Gravity concentration

Introduction
Gravity methods of separation are used to treat a great variety of materials, ranging from heavy metal sulphides such as galena (sp. gr. 7.5) to coal (sp. gr. 1.3), at particle sizes in some cases below 50 µm.

These methods declined in importance in the first half of the twentieth century due to the development of the froth-flotation process, which allows the selective treatment of low-grade complex ores. They remained, however, the main concentrating methods for iron and tungsten ores and are used extensively for treating tin ores, coal and many industrial minerals.

In recent years, many companies have re-evaluated gravity systems due to increasing costs of flotation reagents, the relative simplicity of gravity processes, and the fact that they produce comparatively little environmental pollution. Modern gravity techniques have proved efficient for concentration of minerals having particle sizes in the 50 µm range and, when coupled with improved pumping technology and instrumentation, have been incorporated in high-capacity plants (Holland-Batt, 1998). In many cases a high proportion of the mineral in an orebody can at least be pre-concentrated effectively by cheap and ecologically acceptable gravity systems; the amount of reagents and fuel used can be cut significantly when the more expensive methods are restricted to the processing of gravity concentrate. Gravity separation of minerals at coarser sizes as soon as liberation is achieved can also have significant advantages for later treatment stages due to decreased surface area, more efficient dewatering, and the absence of adhering chemicals which could interfere with further processing.

Gravity techniques to recover residual valuable heavy minerals in flotation tailings are being increasingly used. Apart from current production, there are many large tailings dumps which could be excavated cheaply and processed to give high value concentrates using recently developed technology.

Principles of gravity concentration
Gravity concentration methods separate minerals of different specific gravity by their relative movement in response to gravity and one or more other forces, the latter often being the resistance to motion offered by a viscous fluid, such as water or air.

It is essential for effective separation that a marked density difference exists between the mineral and the gangue. Some idea of the type of separation possible can be gained from the concentration criterion

\[
\frac{D_h - D_l}{D_l - D_f} = 10.1
\]

where \(D_h\) is the specific gravity of the heavy mineral, \(D_l\) is the specific gravity of the light mineral, and \(D_f\) is the specific gravity of the fluid medium.

In very general terms, when the quotient is greater than 2.5, whether positive or negative, then gravity separation is relatively easy, the efficiency of separation decreasing as the value of the quotient decreases.

The motion of a particle in a fluid is dependent not only on its specific gravity, but also on its size (Chapter 9); large particles will be affected more than smaller ones. The efficiency of gravity processes therefore increases with particle size, and
the particles should be sufficiently coarse to move in accordance with Newton’s law (Equation 9.6). Particles which are so small that their movement is dominated mainly by surface friction respond relatively poorly to commercial high-capacity gravity methods. In practice, close size control of feeds to gravity processes is required in order to reduce the size effect and make the relative motion of the particles specific gravity-dependent.

Gravity separators

Many different machines have been designed and built in the past to effect separation of minerals by gravity, and they are comprehensively reviewed by Burt (1985). Many gravity devices have become obsolete, and only equipment that is used in modern mills will be described in this chapter. Contact details for manufacturers of gravity concentrators can be found in the annual Buyer's Guide of products and services published by Mining Magazine in December each year.

A classification of the more commonly used gravity separators on the basis of feed size range is shown in Figure 1.8.

The dense medium separation (DMS) process is widely used to preconcentrate crushed material prior to grinding and will be considered separately in the next chapter.

It is essential for the efficient operation of all gravity separators that the feed is carefully prepared. Grinding is particularly important in adequate liberation; successive regrinding of middlings is required in most operations. Primary grinding should be performed where possible in open-circuit rod mills, but if fine grinding is required, closed-circuit ball milling should be used, preferably with screens closing the circuits rather than hydrocyclones in order to reduce selective overgrinding of heavy friable valuable minerals.

Gravity separators are extremely sensitive to the presence of slimes (ultra-fine particles), which increase the viscosity of the slurry and hence reduce the sharpness of separation, and obscure visual cut-points. It is common practice in most gravity concentrators to remove particles less than about 10 μm from the feed, and divert this fraction to the tailings, and this can account for considerable loss of values. De-slime is often achieved by the use of hydrocyclones, although if hydraulic classifiers are used to prepare the feed it may be preferable to de-slime at this stage, since the high shear forces produced in hydrocyclones tend to cause degradation of friable minerals.

The feed to jigs, cones, and spirals should, if possible, be screened before separation takes place, each fraction being treated separately. In most cases, however, removal of the oversize by screening, in conjunction with de-slimeing, is adequate. Processes which utilise flowing-film separation, such as shaking tables and tilting frames, should always be preceded by good hydraulic classification in multi-spigot hydrosizers.

Although most slurry transportation is achieved by centrifugal pumps and pipelines, as much as possible should be made of natural gravity flow; many old gravity concentrators were built on hillsides to achieve this. Reduction of slurry pumping to a minimum not only reduces energy consumption, but also reduces slimes production in the circuit. To minimise degradation of friable minerals, slurry pumping velocities should be as low as possible, consistent with maintaining the solids in suspension.

One of the most important aspects of gravity circuit operations is correct water balance within the plant. Almost all gravity concentrators have an optimum feed pulp-density, and relatively little deviation from this density causes a rapid decline in efficiency. Accurate pulp-density control is therefore essential, and this is most important on the raw feed. Automatic density control should be used where possible, and the best way of achieving this is by the use of nucleonic density gauges (Chapter 3) controlling the water addition to the new feed. Although such instrumentation is expensive, it is usually economic in the long term. Control of pulp density within the circuit can be made by the use of settling cones preceding the gravity device. These thicken the pulp, but the overflow often contains solids, and should be directed to a central large sump or thickener. For substantial increase in pulp density, hydrocyclones or thickeners may be used. The latter are the more expensive, but produce less particle degradation and also provide substantial surge capacity. It is usually necessary to recycle water in most plants, so adequate thickener or cyclone capacity should be provided, and slimes build-up in the recycled water must be minimised.
If the ore contains an appreciable amount of sulphide minerals then, if the primary grind is finer than about 300 μm, these should be removed by froth flotation prior to gravity concentration, as they reduce the performance of spirals, tables, etc. If the primary grind is too coarse for effective sulphide flotation, then the gravity concentrate must be reground prior to removal of the sulphides. The sulphide flotation tailing is then usually cleaned by further gravity concentration.

The final gravity concentrate often needs cleaning by magnetic separation, leaching, or some other method, in order to remove other mineral contaminants. For instance, at the South Crofty tin mine in Cornwall, the gravity concentrate was subjected to cleaning by magnetic separators, which removed wolframite from the cassiterite product.

The design and optimisation of gravity circuits is discussed by Wells (1991).

**Jigs**

Jigging is one of the oldest methods of gravity concentration, yet the basic principles are only now beginning to be understood. A mathematical model developed by Jonkers et al. (1998) shows considerable promise in predicting jig performance on a size by density basis.

The jig is normally used to concentrate relatively coarse material and, if the feed is fairly closed sized (e.g. 3–10 mm), it is not difficult to achieve good separation of a fairly narrow specific gravity range in minerals in the feed (e.g. fluorite, sp. gr. 3.2, from quartz, sp. gr. 2.7). When the specific gravity difference is large, good concentration is possible with a wider size range. Many large jig circuits are still operated in the coal, cassiterite, tungsten, gold, barytes, and iron-ore industries. They have a relatively high unit capacity on classified feed and can achieve good recovery of values down to 150 μm and acceptable recoveries often down to 75 μm. High proportions of fine sand and slime interfere with performance and the fines content should be controlled to provide optimum bed conditions.

In the jig the separation of minerals of different specific gravity is accomplished in a bed which is rendered fluid by a pulsating current of water so as to produce stratification. The aim is to dilate the bed of material being treated and to control the dilation so that the heavier, smaller particles penetrate the interstices of the bed and the larger high specific gravity particles fall under a condition probably similar to hindered settling (Lyman, 1992).

On the pulsion stroke the bed is normally lifted as a mass, then as the velocity decreases it tends to dilate, the bottom particles falling first until the whole bed is loosened. On the suction stroke it then closes slowly again and this is repeated at every stroke, the frequency usually varying between 55 and 330 c m⁻¹. Fine particles tend to pass through the interstices after the large ones have become immobile. The motion can be obtained either by using a fixed sieve jig, and pulsating the water, or by employing a moving sieve, as in the simple hand-jig (Figure 10.1).

![Hand Jig](image)

**Figure 10.1 Hand jig**

**The jigging action**

It was shown in Chapter 9 that the equation of motion of a particle settling in a viscous fluid is

\[
m \frac{dx}{dt} = mg - m'g - D \tag{9.1}
\]

where \( m \) is the mass of the mineral grain, \( \frac{dx}{dt} \) is the acceleration, \( g \) is the acceleration due to gravity, \( m' \) is the mass of displaced fluid, and \( D \) is the fluid resistance due to the particle movement.

At the beginning of the particle movement, since the velocity \( x \) is very small, \( D \) can be ignored as it is a function of velocity.

Therefore

\[
\frac{dx}{dt} = \left( \frac{m - m'}{m} \right) g \tag{10.2}
\]
and since the particle and the displaced fluid are of equal volume,

\[
\frac{dx}{dt} = \left( \frac{D_s - D_f}{D_s} \right) g \\
= \left( \frac{1 - D_f}{D_s} \right) g
\]

(10.3)

where \(D_s\) and \(D_f\) are the respective specific gravities of the solid and the fluid.

The initial acceleration of the mineral grains is thus independent of size and dependent only on the densities of the solid and the fluid. Theoretically, if the duration of fall is short enough and the repetition of fall frequent enough, the total distance travelled by the particles will be affected more by the differential initial acceleration, and therefore by density, than by their terminal velocities and therefore by size. In other words, to separate small heavy mineral particles from large light particles a short jiggling cycle is necessary. Although relatively short fast strokes are used to separate fine minerals, more control and better stratification can be achieved by using longer, slower strokes, especially with the coarser particle sizes. It is therefore good practice to screen the feed to jigs into different size ranges and treat these separately. The effect of differential initial acceleration is shown in Figure 10.2.

If the mineral particles are examined after a longer time they will have attained their terminal velocities and will be moving at a rate dependent on their specific gravity and size. Since the bed is really a loosely packed mass with interstitial water providing a very thick suspension of high density, hindered-settling conditions prevail, and the settling ratio of heavy to light minerals is higher than that for free settling (Chapter 9). Figure 10.3 shows the effect of hindered settling on the separation.

The upward flow can be adjusted so that it overcomes the downward velocity of the fine light particles and carries them away, thus achieving separation. It can be increased further so that only large heavy particles settle, but it is apparent that it will not be possible to separate the small heavy and large light particles of similar terminal velocity.

Hindered settling has a marked effect on the separation of coarse minerals, for which longer, slower strokes should be used, although in practice, with coarser feeds, it is improbable that the larger particles have time to reach their terminal velocities.

At the end of a pulsion stroke, as the bed begins to compact, the larger particles interlock, allowing the smaller grains to move downwards through the interstices under the influence of gravity. The fine grains may not settle as rapidly during this consolidation trickling phase (Figure 10.4) as during the initial acceleration or suspension, but if consolidation trickling can be made to last long enough, the effect, especially in the recovery of the fine heavy minerals, can be considerable.
Figure 10.5 Ideal jigging process

At the point of transition between the pulsion and the suction stroke, at point E, the bed will be compacted. Consolidation trickling can now occur to a limited extent. In a closely sized ore the heavy grains can now penetrate only with difficulty through the bed and may be lost to the tailings. Severe compaction of the bed can be reduced by the addition of *hutch water*, a constant volume of water, which creates a constant upward flow through the bed. This flow, coupled with the varying flow caused by the piston, is shown in Figure 10.8. Thus suction is reduced by hutch-water addition, and is reduced in duration; by adding a large quantity of water, the suction may be entirely eliminated. The coarse ore then penetrates the bed more easily and the horizontal transport of the feed over the jig is also improved. However, fines losses will increase, partly because of the longer duration of the pulsion stroke, and partly because the added water increases the speed of the top flow.

Figure 10.6 Movement of the piston in a jig

Figure 10.7 Speed of flow through bed during jig cycle

Increases, the grains will be loosened and the bed will be forced open, or dilated. At, say, point B, the grains are in the phase of hindered settling in an upward flow, and since the speed of flow from B to C still increases, the fine grains are pushed upwards by the flow. The chance of them being carried along with the top flow into the tailings is then at its greatest. In the vicinity of D, first the coarser grains and later on the remaining fine grains will fall back. Due to the combination of initial acceleration and hindered settling, it is mainly the coarser grains that will lie at the bottom of the bed.

Types of jig

Essentially the jig is an open tank filled with water, with a horizontal jig screen at the top, and provided with a spigot in the bottom, or *hutch* compartment, for concentrate removal (Figure 10.9). Current
types of jig are reviewed by Cope (2000). The jig bed consists of a layer of coarse, heavy particles, or ragging, placed on the jig screen on to which the slurry is fed. The feed flows across the ragging and the separation takes place in the jig bed so that grains with a high specific gravity penetrate through the ragging and screen to be drawn off as a concentrate, while the light grains are carried away by the cross-flow to be discarded as tailings. The harmonic motion produced by the eccentric drive is supplemented by a large amount of continuously supplied hutch water, which enhances the upward and diminishes the downward velocity of the water (Figure 10.8).

One of the oldest types of jig is the Harz (Figure 10.10) in which the plunger moves up and down vertically in a separate compartment. Up to four successive compartments are placed in series in the hutch. A high-grade concentrate is produced in the first compartment, successively lower grades being produced in the other compartments, tailings overflowing the final compartment. If the feed particles are larger than the apertures of the screen, jigging “over the screen” is used, and the concentrate grade is partly governed by the thickness of the bottom layer, determined by the rate of withdrawal through the concentrate discharge port.

The Denver mineral jig (Figure 10.11) is widely used, especially for removing heavy minerals from closed grinding circuits, thus preventing over-grinding. The rotary water valve can be adjusted so as to open at any desired part of the jig cycle, synchronisation between the valve and the plungers being achieved by a rubber timing belt. By suitable adjustment of the valve, any desired variation can be achieved, from complete neutralisation of the suction stroke with hydraulic water to a full balance between suction and pulsion.

Conventional mineral jigs consist of square or rectangular tanks, sometimes combined to form two, three, or four cells in series. In order to compensate for the increase in cross-flow velocity over the jig bed, caused by the addition of hutch water, trapezoidal-shaped jigs were developed. By arranging these as sectors of a circle, the modular circular, or radial, jig was introduced, in which the feed enters in the centre and flows radially over the jig bed towards the tailings discharge at the circumference (Figure 10.12).

The main advantage of the circular jig is its very large capacity, and IHC Radial Jigs (Figure 10.13) have been installed on most newly built tin dredges in Malaysia and Thailand since their development in 1970. They are also in use for the treatment of gold, diamonds, iron ore, etc., the largest, of 7.5 m in diameter, being capable of treating up to 300 m$^3$ h$^{-1}$ of feed with a maximum particle size of 25 mm. In the IHC jig, the harmonic motion of the conventional eccentric-driven jig is replaced by an asymmetrical “saw-tooth” movement of the diaphragm, with a rapid upward, followed by a slow downward, stroke (Figure 10.14). This produces a much larger and more constant suction stroke, giving the finer particles more time to settle in the bed, thus reducing their loss to tailings, the jig being capable of accepting particles as fine as 60 microns.

The InLine Pressure Jig (IPJ) is a recent Australian development in jig technology which is finding wide application for the recovery of free gold, sulphides, native copper, tin/tantalum, diamonds and other minerals (Figure 10.15). The IPJ is unique in that it is fully encapsulated and pressurised, allowing it to be completely filled with
Figure 10.11 Denver mineral jig

Figure 10.12 (a) Outline of circular jig; (b) radial jig up to twelve modules

Figure 10.13 IHC modular radial jig
slurry (Gray, 1997). It combines a circular bed with a vertically pulsed screen. Length of stroke and pulsation frequency, as well as screen aperture, can all be altered to suit the application. IPJs are typically installed in grinding circuits, where their low water requirements allow operators to treat the full circulating load, maximising recovery of liberated values. Both concentrates and tailings are discharged under pressure.

In an attempt to recover fine particles using gravity concentration methods, jigs have been developed to make use of centrifugal force. Such jigs have all the elements of a standard jig except that the bed of mineral is rotated at high speed while being pulsed. The Kelsey Centrifugal Jig (KCJ) takes a conventional jig and spins it in a centrifuge. The ability to change the apparent gravitational field is a major departure in the recovery of fine minerals. The main operating variables which are adjusted to control processing of different types of material are centrifugal force, ragging material, and size distribution. The 16 hutch J1800 KCJ can treat over 100 t/h, depending on the application. The use of a J650 KCJ in tin recovery is described by Beniu et al. (1994).

**Coal jigs**

Jigs are widely used coal-cleaning devices, and are preferred to the more expensive dense medium process when the coal has relatively little middlings, or "near-gravity" material, as is often the case with British coals. No feed preparation is required, as is necessary with DMS, and for coals which are easily washed, i.e. those consisting predominantly of liberated coal and denser rock particles, the lack of close density control is not a disadvantage.

Two types of air-pulsated jig – Baum and Batac – are used in the coal industry. The standard Baum jig (Figure 10.16), with some design modifications (Green, 1984; Harrington, 1986), has been used for nearly 100 years, and is still the dominant device. Air under pressure is forced into a large air chamber on one side of the jig vessel causing pulsations and suction to the jig water, which in turn causes pulsations and suction through the screen plates upon which the raw coal is fed, thus causing stratification. Various methods are used to continuously separate the refuse from the lighter coal product, and all the modern Baum jigs are fitted with some form of automatic refuse extraction (Adams, 1983). One form of control incorporates a float immersed in the bed of material. The float is suitably weighed to settle on the dense layer of refuse moving across the screen plates. An increase in the depth of refuse raises the float, which automatically controls the refuse discharge, either by
adjusting the height of a moving gate, or by controlling the pulsating water which lifts the rejects over a fixed weir plate (Wallace, 1979). This system is reported to respond quickly and accurately.

In Britain it is now commonplace for the automatic control system to determine the variations in refuse bed thickness by measuring the differences in water pressure under the screen plates arising from the resistance offered to pulsation. The JigScan control system developed at the Julius Kruttschnitt Mineral Research Centre measures bed conditions and pulse velocity many times within the pulse using pressure sensing and nucleonic technology (Loveday and Jonkers, 2002). Evidence of a change in the pulse is an indicator of a fundamental problem with the jig, allowing the operator to take corrective action. Increased yields of greater than 2 per cent have been reported for JigScan-controlled jigs.

In many situations the Baum jig still performs satisfactorily, with its ability to handle large tonnages (up to 1000 t h\(^{-1}\)) of coal of a wide size range. However, the distribution of the stratification force, being on one side of the jig, tends to cause unequal force along the width of jig screen and therefore uneven stratification and some loss in the efficiency of separation of the coal from its heavier impurities. This tendency is not so important in relatively narrow jigs, and in the United States multiple float and gate mechanisms have been used to counteract the effects. The Batac jig (Zimmerman, 1975) is also pneumatically operated (Figure 10.17), but has no side air chamber like the Baum jig. Instead, it is designed with a series of multiple air chambers, usually two to a cell, extending under the jig for its full width, thus giving uniform air distribution. The jig uses electronically controlled air valves which provide a sharp cut-off of the air input and exhaust. Both inlet and outlet valves are infinitely variable with regard to speed and length of stroke, allowing for the desired variation in pulsation and suction by which proper stratification of the bed may be achieved for differing raw coal characteristics. As a result, the Batac jig can wash both coarse and fine sizes well (Chen, 1980). The jig has also been successfully utilised to produce high-grade lump ore and sinter-feed concentrates from such iron ore deposits which cannot be upgraded by heavy-medium techniques (Hasse and Wasmuth, 1988; Miller, 1991).

**Pinched sluices and cones**

Pinched sluices of various forms have been used for heavy mineral separations for centuries. In its simplest form (Figure 10.18), it is an inclined launder about 1 m long, narrowing from about 200 mm in width at the feed end to about 25 mm at the discharge. Pulp of between 50 and 65% solids enters gently and stratifies as it descends; at the discharge end these strata are separated by various means, such as by splitters, or by some type of tray (Sivamohan and Forssberg, 1985b).
Figure 10.19 shows pinched sluices in operation on an Australian heavy mineral sand concentrator. The fundamental basis for gravity concentration in sluices is described by Schubert (1995).

The Reichert cone is a wet gravity concentrating device designed for high-capacity applications. Its principle of operation is similar to that of a pinched sluice, but the pulp flow is not restricted or influenced by side-wall effect, which is somewhat detrimental to pinched-sluice operation.

The Reichert cone concentrator was developed in Australia in the early 1960s primarily to treat titanium-bearing beach sands, and the success of cone circuits has led to their application in many other fields.

The single unit comprises several cone sections stacked vertically (Figure 10.20), so as to permit several stages of upgrading. The cones are made of fibreglass and are mounted in circular frames over 6 m high. Each cone is 2 m in diameter and there are no moving parts in the unit. A cross-section through a Reichert cone system is shown in Figure 10.21. The system shown is one of many possible systems using double and single cones, together with trays, which direct heavy mineral fractions from the centre draw-off areas of the cones to external collection boxes and also serve to further upgrade the fraction, acting as a sort of pinched sluice.

The feed pulp is distributed evenly around the periphery of the cone. As it flows towards the centre of the cone the heavy particles separate to the bottom of the film. This concentrate is removed by an annular slot in the bottom of the concentrating cone; the part of the film flowing over the slot is the tailings. The efficiency of this separation process is relatively low and is repeated a number of times within a single machine to achieve effective performance. A typical machine system for rougher concentration duties will consist of four double single-cone stages in series, each retreat ing the tailings of the preceding stage. This machine
Reichert cones will produce a concentrate from the upper three stages and the product from the fourth stage as middlings (Anon., 1977). A cone concentration system can be set to produce a marked reverse classification effect, in that the slots tend to reject coarser, lighter particles to tailings, while retaining finer, heavier particles in the concentrate flows.

Reichert cones have a high capacity, operating normally in the range 65–90 t h⁻¹, but in exceptional cases this can be from 40 to 100 t h⁻¹, with a feed density of between 55 and 70% solids by weight. They accept feeds of up to 3 mm in size and can treat material as fine as 30 μm, although they are most efficient in the 100–600 μm size range (Forssberg and Sandström, 1979). In recent years the capacity of cones has been increased by increasing the cone diameter to 3.5 m. The longer deck not only provides improved throughput but also greater upgrading per cone. The equivalent overall metallurgical performance can be achieved using fewer stages in a 3.5 m diameter cone (Richards and Palmer, 1997).

The success of cone circuits in the Australian mineral sand industry has led to their application in other fields. Preconcentration of tin and gold, the recovery of tungsten, and the concentration of magnetite are all successful applications. In many of these applications, cones, due to their high capacities and low operating costs, have replaced spirals and shaking tables.

One of the largest single installations is operated at Palabora Mining Co. in South Africa. Sixty-eight Reichert cones treat 34,000 t d⁻¹ of flotation tailings, after preliminary de-sliming and low intensity...
magnetic separation for removal of magnetite. A complex circuit consists of 48 rougher-scavenger units, each with a six-cone configuration, and 20 cleaner recleaners with eight-cone configurations.

The cone section of the plant produces an upgrading of about 200:1, with a recovery of 85% of +45 μm material. The concentrate is further upgraded on shaking tables to produce final concentrates of uranothorite and baddeleyite.

Cone circuits have been successfully used as preconcentrators ahead of flotation to recover gold and silver values from base metal sulphides in a number of operations in Scandinavia, Papua New Guinea, and Australia.

King (2000) describes the use of the MODSIM flowsheet simulator to evaluate different cone configurations in rougher, scavenger, and cleaner applications.

Spirals

Spiral concentrators have, over numerous years, found many varied applications in mineral processing, but perhaps their most extensive usage has been in the treatment of heavy mineral sand deposits, such as those carrying ilmenite, rutile, zircon, and monazite (see Chapter 13), and in recent years in the recovery of fine coal.

The Humphreys spiral (Figure 10.22) was introduced in 1943, its first commercial application being on chrome-bearing sands (Hubbard et al., 1953). It is composed of a helical conduit of modified semicircular cross-section. Feed pulp of between 15 and 45% solids by weight and in the size range 3 mm to 75 μm is introduced at the top of the spiral and, as it flows spirally downwards, the particles stratify due to the combined effect of centrifugal force, the differential settling rates of the particles, and the effect of interstitial trickling through the flowing particle bed. These mechanisms are complex, being much influenced by the slurry density and particle size. Some workers (Mills, 1978) have reported that the main separation effect is due to hindered settling, with the largest, densest particles reporting preferentially to the concentrate, which forms in a band along the inner edge of the stream (Figure 10.23). Bonsu (1983), however, reported that the net effect is reverse classification, the smaller, denser particles preferentially entering the concentrate band.

Ports for the removal of the higher specific-gravity particles are located at the lowest points in the cross-section. Wash-water, added at the inner edge of the stream, flows outwardly across the concentrate band. The width of concentrate band removed at the ports is controlled by adjustable splitters. The grade of concentrate taken from descending ports progressively decreases, tailings being discharged from the lower end of the spiral conduit.

