CHAPTER 7.2

Column Flotation

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Froth flotation has been used for more than a century to efficiently and economically upgrade fine particles produced by the mining industry. The majority of industrial installations use conventional flotation machines that consist of a series of agitated tanks through which feed slurry is passed. Gas is introduced through a rotor–stator assembly that disperses air bubbles, suspends particles, and promotes bubble–particle collisions. Unfortunately, conventional flotation machines suffer from the contamination of froth products by the hydraulic entrainment of fine gangue particles. This problem is particularly severe when large amounts of ultrafine slimes are present in the feed stream. Nonselective entrainment has historically been minimized through the use of cleaner flotation circuits in which the froth product is repeatedly diluted with clarified water and refloated in multiple stages. This brute force approach of dealing with unwanted entrainment adds to both the cost and complexity of flotation plants.

In the early 1960s, an improved method of minimizing froth entrainment was patented by Canadian researchers Pierre Boutin and Remi Tremblay (1967). This invention consisted of a column-type flotation machine that introduced a countercurrent flow of wash water into a deep froth phase. The wash water practically eliminated the hydraulic entrainment of ultrafine slimes and substantially improved the purity of the froth product using only a single stage of flotation. Dobby (2002) has reported that the inventors conceived of column flotation as an analogy to solvent-in-pulp columns employed in the solvent extraction of uranium ore slurries. In solvent extraction, aqueous diluent is added to the top of the extraction column and allowed to flow countercurrently to rising droplets of solvent to reduce contamination of the solvent phase. The inventors first demonstrated column flotation technology for the removal of fine silica from iron ore using a 5-cm-diameter laboratory column and a 0.3-m-diameter pilot-scale column (Wheeler 1983). The technology was later demonstrated in Canada for copper sulfide flotation at the Opemisca Company concentrator using a 0.2-m^2 test column (Rubinstein 1995).

During the 1970s, numerous other laboratory and pilot-scale field trials were conducted by a newly formed company called the Column Flotation Company of Canada Ltd. (Wheeler 1983). This group, created and led by Don Wheeler, is credited with many of the early technical advancements associated with column technology. However, it was not until 1981 that the first industrially accepted production column was installed by Noranda Mine's Les Mines Gaspé concentrator within a molybdenum cleaner circuit (Coffin and Miszczak 1982). By 1987, the two-stage column circuit installed in this operation had replaced as many as 13 stages of conventional flotation machines (Finch and Dobby 1990). This success helped to spearhead the wider commercial acceptance of column technology as a standard unit operation in many flotation plants throughout the processing industry.

The rapid emergence of column flotation can be demonstrated from the number of scholarly works published on the topic since 1960. As shown in Figure 1, the number of publications associated with column technology increased rapidly during the late 1980s and early 1990s. This interest eventually peaked with the advent of the Canadian Institute of Mining, Metallurgy and Petroleum (CIM) symposium on column flotation in 1996. This was shortly after publication of the landmark book on the subject titled *Column Flotation* by James Finch and Glenn Dobby (1990). This text continues to be considered the most comprehensive source of both fundamental and applied information on column flotation technology. Scholarly interest in column technology waned after the CIM Column '96 symposium, which was largely in response to industry acceptance of the technology as a standard unit operation. Publication frequency started to rise again after 2009, largely because of a resurgence of new studies on column flotation in countries outside North and South America.

TECHNOLOGY

As the name implies, a column flotation machine (Figure 2) consists of a tubular tank with a relatively large height-to-diameter ratio. During operation, reagentized feed slurry

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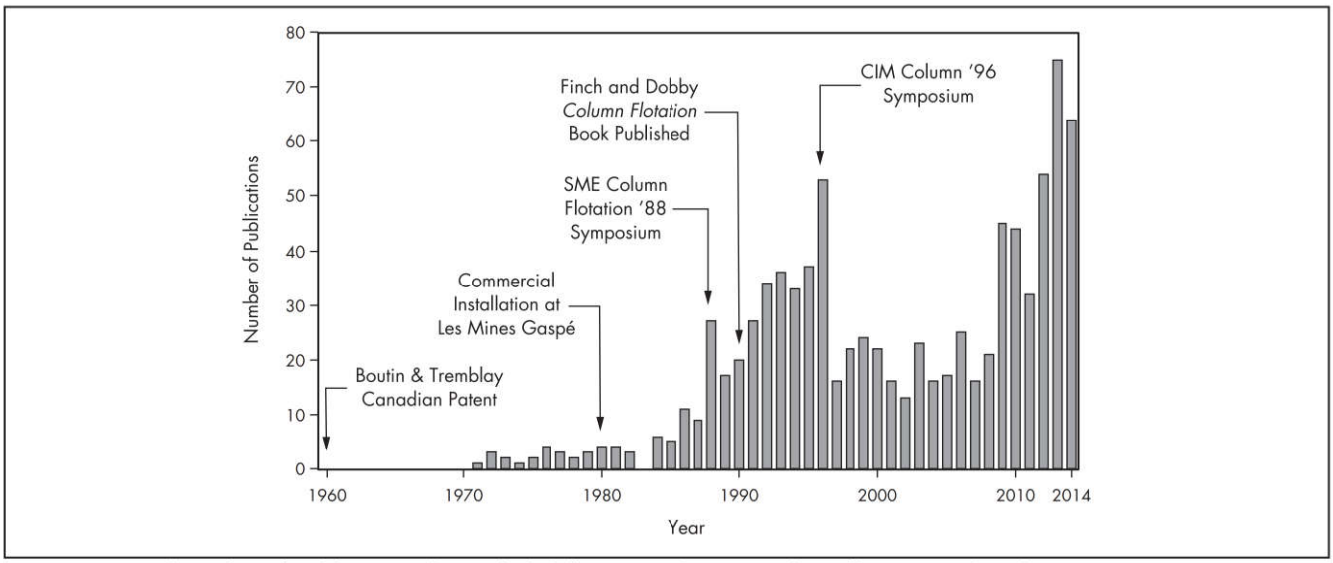


Figure 1 Annual number of publications that included *flotation column* or *column flotation* in the title

is introduced into the upper portion of the column though a header system that distributes the slurry over the column cross section. The slurry flows by gravity down through the column to a discharge port that regulates the underflow via an actuated control valve. The valve is normally configured to hold a pulp level sufficient to create 0.5–1.5 m of froth depth. Compressed air, mixtures of air–slurry or air–water, or other sources of gases are injected into the bottom of the column through various types of gas sparging systems. This arrangement ensures that the feed slurry moves in a countercurrent direction to rising gas bubbles. Hydrophobic particles collide and attach to the bubbles and are carried to the top of the column where a froth bed is formed. Wash water is then added to the froth bed through a distribution system that evenly spreads water over the entire cross section of the froth column. Sufficient wash water is added to prevent fine hydrophilic particles from entering the froth bed. The froth overflows into a collection launder where it reports as a high-grade hydrophobic product. Traditionally, columns do not make use of either an internal rotor–stator mechanism for gas dispersion/particle suspension or paddle mechanisms for froth removal. Gas compressors are normally required, however, to provide the gas flow and energy required for gas dispersion.

COLUMN ZONES

Up to three different aeration zones can be identified within a flotation column (Figure 3). The pulp zone is the region in the column in which bubbles and particles are brought together to form stable aggregates. This zone is located between the air injection point and the pulp level. The gas holdup (ε_p) in the pulp is typically 10%–20% by volume. Particles that do not become attached to bubbles in this zone are eventually discharged from the bottom of the column. The bubble–particle aggregates formed in the pulp zone eventually rise into the stabilized froth zone. This zone, which is located between the wash-water addition point and the pulp level, is largely responsible for the improved selectivity of columns over conventional flotation cells. In this zone, bubble coalescence is retarded by the downward flow of wash water that prevents the thinning of the froth lamella. Consequently, the gas holdup (ε_s) in the stabilized froth is relatively stable with height and is often in the range of 60%–70% by volume, which should be expected for packed spherical bubbles. The downward flow of water through the stabilized froth is commonly referred to as the bias flow. The bias flow stabilizes the froth and allows a deep, loosely packed bed of bubbles to be formed. The nonselective entrainment of gangue particles into this bed is prevented by the bias flow.

In addition, the circulation of floatable particles between the pulp and stabilized froth zones can be beneficial to selectivity. These features allow the performance of the column to closely approach that of the ideal separation by flotation predicted from release analysis (Davis et al. 1995). Although the stabilized froth allows high product grades to be achieved, it limits the particle recovery rate by restricting the amount of gas that can be passed through the column while still maintaining the proper bias flow. The stabilized froth places a constraint on throughput since only those particles that are attached to bubbles can enter the froth. This constraint, which is commonly referred to as the carrying capacity, places a limit on the mass rate of particles reporting to the froth product. Studies suggest that the carrying capacity limitation is particularly severe for the flotation of high-yield systems having fine particle size distributions. The carrying capacity limitation is almost always a major consideration in the design and operation of flotation columns.

The uppermost region in the column is the froth zone. This zone is located between the wash-water addition point and froth overflow lip and is similar to the drained froth phase present in conventional cells. As with conventional machines, the gas holdup (ε_f) in this zone is normally a strong function of height, with the surface reaching a gas fraction given by the gas rate, divided by the combined gas and slurry rate reporting to the froth overflow. This zone may be absent in some columns if the wash water is added above the froth bed using a drip pan or wash box. One purpose of this zone is to control the split of the wash water between the froth product and the stabilized froth zone. An adequate flow of water downward through the stabilized froth zone (i.e., bias flow) must be maintained to prevent the nonselective entrainment

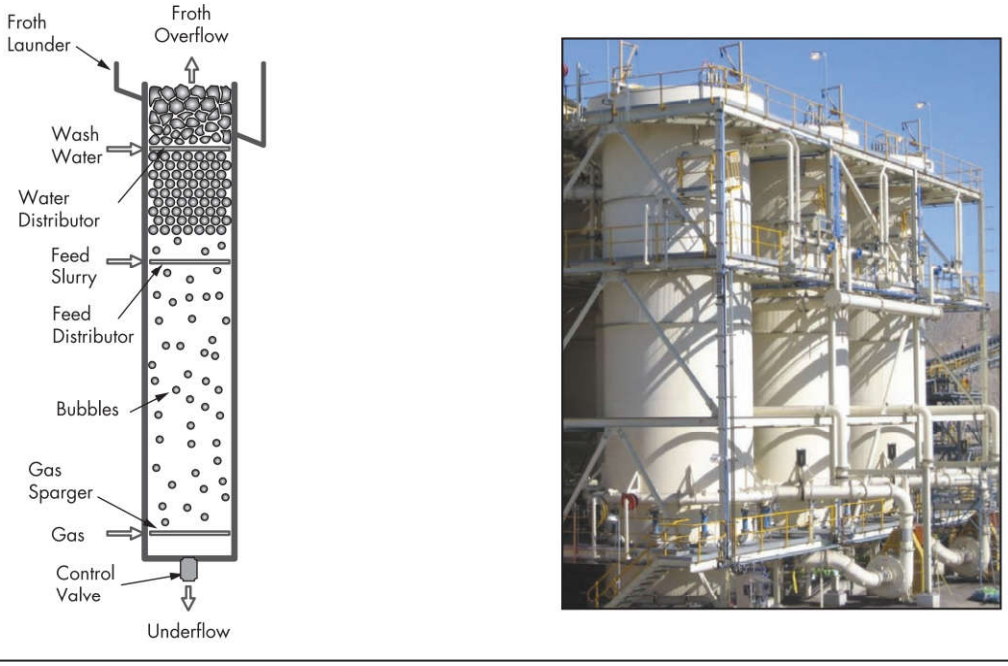


Figure 2 Basic features of a column flotation machine

of fine particles. A simple analysis can be used to show that the bias flow rate can be controlled to some extent by adjusting the aeration rate, wash-water addition point, and wash-water flow rate. Unfortunately, changes in these parameters may also affect the solids loading in the froth. If the froth becomes too heavily loaded with particles, it can become unstable and collapse. This phenomenon, which is referred to as froth overloading, places a limit on the solids content of the froth (Lynch et al. 1981). Potential problems associated with froth overloading must be recognized in the design and operation of columns.

