FILTRATION

1. Introduction

Filtration is a fundamental unit operation in the chemical process industry. In the simplest sense, filters separate a suspension into its component solid and liquid phases. The end result of the filtration may be one of several possible aims:

- Recovery of the liquid phase and discarding of the solid phase.
- Recovery of the solid phase and discarding of the liquid phase.
- Recovery of both the solids and liquid phases.
- Recovery of neither phase, as, eg, when water is cleaned prior to discharge to control water pollution.

The end result is sometimes as important as the material properties in selecting the best filter for the separation.

There are many different types of filtration equipment. In general the filter consists of a *filter housing* that holds a *filter medium* through which the fluid must pass. The fluid discharge from the medium is called *filtrate*. The operation may be performed on incompressible liquids or compressible gases. The physical mechanisms controlling the filtration can vary with the degree of fluid compressibility. This article specifically focuses upon liquid filtration. There are many relevant topics important to filtration that cannot be adequately covered here due to space restrictions. The reader is advised to consult the reference list for more in depth information.

This *Encyclopedia* has separate articles on topics of membrane technology, hollow-fibre membranes, reverse osmosis, and ultrafiltration. Readers should turn to those articles for information on those subjects.

1.1. Types of Liquid Filtration. There are two major types of liquid filtration: surface and depth filtration. In surface filtration, particles are captured on the surface of the filter medium, such as shown in Figure 1. The particles are captured on the surface due to a straining effect, where the particles are too large to fit through the pore openings, or the concentration of particles is large enough that they bridge over the pores.

The filter medium begins the process. The particles captured on the surface of the medium themselves stop and capture subsequent layers of particles. Hence, the cake effectively becomes the filter medium. As the layers of particles collect, a cake forms on the surface of the filter medium. Normally, the filter medium is relatively thin compared to the filter cake.

As the filter cake builds up over time, the resistance of flow through the cake increases. If the filtration is operated at constant pressure, then the flow will decrease over time. In the ideal situation, the filter cake that forms is highly concentrated in the solids and the cake particles do not clog the filter medium. At

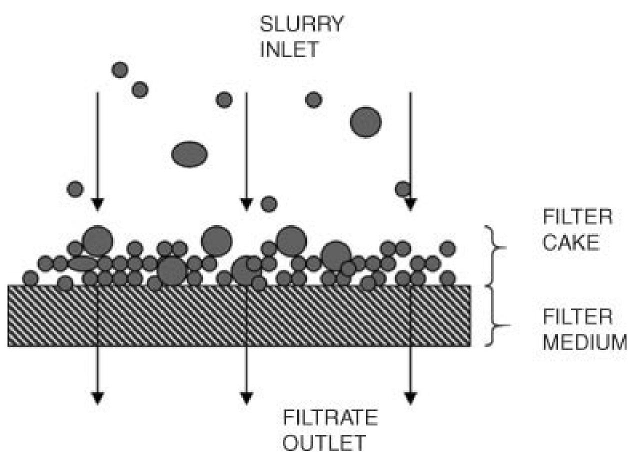


Fig. 1. Surface filtration. The particles in the slurry are stopped at the inlet surface of the filter medium. As the surface becomes covered with particles, the particles themselves stop the approaching particles. The layers of particles build up to form the filter cake.

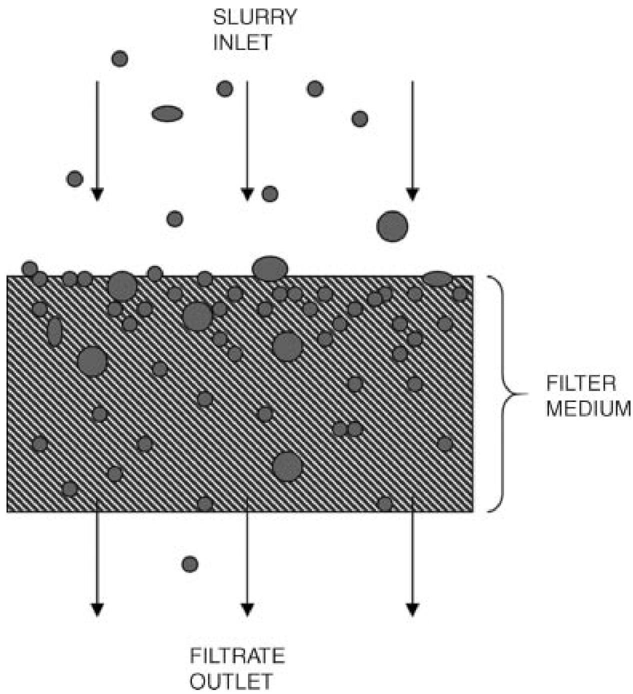


Fig. 2. Depth filtration. Particles in the slurry penetrate into the depth of the filter medium and are captured within the medium.

the end of the filtration the filter cake is separated from the filter medium and is further treated, if necessary, such as by washing and drying.

In depth filtration, the particles penetrate into the medium and are captured within its depth, as shown in Figure 2. Typical depth media have pores that are much larger than the particle sizes it is meant to remove. Ideally the particles penetrate the medium and are captured by attachment to the filter medium fibers or granules due to various interception mechanisms.

As the filter medium fills up with particles the resistance to flow increases and the flow rate decreases in a constant pressure operation. At the end of the filtration, the particles may be separated from the medium by backflow, for a rigid medium, or fluidization, for a granular medium.

To keep the frequency of backwash and volume of wash water down, and to prevent undesired cake formation on the medium surface, depth filtration is normally applied to very dilute suspensions.

1.2. Driving Forces in Filtration Operations. While pressure is the most common mechanism for driving the filtration operation, there are a number of externally applied forces that can be used. The choice of driving force in any specific application must take into account a number of factors. Some generalizations are useful in selecting the type of force (1). These generalizations are summarized here.

Gravity: Filtrations based on gravitational force are simple in design and have very low operating costs. The throughput is relatively small per unit area, hence the equipment is bulky. Cakes formed under gravity may contain an appreciable amount of liquid.

Vacuum: Vacuum filtrations are easily produced by the suction of an ordinary centrifugal pump or indirectly by a vacuum pump on a down stream pressure vessel. The maximum driving pressure difference is ~0.8 bar (~80 kPa). Vacuum filtration is simple to apply and the equipment is of moderate cost. The throughput is better than that obtained by gravity, but large surface areas are still needed, making the equipment bulky. The moisture content of the cakes tends to be relatively high. Vacuum does not work well with volatile liquids.

Pressure: Pressurized filters can achieve higher pressure differentials on the order of 1000 kPa or higher, than are possible with vacuum filters. This greater driving force provides high throughput per unit area and makes it possible to reduce the size of the equipment as compared to vacuum and gravity filters. Pressure filters are enclosed vessels that allow for handling of volatile or hazardous materials, but continuous discharge of the solids as cake is difficult. The produced cakes are usually dryer than those produced by vacuum.

Centrifugal: Centrifugal force provides for the maximum separation. High separation efficiencies can be achieved even with very fine particles. Discharge of cake can be batch or continuous. Very high throughputs are possible relative to the size of the centrifuge, but capital and operational costs can be high. Very low residual moisture content is possible with centrifuges. The cake in a centrifugal filter, depending on the filter design, is often disturbed, causing the cake particles to pack differently than with pressure filter cakes, and hence the cakes to have different properties (permeability and porosity). A related device, the hydrocyclone, is simple in construction and is compact in size. The hydrocyclone has low operating and capital costs, but outputs a thickened slurry instead of forming a cake.

In addition to the above forces, there are secondary mechanisms that may also be applied directly cause the separation or indirectly to enhance one of the other methods.

Magnetic: Magnetic fields are very useful for separating iron and magnetic ores. They can efficiently remove submicron-sized particles. The magnets require periodic cleaning to remove the captured particles. Capital and operational costs can be high and are directly related to the strength of the applied magnetic field. The main advantage is that the separation is highly efficient and can be carried out at high flow rates with minimum head loss across the filter (2).

Acoustic: Acoustic fields are known to cause movement and migration of particles. Accoustic fields applied in beds of glass beads or foam mesh have been shown to be effective in concentrating particles in slurry streams (3). Acoustic fields depend on a difference in compressibilities between the solid particles and the surrounding fluid. Acoustic fields also are energy intensive making them economical only for high-valued materials for which other methods are not effective.

2. Filter Media

Filter media control the filtration process. For a particular slurry, the filter media determine whether the filtration will produce a cake, capture particles within the depth of the media, whether the particles in the slurry will clog the media, or whether the media will be ineffective in particle removal.

Filter media come in a wide range of materials and forms. Some of the major types are listed in Table 1. A successful filter medium often requires a combination of various properties, as listed in Table 2. A discussion of these properties is provided by Purchas (4) and Purchas (5).

Most of the pertinent properties of filter media can be reasoned out with common sense. Table 2 serves as a checklist. When the particles have sizes close to the size of the pore openings the filter medium can clog. Clogging can cause excessive pressure drop to maintain a fixed flow rate. Conversely, if the pressure drop is maintained constant, clogging of the medium will cause the flow rate to rapidly decrease, thus shortening the useful life of the filter medium. To remedy a clogging problem one can go with a more open medium and capture the particles in the depth, or with a more closed medium and capture particles on the surface. If the particles are rigid the latter is the method typically used.

Table 1. **Summary of Filter Media Types and Typical Size Range of Smallest Particles Captured**[a]

Type	Examples	Effective smallest particle retained, μ
solid fabrics	flat wedge-wire screens	100
	wire bound tubes	100
	stacks of rings	5
rigid porous media	ceramics and stoneware	1
	carbon	?
	plastics	?
	sintered metals	3
cartridges	sheet fabrics	3
	yarn wound	2
	bonded beds	2
metal sheets	perforated	100
	woven wire	5
plastic sheets	woven monofilaments	
	fibrillated film	
	porous sheets	
	membranes	0.1
woven fabrics	textile fabrics	10
nonwoven media	felts and needle felts	10
	paper (cellulose or glass fiber)	2
	bonded media	10
	meltblown	<2
	nanofiber sheets	<1
loose media	fibers, powders	<1

[a] See Ref. 1.

Table 2. **Filter Media Properties**[a]

Filtration-specific properties	Machine-oriented properties	Application-oriented properties
smallest particle retained	rigidity	chemical stability
retention efficiency	strength	thermal stability
clean resistance to flow	creep/stretch resistance	biological stability
dirt-holding capacity	stability of edges	dynamic stability
tendency to bind (clog)	resistance to abrasion	adsorptive characteristics
	stability to vibration	wettability
	sealing/gasket function	health/safety characteristics
	ease of fabrication	cost
	dimensions of available supplies	

[a]See Ref. 1.

Alternatives to remedy clogging problems are to use filter-aids as a body feed or a precoat on the filter medium.

Soft deformable gel-like particles can be problematic. Soft particles can deform and spread to cover the pore openings at the surface of the medium and cause blinding of the medium. Normally, gel-like particles can only be effectively filtered at low applied pressure drops with large surface area filter media.

3. Particle Shapes and Particle Sizes

In the context of this article, the term “particle” represents a small body (of sand, salt, soil, silt, etc). Individual solid particles are characterized by their density, size (and size distribution), and shape. Particles of homogeneous solids have the same density as the bulk material. Particles obtained by breaking up a composite solid, such as an ore, may have varying densities. The size and shape for regular particles such as spheres and cubes are easily specified. The size and shapes of irregular particles are not so clear.

Irregular particles of even the same material have a wide range in sizes and shapes. The most common method of characterizing size is sieving. Particles are characterized by the size of the screen opening through which the particles can pass. As shown in Figure 3 particles that are larger than the opening cannot pass. Particles that have at least one dimension smaller than the screen opening may pass through, given sufficient time and opportunity (such as by shaking).

Separating particles by size, known as *classifying*, provides a size distribution of particles. For sieving the particles are weighed to determine the mass fraction that pass through one screen but not through the next smaller screen. Sieving provides a well-defined method for sizing the particles but it may not be the most convenient method in terms of speed, accuracy, or convenience. Screening also has a lower practical limit of $\sim$20 μ. Many methods have been developed using laser light diffraction, light absorption, electrical resistance, stokes settling velocity, and others. Allen (6) gives a good description of particle size analysis. Figure 4 shows the size range of some common particulate materials in contrast with the electromagnetic spectrum wavelength and molecular sizes. The figure also includes some common separation methods for the size ranges.

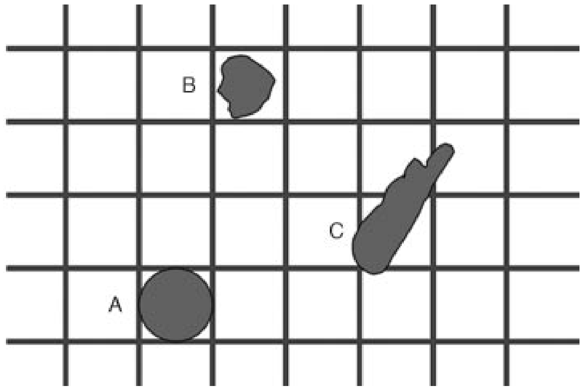


Fig. 3. Particles on screen. The diameter of spherical particle A is the same size as the screen opening. Particles smaller than A will pass through the screen and particles larger than A are captured. Irregular shape particle B is smaller than the screen opening and can pass through the screen. Irregular shape particle C is captured when its largest dimension is oriented so that it cannot fit through the opening. However, particle C could reorient so that its smaller dimensions do fit through the screen.

Particle size and size distribution are important to filter performance. A filter cake of smaller particles usually requires a higher pressure for fluid flow of a given rate than a cake of larger particles. Particles of sizes similar to the pores of a filter medium tend to clog the medium, also causing an increased pressure drop. A slurry with a wide particle size distribution may be problematic to filter because of the difficulty to find a medium that does not clog. For slurries

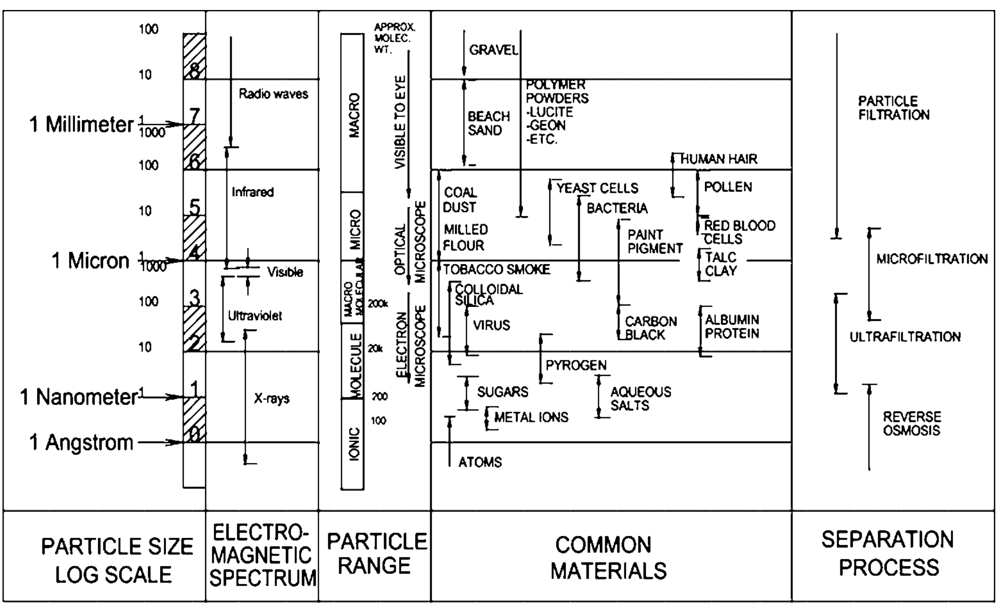


Fig. 4. Relationship of particle size and separation processes.

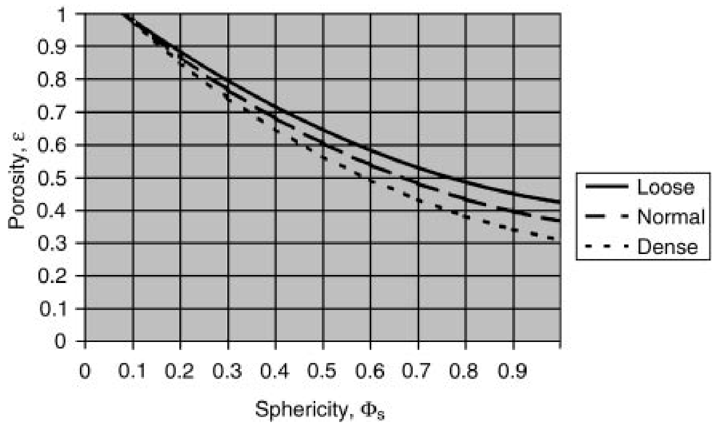


Fig. 5. Porosity, ε, as a function of sphericity, Φ_s, for random-packed beds of uniformly sized particles (8).

with small particles and wide size distributions pretreatment techniques such as flocculation or filter aid body feed may be necessary.

Most particles are not spherical. There are a number of methods to characterize particle shape. One common method to represent particle shape is to relate the shape to that of a sphere (7). The sphericity of a particle is defined as the ratio of the surface area of an equivalent sphere of the same volume of the given particle divided by the surface area of the particle of that particular volume. For a spherical particle the sphericity is one, and for nonspherical particles the sphericity has a value <1.

The definition of sphericity does not capture all of the shape information about the particle, but it is a convenient parameter and it can be used to provide useful correlations. For example, the porosity (void volume fraction) of a packed bed can be related to the sphericity (8), as shown in Figure 5. Figure 5 shows that the porosity for a given sphericity falls in a range of values dependent on whether the particles bed were allowed to loosely settle or if they were tightly packed into a dense bed. Spherical particles form the most concentrated beds, with porosities in the range of 0.3–0.45, and fibrous materials form beds with porosities >0.9 as the most open structures. Sphericity correlations have also been applied to fluidization and to settling processes (9).

4. Volume Fraction Versus Mass Fraction

Some measurement techniques use gravimetric measurements that result in data in terms of mass fractions. As can be seen from Figure 5 the void volume fraction (porosity) depends on the sphericity (shape) of the particles, their size distribution, and their geometric packing arrangement. The densities of the individual particles have no effect on their packing arrangement. Fluid flow through a filter cake depends on the characteristics and dimension of the pores and properties of the liquid and not on the density of the particles. In this context, volume fraction is more useful than mass fraction information.

5. Filter Performance

Similar to other unit operations in chemical engineering, filtration is never 100% complete. Some solids may leave the filter in the filtrate, and some liquid may be entrained in the separated solids. Because the emphasis varies with the application on the separated solid or liquid phases, the two are usually measured separately.

The separation of solids is usually measured by the fractional recovery of how much of the incoming solids are collected by the filter. The separation of the liquid is measured by how much liquid remains with the filter cake, ie, moisture content.

One method that can account for both the conditions of the cake and the filtrate, or more generally to any separation process with one inlet and two outlet streams, is the method of entropy generation. In special cases such an approach can determine criteria for when a separation is possible and the efficiency of the separations (10,11).

With depth filter media, the number of particles that pass through the filter depends on the particle size. Two common methods are available to characterize such filter performance. The first method is the beta ratio (12) and the second method is the quality factor (13).

The beta ratio is the ratio of number concentrations particle of a particular diameter and larger in the feed stream to the filtrate stream, hence the beta ratio is normally greater than unity. The beta ratio is related to Efficiency for a particular particle size, by

$$\beta_i = \frac{1}{1 - E_i} \tag{1}$$

where β_i is the beta ratio of the ith particle size, and E_i is the corresponding efficiency. Here, efficiency is defined as

$$E_i = \frac{c_{\text{in}} - c_{\text{out}}}{c_{\text{in}}} \tag{2}$$

where c_{in} and c_{out} are the number concentrations of the ith particle size in the inlet and outlet stream to the filter medium. For example, a 98% efficient filter has a beta ratio of 50.

The beta ratio may change with time as a filter clogs or bleeds particles. The beta ratio is similar to the grade efficiency curve used to characterize hydrocyclones at steady-state conditions (14).

The beta ratio is normally used to characterize filter media because it is more sensitive to small changes in the higher efficiency range than is the efficiency. The beta ratio and efficiency are often used to define the nominal rating for a depth filter medium. Manufacturers may define the nominal size of their media as the size spherical particle for which the medium captures 98%. Hence if a filter captures 98% of 2-μ spherical particles, then the filter may be given a nominal rating of 2 μ. Different filter media suppliers may use different percentages other than 98% to characterize their media.

The beta ratio and efficiency only considers the capture of particles. The quality factor considers the capture of particles as well as pressure drop. Ideally, the pressure drop through a depth filter is proportional to the thickness of the filter. Conversely, the log of the particle penetration (ratio of the number of particles passing through the filter to the number challenging the filter) is proportional to the thickness of the filter. Hence, the quality factor, defined by

$$QF = \frac{-\ln(c_{\text{out}}/c_{\text{in}})}{\Delta P} \tag{3}$$

is independent of the thickness of the filter medium. The quality factor provides a means for comparing performance between different types and thickness of filter media.

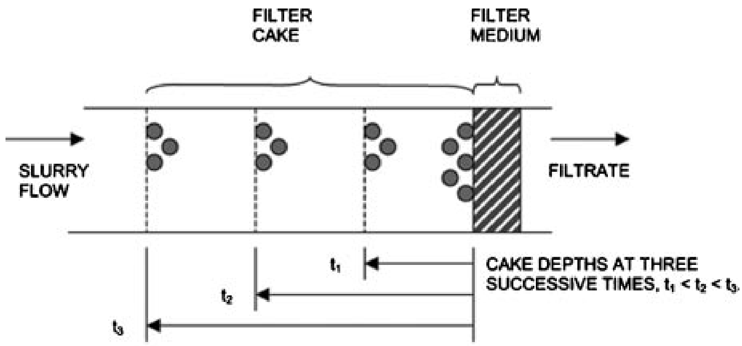
6. Cake Filtration Theory

Cake filtration is an example of flow through a porous medium. Typical questions to be answered are how much cake will be formed from a given slurry; how long will it take; how much pressure must be applied; and how much filtrate will be produced? These questions and others can be answered through analysis of mass and momentum balances on the filter cake, and with experimental data on the material properties.

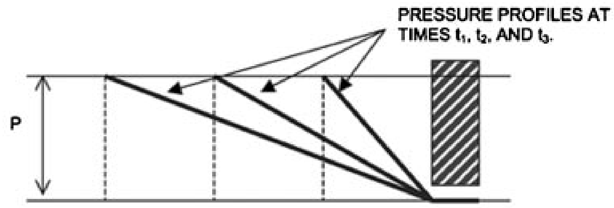
Figure 6 shows typical pressure profiles for compressible and incompressible filter cakes operated at a constant pressure drop over several increments in time (corresponding to several different cake depths). The figure also shows effects of medium resistance and when the medium clogs. As the cake depth increases or the medium clogs, the total resistance to flow increases, and hence the corresponding flow rates decrease.

To increase the filtration rate one naturally will try increasing the pressure drop. While this works in many cases, there are situations in which increasing the applied pressure drop is not advisable. In cases of severe clogging of the filter medium, the resistance to flow can quickly increase to the point that the flow rate reduces to a trickle even with a large applied pressure. Increasing the applied pressure for filtering gel-like particles usually results in squeezing more of the particles into the pores and actually reduces the flow rate below that of lower applied pressures. Super-compactable filter cakes have properties such that the flow rate becomes independent of the applied pressure.

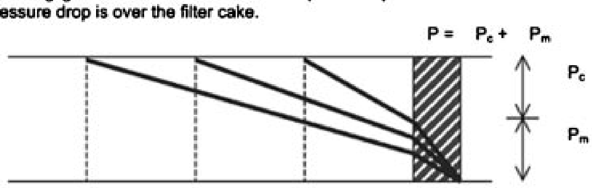
Fig. 6. Typical profiles for a constant pressure filter cake. (**a**) Shows the location of the inlet surface of the filter cake as it moves right to left over time. The filter medium is shown here with exagerated thickness. (**b**) Shows the pressure profiles for an incompressible cake and negligible medium resistance, as compared to (**c**) with a significant medium resistance. (**d**) Shows the typically nonlinear pressure profiles for a compressive filter cake and nonclogging medium in contrast to (**e**), which shows the nonlinear pressure profiles for a clogging medium.



