Chapter 16

Phase Separation Equipment

INTRODUCTION

Phase separation, as its name implies, simply involves the separation of one (or more) phase(s) from another phase. Most industrial equipment used for this class of processes involve the relative motion of the two phases under the action of various external forces (e.g., gravity, electrostatic, and so on).

Consider the following. The separation of solids (particulates) from a gas (usually air) is based on the movement of solid particles through the gas. The objective is often their separation/removal in order for their recovery for economic reasons (e.g., coffee beans, gold dust, and so on) and/or to comply with applicable (environmental) standards and regulations. In order to accomplish this, the particle is subjected to one or more external forces, which are large enough to separate the solid phase from the gas phase during its residence time in the phase separation device.

There are basically five phase separation processes:

1 Gas–Solid (G–S)
2 Gas–Liquid (G–L)
3 Liquid–Solid (L–S)
4 Liquid–Liquid (L–L); immiscible
5 Solid–Solid (S–S)

As one might suppose, the major phase separation process encountered in industry is G–S. As such, the presentation to follow primarily addresses G–S subject matter, including equipment. However, it should be noted that the basic principle of G–S phase separation processes almost universally apply to G–L processes and equipment, and also often apply (but to a lesser degree) to L–S, L–L, and S–S phase separation processes.

Traditional equipment for G–S separation processes include:

1 gravity settlers
2 centrifugal separators (cyclones)

  3 electrostatic precipitators

  4 wet scrubbers

  5 baghouses

Such equipment must be matched to process variables such as flow rate, temperature, nature and concentration of solids, and desired degree of separation. It is also well recognized that no universal separation method exists that will satisfy all problems and conditions. The choice of method depends on many technical and economic factors.

Each separation problem is unique; therefore, some preliminary knowledge is required in order to design compatible equipment. To acquire this knowledge, particulate/phase information must be made available. Particulate properties most fundamental to the performance and choice of the equipment are: particle size, particle-size distribution, shape, structure, density, composition, electrical conductivity, abrasiveness, corrosiveness, flammability, hygroscopic properties, flowability, toxicity, and agglomeration tendencies. Also important is knowledge of the gas stream properties, including temperature, pressure, humidity, density, viscosity, dew point for condensable components, electrical conductivity, corrosiveness, toxicity, composition, and flammability. The process conditions include the allowable pressure drop, electrical power requirements, separation efficiency requirements, particle concentration, and gas volumetric flow rate. The plant factors include maintenance, space limitations, availability of utilities, applicable safety and health protection, potential disposal facilities, and materials of construction. Finally, knowledge of auxiliary equipment is required, including pumps, fans, compressors, motors, ducting, valves, control instrumentation, storage facilities, and conveying equipment.

Although most of the properties and factors considered in the selection and design of separation equipment are self-explanatory, certain introductory G−S dynamics principles merit discussion and are now reviewed. This is followed by a review of the aforementioned G−S separation equipment; the chapter concludes with a discussion of G−L, L−S, L−L, and S−S separation equipment. There is some overlap with these phase separation processes and this is noted in the presentation.

One of the authors[1,2] has employed several definitions for particulates: a small, discrete mass of solid or liquid matter; a fine liquid or a solid particle that is found in the air or emissions; any solid or liquid matter that is dispersed in a gas; or, insoluble solid matter dispersed in a liquid so as to produce a heterogeneous mixture. Further, particulate matter 10 (PM10) is defined as particulate matter with a diameter less than or equal to 10 micrometers (μm) while particulate matter 2.5 (PM2.5) is particulate matter with a diameter less than or equal to 2.5 micrometers (μm).

Particle size is the single most important characteristic that affects the behavior of a particle. The range in size of particles observed in practice is remarkable. Some of the particles collected are as large as raindrops. However, some of the particles created in high-temperature incinerators and metallurgical processes can consist of a few molecules clustered together. These particles cannot be seen by sensitive light microscopes because they are extremely small in size. These sizes approach

those of individual gas molecules (which range from 0.2–1.0 nm). In fact, particles composed of a few molecules clustered together can exist in a stable form. Some of these industrial processes generate particles in the range of 10–100 nm. However, particles in this size range can grow and agglomerate to yield particles in the +100 nm range.

The overall collection/removal process for particulates in a fluid essentially consists of four steps.[3]

1 An external force (or forces) must be applied that enables the particle to develop a velocity that will displace and/or direct it to a collection or retrieval section or area or surface.

2 The particle should be retained at this area with strong enough forces so that it is not re-entrained.

3 As collected/recovered particles accumulate, they are subsequently removed.

4 The ultimate disposition of the particles completes the process.

Obviously, the first is the most important step. The particle collection mechanisms discussed below are generally applicable when the fluid is air; however, they may also apply if the fluid is water, as in the G–L, L–S, and L–L presentations.

The forces listed below are basically the "tools" which may be used for particulate/recovery collection:

1 Gravity settling

2 Centrifugal action

3 Inertial impaction

4 Electrostatic attraction

5 Thermophoresis and diffusiophoresis

6 Brownian motion

Note that all of these collection mechanism forces are strongly dependent on particle size. The above mechanisms 2–4 are briefly described below in Illustrative Examples 16.1–16.3.

ILLUSTRATIVE EXAMPLE 16.1

Describe how a "centrifugal force" is employed for particle capture/recovery.

SOLUTION: So-called "centrifugal force" is one of the collection mechanisms used for particle capture. The shape or curvature of the collector causes the gas stream to rotate in a spiral motion. Larger particles move toward the outside wall by virtue of their momentum. The particles lose kinetic energy there and are separated from the gas stream. Particles are then acted upon by gravitational forces and are collected. Thus, both "centrifugal" and gravitational forces may be responsible for particle collection/recovery. ∎
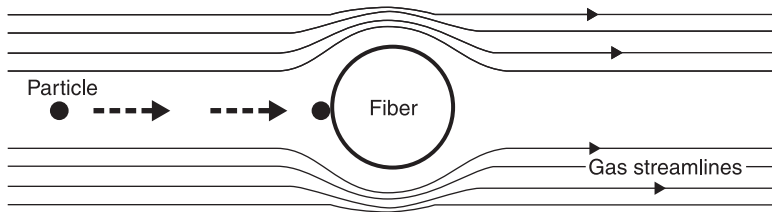
Figure 16.1 Impaction.

ILLUSTRATIVE EXAMPLE 16.2

Describe inertial impaction.

SOLUTION: *Inertial impaction* occurs when an object (e.g., a fiber or liquid droplet), placed in the path of a particulate-laden gas stream, causes the gas to diverge and flow around it. Larger particles, however, tend to continue in a straight path because of their inertia; they may impinge on the obstacle and be collected (as in Fig. 16.1). ■

ILLUSTRATIVE EXAMPLE 16.3

Describe how electrostatic attraction can impact collection/recovery of particles.

SOLUTION: Another primary particle collection mechanism involves electrostatic forces. The particles can be naturally charged, or, as in most cases involving *electrostatic attraction*, be charged by subjecting the particle to a strong electric field. The charged particles migrate to an oppositely charged collection surface. This is the collection mechanism responsible for particle capture in an electrostatic precipitator (see a later section). The electrostatic force F_E experienced by a charged particle in an electric field is given by $F_E = qE_p$ (consistent units) where q = particle charge (not to be confused with volume flowrate) and E_p = the collection field intensity (electric field). ■

Other factors affecting collection mechanisms include:

1 Nonspherical particles
2 Wall effects
3 Multiparticle effects
4 Multidimensional flow

Details on (1)–(4) are available in the literature.[4]

FLUID–PARTICLE DYNAMICS

Most industrial techniques used for the separation of particles from gases involve the relative motion of the two phases under the action of various external forces. The

recovery methods for particulates are based on the movement of solid particles (or liquid droplets) through a gas.

Whenever a difference in velocity exists between a particle and its surrounding fluid, the fluid will exert a resistive force on the particle. Either the fluid (gas) may be at rest with the particle moving through it, or the particle may be at rest with the gas flowing past it. It is generally immaterial which phase (solid or gas) is assumed to be at rest; it is the *relative* velocity between the two that is important. The resistive force exerted on the particle by the gas is called the *drag*. In treating fluid flow through pipes, a *friction factor* term is used in many engineering calculations.[5] An analogous factor, called the *drag coefficient*, is employed in drag force calculations for flow past particles.

Consider a fluid flowing past a stationary solid sphere. If F_D is the drag force and ρ is the density of the gas, the drag coefficient C_D is defined as

$$C_D = \frac{F_D/A_p}{\rho v^2/2g_c} \tag{16.1}$$

where A_p is given by $\pi d_p^2/4$. In the following analysis, it is assumed that

1 the particle is a rigid sphere (with a diameter d_p) surrounded by gas in an infinite medium (no wall or multiparticle effects), and

2 the particle or fluid is not accelerating.

From dimensional analysis, one can show that the drag coefficient is solely a function of the particle Reynolds number, Re:

$$C_D = C_D(\text{Re}) \tag{16.2}$$

where

$$\text{Re} = \frac{d_p v \rho}{\mu} \tag{16.3}$$

and v is the relative velocity, μ is the fluid (gas) viscosity and ρ is the fluid (gas) density. The quantitative use of the equation of particle motion (to be developed shortly) requires numerical and/or graphical values of the drag coefficient. Graphical values are presented in Figure 16.2.

The drag force, F_D, exerted on a particle by a gas at low Reynolds numbers is given in equation form by

$$F_D = 6\pi\mu v a/g_c = 3\pi\mu v d_p/g_c \tag{16.4}$$

Equation (16.4) is known as *Stokes' law* and can be derived theoretically.[5,6] However, keep in mind that Stokes' equation is valid only for very low Reynolds numbers—up to Re $\approx$ 0.5; at Re = 1, it predicts a value for the drag force that is nearly 10% too low. In practical applications, Stokes' law is generally assumed applicable up to a Reynolds number of 2.0. For this "creeping flow" around a particle, one obtains

$$C_D = 24/\text{Re} \tag{16.5}$$
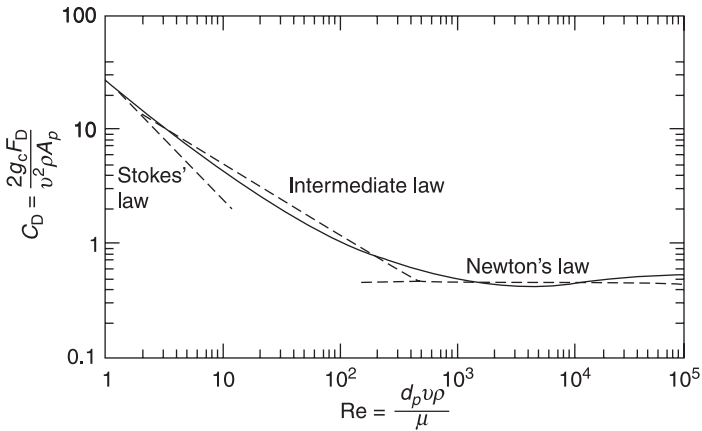
Figure 16.2 Drag coefficient for spheres.

This is the straight-line portion of the log–log plot of C_D vs Re (Fig. 16.2). For higher values of the Reynolds number, it is almost impossible to perform purely theoretical calculations. However, others[1] have managed to estimate, with a considerable amount of effort, the drag and/or drag coefficient at higher Reynolds numbers.