Until relatively recently, all spirals were very similar, based upon the original Humphreys design, which is now obsolete. However, in recent years, there have been considerable developments in spiral technology, and a wide range of devices are now available. The main areas of development have been in the introduction of spirals with only one concentrate take-off, at the bottom of the spiral, and the use of spirals without added wash water. Wash waterless spirals reportedly offer lower cost, easy operation, and simple maintenance, and have been installed at several gold and tin processing plants. Holland-Batt (1995) discusses the design considerations affecting the pitch of the trough and the trough shape, while the modern high capacity spirals which result from careful design are described by Richards and Palmer (1997).
A comprehensive semi-empirical mathematical model of the spiral has been developed by Holland-Batt (1989), while a detailed CFD model of the fluid flow has been developed and validated by Matthews et al. (1998).

Improved spiral concentrator design has offered an efficient and economic alternative to the Reichert cone concentrator (Ferree, 1993). Davies et al. (1991) have reviewed the development of new spiral models and describe the mechanism of separation and the effects of operating parameters. Some of the traditional areas of spiral application are described, together with examples of new applications such as treatment of fine alluvials and tailings, and fine coal recovery. One of the most important developments in fine coal washing was the introduction in the 1980s of spiral separators specifically designed for coal. It is common practice to separate down to 0.5 mm coal using dense medium cyclones (Chapter 11), and below this by froth flotation. Spiral circuits have been installed to process the size range which is least effectively
treated by these two methods, typically 0.1–2 mm (Weale and Swanson, 1991).

Double-spiral concentrators, with two starts integrated into the one space around a common column, have been used in Australia for many years and have also been accepted elsewhere. At Mount Wright in Canada 4300 double-start spirals are upgrading specular hematite ore at 6900t h⁻¹ at 86% recovery (Hyma and Meech, 1989).

Spirals are made with slopes of varying steepness, the angle affecting the specific gravity of separation, but having little effect on the concentrate grade and recovery. Shallow angles are used, for example, to separate coal from shale, while steeper angles are used for normal heavy mineral-silica separations. The steepest angles are used to separate heavy minerals from heavy waste minerals, for example zircon (sp. gr. 4.7) from kyanite and staurolite (sp. gr. 3.6). Capacity ranges from 1 to 3 t h⁻¹ on low slope spirals to about double this for the steeper units. Spiral length is usually five or more turns for roughing duty and three turns in some cleaning units. Because treatment by spiral separators involves a multiplicity of units, the separation efficiency is very sensitive to the pulp distribution system employed. Lack of uniformity in feeding results in substantial falls in operating efficiency and can lead to severe losses in recovery, this is especially true with coal spirals (Holland-Batt, 1993).

**Shaking tables**

When a flowing film of water flows over a flat, inclined surface the water closest to the surface is retarded by the friction of the water absorbed on the surface; the velocity increases towards the water surface. If mineral particles are introduced into the film, small particles will not move as rapidly as large particles, since they will be submerged in the slower-moving portion of the film. Particles of high specific gravity will move more slowly than lighter particles, and so a lateral displacement of the material will be produced (Figure 10.24).

The flowing film effectively separates coarse light particles from small dense particles, and this mechanism is utilised to some extent in the shaking-table concentrator (Figure 10.25), which is perhaps the most metallurgically efficient form of gravity concentrator, being used to treat the smaller, more difficult flow-streams, and to produce finished concentrates from the products of other forms of gravity system.

![Figure 10.24 Action in a flowing film](image_url)

![Figure 10.25 Shaking table](image_url)
It consists of a slightly inclined deck, A, on to which feed, at about 25% solids by weight, is introduced at the feed box and is distributed along C; wash water is distributed along the balance of the feed side from launder D. The table is vibrated longitudinally, by the mechanism B, using a slow forward stroke and a rapid return, which causes the mineral particles to "crawl" along the deck parallel to the direction of motion. The minerals are thus subjected to two forces, that due to the table motion and that, at right angles to it, due to the flowing film of water. The net effect is that the particles move diagonally across the deck from the feed end and, since the effect of the flowing film depends on the size and density of the particles, they will fan out on the table, the smaller, denser particles riding highest towards the concentrate launder at the far end, while the larger lighter particles are washed into the tailings launder, which runs along the length of the table. Figure 10.26 shows an idealised diagram of the distribution of table products. An adjustable splitter at the concentrate end is often used to separate this product into two fractions - a high-grade concentrate and a middlings fraction. Maximum height on the feed side, till they die out near the opposite side, part of which is left smooth (Figure 10.26). In the protected pockets behind the riffles the particles stratify so that the finest and heaviest particles are at the bottom and the coarsest and lightest particles are at the top (Figure 10.27). Layers of particles are moved across the riffles by the crowding action of new feed and by the flowing film of wash water. Due to the taper of the riffles, progressively finer sized and higher density particles are continuously being brought into contact with the flowing film of water that tops the riffles. Final concentration takes place at the unriffled area at the end of the deck, where the layer of material is at this stage usually only one or two particles deep.

The significance of the many design and operating variables and their interactions have been reviewed by Sivamohan and Forssberg (1985a), and the development of a mathematical model of a shaking table is described by Manser et al. (1991). The separation on a shaking table is controlled by a number of operating variables, such as wash water, feed pulp density, deck slope, amplitude, and feed rate, and the importance of these variables in the model development is discussed.

Many other factors, including particle shape and the type of deck, play an important part in table separations. Flat particles, such as mica, although light, do not roll easily across the deck in the water film; such particles cling to the deck and are carried down to the concentrate discharge. Likewise, spherical dense particles may move easily in the film towards the tailings launder.

The table decks are usually constructed of wood, lined with materials with a high coefficient of friction, such as linoleum, rubber, and plastics. Decks made from fibreglass are also used which, although more expensive, are extremely hard wearing. The riffles on such decks are incorporated as part of the mould.
Particle size plays a very important role in table separations; as the range of sizes in a table feed increases, the efficiency of separation decreases. If a table feed is made up of a wide range of particle sizes, some of these sizes will be cleaned inefficiently. As can be seen from Figure 10.26, in an idealised separation, the middlings produced are not "true middlings", i.e. particles of associated mineral and gangue, but relatively coarse dense particles and fine light particles. If these particles are returned to the grinding circuit, together with the true middlings, then they will be needlessly reground.

Since the shaking table effectively separates coarse light from fine dense particles, it is common practice to classify the feed, since classifiers put such particles into the same product, on the basis of their equal settling rates. In order to feed as narrow a size range as possible on to the table, classification is usually performed in multi-spigot hydrosizers (Chapter 9), each spigot product, comprising a narrow range of equally settling particles, being fed to a separate set of shaking tables. A typical gravity concentrator employing shaking tables (Figure 10.28) may have an initial grind in rod mills in order to liberate as much mineral at as coarse a size as possible to aid separation. The hydrosizer products feed separate sets of tables, the middlings being reground before returning to the hydrosizer. Riffled tables, or sand tables, normally operate on feed sizes in the range 3 mm to 100 μm, and the hydrosizer overflow, consisting primarily of particles finer than this, is usually thickened and then distributed to tables whose decks have a series of planes rather than ripples and are designated slime tables.

Dewatering of the hydrosizer overflow is often performed by hydrocyclones, which also remove particles in the overflow smaller than about 10 μm, which will not separate efficiently by gravity methods due to their extremely slow settling rates. Successive stages of regrinding are a feature of many gravity concentrators. The mineral is separated at all stages in as coarse a state as possible in order to achieve reasonably fast separation and hence high throughputs.

The capacity of a table varies according to size of feed particles and the concentration criteria. Tables can handle up to 2 t h⁻¹ of 1.5 mm sand and perhaps 1 t h⁻¹ of fine sand. On 100–150 μm feed materials, table capacities may be as low as 0.5 t h⁻¹. On coal feeds, however, which are often tabled at sizes of up to 15 mm, much higher capacities are common. A normal 5 mm raw coal feed can be tabled with high efficiency at 12.5 t h⁻¹ per deck, whilst tonnages as high as 15.0 t h⁻¹ per deck are not uncommon when the top size in the feed is 15 mm (Terry, 1974).

The introduction of double and triple-deck units (Figure 10.29) has improved the area/capacity ratio at the expense of some flexibility and control. Separation can be influenced by the length of stroke, which can be altered by means of a handwheel on the vibrator, or head motion, and by the reciprocating speed (Figure 10.30). The length of stroke usually varies within the range of 10–25 mm or more, the speed being in the range 240–325 strokes per minute. Generally, a fine feed requires a higher speed and shorter stroke which increases in speed as it goes forward until it is jerked to a halt before being sharply reversed, allowing the particles to slide forward during most of the backward stroke due to their built-up momentum.

The quantity of water used in the feed pulp varies, but for ore tables normal feed dilution is 20–25% solids by weight, while for coal tables pulps of 33–40% solids are used. In addition to the water in the feed pulp, clear water flows over the table for final concentrate cleaning. This varies from a few litres to almost 1001 min⁻¹ according to the nature of the feed material. Tables slope from the feed to the tailings discharge end and the correct angle of incline is obtained by means of a handwheel. In most cases the line of separation is clearly visible on the table so this adjustment is easily made.
The table is slightly elevated along the line of motion from the feed end to the concentrate end. The moderate slope, which the high-density particles climb more readily than the low-density minerals, greatly improves the separation, allowing much sharper cuts to be made between concentrate, middlings, and tailings. The correct amount of end elevation varies with feed size and is greatest for the coarsest and highest gravity feeds. The end elevation should never be less than the taper of the riffles, otherwise there is a tendency for water to flow out towards the ripple tips rather than across the riffles. Normal end elevations in ore tabling range from a maximum of 90 mm for a very heavy, coarse sand to as little as 6 mm for an extremely fine feed.

Ore-concentrating tables are used primarily for the concentration of minerals of tin, iron, tungsten, tantalum, mica, barium, titanium, zirconium, and, to a lesser extent, gold, silver, thorium, uranium, and others. Tables are now being used in the recycling of electronic scrap to recover precious metals.

Pneumatic tables

Originally developed for seed separation, pneumatic or air tables have an important use in the treatment of heavy mineral sand deposits, and in the upgrading of asbestos, and in applications where water is at a premium. Pneumatic tables use a throwing motion to move the feed along a flat riffled deck, and blow air continuously up through a porous bed. The stratification produced is somewhat different from that of wet tables. Whereas in wet tabling the particle size increases and the density decreases from the top of the concentrate band to the tailings, on an air table both particle size and density decrease from the top down, the coarsest particles in the middlings band having the lowest density. Pneumatic tabling is therefore similar in effect to hydraulic classification. They are commonly used in combination with wet tables to clean zircon concentrates, one of the products obtained from heavy mineral sand deposits. Such concentrates are often contaminated with small amounts of fine silica, which can effectively be separated from the coarse zircon particles by air tabling. Some fine zircon may be lost in the tailings, and can be recovered by treatment on wet shaking tables.

Duplex concentrator

This machine was originally developed for the recovery of tin from low grade feeds, but has a wider application in the recovery of tungsten, tantalum, gold, chromite and platinum from fine feeds (Pearl et al., 1991). Two decks are used alternately to provide continuous feeding, the feed slurry being fed onto one of the decks, the lower density minerals running off into the discharge launder, while the heavy minerals remain on the deck. The deck is washed with water after a preset time, in order to remove the gangue minerals, after which the deck is tilted and the concentrate...
is washed off. One table is always concentrating, while the other is being washed or is discharging concentrates. The concentrator has a capacity of up to $5 \text{ t h}^{-1}$ of $-100 \mu\text{m}$ feed producing enrichment ratios of between 20 and 500, and is available with various sizes and numbers of decks.

**Mozley Laboratory Separator**

This flowing film device, which uses orbital shear, is now used in many mineral processing laboratories, and is designed to treat small samples (100 g) of ore, allowing a relatively unskilled operator to obtain information for a recovery grade curve within a very short time (Anon., 1980).

Cordingly et al. (1994) used a Mozley Laboratory Mineral Separator to obtain data for the optimisation of both gravity circuit performance and comminution requirements with respect to liberation size at the Wheal Jane tin concentrator.

**Centrifugal concentrators**

The Kelsey centrifugal jig was described above. Other non-jig centrifugal separators have also been developed over the last 20 years.

The *Knelson* concentrator is a compact batch centrifugal separator with an active fluidised bed to capture heavy minerals (Knelson, 1992; Knelson and Jones, 1994). Unit capacities range from laboratory scale to 150 tonnes of solids per hour. A centrifugal force up to 200 times the force of gravity acts on the particles, trapping denser particles in a series of rings located in the machine, while the gangue particles are flushed out. The Knelson concentrator can treat particles ranging in size from 10 microns to a maximum of 6 mm. It is generally used for feeds in which the dense component to be recovered is a very small fraction of the total material, less than $500 \text{ g/t}$ (0.05% by weight).

Feed slurry is then introduced through the stationary feed tube and into the concentrate cone. When the slurry reaches the bottom of the cone it is forced outward and up the cone wall under the influence of centrifugal force. Fluidisation water is introduced into the concentrate cone through a series of fluidisation holes. The slurry fills each ring to capacity to create a concentrating bed. Compaction of the bed is prevented by the fluidisation water. The flow of water that is injected into the rings is controlled to achieve optimum bed fluidisation. High specific gravity particles are captured and retained in the concentrating cone. When the concentrate cycle is complete, concentrates are flushed from the cone into the concentrate launder. Under normal operating conditions, this automated procedure is achieved in under 2 min in a secure environment.

The batch Knelson Concentrator has been widely applied in the recovery of gold, platinum, silver, mercury, and native copper.

The *Falcon SB* concentrator is another spinning fluidised bed batch concentrator (Figure 10.31). It is designed principally for the recovery of free gold in grinding circuit classifier underflows where a very small (<1%) mass pull to concentrate is required. The feed first flows up the sides of a cone-shaped bowl, where it stratifies according to particle density before passing over a concentrate bed fluidised from behind by back pressure water. The bed retains dense particles such as gold, and lighter gangue particles are washed over the top.
Periodically the feed is stopped and the concentrate rinsed out. Rinsing frequency, which is under automatic control, is determined from grade and recovery requirements. Accelerations of up to 300 times the force of gravity are used in the SB machine (McAlister and Armstrong, 1998). The SB5200 can treat nearly 400 t h\(^{-1}\) of feed solids, depending on the specific application.

The principle of the Mozley Multi-Gravity Separator (MGS) can be visualised as rolling the horizontal surface of a conventional shaking table into a drum, then rotating it so that many times the normal gravitational pull can be exerted on the mineral particles as they flow in the water layer across the surface. Figure 10.32 shows a cross-section of the pilot scale MGS. The Mine Scale MGS consists of two slightly tapered open-ended drums, mounted “back to back”, rotating at speeds variable between 90 and 150 rpm, enabling forces of between 5 and 15 g to be generated at the drum surfaces. A sinusoidal shake with an amplitude variable between 4 and 6 cps is superimposed on the motion of the drum, the shake imparted to one drum being balanced by the shake imparted to the other, thus balancing the whole machine. A scraper assembly is mounted within each drum on a separate concentric shaft, driven slightly faster than the drum but in the same direction. This scrapes the settled solids up the slope of the drum, during which time they are subjected to counter-current washing before being discharged as concentrate at the open, outer, narrow end of the drum. The lower density minerals, along with the majority of the wash water, flow downstream to discharge as tailings via slots at the inner end of each drum. The MGS has been used to effect improvements in final tin concentrate grade by replacing column flotation as the final stage of cleaning (Turner and Hallewell, 1993).

Gold ore concentrators

Although most of the gold from gold mines worldwide is recovered by dissolution in cyanide solution, a proportion of the coarse (+75 \(\mu\)m) gold is recovered by gravity separators. It has been argued that separate treatment of the coarse gold in this way constitutes a security risk and increases costs. Gravity concentration can remain an attractive option only if it can be implemented with very low capital and operating costs. A test designed to characterise the gravity recoverable gold (GRG) in the main circulating load of a grinding circuit has been described by Laplante et al. (1995).
The coarse gold must be concentrated in the grinding circuit to prevent it from being flattened into thin platelets (see Chapter 7). The rise of new gravity concentration devices such as the Knelson, Falcon, and IPJ have made this possible. In certain cases where sulphide minerals are the gravity gold carrier, flash flotation combined with modern gravity concentration technology provides the most effective gold recovery (Laplante and Dunne, 2002).

The deposits of the Yukon use sluice boxes to treat large tonnages of low grade gravels. Kelly et al. (1995) show that the simple sluice box can be a relatively efficient gravity concentrator provided they are correctly operated.

References


Dense medium separation (DMS)

Introduction

Dense medium separation (or heavy medium separation (HMS), or the sink-and-float process) is applied to the pre-concentration of minerals, i.e. the rejection of gangue prior to grinding for final liberation. It is also used in coal preparation to produce a commercially graded end-product, clean coal being separated from the heavier shale or high-ash coal.

In principle, it is the simplest of all gravity processes and has long been a standard laboratory method for separating minerals of different specific gravity. Heavy liquids of suitable density are used, so that those minerals lighter than the liquid float, while those denser than it sink (Figure 11.1).

Since most of the liquids used in the laboratory are expensive or toxic, the dense medium used in industrial separations is a thick suspension, or pulp, of some heavy solid in water, which behaves as a heavy liquid.

The process offers some advantages over other gravity processes. It has the ability to make sharp separations at any required density, with a high degree of efficiency even in the presence of high percentages of near-density material. The density of separation can be closely controlled, within a relative density of ±0.005 kg l⁻¹ and can be maintained, under normal conditions, for indefinite periods. The separating density can, however, be changed at will and fairly quickly, to meet varying requirements. The process is, however, rather expensive, mainly due to the ancillary equipment needed to clean the medium and the cost of the medium itself.

Dense medium separation is applicable to any ore in which, after a suitable degree of liberation by crushing, there is enough difference in specific gravity between the particles to separate those which will repay the cost of further treatment from those which will not. The process is most widely applied when the density difference occurs at a coarse particle size, as separation efficiency decreases with size due to the slower rate of settling of the particles. Particles should preferably be larger than about 4 mm in diameter, in which case separation can be effective on a difference in specific gravity of 0.1 or less.

Separation down to 500 μm, and less, in size can, however, be made by the use of centrifugal separators. Providing a density difference exists, there is no upper size limit except that determined by the ability of the plant to handle the material.

Dense medium separation is possible with ores in which the minerals are coarsely aggregated. If the values are finely disseminated throughout the host rock, then a suitable density difference between the crushed particles cannot be developed by coarse crushing.
Preconcentration is most often performed on metalliferous ores which are associated with relatively light country rock. Thus finely disseminated galena and sphalerite often occur with pyrite as replacement deposits in rocks such as limestone or dolomite. Similarly in some of the Cornish tin ores the cassiterite is found in lodes with some degree of banded structure in which it is associated with other high specific-gravity minerals such as the sulphides of iron, arsenic, and copper, as well as iron oxides. The lode fragments containing these minerals therefore have a greater density than the siliceous waste and will, therefore, allow early separation. Wall rock adjacent to the lode may likewise be disposed of and will in many cases form the majority of the waste generated, since the working of narrow lodes often involves the removal of waste rock from the walls to facilitate access. Problems may arise if the wall rock is mineralised with low-value, high-density minerals such as iron oxides and sulphides, a situation which is often encountered. A good example is a wolfram mine in France where the schist wall rock was found to contain pyrrhotite which raised its density to such an extent that dense medium separation to preconcentrate the wolfram lode material was impossible. As a result, the whole of the run-of-mine ore had to be comminuted in order to obtain the wolfram.

The dense medium

Liquids

Heavy liquids have wide use in the laboratory for the appraisal of gravity-separation techniques on ores. Heavy liquid testing may be performed to determine the feasibility of dense medium separation on a particular ore, and to determine the economic separating density, or it may be used to assess the efficiency of an existing dense medium circuit by carrying out tests on the sink and float products. The aim is to separate the ore samples into a series of fractions according to density, establishing the relationship between the high and the low specific gravity minerals.

Tetrabromoethane (TBE), having a specific gravity of 2.96, is commonly used and may be diluted with white spirit or carbon tetrachloride (sp. gr. 1.58) to give a range of densities below 2.96.

Bromoform (sp. gr. 2.89) may be mixed with carbon tetrachloride to give densities in the range 1.58–2.89. For densities of up to 3.3, diiodomethane is useful, diluted as required with triethyl orthophosphate. Aqueous solutions of sodium polytungstate have certain advantages over organic liquids, such as being virtually non-volatile, non-toxic and of lower viscosity, and densities of up to 3.1 can easily be achieved (Anon., 1984).

Clerici solution (thallium formate-thallium malonate solution) allows separation at densities up to 4.2 at 20°C, or 5.0 at 90°C. Separations of up to 18 kg l⁻¹ can be achieved by the use of magnetohydrostatics, i.e. the utilisation of the supplementary weighting force produced in a solution of a paramagnetic salt or ferrofluid when situated in a magnetic field gradient. This type of separation is applicable primarily to non-magnetic minerals with a lower limiting particle size of about 50 μm (Parsonage 1977; Domenico et al., 1994).

Many heavy liquids give off toxic fumes and must be used with adequate ventilation: the Clerici liquids are extremely poisonous and must be handled with extreme care. The use of pure liquids on a commercial scale has therefore not been found practicable, and industrial processes employ finely ground solids suspended in water.

Suspensions

Below a concentration of about 15% by volume, finely ground suspensions in water behave essentially as simple Newtonian fluids. Above this concentration, however, the suspension becomes non-Newtonian and a certain minimum stress, or yield stress, has to be applied before shear will occur and the movement of a particle can commence. Thus small particles, or those close to the medium density, are unable to overcome the rigidity offered by the medium before movement can be achieved. This can be overcome to some extent either by increasing the shearing forces on the particles, or by decreasing the apparent viscosity of the suspension. The shearing force may be increased by substituting centrifugal force for gravity. The viscous effect may be decreased by agitating the medium, which causes elements of liquid to be sheared relative to each other. In practice the medium is never static, as motion is imparted to it by paddles, air, etc., and also by the sinking material itself. All these factors, by reducing the yield stress, tend to bring the parting
or separating density as close as possible to the density of the medium in the bath.

In order to produce a stable suspension of sufficiently high density, with a reasonably low viscosity, it is necessary to use fine, high specific-gravity solid particles, agitation being necessary to maintain the suspension and to lower the apparent viscosity. The solids comprising the medium must be hard, with no tendency to slime, as degradation increases the apparent viscosity by increasing the surface area of the medium. The medium must be easily removed from the mineral surfaces by washing, and must be easily recoverable from the fine-ore particles washed from the surfaces. It must not be affected by the constituents of the ore and must resist chemical attack, such as corrosion.

Galena was initially used as the medium and, when pure, it can give a bath density of about 4.0. Above this level, ore separation is slowed down by the viscous resistance. Froth flotation, which is an expensive process, was used to clean the contaminated medium, but the main disadvantage is that galena is fairly soft and tends to slime easily, and it also has a tendency to oxidise, which impairs the flotation efficiency.

The most widely used medium for metalliferous ores is now ferrosilicon, whilst magnetite is used in coal preparation. Recovery of medium in both cases is by magnetic separation.

Magnetite (sp. gr. 5.1) is relatively cheap, and is used to maintain bath densities of up to 2.5 kg l\(^{-1}\). Ferrosilicon (sp. gr. 6.7–6.9) is an alloy of iron and silicon which should contain not less than 82% Fe and 15–16% Si (Collins et al., 1974). If the silicon content is less than 15%, the alloy will tend to corrode, while if it is more than 16% the magnetic susceptibility and density will be greatly reduced. Losses of ferrosilicon from a dense medium circuit vary widely, from as little as 0.1 to more than 2.5 kg/t of ore treated, the losses, apart from spillages, mainly occurring in magnetic separation and by the adhesion of medium to ore particles. Corrosion usually accounts for relatively small losses, and can be effectively prevented by maintaining the ferrosilicon in its passive state. This is normally achieved by atmospheric oxygen diffusing into the medium, or by the addition of small quantities of sodium nitrite (Stewart and Guerney, 1998).

Milled ferrosilicon is produced in a range of size distributions, from 30 to 95% – 45 μm, the finer grades being used for finer ores and centrifugal separators. The coarser, lower viscosity grades can achieve medium densities up to about 3.3. Atomised ferrosilicon consists of rounded particles which produce media of lower viscosity and can be used to achieve densities as high as 4 (Myburgh, 2002; Dunglison et al., 1999).

**Separating vessels**

Several types of separating vessel are in use, and these may be classified into gravitational ("static-baths") and centrifugal (dynamic) vessels. There is an extensive literature on the performance of these processes, and effective mathematical models are now being developed, which can be used for simulation purposes (Napier-Munn, 1991).

**Gravitational vessels**

Gravitational units comprise some form of vessel into which the feed and medium are introduced and the floats are removed by paddles, or merely by overflow. Removal of the sinks is the most difficult part of separator design. The aim is to discharge the sinks particles without removing sufficient of the medium to cause disturbing downward currents in the vessel.