SUPERFICIAL RATES

Because of the unique design of column cells, the flow rates passing through the machine are often reported as specific rates or superficial velocities. These values are calculated by dividing the volumetric flow rate or dry solids tonnage rate by the cross-sectional area of the column. In these calculations, the area may be reduced to correct for phenomena such as the holdup of gas within the column. The use of superficial velocities makes it easier to compare flow values for columns of different diameters and is generally more appropriate for describing the complex mechanisms that are the focus of column design calculations. Figures 4 and 5 show the superficial rates corresponding to different volumetric flow rates and dry solid tonnages, respectively, for columns of different diameters.

DESIGN VARIATIONS

Standard Designs

Column flotation machines are designed, manufactured, and serviced by a variety of equipment vendors. While homemade columns were popular during the early years of column technology development, this trend diminished over the past two decades in response to well-documented failures associated with poorly designed units. An industry rule-of-thumb such as "column height = 0.9 × crane height" is obviously inadequate

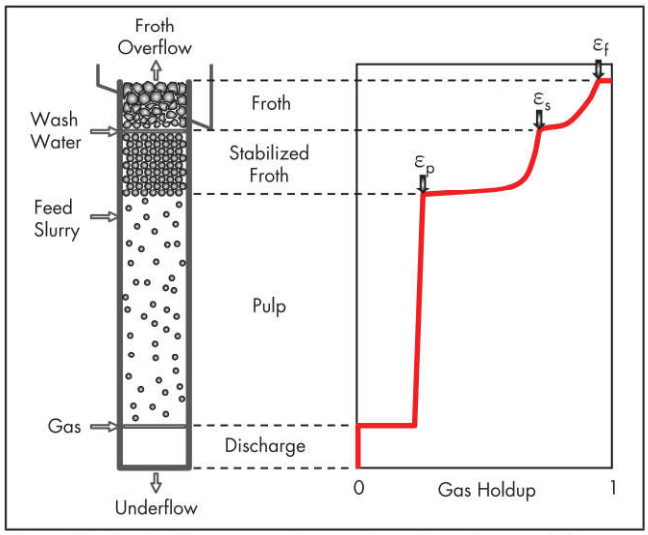


Figure 3 Gas holdup zones in a column flotation machine

for designing a modern column flotation system. Today's leading manufacturers of standard column systems include Eriez, Metso, and Minnovex/SGS. The machines offered by these companies use a traditional tank design with a large height-to-diameter ratio. Tank designs are most commonly circular because of savings on construction costs, although square and rectangular columns are also in commercial use. Tanks are now offered with diameters up to 5 m and heights in excess of 15–20 m. Notable differences between column manufacturers primarily involve vendor-specific designs of systems for gas sparging, slurry/water/gas distribution, and process control. Some vendors also offer internal baffles that are designed to reduce unwanted internal mixing and short-circuiting of feed. Applications of standard columns include the flotation

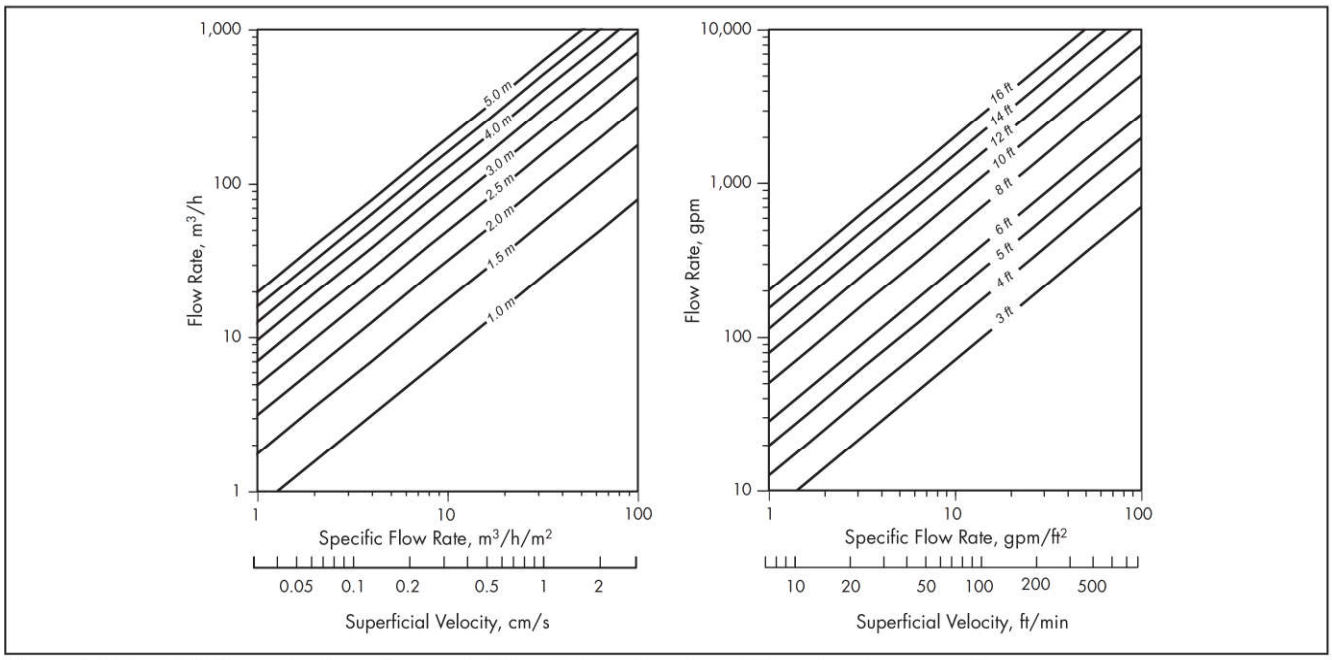


Figure 4 Volumetric flow rates for columns of different diameters, in metric and customary units

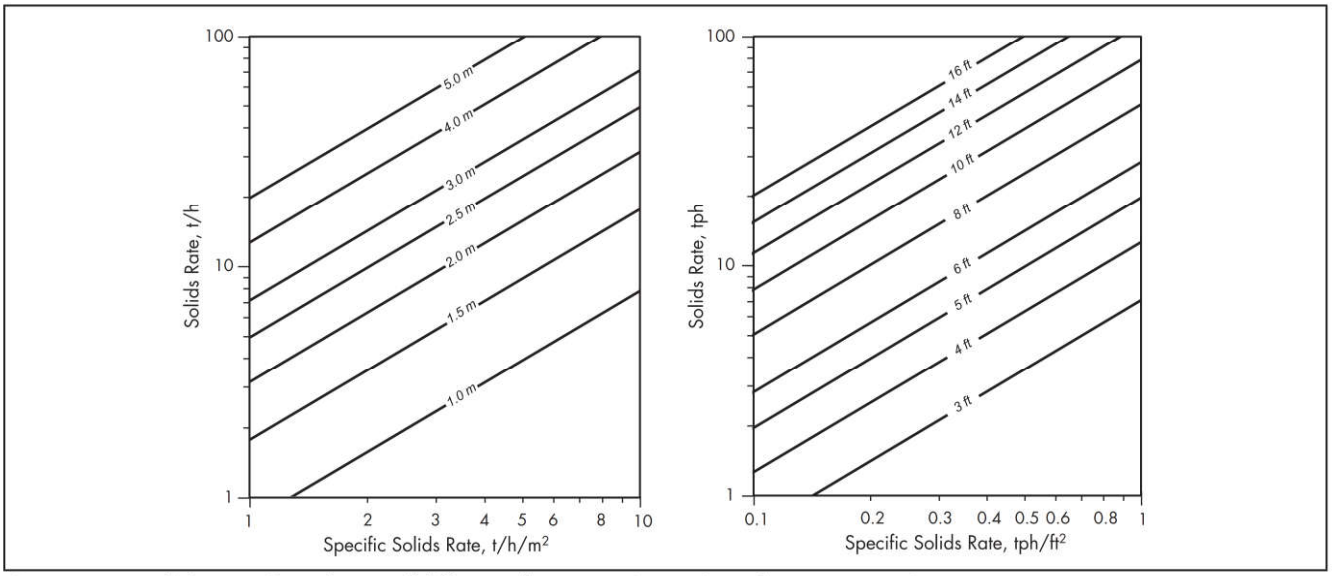


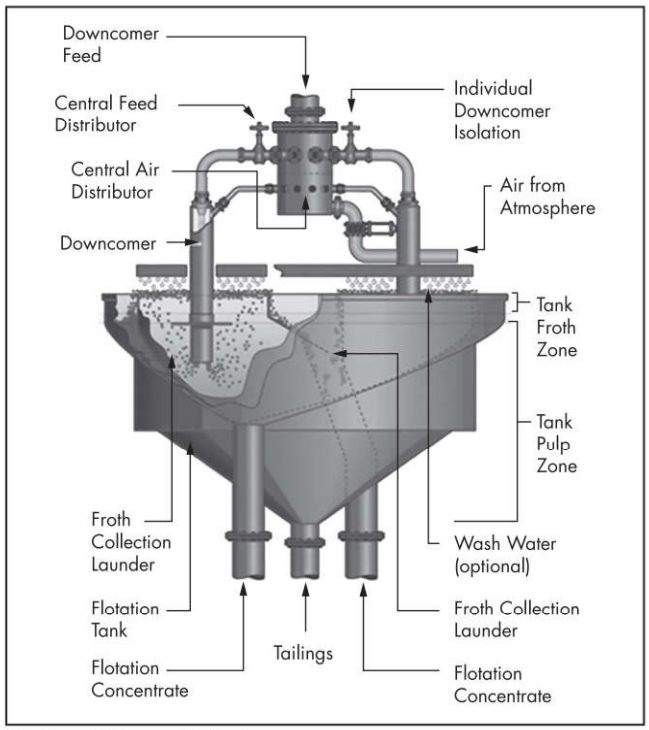
Figure 5 Dry solids rates for columns of different diameters, in metric and customary units

of particulates such as metallic and nonmetallic minerals, fine coal, solvent extraction organics, and bitumen.

Low-Profile Designs

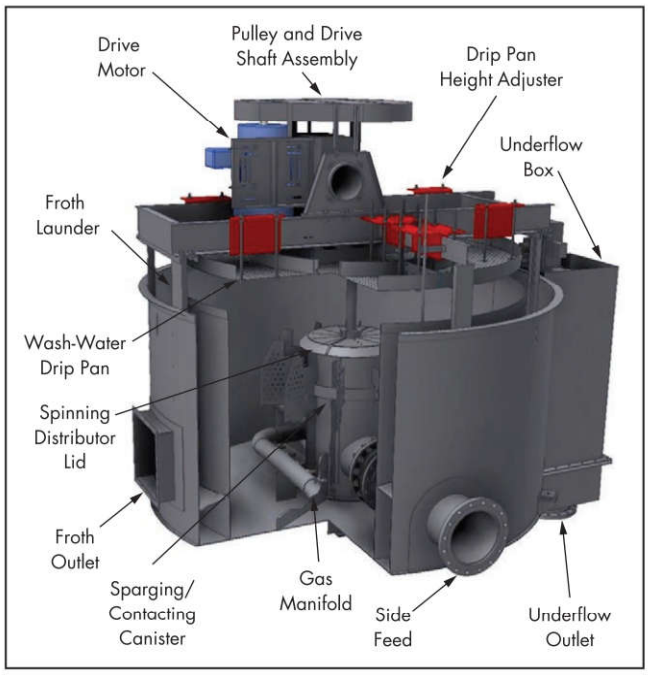
Large-scale flotation columns can pose design challenges because of massive tank dimensions and associated large foundation loads. To combat this issue, several low-profile designs have been developed that incorporate column froth washing technology. Two popular examples of these machines include the Jameson Cell and StackCell technologies. The Jameson Cell (Figure 6) passes pressurized feed slurry though a series of downcomer tubes mounted vertically above the tank. The negative hydrostatic head within the downcomer draws gas into the feed slurry. Gas is dispersed into small bubbles by slurry injected through a wear-resistant nozzle in the top of the downcomer. The flow conditions create a slurry–gas mixture with a high gas holdup inside the downcomer that promotes efficient bubble–particle contacting. The mixture passes into a low-profile flotation tank where a deep froth bed is formed that is suitable for water washing.

