m. Units are constructed from steel and include a cylindrical top portion, an inverted cone bottom section and an angled rake system mounted on a central drive head. Suspension throughputs are limited to about 4000 $m^3\ h^{-1}$.

10.6.2.2 Deep Cone

The deep cone thickener is again broadly similar in form to a conventional thickener (see Fig. 10.31) but the sides of the cylindrical section are longer in relation to the diameter and the inverted cone has much steeper angle in the region of 37°. Although originally conceived on a relatively small scale, units with diameters up to 24 m are now commonly available to process suspensions at throughputs of 850+$m^3\ h^{-1}$; a 40–45 m diameter version is also available. A paddle/rake system rotating at speeds between 0.25 and 2 rpm is usually added to aid the thickening process and facilitate final sludge/paste concentrations of 70+ % w/w. Although deep cone thickeners are relatively cheap to install and occupy a relatively small floor plan area compared to their throughput the operating costs can be higher. The flocculants required to promote efficient settling and operation are generally expensive and raised power inputs may be needed to maintain the stirring action of the paddle through (the often shear thinning) pastes that are characteristic of the high viscosity underflows.

10.6.2.3 Lamella

The lamella separator is characterised by an essentially rectangular tank containing a series of closely spaced rectangular plates inclined at an angle of ~50° to the horizontal (see Fig. 10.32). The plates, which effectively increase the available settling area, allow the sedimenting solids from the feed to slide down their upper surfaces towards a sludge hopper. The clarified liquid overflow is removed from a suitable opening near to the top of the tank. Commercial designs exist for three basic flow arrangements, namely cross-current, co-current and the most popular counter-current where the feed and clarified liquid flows can be most simply arranged. With plate spacings in the region of 50 mm, lamella separators offer a compact design which may be up to 90 % smaller than an equivalent conventional gravity settler. However, maldistribution and solids re-entrainment problems can sometimes limit their effectiveness, a problem which also occurs in an alternative design incorporating inclined tubes rather than plates.

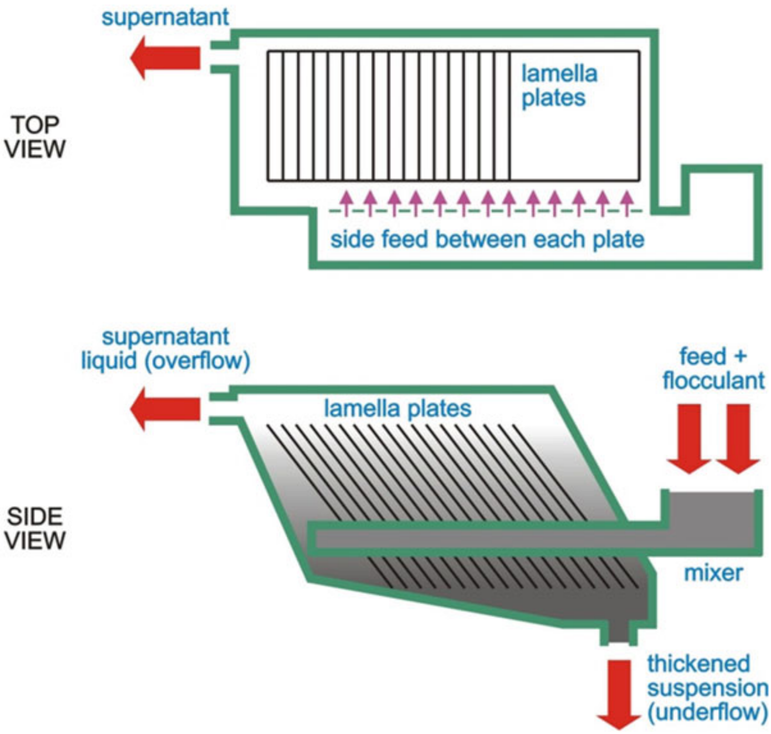


Fig. 10.32 Operating principle of a Lamella separator (Courtesy of Filtration Solutions, UK)

10.6.3 Modelling

Thickener design is traditionally based on either zone sedimentation or compression subsidence in order to establish critical thickener dimensions [9,11].