If a particle is initially at rest in a stationary gas and is then set in motion by the application of a constant external force or forces, the resulting motion occurs in two stages. The first period involves acceleration, during which time the particle velocity increases from zero to some maximum velocity. The second stage occurs when the particle achieves this maximum velocity and remains constant. During the second stage, the particle is not accelerating. The final, constant, and maximum velocity attained is defined as the *terminal settling velocity* of the particle. Most particles can be shown to reach their terminal settling velocity almost instantaneously.[7] For this case, one can show that

$$v = \frac{fd_p^2\rho_p}{18\mu} \tag{16.6}$$

for the Stokes' law range. Equations are also available for higher Reynolds number flow.[1,5,6] Keep in mind that f denotes the external force per unit mass of particle. One consistent set of units for the equations above is ft/s^2 for f, ft for d_p, lb/ft^3 for ρ, lb/ft·s for μ, and ft/s for v.

Another important consideration involves the Cunningham correction factor (CCF).[8] At very low values of the Reynolds number, when particles approach sizes comparable to the mean free path of the fluid molecules, the medium can no longer be regarded as continuous. For this condition, particles can fall between the molecules at a faster rate than predicted by the aerodynamic theories that led to the previous standard drag coefficients. To allow for this "slip," Cunningham introduced a multiplying correction factor (C) to the velocity in Stokes' law.

ILLUSTRATIVE EXAMPLE 16.4

A fly ash particle settles through air. Calculate the particle's terminal velocity and determine how far it will fall in 30 s. Assume that the particle is spherical. Use the data provided below:

> Fly ash particle diameter $= 40\ \mu m$
> Air temperature and pressure $= 238°F$, 1 atm
> Specific gravity of fly ash $= 2.31$

SOLUTION: For the problem at hand, the particle density is calculated using the specific gravity given:

$$\rho_p = (2.31)(62.4)$$
$$= 144.14\ \text{lb/ft}^3$$

The density of air is

$$\rho = P(\text{MW})/RT$$
$$= (1)(29)/(0.7302)(238 + 460)$$
$$= 0.0569\ \text{lb/ft}^3$$

The viscosity of air is

$$\mu = 0.021\ \text{cP}$$
$$= 1.41 \times 10^{-5}\ \text{lb/ft} \cdot \text{s}$$

For a d_p of 40 μm:

$$v = \frac{g d_p^2 \rho_p}{18\mu}; \quad g = \text{gravity force} \tag{16.6}$$

$$= \frac{(32.2)[(40)/(25{,}400)(12)]^2(144)}{(18)(1.41 \times 10^{-5})}$$

$$= 0.315\ \text{ft/s}$$

The distance, D, that the fly ash particle will fall in 30 s may now be calculated,

$$D = (30)(0.315); \quad d_p = 40\ \mu m$$
$$= 9.45\ \text{ft}$$ ∎

There are many techniques available for measuring the particle size distribution of particulates. The wide size range covered, from nanometers to millimeters, cannot always be analyzed using a single measurement principle. Added to this are the usual constraints of capital costs vs operating costs, speed of operation, degree of skill required, and most important, the end-use requirement. A common method of specifying large particle sizes is to designate the screen mesh that has an aperture corresponding to the particle diameter. This received treatment in Chapter 14.

Efficiency is the other characteristic quantity which warrants further discussion. The efficiency of a particulate recovery/control device is usually expressed as the percentage of *mass* collected by the unit compared with that entering the unit. It may be

calculated on a particle *number* basis:

$$E_N = \left(\frac{\text{particles collected}}{\text{particles entering}}\right) 100 \qquad (16.7)$$

or on a total mass basis:

$$E = \left(\frac{(\text{inlet loading}) - (\text{outlet loading})}{\text{inlet loading}}\right) \times 100 \qquad (16.8)$$

It is extremely important to distinguish between the two. Larger particles, which possess greater mass and are more easily recovered in a device, will contribute more significantly to the efficiency calculated on a mass or weight basis.

ILLUSTRATIVE EXAMPLE 16.5

Consider an aerosol volume that contains 100 1-μm particles and 100 100-μm particles. If the efficiency of separation is 90% for 1-μm particles and 99% for 100-μm particles, calculate E_N and E.

SOLUTION: On a particle count basis, 90 1-μm and 99 100-μm particles will be removed out of a total of 200. This gives a particle count efficiency of

$$E_N = \left(\frac{189}{200}\right) 100 = 94.5\%$$

On a mass basis, however, if a 1-μm particle has unit mass, a 100-μm particle has 10^6 mass unit. The mass efficiency is then given by

$$E = \left(\frac{90(1) + 99(10^6)}{100(1) + 100(10^6)}\right) 100 = 99\%$$

Any expression of the efficiency of a particulate recovery device is therefore of little value without a careful description of the size spectrum of particles involved.   ∎

GAS–SOLID (G–S) EQUIPMENT

As noted above, this chapter will key on this section. The following G–S equipment receive treatment:

    1 Gravity (settling) settlers

    2 Cyclones

    3 Electrostatic precipitators

    4 Venturi scrubbers

    5 Baghouses

Equipment description and pertinent design/predictive equations are included below.[1] Illustrative examples complement the presentation.

Gravity Settlers

The gravity settler was one of the first devices used to separate particulates from gases (primarily) and other fluids. It is an expansion chamber in which the gas velocity is reduced, thus allowing the particle to settle out under the action of gravity. One primary feature of this device is that the external force causing separation of the particles from the gas stream is provided free by nature. This chamber's use in industry, however, is generally limited to the removal of larger-sized particles, for example, $>50\ \mu m$ in diameter.

    Inertial collectors, on the other hand, depend on another effect, in addition to gravity, to lead to a successful separation process. This other mechanism is an inertial or momentum effect. It arises by changing the direction of the velocity of the gas and imparting a downward motion to the particle. From a calculational point of view, this induced particle motion is superimposed on the motion arising as a result of gravity.

    An *elutriator* is a slight modification of the gravity settler. The unit consists of one or more vertical tubes or towers through which the dust-laden gas passes upward at a given velocity. The larger particles that settle at a velocity higher than that of the rising fluid air are collected at the bottom of the tube while the smaller particles are carried out the top. In order to vary the fluid velocity, several columns of different diameters may be used in series to bring about more refined separation.

    For capture to occur in a gravity settler, the particle must reach a collection surface $a'b'c'd'$ in Figure 16.3 during its residence time in the unit. Theodore[1] has shown that

$$d_p = (18\mu q/g\rho_p BL)^{0.5} \tag{16.9}$$

The particle diameter above represents a limiting value since particles with diameters equal to or greater than this value will reach the collection surface and particles with diameters less than this value will escape from the unit. This limiting particle diameter may ideally be thought of as the *minimum diameter* of a particle that will automatically
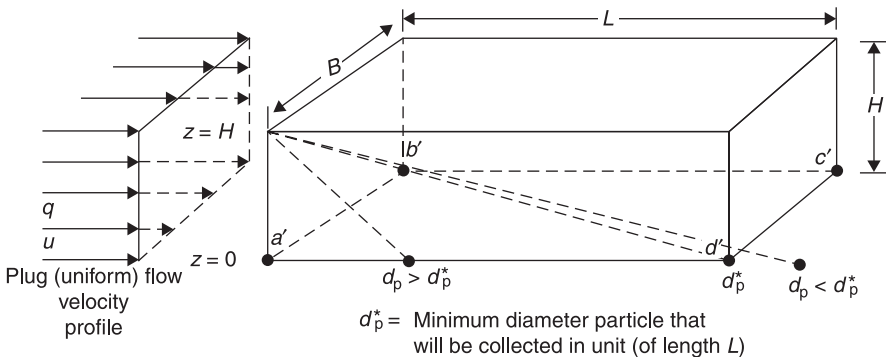

Figure 16.3   Gravity settler nomenclature.

be captured for the above conditions. This diameter is normally denoted by d_p^* or $d_p(\text{min})$.

Collection efficiencies of 100% were used to derive the equations for d_p^*. The collection efficiency, E, for a monodispersed aerosol (particulates of one size) can be shown to be

$$E = \left[\frac{g\rho_p BL}{18\mu q}\right] d_p^2 \tag{16.10}$$

The term in brackets in Equation (16.10) is often multiplied by a dimensionless empirical factor to correlate theoretical efficiencies with experimental data. If no information is available, it is suggested that 0.5 be used. Thus, Equation (16.10) can be written

$$E = 0.5\left[\frac{g\rho_p BL}{18\mu q}\right] d_p^2 \tag{16.11}$$

The process design variables for a settling chamber consist of length (L), width (B), and height (H). These parameters are usually chosen by the chamber manufacturer in order to remove all particles above a specified size. The chamber's design must provide conditions for sufficient particle residence time to capture the desired particle size range. This can be accomplished by keeping the velocity of the exhaust gas through the chamber as low as possible. If the velocity is too high, dust re-entrainment will occur. However, the design velocity must not be so low as to cause the design of the chamber volume to be exorbitant. Consequently, the units are designed for gas velocities in the range $1-10$ ft/s ($0.305-3.05$ m/s).

The development in this section only provides the theoretical collection efficiency of a settling chamber for a single-sized particle. Since the gas stream entering a unit consists of a distribution of particles of various sizes, a *fractional efficiency* curve must be used to determine the overall collection efficiency. This is simply a curve or equation describing the collection efficiency for particles of various sizes. As noted earlier, the overall efficiency can then be calculated using

$$E = \Sigma(E_i)(w_i) \tag{16.12}$$

where E = overall collection efficiency

E_i = fractional efficiency of a specific size particle

w_i = mass fraction of a specific size particle (in range)

ILLUSTRATIVE EXAMPLE 16.6

A hydrochloric acid mist in air at 25°C is to be collected in a gravity settler. You are requested to calculate the smallest mist droplet (spherical in shape) that will definitely be collected by the settler. Assume the acid concentration to be uniform through the inlet cross section of the unit and Stokes' law applies. Operating data and information on the gravity settler are given below:

Dimensions of gravity settler = 30 ft wide, 20 ft high, 50 ft long

Actual volumetric flow rate of acidic gas = 50 ft^3/s

Specific gravity of acid $= 1.6$
Viscosity of air $= 0.0185\,\text{cP} = 1.243 \times 10^{-5}\,\text{lb/ft} \cdot \text{s}$
Density of air $= 0.076\,\text{lb/ft}^3$

SOLUTION: For the problem at hand, first determine the density of the acid mist:

$$\rho_p = (62.4)(1.6)$$

$$= 99.84\,\text{lb/ft}^3$$

Calculate the minimum particle diameter in both feet and micrometers using Equation (16.6), assuming that Stokes' law applies:

$$d_p = \left(\frac{(18)(1.243 \times 10^{-5})(50)}{(32.2)(99.84)(30)(50)}\right)^{1/2}$$

$$= 4.82 \times 10^{-5}\,\text{ft}$$

There are $3.048 \times 10^5\,\mu\text{m}$ in 1 ft. Therefore

$$d_p = 14.7\,\mu\text{m}$$

∎

Cyclones

Cyclones provide a relatively low-cost method of removing particulate matter from exhaust gas streams. Cyclones are somewhat more complicated in design than simple gravity settling systems and their removal efficiency is accordingly much higher than that of settling chambers. However, cyclones are not as efficient as elec-trostatic precipitators, baghouses, and venturi scrubbers, but are often installed as pre-cleaners before these more effective devices.