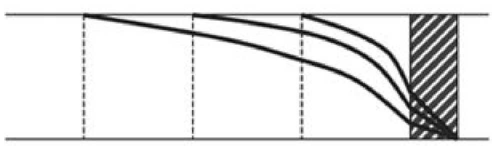
(a) Filter cake and medium at three different moments in time. The slurry enters from the left side and the filtrate exits at the right side. The filter cake grows right to left over time. The subsequent diagrams show pressure profiles corresponding to the filter cakes at the three moments in time shown here.



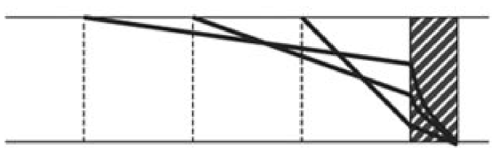
(b) Pressure profiles for constant pressure filtration of an incompressible filter cake and negligible medium resistance. The pressure profiles are linear and the total pressure drop is over the filter cake.



(c) Pressure profiles for a constant pressure filtration of an incompressible filter cake and nonnegligible constant medium resistance. The total pressure drop equals the sum of the pressure drops over the filter cake and the filter medium. For a constant pressure filtration the pressure drop over the medium decreases as the cake depth increases and cake contributes to more of the total pressure drop.



(d) Pressure profiles for a constant pressure filtration of a compressible filter cake. The pressure profiles are nonlinear with the largest gradient in the pressure occurring in the cake at the cake-medium boundary.



(e) Pressure profiles for a constant pressure filtration of an incompressible filter cake and a clogging or plugging filter medium. The filter medium may initially take only a small fraction of the pressure drop but over time the particles in the cake penetrate into the filter medium and close pores that cause the medium pressure drop to become significant.

The location of the largest pressure gradient usually controls the filtration process. In a clogging cake, the largest pressure gradient normally occurs in the filter medium. In this case, to improve the filtration rate requires either selecting a different filter medium that does not clog or changing the cake properties, such as with a body feed, so that the cake particles do not penetrate and clog the medium. For a compactable filter cake, the largest pressure gradient usually occurs in the cake at the cake-medium boundary. The filtration rate may be increased by altering the properties of the cake, such as with a body feed, to reduce the compression, change the nature of the particles to be more rigid and spherical like, or alter the fluid chemistry to reduce particle–particle repulsive forces (represented by the zeta potential).

The particular method used to improve the filtration rate depends on the underlying mechanisms that cause the clogging of the medium or compaction of the cake. To better understand and predict the performance of filter cakes several theories are briefly discussed in the following sections.

6.1. Darcy's Law for Flow Through Porous Media. In the mid-1800s, Darcy developed an empirical expression for describing the flow of water through porous sand beds (15). This relation has limited a priori predictability, but it does serve as a useful reference for comparing the rate of flow through porous materials such as filter cakes and filter media.

In its customary present day form, the macroscopic Darcy's equation is

$$\frac{Q}{A} = \frac{k}{\mu}\frac{\Delta P}{L} \tag{4}$$

where ΔP is the pressure drop across a porous material of thickness L and cross-sectional area A. The fluid has viscosity μ and flows at rate Q through the material. The permeability, k, is defined by equation (4) to relate the pressure drop to the flow rate. Typical values of permeability for common materials are listed in Table 3.

6.2. Ergun Equation to Predict Permeabillity. From Table 3, one can deduce the general trend that as particle size decreases so does the permeability. Coulson and co-workers (16) show that for regular shaped particles, such as spheres, the permeability also depends on surface area of the particles and the void fraction. When the flow rate is high enough, the permeability also becomes a function of the fluid velocity.

Table 3. **Typical Values of Permeability for Some Common Materials**

Material	Permeability, m^2
clean sand, sand–gravel mixture	10^{-12}–10^{-9}
fiber mats, paper	10^{-11}–10^{-8}
fine sand, silt, loam	10^{-16}–10^{-12}
peat	10^{-13}–10^{-11}
filter aids (diatomateous earth, Perlite, etc)	10^{-14}–10^{-12}
clay	10^{-16}–10^{-11}
granite	10^{-20}–10^{-18}

One popular correlation that accounts for all of these factors that control the permeability is the Ergun equation. The derivation of the Ergun equation is described elsewhere (17). In its derivation, the particles are assumed to be spherical and the surface area effect is accounted for by relating specific surface area to the particle diameter. The void fraction of the porous material is explicitly introduced into the correlation in the derivation. The Ergun equation applies to porous material with void fractions <0.9. The Ergun equation is modified to irregularly shaped particles through the sphericity parameter, Φ_s, described previously (7). The Ergun equation is commonly written as, with correction by MacDonald and co-workers (18).

$$\frac{\Delta P}{L} = \frac{180 V_o \mu}{\Phi_s^2 d_p^2} \frac{(1-\varepsilon)^2}{\varepsilon^3} + \frac{1.80 \rho V_o^2}{\Phi d_p} \frac{(1-\varepsilon)}{\varepsilon^3} \tag{5}$$

where ε is the void fraction (porosity). Sphericity is defined by

$$\Phi_s = \frac{6/d_p}{s_p/v_p} \tag{6}$$

where s_p and v_p are the area and volume of the particle. The d_p is the nominal particle diameter defined by

$$d_p = \sqrt[3]{\frac{6 v_p}{\pi}} \tag{7}$$

In the Ergun equation, V_0 is the approach velocity for an empty bed, equivalent to

$$V_o = \frac{Q}{A} \tag{8}$$

By comparing the Ergun equation to Darcy's law, we can deduce that the permeability is related to the particle sizes, porosity, and flow rate by

$$k = \frac{\Phi_s^2 d_p^2 \varepsilon^3}{(1-\varepsilon)^2} \frac{1}{\left(180 + 1.80 \frac{\rho \Phi_s d_p}{\mu(1-\varepsilon)} V_o\right)} \tag{9}$$

When the flow rate is small enough the velocity term in the bracket on the right side of equation (9) is negligible and the permeability becomes is only dependent on the particle size, particle shape, and porosity.

The Ergun equation applies for porosities <0.9. Above 0.9, the Langmuir model is recommended (19). The Ergun equation also does not work well for fibers and plate-like objects (low sphericities). For fibrous materials other correlations are available (20).

6.3. Basic Two-Resistance Model for Cake Filtration. Consider that the filter cake apparatus consists of an assembly that holds the filter medium and the filter cake that forms on the surface of the medium. The total pressure drop over the apparatus, neglecting fittings and other hardware, is the sum of the pressure drop across the cake plus the pressure drop across the medium.

$$\Delta P = \Delta P_c + \Delta P_m \tag{10}$$

where the subscripts c and m refer to the cake and medium, respectively. Instead of permeability we can define the resistance to flow, R, through a porous material by the expression

$$\Delta P = \mu R V_o \tag{11}$$

and where the resistance is related to the permeability by $R = k^{-1}$. The total pressure drop over the filter cake becomes

$$\begin{aligned} \Delta P &= \mu R_c V_o + \mu R_m V_o \\ &= \mu(R_c + R_m)V_o \end{aligned} \tag{12}$$

The approach velocity is related to the filtrate volume, V, by

$$V_o = \frac{1}{A}\frac{dV}{dt} \tag{13}$$

The resistance of the cake is directly proportional to the mass of solids deposited per unit area of the filter. The proportionality constant, α, is defined as the specific cake resistance. The specific resistance is related to the cake resistance and the filtrate volume by the expression (21)

$$R_c = \frac{\alpha c V}{A} \tag{14}$$

and c is a concentration corrected for the presence of the liquid in the cake, related by

$$c = \frac{\rho^l s}{1 - ms} \tag{15}$$

$$m = \frac{\text{mass of wet cake}}{\text{mass of dry cake}} = \frac{\varepsilon_c \rho^l + (1 - \varepsilon_c)\rho^s}{(1 - \varepsilon_c)\rho^s} \tag{16}$$

$$s = \frac{\text{mass of solids in slurry}}{\text{total mass of slurry}} = \frac{(1 - \varepsilon_{\text{slurry}})\rho^s}{\varepsilon_{\text{slurry}}\rho^l + (1 - \varepsilon_{\text{slurry}})\rho^s} \tag{17}$$

Here ρ^s and ρ^l refer to the solid and liquid material intrinsic densities. ε_c and $\varepsilon_{\text{slurry}}$ are the volume fractions of the liquid in the cake and slurry, respectively.

By combining equations (12)–(17), we obtain the classical two resistance model for cake filtration:

$$\frac{1}{A}\frac{dV}{dt} = \frac{A\Delta P}{\mu(\alpha cV + AR_m)} \tag{18}$$

This expression is readily integrated for a constant pressure drop to obtain a parabolic relation

$$\alpha cV^2 + 2VAR_m = 2A^2\Delta Pt \tag{19}$$

between the filtrate volume and time.

Equation (18) has received much attention in literature and can be used to interpret experimental data (21–24). It may be rearranged to give

$$\frac{\Delta P}{\mu(1/A)(dV/dt)} = \alpha\frac{cV}{A} + R_m \tag{20}$$

from which a plot of $\Delta P/\mu(1/A)(dV/dt)$ versus cV/A for an incompressible cake and nonclogging medium gives a line with an intercept of R_m and slope of α. With recent developments in computer software, equation (19) may also be fitted to experimental data without relying on problematic measurements of the filtrate volume rate data in the early stages of the constant pressure filtrations.

6.4. Compactable Cake Filtration Modeling. When the cake is incompressible, the pressure profiles are linear as shown in Figure 6 and the plot of equation (20) gives a constant slope, hence constant α. When the cake is compactable, under pressure it compresses, and the slope in equation (20) is not constant. A number of relations have been proposed to account for changing cake behavior. Typically, power law type relations are employed, such as (21,25)

$$(1 - \varepsilon_c) = (1 - \varepsilon_{co})\left(1 + \frac{P_s}{P_a}\right)^{\beta} \tag{21}$$

$$\alpha = \alpha_o\left(1 + \frac{P_s}{P_a}\right)^{n} \tag{22}$$

where ε_{co} and α_0 are the cake porosity and specific resistance at zero stress at the inlet surface of the filter cake. P_a, β, and n are fitted material parameters. P_s is referred to as the solids compressive stress, related to the fluid pressure and the cake pressure drop by

$$\Delta P_c = (P - P_o) + P_s \tag{23}$$

where P_0 is the fluid pressure at the filter cake exit (at the cake—medium boundary).

The behavior of filter cakes can be classified according to the magnitudes of parameter n (26). When n is nearly zero the cake is incompressible, when n is in the range $0.4 < n < 0.7$ the cake is moderately compactable. When n is in the range $0.7 < n < 0.8$ the cake is highly compactable. Finally, when $n > 1$ the cake is supercompactable.

Interpretation of experimental data to determine the fitted parameters in equations (21) and (22) are possible by applying some simplifying assumptions and integrating the governing equation, equation (18) (26). Smiles (27) combines the effects of compactability with the drag mechanism to formulate the governing equation in the form of the diffusion equation in material coordinates that requires only one material constitutive relation. Robust numerical methods, such as Genetic Algorithms, may also be applied to fit filtration data directly to fitted parameters (28).

To predict cake compressibility one can use a compression permeability cell (C-P cell) (24). The intent of the C-P cell is to apply a known normal stress to the solid cake material in the cell while simultaneously recording pressure drop and flow rate data. The parameters in equations (21) and (22) can be determined from this data, assuming that the local porosity and local specific resistance in the C-P cell are the same as in the filter cake. The C-P cell has limitations and the wall friction leads to nonuniformities in the cake structure and the distribution of the local stresses.

6.5. Advanced Cake Filtration Modeling. Advanced cake filtration models normally account for other mechanisms that affect the cake formation, such as sedimentation (29) or geometric affects (30,31). Advanced models also allow for modeling of non-Newtonian behavior such as power-law fluids (32,33) or yield-stress fluids (34).

Multiphase continuum theory obtained from volume averaging of the single-phase continuum equations provides a rigorous and complete set of equations to account for mass, momentum, energy, and chemical species for modeling flows through porous media (35). For isothermal one-dimensional (1D) cake filtration with no chemical reactions and no mass transfer, the continuum equations reduce to a form analogous to the basic model described previously (36,37). The continuum equations are easily applied to more complex processes, such as cake filtration combined with expression, sedimentation, or chemical reactions through appropriate use of boundary conditions and constitutive functions (38).

The disadvantage to the volume averaged approach is it requires effort to learn the notation and to understand how to apply the concepts at different scales of phenomena. Simplifying assumptions are usually required to reduce the equations to forms that can be integrated and usefully applied. Often the set of partial differential equations are too complex to obtain analytical solutions. The volume averaging theory is less intuitive than the ad hoc approach.

Continuum modeling of cake filtration, whether from volume averaging theory or from an *ad hoc* approach, requires appropriate constitutive relations. Constitutive theory has been applied to the multiphase continuum equations to generalize the constitutive equations (39). Other constitutive relations can be determined using approaches applied in other disciplines such as soil mechanics to account for phenomena such as creep and swelling (40,41).

Another advanced approach to model cake filtration is by a pore-structure model. Pore-structure models may represent the porous media as an array of spherical particles or as an array of capillary tubes and tube network (20). The pore-structure model has the advantage that the single-phase rheological properties of the fluid are applied to the flow through the channels as defined by the structure of the solid phase. Pore structure models circumvent the need for constitutive equations at the mixture scale by modeling the system at the single-phase scale. Disadvantages to this approach are the pore structure may not represent the particular medium of interest and they tend to be computationally intensive and only a limited number of pore structures have been investigated.

7. Depth Filtration Theory

Depth filtration occurs when the particles in the slurry are much smaller than the pore openings in the filter medium. The particles penetrate into the filter medium and are captured on the surfaces of the pore walls. Depth filtration is typically applied to polish fluids such as beverages.

The basic equation for modeling depth filtration is

$$\frac{\partial c}{\partial z} = -\lambda c \tag{24}$$

where

c = concentration of particles in the liquid passing through the filter
λ = filter coefficient
z = distance into the filter medium from the filter surface.

At the start of a filtration, assuming uniform medium properties, equation (24) integrates to give the concentration profile of particles in the liquid phase as

$$\frac{c}{c_o} = e^{(-\lambda_o z)} \tag{25}$$

where

c_0 = concentration of particles in the liquid at the inlet surface
λ_0 = clean medium filter coefficient.

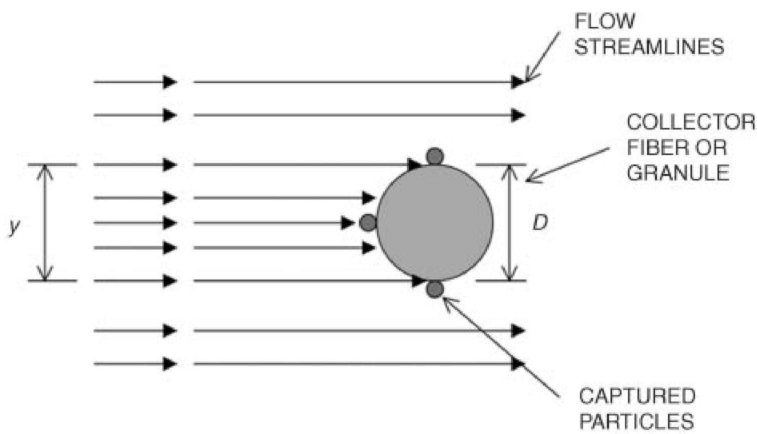
Empirically the filter coefficient can be related to the deposition of particles in the filter through a polynomial expression such as

$$\frac{\lambda}{\lambda_o} = 1 + a_1\sigma + a_2\sigma^2 + a_3\sigma^3 \tag{26}$$

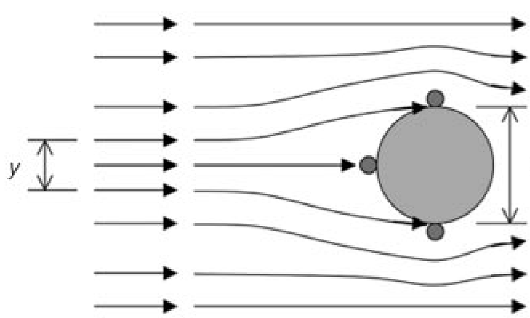
where a_1, a_2, and a_3 are empirical constants dependent on the concentration of the solids, the geometric properties of the filter medium, and the material properties of the fluid, particles, and medium (42).

7.1. Single Collector Efficiency. The filter coefficient may be modeled as the combination of many single collectors per unit volume. The single collectors may be individual granules (43) or fibers (13). In the work by Brown (13), the theory is presented for gas-phase filtration, but the fundamental concepts apply to both gas and liquid filtration.

The filter coefficient is defined by equation (24) as the proportionality constant between the rate of capture of particles per unit length of depth into the filter medium and the concentration of particles in the fluid phase. A collector with an efficiency of unity removes from the fluid stream all particles that lie within the volume swept by its projected area and the velocity vector, assuming rectilinear flow, as shown in Figure 7(**a**). For real filter media, the efficiency is reduced because the flow streamlines bend around the collector and carry some of the particles with them [Fig. 7(**b**)].



(a) Ideal depth medium, 100% efficient capture. Capture efficiency given by $y/D = 1$ where y represents the area through which the flow streamlines pass that intersect the collector. In the idealized case the area of intersecting streamlines equals the projected area of the collector, represented by D.



(b) Real depth medium, the area of streamlines that intersect the collector is smaller than the projected area, due to curvature of the streamlines, hence the efficiency is 100%, given by $y/D < 1$.

Fig. 7. (**a**) An idealized medium would capture all particles passing through the projected area of the collector. In a real medium, (**b**), only a fraction of the streamlines passing through the projected area intersect the collector.

For fibers, the filter coefficient, λ, is related to the single collector efficiency by

$$\lambda = \frac{4(1-\varepsilon)E}{\pi D} \tag{27}$$

for a filter medium of fibers of diameter D and porosity ε.

7.2. Mechanisms of Capture. There are a number of mechanisms that contribute to the capture of particles by a collector. Most of the mechanism models assume that if a particle comes into physical contact with a collector surface that the particle is captured. Particle bounce and particle release are possible but are often ignored.

The five primary mechanisms of single collector capture of a particle are

1. **Direct Interception:** Direct interception occurs when a particle follows a fluid streamline and the streamline flows close enough to the collector surface that the particle touches the surface and is captured (Fig. 8).
2. **Inertial Impaction:** Inertial impaction occurs when a particle's inertia causes the particle to continue on a forward trajectory and to not follow the fluid flow streamlines bending around a collector surface (Fig. 8).
3. **Brownian Diffusion:** Very small particles of low mass are influenced by collisions with fluid molecules. This effect is known as Brownian diffusion. These collisions can cause small particles to diffuse from a stream line that would carry the particle past a collector into a streamline that passes close enough to a collector that the particle is captured (Fig. 8).
4. **Sedimentation:** If the particle density is significantly greater than the fluid density, then particles will settle out of the flow in the direction of gravitational force.
5. **Electrostatic Forces:** Particles and filter media surfaces often carry electrostatic charges. These charges may cause repulsion or attraction of particles in the fluid to the surface of the filter collectors.

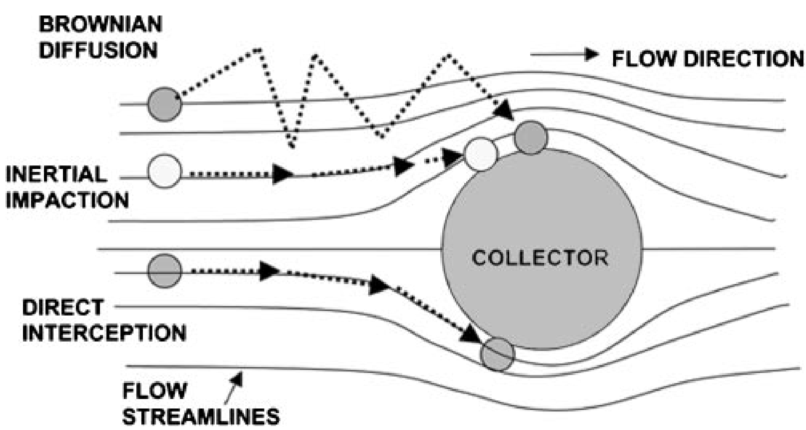


Fig. 8. Diagram illustrating particle trajectories and capture on a collector due to the mechanisms of direct interception, inertial impaction, and Brownian diffusion.

The efficiency of each mechanism may be predicted with appropriate correlations, as given by Brown (13) and Tien (43). For example, the single collector efficiency for a fiber at high Stokes Number is given by

$$E_I = 1 - \frac{0.805}{St} \tag{28}$$

where the Stoke's number is defined as

$$St = \frac{d_p^2 \rho^l V_o}{18 \mu d_f} \tag{29}$$

All of the mechanisms can occur at the same time and contribute to the overall collector efficiency. When the collection mechanisms occur independently, the overall capture efficiency is given by

$$1 - E = \prod_{i=1}^{j} (1 - E_i) \tag{30}$$

7.3. Most Penetrating Particle Size. Of interest to many applications is the most penetrating particle size. The most penetrating particle size is the size particle for which the overall net filter efficiency is lowest.

As particle sizes increase, the single collector efficiencies tend to decrease. However, as the particle size increases a straining mechanism also contributes to the net filter efficiency. Figure 9 illustrates how the combination of these two different types of mechanisms contributes to the overall net filter efficiency. The net efficiency usually has a minimum for particles of diameters in the range of 100–800 nm.

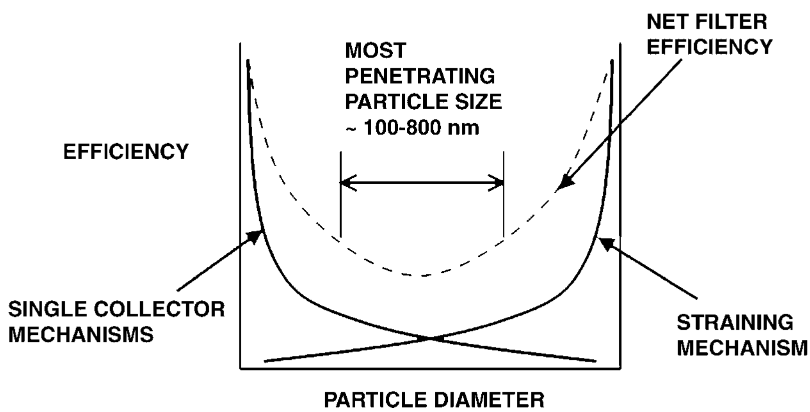


Fig. 9. The combined Single Collector Efficiencies and Straining Efficiencies contribute to the overall net filter efficiency as a function of particle size. The minimum in the net efficiency corresponds with the most penetrating particle size. This usually occurs for particles of size in a range of 100 to 800 nm.

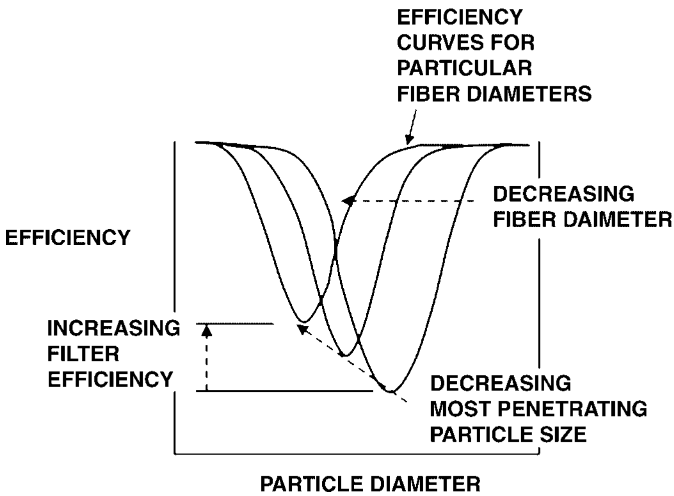


Fig. 10. Filter efficiency as a function of fiber diameter. As the fiber diameter decreases the most penetrating particle size decreases and the filter efficiency increases.