Zone sedimentation models are characterised by the approaches of Coe and Clevenger [4] and Talmage and Fitch [26], and rely upon graphical analysis of batch settling curves (Fig. 10.4); see also Kynch [13]. Both require jar settling experiments to be performed, although the method of Talmage and Fitch is generally preferred as (i) the amount of laboratory work required is much reduced and (ii) thickener capacity is underestimated such that there is an over-design in terms of floor plan area. Although there are exceptions, the model of Talmage and Fitch is generally best suited to unflocculated suspensions.

In flocculated systems, the flocs are essentially a networked structure of particles, and may have a very high porosity and be of considerable size. Therefore, even at low concentrations flocs may not settle as separate entities. As settling proceeds a compressive stress is developed in the forming sediment, so the flocs are not supported solely by their hydrodynamic drag. A suspension in compression should exhibit a compressive yield value which is a function of the solids concentration. These arguments underpin compression models such as those proposed by Michaels and Bolger [17], and Landman and White [14].

The most appropriate approach is often based on the material being separated. For instance, in the gravity thickening of biological sludges there is a strong emphasis on the compression mode of thickening whereas the design of thickeners for metallurgical and chemical suspensions relies heavily on behaviour in the settling zone. It is common to employ the theory of Talmage and Fitch to predict thickener area requirements from laboratory tests. However, when the feed suspensions are flocculated it is probably more appropriate to model the process on the basis of compression subsidence. The interested reader is directed toward the recently published work of Concha [5] which provides more detail on the modelling of gravity sedimentation.

10.7 Membrane Filters

Whilst it is evident that many of the separators in Sects. 10.2, 10.3, 10.4, 10.5, and 10.6 can effectively separate particles of less than 10 μm, when there is a significant fraction of fine material in the feed suspension, then its separation can become significantly more challenging. To be able to separate sub-micron and nano-sized materials it is common practice to use membranes. The principal characteristics of the most pertinent variants are shown in Table 10.7.

In the context of this chapter, membrane filters can be classified according to whether they are operated in a deadend mode (e.g. clarification and sterile filtration, where the feed moves normal/directly toward the membrane) or a crossflow mode

Table 10.7 Typical characteristics of some membrane separations

	MF	UF	NF	RO
Separation principle	Size	Size, charge	Size, charge, affinity	Size, charge, affinity
Typical rejected species	Silts, bacteria, cysts	Proteins, viruses, endotoxins	Sugars, pesticides	Salts, sugars
Separated size (μm)	0.1–20	0.001–0.1	~0.001	<0.0001
Pressure (bar)	0.2–1	1–5	3–15	10–60
Flux ($m^3\ m^{-2}\ day^{-1}\ bar^{-1}$)	>2	0.2–3	0.05–0.5	0.02–0.2
Recovery	90–99.99	80–98	50–95	30–90

MF is microfiltration, *UF* is ultrafiltration, *NF* is nanofiltration and *RO* is reverse osmosis

(e.g. thickening, where the feed moves tangential to the membrane to limit material deposition at the surface). Crossflow filters, with membrane areas at the bench scale of a few cm^2 to whole plants containing 1000s of m^2, are then differentiated according to the pore size in the membrane or according to the size of contaminant they will remove from the process stream. In crossflow (or low shear) filters the filter surface is stationary, and these are distinguished from dynamic (or high shear) devices which usually contain a moving surface. MF, UF, NF and RO units are similar in general form and by way of example some features of MF and UF are described in this section. The interested reader is directed to texts such as Cheryan [3] and Schaefer et al. [21] for more specific details.

10.7.1 Low Shear Crossflow

Low shear crossflow filters units usually comprise either a single membrane module or several modules arranged in a series configuration. The feed suspension is pumped at a constant rate and pressure into the module(s) and caused to flow tangential to the stationary semi-permeable membrane surface(s) at a typical linear velocity of 1–2 m s^{-1}. The shearing action at the membrane surface(s) limits material deposition to produce a relatively rapid permeate (liquid) flux decline toward the start of filtration followed by a near constant separation rate. In normal operation the permeate is collected and the retentate of thickened suspension is recirculated until the desired solids concentration is achieved, or pumping can no longer be performed satisfactorily.