The basic separation principle is simple. Particles enter the device with the flowing gas; the gas stream is forced to turn but the larger particles have more momen-tum and cannot turn with the gas. These larger particles impact and fall down the cyclone wall to be collected in a hopper. The gas stream actually turns a number of times in a helical pattern, much like the funnel of a tornado. The repeated turnings provide many opportunities for particles to pass through the streamlines, thus hitting the cyclone wall. There are other variations in the design of cyclones. They are usually characterized by where the gas enters and exits the cyclone body (tangentially, axially, or peripherally). There are four major parts to a cyclone: the inlet, the cyclone body, the dust discharge system, and the outlet. All affect the overall efficiency of the cyclone.

A type of parallel arrangement uses the axial entry cyclone. Arrangements of high-efficiency, small-diameter axial cyclones can provide increases in collection efficiency with corresponding reductions in pressure drop, space, and cost. Such a unit is defined as a *multiclone*. Pressure drops commonly range from 4−6 inches (10−15 cm) of water.

Objects moving in circular paths tend to move away from the center of their motion. The object moves outward as if a force is pushing it out. This apparent force is known as *centrifugal force*. Note that, in actuality, centrifugal forces do not

exist—instead, physicists refer to the *centripetal* forces acting on an object. As noted above, the whirling motion of the gas in a cyclone causes particulate capture.

Three important parameters can be used to characterize cyclone performance:

$$d_{pc} = \text{cut diameter}$$
$$E = \text{overall collection efficiency}$$
$$\Delta P = \text{pressure drop}$$

Equations involving each of these parameters are provided below. The equations should be used with caution, however, since there are strict limitations on their applicability.

The *cut diameter* is defined as the size (diameter) of particles collected with 50% efficiency. It is a convenient way of defining efficiency for a control device since it provides information on the effectiveness for a particle size range. A frequently used expression for cut diameter is

$$d_{pc} = \left(\frac{9\mu B_c}{2\pi N v_i (\rho_p - \rho)} \right)^{0.5} \tag{16.13}$$

where μ = viscosity, lb/ft · s (Pa · s)

N = effective number of turns (5–10 for the common cyclone)

v_i = inlet gas velocity, ft/s (m/s)

ρ_p = particle density, lb/ft^3 (kg/m^3)

ρ = gas density, lb/ft^3 (kg/m^3)

B_c = inlet width, ft (m)

The cut diameter, d_{pc} or $[d_p]_{\text{cut}}$, is a characteristic of the recovery/control device. Figure 16.4 shows a size efficiency curve and indicates the cut diameter and the *critical diameter*, $[d_p]_{\text{crit}}$, the particle size collected at 100% efficiency. As values of $[d_p]_{\text{crit}}$ are difficult to obtain from such curves, the cut size is often determined instead.
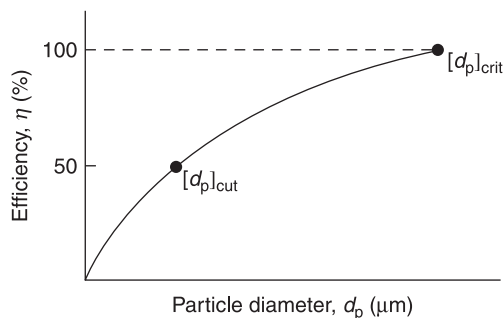
Figure 16.4   Typical size efficiency curve.

A number of formulas exist for the calculation of the cut diameter and critical diameter. A value of N, the number of turns, must be known in order to solve Equation (16.13) for $[d_p]_{cut}$. Given the volumetric flow rate, inlet velocity, and dimensions of the cyclone, N can be easily calculated. Values of N can vary from 1–10, with typical values in the 4–5 range.

A number of equations have been developed for determining the fractional cyclone efficiency E_i for a given size particle. As noted earlier, *fractional efficiency* is defined as the fraction of particles of a given size collected in the cyclone, compared to those of that size entering the cyclone. The most popular method of calculating cyclone fractional efficiency and overall efficiency was developed by Lapple.[9,10] Lapple first computed the ratios $d_p/[d_p]_{cut}$, the particle diameter vs the cut diameter ratio as determined from Equation 16.13. He found that cyclone efficiency correlates in a general way with this ratio. For a typical cyclone, efficiency will increase as the ratio increases as provided by Lapple in Figure 16.5.

As a universal curve for common cyclones, the preceding correlation has been found to agree reasonably well with experimental data. To calculate fractional efficiencies, the procedure presented below should be completed.

d_p range	wt fraction in range	$d_p/[d_p]_{cut}$	E_i for each d_p from experiment or Lapple's method, %	wt fraction $\times E_i$
			Lapple Calculation Procedure	

The sum of these products in the rightmost section of the box will yield the overall efficiency. Theodore[11] provides more detailed information.

The pressure drop across a cyclone is an important parameter to the purchaser of such equipment. Increased pressure drop means greater costs for power to move an exhaust gas through the device. With cyclones, an increase in pressure drop usually means that there will be an improvement in collection efficiency. One exception to
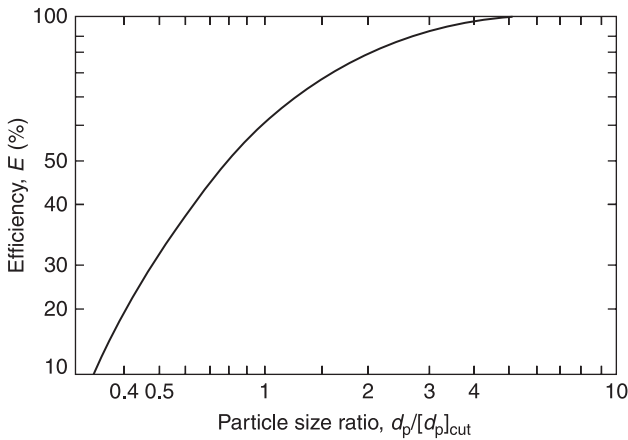
Figure 16.5   Cyclone collection efficiency vs particle size ratio.

this is the use of pressure recovery devices attached to the exit tube; these reduce the pressure drop but do not adversely affect collection efficiency. For these reasons, there have been many attempts to predict pressure drops from design variables. The concept is that having such an equation, one could work back and optimize the design of new cyclones.

The most popular of the empirical pressure drop equation has the form

$$\Delta P = K_c \rho v_i^2; \quad \text{consistent units} \tag{16.14}$$

where K_c = a proportionality factor.

If ΔP is measured in inches of water, K_c can vary from 0.013–0.024, with 0.024 the norm. Velocities for cyclones range from 20–70 ft/s (6–21 m/s), although common velocities range from 50–60 ft/s (15–18 m/s). At velocities greater than 80 ft/s (24 m/s), turbulence increases in the cyclone and efficiency may actually decrease.

Pressure drops for single cyclones vary depending on both size and design. Common ranges are listed below:

Low-efficiency cyclones	2–4 in H_2O (5–10 cm H_2O)
Medium-efficiency cyclones	4–6 in H_2O (10–15 cm H_2O)
High-efficiency cyclones	8–10 in H_2O (20–25 cm H_2O)

ILLUSTRATIVE EXAMPLE 16.7

An engineer is requested to determine the cut size diameter and overall collection efficiency of a cyclone given the particle size distribution of a dust from a cement kiln. Particle size distribution and other pertinent data are provided in Table 16.1.

Additional data:

Gas viscosity = 0.02 cP
Specific gravity of the particles = 2.9
Inlet gas velocity to cyclone = 50 ft/s

Table 16.1   Particle Size Distribution Data for Illustrative Example 16.7

Average particle size in range d_p, μm	Weight percent
1	3
5	20
10	15
20	20
30	16
40	10
50	6
60	3
>60	7

Effective number of turns within cyclone $= 5$

Cyclone diameter $= 10$ ft

Cyclone inlet width $= 2.5$ ft

SOLUTION: Lapple's method provides the collection efficiency as a function of the ratio of particle diameter to cut diameter, as presented in Figure 16.5. One may also use the Theodore–De Paola equation[1,11,12]

$$E = \frac{1.0}{1.0 + (d_{pc}/d_p)^2} \qquad (16.15)$$

in place of Figure 16.5.

   For the problem at hand, determine the value of $\rho_p - \rho$:

$$\rho_p - \rho \cong \rho_p$$

$$= (2.9)(62.4)$$

$$= 181 \text{ lb/ft}^3$$

Calculate the cut diameter:

$$d_{pc} = \left(\frac{9\mu B_c}{2\pi N v_i (\rho_p - \rho)} \right)^{1/2} \qquad (16.13)$$

Substituting,

$$d_{pc} = \left(\frac{(9)(0.02)(6.72 \times 10^{-4})(2.5)}{(2\pi)(5)(50)(181)} \right)^{1/2}$$

$$= 3.26 \times 10^{-5}$$

$$= 9.94 \text{ } \mu\text{m}$$

Table 16.2   Results for Illustrative Example 16.7

d_p, μm	w_i	d_p/d_{pc}	E_i, %	$w_i E_i$, %
1	0.03	0.10	0	0.0
5	0.20	0.5	20	4.0
10	0.15	1.0	50	7.5
20	0.20	2.0	80	16.0
30	0.16	3.0	90	14.4
40	0.10	4.0	93	9.3
50	0.06	5.0	95	5.7
60	0.03	6.0	98	2.94
>60	0.07	—	100	7.0

Table 16.2 is generated using Lapple's method. Slightly more accurate results can be obtained by employing the Theodore–De Paola equation. The overall collection efficiency is therefore

$$E = \Sigma w_i E_i = 0 + 4 + 7.5 + 16 + 14.4 + 9.3 + 5.7 + 2.94 + 7$$

$$= 66.84\%$$

$$= 0.6684 \qquad \blacksquare$$

Electrostatic Precipitators

Electrostatic precipitators (ESPs) are unique among gas cleaning equipment in that the forces separating the particulates from the gas stream are applied directly to the particulates themselves, and hence the energy required to effect the separation may be considerably less than for other types of gas cleaning apparatus. Gas pressure drops through the precipitator may be of the order of 1 inch of water or less as compared with pressures of up to 10–100 inches of water for scrubbers and baghouses. This fundamental advantage of electrostatic precipitation has resulted in its widespread use in applications where large gas volumes are to be handled and high efficiencies are required for collection of small particles.

Burning low-sulfur coal produces a fly ash that has a high resistivity and is difficult to collect. The problem, similar to that for the smelter dusts, is that there is insufficient sulfur trioxide in the gas, which is often corrected by conditioning the gas with this chemical. This resistivity problem is discussed below and later in this chapter.

The electrostatic precipitation process consists of three fundamental steps:

1 Particle charging

2 Particle collection

3 Recovery/removal of the collected dust

Particle charging in precipitators is accomplished by means of a corona, which produces ions that become attached to the particles. Generation of a corona requires the development of a highly non-uniform electric field—a condition that occurs near the wire when a high voltage is applied between the wires and collection electrodes. The electric field near the wire accelerates electrons present in the gas to velocities sufficient to cause ionization of the gas in the region near the wire. The ions produced as a result of the corona migrate toward the collection electrode, and in the process, collide with and become attached to particles suspended in the gas stream. The attachment of ions results in the buildup of an electric charge, the magnitude of which is determined by the number of attached ions.

The charge on the particles in the presence of an electric field results in a force in the direction of the collection electrode. The magnitude of the force is dependent on the charge and the field. This force causes particles to be deposited on the collection electrode where they are held by a combination of mechanical, electrical, and molecular forces.

Once recovered, particles can be removed by coalescing and draining in the case of liquid aerosols, or by periodic impact or rapping in the case of solid material. In the latter case, a sufficiently thick layer of dust must be collected so that it falls into the hopper or bin in coherent masses (effectively like a sheet) to prevent excessive re-entrainment of the material into the gas stream.