Another low-profile column design is the StackCell technology (Figure 7). During operation, feed slurry enters the separator through either a bottom-fed or side-fed feed nozzle at which point low-pressure air is added. The slurry travels into an internal pre-aeration sparging device that provides significant shear and contacting prior to arrival into the separation



Courtesy of Glencore Technology

Figure 6 Low-profile Jameson Cell flotation



Courtesy of Eriez Flotation Division

Figure 7 Low-profile StackCell flotation

chamber. Intense bubble–particle contacting is conducted in the aeration chamber prior to injection into the primary tank where phase separation of the pulp and froth occurs. Studies suggest that the high-intensity agitation is necessary for ultrafine particle flotation. A liquid slurry level is maintained inside the tank to provide a deep froth that can be washed, thereby providing a high-grade float product. The low-profile design allows the units to be easily stacked in series or placed ahead of existing conventional or column flotation cells.

FROTH WASHING

Bias Flow

The reduction in hydraulic entrainment by froth washing makes column flotation an attractive alternative to multistage cleaning. For the wash water to work effectively, a relatively deep froth (>1 m) is typically required and the flow of wash water should exceed the volumetric flow to the froth product. The difference between these two flow rates is normally referred to as the bias flow. The bias flow represents the amount of water that passes down through the stabilized froth to prevent fine hydrophilic slimes from entering the froth. As shown in Figure 8, a positive bias represents the desirable condition in which little or no water in the feed reports to the froth product, while a negative bias represents the undesirable condition in which a portion of the water in the feed reports to the froth product. Ideally, less than about 1% of the feed water will report to the froth product if the wash-water flow rate is properly controlled. In theory, a wash-water flow rate just above the flow rate of water demanded by the froth should provide a positive bias and prevent entrainment. In practice, however, mixing within the froth often requires that the wash-water flow rate be greater than the flow rate of water reporting with the froth phase. On the other hand, an excessive wash-water rate

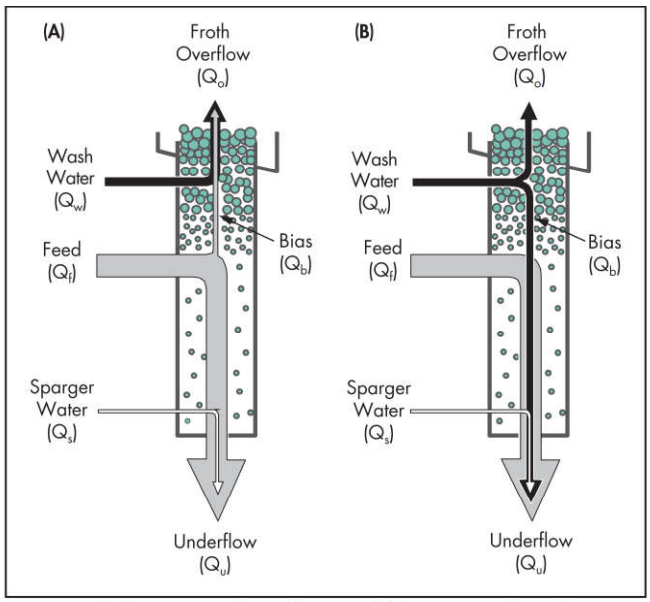


Figure 8 (A) Negative bias flow and (B) positive bias flow

must also be avoided since high water rates create an undesirable reduction in recovery in response to a reduction in slurry retention time. A very high wash-water flow rate can destabilize the froth by stripping/diluting the frother concentration. High wash-water flow rates can also have a detrimental impact on product grade by increasing axial froth mixing and reducing the wash-water effectiveness (Yianatos et al. 1988).

Design Criteria

Data collected from columns currently operating in the minerals industry suggest that a wash-water flow rate of

6.0–8.5 m^3/h (2.5–3.5 gpm/ft^2) of column cross-sectional area is normally adequate in a well-designed column to maintain a positive bias flow. However, the volumetric demand for wash water can be influenced by many interrelated factors that are associated with froth stability. A convenient method for evaluating the demand for wash water is to calculate the number of froth dilution washes (N_d), which is given by

$$N_d = W_w/W_o \quad \textbf{(EQ 1)}$$

where W_w and W_o are the volumetric flow rates of wash water and froth overflow water, respectively. The effect of N_d on the quality of a froth product is shown in Figure 9. A value of $N_d = 1$ indicates that just enough wash water is being added to replace the water reporting to the froth product. However, for reasons described previously, the data show that a value in the range of $N_d = 1.5$ or higher is often required in practice to avoid deterioration of the froth product by nonselectively entrained solids. The data also indicate that the quality of the froth product is much less consistent when operating under negative bias conditions ($N_d < 1$) because fluctuations in froth stability change the amount of water and entrained solids entering the froth product.

The relationship between the volumetric flow of wash water (W_w), mass flow rate of dry solids reporting to the froth overflow (M_o), and the percent solids (S) of the froth product is given by

$$W_w = N_d\,(M_o)\,(100 - S)/S \quad \textbf{(EQ 2)}$$

This relationship, which is plotted in Figure 10, effectively illustrates the rapid increase in wash-water demand as the froth solids content is reduced or dry solids rate of floatable particles is increased. A less stable froth will increase the froth solids content and will require less wash water. If the solids content is too high, however, an overloaded froth condition will occur that may limit froth mobility and capacity. Likewise, a too stable froth can lead to a negative bias condition as a result of the lower solids content. A highly stable froth can also require a water flow rate in excess of that which can penetrate the rising three-phase mixture of gas, solids, and water, thereby resulting in poor-quality froth even at high wash-water addition rates.

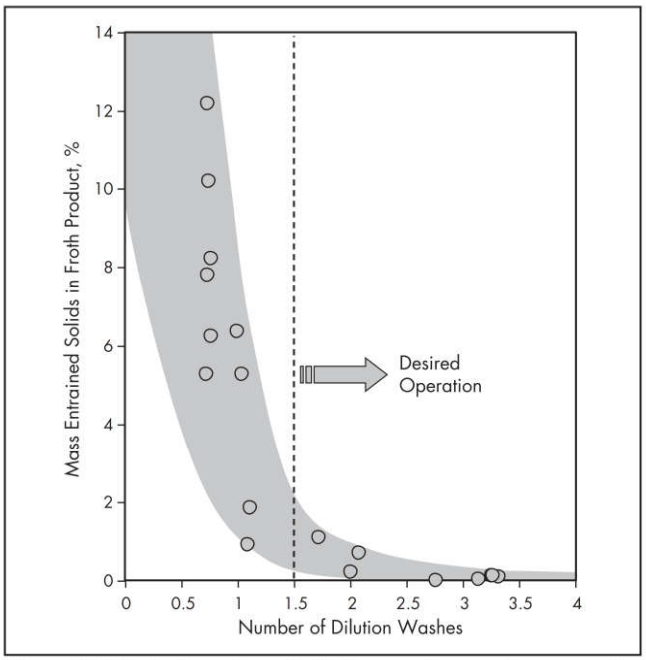


Figure 9 Effect of number of dilution washes on the amount of solids entrained into a froth

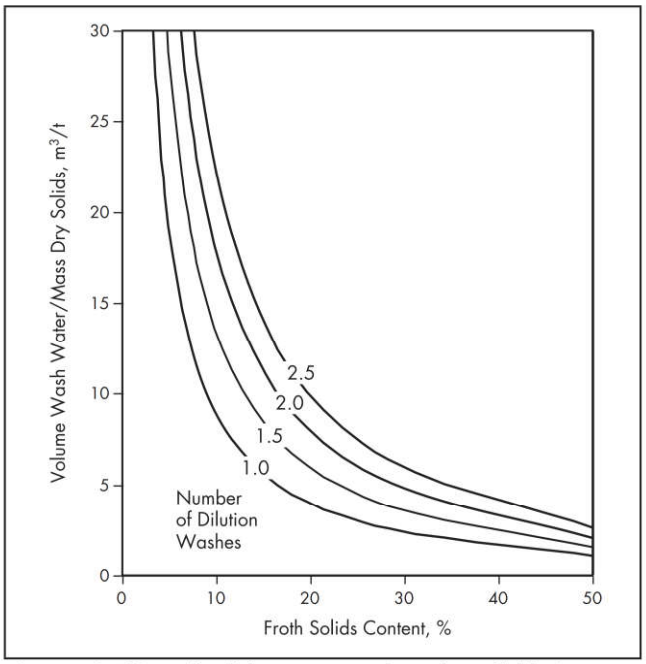


Figure 10 Effect of solids content and number of dilution washes on froth water demand

Water Distributors

The design of the wash-water distribution system can have a large impact on the effectiveness of the wash-water addition. Unfortunately, no standard design currently exists. A common layout for circular columns involves concentric rings of piping into which numerous small holes have been drilled for water distribution (Figure 11). For these systems, the distribution piping can be submerged below the cell lip so that a drained froth can form above the distributor and a stabilized froth below the distributor. This positioning allows the depth of the drained froth and the extent of froth drainage to be varied by raising or lowering the distributor; that is, the vertical position of the water distributor controls the split of water between the froth and bias streams. Multiple distributors placed at different depths may also be incorporated into the design to control froth stability and to balance the water flows (Yoon et al. 1992). For circular-shaped distribution rings, the inner rings are frequently placed above the outer rings to maintain the fluidity and mobility of the center of the froth. The multilevel distribution system can also be adjusted online to accommodate day-to-day variations in froth stability. More commonly, however, operators have found that it is necessary to locate the water distributor just above the top of the froth so that plugging problems can be visually monitored and corrected. This configuration, which eliminates the drained froth zone, also makes it possible to use drip pans instead of distribution pipes. A drip pan, which consists of an open-top perforated tray (Figure 12), is easier to routinely clean from above

Figure 11 Pipe-ring wash-water distributor

Figure 12 Drip-pan wash-water distributor

using a simple water hose and brush. The major shortcoming of this design is that it does not allow the wash-water distributor to be raised or lowered to control the bias flow. Regardless of which design is used, the water should be gently introduced into the froth at low velocity to avoid jets that would disrupt the even upward flow of the froth (Finch and Dobby 1990).

The number, size, and spacing pattern of holes in the water distributor can vary considerably depending on wash-water demand, column area, and vendor-specific preferences. Typical hole sizes range from about 5–10 mm in diameter with center-to-center spacings ranging from 2–15 cm. The spacing pattern used should be chosen to provide an even water distribution over the entire froth area (Neethling et al. 2006). Failure to do so can create localized regions of positive and negative bias within the same froth that are detrimental to separation performance. On the other hand, work performed by Neethling and Cilliers (2001) suggests that washing performance can be enhanced by adding more water to the center of the column. Therefore, field adjustments by the column manufacturer are often required to fine-tune and optimize the performance of the wash-water distributor system for each industrial application.

SPARGING SYSTEMS

Superficial Bubble Surface Area Rate

The gas sparging system is the most important component in the design of a flotation column. An effective gas sparging system must be capable of producing small, uniformly sized bubbles at a desired aeration rate (Xu and Finch 1989; Groppo and Parekh 1992; Huls et al. 1991). Small bubbles can improve flotation kinetics by enhancing the capture efficiency of fine particles (Dobby and Finch 1986a; Yoon and Luttrell 1986). In addition, smaller bubbles may increase the ability of the froth to retain and carry solids by increasing the total interfacial flux of bubble surface area through the froth phase (Finch and Dobby 1990). An ideal sparging system should also be nonplugging and wear-resistant and allow for easy, online servicing.