The *Wemco cone separator* (Figure 11.2) is widely used for ore treatment, since it has a relatively high sinks capacity. The cone, which has a diameter of up to 6 m, accommodates feed particles of up to 10 cm in diameter, with capacities of up to 500 t h\(^{-1}\).

The feed is introduced on to the surface of the medium by free-fall, which allows it to plunge several centimetres into the medium. Gentle agitation by rakes mounted on the central shaft helps keep the medium in suspension. The float fraction simply overflows a weir, whilst the sinks are removed by pump or by external or internal air lift.

*Drum separators* (Figure 11.3) are built in several sizes, up to 4.3 m diameter by 6 m long, with maximum capacities of 450 t h\(^{-1}\), and can treat feeds of up to 30 cm in diameter. Separation is accomplished by the continuous removal of the sink product through the action of lifters fixed to the inside of the rotating drum. The lifters empty into the sink launder when they pass the horizontal position. The float product overflows a weir at
Figure 11.2 Wemco cone separator. (a) Single-gravity, two-product system with torque-flow-pump sink removal; (b) Single-gravity, two-product system with compressed-air sink removal

the opposite end of the drum from the feed chute. Longitudinal partitions separate the float surface from the sink-discharge action of the revolving lifters.

The comparatively shallow pool depth in the drum compared with the cone separator minimises settling out of the medium particles giving a uniform gravity throughout the drum.

Where single-stage dense-medium treatment is unable to produce the desired recovery, two-stage separation can be achieved in the two-compartment drum separator (Figure 11.4), which is, in effect, two drum separators mounted integrally and rotating together, one feeding the other. The lighter medium in the first compartment separates a pure float product. The sink product is lifted and
Figure 11.4 Two-compartment drum separator

conveyed into the second compartment where the middlings and the true sinks are separated.

Although drum separators have very large sinks capacities, and are inherently more suited to the treatment of metallic ores, where the sinks product is normally 60–80% of the feed, rather than to coal, where the sinks product is only 5–20% of the feed, they are very commonly used in the coal industry because of their simplicity, reliability, and relatively small maintenance needs. A mathematical model of the DM drum has been developed by Baguley and Napier-Munn (1996).

The Drewboy bath is widely used in the UK coal industry because of its high floats capacity (Figure 11.5).

The raw coal is fed into the separator at one end, and the floats are discharged from the opposite end by a star-wheel with suspended rubber, or chain straps, while the sinks are lifted out from the bottom of the bath by a radial-vaned wheel mounted on an inclined shaft. The medium is fed into the bath at two points – at the bottom of the vessel and with the raw coal – the proportion being controlled by valves.

The Norwalt washer was developed in South Africa, and most installations are to be found in that country. Raw coal is introduced into the centre of the annular separating vessel, which is provided with stirring arms (Figure 11.6).

The floats are carried round by the stirrers, and are discharged over a weir on the other side of the vessel, being carried out of the vessel by the medium flow. The discard sinks to the bottom of the vessel and is dragged along by scrapers attached to the bottom of the stirring arms, and is discharged via a hole in the bottom of the bath into
a sealed elevator, either of the wheel or bucket type, which continuously removes the sinks product.

The *Teska Bath*, developed in Germany, uses a rotating bucket wheel to remove coal reject.

### Centrifugal separators

Cyclone dense medium separators have now become widely used in the treatment of ores and coal. They provide a high centrifugal force and a low viscosity in the medium, enabling much finer separations to be achieved than in gravitational separators. Feed to these devices is typically de-sludged at about 0.5 mm, to avoid contamination of the medium with slimes, and to minimise medium consumption. A finer medium is required than with gravitational vessels, to avoid medium instability. In recent years work has been carried out in many parts of the world to extend the range of particle size treated by centrifugal separators, particularly those operating in coal preparation plants, where advantages to be gained are elimination of de-sludging screens and reduced froth flotation of the screen undersize, as well as more accurate separation of fine coal. Froth flotation has little effect on sulphur reduction, whereas pyrite can be removed, and oxidised coal treated by DMS. The work has shown that good separations can be achieved for coal particles as fine as 0.1 mm, but below this size separation efficiency is very low. Since a typical British coal can contain 10% material less than 0.1 mm, froth flotation must be retained to clean these finer fractions, although DMS with no de-sludging has been performed in the United States (Anon., 1985). Tests on a lead-zinc ore have shown that good separations can be achieved down to 0.16 mm using a centrifugal separator (Ruff, 1984). These and similar results elsewhere, together with the progress made in automatic control of medium consistency, add to the growing evidence that DMS can be considered for finer material than had been thought economical or practical until recently. As the energy requirement for grinding, flotation, and dewatering is often up to ten times that required for DMS, a steady increase of fines pre-concentration plants is likely.

By far the most widely used centrifugal DM separator is the cyclone (Figure 11.7) whose principle of operation is similar to that of the conventional hydrocyclone (Chapter 9). The commonest form of DM cyclone is that developed by the Dutch State Mines in the 1940s, which has an included
cone angle of 20°. Cyclones typically treat ores and coal in the range 0.5–40 mm. The largest cyclones now exceed 1 m in diameter and are capable of throughputs in coal preparation of over 250 t/h (Lee et al., 1995).

The ore or coal is suspended in the medium and introduced tangentially to the cyclone either via a pump or it is gravity-fed. Gravity feeding requires a taller and therefore more expensive building but achieves a more consistent flow and less pump wear and ore degradation. The dense material (reject in the case of coal, product in the case of iron ore) is centrifuged to the cyclone wall and exits at the apex. The light product “floats” to the flow around the axis and exits via the vortex finder.

Mathematical models of the DM cyclone for coal have been developed by King and Juckes (1988) and Wood et al. (1987) and for minerals by Scott and Napier-Munn (1992). A more general model has been reported by Dunglison and Napier-Munn (1997).

The Vorsyl separator (Figure 11.8) is used in many coal-preparation plants for the treatment of small coal sizes up to about 50 mm at feed rates of up to 120 t h⁻¹ (Shaw, 1984). The feed to the separator, consisting of de-slimed raw coal, together with the separating medium of magnetite, is introduced tangentially, or more recently by an involute entry, at the top of the separating chamber, under pressure. Material of specific gravity less than that of the medium passes into the clean coal outlet via the vortex finder, while the near gravity material and the heavier shale particles move to the wall of the vessel due to the centrifugal acceleration induced. The particles move in a spiral path down the chamber towards the base of the vessel where the drag caused by the proximity of the orifice plate reduces the tangential velocity and creates a strong inward flow towards the throat. This carries the shale, and near gravity material, through zones of high centrifugal force, where a final precise separation is achieved. The shale, and a proportion of the medium, discharge through the throat into the shallow shale chamber, which is provided with a tangential outlet, and is connected by a short duct to a second shallow chamber known as the vortexextractor. This is also a cylindrical vessel with a tangential inlet for the medium and reject and an axial outlet. An inward spiral flow to the outlet is induced, which dissipates the inlet pressure energy and permits the use of a large outlet nozzle without the passing of an excessive quantity of medium.

The LARCODEMS (Large Coal Dense Medium Separator) was developed to treat a wide size range of coal (−100 mm) at high capacity in one vessel (Shah, 1987). It has also been used in concentrating iron ore. The unit (Figure 11.9) consists of a cylindrical chamber which is inclined at approximately 30° to the horizontal. Feed medium at the required relative density is introduced under pressure, either by pump or static head, into the involute tangential
it could have a dramatic effect on the design and construction of future coal preparation plants.

The *Dyna Whirlpool* separator is similar to the LARCODEMS, and is used for treating fine coal, particularly in the Southern Hemisphere, as well as diamonds, fluorspar, tin, and lead-zinc ores, in the size range 0.5–30 mm (Wills and Lewis, 1980).

It consists of a cylinder of predetermined length (Figure 11.10), having identical tangential inlet and outlet sections at either end. The unit is operated in an inclined position and medium of the required density is pumped under pressure into the lower outlet. The rotating medium creates a vortex throughout the length of the unit and leaves via the upper tangential discharge and the lower vortex outlet tube. Raw feed entering the upper vortex tube is sluiced into the unit by a small quantity of medium and a rotational motion is quickly imparted by the open vortex. Float material passes down the vortex and does not contact the outer walls of the unit, thus greatly reducing wear. The floats are discharged from the lower vortex outlet tube. The heavy sink particles of the feed penetrate the rising medium towards the outer wall of the unit and are discharged with medium through the sink discharge pipe. Since the sinks discharge is close to the feed inlet, the sinks are removed from the unit almost immediately, again reducing wear.
considerably. Only near-gravity particles which are separated further along the unit actually come into contact with the main cylindrical body. The tangential sink discharge outlet is connected to a flexible sink hose and the height of this hose may be used to adjust back pressure to finely control the cut-point. The capacity of the separator can be as high as 100 t h⁻¹, and it has several advantages over the DSM cyclone. Apart from the reduced wear, which not only decreases maintenance costs but also maintains performance of the unit, operating costs are lower, since only the medium is pumped. The unit has a much higher sinks capacity and can accept large fluctuations in sink/float ratios (Hacioglu and Turner, 1985).

The Tri-Flo separator (Figure 11.11) can be regarded as two Dyna Whirlpool separators joined in series, and has been installed in a number of coal, metalliferous, and non-metallic ore treatment plants (Burton et al., 1991; Kitsikopoulos et al., 1991; Ferrara et al., 1994). Involute medium inlets and sink outlets are used, which produce less turbulence than tangential inlets.

The device can be operated with two media of differing densities in order to produce sink products of individual controllable densities. Two-stage treatment using a single medium density produces a float and two sinks products with only slightly different separation densities. With metalliferous ores, the second sink product can be regarded as a scavenging stage for the dense minerals, thus increasing their recovery. This second product may be recrushed, and, after de-sliming, returned for retreatment. Where the separator is used for washing coal, the second stage cleans the float to produce a higher grade product. Two stages of separation also increase the sharpness of separation.

DMS circuits

Although the separating vessel is the most important element of a DMS process, it is only one part of a relatively complex circuit. Other equipment is required to prepare the feed, and to recover, clean, and re-circulate the medium (Symonds and Malbon, 2002).

The feed to a dense medium circuit must be screened to remove fine ore, and slimes should be removed by washing, thus alleviating any tendency which such slime content may have for creating sharp increases in medium viscosity.

The greatest expense in any dense medium circuit is the provision for reclaiming and cleaning the medium which leaves the separator with the sink and float products. A typical circuit is shown in Figure 11.12.
The sink and float fractions pass on to separate vibrating drainage screens where more than 90% of the medium in the separator products is recovered and pumped back via a sump into the separating vessel. The products then pass under washing sprays where substantially complete removal of medium and adhering fines is accomplished. The finished float and sink products are discharged from the screens for disposal or further treatment.

The undersize products from the washing screens, consisting of medium, wash water, and fines, are too dilute and contaminated to be returned directly as medium to the separating vessel. They are treated individually as shown, or together, by magnetic separation, to recover the magnetic ferrosilicon or magnetite from the non-magnetic fines. Reclaimed, cleaned medium is thickened to the required density by a centrifugal or spiral density, which continuously returns it to the DMS circuit. The densified medium discharge passes through a demagnetising coil to ensure a non-flocculated, uniform suspension in the separating vessel.

Most large DMS plants include automatic control of the feed medium density. This is done by densifying sufficient medium to cause the medium density to rise, measuring the feed density with a gamma attenuation gauge, and using the signal to adjust the amount of water added to the medium to return it to the correct density.

The major costs in DMS are power (for pumping) and medium consumption. Medium losses can account for 10–35% of total costs. They are principally due to adhesion to products and losses from the magnetic separators, though the proportions will depend on the size and porosity of the ore, the characteristics of the medium solids, and the plant design (Napier-Munn et al., 1995). Losses increase for fine or porous ore, fine media, and high operating densities.

Correct sizing and selection of equipment, together with correct choice of design parameters such as rinsing water volumes, are essential. As effluent water always contains some entrained medium, the more of this that can be recycled back to the plant the better (Dardis, 1987). Careful attention should also be paid to the quality of the medium used, Williams and Kelsall (1992) having shown that certain ferrosilicon powders are more prone to mechanical degradation and corrosion than others.

Medium rheology is critical to efficient operation of dense medium systems (Napier-Munn, 1990), although the effects of viscosity are difficult to quantify (Reeves, 1990; Dunglison et al., 1999). Management of viscosity includes selecting the correct medium specifications, minimising operating density, and minimising the content of clays and other fine contamination (Napier-Munn and Scott, 1990). If the amount of fines in the circuit reaches a high proportion due, say, to inefficient screening of the feed, it may be necessary to divert an increased amount of medium into the cleaning circuit. Many circuits have such a provision, allowing medium from the draining screen to be diverted into the washing screen undersize sump.

Typical dense medium separations

The most important use of DMS is in coal preparation where a relatively simple separation removes the low-ash coal from the heavier high-ash discard and associated shales and sand-stones.

DMS is preferred to the cheaper Baum jig method of separation when washing coals with a relatively large proportion of middlings, or near-density material, since the separating density can be controlled to much closer limits.

British coals, in general, are relatively easy to wash, and jigs are used in many cases. Where DMS is preferred, Drum and Drewboys separators are most widely used for the coarser fractions, with DSM cyclones and Vorsyl separators being preferred for the fines. DMS is essential with most Southern Hemisphere coals, where a high middlings fraction is present. This is especially so with the large, low-grade coal deposits found in the former South African Transvaal province. Drums and Norwalt baths are the most common separators utilised to wash such coals, with DSM cyclones and Dyna Whirlpools being used to treat the finer fractions.

At Amcoal’s Landau Colliery in the Transvaal, a two-density operation is carried out in order to produce two saleable products. After preliminary screening of the run-of-mine coal, the coarse (+7 mm) fraction is washed in Norwalt bath separators, utilising magnetite as the medium to give a separating density of 1.6. The sinks product from
this operation, consisting predominantly of sand and shales, is discarded, and the floats product is routed to Norwalt baths operating at a lower density of 1.4. This separation stage produces a low-ash floats product, containing about 7.5% ash, which is used for metallurgical coke production, and a sinks product, which is the process middlings, containing about 15% ash, which is used as power-station fuel. The fine (0.5–7 mm) fraction is treated in a similar two-stage manner utilising Dyna Whirlpool separators.

In metalliferous mining, DMS is used in the preconcentration of lead–zinc ores, in which the disseminated sulphide minerals often associate together as replacement bandings in the light country rock, such that marked specific gravity differences between particles crushed to fairly coarse sizes can be exploited.

A dense medium plant was incorporated into the lead–zinc circuit at Mount Isa Mines Ltd., Australia, in 1982 in order to increase the plant throughput by 50%. The ore, containing roughly 6.5% lead, 6.5% zinc, and 200 ppm silver, consists of galena, sphalerite, pyrite, and other sulphides finely disseminated in distinct bands in quartz and dolomite (Figure 11.13). Liberation of the ore into particles which are either sulphide-rich or predominantly gangue begins at around –50 mm, and becomes substantial below 18 mm.

The plant treats about 800 t h\(^{-1}\) of material, in the size range 1.7–13 mm by DSM cyclones, at a separating density of 3.05 kg l\(^{-1}\), to reject 30–35% of the run-of-mine ore as tailings, with 96–97% recoveries of lead, zinc, and silver to the pre-concentrate. The pre-concentrate has a 25% lower Bond Work Index, and is less abrasive because the lower specific gravity siliceous material mostly reports to the rejects. The rejects are used as a cheap source of fill for underground operations. The plant is extensively instrumented, the process control strategy being described elsewhere (Munro et al., 1982).

DMS is also used to pre-concentrate tin and tungsten ores, and non-metallic ores such as fluorite, barite, etc. It has a very important use in the pre-concentration of diamond ores, prior to recovery of the diamonds by electronic sorting (Chapter 14) or grease-tabling (Chaston and Napier-Munn, 1974; Rylatt and Popplewell, 1999). Diamonds are the lowest grade of all ores mined, and concentration ratios of several million to one must be achieved. DMS produces an initial enrichment of the ore in the order of 100–1000 to 1 by making use of the fact that diamonds have a fairly high specific gravity (3.5), and are relatively easily liberated from the ore, since they are loosely held in the parent rock. Gravity and centrifugal separators are utilised, with ferrosilicon as the medium, and separating densities of between 2.6 and 3.0 are used. Clays in the ore sometimes present a problem by increasing the medium viscosity, thus reducing separating efficiency and the recovery of diamonds to the sinks.

DMS is also used for upgrading low grade iron ores for blast furnace feed. Both gravity and

---

**Figure 11.13** Mount Isa ore. Bands of sulphide minerals in carbonaceous host rock
centrifugal separators are used, and in some cases the medium density can exceed 4 (Myburgh, 2002).

**Laboratory heavy liquid tests**

Laboratory testing may be performed on ores in order to assess the suitability of dense medium separation and other gravity methods, and to determine the economic separating density.

Liquids covering a range of densities in incremental steps are prepared, and the representative sample of crushed ore is introduced into the liquid of highest density. The floats product is removed and washed and placed in the liquid of next lower density, whose float product is then transferred to the next lower density and so on. The sinks product is finally drained, washed, and dried, and then weighed, together with the final floats product, to give the density distribution of the sample by weight (Figure 11.14).

Care should be taken when evaluating ores of fine particle size that sufficient time is given for the particles to settle into the appropriate fraction. Centrifuging is often carried out on fine materials to reduce the settling time, but this should be done with care, as there is a tendency for the floats to become entrained in the sinks fraction. Unsatisfactory results are often obtained with porous materials, such as magnesite ores, due to the entrainment of liquid in the pores, which changes the apparent density of the particles.

After assaying the fractions for metal content, the distribution of material and metal in the density fractions of the sample can be tabulated. Table 11.1 shows such a distribution from tests performed on a tin ore. The computations are easily accomplished in a spreadsheet. It can be seen from columns 3 and 6 of the table that if a separation density of 2.75 was chosen, then 68.48% of the material, being lighter than 2.75, would be discarded as a float product, and only 3.81% of the tin would be lost in this fraction. Similarly, 96.19% of the tin would be recovered into the sink product, which accounts for 31.52% of the original total feed weight.

The choice of optimum separating density must be made on economic grounds. In the example shown in Table 11.1, the economic impact of rejecting 68.48% of the feed to HMS on downstream performance must be assessed. The smaller throughput will lower grinding and concentration operating costs, the impact on grinding energy and steel costs often being particularly high. Against these savings, the cost of operating the DMS plant and the effect of losing 3.81% of the run-of-mine tin to floats must be considered. The amount of recoverable tin in this fraction has to be estimated, together with its subsequent loss in smelter revenue. If this loss is lower than the saving in overall milling costs, then DMS is economic. The optimum density is that which maximises the difference between overall reduction in milling costs per tonne of run-of-mine ore and loss in smelter revenue. Schena et al. (1990) have analysed the economic choice of separating density and have developed computer software for the evaluation.

Heavy liquid tests are important in coal preparation in order to determine the required density of separation and the expected yield of coal of the required ash content. The “ash” content refers to the amount of incombustible material in the coal. Since coal is lighter than the contained minerals, the higher the density of separation the higher is the yield:

\[
\text{yield} = \frac{\text{weight of coal floats product} \times 100\%}{\text{total feed weight}}
\]

but the higher is the ash content. The ash content of each density fraction from heavy liquid testing is determined by taking about 1 g of the fraction, placing it in a cold well-ventilated furnace, and

**Figure 11.14** Heavy liquid testing
slowly raising the temperature to 815°C, maintaining the sample at this temperature until constant weight is obtained. The residue is cooled and then weighed. The ash content is the mass of ash expressed as a percentage of the initial sample weight taken.

Table 11.2 shows the results of heavy liquid tests performed on a coal sample. The coal was separated into the density fractions shown in column 1, and the weight fractions and ash contents are tabulated in columns 2 and 3 respectively. The weight per cent of each product is multiplied by the ash content to give the ash product (column 4).

The total floats and sinks products at the various separating densities shown in column 5 are tabulated in columns 6–11. To obtain the cumulative per cent for each gravity fraction, columns 2 and 4 are cumulated from top to bottom to give columns 6 and 7 respectively. Column 7 is then divided by column 6 to obtain the cumulative per cent ash (column 8). Cumulative sink ash is obtained in essentially the same manner, except that columns

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<th>(3)</th>
<th>(4)</th>
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<td>Distribution (% Sn)</td>
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<td>Sp. gr. fraction</td>
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<td>Ash %</td>
<td>Ash product</td>
<td>Separating density</td>
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<td>Cumulative sink (Discard)</td>
<td>Wt %</td>
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Figure 11.15  Typical coal washability curves

2 and 4 are cumulated from bottom to top to give columns 9 and 10 respectively.

The results of Table 11.2 are plotted in Figure 11.15 as typical washability curves.

Suppose an ash content of 12% is required in the coal product. It can be seen from the washability curves that such a coal would be produced at a yield of 55% (cumulative per cent floats), and the required density of separation is 1.465.

The difficulty of the separation in terms of operational control is dependent mainly on the amount of material present in the feed which is close to the required density of separation. For instance, if the feed were composed entirely of pure coal at a sp. gr. of 1.3 and shale at a density of 2.7, then the separation would be easily carried out over a wide range of operating densities. If, however, the feed consists of appreciable middlings, and much material is present very close to the chosen separating density, then only a small variation in this density will seriously affect the yield and ash content of the product.

The amount of near-gravity material present is sometimes regarded as being the weight of material in the range ±0.1 or ±0.05 kg l⁻¹ of the separating density, and separations involving feeds with less than about 7% of ±0.1 near-gravity material are regarded by coal preparation engineers as being fairly easy to control. Such separations are often performed in Baum jigs, as these are cheaper than dense medium plants, which require expensive media-cleaning facilities, and no feed preparation, i.e. removal of the fine particles by screening, is required. However, the density of separation in jigs is not as easy to control to fine limits, as it is in dense medium baths, and for near-gravity material much above 7%, dense medium separation is preferred.

Heavy liquid tests can be used to evaluate any gravity separation process on any ore, and Table 11.3 can be used to indicate the type of separator which could effect the separation in practice (Mills, 1978).

Table 11.3 takes no account of the particle size of the material and experience is therefore required in its application to heavy liquid results, although some idea of the effective particle size range of gravity separators can be gained from Figure 11.8. The throughput of the plant must also be taken into account with respect to the type of separator chosen. For instance, if a throughput of only a few tonnes per hour is envisaged, there would be little point in
installing Reichert cones, which are high-capacity units, operating most effectively at about 70 t h\(^{-1}\).

### Efficiency of dense medium separation

Laboratory testing assumes perfect separation and, in such batch tests, conditions are indeed close to the ideal, as sufficient time can be taken to allow complete separation to take place.

In a continuous production process, however, conditions are usually far from ideal and particles can be misplaced to the wrong product for a variety of reasons. The dominant effect is that of the density distribution of the feed. Very dense or very light particles will settle through the medium and report to the appropriate product quickly, but particles of density close to that of the medium will move more slowly and may not reach the right product in the time available for the separation. In the limit, particles of density the same as, or very close to, that of the medium will follow the medium and divide in much the same proportion.

Other factors also play a role in determining the efficiency of separation. Fine particles generally separate less efficiently than coarse, again because of their slower settling rates. The properties of the medium, the design and condition of the separating vessel, and the feed conditions, particularly feed rate, will all influence the separation.

The efficiency of separation can be represented by the slope of a *Partition* or *Tromp* curve, first introduced by K.F. Tromp (1937). It describes the separating efficiency for the separator whatever the quality of the feed and can be used for estimation of performance and comparison between separators.

The partition curve relates the *partition coefficient* or *partition number*, i.e. the percentage of the feed material of a particular specific gravity which reports to either the sinks product (generally used for minerals) or the floats product (generally used for coal), to specific gravity (Figure 11.16). It is exactly analogous to the classification efficiency curve, in which the partition coefficient is plotted against size rather than specific gravity.

![Figure 11.16 Partition or Tromp curve](Image)

**Figure 11.16** Partition or Tromp curve

The ideal partition curve reflects a perfect separation in which all particles having a density higher than the separating density report to sinks, and those lighter report to floats. There is no misplaced material.

The partition curve for a real separation shows that efficiency is highest for particles of density far from the operating density and decreases for particles approaching the operating density.
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The area between the two curves is called the “error area” and is a measure of the degree of misplacement of particles to the wrong product.

Many partition curves give a reasonable straight-line relationship between the distribution of 25 and 75%, and the slope of the line between these distributions is used to show the efficiency of the process.

The probable error of separation or the Ecart probable (Ep) is defined as half the difference between the density where 75% is recovered to sinks and that at which 25% is recovered to sinks, i.e. from Figure 11.16,

\[ Ep = \frac{A - B}{2} \]

The density at which 50% of the particles report to sinks is shown as the effective density of separation, which may not be exactly the same as the medium density, particularly for centrifugal separators, in which the separating density is generally higher than the medium density.

The lower the Ep, the nearer to vertical is the line between 25 and 75% and the more efficient is the separation. An ideal separation has a vertical line with an Ep = 0 whereas in practice the Ep usually lies in the range 0.01–0.10.

The Ep is not commonly used as a method of assessing the efficiency of separation in units such as tables, spirals, cones, etc., due to the many operating variables (wash water, table slope, speed, etc.) which can affect the separation efficiency. It is, however, ideally suited to the relatively simple and reproducible DMS process. However care should be taken in its application, as it does not reflect performance at the tails of the curve, which can be important.