The physical arrangement of precipitators differ depending on the type of application. Wire and cylinder electrodes are used in some applications; however, for reasons of space economy, most commercial precipitators use plates as collection electrodes.

The majority of precipitators are constructed so that the charging and collection steps take place within the same region. Precipitators of this type are termed

single-stage. For some applications, charging takes place in one section which is followed by a section consisting of alternately charged plates. The collecting electric field is established independently of the corona field and such precipitators are termed *two-stage*.

Once a particle is charged, it migrates toward the grounded collection electrode. An indicator of particle movement toward the collection electrode is denoted by the symbol *w* and is called the *particle migration velocity* or *drift velocity*. The particle migration velocity can be calculated, but most ESPs are designed using a particle migration velocity based on field experience rather than theory. Typical particle migration velocity rates such as those listed in Table 16.3 have been published by various ESP vendors. These values can be used to estimate the collection efficiency of the ESP.

Probably the best way to gain insight into the process of electrostatic precipitation is to study the relationship known as the *Deutsch–Anderson equation*. This equation is used to determine the collection efficiency of the precipitator under ideal conditions. The simplest form of the equation is

$$E = 1 - e^{-(wA/q)} \qquad (16.16)$$

where E = fractional collection efficiency of the precipitator

A = effective collecting plate area of the precipitator, ft^2 (m^2)

q = gas flow rate through the precipitator, acfs (acms) [actual ft^3/s (actual m^3/s)]

w = migration velocity, ft/s (m/s)

Table 16.3   Typical Precipitation Rate Parameters for Various Applications

Application	Particle migration velocity	
	ft/s	cm/s
Utility fly ash	0.13–0.67	4.0–20.4
Pulverized coal fly ash	0.33–0.44	10.1–13.4
Pulp and paper mills	0.21–0.31	6.4–9.5
Sulfuric acid mist	0.19–0.25	5.8–7.62
Cement (wet process)	0.33–0.37	10.1–11.3
Cement (dry process)	0.19–0.23	6.4–7.0
Gypsum	0.52–0.64	15.8–19.5
Smelter	0.06	1.8
Open-hearth furnace	0.16–0.19	4.9–5.8
Blast furnace	0.20–0.46	6.1–14.0
Hot phosphorus	0.09	2.7
Flash roaster	0.25	7.6
Multiple hearth roaster	0.26	7.9
Catalyst dust	0.25	7.6
Cupola	0.10–0.12	3.0–3.7

This equation has been used extensively for many years for theoretical collection efficiency calculations. Unfortunately the equation is not scientifically valid, and there are a number of operating parameters that can cause the results to be in error by a factor of ≥ 2. The Deutsch–Anderson equation neglects three significant process variables:

1 It completely ignores the fact that dust re-entrainment may occur during the rapping process.

2 It assumes that the particle size and, consequently, the migration velocity is uniform for all particles in the gas stream.

3 It assumes that the gas flow rate is uniform everywhere across the precipitator and that particle seepage through the hopper section does not occur.

Therefore, this equation should only be used for making preliminary estimates of precipitation collection efficiency.

When the desired collection efficiency and gas flow rate are specified, the required collecting area can be determined from the Deutsch–Anderson equation once an appropriate precipitation rate parameter has been chosen. More recently, better correlations with field data on high-efficiency ESPs have been obtained by raising the exponential term in Equation (16.16) to a power m using existing values of w, i.e.,

$$E = 1 - e^{-(wA/q)^m} \tag{16.17}$$

This provides a more accurate prediction of performance at high efficiency levels, but can become too pessimistic in certain situations. Typical values of m range between 0.4–0.7, with 0.5 as the norm. Equation (16.17) is referred to as the *Matts–Ohnfeldt equation*.

These and many other models have been proposed to predict particulate collection in an electrostatic precipitator, and while special advantages can be argued for each depending on which critical variables they account for, the fact remains that no model exists which accounts for all the variables that describe migration velocity in all situations. This is attributable mainly to the large number of inter-related variables that exist. In the final analysis, the goal is still to determine the correct amount of collecting surface area, which usually depends on the proper selection of w.

ILLUSTRATIVE EXAMPLE 16.8

A small coal-fired power plant sends 2400 acfm through its electrostatic precipitation. The particle migration velocity is known to be 0.35 ft/s. What is the collection area if the overall ESP efficiency is 99.78%?

SOLUTION: Apply the Deutsch–Anderson equation:

$$E = 1 - e^{-(Aw/q)} \tag{16.16}$$

Substituting, one obtains

$$0.9978 = e^{-(A)(0.35)/(2400/60)}$$

Solving for A yields

$$A = 699 \text{ ft}^2$$

■

Venturi Scrubbers

As the name "scrubber" implies, wet collectors or wet scrubbers are devices that use a liquid for separating particles (or polluted gases) from a gas stream. Water sprays can be injected into the gas stream; gas can be forced to pass through sheets or films of liquid; or, the gas can move through beds of plastic spheres covered with liquid. Each of these techniques can effectively collect/remove particulate matter from process gases. They can also effectively remove gases such as HCl or SO_2, but removal/recovery conditions must be right. In addition, gas–liquid contact can bring about gas conditioning, and to a lesser extent, liquid conditioning.

In many cases, the best conditions for removing particulate matter are the poorest for removing gases. In this section, emphasis will be placed on the design and application of wet scrubbers, with particular emphasis on venturi scrubbers for the recovery of particulate matter. Optimum operating conditions for particulate matter removal will also be discussed.

In a typical venturi, the velocity at the throat must increase in order to make up for the decrease in area at the throat. Velocities at such a constriction can range from 200–800 ft/s (from 61–244 m/s). If water is introduced into the throat, the gas is forced to move at a high velocity that will shear the water droplets. Particles in the gas stream then impact onto the droplets produced. Moving a large volume of gas through a small constriction gives a high-velocity flow, but also a large pressure drop across the system. The collection efficiency for most particles increases with increased velocities (and corresponding increased pressure drops) since the water is sheared into more smaller droplets than at lower velocities. The large number of small droplets, combined with the turbulence in the throat section, provides numerous impaction targets for particle collection.

A number of performance and design equations have been developed from basic particle movement principles (theory) to explain the action of wet scrubbing systems. Many of these start from firm scientific concepts, but only give qualitative results when predicting collection efficiencies or pressure drops. The interaction of particulate matter having a given particle size distribution with water droplets having another size distribution is not easy to express in quantitative terms. As a result of this complexity, experimentally determined parameters are usually required in order to perform engineering calculations.

One of the more popular and widely used collection efficiency equations is that originally suggested by Johnstone[13]

$$E = 1 - e^{-kR(\psi)^{0.5}} \tag{16.18}$$

where E = efficiency, fractional

ψ = inertial impaction parameter, dimensionless

$R =$ liquid-to-gas ratio, gal/1000 acf or gpm/1000 acfm

$k =$ correlation coefficient, the value of which depends on the system geometry and operating conditions, typically 0.1–0.2 acf/gal

The term ψ is given by

$$\psi = \frac{C d_p^2 \rho_p v_t}{9 \mu_G d_0} \tag{16.19}$$

where $d_p =$ particle diameter, ft

$\rho_p =$ particle density, lb/ft^3

$v_t =$ throat velocity, ft/s

$\mu_G =$ gas viscosity, lb/ft · s

$d_0 =$ mean droplet diameter, ft

$C =$ Cunningham correction factor (CCF), dimensionless

Values for the CCF are available in the literature.[1,5,6] However, this correction is usually neglected in scrubber calculations, but the effect becomes more pronounced as the particle size decreases, particularly below 1 μm. The mean droplet diameter, d_0, for standard air and water in a venturi scrubber is given by the Nukiyama– Tanasawa relationship:

$$d_0 = \frac{16,400}{v_t} + 1.45 R^{1.5} \tag{16.20}$$

Available data indicate that venturis often operate with pressure drops in the 30–100 in H$_2$O range. Liquid-to-gas ratios for venturi scrubbers are usually in the range of 5–20 gal/1000 ft^3 of gas. At many facilities, liquid-to-gas ratios ranging from 7–45 gal/1000 ft^3 of gas have been reported. In many cases, a minimum ratio of 7.5 gal/1000 ft^3 is needed to ensure that adequate liquid is supplied to provide good gas "sweeping". Gas velocities for venturi scrubbers are in the 100–400 ft/s range. The low end of this range, 100–150 ft/s, is typical of power plant applications, while the upper end of the range has been applied to lime kilns and blast furnaces.

ILLUSTRATIVE EXAMPLE 16.9

Calculate the overall efficiency of N scrubbers in parallel, assuming that the volumetric flow rates in each scrubber are $q_1, q_2, \ldots, q_N$ and the corresponding efficiencies are $E_1, E_2, \ldots, E_N$, respectively. Assume that the gas is sufficiently well mixed so that the particle concentration (particles/volume) is the same at the inlet of each scrubber. Express the result in terms of the q values and the corresponding E values (see Fig. 16.6).

SOLUTION: If c_1 is entering and c_{10} is leaving,

$$E_i = 1 - \frac{c_{i0}}{c_i}$$
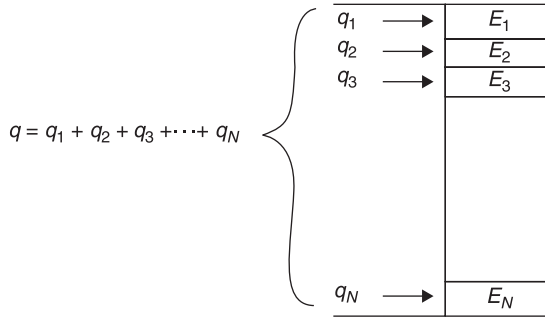
Figure 16.6 N scrubbers in parallel.

or

$$c_{i0} = c_i(1 - E_i)$$

Thus,

$$E = 1 - \frac{c_{10}q_1 + c_{20}q_2 + \cdots + c_{N0}q_N}{c_1q_1 + c_2q_2 + \cdots + c_Nq_N}$$

$$= 1 - \frac{c_1(1 - E_1)q_1 + c_2(1 - E_2)q_2 + \cdots + c_N(1 - E_N)q_N}{c_1q_1 + c_2q_2 + \cdots + c_Nq_N}$$

$$= \frac{\sum q_i c_i E_i}{\sum q_i c_i} \tag{16.21}$$

If $c_1 = c_2 = \cdots = c_N$, the equation above reduces to

$$E = \frac{\sum q_i E_i}{\sum q_i}$$

$$= \frac{\sum q_i E_i}{q} \tag{16.22}$$

∎

ILLUSTRATIVE EXAMPLE 16.10

Calculate the overall efficiency of three venturi scrubbers operating in *parallel* and treating 10,000 acfm of gas. Data are provided in Table 16.4.

Table 16.4 Data for Three Scrubbers in Parallel

Scrubber	q, acfm	c, gr/ft^3	E
1	2500	2.0	0.996
2	5000	4.0	0.985
3	2500	2.0	0.996

Note: gr = grain

SOLUTION: Employ a modified form of the Equation (16.21)

$$E = 1 - \left[\Sigma(q_i c_i P_i)/\Sigma(q_i c_i)\right] \tag{16.23}$$

where P_i = fractional penetration = $1 - E_i$

For the flow rates,

$$\text{Inlet} = (2)(2500)(2.0) + (5000)(4.0) = 30{,}000\,\text{gr/min}$$
$$\text{Outlet} = (2)(2500)(2.0)(0.004) + (5000)(4.0)(0.015) = 340\,\text{gr/min}$$

Thus,

$$E = \frac{30{,}000 - 340}{30{,}000}$$
$$= 0.9887 = 98.87\%$$

∎

ILLUSTRATIVE EXAMPLE 16.11

A consulting firm has been requested to calculate the throat area of a venturi scrubber to operate at a specified collection efficiency. Pertinent data are given below.