A key measure of sparger performance is the superficial bubble surface area rate (S_b). This parameter, which has the units of 1/time, is defined as the total surface area of bubbles per unit of time passing through a given column cross-sectional area (Finch and Dobby 1990). This parameter can be calculated using the following equation:

$$S_b = (Q_g/A_c) / (\pi D_b 3/6) \times \pi D_b^2 = 6Q_g/(A_c D_b) = 6\ J_g/D_b \quad \text{(EQ 3)}$$

where Q_g is the gas flow rate, A_c is the cross-sectional area of the column, D_b is the bubble diameter, and J_g is the superficial gas rate. This expression shows that amount of bubble surface area available for collecting hydrophobic particles is directly proportional to the ratio of superficial gas rate to bubble size. The impact of S_b on flotation recovery is illustrated by the test results given in Figure 13 (Kohmuench et al. 2004). The data for this case show that recovery increases sharply as S_b increases to about 80 s^{-1} and reaches a plateau at 120 s^{-1}. The data also indicate that the relationship between recovery and S_b is generally independent of sparger type. This finding agrees with other studies showing a linear correlation between S_b and flotation rate constant (Gorain et al. 1997, 1998). The problem is that many commercial sparging systems simply cannot produce high S_b values. Field tests also suggest that operation at S_b values above 120 s^{-1} can also produce a poor separation because of excessively wet froths that cannot be effectively washed. A proper combination of gas rate and bubble size will generally provide a gas holdup in the flotation pulp in the range of 12% to 18% by volume.

Sparger Types

Three primary types of column sparging systems have been used commercially: (1) porous bubblers fabricated from filter cloth, punctured rubber tubes, or sintered metals/plastics; (2) high-velocity gas or gas–water injectors; and (3) dynamic mixers that circulate gas–slurry mixtures through static mixing elements. Details of these systems have been described in the technical literature (Clingan and McGregor 1987; Xu and Finch 1989; Huls et al. 1991; Yoon et al. 1992; Groppo and Parekh 1992; Finch 1995; Rubinstein 1995).

Static Spargers

The first column flotation machines evaluated in industry used internal (static) porous bubblers fabricated from porous materials such as filter cloth, sintered metal/ceramic beads, or perforated rubber tubes. Porous metal and permeable ceramic tubes

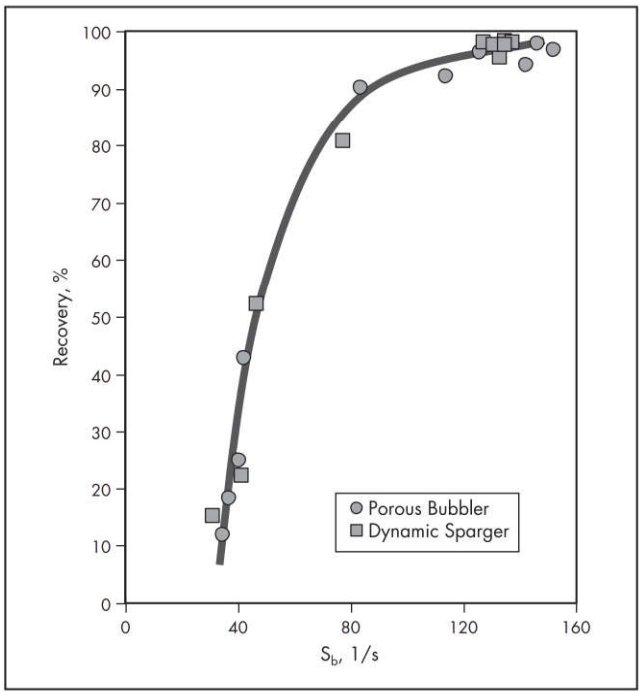
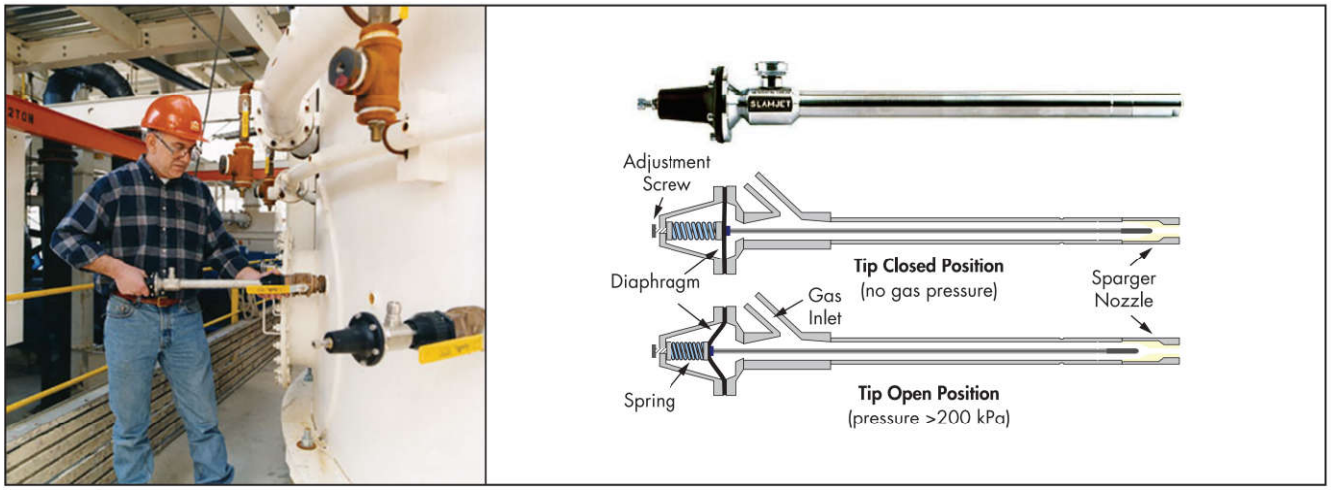


Figure 13 Effect of superficial bubble surface area rate on recovery for different spargers

can be used to produce small air bubbles, but they are prone to plugging and must be frequently removed and cleaned. This is a time-consuming and costly exercise that requires the column to be removed from service and drained. Plugging can also be a problem for static spargers that are designed to pass mixtures of compressed gas and pressurized water through small orifices since plant water and gas supplies are often contaminated with solids. Another issue with porous bubblers is scale-up capability. Xu and Finch (1989) and Clingan and McGregor (1987) showed that porous spargers should be scaled up by maintaining a constant ratio (R_s) between the column cross-sectional area and the sparger surface area. Unfortunately, this ratio can be difficult to maintain in industrial practice. For example, a 5-cm-diameter laboratory column equipped with a single 2-cm-diameter by 5-cm-long porous tube sparger would give $R_s = 0.625$. As such, a single 2-m-diameter full-scale column (3.14 m^2 of area) would require 5 m^2 of sparger surface area to perform similarly to the laboratory column. Installation of such a large area of porous spargers would be difficult in practice because of the space restrictions inside a column. This configuration would also present challenges in terms of plugging caused by slurry scaling and dirty gases.

High-Velocity Injectors

To overcome serious issues associated with static spargers, several high-velocity gas injection systems were developed for column sparging. One of the earliest systems, developed by the U.S. Bureau of Mines, passed gas through small (1–3 mm diameter) holes that were equally spaced along injector tubes that passed through the column wall and across the entire column diameter. A very small amount of pressurized clean water was added to the gas flow to improve bubble dispersion at the injector tips. This system was soon surpassed by a new generation of high-velocity gas injectors developed by groups such as Cominco and Minnovex/SGS. These new designs consisted of removable air lances with a single gas outlet at the end of the sparger tip. When operating properly, gas discharges from the tip at a sonic or near-sonic velocity. The intense turbulence created by the jet disperses the gas into fine bubbles. The new designs performed well but were still plagued with maintenance problems created by plugging during shutdown. Canadian Process Technologies solved this problem with the introduction of the SlamJet sparger (Figure 14). This technology incorporates a spring-loaded diaphragm that automatically drives a rod outward to close the end of the sparger tip when the gas pressure drops below a minimum set point (usually 200 kPa). The self-closing mechanism eliminates any backflow of slurry into the aeration system, thereby avoiding the plugging problems. Worldwide, more than 1,000 of these sparger systems are in use today for minerals processing applications. The recommended range of gas flow rates that can be generated per sparger tip is shown in Figure 15.



Courtesy of Eriez Flotation Division

Figure 14 Externally serviceable, self-closing, high-velocity gas sparger

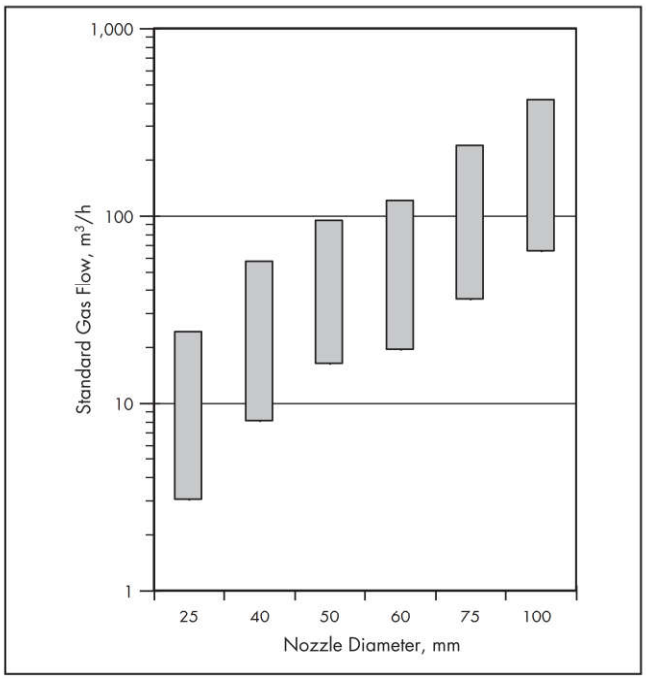


Figure 15 Recommended gas flow ranges for high-velocity gas injectors

Dynamic Spargers

Dynamic spargers are another popular design of gas aeration systems for column flotation machines. This type of sparger design uses a centrifugal pump to circulate slurry from the bottom section of the column through high-turbulence in-line contactors into which compressed gas is also injected. Depending on the manufacturer, the slurry–gas contactors may consist of static mixers, orifice plates, or venturi tubes. The intense mixing within the contactor produces small (<0.8 mm) bubbles and promotes bubble–particle contacting. Industrial studies indicate that these devices typically generate higher S_b values than other types of column spargers, thereby offering a higher capacity and/or recovery of fine particles (Pyecha et al. 2005).