Construction of partition curves

The partition curve for an operating dense medium vessel can be determined by sampling the sink and float products and performing heavy liquid tests to determine the amount of material in each density fraction. The range of liquid densities applied must envelope the working density of the dense medium unit. The results of heavy liquid tests on samples of floats and sinks from a vessel separating coal (floats) from shale (sinks) are shown in Table 11.4. The calculations are easily performed in a spreadsheet.

Columns 1 and 2 are the results of laboratory tests on the float and sink products and columns 3 and 4 relate these results to the total distribution of the feed material to floats and sinks which must be determined by weighing the products over a period of time. The weight fraction in columns 3 and 4 can be added together to produce the reconstituted feed weight distribution in each density fraction (column 5). Column 6 gives the nominal specific gravity of each density range, i.e. material in the density range 1.30–1.40 is assumed to have a specific gravity lying midway between these densities – 1.35.

The partition coefficient (column 7) is the percentage of feed material of a certain nominal specific gravity which reports to sinks, i.e.

\[ \frac{\text{column 4}}{\text{column 5}} \times 100\% \]

It can also be determined by applying the two-product formula (Chapter 3) to the density distributions of feed, sinks, and floats, if all three are available and accurate.

The partition curve can then be constructed by plotting the partition coefficient against the nominal specific gravity, from which the probable error of separation of the vessel can be determined.

An alternative, rapid, method of determining the partition curve of a separator is to utilise density tracers. Specially developed colour-coded plastic tracers can be fed to the process, the partitioned products being collected and hand sorted by density (colour). It is then a simple matter to construct the partition curve directly by noting the proportion of each density of tracer reporting to either the sink or float product. Application of tracer methods has shown that considerable uncertainties can exist in experimentally determined Tromp curves unless an adequate number of tracers is used, and Napier-Munn (1985) presents graphs that facilitate the selection of sample size and the calculation of confidence limits. A system in operation in a US coal preparation plant uses sensitive metal detectors that automatically spot and count the number of different types of tracers passing through a stream. The tracers, of various size and density, are selectively fed into the feed stream of a Baum jig by a computer-controlled dispensing system, allowing the jig’s real-time performance to be assessed (Chironis, 1987).
<table>
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<th>Specific gravity fraction</th>
<th>Floats analysis (wt %)</th>
<th>Sinks analysis (wt %)</th>
<th>Floats % of feed</th>
<th>Sinks % of feed</th>
<th>Reconstituted feed (%)</th>
<th>Nominal sp. gr.</th>
<th>Partition coefficient</th>
</tr>
</thead>
<tbody>
<tr>
<td>-1.30</td>
<td>83.34</td>
<td>18.15</td>
<td>68.83</td>
<td>3.15</td>
<td>71.98</td>
<td>4.39</td>
<td></td>
</tr>
<tr>
<td>1.30–1.40</td>
<td>10.50</td>
<td>10.82</td>
<td>8.67</td>
<td>1.89</td>
<td>10.56</td>
<td>1.35</td>
<td>17.80</td>
</tr>
<tr>
<td>1.40–1.50</td>
<td>3.35</td>
<td>9.64</td>
<td>2.77</td>
<td>1.68</td>
<td>4.45</td>
<td>1.45</td>
<td>37.75</td>
</tr>
<tr>
<td>1.50–1.60</td>
<td>1.79</td>
<td>13.33</td>
<td>1.48</td>
<td>2.32</td>
<td>3.80</td>
<td>1.55</td>
<td>61.05</td>
</tr>
<tr>
<td>1.60–1.70</td>
<td>0.30</td>
<td>8.37</td>
<td>0.25</td>
<td>1.46</td>
<td>1.71</td>
<td>1.65</td>
<td>85.38</td>
</tr>
<tr>
<td>1.70–1.80</td>
<td>0.16</td>
<td>5.85</td>
<td>0.13</td>
<td>1.02</td>
<td>1.15</td>
<td>1.75</td>
<td>88.70</td>
</tr>
<tr>
<td>1.80–1.90</td>
<td>0.07</td>
<td>5.05</td>
<td>0.06</td>
<td>0.88</td>
<td>0.94</td>
<td>1.85</td>
<td>93.62</td>
</tr>
<tr>
<td>1.90–2.00</td>
<td>0.07</td>
<td>4.34</td>
<td>0.06</td>
<td>0.75</td>
<td>0.81</td>
<td>1.95</td>
<td>92.68</td>
</tr>
<tr>
<td>+2.00</td>
<td>0.42</td>
<td>24.45</td>
<td>0.35</td>
<td>4.25</td>
<td>4.60</td>
<td>92.39</td>
<td></td>
</tr>
<tr>
<td>Totals</td>
<td>100.00</td>
<td>100.00</td>
<td>82.60</td>
<td>17.40</td>
<td>100.00</td>
<td>92.39</td>
<td></td>
</tr>
</tbody>
</table>

Partition curves can be used to predict the products that would be obtained if the feed or separation gravity were changed. The curves are specific to the vessel for which they were established and are not affected by the type of material fed to it, provided:

(a) The feed size range is the same – efficiency generally decreases with decrease in size; Figure 11.17 shows typical efficiencies of bath (drum, cone, etc.) and centrifugal (cyclone, DWP, etc.) separators versus particle size. It can clearly be seen that, in general, below about 10 mm, centrifugal separators are better than baths;

(b) the separating gravity is in approximately the same range – the higher the effective separating density the greater the probable error, due to the increased medium viscosity. It has in fact been shown that the $Ep$ is directly proportional to the separating density, all other factors being the same (Gottfried, 1978);

(c) the feed rate is the same.

The partition curve for a vessel can be used to determine the amount of misplaced material which will report to the products for any particular feed material. For example, the distribution of the products from the tin ore, which was evaluated by heavy liquid tests (Table 11.1) can be determined for treatment in an operating separator. Figure 11.18 shows a partition curve for a separator having an $Ep$ of 0.07.

The curve can be shifted slightly along the abscissa until the effective density of separation corresponds to the laboratory evaluated separating density of 2.75. The distribution of material to sinks and floats can now be evaluated, e.g. at a nominal specific gravity of 2.725, 44.0% of the material reports to the sinks and 56.0% to the floats.

The performance is evaluated in Table 11.5. Columns 1, 2, and 3 show the results of the heavy liquid tests, which were tabulated in Table 11.1. Columns 4 and 5 are the distributions to sinks.

![Figure 11.17 Effect of particle size on efficiency of dense media separators](image)
and floats respectively, obtained from the partition curve. Column 6 = column 1 \times column 4, and column 9 = column 1 \times column 5. The assay of each fraction is assumed to be the same whether or not the material reports to sinks or floats (columns 2, 7, and 10). Columns 8 and 11 are then calculated as the amount of tin reporting to sinks and floats in each fraction (columns 6 \times 7 and 9 \times 10) as a percentage of the total tin in the feed (sum of columns 1 \times 2).

The total distribution of the feed to sinks is the sum of all the fractions in column 6, i.e. 40.26%, while the recovery of tin into the sinks is the sum of the fractions in column 8, i.e. 95.29%. This compares with a distribution of 31.52% and a recovery of 96.19% of tin in the ideal separation.

This method of evaluating the performance of a separator on a particular feed is tedious and is ideal for a spreadsheet, providing that the partition numbers for each density fraction are known. These can be represented by a suitable mathematical function. There is a large literature on the selection and application of such functions. Some are arbitrary, and others have some theoretical or heuristic justification. The key feature of the partition curve is its S-shaped character. In this it bears a passing resemblance to a number of probability distribution functions and indeed the curve can be thought of as a statistical description of the DMS process, describing the probability with which a particle of given density (and other characteristics) reports to the sink product. Tromp himself recognised this in suggesting that the amount of misplaced material relative to a suitably transformed density scale was normally distributed, and Jowett (1986) showed that a partition curve for a process controlled by simple probability factors should have a normal distribution form.

However, many real partition curves do not behave ideally as does the one illustrated in Figure 11.16. In particular they are not asymptotic to 0 and 100% but exhibit evidence of short-circuit flow to one or both products. Stratford

Table 11.5 Tin ore evaluation

<table>
<thead>
<tr>
<th>Specific gravity fraction</th>
<th>Nominal sp. gr.</th>
<th>Feed</th>
<th>Distribution to (%)</th>
<th>Sinks</th>
<th>Floats</th>
<th>(9)</th>
<th>(10)</th>
<th>(11)</th>
</tr>
</thead>
<tbody>
<tr>
<td>(1) Wt %</td>
<td>(2) % Sn</td>
<td>(3) % Dist.</td>
<td>(4) Sinks</td>
<td>(5) Floats</td>
<td>(6) Wt %</td>
<td>(7) % Sn</td>
<td>(8) % (feed) distribution</td>
<td>(9) Wt %</td>
</tr>
<tr>
<td>-2.55</td>
<td>-</td>
<td>1.57</td>
<td>0.003</td>
<td>0.004</td>
<td>0.00</td>
<td>100.00</td>
<td>0.00</td>
<td>0.003</td>
</tr>
<tr>
<td>2.55-2.60</td>
<td>2.575</td>
<td>9.22</td>
<td>0.04</td>
<td>0.33</td>
<td>6.0</td>
<td>94.00</td>
<td>0.55</td>
<td>0.04</td>
</tr>
<tr>
<td>2.60-2.65</td>
<td>2.625</td>
<td>26.11</td>
<td>0.04</td>
<td>0.93</td>
<td>13.5</td>
<td>86.5</td>
<td>3.52</td>
<td>0.04</td>
</tr>
<tr>
<td>2.65-2.70</td>
<td>2.675</td>
<td>19.67</td>
<td>0.04</td>
<td>0.70</td>
<td>27.0</td>
<td>73.0</td>
<td>5.31</td>
<td>0.04</td>
</tr>
<tr>
<td>2.70-2.75</td>
<td>2.725</td>
<td>11.91</td>
<td>0.17</td>
<td>1.81</td>
<td>44.0</td>
<td>56.0</td>
<td>5.24</td>
<td>0.17</td>
</tr>
<tr>
<td>2.75-2.80</td>
<td>2.775</td>
<td>10.92</td>
<td>0.34</td>
<td>3.32</td>
<td>63.0</td>
<td>37.0</td>
<td>6.88</td>
<td>0.34</td>
</tr>
<tr>
<td>2.80-2.85</td>
<td>2.825</td>
<td>7.87</td>
<td>0.37</td>
<td>2.60</td>
<td>79.5</td>
<td>20.5</td>
<td>6.26</td>
<td>0.37</td>
</tr>
<tr>
<td>2.85-2.90</td>
<td>2.875</td>
<td>2.55</td>
<td>1.30</td>
<td>2.96</td>
<td>90.5</td>
<td>9.5</td>
<td>2.32</td>
<td>1.30</td>
</tr>
<tr>
<td>+2.90</td>
<td>-</td>
<td>10.18</td>
<td>9.60</td>
<td>87.34</td>
<td>100.00</td>
<td>0.00</td>
<td>10.18</td>
<td>9.60</td>
</tr>
</tbody>
</table>

Totals | 100.00 | 1.12 | 100.00 | 40.26 | 2.65 | 95.29 | 59.74 | 0.09 | 4.71 |
and Napier-Munn (1986) identified four attributes required of a suitable function to represent the partition curve:

1. It should have natural asymptotes, preferably described by separate parameters.

2. It should be capable of exhibiting asymmetry about the separating density; i.e., the differentiated form of the function should be capable of describing skewed distributions.

3. It should be mathematically continuous.

4. Its parameters should be capable of estimation by accessible methods.

A two-parameter function asymptotic to 0 and 100% is the Rosin-Rammler function, originally developed to describe size distributions (Tarjan, 1974):

\[ P_i = 100 - 100 \exp \left[ -\left( \frac{\rho_i}{\rho} \right)^m \right] \]  

(11.1)

In this form, \( P_i \) is the partition number (feed reporting to sinks, %), \( \rho_i \) is the mean density of density fraction \( i \), and \( a \) and \( m \) are the parameters of the function; \( m \) describes the width of the curve (high values of \( m \) indicating more efficient separations). Partition curve functions are normally expressed in terms of the normalised density, \( \frac{\rho_i}{\rho_{50}} \), where \( \rho_{50} \) is the separating density. The normalised curve is generally independent of cut-point and medium density, but dependent on particle size. Inserting this into Equation 11.1, and noting that \( P = 50 \) for \( \rho = \rho_{50} \) (\( \rho / \rho_{50} = 1 \)), gives:

\[ P_i = 100 - 100 \exp \left[ -\ln 2 \left( \frac{\rho_i}{\rho_{50}} \right)^m \right] \]  

(11.2)

One of the advantages of Equation 11.2 is that it can be linearised so that simple linear regression can be used to estimate \( m \) and \( \rho_{50} \) from experimental data:

\[ \ln \left( \frac{100 - P_i}{\ln 2} \right) = m \ln \rho_i - m \ln \rho_{50} \]  

(11.3)

Gottfried (1978) proposed a related function, the Weibull function, with additional parameters to account for the fact that the curves do not always reach the 0 and 100% asymptotes due to short-circuit flow:

\[ P_i = 100 - 100 \left[ f_0 + c \exp \left( -\left( \frac{\rho_i}{\rho_{50} - x_0} \right)^a \right) \right] \]  

(11.4)

The six parameters of the function \((c, f_0, \rho_{50}, x_0, a, and b)\) are not independent, so by the argument of Equation 11.2 \( x_0 \) can be expressed as:

\[ x_0 = 1 - \left[ b \ln \left( \frac{c}{0.5 - f_0} \right) \right]^{1/a} \]  

(11.5)

In this version of the function, representing percent of feed to sinks, \( f_0 \) is the proportion of high density material misplaced to floats, and \( 1 - (c + f_0) \) is the proportion of low density material misplaced to sinks, so that \( c + f_0 \leq 1 \). The curve therefore varies from a minimum of 100[1 - (c + f_0)] to a maximum of 100(1 - f_0).

The parameters of Equation 11.4 have to be determined by non-linear estimation. Non-linear optimisation routines are available in spreadsheets. First approximations of \( c, f_0 \) and \( \rho_{50} \) can be obtained from the curve itself.

King and Juckes (1988) utilised Whitman’s classification function with two additional parameters to describe the short-circuit flows or by-pass:

\[ P_i = \beta + (1 - \alpha - \beta) \left( \frac{e^{\beta \rho_i/\rho_{50}} - 1}{e^{\beta \rho_i/\rho_{50}} + e^{\beta}} \right) \]  

(11.6)

Here, for \( P = \) proportion to underflow, \( \alpha \) is the fraction of feed which short-circuits to overflow and \( \beta \) is the fraction of feed which short-circuits to underflow; \( b \) is an efficiency parameter, with high values of \( b \) indicating high efficiency. Again the function is non-linear in the parameters.

The Ep (Figure 11.16) can be predicted from these functions by substitution for \( \rho_{25} \) and \( \rho_{75} \). Scott and Napier-Munn (1992) showed that for efficient separations (low Eps) without short-circuiting, the partition curve could be approximated by:

\[ P_i = \frac{1}{1 + \exp \left[ \ln 3(\rho_{50} - \rho_i) \right]} \]  

(11.7)

**Organic efficiency**

The term *organic efficiency* is often used to express the efficiency of coal preparation plants. It is defined as the ratio (normally expressed as a percentage) between the actual yield of a desired product and the theoretical possible yield at the same ash content.

For instance, if the coal, whose washability data is plotted in Figure 11.15, produced an operating
yield of 51% at an ash content of 12%, then, since the theoretical yield at this ash content is 55%, the organic efficiency is equal to 51/55, or 92.7%.

Organic efficiency cannot be used to compare the efficiencies of different plants, as it is a dependent criterion, and is much influenced by the washability of the coal. It is possible, for example, to obtain a high organic efficiency on a coal containing little near-gravity material, even when the separating efficiency, as measured by partition data, is quite poor.

References


Anon. (1985). Feeding to zero: Island Creek’s experience in Kentucky, Coal Age, 90(Jan.), 66.


Froth flotation

Introduction

Flotation is undoubtedly the most important and versatile mineral processing technique, and both its use and application are continually being expanded to treat greater tonnages and to cover new areas. Originally patented in 1906, flotation has permitted the mining of low-grade and complex ore bodies which would have otherwise been regarded as uneconomic. In earlier practice the tailings of many gravity plants were of a higher grade than the ore treated in many modern flotation plants.

Flotation is a selective process and can be used to achieve specific separations from complex ores such as lead–zinc, copper–zinc, etc. Initially developed to treat the sulphides of copper, lead, and zinc, the field of flotation has now expanded to include platinum, nickel, and gold-hosting sulphides, and oxides, such as hematite and cassiterite, oxidised minerals, such as malachite and cerussite, and non-metallic ores, such as fluorite, phosphates, and fine coal.

Principles of flotation

Flotation is a physico-chemical separation process that utilises the difference in surface properties of the valuable minerals and the unwanted gangue minerals. The theory of froth flotation is complex, involving three phases (solids, water, and froth) with many subprocesses and interactions, and is not completely understood. The subject has been reviewed comprehensively by a number of authors (Sutherland and Wark, 1955; Glembotskii et al., 1972; King, 1982; Leja, 1982; Ives, 1984; Jones and Woodcock, 1984; Fuerstenau et al., 1985; Crozier, 1992; Laskowski and Poling, 1995; Harris et al., 2002; Johnson and Munro, 2002; Rao, 2004), and will only be dealt with briefly here.

The process of material being recovered by flotation from the pulp comprises three mechanisms:

1. Selective attachment to air bubbles (or “true flotation”).
2. Entrainment in the water which passes through the froth.
3. Physical entrapment between particles in the froth attached to air bubbles (often referred to as “aggregation”).

The attachment of valuable minerals to air bubbles is the most important mechanism and represents the majority of particles that are recovered to the concentrate. Although true flotation is the dominant mechanism for the recovery of valuable mineral, the separation efficiency between the valuable mineral and gangue is also dependent on the degree of entrainment and physical entrapment. Unlike true flotation, which is chemically selective to the mineral surface properties, both gangue and valuable minerals alike can be recovered by entrainment and entrapment. Drainage of these minerals occurs in the froth phase and controlling the stability of this phase is important to achieve an adequate separation. In industrial flotation plant practice, entrainment of unwanted gangue can be common and hence a single flotation stage is uncommon. Often several stages of flotation (called “circuits”) are required to reach an economically acceptable quality of valuable mineral in the final product.

True flotation utilises the differences in physico-chemical surface properties of particles of various minerals. After treatment with reagents, such differences in surface properties between the minerals within the flotation pulp become apparent and, for flotation to take place, an air bubble must be able to attach itself to a particle, and lift it to the water surface. Figure 12.1 illustrates the principles of flotation in a mechanical flotation cell. The agitator provides enough turbulence in the pulp phase to promote collision of particles and bubbles which
results in the attachment of valuable particles to bubbles and their transport into the froth phase for recovery.

The process can only be applied to relatively fine particles, because if they are too large the adhesion between the particle and the bubble will be less than the particle weight and the bubble will therefore drop its load. There is an optimum size range for successful flotation (Trahar and Warren, 1976; Crawford and Ralston, 1988; Finch and Dobby, 1990).

In flotation concentration, the mineral is usually transferred to the froth, or float fraction, leaving the gangue in the pulp or tailing. This is direct flotation and the opposite is reverse flotation, in which the gangue is separated into the float fraction.

The function of the froth phase is to enhance the overall selectivity of the flotation process. The froth achieves this by reducing the recovery of entrained material to the concentrate stream, while preferentially retaining the attached material. This increases the concentrate grade whilst limiting as far as possible the reduction in recovery of valuables. The relationship between recovery and grade is a trade-off that needs to be managed according to operational constraints and is incorporated in the management of an optimum froth stability. As the final separation phase in a flotation cell, the froth phase is a crucial determinant of the grade and recovery of the flotation process.

The mineral particles can only attach to the air bubbles if they are to some extent water-repellent, or hydrophobic. Having reached the surface, the air bubbles can only continue to support the mineral particles if they can form a stable froth, otherwise they will burst and drop the mineral particles. To achieve these conditions it is necessary to use the numerous chemical compounds known as flotation reagents, (Ranney, 1980; Crozier, 1984; Suttill, 1991; Nagaraj, 1994; Fuerstenau and Somasundaran, 2003).

The activity of a mineral surface in relation to flotation reagents in water depends on the forces which operate on that surface. The forces tending to separate a particle and a bubble are shown in Figure 12.2. The tensile forces lead to the development of an angle between the mineral surface and the bubble surface. At equilibrium,

\[ \gamma_{s/a} = \gamma_{s/w} + \gamma_{w/a} \cos \theta \]  

(12.1)

where \( \gamma_{s/a} \), \( \gamma_{s/w} \) and \( \gamma_{w/a} \) are the surface energies between solid and air, solid and water and water and air, respectively, and \( \theta \) is the contact angle between the mineral surface and the bubble.

The force required to break the particle–bubble interface is called the work of adhesion, \( W_{s/a} \), and

---

**Figure 12.1** Principle of froth flotation

**Figure 12.2** Contact angle between bubble and particle in an aqueous medium
is equal to the work required to separate the solid–air interface and produce separate air–water and solid–water interfaces, i.e.

$$W_{s/a} = \gamma_{w/a} + \gamma_{s/w} - \gamma_{s/a} \quad (12.2)$$

Combining with Equation 12.1 gives

$$W_{s/a} = \gamma_{w/a}(1 - \cos \theta) \quad (12.3)$$

It can be seen that the greater the contact angle, the greater is the work of adhesion between particle and bubble and the more resilient the system is to disruptive forces. The hydrophobicity of a mineral therefore increases with the contact angle; minerals with a high contact angle are said to be aerophilic, i.e. they have a higher affinity for air than for water. The terms hydrophobicity and floatability are often used interchangeably. Hydrophobicity, however, refers to a thermodynamic characteristic, whereas floatability is a kinetic characteristic and incorporates other particle properties affecting amenability to flotation (Leja, 1982; Laskowski, 1986; Woods, 1994).

Most minerals are not water-repellent in their natural state and flotation reagents must be added to the pulp. The most important reagents are the collectors, which adsorb on mineral surfaces, rendering them hydrophobic (or aerophilic) and facilitating bubble attachment. The frothers help maintain a reasonably stable froth. Regulators are used to control the flotation process; these either activate or depress mineral attachment to air bubbles and are also used to control the pH of the system. Useful reviews of flotation reagents and their behaviour include those of Crozier (1984); Somasundaran and Sivakumar (1988), Ahmed and Jameson (1989), Adkins and Pearse (1992), Nagaraj (1994), Buckley and Woods (1997), and Ralston et al. (2001).

**Classification of minerals**

All minerals are classified into polar or non-polar types according to their surface characteristics. The surfaces of non-polar minerals are characterised by relatively weak molecular bonds. The minerals are composed of covalent molecules held together by van der Waals forces, and the non-polar surfaces do not readily attach to the water dipoles, and in consequence are hydrophobic. Minerals of this type, such as graphite, sulphur, molybdenite, diamond, coal, and talc, thus have high natural floatabilities with contact angles between 60 and 90°. Although it is possible to float these minerals without the aid of chemical agents, it is universal to increase their hydrophobicity by the addition of hydrocarbon oils or frothing agents. Creosote, for example, is widely used to increase the floatability of coal. Use is made of the natural hydrophobicity of diamond in *grease tabling*, a classical method of diamond recovery which is still used in some plants. The pre-concentrated diamond ore slurry is passed over inclined vibrating tables, which are covered in a thick layer of petroleum grease. The diamonds become embedded in the grease because of their water-repellency, while the water-wetted gangue particles are washed off the table. The grease is skimmed off the table either periodically or continuously, and placed in perforated pots (Figure 12.3), which are immersed in boiling water. The grease melts, and runs out through the perforations, and is collected and re-used, while the pot containing

![Figure 12.3 Diamond recovery grease table](image-url)
the diamonds is transported to the diamond-sorting section.

Minerals with strong covalent or ionic surface bonding are known as polar types, and exhibit high free energy values at the polar surface. The polar surfaces react strongly with water molecules, and these minerals are naturally hydrophilic.

The polar group of minerals have been subdivided into various classes depending on the magnitude of polarity (Wrobel, 1970), which increases from groups 1 to 5 (Table 12.1). Minerals in group 3(a) can be rendered hydrophobic by sulphidisation of the mineral surface in an alkaline aqueous medium. Apart from the native metals, the minerals in group 1 are all sulphides, which are only weakly polar due to their covalent bonding, which is relatively weak compared to the ionic bonding of the carbonate and sulphate minerals. In general, therefore, the degree of polarity increases from sulphide minerals, through sulphates, to carbonates, halites, phosphates, etc., then oxides-hydroxides, and, finally, silicates and quartz.

Collectors

Hydrophobicity has to be imparted to most minerals in order to float them. In order to achieve this, surfactants known as collectors are added to the pulp and time is allowed for adsorption during agitation in what is known as the conditioning period. Collectors are organic compounds which render selected minerals water-repellent by adsorption of molecules or ions on to the mineral surface, reducing the stability of the hydrated layer separating the mineral surface from the air bubble to such a level that attachment of the particle to the bubble can be made on contact.