Particulate deposition, fouling and adsorption of molecular species at the membrane surfaces often lead to lower than expected permeate fluxes. Whilst chemical cleaning and periodic backflushing/backpulsing with permeate or compressed gas can temporarily increase fluxes, large installed membrane areas may be required to achieve the desired separation rates. This, in conjunction with the pumping duty, means that both capital and operating costs of membrane units can be higher.

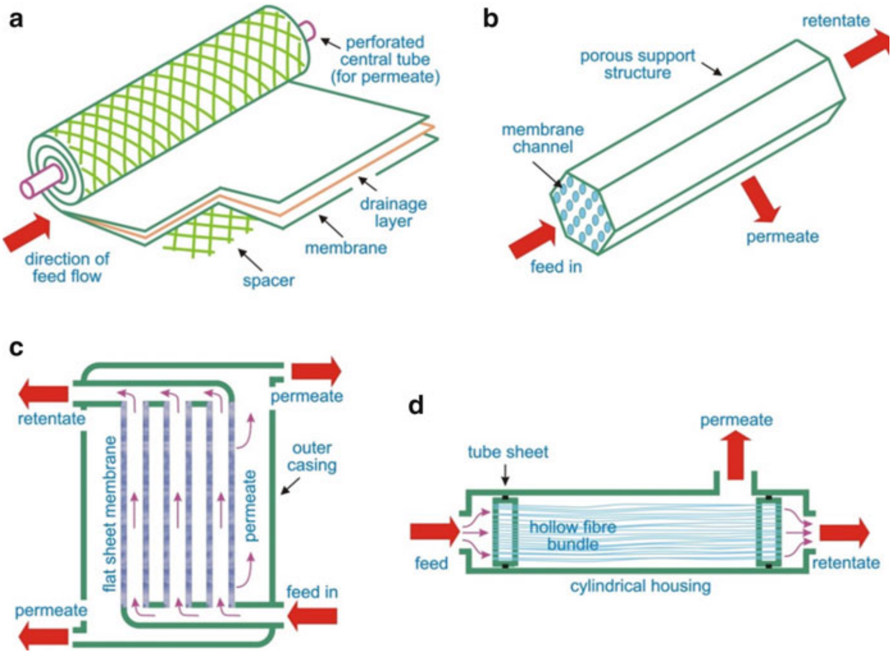


Fig. 10.33 Examples of common crossflow membrane arrangements. (**a**) Spiral wound; (**b**) tubular monolith; (**c**) plate and frame; (**d**) hollow fibre (Courtesy of Filtration Solutions, UK)

However, such costs are frequently offset by the ability to perform separations that are difficult, if not impossible, to achieve economically by other means and both ultra- and micro- filters are becoming the technology of choice in several industrial sectors.

In UF, which is typically used to separate macromolecules, viruses, bacteria, colloids and very fine suspended particles, the membranes are almost exclusively of an asymmetric, microporous construction and available with pore ratings in the range 0.001–0.02 μm. These are manufactured as microporous structures from a range of polymers and ceramics and formed as either flat sheets or tubes for use in one of four basic arrangements (see Fig. 10.33 and Table 10.8). Ultrafilters are usually operated as multiple-pass thickeners in either batch or continuous ‘feed and bleed’ modes (see Fig. 10.34) where the latter can be cascaded *ca*. three to seven times to produce a multi-stage recycle configuration.

Plate and frame: Flat porous plates covered with polymeric membrane material are assembled with alternate hollow spacers to produce a crossflow system where feed moves through the annular spaces between adjacent membrane surfaces. Although now largely superseded by other designs, this variant is still available with membrane areas up to 80 m^2.