Two popular designs of dynamic spargers are the Microcel and CavTube sparging systems (Figure 16). The Microcel technology uses in-line static mixers for gas dispersion and to scavenge any remaining floatable particles from the underflow prior to discharge. This configuration is designed to operate at relatively low back pressures, typically in the range of 100–250 kPa. The flow rate of circulated slurry is typically adjusted to provide a gas–slurry mixture with less than 40%–50% gas by volume. Figure 17 shows the recommended range of gas flow rates that can be effectively generated per sparger. The CavTube system is similar in function but is designed to operate at higher pressures sufficient to supersaturate the

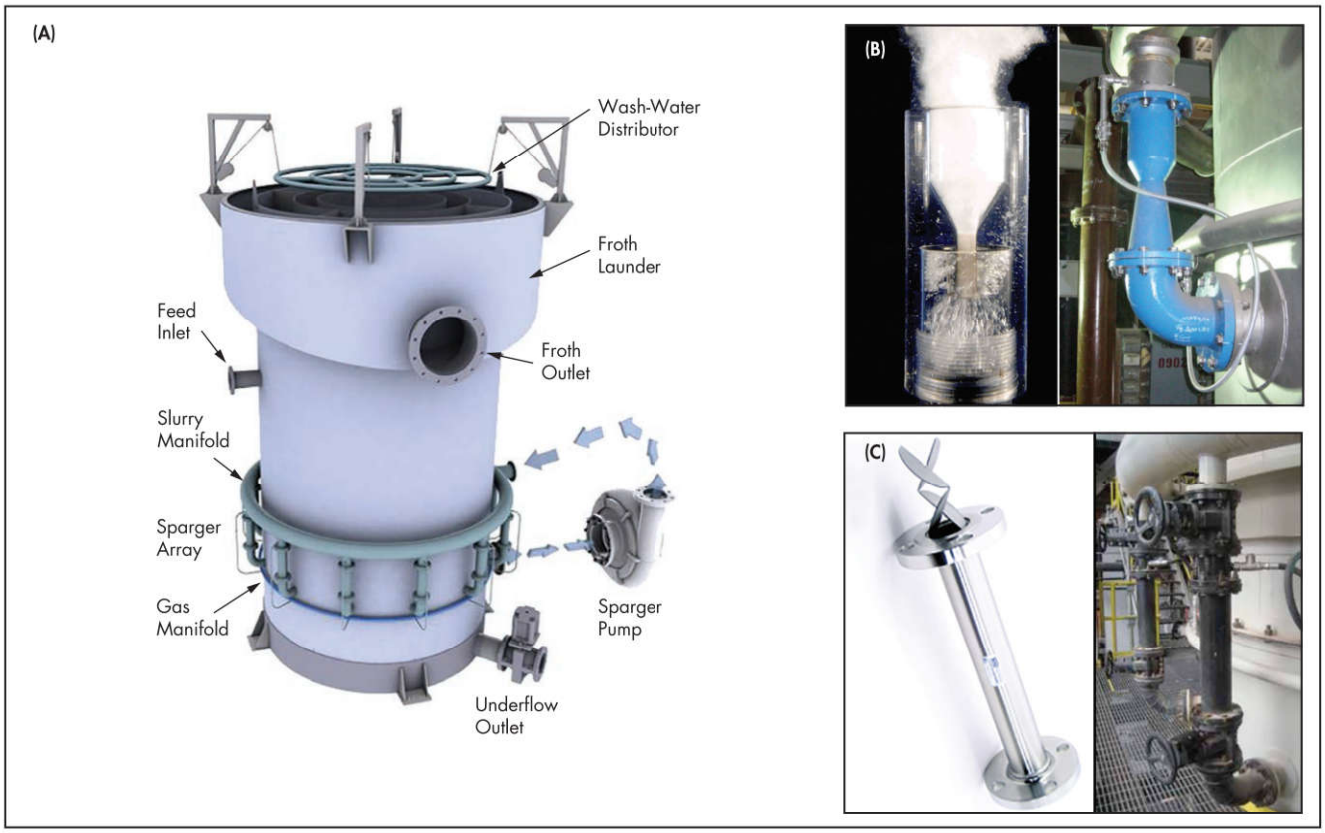


Figure 16 (A) Dynamic sparger system, including (B) CavTube and (C) Microcel systems

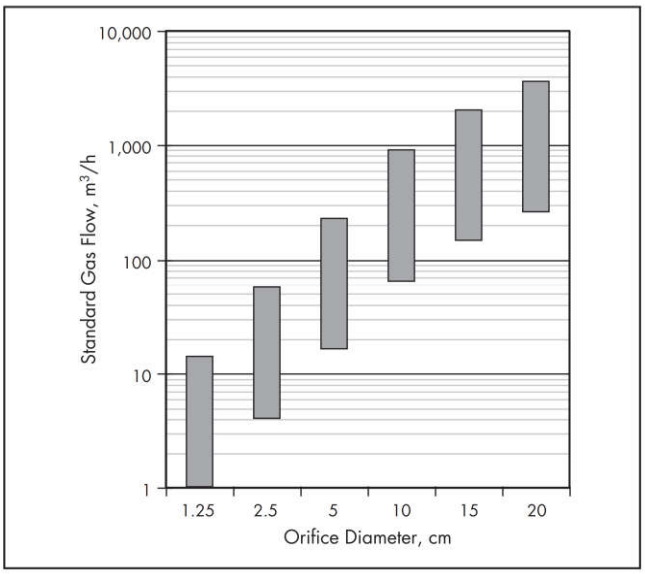


Figure 17 Recommended gas flow ranges for dynamic sparger

slurry with gas and to nucleate very tiny bubbles on the surfaces of hydrophobic particles. Although the nucleated gas bubbles represent less than 2% of the total introduced gas volume (Wasmund 2014), these tiny patches of gas enhance flotation kinetics by increasing the likelihood of successful bubble–particle attachments (Fan and Tao 2010). More than 200 of these systems are currently in operation in the minerals processing industry. CavTube spargers have also been used to pre-aerate the feed slurry to flotation columns and other types of flotation cells. Experimental studies have shown that CavTube pre-aeration can effectively improve flotation recovery, increase capacity, and reduce reagent consumption (Zhou et al. 1997; Honaker et al. 2011; Ahmed 2013).

Compressor Sizing

The majority of column cell installations require the use of compressed gas. Consequently, compressor sizing is an important task in the design of column flotation machines. Column manufacturers frequently report gas flow rates as a standard volumetric flow per time. This value is only valid for dry air at a pressure of 1 atm and temperature of 20°C. The actual flow rate specified by compressor manufacturers is typically reported in terms of inlet conditions (free air). Although this amount of air enters the compressor, it is not necessarily the amount of air delivered to the column because of compressor seal leakage. As a result, the actual flow may be only 95% of the inlet flow. Furthermore, corrections to the gas flow rate must be made to account for differences in elevation (atmospheric pressure) and humidity. Air temperature generally has little impact on the capacity of an oil-flooded screw compressor but may affect the performance of an air-cooled compressor (Sullair Corporation 1992). These complications generally require that professionals be consulted to ensure that the compressor is properly sized for the specified air requirements. However, as a sanity check, operators should expect to require about 0.1 kW of compressor power for each cubic meter per hour of gas flow demand. Caution should also be used during the metering of gas flow rates. A properly designed system should be equipped with a flowmeter that is calibrated to read correctly at a specified operating pressure. The operating pressure should be held constant by placing a pressure regulator ahead of the flowmeter. By placing the airflow control valve after the flowmeter, the flowmeter will always operate at its design pressure. If the flowmeter is placed after the control valve, then the operating pressure and true gas flow rate are both unknown. Improper metering of the gas flow rate can be a particularly serious problem when laboratory and pilot-scale tests are conducted to collect scale-up information.

SCALE-UP AND DESIGN

Flotation Kinetics

Froth flotation is commonly modeled as a first-order kinetic process. As such, the recovery (R) of floatable particles can be mathematically related to the residence time (τ), rate constant (k), and Peclet number (Pe) using Levenspiel's (1972) equation:

$$R = \frac{1 - 4A^{Pe/2}}{[(1+A)^2]^{APe/2} - [(1-A)^2]^{-APe/2}} \quad \textbf{(EQ 4)}$$

$$A = \sqrt{1 + 4k\tau Pe} \quad \textbf{(EQ 5)}$$

This relationship shows that recovery is a function of two dimensionless groups, k and Pe (Dobby and Finch 1985). For the recovery of particles from a flotation pulp, k has been shown (Finch and Dobby 1990) to be theoretically given by

$$k = ¼\ P\ S_b \quad \textbf{(EQ 6)}$$

which shows that the rate of flotation is proportional to the probability of attachment (P) and the superficial bubble surface area rate (S_b). Since k is normally assumed to remain unchanged during scale-up, both P and S_b must be held constant in both the small- and large-scale units. The mean residence time (τ_p) of small particles in the flotation column can be estimated from the active volume (V_p) of the flotation pulp and the volumetric flow rate of slurry to the underflow (Q_u) using

$$\tau_p = \frac{V_p}{Q_u} = H_p\left(\frac{1-\varepsilon_p}{J_u}\right) \quad \textbf{(EQ 7)}$$

in which H_p is the vertical height between the sparger inlet and base of the stabilized froth, ε_p is the gas holdup in the flotation pulp, and J_u is the underflow superficial velocity. This expression assumes that the residence time of slurry and particles is equivalent, and, hence, a correction is occasionally required if particle settling velocities become high. Finally, Pe is a dimensionless group that represents the degree of axial mixing within the flotation pulp. For column cells, Pe varies with both geometry and flow conditions (Dobby and Finch 1985) and may be estimated from an empirical expression derived by Mankosa et al. (1992):

$$Pe = 0.70\left(\frac{H_p}{D}\right)^{0.63}\left(\frac{J_u}{J_g}\right)^{0.30} \quad \textbf{(EQ 8)}$$

This expression suggests that columns become increasingly mixed in response to a reduction in either the height-to-diameter ratio or slurry-to-gas flow ratio.

Figure 18 shows an example of the relationship between recovery and residence time for a flotation column. In each test, the volumetric feed flow rate to the column was steadily increased to reduce the mean residence time. As shown, a good correlation was obtained between the values predicted

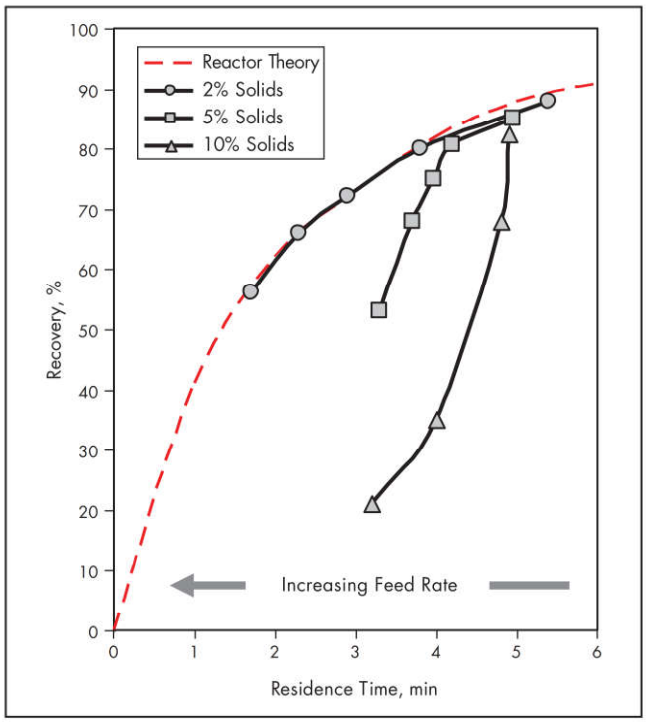


Figure 18 Froth overload condition created by increasing feed solids content

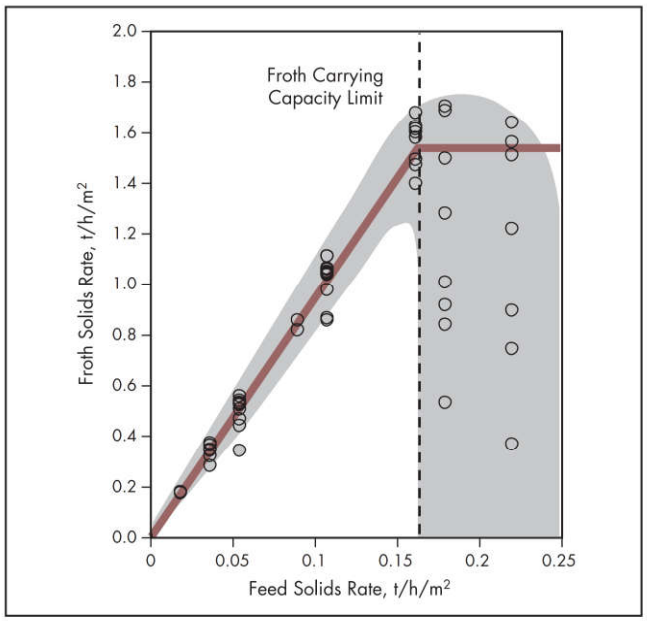


Figure 19 Graphical method for identifying the froth carrying capacity

from Equation 4 and the experimental data obtained using feed slurry with a very low solids content of 2%. The low solids content ensured that the column was limited by residence time and not by froth solids overloading. However, as the solids content of the feed stream was increased to 5% solids, the recovery dropped sharply and deviated substantially from the Levenspiel (1972) relationship. This deviation was even more pronounced when the solids content was doubled to 10% solids. The drop in recovery with increasing solids content can be attributed to limitations associated with froth carrying capacity. When this occurs, there is insufficient bubble surface area to carry all of the floatable particles collected in the pulp phase into the froth phase.