> Volumetric flow rate of process gas stream = 11,040 acfm (at 68°F)
> Density of dust = 187 lb/ft³
> Liquid-to-gas ratio = 2 gal/1000 ft³
> Average particle size = 3.2 μm (1.05×10^{-5} ft)
> Average water droplet size = 48 μm (1.575×10^{-4} ft)
> Johnstone scrubber coefficient $k = 0.14$
> Required collection efficiency = 98%
> Viscosity of gas = 1.23×10^{-5} lb/(ft · s)
> CCF = 1.0

SOLUTION: Calculate the inertial impaction parameter ψ from Johnstone's equation:

$$E = 1 - e^{-kR\psi^{1/2}}$$
$$0.98 = 1 - e^{-(0.14)(2)\psi^{1/2}} \tag{16.18}$$

Solving for ψ, one obtains

$$\psi = 195.2$$

From the calculated value of ψ above, back calculate the gas velocity v at the venturi throat:

$$\psi = \frac{\rho_p v_t d_p^2}{9 d_0 \mu_G} \tag{16.19}$$

$$v_t = \frac{9\psi d_0 \mu}{\rho_p d_p^2} = \frac{(9)(195.2)(1.575 \times 10^{-4})(1.23 \times 10^{-5})}{(187)(1.05 \times 10^{-5})^2}$$

$$= 165.1\,\text{ft/s}$$

Calculate the throat area S using the gas velocity at the venturi throat v_t:

$$S = q/v_t = (11{,}040)/[(60)(165.1)]$$

$$= 1.114\,\text{ft}^2$$

Note that approximately $10\,\text{ft}^2$ of throat area is generally required to treat $10{,}000–20{,}000$ acfm. ∎

Baghouses[19]

One of the oldest, simplest, and most efficient methods for recovering solid particulate from gas streams is by filtration through fabric media. The fabric filter is capable of providing high collection efficiencies for particles as small as $0.01\,\mu\text{m}$ and will remove a substantial quantity of those particles as small as $0.01\,\mu\text{m}$. In its simplest form, the industrial fabric filter consists of a woven or felted fabric through which dust-laden gases are forced. A combination of factors results in the collection of particles on the fabric filters. When woven fabrics are used, a dust cake eventually forms; this, in turn, acts predominantly as a sieving mechanism. When felted fabrics are used, this dust cake is minimal or almost non-existent and the primary filtering mechanisms are a combination of inertial forces, impingement, and so on. These are essentially the same mechanisms that are applied to particle collection on wet scrubbers, where the collection media is in the form of liquid droplets rather than solid fibers.

As particles are collected, the pressure drop across the fabric filtering media increases. Owing in part to fan limitations, the filter must be cleaned at predetermined intervals. Material is removed from the fabric by gravity and/or mechanical means. The fabric filters or bags are usually tubular or flat.

The structure in which the bags are located (hang) is referred to as a *baghouse* and the number of bags in a baghouse may vary from less than a dozen to several thousand. Quite often, when great numbers of bags are involved, the baghouse is compartmentalized so that one compartment may be cleaned while others are still in service.

The basic filtration process may be conducted in many different types of fabric filters in which the physical arrangement of hardware and the method of removing collected material from the filter media will vary. The essential differences may be related, in general, to the following:

1 Type of fabric
2 Cleaning mechanism(s)
3 Equipment geometry
4 Mode of operation

The number of variables necessary to design a fabric filter is very large. Since fundamentals cannot treat all of these factors in the design and/or prediction of performance of a filter, this determination is basically left up to the experience and judgment of the design engineer. In addition, there is no one formula that can determine whether a fabric filter application is feasible. A qualitative description of the filtration process is possible, although quantitatively, the theories are far less successful. Theory, coupled with some experimental data, can help predict the performance and design of the unit.

The state-of-the-art of engineering process design is the selection of filter medium, superficial velocity, and cleaning method that will yield the best economic compromise. Industry relies on certain simple guidelines and calculations, which are usually considered proprietary information, to achieve this. Despite the progress in developing pure filtration theory, and in view of the complexity of the phenomena, the most common methods of correlation are based on predicting a form of a final equation that can be verified by experiment.

The gas-to-cloth (G/C) ratio is a measure of the amount of gas passed through each square foot of fabric in the baghouse. It is given in terms of the number of cubic feet of gas per minute passing through one foot of cloth. In other words, the G/C ratio = gas volume rate/cloth area. Also note that this velocity is not the actual velocity through the openings in the fabric, but rather the apparent velocity of the gas approaching the cloth. As the G/C ratio increases, pressure drop (ΔP) also increases. In the United States, pressure drop in baghouse applications is generally measured in inches of water (in H_2O).

Despite several sophisticated formulas that have been developed, there is no satisfactory set of published equations that allows a designer to accurately calculate the efficiency of a prospective baghouse. However, there are three more heuristic formulas worth mentioning that can help baghouse designers:

$$\text{Gross gas/cloth ratio} = \frac{\text{total inlet gas volume rate}}{\text{total filter cloth area in collector}} \tag{16.24}$$

$$\text{Net gas/cloth ratio} = \frac{\text{total inlet gas volume rate} + \text{cleaning volume rate}}{\text{on stream cloth}} \tag{16.25}$$

$$\text{Units gas/cloth ratio} = \frac{\text{gas volume rate}}{\text{cloth area}} = \frac{\text{ft}^3/\text{min}}{\text{ft}^2} = \text{ft/min} \tag{16.26}$$

As mentioned earlier, baghouse design is still very much an artform and nowhere is this more evident than in the selection of G/C ratios. Factors influencing the G/C ratio include the cleaning method, filter media, particle size and distribution, particle density, particle loading, and other factors unique to each situation. Because of their variability, it has never been possible to satisfactorily quantify all of these factors for every application.

Once the G/C ratio, the cleaning method, and the filter medium have been selected, the essence of the flange-to-flange design selection process is complete. The only major consideration remaining is the baghouse material of construction. Typically, this is mild steel. In some specialty applications, stainless steel units are employed.

An equation that can be used for determining the collection efficiency of a baghouse is[14]

$$E = 1 - e^{-(\psi L + \phi t)} \tag{16.27}$$

where $\psi =$ constant based on fabric, ft^{-1}
        $\phi =$ constant based on cake, s^{-1}

t = time of operation to develop the cake thickness, s

L = fabric thickness, ft

E = collection efficiency, dimensionless

The exit concentration (w_e) for the combined resistance system (the fiber and the cake) is

$$w_e = w_i e^{-(\psi L + \phi t)} \tag{16.28}$$

where w_e = exit concentration, lb/ft^3

w_i = inlet concentration, lb/ft^3

A variation on Darcy's formula for the flow of fluid through a porous bed has been developed for the flow of gases through a filter medium. The basic Darcy equation can be used to predict the pressure drop for an operating fabric filter with accumulated cake:

$$\Delta P = S_E v + K_2 c_1 v^2 t \tag{16.29}$$

where ΔP = pressure drop, in H_2O

S_E = effective residual drag, in H_2O

v = velocity, fpm

K_2 = specific cake coefficient

The effect of bag failure on baghouse efficiency can be described by the following equation developed by Theodore and Reynolds:

$$P_t^* = P_t + P_{tc}$$

$$P_{tc} = \frac{0.582(\Delta P)^{0.5}}{\phi} \tag{16.30}$$

$$\phi = \frac{q}{LD^2(T + 460)^{0.5}}$$

where P_t^* = penetration after bag failure

P_t = penetration before bag failure

P_{tc} = penetration correction term; contribution of broken bags to P_t^*

ΔP = pressure drop, in. H_2O

ϕ = dimensional parameter (different from ϕ in Eq. 16.28)

q = volumetric flow rate of contaminated gas, acfm

L = number of broken bags

D = bag diameter, inches

T = temperature, °F

Refer to the literature for a detailed development of Equation (16.30).[15,16]

ILLUSTRATIVE EXAMPLE 16.12

A baghouse has been used to clean a particulate gas stream for nearly 30 years. There are 600 8-inch diameter bags in the unit and 50,000 acfm of dirty gas at 250°F enters the baghouse with a loading of 5.0 gr/ft^3. The outlet loading is 0.03 gr/ft^3. Local Environmental Protection Agency (EPA) regulations state that the outlet loading should not exceed 0.40 gr/ft^3. If the system operates at a pressure drop of 6.0 in H$_2$O, how many bags can fail before the unit is out of compliance? The Theodore–Reynolds equation applies and all contaminated gas emitted through the broken bags may be assumed the same as that passing through the tube sheet thimble.

SOLUTION: Calculate the efficiency E and penetration P_t before the bag failure(s):

$$E = [(\text{inlet loading}) - (\text{outlet loading})]/(\text{inlet loading})$$
$$= (5.0 - 0.03)/(5.0)$$
$$= 0.9940 = 99.40\%$$
$$P_t = 1 - 0.9940$$
$$= 0.0060 = 0.60\%$$

The efficiency and penetration P_t^* based on regulatory conditions are

$$E = (5.0 - 0.4)/5.0$$
$$= 0.920 = 92.0\%$$
$$P_t^* = 1 - 0.920$$
$$= 0.0800 = 8.00\%$$

The penetration term P_{tc} associated with the failed bags is then

$$P_{tc} = 0.0800 - 0.0060$$
$$= 0.0740$$

Since

$$P_{tc} = \frac{0.528(\Delta P)^{0.5}}{\phi} \tag{16.30}$$

and

$$\phi = \frac{q}{LD^2(T + 460)^{0.5}}$$

If neither the flow nor the concentration is uniformly distributed, the general equation developed in the previous section for the efficiency in a compartmentalized baghouse is used:

$$E = 1 - \frac{\sum c_i(1 - E_i)q_i}{\sum c_i q_i}$$

Substituting (see data), one obtains

$$E = 1 - \frac{(2)(3.8)(2500)(1 - 0.995) + (1)(4.25)(4000)(1 - 0.99)}{36,000}$$
$$= 0.9926 = 99.26\%$$

Note that the penetration has increased by slightly over 10%.   ∎

GAS–LIQUID (G–L) EQUIPMENT

The same equipment employed for G–S systems (not including venturi/wet scrubbers) may also be used for G–L systems. Gravity, centrifugal action, electrostatic effects, and filtering will perform the same separation function with G–L systems.

Gas–liquid phase separation is typically required following some other separation process such as absorption, evaporation, gas–liquid reactions, and condensation. For example, mist eliminators are almost always employed with absorbers (see Chapter 10). Before separation techniques or equipment can be selected, the parameters of the separation must be defined. Information on flow rates and flow ratios of the phases, and dispersed-phase particle size (droplet or bubble size) should be known or estimated.

The most common G–L separation processes usually involve the liquid dispersed in a gas such as the aforementioned absorbers as well as distillation columns, evaporators, or condensers. The usual objective is to prevent environmental emissions, downstream process contamination, or more importantly to recover a valuable material. Droplet size is the key to selecting the most effective separation technique, but size is difficult to measure. It must also be measured at the point where the separation is to be made owing to coalescence of collected particles. For new systems or where information of particle size and particle size distribution are not available, predictions or data from existing units can be used. Additional information is available in the literature.[17]

ILLUSTRATIVE EXAMPLE 16.13

A sodium hydroxide spray in air at 30°C is to be collected in a G–L separation unit that is essentially a gravity settler. The unit is 30 ft wide, 15 ft high, and 40 ft long. The volumetric flow rate of the gas is $42\,\text{ft}^3/\text{s}$. Calculate the smallest mist droplet (spherical in shape) that will be entirely recovered by the settler. The specific gravity of the mist droplets may be assumed to be equal to 1.21.