Graham and co-workers (44) recently reported that for fibrous media, as the fiber diameter decreases the most penetrating particle size decreases, as indicated in Figure 10.

7.4. Advanced Modeling of Depth Filtration. As in cake filtration, depth filtration may be modeled using the multiphase continuum theory derived from volume averaging of the single phase equations. In this approach, the captured particles may be treated as a separate phase (hence a three phase system: the liquid as one phase, the particles carried in the liquid and captured in the medium as a second phase, and the medium itself as the third phase). Or, the particles can be modeled analogously as a chemical species.

A more complete analysis of depth filtration can be obtained by solving the species balance along with the mass and momentum balances from multiphase continuum theory to model the performance of the depth filter including the flow rate, pressure drop, and shadow effects on the capture efficiency and permeability (45–47).

8. Stages of Solid–Liquid Separations

Solid–Liquid separations often involve a wide variation in material properties and operating conditions. One item of equipment or process technique often is not sufficient to achieve the desired separation. Tiller (48) viewed solid–liquid separation processes as consisting of one or more of four possible stages, as diagramed in Figure 11.

The stages are defined based on the intent of the operation and not necessarily on the type of equipment involved. Some equipment may be used in more than one stage. A summary of the four stages is provided here. More detailed

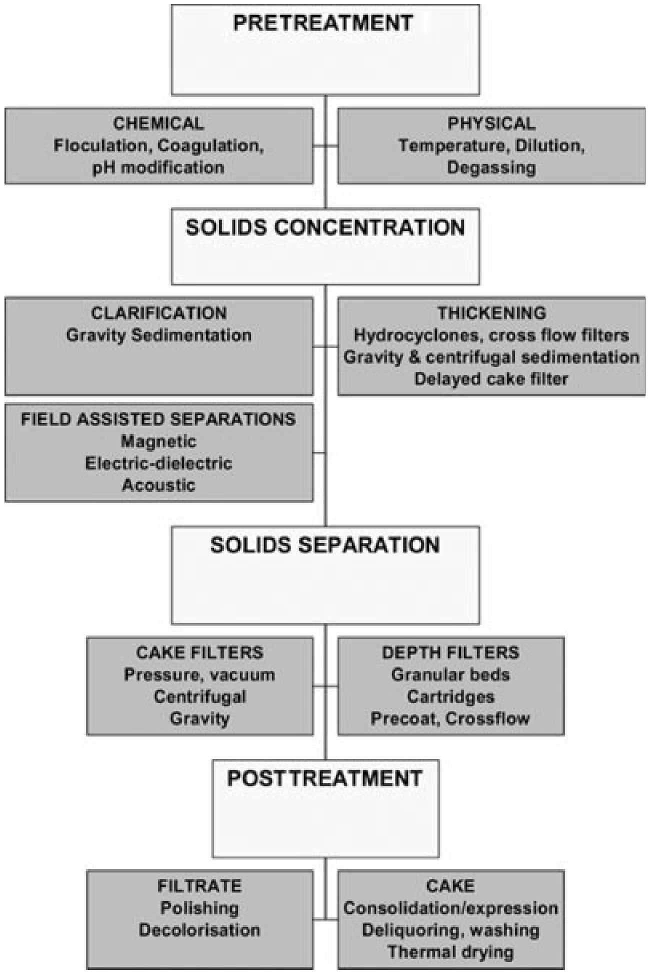


Fig. 11. Stages of solid–liquid separations. A number of the separation processes may be used alternatively as pretreatment, solid concentration, or solid separation techniques.

information are provided in literature such as in publications by Purchas (1), Wakeman and Tarleton (21), La Mer and Healy (49), Gregory (50), and Moody (51).

8.1. Pretreatment. Pretreatment is normally used with slurries that are difficult to separate. Such slurries include those that have slime or gel particles, very fine particles that tend to clog and block entrances to pores in a filter medium, or the surface chemistry of the particles may cause them to stick to the filter medium. The intent of the pretreatment is to change the slurry properties, by changing the properties of the liquid or the solid particles, to improve the separation by a subsequent process.

Pretreatment may be separated into physical or chemical treatments. For physical treatments, the simplest approach may be to change the viscosity. The flow rate through a filter cake modeled by equation (12) shows that the flow rate is inversely proportional to the liquid viscosity. This means that if the liquid viscosity is reduced by one-half the filtration rate can be doubled. Liquid viscosity is strongly a function of temperature; hence if the materials are not thermally sensitive one way to reduce the viscosity is to raise the temperature of the fluid. Another useful approach is to dilute the liquid with a less viscous liquid.

Another pretreatment for the liquid is degassing. As the liquid flows through a filter cake the absolute pressure of the liquid decreases. If bubbles are in the liquid these bubbles expand and can block pores in the filter cake. Chase and Willis (52) measured and compared the resistance to fluid flow through a filter cake with and without the presence of air bubbles. Their results showed a significant increase in resistance due to the presence of the air bubbles.

Dissolved gasses may not appear in the feed stream to a filter but as the absolute pressure drops the liquid may degas and bubbles may form within the cake or the medium. This is particularly troublesome for vacuum filters because the pressure is reduced below atmospheric pressure. There are three methods to reduce the formation of bubbles: increase the upstream pressure of the filtration pressure, so that the downstream pressure is high enough to prevent gas dissolution from the liquid; thermal or vacuum degassing of the liquid prior to filtration; and steam stripping of the liquid prior to filtration.

Physical pretreatment of the solid particles include crystallization to increase the particle sizes, freezing or other phase changes to increase size or improve shape; and the application of filter aids. Filter aids are materials such as diatomaceous earth, may be used to precoat the filter medium (in essence, change the surface of the filter medium to prevent clogging), or by adding a body feed. The filter aid material particles are usually larger than the particles in the slurry and serve to capture or trap the smaller particles, yet the filter aid material is highly porous and allows the fluid to pass through.

Chemical pretreatment of the slurry is mostly concerned with causing the particles to agglomerate to form larger particles. There are two primary ways that this can be accomplished. The first is by means of coagulation in which very fine particles of colloidal size adhere directly to each other as a consequence of Brownian motion. For coagulation to occur normally the mutually repulsive electrical surface forces must be depressed by the addition of ions to the slurry.

The second method is by flocculation. Flocculation is the formation of more open aggregates that form by high molecular weight polymers acting as bridges between particles.

8.2. Solids Concentration. Concentrated slurries are more likely to form filter cakes, by particle bridging across pore spaces, and thus reduce filter medium clogging. Concentrating slurries also has an economic advantage. The operating cost of pumping a slurry through a filter cake is directly related to the volume of fluid that must pass through the cake. Taking a feed stream that is 1% solids by volume and concentrating it to 10% reduces the liquid content by $\sim$90.9%.

A number of methods may be used for concentrating solids. Some of the more common are cross-flow filters, hydrocyclones, and gravity settlers. More recently, other field-assisted techniques have been developed that may be useful for thickening.

8.3. Solids Separation. Solids separations involve filters. In the context of this article the filters are classified as cake filters or depth filters. There are many types and variations on each of these categories. Some of these types are listed in Figure 11. Descriptions of some of these filters are provided in the later sections.

8.4. Posttreatment. Posttreatment is used to improve on the quality of the final product in the filtration. If the product is a low turbidity liquid, then posttreatment may require polishing the filtrate to remove particles to below a specified standard. Posttreatment of filter cakes commonly includes deliquoring, washing, and drying. Deliquoring physically reduces the liquid content, such as by expression. Washing is done to remove soluble components from the cake. Drying is done to further reduce the moisture content of the filter cake.

9. Filter Cycles

Cake filtration by its nature is a batch operation. There are a few methods, such as rotary vacuum filters, that can be designed for continuous throughput. The type of filter, its cycle rate, and its size all impact on production rate. This production rate is important for upstream and downstream operations in a process plant.

9.1. Continuous Filter Cycles. A horizontal rotating tilting pan filter (Fig. 12) operates quasicontinuously by sequentially filtering, deliquoring, washing, drying, and removing the filter cake from the pan in a series of stages. The pans are lined on the bottom with the filter media. The slurry is loaded into the

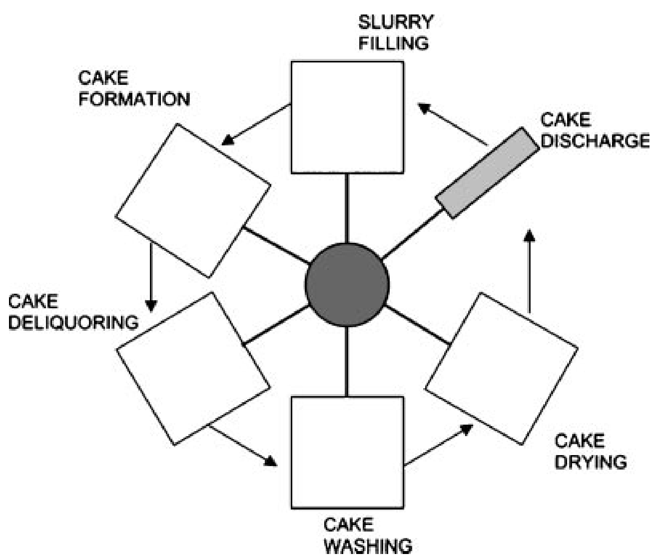


Fig. 12. Simplified sketch top view of a horizontal tilting-pan filter. The pan at the top is filled with slurry. As the apparatus rotates the cake is formed under vacuum. The formed cake is deliquored, washed, and dried. The final step is cake discharge. After discharge the pan returns to the filling location.

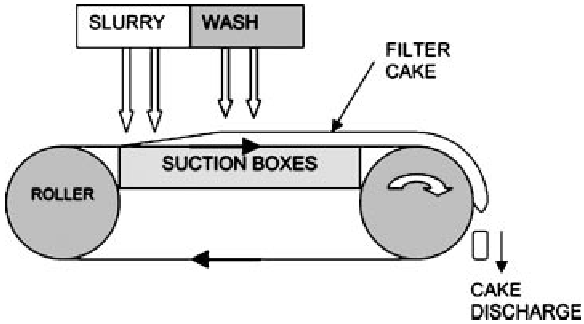


Fig. 13. Simplified sketch of a side view of a horizontal belt filter.

first pan and a vacuum is pulled on the bottom of the pan to form the cake. In the subsequent stages, the cake is processed until it is finally discharged by tilting the pan for the cake to fall out.

The rotation rate of the filter must be tuned to the rates required for the various stages. The rotation rate establishes the frequency at which a dried cake is discharged from one of the pans.

If such a filter has n pans, each pan produces a dry cake of volume V_c(m^3), and the rotation rate is ω (rev/min) then the production rate (m^3/min) of dried cake from the filter is given by

$$Q_c = nV_c\omega \tag{31}$$

If the dry cake has a porosity ε_c and the solids have intrinsic density ρ^s (kg/m^3) then the mass rate of production of dry cake material, $\bullet_c$ (kg/min), is

$$\dot{M}_c = (1 - \varepsilon_c)\rho^s Q_c = (1 - \varepsilon_c)\rho^s nV_c\omega \tag{32}$$

Rotary drum filters and horizontal belt filters can be operated continuously and their analyses are similar. Considering the horizontal belt filter shown in the diagram in Figure 13, the filter cake is formed on the surface of the belt by vacuum, is washed and deliquored.

If a belt filter of width L(m) forms a cake of volume V_c(m^3/m^2) per unit area of belt, and the belt travels at a rate of s (m/min) then the production mass production rate for the belt filter is

$$\dot{M}_c = (1 - \varepsilon_c)\rho^s \hat{V}_c L s \tag{33}$$

A similar type of expression can be derived for the rotary drum filter.

9.2. Batch Filter Cycles. A cycle for a batch filter includes a combination of steps, such as cake formation, deliquoring, washing, expression, drying, discharge, and cleaning. The total time for such an operation is the sum of all of the times for each individual step

$$t_{\text{cycle}} = \sum_{i=1}^{j\,\text{steps}} t_i \tag{34}$$

If the total dry cake volume produced at the end of the operation is V_c then the filter production rate is

$$\dot{M}_c = (1 - \varepsilon_c)\rho^s V_c / t_{\text{cycle}} \tag{35}$$

Methods for optimizing the filtration time are reported by Wakeman and Tarleton (21). If t_f is the cake formation time and t_d is all of the other nonfiltration time steps grouped together into a downtime, then the optimum filtration time can be calculated. For a constant pressure filtration with a constant the medium resistance, R_m, and constant specific cake resistance, then the optimum cake formation time is

$$t_{f_{\text{opt}}} = t_d \left[1 + \sqrt{\frac{2\mu R_m}{\alpha c \Delta P t_d}} \right] \tag{36}$$

When the medium resistance, R_m, is small compared to the specific cake resistance, α, then the second term in equation (36) is small and the optimum cake formation time is equal to the downtime. For any other case the optimum time is always greater than the downtime. Hence, the filtration formation time should be greater than or equal to the nonfiltration downtime (53).

10. Solid–Liquid Separation Equipment Selection

10.1. Introduction. Solid–liquid separations (SLS) are often viewed as a necessary evil that must be tolerated to produce a product, clarify liquids, dispose of wastes, recover raw materials, decontaminate process fluids, etc. Because of these multiple objectives, a plant engineer is faced with a large array of SLS equipment from which to choose. Compounding the selection problem, various industries have adopted certain specialized equipment. The pharmaceutical industry, eg, rarely uses chemical process equipment, and vice versa. Furthermore, standards vary from industry to industry. Vendors may intensify the selection problem by publishing ambiguous or misleading information. A number of equipment selection guides have been published to help a plant engineer through the maze to select SLS equipment.

10.2. Selection Guides. The typical plant engineer usually relies on intuition and vendors' recommendations when selecting SLS equipment. This is fine as long as the vendor can deliver a workable solution to the SLS problem with adequate performance and reasonable cost. However, the plant engineer must be wary of this approach because a vendor may not have a grasp of the total process and upstream or downstream factors that may be important in the selection. Also, a vendor may have a bias toward a particular type of technology and not consider other equipment options that may be more appropriate for a particular application. A plant engineer needs an independent means of selecting equipment to help avoid these pitfalls. There are many good references (1,2,21,54–57) that provide a comprehensive discussion of the selection process. Many of these references are based on the work by Grace (58–60) and Tiller (48,61–68).

The most useful selection criteria that the authors have adopted are based on Tiller's work (48,66,67); Porter and co-workers cake formation rate (69); Svarovsky's simplified scheme (70); Day's charts for centrifuge selection (71); and finally Purchas' guide in his recent SLS texts (1,54). Of these, Purchas' and Tiller's guides are most comprehensive in that they cover the entire SLS spectrum, whereas the others are more limited in scope. Other schemes are generally pertinent to a particular industry, and hence, are quite limited, eg, *Metalworking Fluids' Filter Selection Guide* (72).

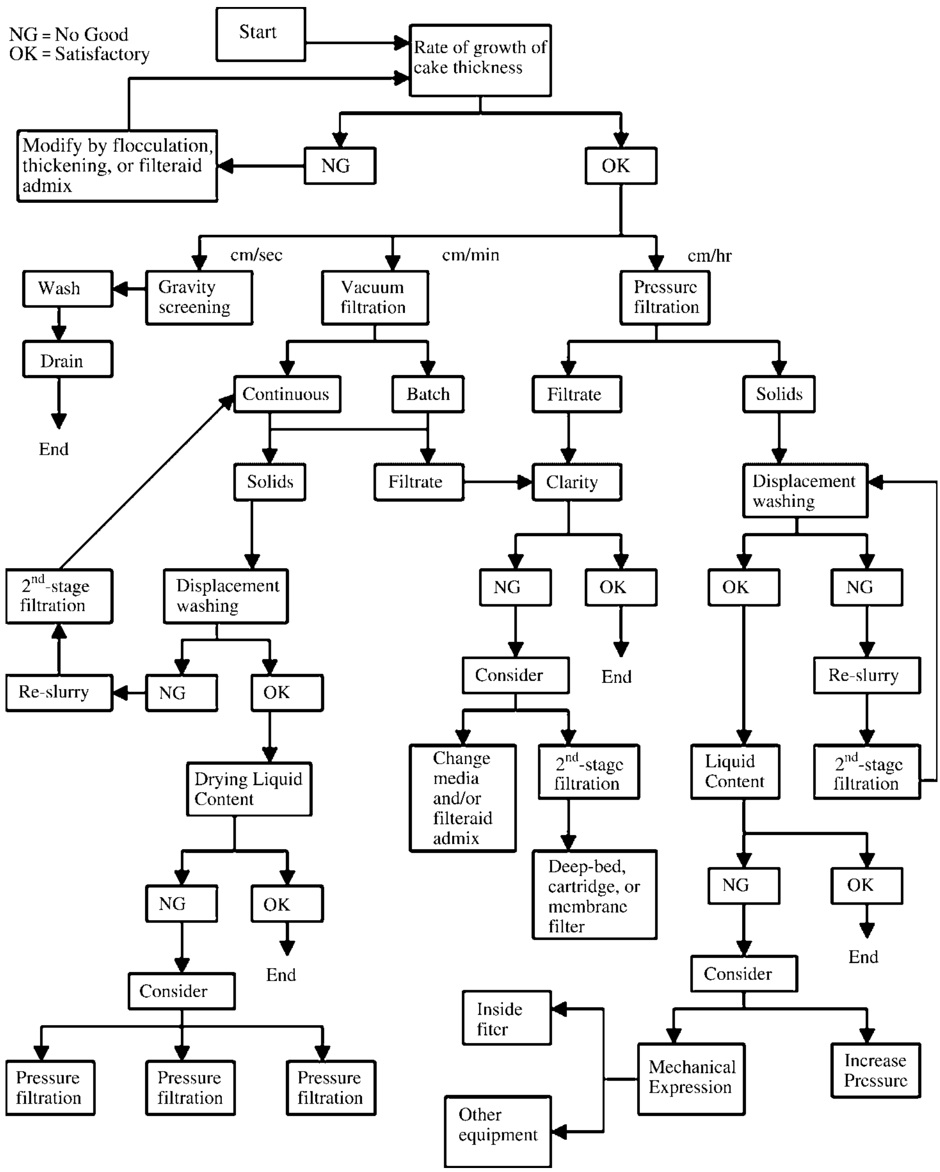


Fig. 14. Tiller's 1978 Guide. Decision Network for SLS Equipment Selection. This guide shows the steps to take in solving an SLS problem (66,67).

As an example, Tiller's guide (66,67), Figure 14 is a useful extension of Porter, and co-workers (69) cake formation rate criteria. Experimental data are needed to establish cake growth rate. You start at the top of the decision tree, answer questions and follow the appropriate arrows until you get to the final selection box. This guide can also be combined with Tiller's simplified equipment selection list (48), and the stages and methods of SLS, Figure 11, to provide a fairly comprehensive and reasonably accurate scheme for SLS selection.

On the other hand, Purchas' guide (1,54) provides the most useful and comprehensive selection scheme, since it also includes clarification as well as centrifugation, screening, and membrane-type operations. It is quite useful to select the initial equipment choices suitable for the plant's particular SLS problem. In addition, it often shows that many options are often available, but that only a few are suitable and that testing should be limited to these few. The authors have repeatedly used Purchas' guide with quite satisfactory results.

Rather than reproducing all of these guides here in this document, the reader is encouraged to turn to the original references. Provided in Table 4 is a list of SLS selection guides that the authors have found useful. The first two categories listed in Table 4 are general guides that cover a wide range of SLS equipment. The remaining categories are more specific to specialized equipment types.

The last category in Table 4, Computer-aided programs, refer to general guides that can be programmed on a computer that enable the user to narrow the selection process rationally and quickly. This is quite important since most equipment manufacturing companies' representatives are usually only interested in selling their particular SLS equipment. However, the user can still use the guides suggested here in conjunction with equipment vendors to at least narrow the choices since there are several thousand of SLS devices from which to choose.

Table 4. **References of SLS Selection Guides**

Category	Cited references
general purpose guides	1,2,48,54,56,66,67,69,70,72,73,
new trends in SLS equipment	74
sludge dewatering	75
automatic pressure filters	76
filter presses	77
candle filters and clarification reusable filters	78
centrifuges	71,79,80
hydrocyclones	81
gravity filters	82
vacuum filters	83
general pressure filters	84
cross-flow filters	85
deep bed filters	86
computer-aided programs	21,87,88

11. Filter Media Selection

11.1. Introduction–Selection. Media selection can be one of the most important items in the SLS selection process, but is rarely given much attention, considering that the medium controls the initial filtration process (refer to be section Filter Media). In addition, it can dominate the entire process if clogging occurs, if a cake will form, whether particles are retained in the media, and whether the cake will release adequately (also refer to Tables 1 and 2 in the section Filter Media). An excellent overview of filter media of all types can be found in Purchas' handbook (5), while general discussion of media selection and definitions are presented in (1,89,90). Specific media such as general nonwovens (91), filter press cloths (92), ceramics (93), flash-spin Tyvek® (94,95), and wire mesh (96) are examples of the wide array of available media. Purchas (4,97) and Shoemaker (98) discuss the vast media choices available, and relate these to various properties of machine-, application-, or filtration-orientated properties (4,21). The machine-orientated properties are normally discussed with the media supplier; while the filtration and application properties must be decided primarily by the user although some filtration-specific properties can be obtained from the media supplier (4,5,21,99).

11.2. Types of Media. There are many types of media. In addition to the ones mentioned above, there are also deep-bed grains, sintered metals (powder, mesh, and fibers), membranes, filter papers, needle felts, a whole array of nonwovens, perforated and wedge-wire metals, as well as filter aids of diatomaceous earth (DE), perlite, cellulose fibers, coal fines, and calcined rice hull ash (5). Many of these woven–nonwoven, sintered, or paper filter media can be fabricated into filter bags, socks (77), and cartridges (5). The breadth of this topic precludes a complete discussion here, and the reader is referred to the many references cited, especially (5).

11.3. Testing. There are hundreds of test methods for evaluation and characterizing filter media (5,12), but the most prevalent today are air permeability coupled with porometry (ie, pore size distribution—90,91,100,101) with automated porometers (100,101). A thorough discussion of this technique from a practical as well as a theoretical viewpoint can be found in (12,100–104). Applications are discussed for needlefelts and nonwovens (91), filter press clothes (94,105), track-etched membranes (100,101,106), and for Tyvek flash-spun nonwovens (95). Additional test methods and details can be found in (5).

11.4. Summary. As mentioned above, a wide variety of filter media exist today for almost any application. Selection involves many factors and usually requires test work in order to obtain the best medium for the user's application (5,12,91,101). The medium selected must satisfy all of the user's needs as well as provide long, economical life.

12. Vacuum Filters

In vacuum filters, the driving force for filtration results from a suction on the filtrate side of the medium. Although the theoretical pressure drop available for vacuum filtration is 1 bar in practice it is often limited to 0.7–0.8 bar (532–608 mm Hg).

Vacuum filters include the only truly continuous filters built in large sizes that can provide for washing, drying, and other process requirements. Vacuum filters are available in a variety of types, and are usually classified as either batch operated or continuous. An important distinguishing feature is the position of the filtration area with respect to gravity, ie, horizontal or nonhorizontal filtering surface. Vacuum filters generally produce lower cake solids than pressure filters.