Froth Carrying Capacity

Although the carrying capacity restriction can often be ignored for conventional flotation circuits, it is important in the design of column cells, largely because the specific surface area of the cell (ratio of the cross-sectional area to the volume) is much higher for conventional cells. Column cells have specific surface areas in the range of 0.02–0.08 m^2/m^3, while those for conventional cells are in the 0.4–2.5 m^2/m^3 range. Theoretical studies indicate that carrying capacity, which is normally reported in terms of the rate of concentrate floated per unit cross-sectional area, is linearly related to the size and density of particles in the froth (Espinosa-Gomez et al. 1988; Sastri 1996). Froth carrying capacity may be estimated from laboratory and pilot-scale flotation tests by conducting experiments as a function of feed solids content (Finch and Dobby 1990). As shown in Figure 19, this procedure can be used to produce a plot of froth solids rate versus feed solids rate (Patwardhan and Honaker 2000). Prior to reaching the carrying capacity, an increase in feed solids rate produces a corresponding increase in the froth solids rate. Eventually, the froth solids rate reaches a plateau defined by the carrying capacity. The production rate of floatable solids becomes very sensitive to changes in operating parameters such as gas rate and frother dosage upon reaching this inflection point. For the case shown, a carrying capacity of 1.55 $t/h/m^2$ of column cross-sectional area should be expected. Table 1 shows the range of carrying capacities that have been observed for different mineral systems treated by column flotation. These values should be used with caution, however, because factors such as particle size and density can greatly alter these values (Espinosa-Gomez et al. 1988).

Scale-Up Projections

The methodology involved in the scale-up of laboratory and pilot-scale column data has been described in detail by Finch and Dobby (1990). Unfortunately, scale-up can be a challenging task for column designers. Many of the parameters used in these calculations may or may not remain constant in the full-scale column. For example, field observations suggest

Table 1 Examples of froth carrying capacities observed for different mineral systems

Mineral System	Carrying Capacity, $t/h/m^2$
Coal	1.2
Copper	2.5
Lead	3.2
Molybdenum	3.7
Nickel	1.7
Oxides	4.4
Phosphates	5.0
Silica	5.0
Zinc	1.7

that the bubble size distributions in full-scale columns are often coarser than those used in the scale-up experiments. The buildup of recycle streams in the feed and contaminants in the process water can also impact the pulp chemistry and froth stability. Reputable manufacturers of column machines have long recognized these unforeseen issues and have developed proprietary corrections based on their historical experiences. Therefore, while a first-pass estimation procedure for column sizing is provided herein, column scale-up is generally best handled by experts working in the field.

The scale-up procedure for column cells primarily involves the determination of column diameter and height. For a given installation, the total cross-sectional area of the cells is normally dictated by the carrying capacity. This relationship may be expressed as

$$D = \sqrt{\frac{4M_f Y}{\pi C}} \qquad \textbf{(EQ 9)}$$

where D is the column diameter, M_f is the mass feed rate of dry solids, Y is the expected mass yield of solids to the froth product, and C is the froth carrying capacity. For a first-pass estimate, it is generally more convenient to calculate the mass rate of froth overflow solids for a full-scale column ($M_{o,full}$) to that for a small laboratory or pilot-scale test column ($M_{o,lab}$) using

$$M_{o,full} = M_{o,lab}\left(\frac{D_{full}}{D_{lab}}\right)^n \qquad \textbf{(EQ 10)}$$

where D_{full}/D_{lab} is the ratio of the diameters of the full-scale and laboratory test columns, respectively, and n is a scaling power factor. Historical field work indicates that an area-based scale-up (n~2) is valid for many industrial mineral flotation systems, provided that the column is well designed with adequate internal launder systems to minimize froth travel distances. According to this relationship, a 3-m-diameter column would have a theoretical froth solids capacity approaching 3,600 times that of a 5-cm-diameter test column. However, an area-based scale-up calculation appears to contradict data reported by Amelunxen (1990) suggesting that froth mobility problems make the column circumference the appropriate scale-up parameter (i.e., n~1). In practice, a value of $1 < n < 2$ may be used by column manufacturers based on specifications related to the launder design and information available from experimental testing and/or proprietary databases.

After determining the column diameter, the height of the column pulp zone must be estimated. This height, which is normally taken as the distance between the base of the froth and the air injection point, must be large enough to ensure that adequate residence time exists to achieve the target recovery dictated by the carrying capacity. Detailed procedures have been reported in the literature for determining this height (Dobby and Finch 1986b). However, as a first-pass estimate, the height of the collection zone for a full-scale column (H_{full}) relative to the height of a smaller test column (H_{lab}) can be determined from

$$H_{full} = H_{lab}\left(\frac{\tau_{full}}{\tau_{lab}}\right) = H_{lab} Z \qquad \textbf{(EQ 11)}$$

where $Z = \tau_{full}/\tau_{lab}$, which is the ratio of the mean particle residence times for the full-scale plant column and laboratory-scale test column, respectively. It cannot be automatically assumed that Z = 1 since the full-scale column is typically more mixed and, hence, requires a longer residence time to achieve the same recovery. Therefore, for first-pass estimations, the plot provided in Figure 20 can be used to estimate the Z multiplier on residence time required for the full-scale column, provided that the flotation rate constant remains unchanged during scale-up. For example, consider a 5-cm-diameter laboratory column with a pulp height of 3.5 m. If the test column is near plug-flow (Pe = 0.05), then a target recovery of 86% in the test column would correspond to $k\tau = 2.2$ for this set of operating conditions. Likewise, an 85% recovery in the full-scale column (Pe = 4) would require $k\tau = 5.2$. Therefore, the height of the full-scale column pulp zone should be 5.2/2.2 = 2.4 times the laboratory pulp height to maintain the same recovery. This calculation dictates a full-scale column pulp height of 9.1 m (i.e., 3.5 × 2.4 = 8.4 m). An additional 1 m of column height should be added to the top of the pulp zone to account for the froth zones. Another half diameter (1.5 m) should also be added to the bottom of the pulp zone for underflow removal and sparger systems. (The height of the bottom section may vary significantly, however, depending on the mechanical design of the column.) Based on these estimates, the total estimated column height for this example would be 10.9 m (8.4 + 1 + 1.5).

Unfortunately, the Pe values for the small-scale test column and full-scale production column are usually not known in advance. In such cases, a conservative estimate for pulp height can be obtained by assuming that the test column is plug-flow (Pe = 0) and the full-scale column is perfectly mixed (Pe = ∞). In this case, the Z ratio can be mathematically derived as

$$Z = \frac{R}{(R-1)\ln(1-R)} \qquad \textbf{(EQ 12)}$$

where R is the desired fractional recovery (D. Catarious, personal communication; Luttrell et al. 2000). For the current example, this expression gives a scale factor of $Z = H_{full}/H_{lab} = 3.1$, which corresponds to a full-scale column pulp height of 10.9 m. While this pulp height is greater than the 8.4 m height predicted by the more rigorous approach, this shortcut method provides a quick means of estimating maximum column height for first-pass designs. For convenience, Equation 12 is plotted along with Equation 4 in Figure 20. Field data obtained from operating columns suggest that the shortcut method is reasonable provided that the height of the smaller test column is very close to that required to reach the carrying capacity limit.

INDUSTRIAL PRACTICE

Iron Ore Flotation

Columns have become widely accepted in the iron ore industry over the past two decades. Reverse flotation using columns has proven to be an economical and effective method for reducing the concentrate silica content to well below 2%. Additionally, the use of wash water has provided a means of obtaining low concentrate silica levels while keeping iron losses to a minimum. Furthermore, total energy per unit volume as well as capital and installation costs favor columns (Murdock et al. 1991). The Brazilian iron ore industry has led the world in adopting column technology for reducing the silica content of iron pellets (Sandvik et al. 1991). In fact, Samarco Mineração S.A., the first Brazilian producer to use columns, installed columns to increase capacity as part of a plant expansion program in the early 1990s. Since that time, additional columns have been installed for the recovery of fine iron from a desliming

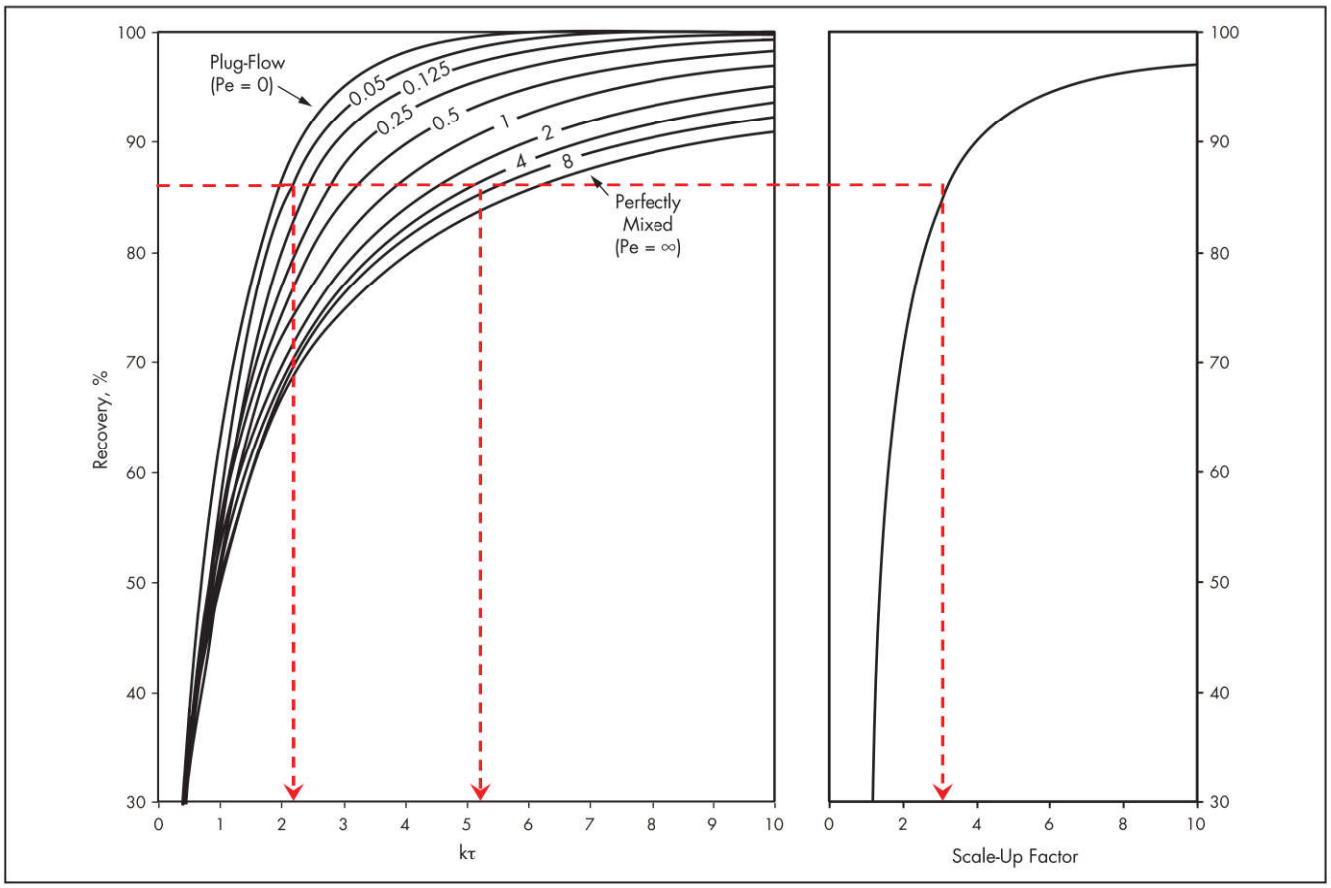


Figure 20 Estimation of the residence time scale-up factor for column flotation

circuit and for plant expansion programs. Many plants require the flexibility to produce more than one iron product, such as blast furnace (2.2% silicon dioxide [SiO_2]) and direct reduction (0.8% SiO_2) grades using the same equipment. When the higher silica blast furnace products are produced, the grinding requirements for liberation generally decrease and therefore the plants can operate at increased capacities. The higher feed rate affects the feed size distribution and must be taken into account in the circuit design. The following section includes examples of some of the flow-sheet configurations adopted by iron ore producers.