SOLUTION: The important property data are tabulated below:

$$\mu = 0.0185\,\text{cP} = 1.245 \times 10^{-5}\,\text{lb/ft}\cdot\text{s}$$

$$\rho = 0.0728\,\text{lb/ft}^3$$

Assume that Stokes' law applies. The describing equation is

$$d_p(\text{min}) = \left(\frac{18\mu q}{g\rho_p BL}\right)^{0.5} \tag{16.9}$$

Substituting the data, one obtains

$$d_p(\text{min}) = \left[\frac{18(1.245 \times 10^{-5}\,\text{lb/ft}\cdot\text{s})(42\,\text{ft}^3/\text{s})}{(32.2\,\text{ft/s}^2)(1.21 \times 62.4\,\text{lb/ft}^3)(30\,\text{ft})(40\,\text{ft})}\right]^{0.5}$$

$$= 5.68 \times 10^{-5}\,\text{ft}$$

The reader should verify that the Stokes' law assumption is valid. The reader may also choose to note the similarities between this example and Illustrative Example 16.6.   ∎

ILLUSTRATIVE EXAMPLE 16.14

Calculate the smallest droplet in Illustrative Example 16.13 assuming the flow rate is halved.

SOLUTION: If q is half of that in the previous example, then simply divide the answer by $\sqrt{2}$; the basis for this is provided below.

$$d_p(\min_1) = \left(\frac{18\mu q_1}{g\rho_p BL}\right)^{0.5}$$

$$d_p(\min_2) = \left(\frac{18\mu q_2}{g\rho_p BL}\right)^{0.5}$$

with

$$q_2 = \tfrac{1}{2}q_1$$

Substituting above yields

$$d_p(\min_2) = \left(\frac{18\mu q_1}{2g\rho_p BL}\right)^{0.5} = \frac{1}{\sqrt{2}}\left(\frac{18\mu q_1}{g\rho_p BL}\right)^{0.5}$$

Therefore,

$$d_p(\min_2) = \frac{d_p(\min_1)}{\sqrt{2}}$$

$$= \frac{5.68 \times 10^{-5}\text{ ft}}{\sqrt{2}} = 4.016 \times 10^{-5}\text{ ft}$$

∎

ILLUSTRATIVE EXAMPLE 16.15

Two mist eliminators operate in series with droplet recovery efficiencies of 90% and 99.5%, respectively. Calculate the overall efficiency of the two units.

SOLUTION: This problem is best solved by noting the overall penetration given by the product of the fractional penetration for each device where

$$\text{For two units, } P = P_1 P_2; \quad P = 1.0 - E \tag{16.31}$$

According to the data given

$$P_1 = 1 - 0.9$$
$$= 0.1$$

and

$$P_2 = 1 - 0.995$$
$$= 0.005$$

Therefore,

$$P = (0.1)(0.005)$$
$$= 0.0005; \quad \text{fractional basis}$$
$$= 0.05; \quad \text{percent basis}$$

and

$$E = 1 - 0.0005 = 0.9995 = 99.95\% \qquad \blacksquare$$

LIQUID–SOLID (L–S) EQUIPMENT

Liquid–solid separation that occurs under the action of gravity can be divided into two categories: Type I and Type II. Type I settling involves dilute mixtures where one can assume no solid particle-to-solid particle interaction. Type II settling involves mixtures that are sufficiently concentrated with the solids that they effectively settle as a mass, as with bulk filtration. Additional details follow.

For Type I settling, all particles settle independently; consequently, if a particle size and size distribution is known, the settling rate of an individual particle can be calculated. The L–S unit can then be designed employing the same procedures provided for G–S equipment. Type II settling behavior generally produces a rather sharp line of demarcation between the clear liquid overflow and the settling solids. For these systems, the design is primarily dictated by the thickening capability of the unit, although the design must be adequate to provide sufficient overflow area to clarify the liquid overflow. Settlers for this category of mixtures are normally referred to as thickeners. They are sometimes constructed as a rectangular basin; however, they are most often of circular cross-section.

This section examines three industrial separation techniques that exploit the density difference between a liquid and a solid. The driving force in these processes is usually the result of gravity, centrifugal action, and/or buoyant effects. Unlike the earlier section that treated gas-particle separation, this presentation will address topics keying on liquid–solid separation. Filtration is treated briefly at the end of this section.

Sedimentation

Gravity sedimentation is a liquid–solid process that separates, under the effect of gravity, a feed slurry into an underflow slurry of higher solids concentration and an overflow of substantially clearer liquid. A difference in density between the solids and the suspended liquid is, as indicated, a necessary prerequisite.

Nearly all commercial equipment for continuous sedimentation is built with relatively simple settling tanks. Distinction is commonly made depending on the purpose of the separation. If the clarity of the overflow is of primary importance, the process is called *clarification* and the feed slurry is usually dilute. If a thick underflow is the primary aim, the process is called *thickening* and the feed slurry is usually more concentrated.[18]

The most commonly used thickener is the circular-basin type (shown in Fig. 16.7). The flocculant-treated feedstream enters the central feed well, which
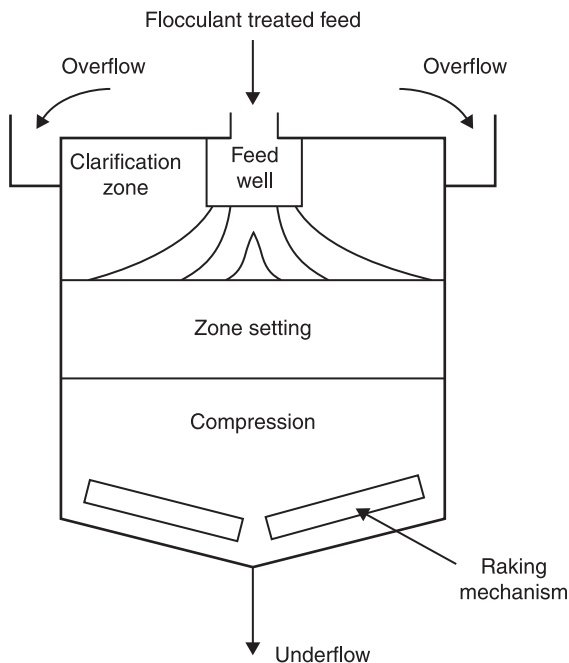
Figure 16.7   The circular-basin continuous thickener.

dissipates the stream's kinetic energy and disperses it gently into the thickener. The feed finds its height in the basin where its density matches the density of the suspension inside (the concentration increases downward in an operating thickener, giving stability to the process) and spreads out at that level. The settling solids move downward as does some liquid, with the liquid amount being determined by the underflow withdrawal rate. Most of the liquid goes upward and into the overflow. Typically, a thickener has three operating zones: *clarification*, *zone settling*, and *compression* (see Fig. 16.7).[18]

Gravity clarifiers sometimes resemble circular thickeners but more often are rectangular basins with the feed at one end and overflow at the other. Settled solids are pushed to a discharge trench by paddles or blades on a chain mechanism. Flocculent may be added prior to the clarifier. Conventional thickeners are also used for clarification; however, the typically low feed concentrations hinder the benefits of zone settling and so the basin area is based on clarification-zone demands.[18] Thus, in a general sense, sedimentation is the process of removing solid particles heavier than water by gravity settling. It is the oldest and most widely used unit operation in water and wastewater treatment. Note that the terms sedimentation, settling, and clarification are often used interchangeably; the unit sedimentation basin may also be referred to as a sedimentation tank, clarifier, settling basin, or settling tank.

In wastewater treatment, sedimentation is used to remove both inorganic and organic materials that are settleable in continuous flow conditions. It removes grit,

particulate matter in the primary settling tank, and chemical flocculants from a chemical precipitation unit. Sedimentation is also used for solids concentration in sludge thickeners.[19]

Details on discrete, or individual, particle settling was presented earlier in this chapter. Here, Stokes' law assumes that particles are present in relatively dilute concentrations and in a fluid medium of relatively large cross-section. When there are a large number of particles, the particles in close proximity will retard other particles. This is termed *hindered settling*. The effect is not significant at volumetric concentrations below 0.1%. However, when the particle diameter becomes appreciable with respect to the diameter of the container in which it is settling, the container walls will exert an additional retarding effect known as the *wall effect*.

Farag[20] has reviewed and developed equations for the above situation. For hindered flow, the settling velocity is less than would be calculated from Stokes' law. The density (employing SI units) of the fluid phase becomes the bulk density of the slurry, ρ_m, which is defined as follows:

$$\rho_m = \varepsilon_f \rho_f + (1 - \varepsilon_f)\rho_s \tag{16.32}$$

where ρ_m is the density of slurry in kg (solid plus liquid)/m^3 and ε_f is the volume fraction of the liquid in the slurry mixture. The density difference is then:

$$\rho_p - \rho_m = \rho_p - [\varepsilon_f \rho_f + (1 - \varepsilon_f)\rho_p] = \varepsilon_f(\rho_p - \rho_f) \tag{16.33}$$

The effective viscosity of the mixture, μ_m, is defined as:

$$\mu_m = \mu_f b \tag{16.34}$$

where μ_f is the pure fluid viscosity and b is a dimensionless correction factor that is a function of ε_f. The term b can be evaluated from Equation (16.35)

$$b = 10^{1.82(1-\varepsilon_f)}$$
$$\log_{10} b = 1.82(1 - \varepsilon_f) \tag{16.35}$$

The settling velocity, v (with respect to the unit) is ε_f times the velocity calculated by Stokes' law. Substituting Equations (16.33) and (16.34) into Stokes' law terminal velocity equation and multiplying by ε_f for the relative velocity effect leads to:

$$v = \frac{gd_p^2(\rho_p - \rho_f)}{18\mu_f}\frac{\varepsilon_f^2}{b} \tag{16.36}$$

The Reynolds number, based on the settling velocity relative to the fluid is then:

$$\mathrm{Re} = \frac{\rho_m v d_p}{\mu_m \varepsilon_f} = \frac{gd_p^2(\rho_p - \rho)\rho_m}{18\mu_f^2}\frac{\varepsilon_f}{b^2} \tag{16.37}$$

When the Reynolds number is less than 1, the settling is in the Stokes' law regime.

When the diameter, d_p, of the particle becomes appreciable with respect to the diameter of the container, D_w, the terminal velocity is reduced. This is termed the aforementioned wall effect. In the case of settling in the Stokes' law regime, the calculated terminal velocity is multiplied by a correction factor, k_w.[20]

ILLUSTRATIVE EXAMPLE 16.16

Glass spheres are settling in water at 20°C. The slurry contains 60 wt% solids and the particle diameter is 0.1554 mm. The glass density is 2467 kg/m³. Find the volume fraction, ε_f, of the liquid and the bulk density of the slurry, ρ_m. For water at 20°C, $\rho_f = 998$ kg/m³ and $\mu_f = 0.001$ kg/m · s.