A number of vacuum filter types use a horizontal filtering surface with the cake forming on top. This arrangement offers a number of advantages: gravity settling can take place before the vacuum is applied, and in many cases may prevent excessive blinding of the cloth due to action of a precoat formed by the coarser particles; if heavy or coarse materials settle out from the feed they do so onto the filter surface, and can be filtered; and fine particle penetration through the medium can be tolerated because the initial filtrate can be recycled back onto the belt. Top-feed filters are ideal for cake washing, cake dewatering, and other process operations such as leaching. Horizontal filter surfaces also allow a high degree of control over cake formation. Allowances can be made for changed feeds and/or different cake quality requirements. This is particularly true of the horizontal belt vacuum filters. With these units the relative proportions of the belt allocated to filtration, washing, drying, etc, as well as the belt speed and vacuum quality, can be easily altered to suit process changes.

There are, however, two significant drawbacks to horizontal filters, ie, such filters usually require large floor areas, their capital cost is high, and they are somewhat difficult to enclose for hazardous materials. Saving in floor area as well as in installed cost can be made by using a filter with vertical or other nonhorizontal filtration surfaces but at the cost of losing most, if not all, of the advantages of horizontal filters (2,70).

12.1. Nutsche Filter. The nutsche filter (Fig. 15) is simply an industrial-scale equivalent of the laboratory Buechner funnel. Nutsche filters consist of cylindrical tanks divided into two compartments of roughly the same size by a

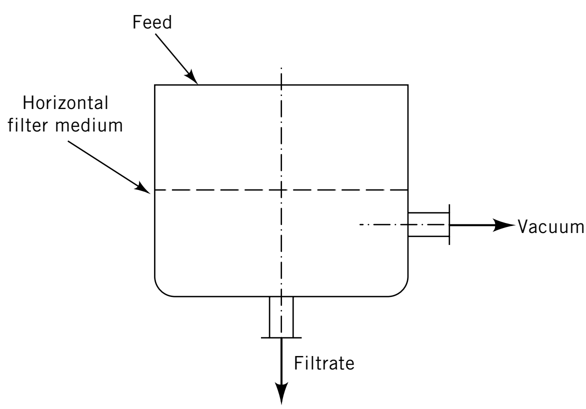


Fig. 15. Schematic diagram of a nutsche filter.

horizontal medium supported by a filter plate. Vacuum is applied to the lower compartment, into which the filtrate is collected. It is customary to use the term nutsche only for filters that have sufficient capacity to hold the filtrate from one complete charge. The cake is removed manually or sometimes by reslurrying.

These filters are particularly advantageous when it is necessary to keep batches separate, and when extensive washing is required, although separation of wash from the mother liquor is difficult. Washing is carried out either by displacement or by reslurrying, the latter being more conveniently performed in the enclosed agitated nutsches described separately. If cake cracking occurs, the cracks can be closed manually for the subsequent washing.

Nutsche filters are simple in design, but laborious in cake discharge. They are prone to high amounts of wear due to the digging out operation. They are completely open and quite unsuited for dealing with inflammable or toxic materials. For such applications enclosed agitated or pressure nutsche filters are used. Throughputs are limited but the range of solids that can be filtered is very wide, ranging from easily filterable precipitates to dyestuffs, chemicals, etc.

12.2. Enclosed Agitated Vacuum Filters. These filters, often called mechanized nutsches, are circular vessels provided with a cover through which passes a shaft carrying a stirrer. The stirrer can sweep the whole area of the filter cake and can be lowered or raised vertically as required. During the filtration process the stirrer is in the top position, in motion if necessary to prevent the formation of an unequal layer of cake with fast settling suspensions. The suspension is fed through a connecting piece in the cover. When filtration is completed, any cracks that might occur in the filter cake can be smoothed over by lowering the stirrer on top of the cake and pasting over the cake to cover the cracks. The stirrer is rotated in the reverse direction for this operation and the pressing of the cake by the rotor blades might result in an additional dewatering, particularly with thixotropic cakes.

The filter cake can then be washed either by displacement or by reslurrying. Reslurrying is easily accomplished using the stirring action of the rotor blades when the rotor is lowered into the cake. The cake may also be dried *in situ* by the passage of hot air through it, or may be steam distilled for the recovery of solvent.

For discharge of the cake, a discharge door is provided at the edge and the cake is moved by the rotor toward the door, the stirrer blades gradually being lowered onto the surface so that a small depth of solids is scraped away. Alternatively, the cake may be reslurried and pumped away.

Enclosed agitated filters are useful when volatile solvents are in use or when the solvent gives off toxic vapor or fume. Another significant advantage is that their operation does not require any manual labor. Control can be manual or automatic, usually by timers or by specific measurements of the product. Filtration areas up to 10 m^2 are available and the maximum cake thickness is 1 m. Applications are mainly in the chemical, pharmaceutical, and foodstuff industries for the recovery of solvents.

The pressure version of the enclosed agitated filter is known as the Rosenmund filter; it uses a screw conveyor to convey the cake to a central cake discharge hole. Other vendors now also supply these.

12.3. Vacuum Leaf Filter. The vacuum leaf, or Moore, filter consists of a number of rectangular leaves manifolded together in parallel and connected to a vacuum or compressed air supply by means of a flexible hose (Fig. 16). Each leaf is composed of a light pervious metal backing, usually of coarse wire grid or expanded metal set in a light metal frame and covered on each surface with filter cloth or woven wire cloth. The leaves, which are carried by an overhead crane during the filtration sequence, are dipped successively into a feed slurry tank, where the filtration takes place; a holding tank, where washing occurs; and a cake-receiving container, where cake discharge is performed by back-blowing with compressed air. An alternative arrangement is to move the tanks rather than the leaf assembly.

The operating cycle is seldom <2 h, and several sets of frames can be operated in rotation. The cake thickness should be >3 mm, with 9 mm being a typical value. Sluicing of the cake with a jet of compressed air has been used to permit thinner cakes and shorter filtration times. The leaves are spaced sufficiently far apart that there is always clearance between the finished cakes.

Simple design, general flexibility, and good separation of the mother liquor and the wash are important advantages of vacuum leaf filters. On the other hand, they are labor-intensive, require substantial floor space, and introduce the danger of the cake or leafs falling off during transport. Vacuum leaf filters are particularly useful when washing is important, but washing is difficult with very fine solids, such as occur in titanium dioxide manufacture.

12.4. Tipping Pan Filter. This is a nutsche filter with a small filtrate chamber, in the form of a pan built so that it can be tipped upside down to discharge the cake. A separate vessel is used to receive the filtrate; this allows segregation of the mother and wash liquor if necessary.

A variation on this type of filter is the double tipping pan filter, which is a semicontinuous type consisting of two rectangular pans fitted with a filter cloth and pivoted about a horizontal axis. Slurry is first fed onto one pan, which is turned over for cake discharge at the end of the cycle. The second pan is used for filtration while the first is being discharged.

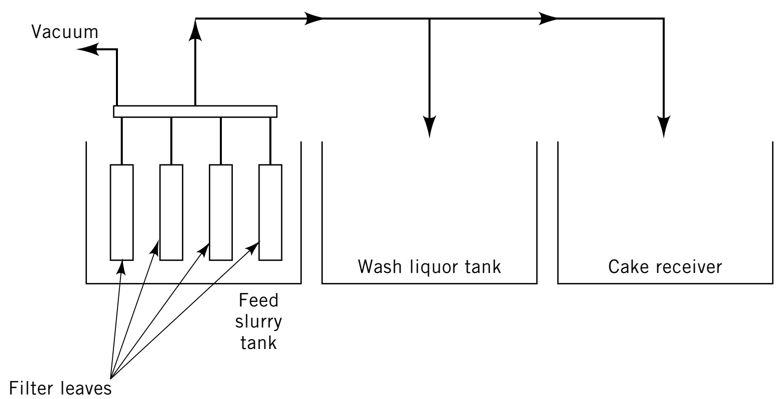


Fig. 16. Schematic diagram of the operation of a Moore vacuum leaf filter.

In general, pan filters are selected for freely filtering solids and thick filter cakes. Cake washing can be introduced easily. Most applications are in the mining and metallurgical industries for small-scale batch filtration.

12.5. Horizontal Rotating Pan Filters. These filters (Fig. 17) represent a further development of the tipping pan filter for continuous operation. They consist of a circular pan rotating around the central filter valve. The pan is divided into wedge-shaped sections covered with the filter medium. Vacuum is applied from below. Each section is provided with a drainage pipe that connects to a rotary filter valve of the same type as in drum filters. This allows each section, as it rotates, to go through a series of operations such as filtration, dewatering, cake washing, and discharge. Two basic designs exist, depending on the method of solids discharge.

The horizontal pan filter with scroll discharge incorporates a spiral scroll in a radial position just ahead of the feed zone. The scroll scrapes the cake off the medium. The need for a clearance between the scroll and the medium can cause incomplete cake discharge and consolidation of the remaining heel, due to the action of the scroll. This problem may be overcome by injection of compressed air in the reverse direction through the cloth under the feed zone or by sucking air through the cloth in the opposite direction, combined with cloth washing. Filtration areas range from 0.5 to 100 m^2.

An important variation of this filter is based on replacing the rigid outer wall necessary for containing the feed and the cake on the rotating table by an endless rubber belt. The belt is held under tension and rotates with the table. It is in contact with the table rim except for the sector where the discharge screw is positioned, and where the belt is deflected away from the table to allow the solids to be pushed off the table. The cloth can also be washed in this section by high pressure water sprays. This filter, recently developed in Belgium, is available in sizes up to 250 m^2, operated at speeds of 2 min/revolution, and cake thicknesses up to 200 mm.

In horizontal rotary tilting pan filters, the wedge-like compartments are arranged as independent pans. Each is connected to the center valve by a swivel pipe joint that inverts the pan as it passes the discharge point. Air blowback is

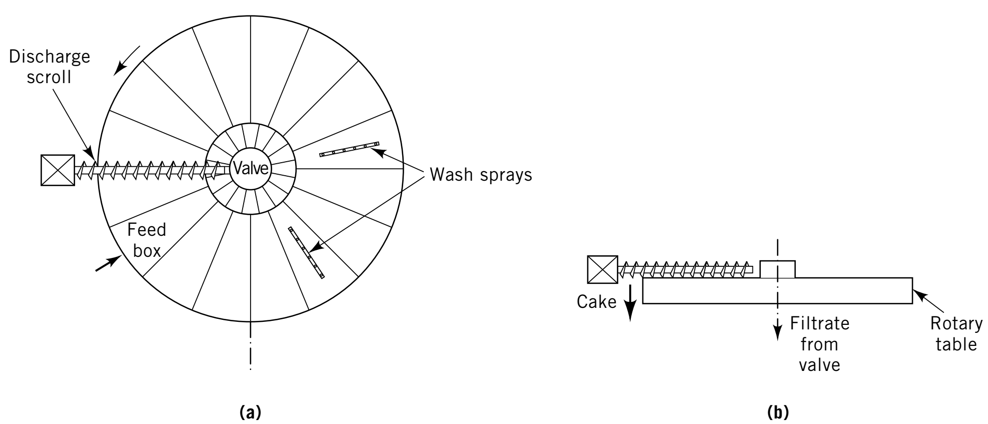


Fig. 17. Horizontal rotating pan filter: (**a**) plan view; (**b**) elevation.

often used to assist the cake discharge. The units can also be adapted for cloth washing. Typical applications include filtration of gypsum from phosphoric acid and many mineral processing uses. Areas from 15 to 250 m^2 are available. The tilting pan is more expensive, requires more floor space, and has higher maintenance than the horizontal rotating pan filters. However, its advantages include excellent cake discharge and control of wash liquor, and the availability of larger sizes.

12.6. Horizontal Belt Vacuum Filters. This type of filter (Fig. 18) is another development of the pan filter idea. A row of vacuum pans arranged along the path of an endless horizontal belt was the original patented design. This has been superseded by the horizontal belt vacuum filter, which resembles a belt conveyor in appearance. The top strand of the endless belt is used for filtration, cake washing, and drying, whereas the bottom return is used for tracking and washing of the cloth. There is appreciable flexibility in the relative areas allocated to filtration, washing, and drying. Hooded enclosures are available wherever necessary. Modular construction of many designs allows field assembly as well as future expansion if process requirements change.

Horizontal belt filters are well suited to either fast or slowly draining solids, especially where washing requirements are critical. Multistage countercurrent washing can be effectively carried out due to the sharp separation of filtrates available. Horizontal belt vacuum filters are classified according to the method employed to support the filter medium.

One common design is typified by a rubber belt mounted in tension. The belt is grooved to provide drainage toward its center. Covered with cloth, the belt has raised edges to contain the feed slurry, and is dragged over stationary vacuum boxes located at the belt center. Wear caused by friction between the belt and the vacuum chamber is reduced by using replaceable, secondary wear belts made of a suitable materials such as (PTFE), etc.

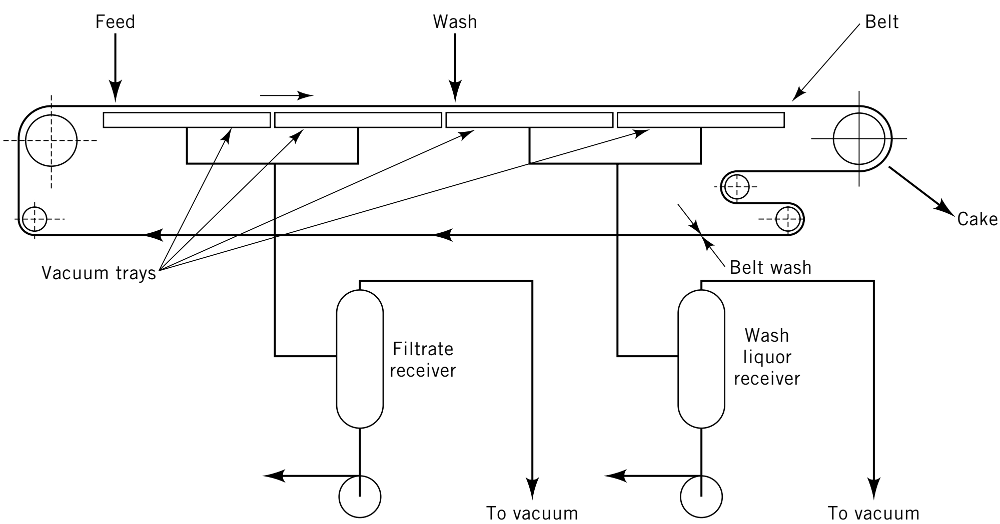


Fig. 18. Schematic diagram of a horizontal belt vacuum filter.

The rubber belt filters are available in large capacities with filtration areas up to 200 m^2 or more. They can be run at very high belt speeds, up to 30 m/min, when handling fast-filtering materials such as mineral slurries. The main disadvantages of rubber belt filters are the high replacement cost of the belts, the relatively low vacuum levels, and limitations on the type of rubber that can be used in the presence of certain solvents. The vacuum pumps are usually sized to give from 2 to 20 m/min air velocity over the filtration area, at $\sim$0.7–0.8 bar, depending on the filterability of the solids. The solids capacity ranges from 10 to 500 kg of solids per m^2 of filtration area per hour with cake thicknesses from 3 to 150 mm.

Another type of horizontal belt vacuum filter uses reciprocating vacuum trays mounted under a continuously traveling filter cloth. The trays move forward with the cloth as long as the vacuum is applied and return quickly to their original position after the vacuum is released. This overcomes the problem of friction between the belt and the trays because there is no relative movement between them while the vacuum is being applied. The mechanics of this filter are rather complex, and the equipment is expensive and requires intensive maintenance. A range of solvents can be used. Widths up to 2 m and areas up to 75 m^2 are available. The cloth can be washed on both sides.

The indexing cloth machines are a further development. In these, the vacuum trays are stationary and the cloth is indexed by means of a reciprocating discharge roll. During the time the vacuum is applied, the cloth is stationary on the vacuum trays. When the vacuum is cut off and vented, the discharge roll advances rapidly, moving the cloth forward. The cycle is then repeated. As with the reciprocating tray types, the cloth can be washed on both sides. The cake discharges by gravity at the end of the belt when it travels over the discharge roll. The primary advantages of this filter are its simple design and low maintenance costs. The main disadvantage is the difficulty of handling very fast-filtering materials on a large scale. Areas up to 93 m^2 are available.

Some horizontal belt vacuum filter designs incorporate a final compression stage for maximum mechanical dewatering. This is achieved by another compression belt that presses down on the cake formed in the preceding conventional filtration stage.

12.7. Rotary Vacuum Drum Filters. This is the most popular vacuum filter. There are many versions available and they all incorporate a drum that rotates slowly, $\sim$1–10 min/revolution, about its horizontal axis and is partially submerged in a slurry reservoir (Fig. 19). The perforated surface of the drum is divided into a number of longitudinal sections of $\sim$20 mm in thickness. Each section is an individual vacuum chamber, connected through piping to a central outlet valve at one end of the drum. The drum surface is covered with a cloth filter medium and the filtration takes place as each section is submerged in the feed slurry. A rake-type slowly moving agitator is used to keep the solids in suspension in the slurry reservoir without disturbing the cake formation. The agitator usually has a variable speed drive.

Filtration can be followed by dewatering, washing, and drying. In some applications, compression rolls or belts are used to close possible cracks in the cake before washing or to further dewater the cake by mechanical compression. Final dryness of the cake can also be enhanced by fitting a steam hood. Systems

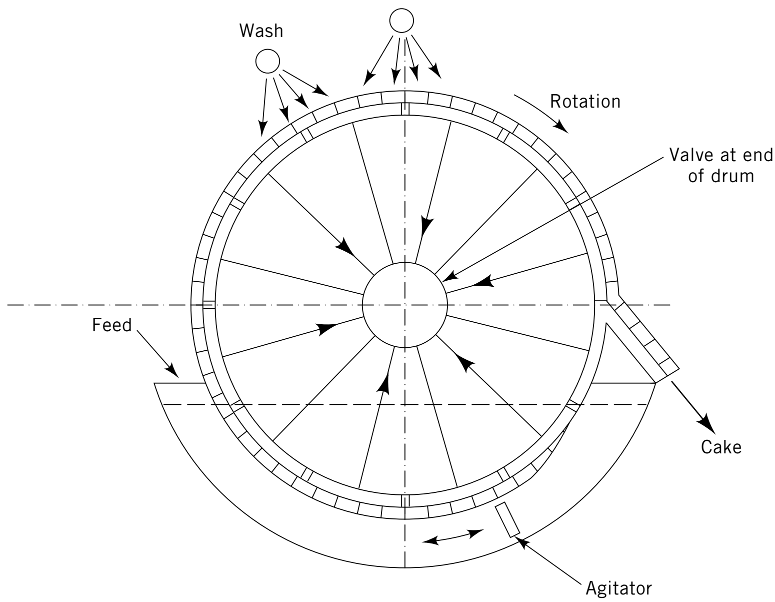


Fig. 19. Schematic diagram of a rotary vacuum drum filter.

of cake discharge, all of which can be assisted by air blowback, include simple knife discharge, advancing knife discharge with precoat filtration, belt or string discharge, and roller discharge. The type of discharge selected depends on the nature of the material being filtered.

The scraper or knife discharge consists of a blade that removes the cake from the drum by direct contact with the filter cake. It is normally used for granular materials with cake thickness greater than ~6 mm. In order not to damage the filter cloth, a safety distance of 1–3 mm between the blade and the cloth must be observed. If the residual layer is made not of filter aid but of the product, there is danger of its blocking by fine particles and by successive consolidation by the scraper blade.

The string discharge arrangement uses a number of parallel strings tied completely around the filter at a pitch of 1–2 cm, passing over the discharge and return rolls. As the strings leave the drum before the discharge point, they lift the filter cake from the medium and discharge it at the discharge roll. This type of discharge is recommended for gelatinous or cohesive cakes. Wire, chains, or coil springs have also been used. The coil spring discharge has been successfully applied to fibrous solids such as waste from a pulp mill or highly flocculated slurries such as in sewage treatment.

The cloth belt discharge is based on taking the cloth endless belt off the drum in the same way as with the string discharge. The advantage here is the ease of washing both sides of the cloth before the cloth returns to the drum. The disadvantage is in the need for an additional control device for the guidance of the cloth.

Roll discharge systems use a roll, which rotates at a slightly greater peripheral speed than the cake and is in contact with the cake. The cake is transferred from the drum to the discharge roll by adhesion which, by giving the roll a rough surface or because of the presence of residual cake on the roll, is designed to be greater to the roll than to the drum. The cake is usually removed from the roll by a knife. This type of discharge is designed to discharge tacky cakes that cannot be handled effectively by either of the previously described discharge designs. The cake thickness is small here, from 0.5 to 3 mm.

Multicompartment drum filters range in size from ~1 m^2 to >100 m^2; they are widely used in mineral and chemical processing, in the pulp and paper industry, and in sewage and waste materials treatment.

Despite their theoretically poor washing performance, due to uneven wash distribution and excessive run-off because the filter surface is not horizontal, many multicompartment drum filters continue to be used as cake washing filters. Effective washing of the filter cloth can be done only with the belt discharge type, where the cloth leaves the drum for a brief period and can thus be washed on both sides.

Because in the most common bottom feed version the feed suspension enters the drum from the bottom, fast settling slurries are not suitable for most rotary drum filters; nor are very fine slurries because of inevitable penetration problems.

In units designed to use a precoat filter aid, the drum can be evacuated over the full 360° and fitted with an advancing knife system that continuously shaves off the deposited solids together with a thin layer of the precoat. The precoat has to be renewed periodically.

No internal piping and no conventional filter valve are needed with single-cell drum filters where the entire drum also operates under vacuum. The cake discharge is effected by air blowback from an internal stationary shoe mounted inside the drum at the point of discharge. There are very close tolerances between the inside surface of the drum and the shoe in order to minimize the leakage. The inside of the drum acts as a receiver for the separation of air and filtrate; conventional multicompartment drum filters require a separate external receiver. This type of filter permits operation of the filter with thin cakes so that high drum speeds, up to 26 rpm, can be used and high capacities can be achieved. Sizes up to 14 m^2 are available.

In most rotary drum filters, the submergence of the drum is usually about one-third of its circumference. Greater submergence is achieved in units equipped with submerged bearings, although this is more costly.

Another option available with rotary vacuum drum filters is full enclosure. This enables operation under nitrogen or other atmospheres, for reasons such as safety, prevention of vapor loss, etc. Enclosure may also be used to prevent contamination of the material being filtered or to confine the spray from washing nozzles. The rotary drum filter also can be enclosed in a pressure vessel and operated under pressure.

Disadvantages of the bottom feed arrangement can be overcome by using top-feed drum filters, which use the nearly horizontal surface on the top of the drum. The area available for filtration is small, however, and such filters are reserved for fast settling solids that dewater readily, and applications that permit precoating. Top-feed arrangements are common in the brewing industry. Drying may also be carried out in top-feed filters in totally enclosed systems.

12.8. Rotary Vacuum Disk Filters. An alternative to the drum filter is the disk filter, which uses a number of disks mounted vertically on a horizontal shaft and suspended in a slurry reservoir. This arrangement provides a greater surface area for a given floor space, by as much as a factor of 4, but cake washing is more difficult and cloth washing virtually impossible.

As in the case of drum filters, each disk is divided into a number of separate segmental sectors, normally 12 but sometimes up to 30, that have suitable drainage and cloth support, and the sectors are connected to a filter valve similar to the type used in drum filters. It is necessary to operate disk filters at high submergence because a sector must be completely submerged during cake formation. There is also a top limit to the liquid level to allow the scraper to discharge the cake under gravity. The higher submergence, 40% or even higher (up to 55%), reduces the area available for dewatering.

In conventional disk filters, cake discharge is usually performed by a scraper blade, for cakes thicker than 10 mm, or sometimes by a tapered roll; air blowback is often used to assist the discharge. High-pressure sprays also have been used for cake discharge.

As with drum filters, the slurry reservoir has to be agitated to prevent settling. Rather than using one large vat for all the disks, some designs provide individual low volume, closely fitting vats for each disk, thus avoiding the need for an agitator.