Collector molecules may be ionising compounds, which dissociate into ions in water, or non-ionising compounds, which are practically insoluble, and render the mineral water-repellent by covering its surface with a thin film.

Ionising collectors have found very wide application in flotation. They have complex molecules which are asymmetric in structure and are heteropolar, i.e. the molecule contains a non-polar hydrocarbon group and a polar group which may be one of a number of types. The non-polar hydrocarbon radical has pronounced water-repellent properties, whereas the polar group reacts with water.

Ionising collectors are classed in accordance with the type of ion, anion or cation that produces the water-repellent effect in water. This classification is given in Figure 12.4.

The structure of sodium oleate, an anionic collector in which the hydrocarbon radical, which
Froth flotation

Collectors

Non-ionising
Liquid, non-polar hydrocarbons
which do not dissociate in water

Ionising

Cationic
Cation is water-repellent.
Based on pentavalent nitrogen

Anionic

Based on organic and sulfo acid groups

Oxyhydroxyl

Based on bivalent sulphur

Sulphhydryl

Based on pentavalent nitrogen

Carboxylic
Sulphates
Sulphonates

Xanthates
Dithiophosphates

(Fatty acids and soaps)

Figure 12.4 Classification of collectors (after Glembotskii et al., 1972)

Figure 12.5 Structure of sodium oleate

does not react with water, constitutes the non-polar part of the molecule, is shown in Figure 12.5.

Amphoteric collectors possess a cationic and an anionic function, depending on the working pH, and have been used to treat sedimentary phosphate deposits (Houot et al., 1985) and to improve the selectivity of cassiterite flotation (Broekaert et al., 1984).

Because of chemical, electrical, or physical attraction between the polar portions and surface sites, collectors adsorb on the particles with their non-polar ends orientated towards the bulk solution, thereby imparting hydrophobicity to the particles (Figure 12.6). They are usually used in small amounts, substantially those necessary to form a monomolecular layer on particle surfaces (starvation level), as increased concentration, apart from the cost, tends to float other minerals, reducing selectivity. It is always harder to eliminate a collector already adsorbed than to prevent its adsorption.

An excessive concentration of a collector can also have an adverse effect on the recovery of the valuable minerals, possibly due to the development of collector multi-layers on the particles, reducing the proportion of hydrocarbon radicals orientated into the bulk solution. The hydrophobicity of the
particles is thus reduced, and hence their floatability. The flotation limit may be extended without loss of selectivity by using a collector with a longer hydrocarbon chain, thus producing greater water-repulsion, rather than by increasing the concentration of a shorter chain collector. However, chain length is usually limited to two to five carbon atoms, since the solubility of the collector in water rapidly diminishes with increasing chain length and, although there is a corresponding decrease in solubility of the collector products, which therefore adsorb very readily on the mineral surfaces, it is, of course, necessary for the collector to ionise in water for chemisorption to take place on the mineral surfaces. Not only the chain length but also the chain structure, affects solubility and adsorption (Smith, 1989); branched chains have higher solubility than straight chains.

It is common to add more than one collector to a flotation system. A selective collector may be used at the head of the circuit, to float the highly hydrophobic minerals, after which a more powerful, but less selective one, is added to promote recovery of the slower floating minerals.

Anionic collectors

These are the most widely used collectors in mineral flotation and may be classified into two types according to the structure of the polar group (Figure 12.4). Oxyhydril collectors have organic and sulpho-acid anions as their polar groups and, as with all anionic collectors, the cation takes no significant part in the reagent–mineral reaction.

Typically, oxyhydril collectors are organic acids or soaps. The carboxylates are known as fatty acids, and occur naturally in vegetable oils and animal fats from which they are extracted by distillation and crystallisation. The salts of oleic acid, such as sodium oleate (Figure 12.5) and linoleic acid, are commonly used. As with all ionic collectors, the longer the hydrocarbon chain length, the more powerful is the water-repulsion produced, but solubility decreases. Soaps (the salts of fatty acids), however, are soluble even if the chain length is long. The carboxylates are strong collectors, but have relatively low selectivity. They are used for the flotation of minerals of calcium, barium, strontium, and magnesium, the carbonates of non-ferrous metals, and the soluble salts of alkali metals and alkaline earth metals (Finch and Riggs, 1986).

The sulphates and sulphonates are used more rarely. They possess similar properties to fatty acids, but have lower collecting power. However, they have greater selectivity and are used for floating barite, celestite, fluorite, apatite, chromite, kyanite, mica, cassiterite, and scheelite (Holme, 1986).

The oxyhydril collectors have been used to float cassiterite, but have now been largely replaced by other reagents such as arsenic and phosphonic acids and sulphosuccinamates (Broekaert et al., 1984; Collins et al., 1984; Baldauf et al., 1985).

The most widely used collectors are of the sulphydryl type where the polar group contains bivalent sulphur (thio compounds). They are very powerful and selective in the flotation of sulphide minerals (Avotins et al., 1994). The mercaptans (thiols) are the simplest of thio compounds, having the general formula RS\(^-\)Na (or K)\(^+\), where R is the hydrocarbon group. They have been used as selective collectors for some of the more refractory sulphide minerals (Shaw, 1981). The most widely used thiol collectors are the xanthogenates (technically known as the xanthates) and the dithiophosphates (Aerofloat collectors). The xanthates are the most important for sulphide mineral flotation. They are prepared by reacting an alkali hydroxide, an alcohol and carbon disulphide:

\[ \text{ROH} + \text{CS}_2 + \text{KOH} = \text{RO.CS.SK} + \text{H}_2\text{O} \]  

(12.4)

where R is the hydrocarbon group and contains normally one to six carbon atoms, the most widely used xanthates being ethyl, isopropyl, isobutyl, amy, and hexyl. Sodium ethyl xanthate is typical and has the structure shown in Figure 12.7. The anion consists of a hydrocarbon non-polar radical and a connected polar group. Although the cation (sodium or potassium) plays no part in the reactions leading to mineral hydrophobicity, it has been shown (Ackerman et al., 1986) that the sodium

![Figure 12.7 Structure of sodium ethyl xanthate](image-url)
form of the alkyl xanthates decreases in efficacy with age, probably due to water absorption from the atmosphere, whereas the potassium salts are not affected by this problem.

The dithiophosphates have pentavalent phosphorus in the polar group, rather than tetravalent carbon (Figure 12.8).

![Figure 12.8 Dithiophosphates](image)

The reaction between sulphide minerals and sulphydryl collectors is complex and various mechanisms have been proposed (Yoon and Basilio, 1993). Xanthates are assumed to adsorb on sulphide mineral surfaces due to chemical forces between the polar group and the surface, resulting in insoluble metal xanthates, which are strongly hydrophobic. Mechanisms involving the formation and adsorption of dixanthogen, xanthic acid, etc., have also been proposed. It has been established that the sulphide is not joined to the collector anions without the previous action of oxygen. The solubilities of sulphide minerals in water are very low, suggesting that sulphides should be relatively inert in aqueous solution. However, they are thermodynamically unstable in the presence of oxygen, and surface oxidation to S\(^2^-\), S\(_2\)O\(_3\)^2-\(^-\), and SO\(_4\)^2-\(^-\) can occur, depending on the Eh-pH conditions. Figure 12.9 shows the Eh-pH (Pourbaix) diagram for galena. At cathodic potential, the surface of galena is converted to lead, and sulphide ions pass into solution. Under anodic conditions (i.e. when cathodic reduction of oxygen occurs, e.g. \(\frac{1}{2}O_2 + H_2O + 2e \rightarrow 2OH^-\)), lead will dissolve or form oxidised metal species on the surface, depending on the pH. The initial oxidation of sulphide leads to the formation of elemental sulphur, e.g. in acid solution:

\[ MS \rightarrow M^{2+} + S^0 + 2e \]  \hspace{1cm} (12.5)

with its equivalent in neutral or alkaline solution:

\[ MS + 2H_2O \rightarrow M(OH)_2 + S^0 + 2H^+ + 2e \]  \hspace{1cm} (12.6)

The presence of elemental sulphur in the mineral surface can lead to hydrophobicity, and the mineral may be floated in the absence of collectors, although control of these redox conditions is difficult in practice. Usually the cathodic reduction of oxygen is strong enough to provide a sufficient electron sink for oxidation of the sulphide mineral surface to oxy-sulphur species, which are not hydrophobic, and so collectors are required to promote flotation. The oxidation products are more soluble than the sulphides, and the reaction of xanthates and other thiol collectors with these products by an ion-exchange process is the major adsorption mechanism for the flotation of sulphides (Shergold, 1984). For instance, if the sulphide surface oxidises to thiosulphate, the following reactions can occur:

\[ 2MS + 2O_2 + H_2O \rightarrow MS_2O_3 + M(OH)_2 \]

and

\[ MS_2O_3 + 2ROCS_2^- \rightarrow M(ROCS_2)_2 + S_2O_3^{2-} \]

or

\[ 2MS + 4ROCS_2^- + 3H_2O \rightarrow 2M(ROCS_2)_2 + S_2O_3^{2-} + 6H^+ \]

\[ + 8e \text{ (anodic reaction)} \]  \hspace{1cm} (12.7)

The insoluble metal xanthate so formed renders the mineral surface hydrophobic. However, strong
oxidising conditions can lead to the formation of sulphates, e.g.:

\[
MS + 2\text{ROCS}_2^- + 4\text{H}_2\text{O} \rightarrow M(\text{ROCS}_2)_2 + \text{SO}_4^{2-} + 8\text{H}^+ + 8e^- \quad (12.8)
\]

Although sulphates react strongly with xanthates, they are relatively soluble in aqueous solution, and so do not form stable hydrophobic surface products, the metal xanthates formed tending to scale off the mineral.

The solubilities of the hydrophobic xanthates of copper, lead, silver, and mercury are very low, but the xanthates of zinc and iron are much more soluble. Typically, ethyl xanthates are only weak collectors of pure sphalerite, but replacement of the crystal lattice zinc atoms by copper improves the flotation properties of the mineral. The alkali earth metal xanthates (calcium, barium, magnesium) are very soluble and xanthates have no collector action on the minerals of such metals, nor on oxides, silicates, or aluminosilicates, which permits extremely selective flotation of sulphides from gangue minerals.

Xanthates are used as collectors for oxidised minerals such as malachite, cerussite, and anglesite, and for native minerals such as gold, silver, and copper. Comparatively high concentrations are necessary with the oxidised minerals, and often higher xanthates such as amyl are preferred.

Xanthates and similar compounds tend to oxidise fairly easily, which can lead to complications in flotation. After a few months of storage, they develop a strong odour and a deeper colour due to the formation of "dixanthogen", e.g. with potassium ethyl xanthate:

\[
\begin{align*}
2[C_2H_5-O-C\equiv S]+\frac{1}{2}O_2+\text{CO}_2 & \\
S-\text{K} & \\
S=C-O-C_2H_5+K_2\text{CO}_3 & \\
S=C-O-C_2H_5 & \\
\end{align*}
\]

Dixanthogens and similar products of oxidation are themselves collectors (Jones and Woodcock, 1983), and their formation can lead to loss of selectivity and control in complex flotation circuits.

Xanthates also form insoluble metal salts with ions of copper, lead, and other heavy metals which may be present in the slurry, which reduces the effectiveness of the collector. By using alkaline conditions, preferably as early as the grinding circuit, these heavy metal ions can be precipitated as relatively insoluble hydroxides. Alkaline conditions also inhibit xanthate breakdown, which proceeds more rapidly as the pH is lowered:

\[
\text{H}^+ + \text{ROCS}_2^- \leftrightarrow \text{HX} \rightarrow \text{ROH} + \text{CS}_2 \quad (12.10)
\]

With xanthic acid (HX) and xanthate ions in equilibrium, the unstable xanthic acid decomposes to alcohol and carbon disulphide.

Dithiophosphates are not as widely used as the xanthates, but are still important reagents in practice. They are comparatively weak collectors, but give good results in combination with xanthates. They are often used in the separation of copper from lead sulphides, as they are effective selective collectors for copper sulphide minerals.

It appears that the water repulsion imparted to the mineral surface is due to the formation of an oxidation product of the dithiophosphate collector which adsorbs on to the mineral surface. Thus, as with xanthates, the presence of oxygen, or another oxidising agent, is essential for flotation. Strong oxidising conditions destroy the hydrophobic substances and are thus undesirable, while oxidisation of the mineral surface itself may impede collector adsorption.

Hartati et al. (1997) described the properties of monothiphosphate (MTP) and showed how this collector dramatically altered the collecting property of dithiophosphate (DTP) when one of the S atoms was replaced by an O atom particularly in the flotation of gold in porphyry copper ores. They showed that MTP achieved selectivity in the flotation of gold against pyrite at an alkaline environment.

Various reviews of the interaction between xanthates, dithiophosphates, other thiol collectors, and mixtures thereof with sulphide mineral surfaces have been made (Klimpel, 1986; Woods and Richardson, 1986; Aplan and Chander, 1987; Crozier, 1991; Adkins and Pearse, 1992; Bradshaw, 1997) and a list of the common thio collectors is given in Table 12.2, including references which provide more detail on these extremely important reagents.
<table>
<thead>
<tr>
<th>Reagent</th>
<th>Formula</th>
<th>pH range</th>
<th>Main uses</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>O-alkyl dithiocarbonates (Xanthates)</td>
<td>R-O-C=S</td>
<td>8-13</td>
<td>Flotation of sulphides, oxidised minerals such as malachite, cerussite, and elemental metals</td>
<td>Leja (1982); Rao (1971)</td>
</tr>
<tr>
<td></td>
<td>(or Na+)</td>
<td></td>
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</tr>
<tr>
<td></td>
<td>R--O--S--K+</td>
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<td></td>
<td>(or Na+)</td>
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</tr>
<tr>
<td>Dialkyl dithiophosphates (Aerofloats)</td>
<td>R-O-P=S</td>
<td>4-12</td>
<td>Selective flotation of copper and zinc sulphides from galena</td>
<td>Mingione (1984)</td>
</tr>
<tr>
<td></td>
<td>(or Na+)</td>
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<tr>
<td></td>
<td>R--O--S--K+</td>
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<td></td>
<td>(or Na+)</td>
<td></td>
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</tr>
<tr>
<td>Dialkyl dithiocarbamate</td>
<td>R-N=C=S</td>
<td>5-12</td>
<td>Similar properties to xanthates, but more expensive</td>
<td>Jiwu et al. (1984)</td>
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<tr>
<td></td>
<td>(or Na+)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>R--S--K+</td>
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<tr>
<td></td>
<td>(or Na+)</td>
<td></td>
<td></td>
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</tr>
<tr>
<td>Isopropyl thionocarbamate (Minerec 1661/Z-200)</td>
<td>(CH₃)₂CH--O--C--N</td>
<td>4-9</td>
<td>Selective flotation of copper sulphides from pyrite</td>
<td>Ackerman et al. (1984)</td>
</tr>
<tr>
<td></td>
<td>O--C--N</td>
<td></td>
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<tr>
<td></td>
<td>(CH₃)₂CH</td>
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<td></td>
<td>C₂H₅</td>
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</tr>
<tr>
<td>Mercaptobenzothiazole (R404/425)</td>
<td>N=C=S</td>
<td>4-9</td>
<td>Flotation of tarnished or oxidised lead and copper minerals. Floats pyrite at pH 4-5</td>
<td>Fuerstenau and Raghavan (1986)</td>
</tr>
<tr>
<td></td>
<td>(or Na+)</td>
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<td></td>
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</tbody>
</table>
Chelating reagents have potential as flotation collectors, in view of their ability to form stable, selective compounds with the cations present on mineral surfaces (Somasundaran et al., 1993; Marabini, 1994). They are highly specific complexing reagents consisting of large organic molecules capable of bonding to the metal ion via two or more functional groups. However, despite several successful attempts at laboratory scale to demonstrate their effectiveness, the number of commercial plants using these reagents is relatively insignificant, mainly due to their prohibitive cost.

Cationic collectors

The characteristic property of this group of collectors is that the water-repulsion is produced by the cation where the polar group is based on pentavalent nitrogen, the amines (Figure 12.10) being the most common (Gefvert, 1986; Zachwieja, 1994). The anions of such collectors are usually halides, or more rarely hydroxides, which take no active part in the reaction with minerals.

\[
\begin{align*}
R & \quad R \\
\text{Cation} & \quad \text{Anion} \\
\text{OH} & \\
\end{align*}
\]

Figure 12.10 Cationic amine collector

Unlike the xanthates, the amines are considered to adsorb on mineral surfaces primarily due to electrostatic attraction between the polar head of the collector and the charged electrical double layer on the mineral surface. Such forces are not as strong or irreversible as the chemical forces characteristic of anionic collectors, so these collectors tend to be relatively weak in collecting power.

Cationic collectors are very sensitive to the pH of the medium, being most active in slightly acid solutions and inactive in strongly alkaline and acid media. They are used for floating oxides, carbonates, silicates, and alkali earth metals such as barite, carnallite, and sylvite. The primary amines (i.e. those where only one hydrocarbon group is present with two hydrogen atoms) are strong collectors of apatite and they can selectively float sedimentary phosphates from calcareous ores. The collector requirement can be reduced by adding a non-polar reagent such as kerosene, which is co-adsorbed on the mineral surface. Since the zeta potentials of both apatite and dolomite are negative in the relevant pH range, the selective flotation of the phosphate may not be interpreted solely by the electrostatic model of adsorption, and experimental evidence for chemical interaction has been presented (Soto and Iwasaki, 1985).

Frothers

Frothers are added to stabilise bubble formation in the pulp phase, to create a reasonably stable froth to allow selective drainage from the froth of entrained gangue, and to increase flotation kinetics. The importance of the froth phase to flotation performance is being increasingly recognised and the factors affecting froth stability are being extensively researched (Harris 1982; Melo and Laskowski 2003, 2005; Hatfield et al., 2004; Barbian et al., 2005).

Plant practice involving frothers has been reviewed by Crozier and Klimpel (1989).

Frothers are in many respects chemically similar to ionic collectors, and, indeed, many of the collectors, such as oleates, are powerful frothers, being in fact too powerful to be used as efficient frothers, since the froths which they produce can be too stable to allow efficient transport to further processing. Froth build-up on the surfaces of thickener and excessive frothing of flotation cells are problems occurring in many mineral processing plants. A good frother should have negligible collecting power, and also produce a froth which is just stable enough to facilitate transfer of floated mineral from the cell surface to the collecting launder.

Frothers are generally heteropolar surface-active organic reagents, capable of being adsorbed on the air–water interface. When surface-active molecules react with water, the water dipoles combine readily with the polar groups and hydrate them, but there is practically no reaction with the non-polar hydrocarbon group, the tendency being to force the latter into the air phase. Thus the heteropolar structure of the frother molecule leads to its adsorption, i.e. the molecules concentrate in the surface layer with the non-polar groups oriented towards the air and the polar groups towards the water (Figure 12.11).
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Water

Figure 12.11 Action of the frother

Frothing action is thus due to the ability of the frother to adsorb on the air–water interface because of its surface activity and to reduce the surface tension, thus stabilising the air bubble.

Frothers must be to some extent soluble in water, otherwise they would be distributed very unevenly in an aqueous solution and their surface-active properties would not be fully effective. The most effective frothers include in their composition one of the following groups:

- Hydroxyl: $\text{OH}$
- Carboxyl: $\text{C}=\text{O}$
- Carbonyl: $\text{C}==\text{O}$
- Amino group: $\text{NH}_2$
- Sulpho group: $\text{SO}_2\text{OH}, \text{HSO}_3\text{OH}$

The acids, amines, and alcohols are the most soluble of the frothers. The alcohols ($\text{OH}$) are the most widely used, since they have practically no collector properties, and in this respect are preferable to other frothers, such as the carboxyls, which are also powerful collectors; the presence of collecting and frothing properties in the same reagent may make selective flotation difficult. Frothers with an amino group and certain sulpho group frothers also have weak collector properties.

Pine oil, which contains aromatic alcohols, the most active frothing component being terpineol, $\text{C}_{10}\text{H}_{17}\text{OH}$, has been widely used as a frother. Cresol (cresylic acid), $\text{C}_8\text{H}_7\text{OH}$, has also had wide use.

A wide range of synthetic frothers, based mainly on high molecular-weight alcohols, are now in use in many plants. They have the important advantage over industrial products such as pine oil and cresol in that their compositions are much more stable, which makes it easier to control the flotation process and improves performance. A widely used synthetic alcohol frother is methyl isobutyl carbinol (MIBC). Another range of synthetic frothers are based on polyglycol ethers, and have been found to be very effective. They are marketed under various names, such as Cytec Oreprep 549 and Cytec Aerofroth 65. Frothers based on polyglycols are also used, and it is not uncommon to blend all three chemical groups – alcohols, polyglycol ethers, and polyglycols – together to provide a specific frother for a particular flotation circuit (Riggs, 1986).

The alcohol groups provide a selective, often brittle, froth, which allows good control and materials transfer through the launders and pumps. The glycol ether group is stronger, with more persistence than the alcohol groups, while the polyglycols are the strongest surface active frothers utilised. They are very effective in maximising load support with coarse grinds and high grade feeds, at all pH ranges.

Although frothers are generally surface-active reagents, it has been shown that surface-inactive reagents, such as diacetone alcohol and ethyl acetal, behave as frothers in solid–liquid–air systems, although not in two-phase liquid–air systems (Lekki and Laskowski, 1975). Molecules of these reagents have two polar groups and are readily soluble in water. They adsorb on solid surfaces but do not appreciably change their hydrophobicity. When the mineral surface, on which the surface inactive frother is adsorbed, is approached by an air bubble, the molecules reorientate and produce a sufficiently stable three-phase froth. Being surface inactive, these reagents do not reduce surface tension, and apart from the slight reduction due to collectors, the forces available for flotation are maintained at their maximum.

Regulators

Regulators, or modifiers, are used extensively in flotation to modify the action of the collector, either by intensifying or by reducing its water-repellent effect on the mineral surface. They thus make collector action more selective towards certain minerals. Regulators can be classed as activators, depressants, or pH modifiers.
Activators

These reagents alter the chemical nature of mineral surfaces so that they become hydrophobic due to the action of the collector. Activators are generally soluble salts which ionise in solution, the ions then reacting with the mineral surface.

A classical example is the activation of sphalerite by copper in solution. Sphalerite is not floated satisfactorily by a xanthate collector, since the collector products formed, such as zinc xanthate, are relatively soluble in water, and so do not provide a hydrophobic film around the mineral. Floatability can be improved by the use of large quantities of long-chain xanthates, but a more satisfactory method is to use copper sulphate as an activator, which is readily soluble and dissociates into copper ions in solution. Activation is due to the formation of molecules of copper sulphide at the mineral surface, due to the fact that copper is more electronegative than zinc and therefore ionises less readily:

\[
\text{ZnS} + \text{Cu}^{2+} \leftrightarrow \text{CuS} + \text{Zn}^{2+} \quad (12.11)
\]

The copper sulphide deposited on the sphalerite surface reacts readily with xanthate to form insoluble copper xanthate, which renders the sphalerite surface hydrophobic. Recent work, however, suggests that this simple ion-exchange mechanism may be oversimplified, and Wang et al. (1989a,b) propose a model based on surface oxidation of the mineral and reduction of the activator, surface precipitation of the activator hydroxide and a mixed potential mechanism.

The main use of copper sulphate as an activator is in the differential flotation of lead–zinc ores, where after lead flotation the sphalerite is activated and floated. To some extent, copper ions can also activate galena, calcite, and pyrite. When sphalerite is associated with pyrite or pyrrhotite, selectivity is usually ensured by the high alkalinity (pH 10.5–12) of the pulp, lime being added in conjunction with the copper sulphate activator.

Oxidised minerals of lead, zinc, and copper, such as cerussite, smithsonite, azurite, and malachite, float very inefficiently with sulphidyl collectors and require an extremely large amount, as heavy metal ions dissolved from the mineral lattice must be precipitated as metal xanthate before the collector interacts with the mineral. Adsorption at the mineral surface is also poor, the collector coating being readily removed by particle abrasion. Such minerals are activated by the use of sodium sulphide or sodium hydrosulphide (Fuerstenau et al., 1985; Malghan, 1986). Large quantities of up to 10 kg t⁻¹ of such “sulphidisers” may be required, due to the relatively high solubilities of the oxidised minerals.