SOLUTION: Start by assuming a basis of 100 kg of slurry. To determine the volume fraction of the liquid, divide the mass of liquid by its density. Since 100 kg of slurry is the basis and the slurry contains 60 wt% solid,

$$m_f = 40 \, \text{kg}$$

The volume of the fluid (water) is

$$V_f = \frac{m_f}{\rho_f} = \frac{40}{998} = 0.040 \, \text{m}^3$$

Similarly,

$$m_p = 60 \, \text{kg}$$

$$V_p = \frac{m_p}{\rho_p} = \frac{60}{2467}$$

$$= 0.0243 \, \text{m}^3$$

Therefore,

$$V = V_f + V_p = 0.040 + 0.0243$$

$$= 0.0643 \, \text{m}^3$$

and

$$\varepsilon_f = \frac{V_f}{V} = \frac{0.040}{0.643}$$

$$= 0.622$$

and

$$\varepsilon_p = 1 - \varepsilon_f$$

$$= 0.378$$

Calculate ρ_m from Equation (16.32).

$$\rho_m = \varepsilon_f \rho_f + \varepsilon_p \rho_p$$

Substituting,

$$\rho_m = 0.622(998) + 0.378(2467)$$

$$= 1553 \, \text{kg/m}^3 \qquad ■$$

ILLUSTRATIVE EXAMPLE 16.17

Refer to Illustrative Example (16.16). Calculate the Reynolds number.

SOLUTION: Employ Equation (16.37) to obtain:

$$\text{Re} = \frac{\rho_m v d_p}{\mu_m \varepsilon_f}$$

Substituting,

$$\text{Re} = \frac{(1553)(0.00153)(1.554 \times 10^{-4})}{(0.0049)(0.622)}$$

$$= 0.121$$

∎

Centrifugation

Centrifugal force is widely used when a force greater than that of gravity is desired for the separation of solids and fluids of different densities, as in settling or for other forms of separation. Two terms need to be defined. A centrifugal force is created by moving a mass in a curved path and is "exerted" in the direction away from the center of curvature of the path. As noted earlier, centripetal force is the force applied to the moving mass in the direction toward the center of curvature that causes the mass to travel in a curved path.

Centrifugation is therefore another process that uses density differences to separate solids from liquids (or an immiscible liquid from other liquids). The feed is subjected to forces that make the solids move radially through the liquid (outward if heavier, inward if lighter). In a sense, centrifugation is an extension of gravity sedimentation to particle sizes and to emulsions that are normally stable in a gravity field. The describing equations developed in an earlier section for particle motion again apply. The gravity force is replaced by a centripetal force, F_c (force/mass)

$$F_c = \frac{r\omega^2}{g_c} = \frac{v_t^2}{g_c r} \tag{16.38}$$

where r is the radius of curvature of the particle or heavier phase, ω the angular velocity and v_t the tangential velocity at the point in question.

As indicated above, centrifugation attempts to increase the particle "settling" velocity many times higher than that of gravity by applying a centripetal force. The ratio of these two forces has been defined by some as the number of "Gs," where

$$G = \frac{r\omega^2}{g} \tag{16.39}$$

Note that the g term in the denominator in Equation (16.39) is the acceleration due to gravity and not the term g_c (the conversion constant) that appears in Equation (16.38).

There are two main classes of centrifugation equipment: cyclones and centrifuges. Cyclones are generally classified as air particulate control/recovery equipment; this unit is primarily used for separating gas–solid systems. Centrifuges are primarily employed for liquid–solid separation; units in this category include basket, tubular,

scroll-type, dish, and multiple chamber. The units may operate in the batch or continuous mode. Details on this equipment are available in the literature.[20–22]

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ILLUSTRATIVE EXAMPLE 16.18

</div>

A particle is spinning in a 3-inch ID (inside diameter) centrifuge with an angular velocity of 30 rad/s. Calculate the number of Gs for the particle.

SOLUTION: Employ Equation (16.39).

$$G = \frac{r\omega^2}{g}$$

Substituting yields:

$$\frac{(3/12)(30)^2}{32.2} = 7.0$$

■

Flotation

Flotation processes are useful for the separation of a variety of species, ranging from molecular and ionic to microorganisms and mineral fines, from one another for the purpose of extraction of valuable products, as well as cleaning of waste waters. They are particularly attractive for separation problems involving very dilute solutions where most other processes usually fail. The success of flotation processes is primarily dependent on the tendency of surface-active species to concentrate at the water–fluid interface and on their capability to make selected non-surface-active materials hydrophobic by means of adsorption on them or association with them. Under practical conditions, the amount of interfacial area available for such concentration is increased by generating air bubbles or oil droplets in the aqueous solution.[22]

Flotation is a gravity separation process based on the attachment of air or gas bubbles to solid (or liquid) particles that are then carried to the liquid surface where they accumulate as float and can be skimmed off. The process consists of two stages: the production of suitably small bubbles and their attachment to the particles. Depending on the method of bubble production, flotation is classified as dissolve-air, electrolytic, or dispersed-air, with dissolve-air primarily employed by industry.[18]

The separation of solid particles into several fractions based upon their terminal velocities is called hydraulic classification. If there are particles of materials of different densities but of the same size, these may also be separated by the method of hydraulic separation or classification, i.e., by placing the particles of different densities in an upward-flowing stream of fluid (the fluid is often water). If the velocity of the water is adjusted so that it lies between the terminal falling velocities (or settling velocities) of the two particles, the slower particles will be carried upward and the particles of higher terminal velocity than the water velocity will move downward, and a separation is thereby attained.

ILLUSTRATIVE EXAMPLE 16.19

It is desired to separate quartz particles (specific gravity = 2.65, with a size range of 40–90 μm) from galena particles (specific gravity = 7.5, with a similar size range of 40–90 μm). The mixture is placed in a rising water flow (density, $\rho = 1000$ kg/m³; viscosity, $\mu = 0.001$ kg/m · s). See Figure 16.8. Calculate the water velocity to obtain pure galena. Will this pure galena be a top or bottom product?

SOLUTION: Calculate the settling velocity of the largest quartz particle with a diameter $d_p = 90$ μm. Assume Stokes' law flow regime applies. Therefore, from Equation (16.6), with the fluid density term ρ_f retained,

$$v_q = \frac{g d_p^2 (\rho_s - \rho_f)}{18 \mu_f} = \frac{9.807(9 \times 10^{-5})^2 (2650 - 1000)}{18(0.001)}$$

$$= 0.0073 \, \text{m/s}$$

Calculating the settling velocity of the smallest galena particle with a diameter $d_p = 4 \times 10^{-5}$ m,

$$v_g = \frac{g d_p^2 (\rho_s - \rho_f)}{18 \mu_f} = \frac{9.807(4 \times 10^{-5})^2 (7500 - 1000)}{18(0.001)} = 0.00567 \, \text{m/s}$$
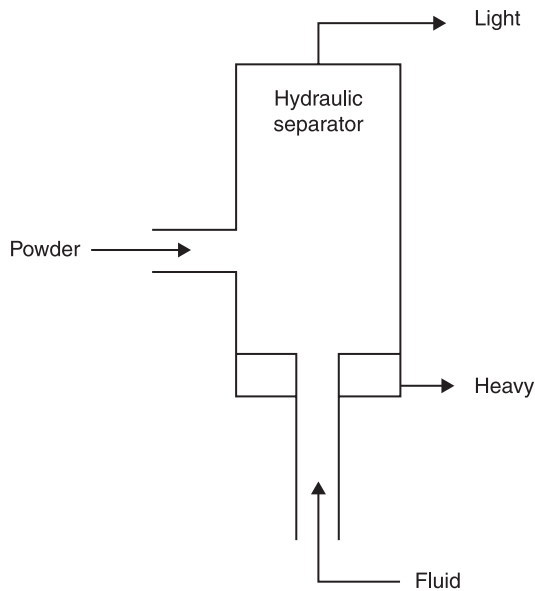
Figure 16.8 Particle separator.

To obtain pure galena, the upward velocity of the water must be equal to or greater than the settling velocity of the largest quartz particle. Therefore,

$$v_w = 0.0073 \, \text{m/s} = 7.3 \, \text{mm/s}$$

Since the water velocity of 7.3 mm/s is greater than the settling velocity of the smallest galena particle, some galena will be washed up with the quartz. One may conclude that pure galena will be the bottom product; the top product will be the quartz plus some galena. ∎

ILLUSTRATIVE EXAMPLE 16.20

Refer to Illustrative Example 16.19. Determine the size range of the galena in the top product.

SOLUTION: To determine the size range of the galena product, calculate the galena particle size that has a settling velocity of 7.3 mm/s. Assume Stokes' law applies

$$v = \frac{g d_p^2 (\rho_s - \rho_f)}{18\mu} \tag{16.6}$$

$$d_p = \sqrt{\frac{18\mu_f v}{g(\rho_s - \rho_f)}} = \sqrt{\frac{18(0.001)(0.0073)}{9.807(7500 - 1000)}} = 4.54 \times 10^{-5} \, \text{m} = 45.4 \, \mu\text{m}$$

The size ranges for the galena are therefore 40–45.4 μm for the top washed product. ∎

Units other than the traditional settler or L–S variety also find application. Dissolved air flotation is another phase separation process used to remove organic or inorganic solids suspended in feed streams such as waste-waters or slurries. The process effectively concentrates the feed by forming a sludge to remove valuable or toxic solids. Since chemical reagents are sometimes used to remove specific inorganic solids, the process can be chemical as well as physical.

Flocculation and precipitation are physicochemical processes. In *flocculation*, fine suspended particles, which are difficult to settle out of the liquid, are brought together by flocculating chemicals to form larger, more easily collected particles. In *precipitation*, metals soluble in the liquid are made insoluble by use of precipitating chemicals. The solids removed by these two processes form a sludge that may be disposed of or treated to recover any valuable material. The operating costs for both processes are dependent on the amount and type of flocculating and precipitating chemicals used.

Filtration is a popular liquid–solid separation process commonly used in industry. In wastewater treatment, it is used to purify the liquid by removing suspended solids. This is usually followed by flocculation or sedimentation for further solids removal. In sludge treatment, it is used to remove the liquid (*sludge dewatering*) and concentrate the solid, thereby reducing the sludge volume. This method is highly competitive with other sludge dewatering processes. Filtration may also be used in treating non-aqueous liquid streams.[5]

In the filtration process, a liquid containing suspended solids is passed through a porous medium. The solids are trapped against the medium and the separation of solids from the liquid results. A filter press is the unit of choice in many applications, particularly if a Type II L–S system is to be treated. For large solid particles, a thick barrier such as sand may be used. This is known as *granular media* filtration. For smaller particles, a fine filter such as a filter cloth may be used. Fluid passage may be induced by gravity, positive pressure, or a vacuum. The filter is cleaned and the solids collected by passing a stream of liquid (often water) in the opposite direction of the stream flow; this is known as *backwashing*. Additional details are available in the literature.[5]

Finally, the size and handling characteristics of the settled solids also have an impact on separator selection, design, and performance. The desired clarity of the liquid and the dryness of the solids are important additional selection criteria. For applications requiring the driest solids, a filtering centrifuge or filter press are the logical choices. Regarding equipment selection, it should be noted that for almost every application, more than one type of unit will accomplish the job. The selection procedure usually involves a study of the characteristics of the solids, process requirements, and consultation with several equipment manufacturers.

LIQUID–LIQUID (L–L) EQUIPMENT

As noted earlier, the applicable theory, equipment, and operation of most phase separation processes are very similar regardless of whether the individual phases are gas, liquid, or solid. However, the separation of immiscible liquid phases (L–L) is different when compared to the others in that the important density difference between the phases is usually very small. Gas–solid (G–S), liquid–solid (L–S), or liquid–gas (L–G) separations typically have significant phase density differences. The separation of phases with a small density difference almost always requires large equipment, application of large forces, unique equipment geometries, or all of the above. As with G–S, L–S, and L–G separation processes, the common mechanisms are gravity, centripetal force, impaction, and electrostatic effects. The design for L–L separations is affected by temperature, pressure, presence of contaminants such as surfactants, and stream mixing effects.