Speed of rotation is relatively high, from 20 to as much as 180 rpm. The disk size varies from 0.5 to 5.3 m, with up to 20 disks assembled on one shaft, providing filtration areas up to 380 m^2. Disk vacuum filters are principally used in mining and metallurgical applications for handling large volumes of free-filtering materials. They have also been successfully operated with cement, starch, sugar, paper and pulp, and flue dusts.

13. Batch Pressure Filters

Excluding variable chamber presses, which rely on mechanical squeezing of the cake and are discussed in a separate section, pressure filters may be grouped into two categories, ie, plate-and-frame filter presses, and pressure vessels containing filter elements. The latter group also includes cartridge filters; these are discussed separately. All of the above pressure filters are suited to handling different types of cake. Pressure vessel filters (leaf-type) handle incompressible or slightly compressible cakes. Filter presses handle both compressible and incompressible cakes, especially with the flexibility potential of membranes. Cylindrical element filters, ie, candle filters, are used for clarification applications, using membrane socks (or tubes), precoat and often body-feed, resulting in cakes that are slightly compressible. Cartridge filters are for clarification only, with little if any cake formed.

13.1. Plate-and-Frame Filter Presses. In the conventional plate-and-frame press (Fig. 20), a sequence of perforated square, or rectangular, plates alternating with hollow frames is mounted on suitable supports and pressed together with hydraulic or screw-driven rams. The plates are covered with a

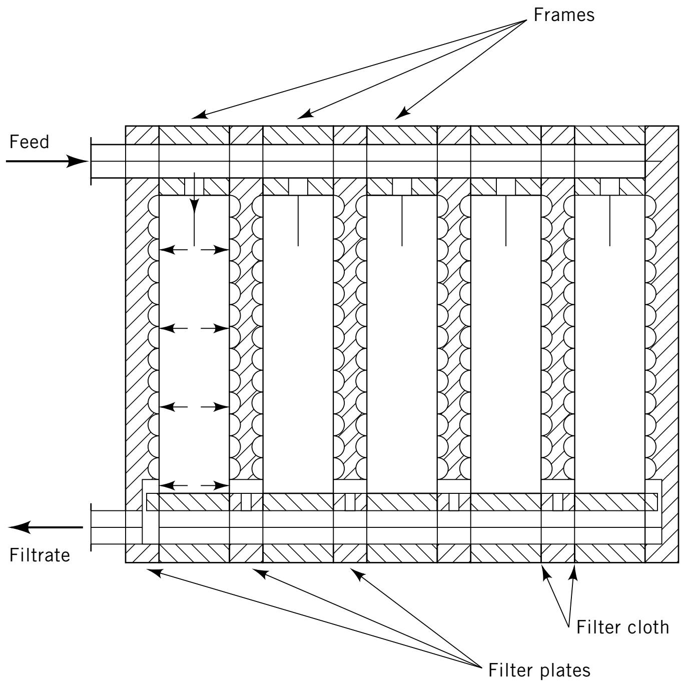


Fig. 20. Principle of plate-and-frame presses.

filter cloth that also forms the sealing gasket. The slurry is pumped into the frames and the filtrate is drained from the plates.

The drainage surfaces are usually made in the form of raised cylinders, square-shaped pyramids, or parallel grooves in materials such as stainless steel, cast iron, rubber of coated metal, polypropylene, rubber, or wood. Designs are available with every conceivable combination of inlet and outlet location, ie, top feed, center feed, bottom feed, corner feed, bottom external feed, and side feed, with a similar profusion of possible positions of discharge points. Each combination has particular advantages, depending on whether washing is required and also on the application and nature of the suspension. The discharge may be through a separate cock on each plate, rather than through a common filtrate port and manifold, to allow observation and sampling of the filtrate from each plate. This enables the operator to spot cloth failure and isolate the plate or segregate the cloudy filtrate. The cocks discharge into open channels or enclosed pipe systems fitted with a sight glass.

Both flush plates and recessed plates can be specified. Recessed plates obviate the need for the frames but are tougher on filter cloths due to the strain around the edges. These presses are more suitable for automation because of the difficulty of the automatic removal of residual cake from the frames in a plate-and-frame press.

Plate sizes range from 150 mm to 2.4 m^2, giving filtration areas up to $\sim$800 m^2. The number of chambers varies up to 120 with exceptions to 200.

Plate-and-frame filters are most versatile since their effective area can be varied simply by blanking off some of the plates. Cake holding capacity can be altered by changing the frame thickness or by grouping several frames together. These filters are available in a variety of semiautomated or fully automated versions that feature mechanical leaf-moving devices, cake removal by vibration, or cake removal by pulling the cloth when the press is open, etc. An operator usually must be present, however, because it is not certain that each and every chamber will discharge its cake unaided every time. Should manual intervention be necessary, the operator must be protected by a suitable safety photoelectric device from injury by the plate-moving mechanism, or by the closing mechanism.

Washing performed in filter presses is either simple or thorough washing. In simple washing, the wash liquid is introduced either through the main feed port or through a separate port into each chamber, and the washing is therefore in the same direction as the filtration process that formed the cake. In thorough washing, the wash liquid enters through a separate port, behind the filter cloth on every other plate, thus passing through the whole thickness of the cake in the chamber. Washing is less efficient with recessed chamber presses than with flush plate frame presses, because of poorer distribution of the wash liquid. In either case the amount of wash liquid needed is high.

Filter media for plate-and-frame presses include various cloths, mats, and paper. Paper filter media usually must be provided with a backing cloth for support.

The typical operating pressure of filter presses is 7–15 bar, although some manufacturers offer presses for 30 bar or higher. As the pressure increases during filtration, it forces the plates apart; this may be offset by a pressure compensation facility offered with some large mechanized presses.

Full mechanization of filter presses started in the late 1950s; this was followed by addition of the mechanical expression, ie, cake squeezing, mechanism. Rubber or plastic membranes are sometimes fitted to compress the cake that is formed by conventional pressure filtration. The membranes normally rest on the plates and have grooves and openings in their surfaces for filtrate collection. They are inflated at the end of the filtration cycle by air, for pressures to $\sim$7 bar, or by water at higher pressures. The membranes should be designed to last up to or in excess of 20,000 cycles. The principal advantages of using mechanical expression of cakes are the additional dewatering usually achieved, the ability to handle thin cakes, and superior cake washing. The main filtration process can be done at lower pressures so that a relatively cheap pump can be used, and the compression by the membrane then goes to higher pressures (107,108).

The automation of filter presses has affected several other advantages and developments. Plate shifting mechanisms have been developed, allowing the cloths to be vibrated; filter cloth washing, on both sides, has been incorporated to counteract clogging and downtimes have been reduced with automation, thus increasing capacities.

The vertical recessed plate automatic press (or tower press) is shown schematically in Figure 21. Unlike the conventional filter press with plates hanging down and linked in a horizontal direction, this filter press has the plates in a horizontal plane placed one upon another. This design offers semicontinuous

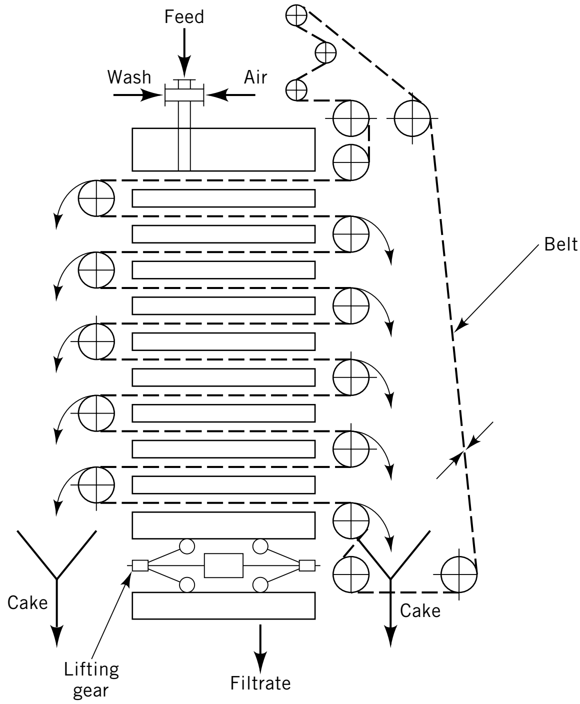


Fig. 21. Vertical automatic filter presses (or tower press).

operation, saving in floor space, and easy cleaning of the cloth, but it allows only the lower face of each chamber to be used for filtration (although the new Hoesch DS tower press allows for double sided filtration (109).

The filter usually has an endless cloth, traveling intermittently between the plates via rollers, to peel off cakes. Unfortunately, if the cloth is damaged anywhere, the whole cloth must be replaced, which is a difficult process. Each time the filter cloth zigzags through the filter, the filtering direction is reversed; this tends to keep the cloth clean. Most of these filters incorporate membranes for mechanical expression, and cakes sometimes stick to the membranes and remain in the chamber after discharge. Because the height of the vertical press makes maintenance difficult, the number of chambers is restricted, usually to 20, with a maximum of 40, with filter areas up to 32 m^2. However, newer designs (eg, Schneider's unit and filtra-Systems' Verti-Press (74,76) offer increased sizes and simpler designs). Some vertical filters (usually called automatic pressure filters or APFs) are available with a separate cloth for each frame. The cloth may be disposable and such filters are designed to operate with or without filter aids (76): and are primarily used in the machine tool coolant and wastewater industries.

The application of filter presses spans virtually all areas of the processing industries due to their versatility. Examples of use include clarification of beer and juices; wastewater and activated sludge filtration in breweries, paper mills, and petrochemical plants; dewatering of fine minerals; lime mud separation; and washing in the sugar industry. Filtration rates are usually from 0.025 to 1 $m^3/m^2/h$. Dry solids handling capacities are <1000 $kg/m^2/h$, with the higher values being more usual with the automatic presses due to their shorter down times.

13.2. Pressure Vessel Filters. The several designs of pressure vessel filters all consist of pressure vessels housing a multitude of leaves or other elements that form the filtration surface and that are mounted either horizontally or vertically. With horizontal leaves most suitable where thorough washing is required, there is no danger of the cake falling off the cloth; with vertical elements, a pressure drop must be maintained across the element to retain the cake. The disadvantage of horizontal leaf types is that one-half the filtration area is lost because the underside of the leaf is not used for filtration because of the danger of the cake falling off. Discharge of the cake also may be more difficult in this case.

The elements or leaves normally consist of a coarse stainless steel mesh over which a fine, often woven metal, mesh or filter cloth is stretched and sealed at the edges. The leaves are in parallel, each connected to a header and, almost without exception, filtration is from the outside in through the medium. These filters are essentially batch operated, and most require the use of a filter aid for precoating to avoid cloudy filtrates and blinding. A separate filtering element is often installed at the bottom of the vessel as a scavenger filter for the filtration of the heel of unfiltered slurry that is still in the vessel at the end of the filtration period. This residual slurry is particularly troublesome with the vertical leaf filters because compressed air cannot be used to complete the filtration of this heel, as the air would preferentially escape through the tops of the leaves as soon as they emerged from the suspension. During the scavenge filtration the main leaves are usually isolated so that compressed air is not lost through them.

Cylindrical Element Filters. These filters, often referred to as candle filters, have cylindrical elements or sleeves mounted vertically and suspended from a header sheet, which divides the filter vessel into two separate compartments (Fig. 22). The filtration takes place on the outside of the sleeves, but in some designs, filtration takes place on the inside generally used for semidry cake discharge (74). The inlet is usually in the bottom section of the vessel and the filtrate outlet in the top section above the header (or tube) sheet.

The tubes generally measure from 25 to 75 mm in diameter, and up to 2.5 m in length. Made from metal or cloth-covered metal or thin membrane-type socks over various tube designs, they provide filtration areas up to 100 m^2. Alternatively, the tubes can be made of stoneware, plastics, sintered metal, or ceramics. The elements may be deliberately made flexible, and tank diameters up to 1.5 m are available. Cake removal is performed by scraping with hydraulically operated scraper rings, by vibration, by turbulent flow bumping, or by backwashing. The mechanical strength of the tubular element makes it ideal for cleaning by the sudden application of reverse pressure. Physical expansion or flexing of the tubular elements on application of the reverse flow aids cake discharge. These filters find wider use where cake washing is not required.

The advantage of candle filters is that as the cake grows on the outside of the tubular elements the filtration area increases and the thickness of a given volume of cake is therefore less than it would be on a flat element. This is of

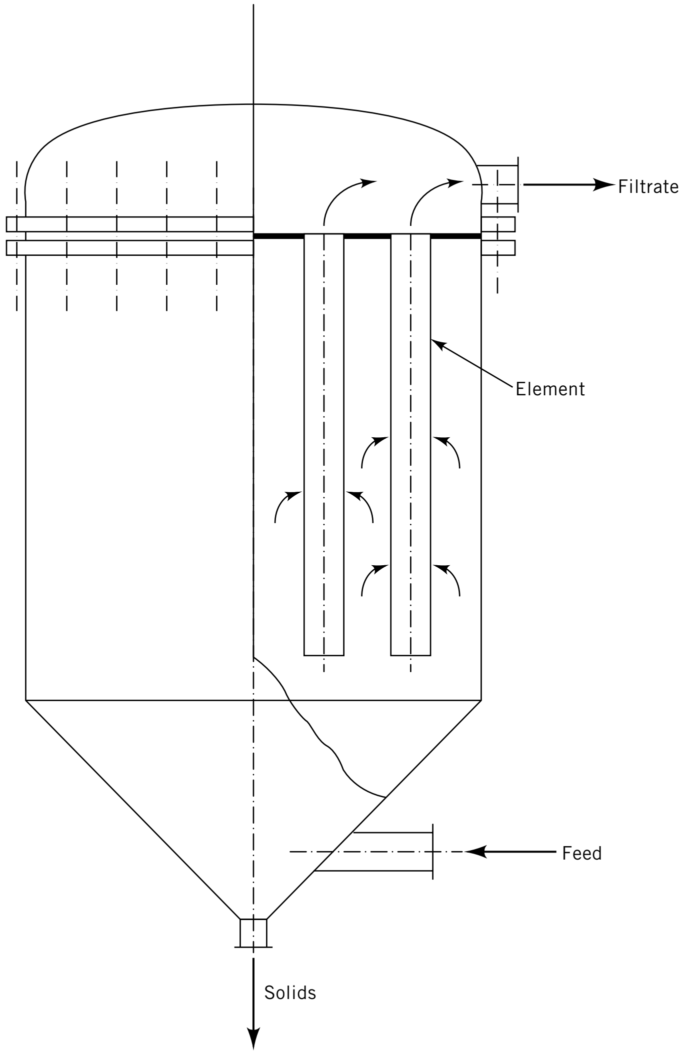


Fig. 22. A cylindrical element (candle) pressure filter.

importance where a thick cake is being formed; the rate of increase in the pressure drop is less with tubular elements.

A new type of candle filter (Amafilter's Cricketfilter, Holland) has been introduced that have flattened candles that reportedly permit more elements per unit volume and allow for better backwashing or backpulsing (110).

An even more unique candle filter (DrM's Fundabae, Switzerland) uses multiple small tubes for enhanced cake removal from their usual woven cloth socks (111). Both of these designs can provide for dry cake discharge.

The pressure filter with tubular elements has also been used as a thickener, when the cake, backwashed by intermittent reverse flow, is redispersed by an agitator at the bottom of the vessel and discharged continuously as a slurry. In

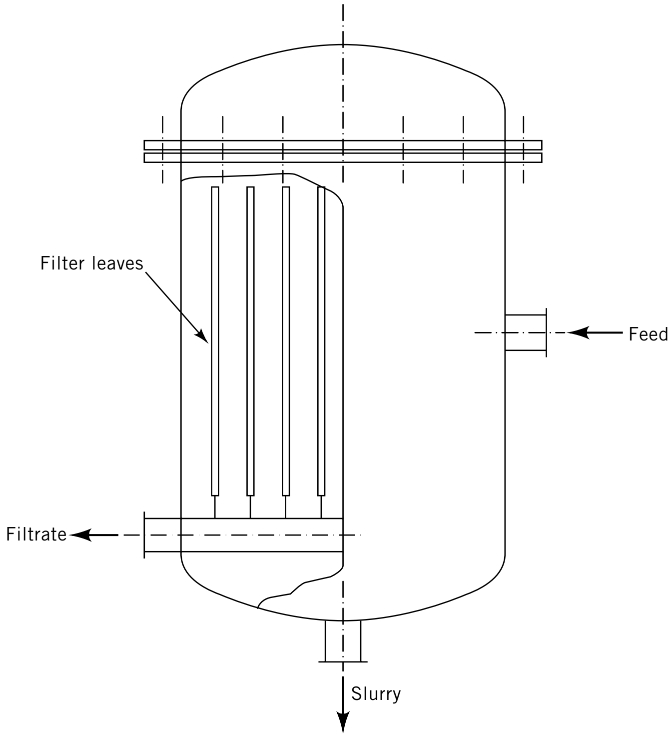


Fig. 23. Vertical vessel, vertical leaf filter.

some cases, the filter cake builds up to a critical thickness and then falls away without blowback.

Vertical Vessel, Vertical Leaf Filters. These are the cheapest of the pressure leaf filters and have the lowest volume/area ratio (Fig. 23). Their filtration areas are limited to $<80\ m^2$. Large bottom outlets, fitted with rapid-opening doors, are used for dry cake discharge, and smaller openings are used for slurry discharge. Wet discharge may be promoted by spray pipes, vibrators, reverse flow, bubble rings, scrapers, etc, while dry discharge is usually caused by vibration. As with all vertical leaf filters, these are not suited for cake washing.

In the Scheibler filter, the filter leaves take the form of bags, each suspended in a rectangular pressure vessel from a horizontal tube that acts as the filtrate outlet. The sides of the bag are prevented from meeting by looped chains that are attached at the top of the loop to the horizontal tube and hang downward. With this method of separating the cloth surfaces by chains, the bags can be wide enough to hang in pleats and the filtering area can thus be as much as three times the area of the frames inside the bags. Filtration areas up to $250\ m^2$ are available with applications being mostly in the chemical industry. An improved version of this Scheibler is now offered by Larox.

Horizontal Vessel, Vertical Leaf Filters. In a cylindrical vessel with a horizontal axis (Fig. 24), the vertical leaves can be arranged either laterally or longitudinally. The latter, less common, arrangement may be designed as the

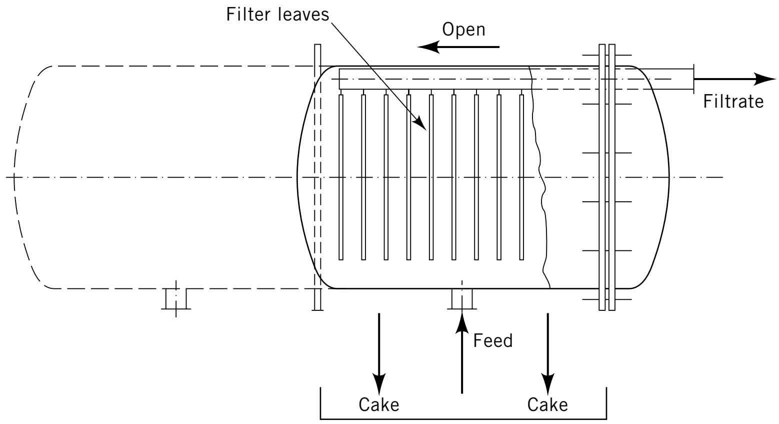


Fig. 24. Horizontal vessel, vertical leaf filter.

vertical vessel, vertical leaf filters but mounted horizontally. Its design is suitable for smaller duties and the leaves can be withdrawn individually through the opening end of the vessel.

Filtration areas up to 120 m^2 are available with the Kelly leaf filter, another longitudinal arrangement and probably the earliest pressure leaf filter. It has been used for the filtration of very viscous liquids such as glycerol and concentrated sugar solutions. The height of the leaves varies according to the space available at their location in the vessel. The leaves are attached to the removable circular front cover, with each leaf having an outlet connection through this cover. The leaves, together with the cover and outlet pipes, are attached to a carriage that can be run into and out of the shell to facilitate cake discharge outside the vessel by air blowback or rapping. This arrangement requires a considerable amount of floor space and to minimize this drawback the filters are often constructed in pairs on a single runway, with their opening ends facing each other. Thus each filter can be opened in turn into the common space between them.

Horizontal vessel filters that have the vertical leaves arranged in a plane perpendicular to the axis of symmetry of the vessel, ie, laterally, have the greatest use because they provide easy access to the leaves. Most of these designs open in a way similar to the Kelly filter. Some designs move the shell rather than the leaf assembly so that the filtrate pipe can remain permanently connected; withdrawal of one leaf or a bundle of leaves at a time is possible. The leaves may be rectangular, circular, or of some other shape, and may be designed to rotate during cake discharge. Sluicing by sprays is used for wet discharge, with or without rotation of the leaves. If the leaves are designed for rotation, they are invariably circular and mounted on a central hollow shaft that serve as the filtrate outlet. Dry cake discharge may be carried out with rotating leaves by application of a scraper blade. If this is to be done without opening the vessel, then the bottom of the vessel must be shaped as a hopper, with a screw conveyor if necessary.

The Vallez filter, originally developed in the United States for the sugar industry, rotates the leaves at $\sim$1 rpm during the filtration operation to keep the solids in suspension and achieve a more uniform cake.

The Sweetland filter, a significant departure from the standard end-opening design, has the cylindrical shell split in a horizontal plane into two parts, where the bottom one-half can be swung open for cake discharge. The upper one-half is rigidly supported and both the feed and the filtrate piping are fixed to it. The lower part is hinged to the upper along one side and is counterbalanced for easy opening. Cake discharge is either by sluicing or by dropping, assisted by some scraping. If much scraping is needed, there is not much advantage in using this type of filter. Because the leaves are stationary, the cake deposited on them may be uneven, with greater mass of cake at the bottom of the leaves.

Generally, the horizontal vessel, vertical filters with leaves arranged laterally can be designed up to filtration areas of 300 m^2. Cake washing is possible but must be carried out with caution since there is a danger of the cake falling off.

Horizontal vessel filters with vertical rotating elements have been under rapid development with the aim of making truly continuous pressure filters, particularly for the filtration of fine coal.

Vertical Vessel, Horizontal Leaf Filters. These filters, like all horizontal leaf filters, are advantageous where the flow is intermittent or where thorough cake washing is required. Filtration areas are limited to $\sim$45 m^2.

The pressure versions of the nutsche filter, which falls into this category, are either simple pressurized filter boxes or more sophisticated agitated nutsches, much the same in design as the enclosed agitated vacuum filters described earlier. These are extremely versatile, batch-operated filters, used in many industries, eg, agrochemistry, pharmaceuticals, or dyestuff and foodstuff production.

An obvious method of increasing the filtration area in the vessel is to stack several plates on top of each other; the plates are operated in parallel. One design, known as the plate filter, uses circular plates and a stack that can be removed as one assembly. This allows the stack to be replaced after the filtration period with a clean stack, and the filter can be put back into operation quickly. The filter consists of dimpled plates supporting perforated plates on which filter cloth or paper is placed. The space between the dimpled plates and the cloth is connected to the filtrate outlet, which is either into the hollow shaft or into the vessel, the other being used for the feed. When the feed is into the vessel, a scavenger plate may have to be fitted because the vessel will be full of unfiltered slurry at the end of the filtration period. This type of filter is available with filtration areas up to 25 m^2 and cakes up to 50 mm thick.

Centrifugal discharge filters form another group in this category. As the name suggests, the cake discharge is accomplished by rotating the stack of plates around the hollow shaft. The cake slides off the plates due to the centrifugal action; sometimes it is necessary to supplement this by sluicing with a suitable liquid, in which case the discharge is wet. The filtrate leaves through the hollow shaft. These filters lend themselves to automation and, as opposed to manually operated leaf filters, they can be operated with short cycle times and very short downtimes, which is economical. Many different designs are available, with various ways of driving the shaft and locations of the electric motor as well as other varying constructional details. Sizes vary up to 65 m^2.