Table 10.8 Relative comparisons between membrane arrangements

Parameter	Spiral wound	Tubular	Plate and frame	Hollow fibre
Availability	UF	UF and MF	UF and MF	UF and MF
Membrane surface per module volume ($m^2\ m^{-3}$)	600	25–50	350–600	600–1200
Investment cost	Medium	High	High	Low
Operating cost	Low	High	Low	Low
Flow control	Fair	Good	Fair	Good
Ease of in-situ cleaning	Poor-fair	Good	Fair	Fair

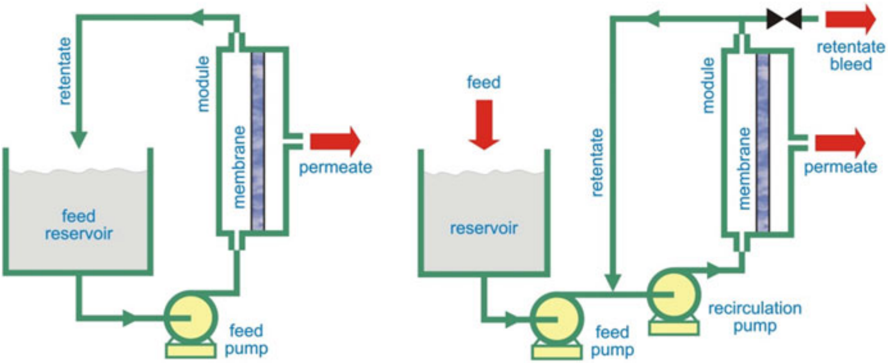


Fig. 10.34 Batch (*left*) and 'feed and bleed' (*right*) membrane plant configurations (Courtesy of Filtration Solutions, UK). The retentate from one 'feed and bleed' loop can be used as the feed to a second loop (and so on) in order to produce a cascaded or multi-stage configuration

Tubular monolith: A thin membrane layer which facilitates the separation is formed on the inside of a more robust, and open, monolith support. In earlier examples the separating layer was formed to constant depth along the length of the monolith, however, newer designs employ a progressively reducing depth that promotes better overall flux performance. A typical monolith is made from alumina or zirconia ceramic and may contain more than 30 individual channels of 4–7 mm diameter and length up to 1 m; some silicon carbide versions can contain over 200 flow channels. The feed passes along the inside of each channel to enable a separation to proceed. As many as 300 individual monoliths can be assembled into a module that is similar in form to a single-pass shell-and-tube heat exchanger.

Hollow fibre: Up to several thousand small diameter (*ca.* 40 μm to 2 mm), hollow tubular membranes are externally sealed at both ends into a larger diameter, solid cylindrical housing. The pressurised feed stream usually flows into this polymeric lumen with the permeate moving radially outward through the fibre walls ('inside-out' filtration). Hollow fibre systems can also be designed to have the feed flow on the outside of the fibres with the permeate collected from the inside of the fibres ('outside-in' filtration). Although hollow fibre ultrafilters offer the advantage of a large membrane area in a small volume, they can be prone to blockage, greater membrane fouling and also cleaning problems.

Spiral wound: Spiral modules are constructed using flat sheet polymeric membranes in the form of a pocket, consisting of two membrane sheets separated by a highly permeable mesh spacer which defines the region for permeate flow. The assembly is sealed using an appropriate epoxy or polyurethane adhesive along three edges. The open side of the pocket is glued to a central perforated tube that is used to collect the permeate flow. Several of these pockets are spirally wound around a single collecting tube using a feed-side mesh as a spacer between the pockets to establish the required feed channel thickness. Like the hollow fibre arrangement, the spiral wound module offers a large membrane area within a small volume but again suffers from potential blocking and cleaning problems.

Microfilters, which are used for the separation of viruses, bacteria, colloids and fine suspended solids, differ from ultrafilters primarily in the pore size range and construction of the membranes used to achieve a separation but are otherwise broadly similar. The polymeric, flat sheet types are usually of a symmetric construction and exhibit either microporous or track-etched forms to facilitate either depth or surface filtration. These membranes are manufactured with pore ratings of 0.02–10 μm. Ceramic and metal microfilters are also available for more extreme duties, e.g. pH range 0–14, as either flat sheet or tubular forms with pore ratings between 0.05 and 8 μm or 0.2 and 20 μm, respectively.

10.7.2 High Shear Crossflow

High shear crossflow filters offer many of the advantages of the low shear filters described in Sect. 10.7.1, but with the potential benefit of higher fluxes. Several variants exist.