In order to design an immiscible liquid separation unit, the average drop size and droplet size range of the dispersed phase should be specified or approximated.

The most common unit employed is a gravity settler (often referred to as a *decanter*) and may be operated in a batch or the more popular continuous mode. The unit may be designed vertically or horizontally. The vertical unit is rarely employed in practice.

The horizontal unit is the usual choice for L–L separation with long vessels the most desirable geometry. In these units, the continuous phase (it may be either the lighter or heavier phase) flows horizontally and perpendicular to the settling phase. Turbulence and turbulent mixing is a problem, but it can be reduced or eliminated by reducing the superficial velocity of the continuous phase. The droplet settling
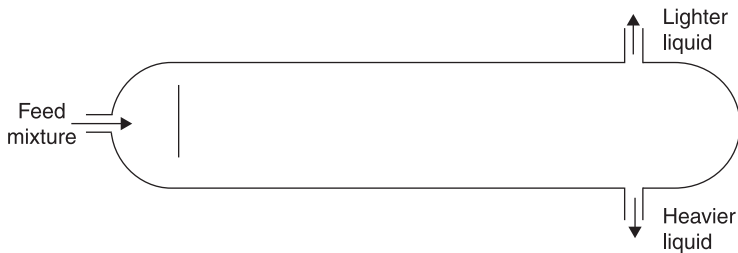
Figure 16.9   A horizontal decanter.

velocity is estimated from any of the equations provided in the Fluid–Particle Dynamics section at the beginning of this chapter but it is usually safe to assume Stokes' law applies.

A basic decanter layout is shown in Figure 16.9. The proper introduction of feed to a gravity separation is one of the keys to its performance. For optimum design, the feed should be:

1 introduced uniformly across the active cross-section of the decanter, and

2 accomplished in a manner so as to leave no residual jetting or turbulence.

Forcing a direction change on the inlet flow can reduce the jet effect; a simple approach is to use a plate just inside the inlet at least two times the nozzle diameter and located one-half nozzle diameter in from the inlet.

Four important decanter design factors include the following:

1 the velocities of both phases should be approximately equal

2 the feed mixture velocity should be below 2 ft/s

3 the flow for both phases should be laminar

4 the discharge velocity should not exceed ten times the superficial velocity in the separator

The time for separation to occur (the "settling" time, either up or down) is obviously critical to the design, operation, and performance of the decanter. The afore-mentioned difficulty in accurately specifying droplet size(s) makes this calculation almost impossible. However, if this information is required, a conservative droplet size of 100 μm (microns) is recommended. As noted above, the droplet "settling" velocity may then be estimated from any of the equations provided in the Fluid–Particle Dynamics section at the beginning of this chapter, and one can usually safely assume Stokes' law applies.

ILLUSTRATIVE EXAMPLE 16.21

Refer to Figure 16.9. If the volumetric feed rate is 26.7 ft^3/h, calculate the length, L, of a decanter, 1.2 ft in diameter, to ensure an average residence time of 18 min.

SOLUTION: The volume, V, of the decanter is

$$V = \left(\frac{\pi D^2}{4}\right) L$$

$$= (0.785)(1.2)^2 L$$

$$= 1.13 L$$

The average residence time, t, is

$$t = \frac{V}{q}$$

$$= \frac{1.13 L}{26.7}$$

$$= 0.0423 L; \text{ h}$$

Set t equal to the required residence time.

$$(18/60) = 0.0423 L$$

$$L = 7.09 \text{ ft}$$ ■

SOLID–SOLID (S–S) EQUIPMENT

This type of phase separation finds limited application in practice. There are two units that have received attention: the high-gradient magnetic separation (HGMS) device and solidification. Each are briefly discussed below.

High-Gradient Magnetic Separation

High-gradient magnetic separation (HGMS) is a phase and component separation process still in the development stage. It could be used to remove magnetic or non-magnetic materials from a variety of feeds such as slurries, sludges, and solids plus aqueous and non-aqueous liquids. Its potential use lies in the removal of ferromagnetic and paramagnetic particulates. It is being used on a small scale in wastewater treatment and coal desulfurization. The process works well when magnetic components make up only a small percentage of the total concentration, and solids make up 10–15% of the total waste stream.

The HGMS process uses a magnetized ferromagnetic filter to separate magnetic or paramagnetic particles from nonmagnetic particles. The filter captures the magnetic material, which is recovered when the filter is cleaned. The filter can also be used to remove nonmagnetic material by treating the feed with a magnetic seed.

Capital investment may be low when only ferromagnetic material is removed but higher for the removal of nonmagnetic material. High-volume applications make the process more economically attractive.

Solidification

Solidification is a process that transforms a feed into a solid product by fixation or encapsulation. In *fixation*, a chemical or physical process and a solidifying agent

are used to solidify the feed. *Encapsulation* is a process in which the feed is surrounded by a binder after it has been solidified by a chemical agent. Both produce a durable and impermeable product.

Solidification has only recently become a popular form of feed treatment. Previously, it had been used solely for treating radioactive wastes because of its high cost compared to other disposal means. Stricter regulations have made solidification a more popular choice of feed treatment. Solidification processes may be separated into five general categories depending on the solidifying agent used:

1 silicate- and cement-based

2 lime-based

3 thermoplastic-based

4 organic polymer-based

5 encapsulation techniques

All five of these processes are used for feed treatment. The cement- and lime-based methods are also used for sludges from stack-gas scrubbers after the sludges are chemically treated to precipitate out the metals. Solidification is unsuitable for treating organics, oxides, and toxic anions because these materials are difficult to solidify. Solidification has been used in treating feeds produced from steel mills, plating and lead-smelting plants, food production sludges, and sulfur residues.

The *silicate-* and *cement-based* methods may be operated as batch or continuous processes. Portland cement and other additives are added to either a wet or sludge-like stream forming an impermeable, rock-like solid. The solidified product may then be used for land reclamation. The degree of solidification and strength of the product depends on the concentration of metals and organics present in the feed. This particular solidification process is the one most used in the United States. Some of its advantages are that the additives are not expensive, the process is well developed, equipment is available, and end product strength and permeability are easily controlled by the amount of cement used. The disadvantages are that feed weight and bulk are increased by the cement, and pretreatment or special additives may be needed to ensure proper solidification.

The *lime-based* process is both chemical and physical. Lime, water, and siliceous material are added to the feed. A product known as *pozzolanic cement* is formed. Additives may be added to increase the strength of the end product. The degree of solidification depends on the reaction of the lime with the other components forming the cement. Once the product is formed, it is often used for landfill, mine reclamation fill, or capping material. The advantages and disadvantages of this process are similar to those of the cement-based process.

Thermoplastic solidification is a two-step process. Initially, the feed, which is first dried, is combined with paraffins, bitumen, and polyethylene at a temperature above 212°F. Upon cooling, the mixture begins to solidify. In the second step, the solid waste is thermoplastically coated and then ready for disposal. The thermoplasticized product is fairly permeable to most aqueous solutions. In addition to toxic inorganics, the process is also used in treating nuclear wastes. The equipment used in

thermoplastic solidification is more specialized and therefore more costly than the equipment used in the cement- or lime-based processes. On the negative side, thermoplastics are flammable and the feed must be dried before processing takes place.

In the *organic polymer* process, a monomer and a catalyst are combined with the feed and the polymer is allowed to form. The product is then containerized and ready for disposal. One popular technique is the urea–formaldehyde process that is used in treating nuclear wastes. Two advantages of the organic polymer process are that it may be used on wet or dry streams and that the end product weighs less than those produced from the cement-, lime-, or thermoplastic-based techniques. The disadvantages are that some organic polymers are biodegradable, the end product must be containerized, and the end product may not be as durable as the end products produced from the other methods.

The *encapsulation* process is somewhat similar to the thermoplastic process. The dried feed is first chemically treated and then coated with a binder, usually polyethylene. The advantages of this process are that the encapsulated feed is very durable and resistant to water and deterioration, and the final product does not need to be stored in containers. The disadvantages are that the process is expensive, only small volumes of feed can be treated, and the sludge must be dried before processing.

The various solidification treatments may be performed on-site, which requires a large investment, or at regional facilities, which requires transportation costs. As with any other option, the choice and final cost of the solidification process will depend upon the type of feed involved, the desired end product, and the investment capabilities of the user.[23]

Finally, the separation of particle sizes using the methods described earlier can also be employed to separate particles of different (chemical) composition. This separation technique can only be employed if the solids in question have different densities but are of the same size.

REFERENCES

1. L. THEODORE, "*Air Pollution Control Equipment Calculations*," John Wiley & Sons, Hoboken, NJ, 2008.
2. L. THEODORE: personal notes, 1982.
3. L. THEODORE: personal notes, 1976.
4. L. THEODORE, "*Nanotechnology: Basic Calculations for Engineers and Scientists*," John Wiley & Sons, Hoboken, NJ, 2006.
5. J. ABULENCIA and L. THEODORE, "*Fluid Flow for the Practicing Engineer*," John Wiley & Sons, Hoboken, NJ, 2009.
6. L. THEODORE: personal notes, 1960.
7. L. THEODORE: personal notes, 1966.
8. C. E. CUNNINGHAM, *Proc. Roy. Soc. London, Ser. A*, **83**, 357, 1910.
9. C. LAPPLE, "*Air Pollution Engineering Manual*," USEPA, AP-40, 94–99, 1951.
10. C. LAPPLE, *Chem. Eng.*, New York City, NY, **58**, 144, 1951.
11. L. THEODORE, "Engineering calculations: cyclone collection systems," *CEP*, New York City, NY, Sept. 18–20, 2005.
12. L. THEODORE and V. DEPAOLA, "Predicting cyclone performance," *J. Air Poll. Control Assoc.*, Pittsburg, PA, **30**, 1132–3, 1980.

13. H. JOHNSTONE et al., *Ist. Chem. Eng.*, New York City, NY, **46**, 1601, 1954.

14. L. THEODORE: personal notes, 1981.

15. L. THEODORE and J. REYNOLDS, "Effect of bag failure on baghouse outlet loading," *J. Air Poll. Control Assoc.*, Pittsburg, PA, 870–872, August 1979.

16. L. THEODORE, "Engineering calculations: baghouse specification and operation simplified," *CEP*, New York City, NY, 22, June, 2000.

17. D. GREEN and R. PERRY, "*Perry's Chemical Engineers' Handbook*," 8th edition, McGraw-Hill, New York City, NY, 2008.

18. L. SVAROUSKY, "Sedimentation, centrifugation and flotation," *Chem. Eng.*, New York City, NY, July 16, 1979.

19. S. LIN, "*Water and Wastewater Evaluations Manual*," McGraw-Hill, New York City, NY, 2001.

20. I. FARAG, "*Fluid Flow*," A Theodore Tutorial, Theodore Tutorials, East Williston, NY, 1996.

21. J. REYNOLDS, J. JERIS, and L. THEODORE, "*Handbook of Chemical and Environmental Engineering Calculations*," John Wiley & Sons, Hoboken, NJ, 2002.

22. R. ROUSSEAU (ed), "*Handbook of Separation Process Technology*," John Wiley & Sons, Hoboken, NJ, 1987.

23. R. B. POJASEK, "Solid-waste disposal: solidification," *Chem. Eng.*, New York City, NY, 140–145, August 13, 1979.

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