Another available design allows discharge of the cake by vibration of the circular plates, which are slightly conical, sloping downward toward the outside

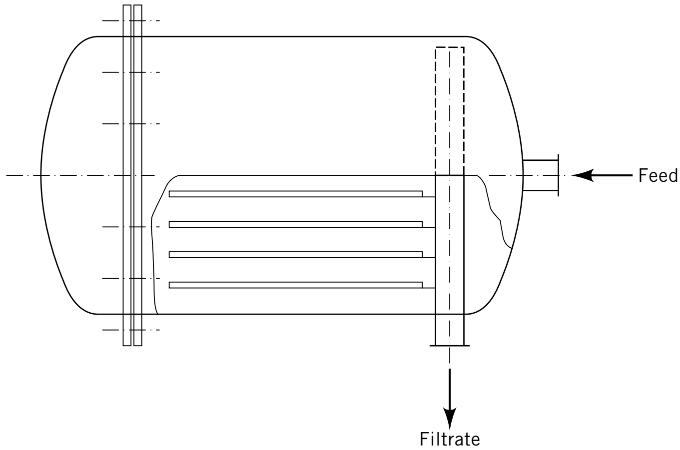


Fig. 25. Horizontal vessel, horizontal leaf filter.

of the plates. This design allows higher pressures to be used as there are no rotating seals necessary.

Horizontal Vessel, Horizontal Leaf Filters. These filters consist of a horizontal cylindrical vessel with an opening at one end (Fig. 25). A stack of rectangular horizontal trays is mounted inside the vessel; the trays can usually be withdrawn for cake discharge, either individually or in the whole assembly. The latter case requires a suitable carriage. One alternative design allows the tray assembly to be rotated through 90° so that the cake can fall off into the bottom part, designed in the shape of a hopper and fitted with a screw conveyor.

The trays may be fitted with rims; this is particularly useful for flooding the trays in washing operations. Scavenger leaves are often used. Filtration areas up to 50 m^2 are available. Like all horizontal leaf filters, horizontal vessel, horizontal leaf filters are particularly suitable when thorough washing is needed.

13.3. Cartridge Filters. Cartridge filters use easily replaceable, tubular cartridges made of paper, sintered metal, woven cloth, needle felts, activated carbon, or various membranes of pore size down to ~0.1 μm. Filtration normally takes place in the direction radially inward, through the outer face of the element, into the hollow core (Fig. 26). Cartridge filtration is limited to liquid polishing or clarification, ie, removing very small amounts of solids, in order to keep the frequency of cartridge replacements down. Typically, suspensions of <0.01% vol concentration of solids can be treated with cartridge filters, and such filters are favored in small-scale manufacturing applications.

Cartridge filters are either depth or surface type, according to where most of the solids separate; the precise demarcation line is difficult to assess. The most common depth cartridge is the yarn-wound type that has a yarn wound around a center core in such a way that the openings closet to the core are smaller than those on the outside. The aim is to achieve depth filtration, which increases the solids-holding capacity of the cartridge. The yarn may be made of any fibrous material, ranging from cotton or glass fiber to the many synthetic fibers such as polypropylene, polyester, nylon, or Teflon. The spun staple fibers are brushed

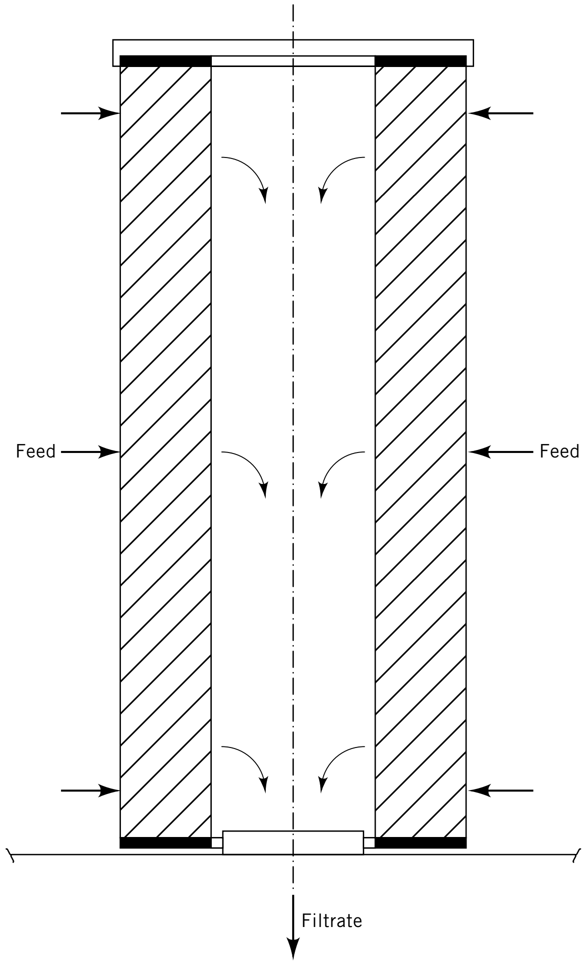


Fig. 26. The principle of cartridge filters.

to raise the nap and this makes the filter medium. The cores are made from polypropylene, phenolic resin, stainless steel, or other metals or alloys. The nominal mm rating of this type of cartridge varies from 0.5 to 100 μm, but are generally only ∼70% efficient. On the other hand, they are the most economical.

Cartridges for higher viscosity liquids are often made of long, loose fibers, again either natural or synthetic, impregnated with phenolic resin. Such bonded cartridges are usually formed into the shape of a thick tube by a filtration technique and do not require a core because they are self-supporting. The porosity of the medium can be graded during the formation process, again to increase the solids-holding capacity. Bonded cartridges are available in somewhat coarser ratings from 10 to 75 μm.

Depth-type cartridges cannot be cleaned but have high solids-holding capacity, and are cheap and robust. Considerable standardization of the cartridge size throughout industry, ∼25, 50, 75, and 100 cm long, 6.3–7.0-cm overall diameter, and 2.5-cm internal diameter, allows testing of different cartridge makes and types in the same cartridge housing.

A common surface cartridge is the pleated paper construction type, which allows larger filtration areas to be packed into a small space. Oil filters in the automobile industry are of this type. The paper is impregnated, for strength, with epoxy or polyurethane resin. Any other medium in sheet form, similar to cellulose paper, such as polypropylene, glass, or a variety of nonwovens, may be used.

The nominal rating ranges from 0.2 to 50 μm. Pleating in the radial direction is usual, but some cartridges have axial pleating, or, in an effort to pack as much filtration surface into a given space as possible, have hollow disks of lenticular shape; up to 3 m^2 of surface in one cartridge is possible. The solids-holding capacity of pleated paper surface cartridges is low but some applications allow the prolonged buildup of solids on the surface, until the pleats are completely filled up.

Another type of cartridge is the edge filter that contains a number of thin disks mounted on a central core and compressed together. The disks are usually made of metal although paper or plastics are used. Filtration takes place on the surface of the cylinder, with the particles unable to pass between the disks. The principal advantage of the edge-type filter is that it is cleanable. This is done by reverse flow, ultrasonics, oscillation, or by scraping the outside surface of the cylinder with a mechanical scraper. Edge filters made from paper disks have been known to retain particles as fine as 1 μm but the metal variety retains solids >50 μm or so. Other designs of cartridges use active carbon, Fuller's earth, sintered metal, or other specialized media.

The most important characteristics of cartridge filters are the filter rating, ie, the largest particle ($\sim$98% nominal and 99.98% absolute retention cut size) that passes through the filter; the relationship between the pressure drop and solids-holding capacity; and the maximum allowable pressure drop beyond which the cartridge fails structurally. Both the retention and the solids-holding capacity depend on filtration velocity and this must be considered when testing cartridges. Thermal or shock stresses can lead to cracking of cartridges, with the subsequent loss of filtrate clarity. Standardized tests are now available to test and validate most cartridges.

The housings of cartridge filters are simple pressure vessels designed for one cartridge, or a number of cartridges in parallel, in multielement filters. Some housings are designed to withstand pressures up to 300 bar. Proper sealing of the elements is a necessary prerequisite of their efficient use. Frequent replacement of cartridges should be facilitated by quick-opening clamping fittings.

Cartridge filters are used to clean power fluids, lubrication oils, DI water, wines, fruit juices, or pharmaceutical liquids. They are also used to protect other equipment, eg, in control systems or automatic valves. Low capital and installation costs, low maintenance costs, simplicity, and compactness are the main advantages of cartridge filters. Running costs are high, especially when disposable cartridges are used. It is most important, therefore, that a full economical analysis, based on reliable cartridge replacement frequency, is carried out before adopting a cartridge filtration system; the low cost of the basic hardware may be deceptive.

13.4. Mechanical Batch Compression Filters. In conventional cake filtration the liquid is expelled from the slurry by fluid pressure in a fixed-volume filtration chamber; in mechanical compression this is achieved by reduction of the volume of the retaining chamber. This compression of either a slurry or a cake, which might have been formed by conventional filtration, offers advantages to industries handling a variety of different materials. Such materials include highly compressible, sponge-like solids; very fine particles such as clays; fibrous pulps; gelatinous mixtures like starch residues or some pharmaceuticals; and flocculated wastewater sludges.

The compressibility of filter cakes is a nuisance from the point of view of the filtration theory. In practice, it means that with increasing pressure cakes become more compact and therefore drier and more difficult to filter. The resistance to flow increases due to reduced porosities, however, and, with some materials, eg, paper mill effluents and municipal wastewater sludges; higher pressures do not necessarily give increased flow rates. In cakes undergoing conventional pressure filtration, the bottom layers closest to the medium are subjected to the highest compression forces whereas the top layers are subjected only to light hydraulic forces and are not compacted so tightly. If a mechanical force is applied to the top of the filter cake, the distribution of pressure through the cake is more uniform. Cakes drier than those formed by using high pumping pressures of the feed suspension can thus be achieved.

Since 1980 (75,107,108), a number of new filters have appeared on the market, utilizing some form of mechanical compression of the filter cake, either after a conventional pressure filtration process or as a substitute for it. In most designs, the compression is achieved by inflating a diaphragm that presses the slurry or the freshly formed filter cake toward the medium, thus squeezing an additional amount of liquid out of the cake.

Other designs squeeze the cake between two permeable belts or between a screw conveyor of diminishing diameter, or pitch, and its permeable enclosure. The available filters that use mechanical compression can be classified into four principal categories, ie, membrane plate presses, tube presses, belt presses, and screw presses.

Membrane plate and tube presses are dealt with here; belt and screw presses are included in the discussion of continuous pressure filters.

The advantages of using mechanical compression with compressible cakes include increased solid content of the cakes, leading to reduced energy requirements if thermal drying has to follow, or to better handling properties; improved washing efficiencies; increased filtration rates; and easier or automatic cake discharge. Invariably, however, the capital cost of such filters is higher than for conventional pressure filters. Whenever a membrane is used for the compression, however, this increases the capital cost and thus, variable chamber filters tend to be more expensive than conventional filters in the same category.

Membrane Plate Presses. Membrane presses are closely related to conventional plate and frame presses. They consist of a recessed plate press in which the plates are covered with an inflatable diaphragm that has a drainage pattern molded into its outside surface. The filter cloth is placed over the diaphragm (Fig. 27). During the first stage of filling the filter chambers with the slurry and conventional pressure filtration, the membrane is pushed against the plate body.

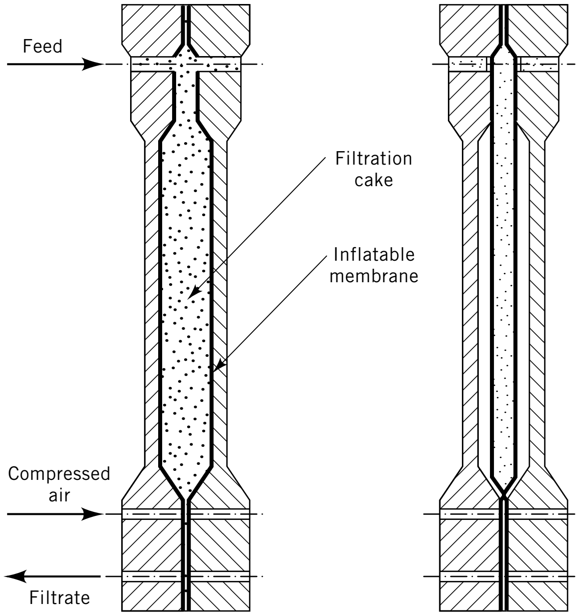


Fig. 27. Membrane plate press, ie, recessed plate with variable chamber press.

When the chambers are relatively full of cake, the feed is terminated and the membranes are inflated by pumping compressed air or hydraulic fluid in between the membranes and the plates; the cake in the chamber is compacted as the membranes expand. Washing of the cake may follow and can be carried out more effectively than in a normal press because the cake is compacted to a more uniform density by squeezing. The resulting advantage is in the reduction of the washing time and washwater requirements. The squeeze pressures vary from 6 to 30 bar. Additional reductions of up to 25% in the moisture content over that obtained with conventional filter presses can be achieved. The membranes are usually made of rubber compounds, polypropylene or (PVDF), which are resistant to solvents.

Another advantage of the membrane plate is its flexibility to cake thickness, ie, thinner cakes can be easily handled with increased cake dryness. Cake release characteristics are also improved by deflation of the membranes prior to cake discharge. Alternating arrangements, in which the membrane plates and the normal recessed plates alternate, have been used to reduce cost.

The plate press filter replaces the pneumatically operated membranes with flexible seals and compression by a hydraulically powered ram. The cake is formed in the chambers between the hollow circular frames carrying the filter medium and the flat rectangular discharge plates. The frames are sealed against the discharge plates by rubber rim seals which form the filtration chambers. As the rim seals are compressed by the hydraulic ram, the cake inside is squeezed. The filter is then opened, the cake adheres to the discharge plates and, as the plates are lowered and raised again, the cake is removed by scrapers.

The process is fully automated and the filtration areas available go up to 20 m^2. No washing of the cake is possible.

The OMD leaf filter (Stella Meta Filters and Industrial Filter) is a vertical leaf filter with a rubber diaphragm suspended between the leaves. The cake that forms on the leaves eventually reaches the diaphragm at which point pump pressure is used to inflate the diaphragm and compress the cake. The cake discharge is by vibration.

A variation of the same principle is the DDS-vacuum pressure filter, which has a number of small disks mounted on a shaft that rotates discontinuously. The cake is formed on both sides of the disks when they are at the bottom position, dipped into the slurry. When the disks come out of the slurry and reach the top position, hydraulically driven pistons squeeze the cake and the extra liquid then drains from both sides of the cake. The cake is removed by blowback with compressed air.

Cake compression by flexible membranes is also used in the new automated vertical presses that use one or two endless cloth belts, indexing between plates [ie, Hoesch, Larox tower presses and Filtra-Systems' Verti-Press (74)].

Filtration and compression take place with the press closed and the belt stationary; the press is then opened to allow movement of the belt for cake discharge over a discharge roller of a small diameter. This allows washing of the belt on both sides (Fig. 21). Cycle times are short, typically between 10 and 30 min, and the operation is fully automated. Sizes up to 32 m^2 are available and the maximum cake thickness is 40 mm.

Washing and dewatering by air displacement of cakes are possible. Applications are in the treatment of minerals, in the sugar industry, and in the treatment of municipal sewage sludge and fillers like talc, clay, calcium carbonate, etc.

The versatility of these batch membrane presses has been further enhanced in the past few years by combination with *in situ* heating and vacuum drying so the entire dewatering–drying process is conducted in one piece of equipment. Some municipal sewage sludges are dried via this process (112). Four vendors are actively pursuing this technology: JWI (USFilter), DryVac, Bertrams, and Eurofilter.

Cylindrical Presses. Another group of filters that utilize the variable chamber principle are those with a cylindrical filter surface. There are two designs in this category, both of which originate from the United Kingdom.

The VC filter (Fig. 28) consists of two concentric hollow cylinders mounted horizontally on a central shaft. The inner cylinder is perforated and carries the filter cloth, the outer cylinder is lined on the inside with an inflatable diaphragm. The slurry enters into the annulus between the cylinders and conventional pressure filtration takes place, with the cake forming on the outer surface of the inner cylinder. The filtration can be stopped at any cake thickness or resistance, as required by the economics of the process, and hydraulic pressure is then applied to the diaphragm that compresses the cake.

As with other filters of this type, washing can be carried out by deflation of the diaphragm and introduction of washwater into the annulus. Reinflation of the membrane then forces the wash liquor through the cake, thus displacing the mother liquor. At the end of the process, the inner cylinder is withdrawn

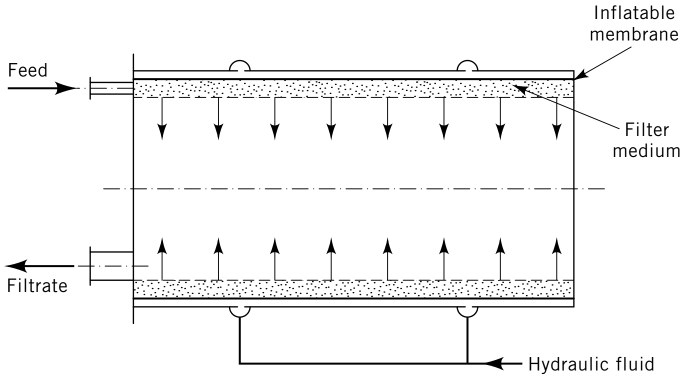


Fig. 28. Principle of the VC filter.

from the outer shell and the cake is either discharged manually or blown off with compressed air. Sizes available range from small, mobile test units of 0.4-m^2 area to large, fully automated machines with 6.1-m^2 filtration area. Choice of two alternative core sizes is offered, giving annuli of 6 or 2.8 cm available for the cake. The hydraulic pressure for operating the membrane goes up to 14 bar. Although originally developed for filtration of dyestuffs, the VC filter has been successfully used for the filtration of gypsum, china clay, cement, industrial effluents, metal oxides, coal washings, nuclear waste, and other slurries.

The ECLP tube press is smaller in diameter and, unlike the VC filter, is operated in a vertical position. It uses compression only, both for the filtration and for squeezing the cake. The space between the cylindrical rubber membrane and the cloth tube is first filled with the slurry and the hydraulically operated membrane is used to drive the liquid through the cloth. It follows, therefore, that this filter is suitable for higher solids concentrations, usually in excess of 10% by weight, in order to obtain the minimum cake thickness necessary for efficient cake discharge of $\sim$4 mm. At the end of the process, the central core is lowered by $\sim$300 mm and the cake is removed by a blast of compressed air from the inside. The hydraulic operating pressures are higher than the VC filter at $\sim$150 bar but the single tube area is only $\sim$1.3 m^2. Multiple tube assemblies are used to treat larger flows. Cake washing is possible but with some solids there is a danger of the cake falling off the inner core while the annulus is being filled with water.

The ECLP tube press was originally developed for the filtration of china clay but has been used with many other slurries such as those in mining, TiO_2, cement, sewage sludge, etc. The typical cycle time is $\sim$4 min or more.

14. Continuous Pressure Filters

A continuous pressure filter may be defined as a filter that operates at pressure drops >1 bar and does not require interruption of its operation to discharge the cake; the cake discharge itself, however, does not have to be continuous. There is little or no downtime involved, and the dry solids rates can sometimes be as high as 1750 kg/m^{2}h with continuous pressure filters.

Most continuous pressure filters available have their roots in vacuum filtration technology. A rotary drum or rotary disk vacuum filter can be adapted to pressure by enclosing it in a pressure cover; however, the disadvantages of this measure are evident. The enclosure is a pressure vessel which is heavy and expensive, the progress of filtration cannot be watched, and the removal of the cake from the vessel is difficult. Other complications of this method are caused by the necessity of arranging for two or more differential pressures between the inside and outside of the filter, which requires a troublesome system of pressure regulating valves.

Despite the disadvantages, the advantages of high throughputs and low moisture contents in the filtration cakes have justified the vigorous development of continuous pressure filters.

Horizontal or vertical vessel filters, especially those with vertical rotating elements, have undergone rapid development with the aim of making truly continuous pressure filters, particularly but not exclusively for the filtration of fine coal. There are basically three categories of continuous pressure filters available, ie, disk filters, drum filters, and belt filters including both hydraulic and compression varieties.

The advantages of continuous pressure filtration are clear and indisputable, particularly with slow-settling slurries and fairly incompressible cakes. Such filters are expensive, both to install and to run, and the most likely applications are either in large-scale processing of products that require thermal drying after the filtration stage, ie, fine coal or cement slurries in the dry process, or in small-scale processing of high value product such as in the pharmaceutical industry.

14.1. Disk Filters. *The McGaskell and Gaudfrin Disk Filters.* One of the earliest machines in this category, the McGaskell rotary pressure filter, is essentially a disk-type filter enclosed in a pressure vessel. The rotating disks are each composed of several wedge-shaped elements connected to a rotary filter valve at the end of the shaft, similar to the vacuum rotary disk filters. Originally designed for the filtration of waxes in the oil industry and equipped for a gradual increase in pressure with cake buildup, the McGaskell rotary pressure filter is said to have produced high filtrate clarity. Pressures up to 7 bar have been used.

The slurry reservoir is divided into pockets or crenellations that have spring-loaded scrapers. The scrapers press against the disks and direct the cake into the spaces between the pockets around the disks that lead to a chute connected to an inner casing in which is placed a worm gear. The worm conveys and compresses the cake, thus squeezing more liquid from it, through a filter cloth. The compressed cake forms a plug around a spring-loaded, tapered discharge valve, and the plug prevents leakage of gas. The cake can be washed but not very effectively. This filter is reported in filtration literature (1,2), but is no longer commonly used.

The Gaudfrin disk filter, designed for the sugar industry and available in France since 1959, is also similar in design to a vacuum disk filter but it is enclosed in a pressure vessel with a removable lid. The disks are 2.6 m in diameter, composed of 16 sectors. The cake discharge is by air blowback, assisted by scrapers if necessary, into a chute where it may be either reslurried and pumped out of the vessel or, for pasty materials, pumped away with a monopump without reslurrying.

The Gaudfrin disk filter is designed for only relatively low pressures of 1 bar on average and it provides for cake washing in two stages, in two separate compartments within the same vessel.

The KDF Filter. The KDF filter (Fig. 29) (Amafilter, Holland) is based on the same principle as disk filters. It was developed for the treatment of mineral raw materials, like coal flotation concentrates or cement slurries, and can produce a filter cake of low moisture content at very high capacities, up to 1750 kg/m^{2}h. The pressure gradient is produced by pressurized air above the slurry level that provides the necessary driving force for the filtration and also is used for displacement dewatering of the cake.

Assemblies of small disks are rotated in a planetary movement around a central screw conveyor. The disks are mounted on six hollow axles and the axles revolve on overhanging bearings from the gearbox at one end of the vessel where they are driven, via a drive shaft, by an electric motor. The filtrate is collected from the disks via the hollow shafts and a filter valve into a large collecting pipe. The hollow shafts also collect the water and air from the dewatering process, in another part of the rotational cycle. The number of disks mounted on the shafts can be adjusted for different materials, depending on the required capacity and the cake thickness to be used.