A typical unit comprises a cylindrical pressure vessel enclosing 12–15 filter leaves of ~0.5 m diameter. The preferred type uses static circular filter elements with solid discs mounted between. The discs are attached to a central shaft rotating at constant speeds up to 2000 rpm. The rotation ensures the generation of relatively high shear forces and local suspension velocities in excess of 10 m s^{-1}. The feed is pumped into the pressure vessel at a rate dependent on its filtration characteristics and separation proceeds to produce a thickened suspension. As the feed thickens it invariably becomes more viscous with the result that significant rotational energy can be transferred to the feed in the form of heat and, perhaps more importantly, higher motor currents are required to turn the central shaft and discs. These disadvantages are offset by the inherent ability of the filter to decouple the shear generated at the separating surface from the overall suspension throughput. A typical filter utilises flat sheet microfiltration membranes or tightly woven, multi-filament filter cloths.

Technical alternatives include versions where the filter media themselves rotate and a variant that employs vibration of the elements to enhance filtration. For the latter up to 100 double-sided, flat disc filter elements separated by thin spacers are clamped together, mounted within a vertically mounted cylindrical vessel and caused to vibrate by a motor drive assembly close to their resonant/natural frequency. Oscillatory motion in the plane of the horizontal elements produces a shear rate up to 150,000 s^{-1} which is many times greater than that observed in typical low shear crossflow filters. In this manner rheologically sensitive feeds can be processed and filter areas up to 200 m^2 can be accommodated.

10.8 Outlook for the Future

Whilst incremental improvements will continue to be made in the mechanical design and operation of separation equipment, many of the recent, and likely future, developments in filter technology centre around improvements to filter media. The introduction of, for instance, composite media which incorporate polymer coatings [16], multi-layer media with complex weaves and combinations of different media types, activated media which facilitate both filtration and adsorptive capacity [24], functionalised media for selective separations (e.g. membranes) and the inclusion of nanofibre webs in otherwise conventional filter media has allowed for improved separation efficiencies at lower pressure drops and reduced operating cost. Although still largely in their infancy, technologies such as 3D printing to create new filter forms by successive layer formation [29] and the use of carbon nanotubes in new membranes for water purification [7] offer great potential going forward.

Developments in filter media have not only been driven by better science and engineering but also by the increased use of computational fluid dynamics (CFD). CFD incorporates a powerful suite of tools for modelling fluid flow that has facilitated not only simulations of filter media [34] but also closer examination of the flow patterns in, for instance, gravity sedimenters and centrifuges. It is likely that CFD will continue to grow in prominence in future years, however, it is important to realise its limits and the ongoing need for experimentation. We are still a considerable way from being able to predict equipment performance from a basic knowledge of equipment characteristics and, more pertinently, fundamental fluid and particle properties and their interactions. This situation is reflected in Sects. 10.2, 10.3, 10.4, 10.5, and 10.6 where there is often a need for empiricism or heuristics in the models presented. Computer programs such as Filter Design Software [8] go some way to providing automated calculations of equipment performance (and selection/scale-up), but until tractable and accurate models of fundamental suspension behaviour are developed it is unlikely that we can progress beyond the widespread use of heuristics and personal experience when designing and specifying filtration and sedimentation equipment.

10.9 Definitions, Abbreviations and Symbols

Reynolds number	dimensionless number describing the ratio of inertial and viscous forces; its value indicates the type of fluid flow (e.g. laminar, turbulent)
Schmidt number	dimensionless number describing the ratio of momentum diffusivity to mass diffusivity

ASCE	American Society of Civil Engineers
AWWA	American Water Works Association
MF	microfiltration
NF	nanofiltration
RO	reverse osmosis
UF	ultrafiltration