In solution, sodium sulphide hydrolyses and then dissociates:

\[
\begin{align*}
\text{Na}_2\text{S} + 2\text{H}_2\text{O} & \leftrightarrow 2\text{NaOH} + \text{H}_2\text{S} \quad (12.12) \\
\text{NaOH} & \leftrightarrow \text{Na}^+ + \text{OH}^- \quad (12.13) \\
\text{H}_2\text{S} & \leftrightarrow \text{H}^+ + \text{HS}^- \quad (12.14) \\
\text{HS}^- & \leftrightarrow \text{H}^+ + \text{S}^{2-} \quad (12.15)
\end{align*}
\]

Since the dissociation constants of Equations 12.14 and 12.15 are extremely low and that of Equation 12.13 is high, the concentration of OH⁻ ions increases at a faster rate than that of H⁺ ions and the pulp becomes alkaline. Hydrolysis and dissociation of sodium sulphide release OH⁻, S²⁻, and HS⁻ ions into solution and these can react with and modify the mineral surfaces. Sulphidation causes sulphur ions to pass into the crystal lattice of the oxidised minerals, giving them a relatively insoluble pseudo-sulphide surface coating and allowing them to be floated by sulphidyl collectors. For example, in the sulphidisation of cerussite, the following reactions take place:

\[
\begin{align*}
\text{Na}_2\text{S} + \text{H}_2\text{O} & \leftrightarrow \text{NaHS} + \text{NaOH} \quad (12.16) \\
\text{PbCO}_3 + 3\text{NaOH} & = \text{H}_2\text{O} + \text{Na}_2\text{CO}_3 + \text{NaHPbO}_3 \quad (12.17) \\
\text{NaHS} + \text{NaHPbO}_2 & = 2\text{NaOH} + \text{PbS} \quad (12.18) \\
or \\
\text{Na}_2\text{S} + \text{PbCO}_3 & = \text{Na}_2\text{CO}_3 + \text{PbS} \quad (12.19)
\end{align*}
\]

The amount of sodium sulphide added to the pulp must be very strictly controlled, as it is a very powerful depressant for sulphide minerals and will, if in excess, depress the activated oxide minerals, preventing collector adsorption. The amount required is dependent on the pulp alkalinity, as an increase in pH causes Equations 12.14 and 12.15 to proceed further to the right, producing more HS⁻ and S²⁻ ions. For
Froth flotation

this reason sodium hydrosulphide is sometimes preferred to sodium sulphide, as the former does not hydrolyse and hence increase the pH. The amount of sulphidiser added should be sufficient only to produce a coherent film of sulphide on the mineral surface, such that xanthate can be adsorbed. With an increase in sulphidiser beyond that required for activation, concentrations of sulphide and hydro-sulphide ions increase. The HS\(^{-}\) ions readily adsorb on the mineral surfaces, giving them a high negative charge, and preventing adsorption of the collector anions. Excess sodium sulphide also removes oxygen from the pulp:

\[ \text{Na}_2\text{S} + 2\text{O}_2 = \text{Na}_2\text{SO}_4 \]  (12.20)

Since oxygen is required in the pulp for the adsorption of sulphydryl collectors on sulphide surfaces, flotation efficiency is reduced.

In the flotation of mixed sulphide-oxidised ores, the sulphide minerals are usually floated first, before sulphidisation of the oxidised surfaces. This prevents the depression of sulphides by sodium sulphide and the sulphidiser is subsequently added to the pulp in stages, in starvation levels. It has recently been suggested (Zhang and Poling, 1991) that the detrimental effects of residual hydro-sulphide can be eliminated by the addition of ammonium sulphate with the hydrosulphide. Use of the relatively inexpensive ammonium sulphate appears to reduce the consumption of the much more expensive sulphidising agent and enhances the activating effect of the hydrosulphide ions. Zhou and Chander (1991) have further suggested that sodium tetrasulphide may be superior to sodium sulphide in terms of flotation response, and propose mechanisms for the reactions.

**Depressants**

Depression is used to increase the selectivity of flotation by rendering certain minerals hydrophilic (water-avid), thus preventing their flotation. They are key to the economic flotation of certain ores such as platinum and nickel sulphides.

There are many types of depressants and their actions are complex and varied, and in most cases not fully understood, making depression more difficult to control than the application of other types of reagent, particularly when the froth phase is also affected by their action (Bradshaw et al., 2005).

Slime coating is an example of a naturally occurring form of depression. Slimes in a crushed and ground ore make flotation difficult, as they coat the mineral particles, retarding adsorption of the collector anions. Excess sodium sulphide also removes oxygen from the pulp:

\[ \text{Na}_2\text{S} + 2\text{O}_2 = \text{Na}_2\text{SO}_4 \]  (12.20)

Since oxygen is required in the pulp for the adsorption of sulphydryl collectors on sulphide surfaces, flotation efficiency is reduced.

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**Inorganic depressants**

Cyanides are widely used in the selective flotation of lead–copper–zinc and copper–zinc ores as depressants for sphalerite, pyrite, and certain copper sulphides. Sphalerite rejection from copper concentrates is often of major concern, as zinc is a penalty element in copper smelting.

It is fairly well established that pure clean sphalerite does not adsorb short-chain xanthates until its surface is activated by copper ions (Equation 12.11). However, copper ions resulting from very slight dissolution of copper minerals present in the ore may cause unintentional activation and prevent selective separation. Cyanide is added to the pulp to adsorb the surface copper and to react with copper in solution forming soluble cyanide complexes. Sodium cyanide is most commonly used, which hydrolyses in aqueous solution to
form free alkali and relatively insoluble hydrogen cyanide:

\[ \text{NaCN} + \text{H}_2\text{O} \leftrightarrow \text{HCN} + \text{NaOH} \quad (12.21) \]

The hydrogen cyanide then dissociates:

\[ \text{HCN} \leftrightarrow \text{H}^+ + \text{CN}^- \quad (12.22) \]

The dissociation constant of Equation 12.22 is extremely low compared with that of Equation 12.21, so that an increase in pulp alkalinity reduces the amount of free HCN, but increases the concentration of CN\(^-\) ions. An alkaline pulp is essential, as free hydrogen cyanide is extremely dangerous. The major function of the alkali, however, is to control the concentration of cyanide ions available for dissolution of the copper to cupro-cyanide:

\[ 3\text{CN}^- + \text{Cu}^{2+} \leftrightarrow [\text{Cu(CN)}_2]^- + \frac{1}{2}\text{C}_2\text{N}_2 \quad (12.23) \]

Apart from the reactions of cyanide with metal ions in solution, it can react with metal xanthates to form soluble complexes, preventing xanthate adsorption on the mineral surface, although this cannot occur until the metal ions in solution have been complexed, according to Equation 12.23. Hence if Cu\(^{2+}\) ions are in solution, the prevention of xanthate adsorption cannot occur unless the ratio of CN\(^-\) ions to Cu\(^{2+}\) ions is greater than 3:1. The greater the solubility of the metal xanthate in cyanide, the less stable is the attachment of the collector to the mineral. It has been shown that lead xanthates have very low solubilities in cyanide, copper xanthates are fairly soluble, while the xanthates of zinc, nickel, gold, and iron are highly soluble. Iron and zinc can, therefore, be very easily separated from lead in complex ores. In the separation of chalcopyrite from sphalerite and pyrite, very close control of cyanide ion concentration is needed. Cyanide should be added sufficiently only to complex the heavy metal ions in solution, and to solubilise the zinc and iron xanthates. Excess cyanide forms soluble complexes with the slightly less soluble copper xanthate, depressing the chalcopyrite.

The depressive effect of cyanide depends on its concentration and on the concentration of the collector and the length of the hydrocarbon chain. The longer the chain, the greater is the stability of the metal xanthate in cyanide solutions and the higher the concentration of cyanide required to depress the mineral. Relatively low concentrations of xanthates with short hydrocarbon chains are therefore used for selective flotation where cyanides are used as depressants.

Cyanides are, of course, extremely toxic and must be handled with great care. They also have the disadvantage of being expensive and they depress and dissolve gold and silver, reducing the extraction of these metals into the froth product. Despite these disadvantages, they are widely used due to their high degree of selectivity in flotation. They also have the advantage of leaving the mineral surface relatively unaffected, so that subsequent activation is relatively simple, although residual cyanide ions in solution may interfere with the activator.

While many plants function efficiently with cyanide alone, in others an additional reagent, generally zinc sulphate, is added to ensure satisfactory depression of sphalerite. If copper ions are present, the introduction of zinc ions can prevent the copper depositing on the sphalerite surface by shifting Equation 12.11 to the left.

However, other, more complex, reactions occur to assist depression and it is considered that cyanide reacts with zinc sulphate to form zinc cyanide, which is relatively insoluble, and precipitates on the sphalerite surface, rendering it hydrophilic and preventing collector adsorption:

\[ \text{ZnSO}_4 + 2\text{NaCN} = \text{Zn(CN)}_2 + \text{Na}_2\text{SO}_4 \quad (12.24) \]

In an alkaline pulp, zinc hydroxide, which adsorbs copper ions, is also formed and it is precipitated on to the sphalerite surface, preventing collector adsorption.

The use of zinc sulphate thus reduces cyanide consumption and cases have been known where depression of sphalerite has been achieved by the use of zinc sulphate alone.

Although cyanides and zinc sulphate are widely used, they do have many disadvantages, for example many concentrators being loath to use cyanides due to environmental problems. Zinc sulphate is effective only at high pH values, at which zinc hydroxide precipitates from solution. There is a need, therefore, for alternative selective depressants. Research on Pb–Zn ores in Yugoslavia has shown that sphalerite depression by zinc sulphate and sodium cyanide can be successfully replaced by a combination of
Froth flotation

ferrosulphate and sodium cyanide (Pavlica et al., 1986). This has the advantage of reducing sodium cyanide consumption, with consequent economic and ecological advantages. Zinc bisulphite used with cyanide in alkaline conditions is being used to treat bulk copper–zinc–iron concentrates at the Cerro Colorado mill in Spain (Ser and Nieto, 1985). This reagent combination was found to give results which were more favourable than those obtained using standard depression techniques, such as cyanide–zinc sulphate, which was found to be very sensitive to variations in the chalcocite content of the ore.

Sphalerite activation can be prevented by eliminating copper ions from the flotation pulp, and in some plants precipitation with hydrogen sulphide or sodium sulphide is carried out.

Sulphur dioxide has developed into a most versatile and almost indispensable conditioning reagent for polysulphide ores. Although widely used as a galena depressant in copper–lead separations, it also deactivates zinc sulphides and enhances the flotation differential between zinc and other base metal sulphides. In copper cleaner and copper–lead separation circuits, very effective zinc rejection is attained through acidification of pulps by injection of SO₂. However, SO₂ cannot be employed when treating ores which contain the secondary copper minerals covellite or chalcocite, since they become soluble in the presence of sulphur dioxide and the dissolved copper ions activate zinc sulphides (Konigsman, 1985). Sulphur dioxide does not appreciably depress chalcopyrite and other copper-bearing minerals. In fact, adsorption of xanthate on chalcopyrite is enhanced in the presence of SO₂, and the addition of SO₂ before xanthate results in effective sphalerite depression while increasing the floatability of chalcopyrite. The use of SO₂ in various Swedish concentrators is discussed by Broman et al. (1985), who point out that SO₂ has the advantage over cyanide in sphalerite depression in that there is little copper depression, and no dissolution of precious metals. However, it is indicated that the use of SO₂ demands adaptation of the other reagent additions, and in some cases a change of collector type is required.

Potassium dichromate (K₂Cr₂O₇) is also used to depress galena in copper–lead separations. The depressive action is due to the chemical reactions between the galena surface and CrO₄⁻ anions, which produces insoluble lead dichromate that increases wettability and prevents flotation.

More than 40% of the Western world’s molybdenum is produced as a by-product from porphyry copper ores. The small amount of molybdenite is collected along with copper in a bulk Cu–Mo concentrate. The two minerals are then separated, almost always by depressing the copper minerals and floating the molybdenite. Sodium hydroxysulphide (or sodium sulphide) is used most extensively, though several other inorganic compounds, such as cyanides, and Noke’s reagent (a product of the reaction of sodium hydroxide and phosphorous pentasulphide), are also used (Nagaraj et al., 1986). Almost all the currently used depressants are inorganic. Numerous organic depressants have been developed over the years, but apart from sodium thioglycolate, none have been successfully commercialised (Agar, 1984).

Polymeric depressants

The use of polymeric depressants has the advantage of being less hazardous than the more widely used inorganic depressants, and interest in their use has been growing (Liu and Laskowski, 1989). Organic reagents such as starch, tannin, quebracho, and dextrin do not ionise in solution, but prevent flotation in a manner similar to a slime coating. They have been used for many years as gangue-mineral depressants, and are used in small amounts to depress talc, graphite, and calcite (Pugh, 1989). Starch and dextrin can also be used as supplementary lead depressants in copper–lead separations. Other applications include the selective depression of polynmetallic sulphide ores in the processing of iron ore (Nyamekye and Laskowski, 1993), as blin- ders in potash flotation (Arsentiev et al., 1988) and the depression of talcaeous gangue minerals in platinum and base metal flotation (Steenberg and Harris, 1984; Liu and Laskowski 1999; Shortridge et al., 2000; Bradshaw et al., 2005; Smeink et al., 2005; Wang et al., 2005).

In the South African platinum group mineral (PGM) industry, polymeric depressants such as carboxymethyl cellulose (CMC) and guar are widely used to depress talcaeous gangue minerals. One of the major differences between these two polysaccharides is that CMC is negatively charged in solution, whereas guar is typically only slightly charged, if at all (Mackenzie, 1986).
The importance of pH

It is evident from the foregoing that pulp alkalinity plays a very important, though very complex, role in flotation, and, in practice, selectivity in complex separations is dependent on a delicate balance between reagent concentrations and pH.

Flotation where possible is carried out in an alkaline medium, as most collectors, including xanthates, are stable under these conditions, and corrosion of cells, pipework etc., is minimised. Alkalinity is controlled by the addition of lime, sodium carbonate (soda ash), and to a lesser extent sodium hydroxide or ammonia. Sulphuric or sulphurous acids are used where a decrease in pH is required.

These chemicals are often used in very significant amounts in almost all flotation operations. Although they are cheaper than collectors and frothers, the overall cost is generally higher with pH regulators per tonne of ore treated than with any other processing chemical. For example, the cost of lime in sulphide mineral flotation is roughly double that of the collector used, so significant operational cost savings can be achieved by the proper choice and use of pH regulators (Fee and Klimpel, 1986).

Lime, being cheap, is very widely used to regulate pulp alkalinity, and is used in the form of milk of lime, a suspension of calcium hydroxide particles in a saturated aqueous solution. Lime or soda ash is often added to the slurry prior to flotation to precipitate heavy metal ions from solution. In this sense, the alkali is acting as a “deactivator”, as these heavy metal ions can activate sphalerite and pyrite and prevent their selective flotation from lead or copper minerals. Since the heavy metal salts precipitated by the alkali can dissociate to a limited extent and thus allow ions into solution, cyanide is often used with the alkali to complex them. Hydroxyl and hydrogen ions modify the electrical double layer and zeta potential (see Chapter 15) surrounding the mineral particles, and hence the hydration of the surfaces and their floatability is affected. With xanthates as collectors, sufficient alkali will depress almost any sulphide mineral, and for any concentration of xanthate there is a pH value below which any given mineral will float, and above which it will not float. This critical pH value depends on the nature of the mineral, the particular collector and its concentration, and the temperature (Sutherland and Wark, 1955). Figure 12.12 shows how the critical pH value for pyrite, galena, and chalcopyrite depends on the concentration of sodium aerofloat collector.

\[\text{Critical pH value} = \frac{\text{concentration of collector (mg/l)} \times \text{pH value}}{1000}\]

It is evident from the curves that using 50 mg l\(^{-1}\) of sodium aerofloat, and a pH value of 8, chalcopyrite can be floated from galena and pyrite. On reducing the pH to 6, the galena can be floated from the pyrite.

Lime can also act as a strong depressant for pyrite and arsenopyrite when using xanthate collectors. Both the hydroxyl and calcium ions participate in the depressive effect on pyrite by the formation of mixed films of Fe(OH), FeO(OH), CaSO\(_4\), and CaCO\(_3\) on the surface, so reducing the adsorption of xanthate. Lime has no such effect with copper minerals, but does depress galena to some extent. In the flotation of galena, therefore, pH control is often affected by the use of soda ash, pyrite and sphalerite being depressed by cyanide.

As was shown earlier, the effectiveness of sodium cyanide and sodium sulphide is governed to such a large extent by the value of pH that these reagents are of scarcely any value in the absence of alkalis. Since, where cyanide is used as a depressant, the function of the alkali is to control the cyanide ion concentration (Equations 12.22 and 12.23), there is for each mineral and given concentration of collector a “critical cyanide ion concentration” above which flotation is impossible. Curves for several minerals are given in Figure 12.13, and it can be seen that chalcopyrite can be floated from pyrite at pH 7.5 and 30 mg l\(^{-1}\) sodium cyanide.
Since, of the copper minerals, chalcopyrite lies closest to pyrite relative to the influence of alkali and cyanide, all the copper minerals will float with the chalcopyrite. Thus, by careful choice of pH value and cyanide concentration, excellent separations are theoretically possible, although in practice other variables serve to make the separation more difficult. Adsorption of xanthate by galena is uninfluenced by cyanide, the alkali alone acting as a depressant.

These sulphur-rich, metal-deficient zones can render the mineral hydrophobic, provided that the local conditions are such that the metal oxides/hydroxides formed by the reaction are solubilised. Excessive oxidation can produce thiosalts (Equation 12.7), and, ultimately, sulphate (Equation 12.8), together with metal ions which may re-adsorb, as hydrolysis products, on to the mineral, producing hydrophilic surfaces.

Buckley et al. (1985) studied the surface oxidation of galena, bornite, chalcopyrite, and pyrrhotite, and found that for each mineral the initial oxidation reaction is the removal of a metal component from the surface region to leave a sulphide with similar structure to the original mineral but with lower metal content. Metal-deficient sulphide layers containing high sulphur–metal ratios are probably stabilised by the underlying mineral because they have the same sulphur lattices. The authors showed that flotation of the minerals could be accomplished without the aid of collectors when a metal-deficient sulphide, rather than elemental sulphur, is formed. Naturally hydrophobic sulphide minerals, such as molybdenite, have such layer structures, the behaviour of these minerals being explained in terms of the work of adhesion of water to the surface being largely determined by dispersion forces, with hydrogen bonding and ionic interactions being small. It is possible that a similar situation exists at the surface of other sulphides where a metal-deficient layer is formed. Although the metal is dissolved at low pH (Equation 12.25), in neutral or alkaline conditions a hydroxy-oxide is formed (Equation 12.26), which could be expected to be hydrophilic. However, collectorless flotation occurs under these conditions, and the authors conclude that the metal oxides are dissolved due to the turbulence in the flotation cells, or are abraded from the mineral surfaces.

The collectorless flotation process has also been tested on six different chalcopyrite ores while monitoring the potentials of the pulp (Luttrell and Yoon, 1984). The results confirmed that collectorless flotation is effective only under oxidising conditions. In addition, the flotation requires that the chalcopyrite surface be relatively free of hydrophilic oxidation products, which can be accomplished by treating the ore pulp with sodium sulphide. The role of sodium sulphide in collectorless flotation was initially thought...
to be one of sulphidising agent. However, the excess \( \text{HS}^-/\text{S}^{2-} \) ions that have not been consumed in sulphidisation may be oxidised to become elemental sulphur or polysulphides, depending on pH, which may deposit on the mineral surface. Thus, the collectorless flotation process using sodium sulphide may provide an external source for these hydrophobic species that could enhance flotation. Collectorless flotation was also found to be pH dependent, becoming more favourable with decreasing pH.

However, as explained by Guy and Trahar (1985), the application of such findings to realistic separations is not straightforward, as the areas of floatability determined from experiments with single sulphides do not necessarily coincide with those determined from experiments with sulphide mixtures. Cations produced by sulphide oxidation may react in different ways in a given system. Apart from modifying the surfaces of some minerals by surface interactions, they may be precipitated as hydroxides which have a profound effect on sulphide floatability.

For instance, pyrite and pyrrhotite occur together in many important ores, and the galvanic interaction between these two minerals and its effect on their floatabilities have been investigated (Nakazawa and Iwasaki, 1985). The galvanic contact decreased the formation of hydroxide or oxide and sulphate species of iron on pyrrhotite, whereas such formations were increased on pyrite. The effect was to improve the floatability of pyrrhotite, while reducing that of pyrite.

The control of redox conditions is complicated not only by the galvanic interactions between the different minerals in the ore but also by the interactions between the minerals and the steel grinding medium (Martin et al., 1991). The reducing conditions at a sulphide mineral surface created by the oxidation of steel in a galvanic interaction can hinder the adsorption of the collector.

Learmont and Iwasaki (1984) have studied the interaction between galena and steel media. They show that iron oxide, hydroxide, or sulphate species form on the galena surface on contact with mild steel, decreasing the galena floatability. The time of contact and aeration conditions affect the severity of the flotation depression. Adam and Iwasaki (1984) showed that the flotation response of pyrrhotite was similarly adversely affected. The formation of hydroxide or oxide and sulphate species of iron through the oxygen reduction reaction at the cathodically polarised surface of pyrrhotite was shown to be the mechanism responsible for the reduced floatability of pyrrhotite, the following reactions being proposed:

\[
\frac{1}{2} \text{O}_2 + \text{H}_2\text{O} + 2\text{e}^- = 2\text{OH}^- \quad \text{(cathode)}
\]

\[
2\text{H}^+ + 2\text{e}^- = \text{H}_2 \quad \text{(cathode)}
\]

\[
\text{FeS} = \text{Fe}^{2+} + \text{S}^{2-} \quad \text{(dissociation)} \rightarrow \text{Fe}_2\text{O(OH)}_3
\]

or

\[
\text{FeOOH} \rightarrow \text{Fe(OH)SO}_4
\]

The formation of an iron hydroxide coating covering the mineral surface reduces mineral floatability.

Due to these many complex interactions, measurement of the pulp oxidation–reduction potential is difficult in a plant environment (Johnson and Munro, 1988; Labonte and Finch, 1988). Electrodes which have different activities for the oxygen reduction reaction, such as platinum and gold, can give rise to different \( E_h \) values, and different sulphides can give rise to different \( E_h \) values in the same solution. Because of these complexities, on-line measurement of \( E_h \) to control redox conditions is still a control strategy of the future, although some concentrators do use such actions based on operating experience, and Outokumpu Oy is developing methods to control the electrochemical potentials of minerals directly in the ore pulps in order to attain the optimum combination of \( E_h \) and pH, as well as the optimum collector addition (Heimala et al., 1988).

Pre-flotation aeration of sulphide pulps has been practised in the Noranda Group (Canada) and other organisations for many years to help depress pyrite and pyrrhotite and promote chalcopyrite and galena (Konigsman, 1985).

The introduction of talc pre-flotation at Woodlawn in Australia had a detrimental effect on copper circuit performance, due to the aeration provided by the talc cells (Williams and Phelan, 1985). The aeration promoted the flotation of the other sulphides, especially galena, relative to that of chalcopyrite. Copper circuit performance was subsequently improved by the addition of a strong reducing agent, sodium sulphide, to the talc flotation tailing.
Nitrogen is used as the carrier gas in a few molybdenum flotation circuits, a reducing potential being used to minimise the consumption of the sulphide depressant which inhibits flotation of copper minerals. Nitrogen has great potential as a carrier gas in other flotation circuits (Martin et al., 1989), apart from chalcopyrite–molybdenite separation, because of its ready availability at smelter sites and its chemical inertness. The latter means that it is unlikely to be consumed by side reactions.

Only recently has there been a revival of interest in studying the mechanism of depression of sulphides. The influence of the strongly reducing nature of sodium hydrosulphide on depressant action has been monitored by means of solution redox-potential measurements, and it would appear that the depressant activity is to some extent electrochemical, the HS⁻ ions, by virtue of their large negative $E_h$, destabilising the coating of thiol collector (Nagaraj et al., 1986). The oxidation–reduction effects in sulphide mineral depression have been reviewed by Chander (1985).

The role of bubble generation and froth performance

In the science of flotation, one of the most critical components within the process is the role of the bubbles. Gorain et al. (1997, 1998) showed that the first order rate constant ($k$) achieved in a variety of industrial flotation cells of different types and sizes operated at a range of different air rates, impeller speeds and froth depths depended on the feed ore floatability ($P$), the bubble surface area flux ($S_b$) generated in the cell and the recovery across the froth phase ($R_f$), in a simple numerical relationship:

$$K = P \cdot S_b \cdot R_f$$  \hspace{1cm} (12.27)

where $k$ = rate constant (s⁻¹); $P$ = floatability (dimensionless); $S_b$ = bubble surface area flux (s⁻¹); $R_f$ = froth recovery (fraction).

Based on these findings, the performance of a flotation unit can be considered to arise from the interaction of a stream property – the particle floatability ($P$) – with parameters that characterise the operating conditions of the pulp and froth zones of the unit ($S_b$ and $R_f$). In other words, the particle floatability is governed by the degree of hydrophobicity (as described earlier), the bubble surface area flux is a key driver within the pulp zone of a given cell, and the froth recovery describes the performance across the froth zone.

The bubble surface area flux, which is the rate at which bubble surface area moves through the cell per unit of cell cross-sectional area, can be measured directly within a cell from the measurements of superficial gas velocity ($J_g$) and the bubble size ($d_b$):

$$S_b = \frac{6 J_g}{d_b}$$  \hspace{1cm} (12.28)

where $J_g$ = the superficial gas velocity (m/s); $d_b$ = the Sauter mean bubble diameter (m)

Both $J_g$ and $d_b$ are measurable using a suitable bubble size analyzer (e.g. Tucker et al., 1994; Hernandez et al., 2002) and a superficial gas velocity probe (Gorain et al., 1996). $S_b$ can also be predicted using a correlation developed by Gorain et al. (1999) using a large number of data sets collected from different base metal flotation plants:

$$S_b = 123 J_g^{0.75} N_s^{0.44} A_s^{-0.10} P_{80}^{-0.42}$$  \hspace{1cm} (12.29)

where $N_s$ = impeller tip speed (rpm); $A_s$ = impeller aspect ratio (impeller width/impeller height) (dimensionless); $P_{80}$ = cell feed 80% passing size (μm).