As the vessel is only about one-half filled with slurry, the disks become coated with the cake when immersed, the cake is dewatered when the disks emerge from the slurry, and scraped or blown off, by reverse blow, into the central conveyor that takes the cake to one end of the vessel. The planetary action and the slow movement of the disks through the feed slurry ensure exceptionally good homogeneity of the cake that is critically important for good dewatering

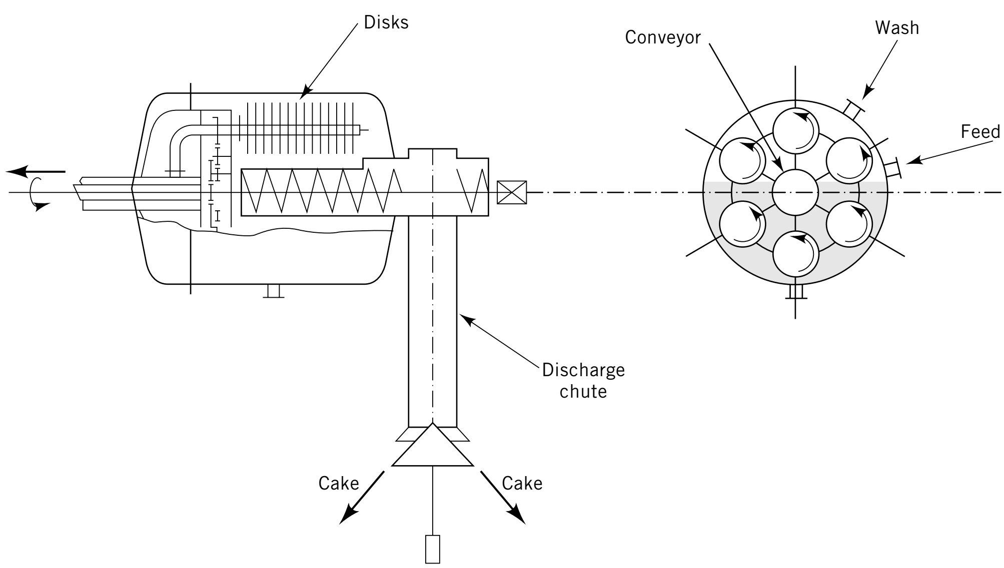


Fig. 29. KDF continuous pressure filter.

characteristics; the typical speed of rotation of the planetary system of shafts is from 0.8 to 1 rpm.

A screw conveyor was used originally to convey the cake, but this has been replaced with a chain-type conveyor. The first prototype used a tapered screw to form a plug before discharge into atmospheric pressure; this has been replaced by compaction in a vertical pipe.

The cake discharge is initiated and stopped by two level indicators inside the vertical pipe. The cake is actually discharged using the pressure inside the vessel; a specially designed, hydraulically operated discharge valve momentarily opens and the cake shoots out. The air pressure used for driving the slurry through the filter is 6 bar and filtration areas are available up to 120 m^2. Cake washing is possible but it has not been reported as actually performed.

The KDF filter was first tested in prototype on a coal mine in northern Germany. It was installed in parallel with existing vacuum filters and it produced filter cakes consistently lower in moisture content by 5–7% than the vacuum filters. Two production models have been installed and operated on a coal mine in Belgium. The filter is controlled by a specially developed computer system; this consists of two computers, one monitoring the function of the filter and all of the detection devices installed, and the other controlling the filtration process. The system allows optimization of the performance, automatic start-up or shut-down, and can be integrated into the control system of the whole coal washing plant.

The KHD Pressure Filter. Another development of the disk filter has been reported (KHD Humboldt Wedag AG, Germany). A somewhat different system, probably a predecessor, was patented (113).

The patented system (113) has stationary disks mounted inside a pressure vessel (horizontal vessel, vertical disks) that is mounted on rollers and can rotate slowly about its axis. A screw conveyor is mounted in the stationary center of rotation; it conveys the cake, which is blown off the leaves when they pass above the screw, to one end of the vessel where it falls into a vertical chute. The cake discharge system involves two linear slide valves that slide the cake through compartments that gradually depressurize it and move it out of the vessel without any significant loss of pressure. The system relies entirely on the cake falling freely from one compartment to another as the valves move across. This may be an unrealistic assumption, particularly with sticky cakes; when combined with lots of sliding contact surfaces that are prone to abrasion and jamming, the practicality of the system is questionable.

Another significant disadvantage of the patented process is the two large running seals involved in the main body of the filter as the vessel rotates around a stationary central arrangement; this seal is another potential source of trouble. This version has little chance of commercial success and has been shelved in favor of a more conventional system of stationary vessel (114).

The newer version, tested with coal slurries in a pilot-plant facility, and with a 90-m^2 version in production, has the rotating disks and all the driving elements inside a stationary vessel. The disks, according to the manufacturer's literature, range from 1300 to 3000 mm in diameter, with up to 10 in one vessel, giving filtration areas up to 480 m^2. The cake discharge is through a rotary lock discharger that has cylindrical compartments rotating around a vertical axis.

Sliding surfaces are involved and the cake is assumed to be nonsticky so that it will fall out when the compartment opens to the atmosphere. The filtration area can be varied by changing the size and number of disks.

The test results reported show the advantages of pressure filtration quite clearly, ie, the dry cake production capacity obtained with the test solids (coal suspensions) was raised 60 or 70% and the final moisture content of the cake reduced by as much as 5–7% by increasing the pressure drop from 0.6 to 2 bar. Further increases in the operating pressure bring about less and less return in terms of capacity and moisture content.

The operating pressure is kept low to reduce air consumption. To obtain high capacities, the disks spin fairly fast (up to 2 or 3 rpm) and this leads to thin cakes that give little resistance to air flow; the only way to keep this flow within economical limits is to reduce the air pressure in the vessel.

14.2. Drum Filters. The rotary drum filter, also borrowed from vacuum filtration, makes relatively poor use of the space available in the pressure vessel, and the filtration areas and capacities of such filters cannot possibly match those of the disk pressure filters. In spite of this disadvantage, however, the pressure drum filter has been extensively developed.

The drum may be mounted in a vertical rather than horizontal vessel, and pressure is created by pumping compressed gas into the vessel. The intake of the compressor may be connected to the filtrate side of the filter and the gas goes around in a closed circuit. The method used to discharge the cake continuously from the high pressure inside the vessel into the atmosphere is the real basis of the design. Variable pitch screws, star valves, alternating or serial decompression chambers, monopumps (for pasty and thixotropic cakes), and other similar devices have been tried with varying degrees of success, depending on the properties of the cake. Another, rather obvious, alternative is the use of two storage vessels into which the cake is alternately discharged at the same pressure as in the filter, the pressure is later released, and the cake discharged from the vessel under atmospheric pressure.

A test unit of a small drum filter has been developed, total filter area of 0.7 m^2 with 30% submergence, housed in a large horizontal pressure vessel. Several interesting concepts have been developed and tested based on this model (115).

The so-called hyperbar vacuum filtration is a combination of vacuum and pressure filtration in a pull–push arrangement, whereby a vacuum pump of a fan generates vacuum downstream of the filter medium, while a compressor maintains higher than atmospheric pressure upstream. If, eg, the vacuum produced is 0.8 bar, ie, absolute pressure of 0.2 bar, and the absolute pressure before the filter is 1.5 bar, the total pressure drop of 1.3 bar is created across the filter medium. This is a new idea in principle but in practice requires three primary movers: a liquid pump to pump in the suspension, a vacuum pump to produce the vacuum, and a compressor to supply the compressed air. The cost of having to provide, install, and maintain one additional primary mover has deterred the development of hyperbar vacuum filtration; only Andritz in Austria offers a system commercially.

TDF Drum Filter. This is a fairly conventional drum filter housed in a vertical pressure vessel. Test data, obtained with the smallest model of only 0.7-m^2

filtration area, is available (116). Larger models have also been announced, ranging up to the filtration area of 46 m^2 and very large vessels. The operating pressures are moderate, up to 0.35 bar, and the drum speeds fairly conventional from 0.3 to 1.5 rpm. The range of dry cake production quoted is from 250 to 650 kg/m^2h for fine coal.

The cake is scraped off with a conventional knife arrangement, then conveyed in a screw conveyor to one end of the vessel where it enters the discharge system. There are four design alternatives depending on the cake to be processed: tapered rotary valve with a horizontal axis; a pump, presumably a monopump; a rotary valve, vertical axis, with a blow-through, similar to the one used in pneumatic conveying; and a vertical pipe compactor similar in design to the Fuller-Kinyon pump in pneumatic conveying.

A plate-type filter, the PDF filter (116), uses a paddle wheel with radial, longitudinal plates covered with filter cloth and manifolded to the filter valve at one end of the vessel, instead of a drum. This filter uses a horizontal pressure vessel, was built to have only 0.75- or 1.5-m^2 area, and operates at 0.25 bar. A central screw conveyor collects the cake blown off the plates and conveys it to the discharge end of the vessel.

The BHS-Fest Filter. A different approach to the use of a drum for pressure filtration is made in the BHS-Fest Filter (Fig. 30). This permits a separate treatment of each filter section, in which the pressure may vary from vacuum to a positive pressure; pressure regulation is much less difficult than in the conventional enclosed drum-type pressure filter.

The BHS-Fest pressure filter has a rotating drum divided into sections. The separating strips project above the filter cloth and thus form cells. The drum is almost completely surrounded by an outer shell and the space between the

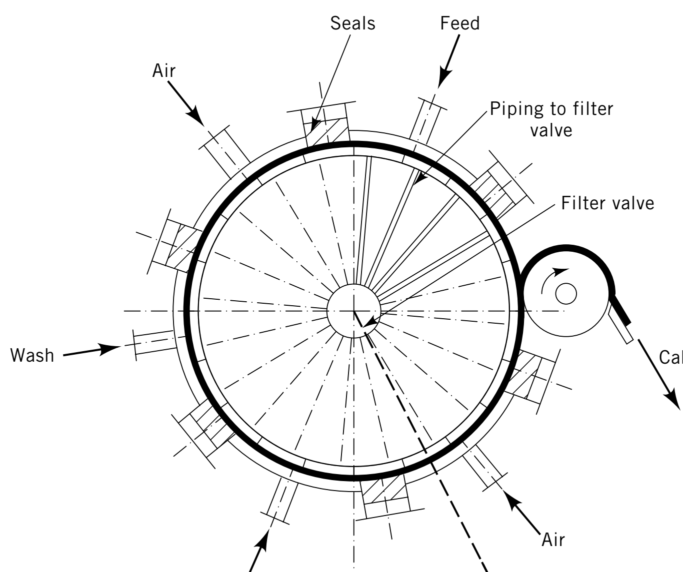


Fig. 30. BHS-Fest pressure filter.

shell and the drum is divided into a number of compartments separated by seals under adjustable pressure. As the drum rotates, each cell on the drum passes progressively through the series of compartments, thereby undergoing different processes such as cake formation, dewatering, cake washing, or cake drying; these can be carried out in several stages under different pressures or even under vacuum.

Cake discharge occurs at atmospheric pressure by the action of a roll or a scraper, assisted by blowback. The cloth may be washed by a spray before the cycle starts again. Filtering areas range up to 8 m^2 and drum diameters up to 2 m. The necessity for large seals limits the operating pressure typically to <3 bar. Cake thickness can be from 2 to 150 mm, depending on machine size, and the speed of drum rotation up to 2 rpm, usually from 0.3 to 1 rpm. Applications occur in the manufacture of pharmaceuticals, dyestuffs, edible oils, and various chemicals and minerals.

14.3. Horizontal Belt Pressure Filters. Horizontal belt filters have a great advantage in cake washing application due to their horizontal filtration surface. In the context of pressure filtration and the requirements of good dewatering, however, they have a significant disadvantage because the cake is not very homogeneous; gravity settling on the belt and the inevitable problems of distribution of the feed suspension over the belt width cause particle stratification and nonhomogeneous cakes.

The Flat-bed pressure filter [or automatic pressure filter, APF- (74,76)] is based on the above principle. The pressure compartment consists of two halves, top, and bottom. The bottom half is stationary while the top half can be raised to allow the belt and the cake to pass out of the compartment, and can be lowered into the belt during the filtration and dewatering stage. The filter can be considered as a horizontal filter press with an indexing cloth; in comparison with a conventional filter press, however, this filter allows only the lower face of the chamber to be used for filtration. There are a number of designs available (76).

The same idea has been described except it does not lift the top half of the pressure compartment but opens and closes end seals for the belt to pass through (74). This filter is proposed for dewatering of fine coal and mineral slurries in particular.

The vertical recessed plate automatic press, shown schematically in Figure 21 and described previously, is another example of a horizontal belt pressure filter. Cycle times are short, typically between 10 and 30 min, and the operation is fully automated. The maximum cake thickness is $\sim$40 mm; washing and dewatering (by air displacement) of cakes is possible. Applications include treatment of mineral slurries, sugar, sewage sludge, and fillers like talc, clay, and calcium carbonate.

14.4. Continuous Compression Filters. The variable chamber principle applied to batch filtration, as described before, can also be used continuously in belt presses and screw presses.

Belt Presses. Belt filter presses combine gravity drainage with mechanical squeezing of the cake between two running belts. The Manor Tower press (Fig. 31), a Swiss invention also available in the United Kingdom, consists of two acutely angled vertically converging filter belts running together downward. The shallow funnel formed between the belts, and the ends sealed by a special

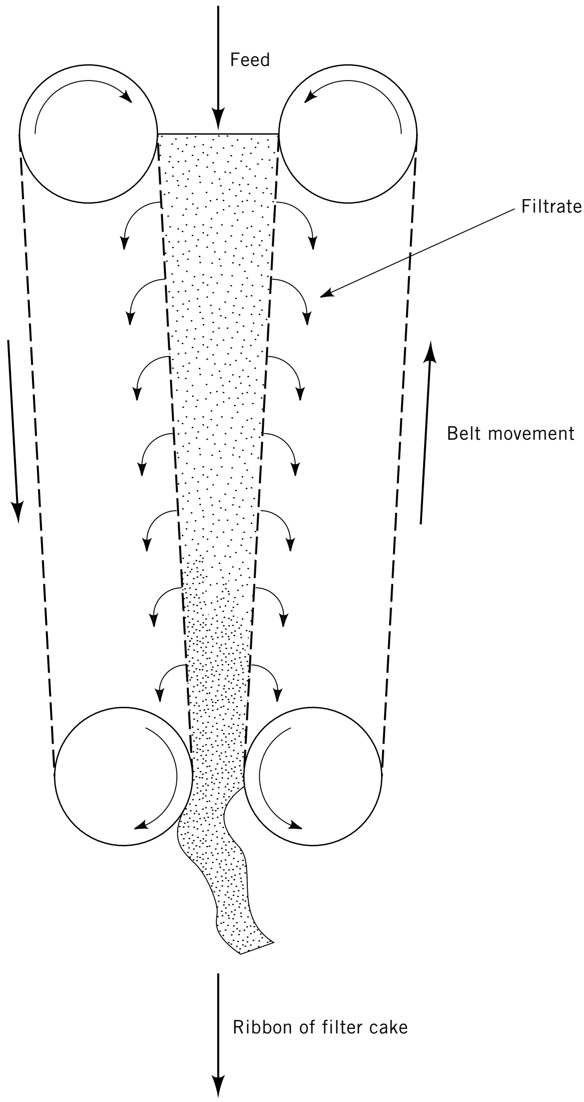


Fig. 31. Manor tower press.

edge-sealing belt, is filled from the top with the slurry to be filtered. The slurry moves with the belts down the vertical narrowing gap where the filtration takes place. The hydrostatic pressure causes the solids to be deposited on the faces of the belts until the two cakes combine to form a continuous ribbon of filter cake that is squeezed in the closing gap at a pressure of 2.5 bar, gauge pressure, and discharged at the bottom. To protect the press against overload, the final gap at the bottom is automatically controlled so the above mentioned pressure is not exceeded.

Originally designed for the continuous filtration of conditioned sewage sludges, as were most of the filter belt presses available, the Manor Tower press is increasingly used for the treatment of paper mill sludge, coal, or flocculated clay slurries.

There are probably >20 designs of horizontal belt filter presses available, most developed during the 1980s. They owe their existence to the availability of cationic polyelectrolytes that promote the release of water from organic sewage sludges, ie, the area where belt filter presses are most used.

The flocculated feed is first introduced onto the horizontal drainage section where the free water is removed by gravity. Sometimes a system of ploughs may be employed to turn the forming cake and allow any free water on the cake surface to drain through the belt mesh. The sludge then is sandwiched between the carrying belt and the cover belt and compression dewatering takes place; liberated water passes through the belts. The third zone, the shear zone, shears the cake by flexing it in opposite directions during passage through a train of rollers in a meander arrangement, to produce drier cake. To prevent the released water from being absorbed back into the cake on the cake release, scrapers or wipers may be installed to remove the water from the outside of the belts.

Belt filter presses are made up to a width of 2.5 m and produce a final solids concentration of the discharge sludge in the range of 12–60%, depending on the type of sludge.

Screw Presses. Another way of achieving compression of the cake is by squeezing in a screw press. This is suitable only for the dewatering of rough organic materials, pastes, sludges, or similar materials, because it does not include a filtration stage. The material is conveyed by a screw inside a perforated cage. The volume available continuously diminishes, either by reducing the pitch of the screw in a cylindrical cage or by reducing the diameter of the screw, in which case the cage is conical, as shown in Figure 32. The cage is either perforated or constructed from longitudinal bars in a split casing. The solids discharge is controlled by a suitable throttling device that controls the operating pressure. Washing or dilution liquid can be injected at points along the length of the cage. The power requirements are generally large. Newer versions are available that are reported to improve performance with sewage sludges at reduced power consumption (117).

Typical applications include dewatering of sugar beet pulp, fish meal, distillers and brewers spent grains, starch residues, fruit, potato starch by-products, grass, maize, leaves, and similar materials.

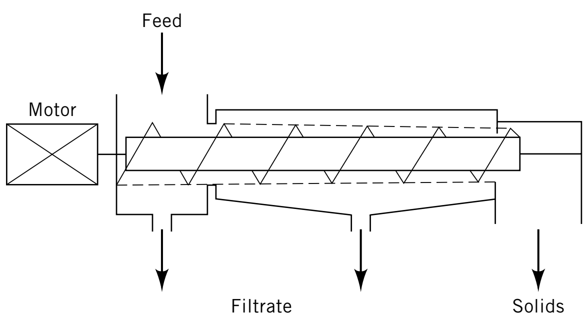


Fig. 32. The principle of a screw press.

14.5. Thickening Pressure Filters. The most important disadvantage of conventional cake filtration is the declining rate due to the increased pressure drop caused by the growth of the cake on the filter medium. A high flow rate of liquid through the medium can be maintained if little or no cake is allowed to form on the medium. This leads to thickening of the slurry on the upstream part of the medium; filters based on this principle are sometimes called filter thickeners.

The methods of limiting cake growth are classified into five groups, ie, removal of cake by mass forces (gravity or centrifugal), or by electrophoretic forces tangential to or away from the filter medium; mechanical removal of the cake by brushes, liquid jets, or scrapers; dislodging of the cake by intermittent reverse flow; prevention of cake deposition by vibration; and cross-flow filtration by moving the slurry tengentially to the filter medium so that the cake is continuously sheared off. The extent of the commercial exploitation of these principles in the available equipment varies, but cross-flow filtration is exploited most often.

Mechanical Cake Removal. This method is used in the American version of the dynamic filter described under cross-flow filtration with rotating elements, where turbine-type rotors are used to limit the cake thickness at low speeds. The Exxflow filter, introduced in the United Kingdom, is described in more detail under cross-flow filtration in porous pipes. It uses, among other means, a roller cleaning system that periodically rolls over a curtain of flexible pipes and dislodges any cake on the inside of the pipes. The cake is then flushed out of the curtain by the internal flow.

Dislodging of Cake by Reverse Flow. Intermittent back-flushing of the filter medium can also be used to control cake growth, leading to filtration through thin cakes in short cycles. Conventional vacuum or pressure filters can be modified to counter the effects of the forces during the back-flush (118,119).

A filter based on continuous backwash by spray has been announced (120); the feed is introduced tangentially into a cylindrical vessel, which has a cylindrical screen in the center. The filtrate leaves through the screen and the solids deposit on it. A rotary spray system mounted on the inside of the screen cylinder goes around continuously at 115 rpm and sprays water radially outward, thus removing the solids that fall down and are removed from the filter through the solids outlet. Full backwashing may become necessary in addition to this spray cleaning and provision is made for this. The medium is made of polypropylene or nylon, available down to 20-μm pore size. Flow rates up to 68 m^3/h can be treated at feed concentrations up to 0.2%. There is a minimum underflow discharge rate of 1.1 m^3/h leading to relatively dilute underflows.

Prevention of Cake Deposition by Vibration. Vibration provides another method for preventing the formation of a dense filter cake. However, commercial exploitation of this principle is uncommon. A notable exception is the vibrating slurry filter (121). Each filter unit consists of three 38-mm diameter tubular screens enclosed in a cylindrical housing and made from wedge-shaped wires. The screen openings available go down to 44 μm (325 mesh). The tubular filter elements are continuously vibrated pneumatically at high frequency and low amplitude. This keeps the oversize particles in suspension and reduces premature screen blinding. The solids are removed by intermittent backwash. Two or

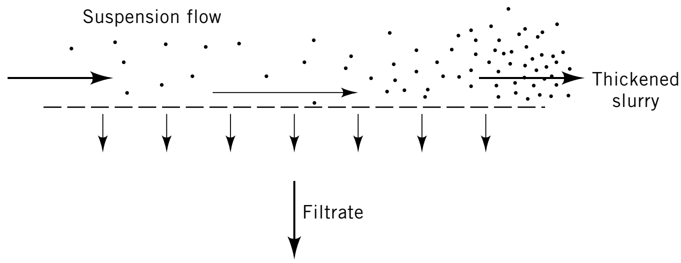


Fig. 33. Schematic diagram of cross-flow filtration.

more filters can be manifolded in parallel to common headers to provide higher flows. In such cases, the service can be continuous, as individual filters can be backflushed while all other filters remain on-stream. The filters are designed for pressures up to 1 bar, the areas available are multiples of 0.3 m^2, and flow rates of 8 m^3/h can be handled by a single unit. Applications include clarification of paper coatings, colloidal gels, ceramic slips, or calcium carbonate slurries.

Cross-Flow Filtration. In conventional filtration, the flow of the suspension is perpendicular, ie, dead end, to the surface of the filter medium, with all the liquid passing through it, except for the small amount of moisture retained in the cake. In cross-flow filtration, the slurry flow is tangential to the filtration medium (Fig. 33) at high relative velocities with respect to the medium. The shear forces in the flow close to the medium continuously remove a part or the whole of the cake and mix it with the remaining suspension. The filtration velocity through the medium, due to the pressure drop across it, is small relative to the velocity of the flow parallel with the medium, typically by three orders of magnitude. There is little tendency for any colloidal and particulate matter to accumulate in the boundary layer. In most cases, however, a layer of fines exists within the boundary layer, leading to the so-called dynamic membrane effect which limits the performance.

As more and more of the filtrate is removed, the slurry gradually thickens and may become thixotropic. The solids content of the thickened slurry may be higher than that obtained with conventional pressure filtration, by as much as 10 or 20%. A range of velocity gradients from 70 to 500 L/s has been suggested as necessary to prevent cake formation and to keep the thickening slurry in a fluid state.

Cross-Flow Filtration with Rotating Elements. The first patented method of limiting cake growth by this means was a cylindrical, rotating filter element, mounted in a cylindrical pressure vessel. The suspension was then pumped into the annulus between the rotor and the vessel (118). More recently, virtually all possible combinations and arrangements for such dynamic filters have been patented. Figure 34 gives an example of one such arrangement. It comprises a pressure vessel with a hollow filter leaf inside, connected to and rotating about a hollow shaft. The slurry is continuously fed at a constant rate into the vessel, and the filtrate passes through the rotating filter medium and out through the hollow shaft. The thickened slurry is discharged via a control valve used to maintain optimum slurry concentration.

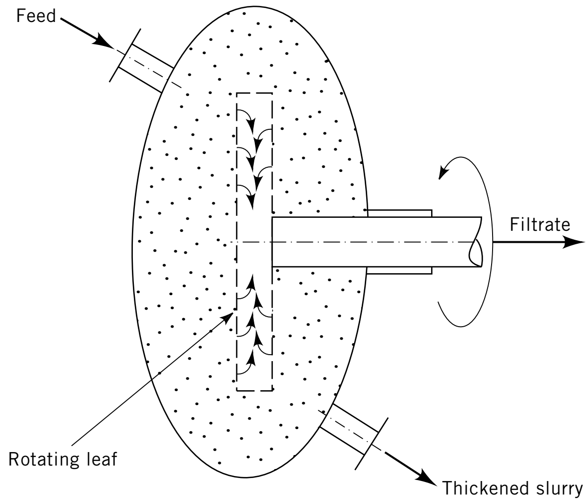


Fig. 34. Kaspar's dynamic filter.