A	filtration area, m^2
A_p	area of cloth in a single tilting pan, m^2
c	effective concentration of solids in feed or suspension, kg m^{-3}
C	cake solids volume fraction
C_e	modified consolidation coefficient, $m^2\ s^{-1}$
C_{e0}	modified consolidation coefficient at unit applied pressure, $m^2\ s^{-1}\ kPa^{-\gamma}$
C_0	cake solids volume fraction at unit applied pressure, $kPa^{-\beta}$
d	diameter of a filter candle, m
d_i	inner diameter of filtering surface on a disc or table, m
d_o	outer diameter of filtering surface on a disc or table, m
D	diameter of drum, basket or vessel, m, or molecular diffusivity of solute, $m^2\ s^{-1}$
D_L	axial dispersion coefficient, $m^2\ s^{-1}$
D_n	dispersion number
F	fraction of solute removed from a filter cake by washing
F_D	drag force on a sphere falling through a fluid, N
g	acceleration due to gravity, m s^{-2}
h	height of centrifuge basket, m
h_B	filter belt width, m
h_D	drum width, m
i	number of drainage surfaces
j_{II}	consolidation area factor applied to 2-dimensional expression
k	permeability, m^2
L	thickness of filter cake or length of centrifuge bowl, m
m_{av}	ratio of mass of wet cake to mass of dry cake
m	mass of particle, kg
M	moisture content of a filter cake (ratio of mass of liquid in cake to total mass of wet cake)
M_s	mass of solids in a filter cake, kg

N	number of disks in a centrifuge
n	compressibility index or efficiency factor in a sedimenting centrifuge
n_d	number of discs
n_p	number of tilting pans
p	pressure, usually hydraulic pressure delivered by the pumping system, Pa
Δp	pressure difference, Pa
p_b	threshold pressure or vacuum, Pa
p_0–p_3	pressure at a given position, Pa
q	$= dV/dt$, volume flow rate of liquid (e.g. filtrate), $m^3\ s^{-1}$
Q	volumetric feed flow rate, $m^3\ s^{-1}$
Q_s	volumetric sludge (underflow) flow rate, $m^3\ s^{-1}$
r	radius of rotation, m
r_c	radius of cake layer in centrifuge, m
r_l	radius of liquid layer (or pool/pond) inside a centrifuge, m
r_0	inner radius of centrifuge bowl/basket, m
R	resistance to fluid flow through a filter medium, m^{-1}
R_1	inner radius of a conical disk, m
R_2	outer radius of a conical disk, m
Re	Reynolds number $= \rho_l u_c x_c/\mu_l$, where u_c and x_c are characteristic velocity and linear dimension, respectively
s	mass fraction of solids in a feed suspension
S	saturation, volume of liquid in a cake per unit volume of the voids
Sc	Schmidt number $= \mu_l/(\rho_l D)$
S_R	reduced saturation
S_0	specific surface of particles, $m^2\ m^{-3}$
S_∞	irreducible saturation
t	time, s
T_c	dimensionless consolidation time
u	superficial velocity, $m^3\ m^{-2}\ s^{-1}$, or particle-fluid relative velocity, $m\ s^{-1}$
u_t	terminal velocity, $m\ s^{-1}$
v_B	linear velocity of a filter belt, $m\ s^{-1}$
V, V_f	filtrate volume, m^3
V_s	solids volume fraction in a suspension or solid/liquid mixture
W	wash ratio (amount of wash liquid passed through a cake per unit amount of liquid in the cake at start of washing)
x	particle diameter/size, m
z	co-ordinate direction, or length of filter devoted to a particular process

α	specific resistance of a filter cake, $m\ kg^{-1}$
α_0	specific resistance at unit applied pressure or at zero applied pressure, $m\ kg^{-1}\ kPa^{-n}$
β	compressibility index
γ	compressibility index
ε	porosity, volume of voids per unit volume of filter cake (or porous medium)

λ compressibility index
μ viscosity of liquid in a feed or filtrate, Pa s
v consolidation index
θ dimensionless time, or conical half angle of a disk, °
ρ, ρ_l density of liquid in a feed or filtrate, kg m^{-3}
ρ_c bulk density of a filter cake, kg m^{-3}
ρ_s density of solids or particles, kg m^{-3}
σ surface tension, N m^{-1}
φ fraction of filter area
ω angular velocity, s^{-1}
ω_0 volume of solids per unit filter area, m^3 m^{-2}

Subscripts

av average value
c referring to consolidation phase
f referring to filtration phase, or to filtrate
l referring to liquid phase
s referring to solid phase
w referring to washing phase
0 initial value, unless otherwise stated
∞ equilibrium value

Superscripts

* dimensionless value (unless otherwise stated)

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