Gorain et al. (1997) and Alexander et al. (2000) showed that the bubble surface area flux was linearly related to the first order rate constant at shallow froth depths. In addition, this relationship was shown to be independent of cell size and operating parameters. This is illustrated in Figure 12.14 which shows that the relationship measured in a pilot scale 60 litre cell was essentially identical to that measured in a parallel Outokumpu 100 m³ tank cell.

At present, there are several techniques available to quantify the froth recovery factor, $R_f$. However, most of these methods are either intrusive in the froth zone or subject to assumptions that cannot be adopted in conventional cell modelling (e.g. no entrainment effects). A method initially developed for batch flotation cells by Feteris et al. (1987) was later modified by Vera et al. (1999) to determine $R_f$ directly from industrial scale flotation froths. In this approach, froth recovery ($R_f$) is estimated by determining the cell recovery at a measured froth depth (and hence the first order rate constant, $k$) to the cell recovery at no froth depth (and hence the collection zone rate constant, $k_c$). The cell recovery at
Figure 12.14 First order rate constant and bubble surface area flux relationship in a 60 l pilot cell and a 100 m³ OK rougher cell (after Alexander et al. (2000))

no-froth depth cannot be measured directly, but can be estimated by extrapolation of results obtained at four or more froth depths.

An alternative technique for determining \( R_f \) in industrial flotation cells was developed by Savassi et al. (1998). A direct measurement of \( R_f \) is achieved by solving a set of mass balance equations with data obtained from a sample of the concentrate and samples taken below the pulp–froth interface. This direct measuring technique is a better approach than the first method, as no variation in the operating conditions of the flotation cell is needed. However, as noted by the authors, the technique is limited to use in roughers as it requires high bubble loads and a significant difference between the grade of the attached and suspended particles. Alexander et al. (2000) proposed a new direct method for measuring froth recovery which is applicable in other sections of the flotation circuit. This was based on the Savassi method of solving the mass balance equations across the froth but extended the technique using improved sampling methods. This is the current technique being used by many metallurgists to measure froth recovery in large industrial flotation cells.

**Entrainment**

The true flotation response has dominated the flotation literature since the separation process was first commercially used in 1905. Jowett (1966) first noted the recovery of fine particles by entrainment in water. Since then measuring techniques and mathematical models to measure and represent the entrainment mechanism have been developed by a number of authors. However, little work has been carried out in industrial scale cells. An exception was research conducted by Johnson (1972) which included industrial cell data to supplement laboratory data. This work showed that the recovery by entrainment is proportional to the feed water recovery to the concentrate. From this finding, the degree of entrainment was defined as the ratio of the recovery of entrained solids to that of water.

Johnson (1972) also showed that the degree of entrainment was a strong function of particle size: entrainment has been shown to be significant below particle sizes of 50 μm (Smith and Warren, 1989). Recently, Savassi et al. (1998) developed an empirical model to describe the relationship between the degree of entrainment and particle size. This model is represented in the equations below:

\[
\text{ENT}_i = \frac{2}{\exp(2.292(d_i/\xi)_{\text{adj}}) + \exp(-2.292d_i/\xi)}
\]

(12.30)

and

\[
\text{adj} = 1 + \frac{\ln(\delta)}{\exp(d_i/\xi)}
\]

(12.31)

where \( \text{ENT}_i \) = mass transfer of entrained particles to the concentrate ÷ mass transfer of water to the concentrate; \( d_i \) = particle size (μm); \( \xi \) = entrainment parameter, or the particle size for which the degree of entrainment is 20% (μm); \( \delta \) = drainage parameter, related to the preferential drainage of coarse particles (dimensionless).
The engineering of flotation

The industrial application of the flotation process has been practised for 100 years. Although the process is effective, industrial flotation practice often requires several stages to produce the product quality desired by the market. These stages are combined in various methods and are referred to as “flotation circuits”. In this section, the stages required in developing a flotation circuit including laboratory and pilot plant flotation testing, the types of circuits currently in practice and the types of flotation cells used are described. Lane et al. (2005) present a useful review of the logical approach to the design of a flotation flowsheet.

Laboratory flotation testing

In order to develop a flotation circuit for a specific ore, preliminary laboratory testwork must be undertaken in order to determine the choice of reagents and the size of plant for a given throughput as well as the flowsheet and peripheral data. Flotation testing is also carried out on ores in existing plants to improve procedures and for development of new reagents.

It is essential that testwork is carried out on ore which is representative of that treated in the commercial plant. Samples for testwork must be representative, not only in chemical composition but also relative to mineralogical composition and degree of dissemination. A mineralogical examination of drill cores or other individual samples should therefore be made before a representative sample is selected. Composite drill core samples are ideal for testing if drilling in the deposit has been extensive; the cores generally contain ore from points widely distributed over the area and in depth. It must be realised that ore bodies are variable and that a representative sample will not apply equally well to all parts of the ore body; it is used therefore for development of the general flotation procedure. Additional tests must be made on samples from various areas and depths to establish optimum conditions in each case and to give design data over the whole range of ore variation.

Characterisation of the flotation response of ore deposits must therefore recognise that the ore deposit could represent a variety of rock types, with different ore mineralogy, textures (fine or coarse grain) and faulting. It is therefore preferable that drill core samples be selected to represent the variations within the ore body. Each sample should be tested separately and the overall value of the deposit is then assessed by compositing the metallurgical responses of each sample mathematically.

Having selected representative samples of the ore, it is necessary to prepare them for flotation testing, which involves comminution of the ore to its optimum particle size. Crushing must be carried out with care in order to avoid accidental contamination of the sample by grease or oil, or with other materials which have been previously crushed. Even in a commercial plant, a small amount of grease or oil can temporarily upset the flotation circuit. Samples are usually crushed with small jaw crushers or cone crushers to about 0.5 cm and then to about 1 mm with crushing rolls in closed circuit with a screen.

Storage of the crushed sample is important, since oxidation of the surfaces is to be avoided, especially with sulphide ores. Not only does oxidation inhibit collector adsorption, but it also facilitates the dissolution of heavy metal ions, which may interfere with the flotation process. Sulphides should be tested as soon as possible after obtaining the sample and ore samples must be shipped in sealed drums in as coarse a state as possible. Samples should be crushed as needed during the testwork, although a better solution is to crush all the samples and to store them in an inert atmosphere.

Wet grinding of the samples should always be undertaken immediately prior to flotation testing to avoid oxidation of the liberated mineral surfaces. Batch laboratory grinding, using ball mills, produces a flotation feed with a wider size distribution than that obtained in continuous closed-circuit grinding; to minimise this, batch rod mills are used which give products having a size distribution which approximates closely to that obtained in closed-circuit ball mills. True simulation is never really achieved, however, as overgrinding of high specific gravity minerals, which is a feature of closed-circuit grinding, is avoided in a batch rod mill. It is also important to understand the effect of grinding media on flotation especially where scale-up is sought (Greet et al., 2005).

A soft dense mineral, such as galena, will be ground finer in closed circuit than predicted by
the batch tests, and its losses due to production of ultra-fine particles may be substantial. Some sulphide minerals, such as sphalerite and pyrite, can be depressed more easily at the coarser sizes produced in batch grinding, but may be more difficult to depress at the finer sizes resulting from closed-circuit grinding. Predictions from laboratory tests can be improved if the mineral recovery from the batch tests is expressed as a function of mineral size rather than overall product size. The optimum mineral size can be determined and the overall size estimated to give the optimum grind size (Finch et al., 1979). This method assumes that the same fineness of the valuable mineral will give the same flotation results both from closed-circuit and batch grinding, irrespective of the differences in size distributions of the other minerals.

It must be appreciated that the optimum grinding size of the particles depends not only on their grain size but also on their floatability. Initial examination of the ore should be made to determine the degree of liberation in terms of particle size in order to estimate the required fineness of grind.

The potential for liberation of the minerals contained in the ore can be determined by characterising the grain sizes of the minerals present. This can be achieved by breaking the drill core samples at a relatively coarse size (typically about 600 microns). This preserves the in situ texture of the samples, including grain size, association, and shape. The texture can be characterised by using a scanning electron microscope configured as a mineral liberation analyser, such as the MLA (Figure 12.15) or the QEMSCAN, as discussed in Chapter 1. Such an analyser can measure the grain sizes and composition of the component minerals of the ore. An example of an MLA image is shown in Figure 12.16.

Testwork should then be carried out over a range of grinding sizes in conjunction with flotation tests in order to determine the optimum flotation feed size distribution. In certain cases, it may be necessary to overgrind the ore in order that the particles are small enough to be lifted by the air bubbles. If the mineral is readily floatable a coarse grind may be utilised, the subsequent concentrate requiring regrinding to further free the mineral from the gangue, before further flotation is performed to produce a high-grade concentrate.

Figure 12.15 An example of an automated mineral liberation analyser – the FEI–JKMRC MLA (Courtesy JKMRC and JKTech Pty Ltd)

Mineral surface analysis

A useful laboratory method is that of contact angle measurement (Laskowski, 1986; Ralston and Newcombe, 1992; Woods, 1994), where, in its simplest form, a clean smooth surface of mineral is placed in distilled water, and a bubble of air from the end of a capillary tube is pressed down upon it. If, after a short time, no adhesion is visible on withdrawal of the bubble, the mineral surface is assumed to be clean, and the collector is then added. If the mineral surface now becomes hydrophobic, adherence of the introduced bubble to the surface results. The contact angle produced across the water phase (Figure 12.2) is a measure of the floatability of the mineral. The method suffers from many disadvantages; it is extremely difficult to obtain a truly representative surface of the mineral of the required size (at least 0.5 cm²). The mineral may not be representative of the naturally liberated surface after the intense polishing required to produce
Froth flotation

Figure 12.16 An image from an MLA showing the mineral grains in a copper-gold ore. The particles are 100–200 μm in size. These images are usually presented in false colour with each colour denoting a mineral or mineral class (Courtesy JKMRC and JKTech Pty Ltd)

a completely clean, flat surface. The method is static, whereas true flotation is dynamic, particles adhering after impact with bubbles rising in the pulp. Contact angle measurements should, therefore, be regarded only as indicators of flotation response.

Several sophisticated analytical techniques are now available for measuring the condition of the mineral surface and the products formed when adding collectors. These can be used both for a fundamental understanding of the processes of surface modification by reagents and for diagnosing particular separation problems or opportunities. They include Time of Flight Mass Spectroscopy (TOF-SIMS) either as a separate technique or in combination with X-Ray photoelectron spectroscopy (XPS) (Piantadosi et al., 2000; Hart et al., 2005, Hope et al., 2005), Infrared External Reflection Spectroscopy (Mielczarski and Mielczarski, 2005), Spectroelectrochemical Raman studies (Goh et al., 2005), and molecular modelling and verification (Rao et al., 2005).

Microflotation tests

Initial floatability tests are often made on the liberated mineral particles, as a means of assessing a range of suitable collectors and regulators, and to determine the effective pH for flotation. In the Hallimond tube technique (Figure 12.17), dynamic conditions prevail. The mineral particles are held on a support of sintered glass inside the tube containing the distilled water and the collector under test. Air bubbles are introduced through the sinter and any hydrophobic mineral particles are

![Hallimond tube diagram]

**Figure 12.17** Hallimond tube
lifted by the bubbles, which burst at the water surface, allowing the particles to fall into the collecting tube. By treating a small weighed sample of pure mineral, or a mixture of pure minerals (e.g. galena and quartz), the weight collected in the tube can be related to the floatability. The Hallimond tube has the advantage of eliminating costly assaying. However, as frothers are not used in the test, it is doubtful whether the method truly simulates industrial flotation.

Other microflotation systems used to evaluate floatability on a microscale include those described by Partridge and Smith (1971), and the UCT microflotation cell (Bradshaw and O'Connor, 1996).

Batch flotation tests
The bulk of laboratory testwork is carried out in batch flotation cells (Figure 12.18), usually with 500 g, 1 kg, or 2 kg samples of ore. The cells are mechanically agitated, the speed of rotation of the impellers being variable, and simulate the large-scale models commercially available. Introduction of air to the cell is normally via a hollow standpipe surrounding the impeller shaft. The action of the impeller draws air down the standpipe, the volume being controlled by a valve and by the speed of the impeller. The air stream is sheared into fine bubbles by the impeller, these bubbles then rising through the pulp to the surface, where any particles picked up are removed as a mineralised froth.

Batch tests are fairly straightforward in practice, but a few experimental points are worth noting:

1. Agitation of the pulp must be vigorous enough to keep all the solids in suspension, without breaking up the mineralised froth column.
2. Conditioning of the pulp with the reagents is often required. This is a period of agitation, varying from a few seconds to 30 min, before the air is turned on, which allows the surface of the mineral particles to react with the reagents.
3. Very small quantities of frother can have marked effects, and stage additions of frother are often needed to control the volume of froth. The froth depth should be between 2 and 5 cm, as very shallow froths entail the risk of losing pulp into the concentrate container. Reduction of the amount of air is sometimes used to limit the amount of froth produced. This should be standardised for comparative tests in order to prevent the introduction of another variable.
4. As a matter of economics, flotation separations are carried out in as dense a pulp as possible consistent with good selectivity and operating conditions. The denser the pulp, the less cell volume is required in the commercial plant, and also less reagent is required, since the effectiveness of most reagents is a function of their concentrations in solution. The optimum pulp density is of great importance, as in general the more dilute the pulp, the cleaner the separation. Most commercial floats are in pulps of 25–40% solids by weight, although they can be as low as 8% and as high as 55%. It must be borne in mind that in batch flotation tests the pulp density varies continuously, from beginning to end, as solids are removed with the froth and water.

![Figure 12.18 Laboratory flotation cell](image-url)
is added to maintain the cell pulp level. This continuous variation changes the concentration of reagents as well as the character of the froth.

(5) As water contains dissolved chemicals which may affect flotation, water from the supply which will be used commercially should be used, rather than distilled water.

(6) Normally only very small quantities of reagent are required for batch tests. In order to give accurate control of their addition rates, they may have to be diluted. Water-soluble reagents can be added as aqueous solutions by pipette, insoluble liquid reagents by graduated dropper or hypodermic needle. Solids may either be emulsified or dissolved in organic solvents, providing the latter do not affect flotation.

(7) Recovery of froth is sensitive to operator technique.

(8) Most commercial flotation operations include at least one cleaning stage, in which froth is refloated to increase its grade, the cleaner tailings often being recycled. Since cleaner tails are not recycled in batch tests, they do not always closely simulate commercial plants. If cleaning is critical, cycle tests may have to be undertaken. These are multiple-step flotation tests designed to measure the effect of circulating materials. The main objectives of cycle tests are to determine:

- The increase in recovery obtained by recirculating cleaner tailings.
- The variation in reagent requirements to compensate for the circulating load of reagents.
- The effect of build-up of slimes or other undesirables which may interfere with flotation.
- The froth handling problems.

Normally at least six cycles are required before the circuit reaches equilibrium and a complete material balance should be made on each cycle. Since the reagents are in solution, it is essential that liquids as well as solids recirculate, so any liquid used to adjust pulp density must be circuit liquid obtained from decantation or filtration steps. Cycle tests are very laborious to carry out, and often the test fails to reach steady state. A method has been developed (Agar and Kipkie, 1978) whereby cycle test behaviour can be predicted from data developed from individual batch tests, and a computer program has been developed to arrive at a steady-state balance for a variety of simulated circuits.

Pilot plant testwork

Laboratory flotation tests provide the basis of design of the commercial plant. Prior to development of the plant, pilot scale testing is often carried out in order to:

(1) Provide continuous operating data for design. Laboratory tests do not closely simulate commercial plants, as they are batch processes.

(2) Prepare large samples of concentrate for survey by smelters, etc., in order to assess the possibility of penalties or bonuses for trace impurities.

(3) Compare costs with alternative process methods.

(4) Compare equipment performance.

(5) Demonstrate the feasibility of the process to non-technical investors.

Laboratory and pilot scale data should provide the optimum conditions for concentrating the ore and the effect of change of process variables. The most important data provided by testwork includes:

(a) The optimum grind size of the ore. This is the particle size at which the most economic recovery can be obtained. This depends not only on the grindability of the ore but also on its floatability. Some readily floatable minerals can be floated at well above the liberating size of the mineral particles, the only upper limit to size being that which the bubbles can no longer physically lift the particles to the surface. The upper size limit is normally around 300 \( \mu \text{m} \). The lower size limit for flotation, at which problems of oxidation and other surface effects occur, is around 5 \( \mu \text{m} \).

(b) Quantity of reagents required and location of addition points.

(c) Pulp density; important in determining size and number of flotation cells.
Flotation time; experimental data gives the time necessary to make the separation into concentrate and tailings. This depends on the particle size and the reagents used and is needed to determine the plant capacity.

Pulp temperature, which affects the reaction rates. Water at room temperature is, however, used for most separations.

The extent of uniformity of the ore; variations in hardness, grindability, mineral content, and floatability should be investigated so that variations can be accommodated in the design.

Corrosion and erosion qualities of the pulp; this is important in determining the materials used to construct the plant.

Type of circuit; many different types of circuit can be used, and laboratory tests should provide data for the design of the best-suited circuit. This should be as basic as possible at this stage. Many flow schemes used in operating plants have evolved over a long period, and attempted duplication in the laboratory is often difficult and misleading. The laboratory procedures should be kept as simple as possible so that the results can be interpreted into plant operation.

A key issue in pilot plant testing is flexibility and consistency of operation. A standardised pilot plant has recently been developed called the floatability characterisation test rig (FCTR). The unit described by Rahal et al. (2000) is a fully automated pilot plant which is designed to move from plant to plant and characterise the floatability of each plant’s ore according to standard procedures. It can be used both for testing modified circuits in existing plants and developing flowsheets for new ores. The FCTR is shown in operation in Figure 12.19.

**Basic flotation circuits**

Commercial flotation is a continuous process. Cells are arranged in series forming a bank (Figure 12.20). Pulp enters the first cell of the bank and gives up some of its valuable minerals as froth; the overflow from this cell passes to the second cell, where more mineralised froth is removed, and so on down the bank, until barren tailings overflow the last cell in the bank. In the case of flotation cells that use weir-type level control, the height of the froth column for each cell is determined by adjusting the height of the tailings overflow weir, the difference in height between this and the cell overflow lip determining the froth depth. In modern tank cells, pulp level is often maintained by adjusting the cell’s tailings discharge with a rubber sleeve pinch valve.

New feed enters the first cell of the bank, the froth column in the first few cells being kept high, since there are plenty of hydrophobic mineral particles to sustain it. The pulp level is raised from cell to cell, as the pulp becomes depleted in floatable minerals, by progressively raising the cell tailings weir height. The last few cells in the bank contain relatively low-grade froths, comprising weakly aerophilic particles. These are the scavengers, usually containing middlings particles, which are often recirculated to the head of the system.

In earlier cell designs, the scavenger cells, having little mineral to sustain a deep froth, have their tailings weirs raised so that pulp is almost overflowing the cell lip. This policy, which is used to remove all weakly floating material (“pulling the cells hard”), ensures maximum recovery from the bank of cells. Excessive circulating loads should be avoided, however, as the rougher feed may be diluted, and the flotation time reduced. In more recent cell designs, as the amount of minerals in the froth decreases (as in the scavenger cells described above) the froth is crowded using “froth crowders”. This design allows the cell to be operated with a slightly deeper froth.

The flowsheet for this basic system is shown in Figure 12.21. This flowsheet can be successfully operated only when the gangue is relatively unfloatable, and it requires extremely careful control to produce an even grade of concentrate if there are any fluctuations in the head grade. A preferable system is to dilute the concentrate from the first few cells of the bank, known as rougher concentrates, and refloat them in cleaner cells, where the level of the pulp is kept low to maintain a deep froth and produce a high-grade concentrate. In this rougher-scavenger-cleaner system (Figure 12.22), the cleaning cells receive a comparatively high-grade feed, whilst the scavenging section can be run with an excess of air so as to obtain maximum recovery. Tailings from the cleaner cells, usually containing aerophilic mineral particles, can be
recirculated to the rougher cells, along with the scavengers. This type of circuit, besides being useful for ores that need the maximum amount of aeration at the end of the bank to produce profitable recovery, is often employed when the gangue has a tendency to float and is difficult to separate from the mineral. In such cases, it may be necessary to utilise one or more recleaning banks of cells (Figure 12.23).

It is worth noting that the diluting water used to lower the pulp density of the cleaning bank passes to the roughing cells and dilutes the primary feed, which should therefore contain a correspondingly smaller portion of water as it leaves the grinding section in order that the dilution of the cleaner tailing may bring it to the correct pulp ratio in the roughing cells.

**Flowsheet design**

In designing a suitable flowsheet for a flotation plant, the primary grind size is of major consideration. This is mainly due to the fact that the flotation response depends on the level of liberation of the minerals in the ore. The target grind size can be estimated based on past experience and from mineralogical evaluation, but laboratory
grind-flotation tests must be conducted to determine optimum conditions. Grind size can be estimated knowing the size of the grains in the ore, and grain size can be estimated using mineral liberation analysers.

The purpose of the primary grind is to promote economic recovery of the valuable ore minerals. Batch tests are performed, utilising various reagent combinations, on samples of ore ground to various degrees. Incremental concentrate samples are weighed and assayed, and the results plotted as recovery–time and recovery–grade curves (Figure 12.24).

Initially, a grind size should be chosen which gives a reasonable rougher grade and recovery within an acceptable flotation time. If grinding is too coarse, some of the valuable mineral, locked in middlings grains, will not be floated. However, excessive flotation times may eventually allow some of these particles to report to the concentrate,
reducing its grade. It is here that the flotation engineers must use their experience in deciding what is, at this stage, a reasonable concentrate grade and flotation time.

As grinding is invariably the greatest single operating cost, it should not be carried out any finer than is justified economically. Later testwork, having improved on the basic flotation scheme, will modify the primary grind size, taking into account the amount of secondary grinding required to reach the specified concentrate grade, and the number of cleaning stages required. Finer grinding should not be performed beyond the point where the NSR for the increment saved becomes less than the operating cost (Steane, 1976).

After determining a suitable primary grind size (which may be modified in later testwork), further tests are performed to optimise reagent additions, pH, pulp densities, etc. Having optimised flotation recovery, testwork is then aimed at producing the required concentrate grade, and determining the flowsheet which must be utilised to achieve this.

As Figure 12.24(a) indicates, most of the valuable mineral floats within a few minutes, whereas it takes much longer for the residual small quantity to float. The rate equation for flotation can be expressed in a general way as follows:

\[ v = -dW/dt = K_n W^n \]  \hspace{1cm} (12.32)

where \( v \) (weight/unit time) is the flotation rate, \( W \) is the weight of floatable mineral remaining in the pulp at time \( t \), \( K_n \) is the rate constant, and \( n \) is the order of the reaction. The kinetics of flotation have been studied by many workers, the majority classifying flotation as a first order reaction \((n = 1)\), others reporting second order kinetics (Mori et al., 1985). Dowling et al. (1985) applied thirteen rate models to batch copper flotation data and evaluated the results using statistical techniques. The flotation of the copper ore was shown to be essentially a first order process, and all the models tested were found to give a reasonably good fit to the experimental data, though some models were clearly better than others.

The first order rate equation is usually expressed as (Lynch et al., 1981):

\[ R = 1 - \exp(-kt) \]  \hspace{1cm} (12.34)

where \( R \) is the cumulative recovery after time \( t \); \( k \) is the first order rate constant \((\text{time}^{-1})\); \( t \) is the cumulative flotation time.

Plots of \( \ln(1 - R) \) versus time should produce straight lines, but such plots are often concave upwards, which has led a number of workers to postulate the presence of fast and slow floating components. Agar (1985) argues that such postulates are false, the non-linear plots resulting from the assumption that the maximum possible recovery is 100%, whereas in practice some floatable material is usually totally irrecoverable, as it may be encased in gangue. A modified first order rate equation of the form:

\[ R = RI[1 - \exp(-kt)] \]  \hspace{1cm} (12.35)

is proposed, where \( RI \) is the maximum theoretical flotation recovery.

The flotation rate constant is dependent on the particle size and the degree of liberation of the mineral, the curve shown in Figure 12.24(a) being a summation of the flotation rates of all the particles within the ore. Figure 12.25 shows the variation of flotation rate constant of an ore as a function of the particle size. Extensive studies of
the influence of particle size on flotation have been made (Trahar and Warren, 1976; Hemmings, 1980; Trahar, 1981).