Cleaner–Scavenger Circuit

Columns are often used as cleaners in combination with mechanical cells. This is the case for a Brazilian flotation plant that originally consisted of four parallel mechanical flotation lines containing a rougher, cleaner, and two scavenging stages. The depletion of reserves at the existing mine and the subsequent development of a new ore zone necessitated the modification of the existing flow sheet because ore from the new mine is harder and slower floating (Sandvik et al. 1991). Several circuit alternatives were considered, including additional conventional cells, a parallel line of columns, and columns as re-cleaners. Pilot-plant test results revealed that the third option provided the best combination of plant throughput and iron recovery. The flow sheet for the modified circuit (Wyslouzil 2009a) is shown in Figure 21. In this approach, cyclone underflow from the desliming circuit feeds the flotation operations with a particle size range of 150 × 10 µm. One column was added as a re-cleaner to each processing line to preserve circuit flexibility. The re-cleaner column feed grade ranges between 1% and 6% silica; therefore, one column stage is sufficient to produce either blast-furnace or direct-reduction concentrate. A scavenger column was also added to the flow sheet to maximize fine iron recovery, particularly during periods when direct reducing-grade iron is being produced. Because these columns have been added to an existing silica flotation plant as re-cleaners, consideration was given to the effects of variations in the roughing and primary circuits. Generally, the fine, fast-floating silica is removed in the roughing stage and therefore the column feed tends to be enriched in coarse silica. Furthermore, any operational problems with the existing plant tend to flow through to the column, resulting in variable feed conditions.

Silica Reduction from Magnetite Concentrate

Columns can also be used in combination with magnetic separation circuits. The flow sheet in Figure 22 shows a 1,200 t/h flotation plant designed to lower the silica content of magnetite concentrate from 7%–9% to 2.2% SiO_2. New feed is ground to 80% passing 150 µm and is fed to primary magnetic separators. The magnetic concentrate is reground to approximately 75% passing 45 µm and is retreated in secondary magnetic separators that produce a concentrate containing 7%–9% SiO_2. The magnetic concentrate is conditioned with caustic starch solution and amine prior to being fed to a single-stage

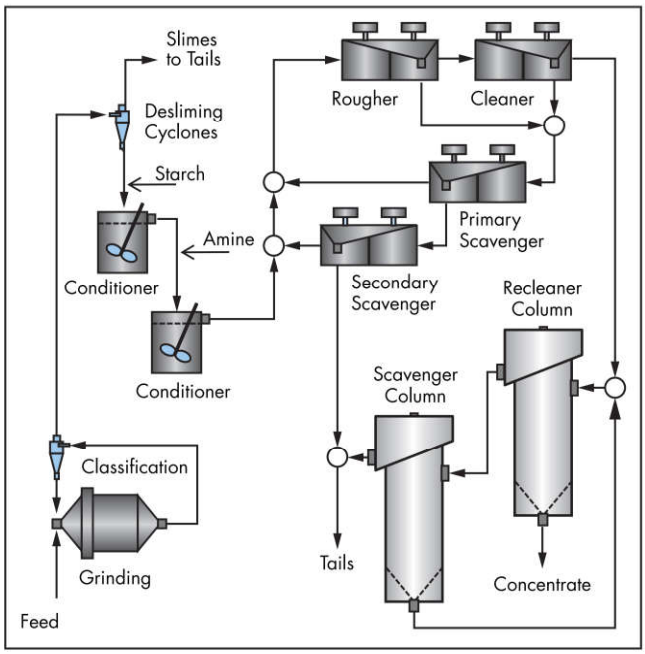


Figure 21 Columns added to an existing iron ore circuit

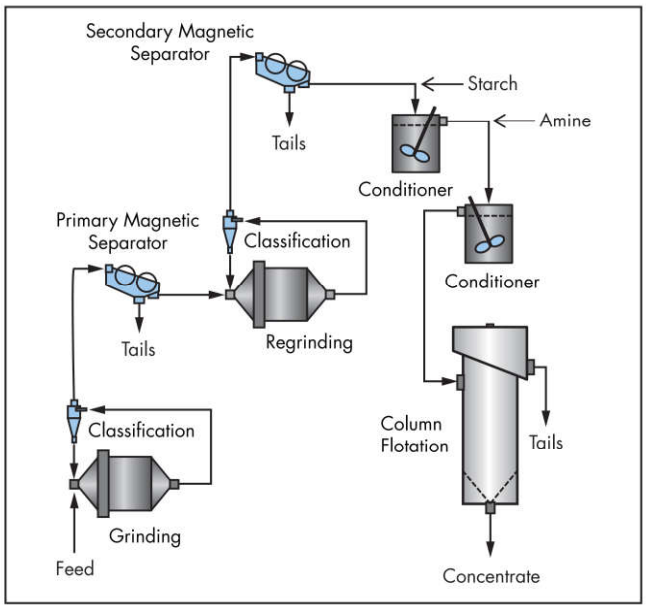


Figure 22 Column flotation circuit for magnetite treatment

flotation column where the silica content is reduced to below 2.2% SiO_2. In this case, the ore quality is highly variable, and therefore it was necessary to design the circuit to tolerate large fluctuations in feed rates. A high level of instrumentation was specified to provide online stream analysis, automatic reagent metering, and the ability to switch groups of columns into a standby mode. As a result, individual columns are automatically bypassed and idled with the reagent addition stopped during periods of decreased feed rates. When the feed increases to normal levels, the procedure is reversed and the columns resume normal operation.

Phosphate Flotation

Column flotation technology has proven to be an efficient method for improving the separation of valuable phosphate particles from the host matrix. As a result, many phosphate producers have installed column flotation systems as a means of boosting production while reducing operating costs. The approach has extended the size range of recoverable apatite particles from about 30 µm down to 5 µm (Wyslouzil 2009b). The high degree of selectivity achieved by this equipment has made it possible to treat material previously considered unrecoverable. A key characteristic of columns is the ability to operate with a deep froth, which is washed with fresh/clarified water, resulting in improved metallurgical performance. In some cases, the wash water can also partially offset the need for depressants, which lowers chemical dosage rates.

Conventional Versus Column

In Brazil, hard-rock phosphate deposits are volcanic in origin. These hard-rock ores contain apatite as the valuable component and, depending on the deposit, are associated with impurities such as barite, hematite, calcite, alumina, and silicates (e.g., quartz, micas, and clays). These ores are typically ground for liberation purposes and then subjected to coarse and fine flotation circuits. Mechanically agitated cells have been the standard choice for many years; however, this

Table 2 Comparison of mechanical and column flotation cells

Circuit Data	Mechanical Cell	Column Cell	% Change
Number of cells	66	6	–91
Power, kW·h/t	108	100	–7
Collector, g/t			
Barite circuit	567	170	–70
Apatite circuit	2,803	404	–86
Barite recovery	95	97	2
Apatite recovery			
Natural fines	50	78	56
Generated fines	65	77	18
Coarse/regrind	71	72	1
Fine product			
P_2O_5	33.5	34.7	4
Fe_2O_3	6.1	5.8	–5
Coarse product			
P_2O_5	36.2	35.9	–1
Fe_2O_3	3.1	3.1	0

equipment often struggles to produce high-grade concentrates because of the fine particle sizes. Fine impurities such as magnetite and silica are easily carried into the froth by hydraulic entrainment, which lowers the quality of the final concentrate. Also, attrition because of the cell agitators results in the continuous generation and release of fresh fines into the circuit. Consequently, fine particle flotation circuits require chemical depressants and multiple cleaning stages to reach acceptable grades. As a result, in the mid-1990s producers began to install column cells based on a desire to improve separation performance, reduce operating costs, and simplify plant circuitry. Table 2 provides an example of the potential benefits that can be realized using column technology. The data compare the results obtained after installation of a column circuit at a major phosphate plant in Brazil. The upgrade reduced the number of

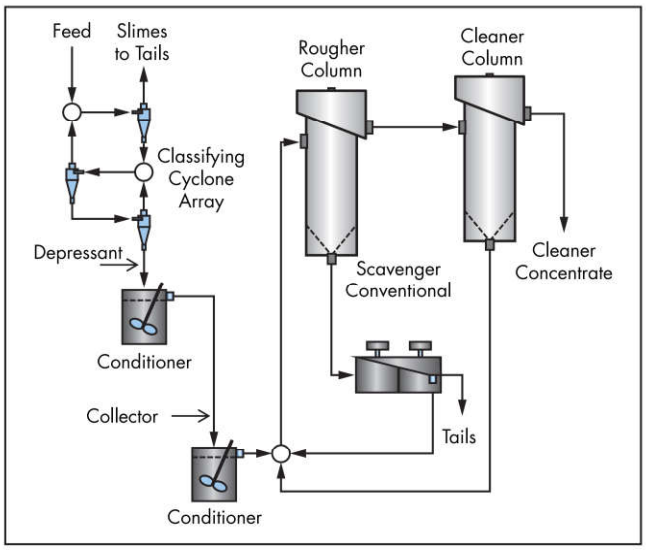
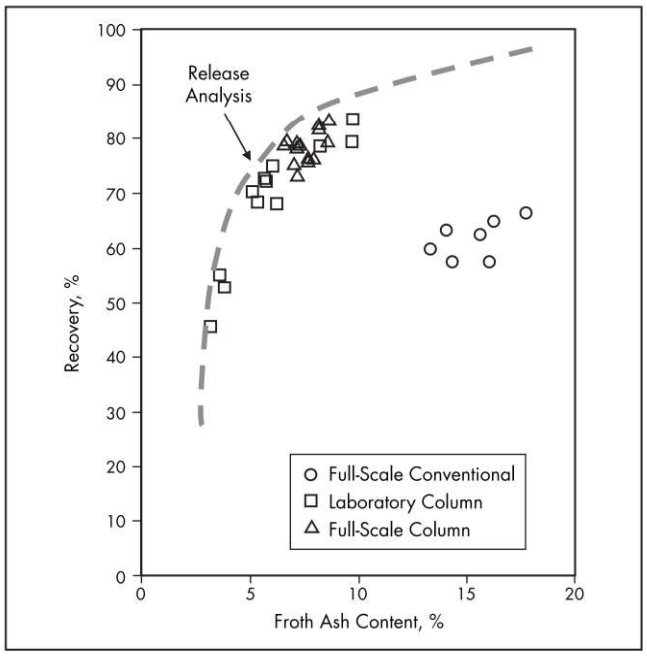


Figure 23 Circuit arrangement for ultrafine phosphate separation



Source: Davis et al. 1995

Figure 24 Comparison of conventional and column flotation relative to release analysis

cells in the plant from 66 machines to six columns. Despite using fewer pieces of equipment, the recovery of coarse and fine apatite and barite was improved. The retrofit also reduced power costs and lowered collector demand.

Ultrafine Flow Sheet

Another column flow sheet for a phosphate operation in Brazil is shown in Figure 23 (Wyslouzil 2009b). In this case, the major source of phosphate is derived from apatite, which comprises approximately 30% of the minerals in the ore zones. The major impurities consist of iron oxides and silicate minerals. The ore is crushed, screened, and then fed to the concentrator, where it is subjected to grinding (rod and ball mills), classification (hydrocyclones), magnetic separation, desliming (hydrocyclones), flotation, and dewatering. Traditionally, the primary desliming stage was designed to remove particles finer than 25–30 μm, which were then discarded. The high surface area and high impurity content associated with this particle size class made it difficult to treat with conventional flotation equipment. The removal of these slimes constituted a major loss of the total phosphate values since they represented as much as 10%–15% of the total reserves. The application of column flotation made it possible to extend the lower size range of particles that can be treated by flotation to 5–10 μm. In this circuit, the primary slimes (–30 μm) are treated in a second stage of hydrocyclones with a size cut of 5–10 μm. The cyclone underflow is then fed to a series of conditioners where the pulp is treated with caustic soda, starch, and a collector prior to column flotation. The flotation cells are arranged in a rougher–scavenger–cleaner circuit with intermediate products recycled internally.