As observed from Figure 34, the cake removal by fluid shear is also aided by centrifugal force. Other arrangements include stationary filtration media and rotating disks to create the shear effects, and rotating cylindrical elements; it has also been shown how such filters can be used for cake washing.

Since its conception, the dynamic filter has been widely reported and further developed. Most European designs are comprised of a multistage disk arrangement (Fig. 35) with both the rotating and stationary elements covered with filter cloth, thus utilizing the space inside the pressure vessel. Such filters have been found (122) to be from 5 to 25 times more productive in mass of dry cake per unit area and time than filter presses for the same moisture content of the final slurry. In some cases, the moisture content with the dynamic filter

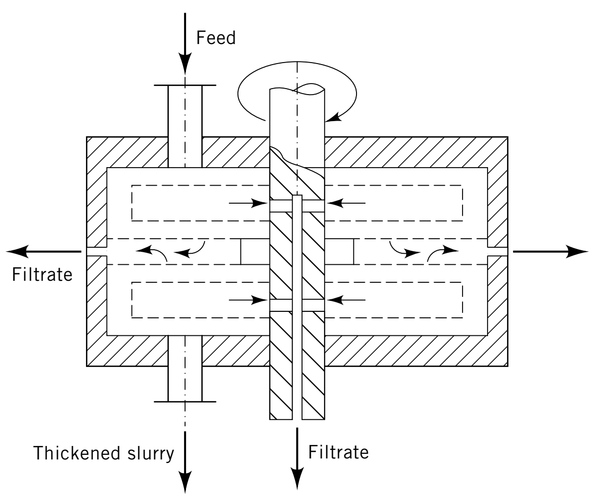


Fig. 35. European dynamic filter.

was actually lower than with a filter press. The maximum productivity was achieved with peripheral disk speeds from 2.8 to 4.5 m/s.

The idea of axial filtration with a cylindrical rotating element is taken a step further in the Escher-Wyss pressure filter (123). In addition to the cylindrical rotating filter surface, it provides a stationary outer filter surface (Fig. 36) so that the suspension, passing through the narrow annulus between the two cylinders, filters through the whole of the wetted surface. Strategically placed

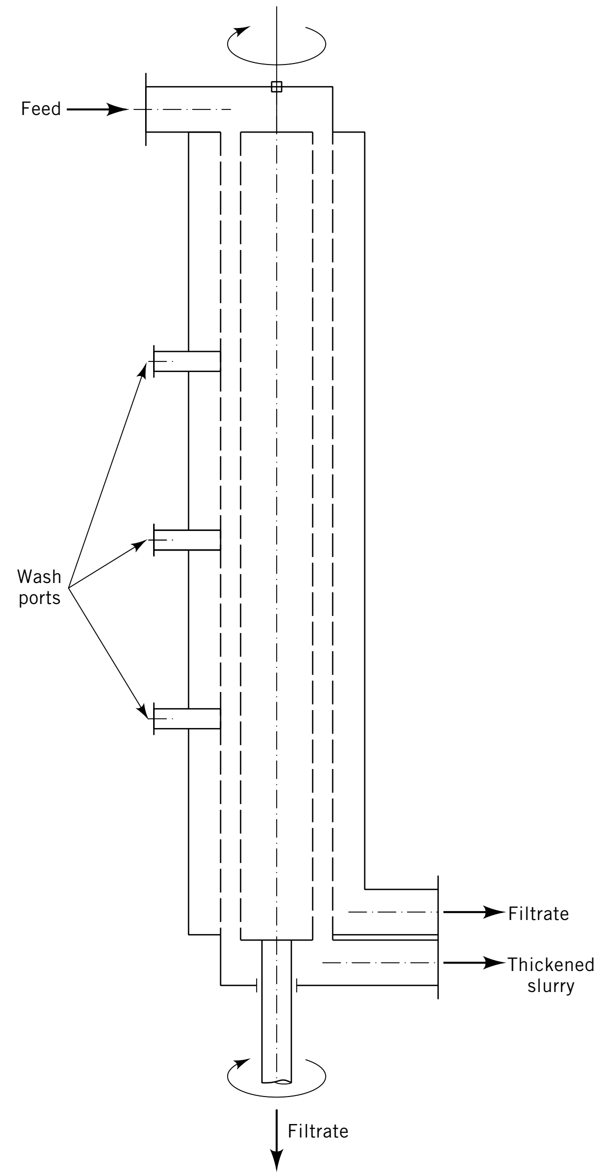


Fig. 36. Escher-Wyss filter.

washwater entry points along the filter length allow continuous washing, equivalent to reslurrying. This filter is particularly suitable for viscous, plastic, or thixotropic slurries.

All the dynamic filters that use rotating filter elements are subject to several disadvantages. First, the filter medium, usually cloth, is stretched by the centrifugal force, and this imposes special requirements on its fastening and stretch resistance. Second, the filtrate is collected into the element radially inward, ie, against the centrifugal force. This reduces filtration velocities, and the necessity for discharge through a hollow shaft makes the filter more complicated. Finally, any solids that may penetrate through the medium might accumulate inside the rotating elements due to centrifugal settling, with the accompanying danger of partial reduction in the area of the cloth available for filtration.

The three disadvantages described can be avoided by using solid elements, instead of permeable ones, which create the shear to prevent or reduce cake formation. Only the stationary surface inside the filter is then available for filtration and this means a reduction in capacity. This is not a problem because the solid disks can be slimmer and the collection of filtrate does not have to be through a hollow shaft.

The American version of the dynamic filter, known as the Artisan continuous filter (Fig. 37), uses such nonfiltering rotors in the form of turbine-type elements. The cylindrical vessel is divided into a series of disk-type compartments, each housing one rotor, and the stationary surfaces are covered with filter cloth. The feed is pumped in at one end of the vessel, forced to pass through the compartments in series, and discharged as a thick paste at the other end. At low rotor speeds the cake thickness is controlled by the clearance between the scraper and the filter medium on the stationary plate, while at higher speeds part of the cake is swept away and only a thin layer remains and acts as the actual medium.

Results of test work with this filter, producing cakes of 1 mm thickness using a 3-mm clearance, have been published (124,125). The cake formed on the medium was generally stable, giving high filtration rates over long periods

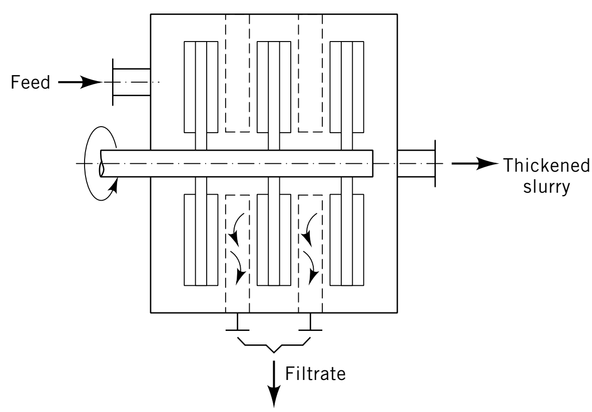


Fig. 37. Artisan continuous filter.

of time, and the precoat-type cake did not blind with time. There was no evidence of any size selectivity of the process; the only exception was conventional filter aids that were preferentially picked up by the rotating fluid. This was attributed to inertial by-passing of the cake by larger particles and reentrainment of the smaller ones. Good control of the agitator speed and operating pressure is recommended in order to maintain a balance between the overall filtration and slurry flow rates so that the critical mud concentration and excessive torque are avoided.

The Artisan continuous filter operates at pressures from 1.5 to 2.0 bar, it allows washing by injection of wash liquid after the initial filtering stages, and the sizes available range from 0.3 to 19 m^2. Abrasive wear on the cloth is claimed to be eliminated by the thin protective layer of cake on the medium.

Because the energy going into the agitation converts mostly into heat, the extruded cake may be quite hot, from 50 to 80°C in some cases. High torque is needed to drive the agitators and most, if not all, of the energy saved by reducing the cake thickness may go back into the agitation. Little information exists about the actual power requirements but a favorable comparison has been made with a solid bowl centrifuge operated with the same calcium carbonate slurry (126). It is indisputable, however, that higher productivity is obtained per unit area of the filter surface than with conventional pressure filters, ie, the solids dry cake yield varies from 20 to 1000 kg/m^2h for slurries such as pigments, calcium carbonate, magnesium hydroxide, and kaolin. Other materials handled by this filter include dyestuffs, polymer slurries, clays, pharmaceuticals, and metal oxides.

Other tests with the American design of the dynamic filter have been reported (127) in which the cake was eliminated completely by spinning the disks faster. No visible abrasion of the filter medium was observed while the filtration velocities could be maintained high, even at lower pressures <0.5 bar. The work showed clearly the effect of speed and number of vanes on the rotors, on filtration velocities, and specific energy input. The specific energy input in kWh/m^3 of the filtrate is shown to increase with rotor speed for any number of rotor vanes. This increase can be minimized by optimization of the number of vanes. This does not justify the use of high speeds in terms of running costs but a smaller filter can be used for the same capacity. An improved version (Bokela's Dyno filter) has recently been introduced (128).

Cross-Flow Filtration in Porous Pipes. Another way of limiting cake growth is to pump the slurry through porous pipes at high velocities of the order of thousands of times the filtration velocity through the walls of the pipes. This is in direct analogy with the now well-established process of ultrafiltration, which itself borders on nanofiltration and reverse osmosis at the molecular level. The three processes are closely related yet different in many respects.

The idea of ultrafiltration has been extended in recent years to the filtration of particles in the micrometer and submicrometer range in porous pipes, using the same cross-flow principle. In order to prevent blocking, thicker flow channels are necessary, almost exclusively in the form of tubes. The process is often called cross-flow microfiltration but the term cross-flow filtration is used here.

The use of various porous tubes for the cross-flow filtration of hydrated alumina and red bauxite was first reported in 1964 (129). The tubes used were ceramic (of 30–40% porosity) and titanium, at specific outputs ranging from

0.3 to 8 m/h. Dead-end filtration was compared with cross-flow, the influence of the slurry flow rate and pressure on the filtrate rate was studied, and a multi-pass model filter resembling a heat exchanger was developed. Mayer (89) also describes the first theoretical and empirical formulas for the total length of tubes needed for a specific application, from known average specific output of filtrate, for the filtration velocity from geometric and physical parameters in dimensionless groups, and for the degree of thickening and the variation in the solids concentration along the tube length.

Results of an investigation into the thickening of kaolin clay in woven fiber polyester hoses was published (130). The results achieved production rates from 0.5 to 1.7 m/h and found temperature and slurry velocity to be the primary variables affecting the filtration velocity. The filtration velocity was proportional to the circulation velocity. The slurry velocities required for particle transport have been studied (131) and the separation of particles using porous stainless steel tubes examined. The relationship between filtration velocity and the slurry velocity was found to be linear, for magnesium carbonate in this case, but not going through the origin. The two constants in the linear equation were found to depend on particle size. Chakrabarti and co-workers (131) also suggests the use of precoating techniques for cross-flow filtration in tubes.

15. Centrifugal Filters

The driving force for filtration in centrifugal filters is centrifugal forces acting on the fluid. Such filters essentially consist of a rotating basket equipped with a filter medium. Similar to other filters, centrifugal filtration does not require a density difference between the solids and the suspending liquid. If such density difference exists sedimentation takes place in the liquid head above the cake. This may lead to particle size stratification in the cake, with coarser particles being closer to the filter medium and acting as a precoat for the fines to follow [McCabe and co-workers (7) has a good overview on this topic].

In centrifuges, in addition to the pressure due to the centrifugal head due to the layer of the liquid on top of the cake, the liquid flowing through the cake is also subjected to centrifugal forces that tend to pull it out of the cake. This makes filtering centrifuges excellent for dewatering applications. From the fundamental point of view, there are two important consequences of these additional dewatering forces. First, Darcy's law and all of the theory based on it is incomplete because it does not take into account the effect of mass forces. Second, pressures below atmospheric can occur in the cake in the same way as in gravity fed deep bed filters. The conventional filtration theory has been modified to make it applicable to centrifugal filters (57).

Due to good performance and high cost, centrifuges are often referred to as the ultimate in SLS. They have parts rotating at high speeds and require high engineering standards of manufacture, high maintenance costs, and special foundations or suspensions to absorb vibrations. Another feature distinguishing the filtering centrifuges from other cake filters is that the particle size range they are applicable to is generally coarser, from 10 μm to 10 mm. In particular, cake filters that move the cake across the filter medium are restricted to using metal screens, which by their very nature are coarse. No cloth can withstand the abrasion due to the cake forced on the cloth and pushed over its surface. Only the fixed-bed, batch-operated centrifuges can use cloth as the filtration medium and be used, therefore, with fine suspensions.

15.1. Fixed-Bed Centrifuges. The simplest of the fixed-bed centrifuges is the perforated basket centrifuge (Fig. 38) that has a vertical axis, a closed bottom, and a lip or overflow dam at the top end. In the industrial versions, the basket housing is often supported by a three-point suspension called the three-column centrifuge.

The basket centrifuge is by no means obsolete; it has found a wide range of application in the filtration of slow draining products that require long feed, rinse, and draining times, and for materials sensitive to crystal breakage or that require thorough washing. It can be applied to the finest suspensions of all filtering centrifuges because filter cloths may be of pore size down to 1 μm. The cloth is supported by a mesh.

The basket centrifuge exists in many different versions. The slurry is fed through a pipe or a rotating feed into the basket. The cake is discharged manually by digging it out, or some versions allow the cloth to be pulled inside out by the center causing the cake to discharge; the axis of rotation has to be horizontal in this case. The latter method is particularly suitable for crystalline or thixotropic materials. Alternatively, the cake can be ploughed out by a scraper that moves into the cake after the basket slows down to a few revolutions per minute. The plough directs the solids toward a discharge opening provided at the bottom of the basket, through which the cake simply falls out. The speed of the bowl varies during the cycle, ie, filtration is done at moderate speed, dewatering at high speed, and cake discharge by plowing at low speed. The frequent changes in speed lead to dead times that limit the capacity.

The plough cannot be allowed to reach too close to the cloth and some residual cake remains. Where this is not acceptable, the cake may be removed by a pneumatic system, by vacuum, or by reslurrying. The cycle can be automated

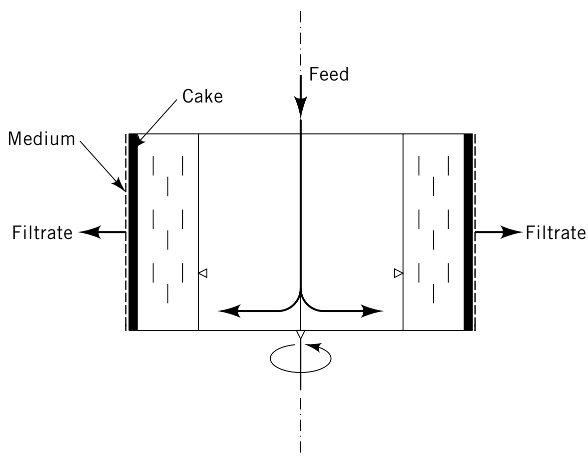


Fig. 38. Perforated basket centrifuge.

and controlled by timers. The maximum speeds of basket centrifuges vary from 800 to 1500 rpm, and basket diameters are in the range from 10 to 1400 mm. A 1200-mm diameter, 750-mm long basket may handle as much as 200 kg of cake in one charge.

The vertical axis of the basket centrifuge may cause some nonuniformity due to the effects of gravity, with the accompanying problems when cake washing is used. This problem can be eliminated by making the axis horizontal. This is known as the peeler centrifuge shown in Figure 39, which is designed to operate at constant speed so that the nonproductive periods of acceleration and deceleration are eliminated. The cake discharge is also affected at full speed, by means of a sturdy knife that peels off the cake into a screw conveyor or a chute in the center of the basket. A narrow-bladed reciprocating knife is usually preferred to the faster full-width blade, because it causes less crystal damage and less severe glazing of the heel surface. The presence of the residual heel is not avoidable in either case, and it may be necessary to recondition it by suitable washes in order to restore its drainage properties. Some designs incorporate cake feeler–leveler mechanisms, which ensure that an even cake is formed.

An interesting improvement of the peeler principle is the addition of a rotating siphon in the form of an additional outer filtrate chamber. This retains the filtrate at a larger diameter than the basket diameter, therefore siphoning higher flow rate through the cloth than would otherwise be achieved. An adjustable siphon pipe controls the total pressure drop across the filter medium. Large increases in throughput are claimed due to the use of the rotating siphon but the additional chamber increases the size of the bowl and similar gains could

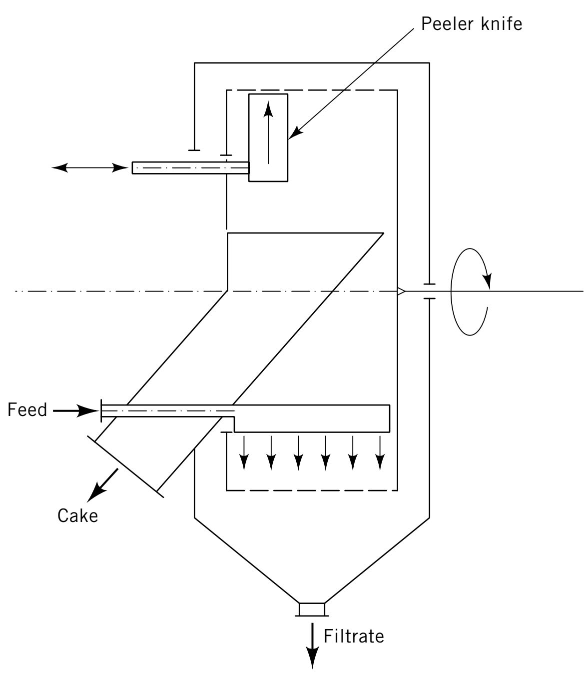


Fig. 39. Peeler centrifuge.

be obtained by simply making the size of the conventional bowl as large as the one with the siphon. It is also not always desirable to lower the absolute pressure in the cake or the medium because cavitation may result due to the release of gas or liquid vapor.

The peeler centrifuge is attractive where filtration and dewatering times are short. The principal application is for high output duties with nonfragile crystalline materials giving reasonable drainage rates that require good washing and dewatering. Basket diameters range from 250 to 2500 mm and speeds from 750 to 4000 rpm.

15.2. Moving Bed Centrifuges. The continuously fed, moving bed machines are available with conical or cylindrical screens. As can be seen in Figure 40, the conical screen centrifuges have a conical basket rotating either on a vertical or horizontal axis, with the feed suspension fed into the narrow end of the cone. If the cone angle is sufficiently large for the cake to overcome its friction on the screen, the centrifuge is self-discharging. Such machines cannot handle dilute slurries and require high feed concentrations of the order of 50% by mass, eg, as occurs in coal dewatering. Different products, however, require different cone angles. Unnecessarily large angles shorten the residence time of the solids on the screen surface and thus lead to poor dewatering. The movement of the solids along the screen may be assisted by vibrating the basket either in the axial direction or, in a tumbler centrifuge, in a tumbling action.

A positive control of the solids on the screen is achieved with the scroll-type conical screen centrifuge. Here the scroll moves slowly in relation to the bowl, conveying the solids along the bowl; variable speed differential drive allows good control of the solids movement. Crystal breakage is more severe due to the action of the scroll. Basket speeds vary from 900 to 3000 rpm, while the basket diameters range from 300 to 1000 mm. Washing is possible with all the conical screen centrifuges but it is not very efficient, giving poor separation of the mother liquor and the wash liquid.

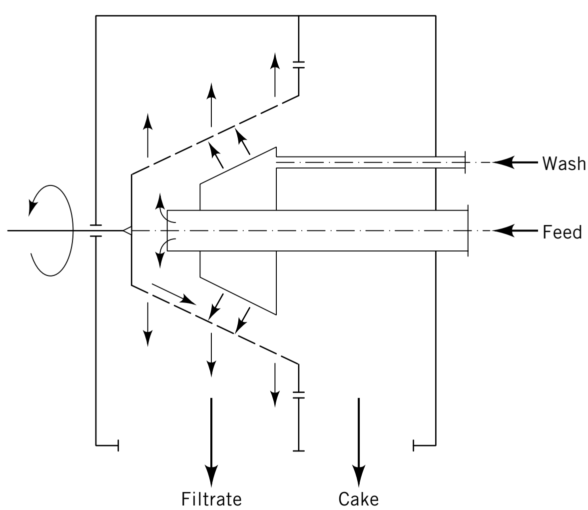


Fig. 40. Conical basket centrifuge.

The pusher-type centrifuge has a cylindrical basket with its axis horizontal. The feed is introduced through a distribution cone at the closed end of the basket (Fig. 41) and the cake is pushed along the basket by means of a reciprocating piston that rotates with the basket. The screen is made of self-cleaning trapezoidal bars of at least 50 μm in spacing, so that only particles larger than ~50 μm can be efficiently separated by this machine. Each stroke of the piston stops just short of the discharge end of the basket, thus allowing a layer of the cake to act as a lip or a dam to effect the axial retention of the feed suspension. If the feedstock is allowed to rise to a height exceeding this remnant cake thickness, the suspension will overflow at the discharge end and rapidly erode a deep furrow in the cake. This is highly undesirable and an overflow limit exists as an upper limit for the feed rate (2,57).

Pusher centrifuges can be made with multistage screens consisting of several steps of increasing diameter, which is advantageous for liquids of high viscosity, or where the cake is soft, plastic, or of high frictional resistance to sliding. It also makes washing better, the washing liquid being introduced over the second stage screen. The transit of the solids over the step reorients the particles.

Pusher centrifuges require high feed concentrations to enable the formation of a sufficiently rigid cake to transmit the thrust of the piston. The diameters vary from 150 to 1400 μm, the stroke frequency from 20 to 100 strokes per minute, and the solids handling capacities up to 40 metric tons per hour or more.

15.3. Scale-Up. The scale-up of filtration centrifuges is usually done on an area basis, based on small-scale tests. A test procedure has been described with a specially designed filter beaker to measure the intrinsic permeability of the cake (7,132). The best test is, of course, with a small-scale model, using the actual suspension. Many manufacturers offer small laboratory models for such tests. The scale-up is most reliable if the basket diameter does not increase by a factor of >2.5 from the small scale. Newer more modern bucket-type centrifuges have found widespread use for filtering centrifuge scale-up (133).

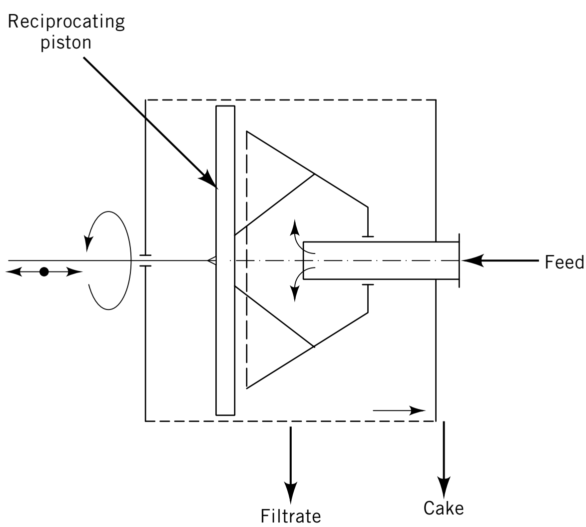


Fig. 41. Pusher centrifuge.

Newer designs are also now available (FIMA and Heinkel) that combine centrifugal dewatering with hyperbaric or steam-assisted drying in order to achieve even lower moisture contents (134,135).

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George G. Chase
The University of Akron

Ernest Mayer
E. I. du Pont Nemours and Company