Results from this circuit modification indicate that an ultrafine phosphate product can be produced containing 33.5% P_2O_5. This represents a 72% recovery of P_2O_5 for the ultrafine circuit and an overall increase in plant yield of 3%–5%. By following this treatment scheme, it is possible to obtain an ultrafine concentrate ideally suited for the production of single superphosphate fertilizers. The fine particle size minimizes the costs of concentrate regrinding at the fertilizer plant, saving additional processing costs. Faced with difficult ores and low-grade deposits, Brazilian phosphate producers have been world leaders in adapting this technology to enhance fine phosphate recovery.

Coal Flotation

Over the past decade, columns have been used extensively in the coal industry as an efficient and economical means of recovering fine coal from preparation plant effluent streams. Coal flotation applications are particularly challenging because the flotation system must cope with a high mass recovery rate and, typically, an extensive amount of ultrafine clay. Under these conditions, a deep froth and high-volume flow of wash water are particularly beneficial. The benefit of wash water is illustrated by the data summarized in Figure 24, which compares column flotation technology with an existing bank of conventional cells. As shown, the product quality is significantly improved using wash water. In fact, column performance agrees with the separation curve predicted by release analysis (Dell 1963; Dell et al. 1972). A release analysis is an indication of the ultimate separation performance attainable for flotation. This figure suggests that columns provide a level of performance that would be difficult to achieve even after multiple stages of cleaning by conventional machines.

Two circuits are typically used for fine coal flotation (Figure 25). These include the traditional minus 0.150-mm by-zero circuit and the 0.150 × 0.045-mm deslime circuit. The by-zero approach incorporates raw coal cyclones that treat –1 mm feed. Generally, these cyclones are configured to cut at a nominal particle top size of 0.15 mm, with the finer fraction reporting directly to flotation. In some markets, the particle size cut can be as coarse as 0.5 mm, although it is recommended to not exceed 0.25 mm for columns.

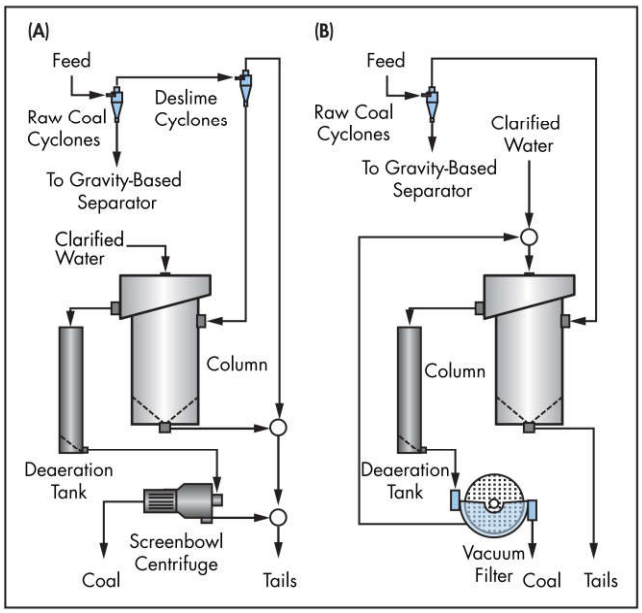


Figure 25 Coal flotation using (A) deslime column–screenbowl circuit and (B) by-zero column–filter circuit

By-Zero Column Circuits

The feed to flotation in a by-zero circuit is relatively dilute and typically ranges between 2.5% and 6.0% solids, by weight. As a result, volume flows in these circuits can be quite large. Consequently, significant cell volume must be available to achieve the desired retention time to ensure adequate combustible recovery. To attain the needed retention time, several large-diameter cells may be required, which can range in height from 8.5–16 m. The ultimate height is typically based on kinetic data generated from laboratory testing but is generally constrained by practical limits as dictated by both economic and site-specific engineering requirements. As a result, the height-to-diameter ratio for an industrial cell ranges from 3:1 to as low as 2:1. Also, because of the high flow volumes found in by-zero circuits, bypassing of feed can occur and is caused by the less-than-ideal mixing found within the large diameter and relatively short tanks commonly found in coal flotation installations. The high degree of mixing allows some material within the cell to report to tails without being influenced by the bubble swarm produced by the sparging system. As the residence time is increased, the probability of bypass is reduced. As a consequence of this relationship, the occurrence of bypass is more often found in traditional by-zero circuits where the volumetric feed rates are relatively high and retention times are typically low. More recent studies show that arranging column cells in series helps to minimize bypass (Stanley et al. 2006).

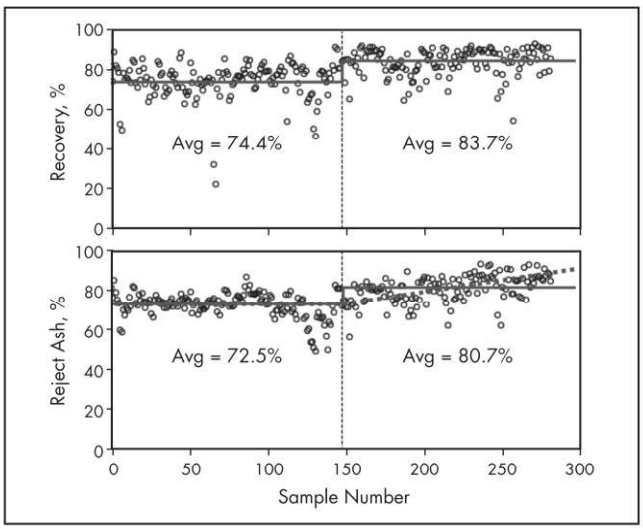
Deslime Column Circuits

The by-zero circuit works well in plants that serve the metallurgical coal market where ash is the primary quality specification; however, studies have shown that a by-zero approach is not the optimum circuitry for plants that are producing coal for the thermal market where other quality specifications such as moisture, heat content, and sulfur are also critical (Luttrell et al. 2009; Bethell and Luttrell 2005). These studies show that the popular deslime circuit is better suited to achieve a separation that maximizes profitability as dictated by a typical utility coal sales agreement. The deslime circuit achieves this rejection by incorporating a secondary bank of 150-mm-diameter cyclones, which is used to further classify the flotation feed at approximately 0.045 mm. This approach is superior for circuits that contain clay-rich flotation feeds where the benefit of wash water can be maximized.

Test results indicate that several benefits to the deslime flotation circuit are realized because of the rejection of high-surface-area particles. These include a lower-ash product, reduced froth moisture, an increase in flotation capacity, and an improvement in downstream froth handling. The improved froth handling is a result of two factors: a coarser float product and a significant reduction in total frother addition. The coarser froth product tends to be less stable and will readily collapse. The lower frother addition rate is a result of the volume split achieved in the smaller 150-mm-diameter cyclones, which substantially reduces the total volume of slurry that must be processed in flotation. Given that frother is added based on total flotation volume flow, this allows high concentrations of surfactant to be realized in the flotation circuit without causing upsets elsewhere in the plant. The obvious secondary benefit is the reduction in size and cost of the flotation and downstream dewatering equipment. Engineering estimates have shown that a deslime circuit may cost less than half of an equivalent by-zero circuit.

Low-Profile Column

More recently, there has been a move toward low-profile flotation cells designed specifically to overcome some of the challenges set forth by both end users and engineering firms with regard to traditional column cell installations. These challenges include cell size, foundation loads, and energy consumption. The benefits of this approach were demonstrated by Davis et al. (2011). In this case, StackCells were installed preceding two parallel column flotation cells to address a circuit-overloading condition. The net change in circuit performance is shown in Figure 26. As shown, the average coal recovery increased from 74.4%–83.7%. Likewise, the average tailings



Source: Davis et al. 2011

Figure 26 Change in plant performance after StackCell installation

ash content increased from 72.5%–80.7%. More recent data indicate that the circuit regularly achieves a combustible recovery exceeding 90%. The improvement is attributed to the optimization of operating variables such as reagent dosage, froth depth, aeration rate, and wash-water addition rate. The increased recovery is significant considering that less than 10% more cell volume was added to the circuit via the installation of the StackCell technology.

Sulfide Flotation

The aforementioned case studies describe applications in which columns are used in complex circuits as well as in stand-alone applications (i.e., coal). However, because of the ability to provide a high upgrade ratio, columns are often used as cleaners after mechanical cells in many applications. Columns are ideally suited for this purpose, particularly in sulfide mineral applications. Because of the high tonnage rate and the low feed grade associated with finely disseminated sulfide ores, conventional flotation cells are typically used for roughing and scavenging operations with the objective of producing a final tailings stream. After this initial upgrading, column cells are then used as cleaners to produce a high-grade concentrate prior to secondary grinding. Likewise, columns can also be used after regrinding to produce final copper and molybdenum concentrates.

The ability to operate a column cell with a deep froth and froth washing typically provides a higher-grade final concentrate. One such example from a currently operating concentrator is shown in Figure 27. In this case, columns were selected to replace a bank of conventional flotation cells. Performance data collected before and after the upgrade show a substantial increase in copper grade while also achieving a slight increase in copper recovery. The latter is attributed to the need to run a lean froth in conventional machines when difficult feed conditions are encountered to achieve the required final concentrate grade. By contrast, the application of froth washing allows the columns to be operated under optimal recovery conditions while maintaining a consistent grade.

A recent report in the technical literature (Leonida 2015) describes the successful application of a cleaner column circuit installed in a copper–gold operation in Australia. According to the article, this operation, completed in 2013, is one of the Asia-Pacific region's premier high-grade copper mines and will produce up to 300,000 t of high-grade copper concentrate annually. The plant was initially designed with all conventional flotation cells. However, after extensive laboratory- and pilot-scale testing, columns were added to the cleaning circuit to improve overall plant metallurgy. Results show an improvement in both product quality and overall circuit copper recovery.

In addition to copper sulfide circuits, columns are also advantageous in applications requiring a high upgrade ratio and/or a consistent final concentrate grade. They are routinely used for lead–zinc concentration, fine gold and silver recovery, and molybdenum concentration. Molybdenum typically occurs with copper sulfides and is recovered as a bulk sulfide flotation product. In this application, as discussed earlier, columns can be used for the bulk sulfide (copper and molybdenum) cleaner separation as well as the copper/molybdenum separation and subsequent cleaning. Columns with dynamic sparging systems are particularly beneficial in this latter application. The high superficial surface area rate (S_b) and the intense contacting of the sparging system are beneficial for ultrafine particle recovery situations.

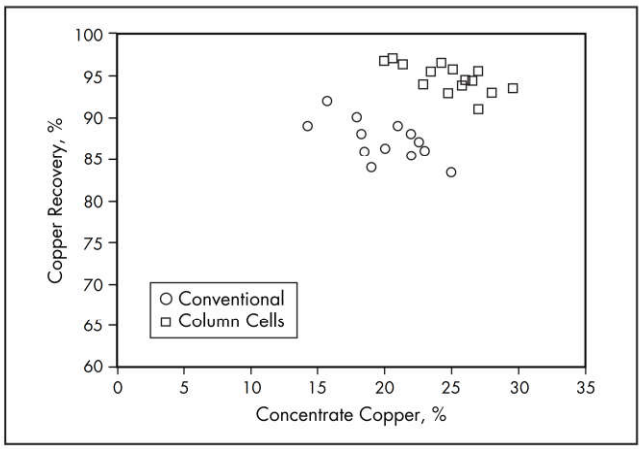


Figure 27 Column data for a copper cleaner application

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