It is clear that the flotation activity of an ore falls off slowly towards the range of fine particle size, mainly due to the increase in number of particles per unit weight and to the deteriorating conditions for bubble–particle contact and effects such as increased surface oxidation of the particles. Flotation activity falls off very rapidly above the optimum particle size, due to the lesser degree of liberation of the minerals and to the decreasing ability of the bubbles to lift the coarse particles. It can be seen that the floated material is composed of a fast floating fraction in the medium-size range and a more reluctant fraction comprising unliberated coarse material and fines. In a commercial flotation circuit the fast floating material will be recovered in the roughing section, while the more reluctant fraction is recovered by scavenging, certain losses into the tailings having to be accepted. Figure 12.26 relates the distribution in terms of the flotation rate constant.

The essential difference between the concentrates from the roughers and scavengers is that the latter comprise both coarse and fine particles while the rougher concentrate consists essentially of an intermediate-size fraction.

The grade of the final cleaner concentrate is dependent on the grade of the rougher concentrate (Figure 12.24(b)) and, in order to reach the specified optimum cleaning grade, it is necessary to keep the rougher grade at a predetermined value. The decision as to where the rougher-scavenger split should be can be made on the basis of batch tests where cumulative concentrate grade is plotted against time (Figure 12.27), the time limit for rougher flotation then being fixed as that required to give a rougher concentrate with a high enough grade to produce the specified final concentrate grade with the chosen number of cleaning stages. The remaining flotation time (Figure 12.24(a)) is for scavenging, this time sometimes being reduced by increasing the severity of the flotation conditions (i.e., increased aeration, addition of more powerful collector) after the removal of the rougher concentrate.

Agar et al. (1980) have argued that the rougher-scavenger split should be made at the flotation time where separation efficiency (Equation 1.1) is maximised. Separation efficiency (SE) reaches a
maximum value when \( \frac{dSE}{dt} \) is zero, so that from Equation 1.3:

\[
\frac{dSE}{dt} = \frac{100m}{f(m-f)} \left[ (c-f) \frac{dC}{dt} + C \frac{dc}{dt} \right] \tag{12.36}
\]

\( = 0 \) at maximum separation efficiency.

\[
\int GdC = Cc \tag{12.37}
\]

Therefore, on differentiating Equation 12.37 with respect to \( t \):

\[
G \frac{dC}{dt} = C \frac{dC}{dt} + c \frac{dC}{dt} \tag{12.38}
\]

and substituting Equation 12.38 in Equation 12.36:

\[
\frac{dSE}{dt} = \frac{100m}{f(m-f)} \left[ c \frac{dC}{dt} - f \frac{dC}{dt} + G \frac{dC}{dt} - c \frac{dC}{dt} \right] \tag{12.39}
\]

Therefore, at maximum separation efficiency, where \( \frac{dSE}{dt} = 0 \), \( G = f \).

This means that at maximum separation efficiency, the grade of concentrate produced is equal to the flotation feed, and after this time the flotation system is no longer concentrating the valuable mineral.

Since separation efficiency = recovery of mineral—recovery of gangue (Equation 1.1), separation efficiency is also maximised when:

\[
\frac{d(Rm-Rg)}{dt} = 0, \quad \text{i.e. when} \quad \frac{dRm}{dt} = \frac{dRg}{dt}
\]

Therefore, at maximum separation efficiency, the rate of flotation of valuable mineral is equal to that of the gangue, and above the optimum flotation time the gangue begins to float faster than the mineral. This optimum flotation time can be calculated from the first order rate Equation (12.35). However, as shown by Agar (1985), this equation has to be modified for batch flotation tests to incorporate a correction factor for time. In batch flotation some hydrophobic solids will have air attached to them during the conditioning period, which causes them to float more rapidly than they would naturally. This causes a positive correction to time zero, as flotation actually started before the air flow was introduced. On the other hand, when air flow commences, several seconds elapse before a full depth of loaded froth is present in the cell, and this gives a negative correction to time zero. Agar’s modified rate equation for batch flotation tests is:

\[
R = RI \{ 1 - \exp \left[ -k(t + b) \right] \} \tag{12.40}
\]

where \( b \) is correction for time zero.

A plot of \( \ln \left[ \frac{RI - R}{RI} \right] \) versus \((t + b)\) should produce a line of slope \(-k\). However, \( RI \) and \( b \) are both unknown. Using experimental data, at the \( q \)th value of \( R \) and \( t \):

\[
\ln \left( \frac{RI - R_q}{RI} \right) + k(t_q + b) = r_q
\]

where \( r_q \) is the residual due to errors in the experimental data.

Hence

\[
r_q^2 = \left[ \ln \left( \frac{RI - R_q}{RI} \right) \right]^2 + k^2(t_q + b)^2 + 2k(t_q + b) \cdot \ln \left( \frac{RI - R_q}{RI} \right)
\]

Therefore, for \( n \) experimental data:

\[
\sum_{q=1}^{n} r_q^2 = \sum_{q=1}^{n} \left[ \ln \left( \frac{RI - R_q}{RI} \right) \right]^2 + k^2 \sum_{q=1}^{n} t_q^2 + nk^2 b^2 + 2k b \sum_{q=1}^{n} t_q \cdot \ln \left( \frac{RI - R_q}{RI} \right) + 2k b \sum_{q=1}^{n} \ln \left( \frac{RI - R_q}{RI} \right)
\]

\[
\sum_{q=1}^{n} r_q^2 \text{ is a minimum when } \frac{\partial}{\partial k} \left( \sum_{q=1}^{n} r_q^2 \right) \text{ and } \frac{\partial}{\partial b} \left( \sum_{q=1}^{n} r_q^2 \right) \text{ are zero,}
\]

i.e. when

\[
\frac{\partial}{\partial k} \left( \sum_{q=1}^{n} r_q^2 \right) = 2k \sum_{q=1}^{n} t_q^2 + 2nk b^2 + 4k b \sum_{q=1}^{n} t_q \cdot \ln \left( \frac{RI - R}{RI} \right) + 2b \sum_{q=1}^{n} \ln \left( \frac{RI - R}{RI} \right) = 0 \tag{12.41}
\]
Equations 12.42 and 12.43 can be solved to give:

\[
\hat{k} = -\frac{\left\{ \sum_{q=1}^{n} t \cdot \ln\left[ (RI - R) / RI \right] \right\}}{n \sum_{q=1}^{n} t^2 - \left( \sum_{q=1}^{n} t \right)^2}
\]

\[
\hat{b} = -\frac{\left\{ \sum_{q=1}^{n} t + \sum_{q=1}^{n} \ln\left[ (RI - R) / RI \right] \right\}}{nk}
\]  

RI can initially be assigned a value of 100, and \( \hat{k} \) and \( \hat{b} \) are calculated from Equations 12.44 and 12.45. Using these values \( \sum_{q=1}^{n} r^2 \) is then calculated from Equation 12.41.

\[
RI \text{ is then decremented and the procedure repeated until values of } \hat{k}, \hat{b}, \text{ and } \hat{RI} \text{ are found which minimise } \sum_{q=1}^{n} r^2.
\]

From Equation 12.40:

\[
dR / dt = RI \cdot k \exp[-k(t + b)]
\]

so that if the computations are performed for mineral (m) and gangue (g), then at optimum flotation time:

\[
RI_m k_m \exp[-k_m (t + b_m)] = RI_g k_g \exp[-k_g (t + b_g)]
\]

from which optimum flotation time

\[
\left[ \ln \frac{RI_m k_m}{RI_g k_g} - k_m b_m + k_g b_g \right] / (k_m - k_g)
\]  

In a complex flotation circuit, the rougher flotation may be divided into stages, each delivering its concentrate into the cleaning circuit at a different location according to its grade.

Thus, the basic flowsheet consisting of cleaners and recleaners may be supplemented by a low-grade cleaning circuit (Figure 12.28).

To ensure the recovery of the weakly aerophilic particles which are passed to the particular section of the cleaning plant, it is essential that the retention time in the cleaning stage is at least that of the corresponding roughing section.

![Figure 12.28 Complex flotation circuit](image-url)
Since the object of the scavenging section is to promote maximum recovery by minimising the losses to tailings, it is advisable to dimension the scavengers generously, to allow not only for the slow-floating character of the particles but also for fluctuations in the circuit. However, it is important to avoid excessive overloading of the system with large volumes of low-grade material, so a compromise must be made in designing the scavenger circuit (Lindgren and Broman, 1976). It may be preferable to have a lower rougher concentrate grade (longer flotation time), and more cleaning stages, thus reducing the volume of scavenger concentrate produced.

This is particularly important in non-metallic flotation, where, due to the generally low ratio of concentration, large circulating loads are often produced. For instance, in the flotation of a low-grade metallic ore, the ratio of concentration may be as high as 50, so that only about 2% of the ore is removed as concentrate, and the circulating loads in the system are of this order of magnitude. Non-metallic ores, however, are often of high grade, and the ratio of concentration can be as low as 2, so that 50% of the ore is removed as concentrate, and very high circulating loads are produced. Control of such circuits can often be facilitated by the addition of a thickener, or agitator, which can act as surge capacity for large changes in circulating load which may arise when changes in ore grade occur.

If cleaning does not give the required concentrate grade, regrinding of the rougher concentrate may be needed, usually being necessary to at least regrind the scavenger concentrate, and sometimes the primary cleaner tailings, before recirculation to the rougher circuit. The purpose of primary grinding is to promote maximum recovery, by rendering most of the valuable mineral floatable so that the bulk of the gangue can be discarded, thus reducing the amount of material that must be further processed. In secondary grinding, or regrinding, the major consideration is the grade of the concentrate.

Regrinding of the middlings products is common practice in flotation plants. Both the scavenger product and the cleaner tailings contain essentially a slow floating, fairly metal-rich fine fraction and a coarse product consisting mainly of unliberated mineral. These products are generally classified if the amount of fines is appreciable, after which the coarse product is reground and returned to the system with the new feed. The fine classified product is either recycled to the rougher circuit, or cleaned in a separate circuit to a grade high enough to be fed to the final concentrate or the main cleaner system.

Regrinding practice depends to a large extent on the ore mineralogy. In certain circumstances, particularly when the mineral is of high floatability and is associated with an unfloatable gangue, it may be economical to grind at a relatively coarse size and regrind the rougher concentrate (Figure 12.29). This is common practice with such minerals as molybdenite, which is readily floatable, when associated with hard, abrasive gangues. Removing the gangue as a tailings at a coarse particle size considerably reduces the energy consumption in the grinding stage.

![Figure 12.29 Regrinding of rougher concentrates](image)

Figure 12.30 shows a circuit used at the Phoenix Copper Div. of Granby Mining Corp., Canada (Hardwicke et al., 1978). The main copper mineral is chalcopyrite, some of which is finely disseminated in the gangue, but it also occurs as complex grains with pyrite. The circuit removes the fairly coarse chalcopyrite early by one-stage rougher-scavenger-cleaning. The middlings from this stage, consisting essentially of the finely disseminated copper minerals, are reground and floated in a completely separate circuit utilising two cleaning stages, thus isolating the flotation of the coarse material (80%–188 μm) from the flotation of the very fine particles (80%–40 μm).

Selective flotation circuits, which concentrate two or more minerals, must incorporate substantial facilities for control. Where, for example, heavy sulphide ores are being treated, it is common for a bulk float to be initially removed. This
isolates the sulphides from the non-sulphides, thus simplifying the subsequent selective separation of each sulphide component, providing that this is not inhibited by the presence of reagents from the bulk float which are adsorbed on the mineral surfaces. If this is serious, direct selective flotation must be used, which is essentially two or more one-product circuits in series, although some plants treating difficult ores use a combination of bulk and selective flotation in the rougher operations. Mineral composition and the degree of intergrowth of the valuable minerals are also important factors. Extremely fine intergrowth inhibits selective flotation separation, and there are some complex ores, containing sulphides of copper, lead, and zinc, which cause extreme difficulties in selective flotation. Figure 12.31 gives an outline of three flowsheets in use for such ores, from an “easy” coarse-grained ore (a) to a “difficult” fine grained ore (c).

Some flotation plants are in operation where more than five concentrates are effectively recovered from a single feed, such operations demanding considerable modification in the chemical nature of the feed pulp for each stage in the total treatment. The pH of the pulp may have to span a range from as low as 2.5 to as high as 10.5 to recover sulphide minerals alone, and further complications can arise if non-sulphide minerals, such as cassiterite, fluorite, barytes, etc., are to be recovered with sulphides.

Figure 12.32 shows a circuit which has been used to treat a complex ore containing copper, zinc, and iron sulphides, and cassiterite disseminated in a siliceous gangue.

A relatively coarse primary grind is used in order to recover as much cassiterite at as coarse a size as possible in the subsequent gravity process. After conditioning with copper sulphate to activate the sphalerite, the relatively large amount of sulphide minerals, which would interfere with the recovery of the cassiterite, is removed by bulk flotation at neutral pH. The bulk rougher concentrate is reground to release finely disseminated cassiterite, after which cleaning is undertaken, the cleaner tailings being recirculated to the head of the system. The bulk cleaner concentrate is conditioned with lime to about pH 11, which depresses the pyrite, and the copper and zinc sulphides are floated and cleaned, leaving the pyrite in the rougher tailings.

A significant problem in connection with flotation circuit design is that of transposing times from batch tests to flotation times in the continuous working circuit. The fundamental difference between a batch test and a continuous process is that every part of the flotation pulp in a batch test remains in the cell for the same length of time, whereas in a cell with continuous flow there is a spread, often quite considerable, in the retention times of different unit volumes. Some of the pulp takes a short cut and passes out of the cell relatively quickly, with the result that flotation is incomplete.
To reduce this problem the desired total flotation volume is divided into smaller portions.

The total cell volume required to give the specified nominal flotation time must, of course, be computed with allowance for the fact that only a part of the actual cell volume is available for the pulp. From the gross volume must be subtracted the volumes of the rotor machinery and stator, the froth layer and the air present in the pulp during the flotation process. Calculations indicate that the net volume in some cases can be as small as 50% of the gross cell volume. It must be remembered, however, that this factor only gives an adequate nominal retention time, without providing a safety margin for partial short-circuits in the flow as mentioned above or for the fluctuations that are liable to occur in the system. A safety factor of two to three is usually applied to laboratory flotation times in order to determine the required cell volume of the full-scale plant.

It should be noted that although increased aeration results in faster flotation, it also results in shorter retention times, as a larger portion of the total volume is occupied by air. There is, therefore,
an optimum rate of air supply to the cells, above which recovery may be reduced. This is not evident from the results of batch tests.

Flotation circuits can now be designed and optimised using computer modelling and simulation software, e.g. JKSimFloat (Harris et al., 2002). This simulator has the capabilities of predicting the flotation performance of a circuit under conditions of changing:

- Feed throughput (assuming that flotation floatability remains constant and residence time varies);
- Bank residence time;
- Cell operating parameter, e.g. air flow rate, froth depth, etc.;
- Circuit stream destination.

Numerous scenarios can be simulated quickly, providing the flotation design engineer with a tool for assessing the optimum circuit flowsheet.

**Circuit flexibility**

The decision having been reached to design a flotation circuit according to a certain scheme, it is necessary to provide for fluctuations in the flow rate of ore to the plant, both large and small, and for minor fluctuations in grade of incoming ore.

The simplest way of smoothing out grade fluctuations and of providing a smooth flow to the plant is by interposing a large storage agitator between the grinding section and the flotation plant:

Grind → Storage Agitator → Flotation Plant

Any minor variations in grade and tonnage are smoothed out by the agitator, from which material is pumped at a controlled rate to the flotation plant. The agitator can also be used as a conditioning tank, reagents being fed directly into it. It is essential to precondition the pulp sufficiently with the reagents before feeding to the flotation banks, otherwise the first few cells in the bank act as an extension of the conditioning system, and poor recoveries result.

Control systems can be installed to maintain the flow rate of the slurry as constant as possible. The control system starts in the grinding circuit where the feed rate of ore to the grinding mills is maintained constant using variable speed feeders. The level of slurry in pump boxes is also maintained constant by automatically adjusting pump speed using variable speed drives. Levels of slurry in flotation cells are maintained constant by using automatic cell level control systems.

Provision must be made to accommodate any major changes in flow rate which may occur; for example, a number of mills may have to be shut
Figure 12.33 Parallel flotation banks

down for maintenance. This is achieved by splitting the feed into parallel banks of cells (Figures 12.33 and 12.20(b)), each bank requiring an optimum flow rate for maximum recovery. Major reductions in flow rate below the designed maximum can then be accommodated by shutting off the feed to the required number of banks. The optimum number of banks required will depend on the ease of control of the particular circuit. More flexibility is built into the circuit by increasing the number of banks, but the problems of controlling large numbers of banks must be taken into account, and in plants that have installed very large unit processes, such as grinding mills, flotation machines, etc., in order to reduce costs and facilitate automatic control, the need for many parallel banks has been reduced.

In designing each flotation bank, the number of cells required must be assessed: should a few large cells be incorporated or many small cells giving the same total capacity?

This is determined by many factors. If a small cell in a bank containing many such cells has to be shut down, then its effect on production and efficiency is not as large as that of shutting down a large cell in a bank consisting of only a few such cells. Maintenance costs, however, tend to be lower with large cells, since there are relatively fewer parts to change in a particular bank.

The desired residence time for maximum economic recovery, which is calculated from laboratory tests, assumes that every particle is given the chance to float in that time. This does not necessarily happen in a continuous process, as it is possible for particles to short-circuit immediately from one cell to the next. This becomes increasingly serious when there are the fewer cells in the bank. Designing a bank with many small cells gives particles which have short-circuited in one or more cells the chance to float in succeeding cells. The designer, therefore, must decide between small cells for greater flexibility and metallurgical performance, and large cells, which have a smaller total capital cost, less floor area per unit volume, and lower operating costs. In eastern Europe, it has been common to install 30 or more machines in a single bank, while in the West the trend is to install very large cells, especially in the roughing stage.

Flotation plants built in the 1970s and 1980s used between eight and fourteen cells in the rougher banks to produce an optimum design, depending on the most economic layout of the plant. This had the effect of limiting the use of 28 m$^3$ (1000 ft$^3$) cells to mills with tonnage throughputs of 15,000 t d$^{-1}$ or higher, although some machine manufacturers, particularly Outokumpu, recommend using the largest cells possible, which reduces the number of mechanisms, in some cases to only two in a bank. There are reports that recovery is unimpaired, or even enhanced, at the same total retention time. As Young (1982) observed, a clear difference of opinion has emerged, which requires further research.

In the 1990s and 2000s the flotation cell suppliers that produced Outokumpu and Wemco cells developed the tank cell designs. These cells can be as big as 150 m$^3$ or more in volume and can treat more that 100,000 tonnes per day of ore, particularly in large copper operations in South America and Asia (Figure 12.20(b)). Tank cells of even greater volumes are under development from flotation cell suppliers (Weber et al., 2005). These cells are able to minimise short-circuiting of slurry by using big tanks with a single tailings discharge that is controlled by rubber sleeved pinch valves. A
bank is typically designed with about eight to ten cells in the rougher section.

Flexibility must be provided relative to the number of cells in a bank producing rougher and scavenger concentrates, in order to allow for changes in the grade of incoming ore. For instance, if the ore grade decreases, it may be necessary to reduce the number of cells producing rougher concentrate, in order to feed the cleaners with the required grade of material. A simple method of adjusting the “cell split” on a bank is shown in Figure 12.34. If the bank shown has, say, twenty cells, each successive four cells feeding a common launder, then by plugging outlet B, twelve cells produce rougher concentrate. Similarly, by plugging outlet A, only eight produce rougher concentrates, and by leaving both outlets free, a ten–ten cell split is produced.

At North Broken Hill in Australia, the lead recleaner concentrate grade was automatically controlled by stabilising the mass flow rate of recleaner feed. An increase in flow rate increased the cleaner concentrate grade, due to the decreased residence time within the cells. Automatically controlled froth diverter trays (Figure 12.35) increased the number of cells producing concentrate to compensate for the increase in feed rate, and the number of cells producing middlings was correspondingly reduced (Figure 12.36).

**Flotation machines**

Although many different machines are currently being manufactured and many more have been developed and discarded in the past, it is fair to state that two distinct groups have arisen: pneumatic and mechanical machines. The type of machine is of
great importance in designing a flotation plant and is frequently the characteristic causing most debate (Araujo et al., 2005; Lelinski et al., 2005).

Pneumatic machines either use air entrained by turbulent pulp addition (cascade cells), or more commonly air either blown in or induced, in which case the air must be dispersed either by baffles or by some form of permeable base within the cell. Generally pneumatic machines give a low-grade concentrate and little operating trouble. Since air is used not only to produce the froth and create aeration but also to maintain the suspension and to circulate it, an excessive amount is usually introduced and for this and other reasons they have been little used.

One of the early developments in the pneumatic field was the Davcra cell (Figure 12.37), which has been claimed to yield equivalent or better performance than a bank of mechanical machines.

![Davcra cell](image)

The cell consists of a tank segmented by a vertical baffle. Air and feed slurry are injected into the tank through a cyclone-type dispersion nozzle, the energy of the jet of pulp being dissipated against the vertical baffle. Dispersion of air and collection of particles by bubbles allegedly occurs in the highly agitated region of tank confined by the baffle. The pulp flows over the baffle into a quiescent region designed for bubble–pulp disengagement. The cell can be used for roughing or cleaning applications on a variety of minerals. Although not widely used, Davcra cells replaced some mechanical cleaner machines at Chambishi copper mine in Zambia, with reported lower operating costs, reduced floor area, and improved metallurgical performance.

A significant development in recent years has been the increasing industrial use of flotation columns. The main advantages of columns include improved separation performance, particularly on fine materials, low capital and operational cost, less plant space demand, and adaptability to automatic control.

A typical configuration of a column is shown in Figure 12.38. It consists of two distinct sections. In the section below the feed point (the recovery section), particles suspended in the descending water phase contact a rising swarm of air bubbles produced by a sparger (Murdock and Wyslouzil, 1991) in the column base. Floatable particles collide with and adhere to the bubbles and are transported to the washing section above the feed point. Non-floatable material is removed from the base of the column as tailing. Gangue particles that are loosely attached to bubbles or are entrained in bubble slipstreams are washed back into the recovery section, hence reducing contamination of the concentrate. The wash water also serves to suppress the flow of feed slurry up the column towards the concentrate outlet. There is a downward liquid flow in all parts of the column.
preventing bulk flow of feed material into the concentrate.

Columns were originally developed in Canada, and were used initially for cleaning molybdenum concentrates. Two-column flotation units were installed in the molybdenum circuit at Mines Gaspe, Canada, in 1980, and excellent results were reported (Cienski and Coffin, 1981). The units replaced mechanical machines in the cleaner banks. Since then many of the copper–molybdenum producers in North America have installed columns for molybdenum cleaning, and their use has been expanded into the roughing, scavenging, and cleaning of a variety of ore types, in many parts of the world. An indication of the interest in columns is that within a period of three years, they have been the subject of two international conferences (Sastry, 1988; Agar et al., 1991), a textbook (Finch and Dobby, 1990), and many other papers (Araujo et al., 2005).

The US Bureau of Mines compared column flotation with conventional flotation on a Montana chromite ore, the results showing that column flotation appears to be a physical improvement in the flotation separation process (McKay et al., 1986). Because of the excellent results achieved, further studies of column flotation were underway on ores containing fluorite, manganese, platinum, palladium, titanium, and other minerals. The United Coal Co. in the United States has also pioneered the use of columns for the flotation of fine coal (Chironis, 1986). It is possible that, due to their froth washing capability, columns may find an increasing use in the future for treating ores that need extensive fine grinding, followed by de-sludging and multi-stage cleaning.

Instrumentation and some degree of automatic control is a necessity for column operation. The methods currently used for the control of columns have been summarised by Moys and Finch (1988).

Flotation columns are usually about 12 m high, with diameters of up to about 3.5 m (round or square, the former being more popular), the importance of height/diameter ratio having been discussed by Yianatos et al. (1988). Several attempts have been made to develop column-type devices with much smaller height/diameter ratios, the Jameson cell (Kennedy, 1990; Cowburn et al., 2005) being a successful example (Figure 12.39 – Harbort et al., 2002). Contact between the feed and the air stream is made in a mixing device at the top of a vertical downcomer. The air–liquid mixture flows downwards to discharge into a shallow pool of pulp in the bottom of a short cylindrical column. The bubbles disengage and rise to the top of the column to overflow into a concentrate launder, while the tails are discharged from the bottom of the vessel. The main advantages of the device are that the overall height of the column is reduced to about 1 m, and the flotation column can be self-inducing with respect to the air supply.

The Jameson cell was developed jointly by Mount Isa Mines Ltd and the University of Newcastle, Australia. The cell was first installed for cleaning duties in base metal operations (Clayton et al., 1991; Harbort et al., 1994) but it has also found uses in other duties including roughing and preconcentrating. The major advantage of the cell is its ability to produce clean concentrates in one stage of operation. It also has a novel application in solvent extraction – electrowinning, where it is being used to recover entrained organic from copper-rich electrolyte (Miller and Readett, 1992) in many of the copper-leaching operations in Arizona and New Mexico in the United States as well as in Mexico.

The Jameson cell has also become widely used in the coal industry in Australia in the 1990s and 2000s. Typical cell layout is shown in Figure 12.40, which shows the fine coal slurry feeding a central distributor, splitting the stream and being treated...
inside the downcomer. The clean coal then overflows as a concentrate from the separation chamber. Froth separators were developed in the USSR in 1961 and had treated 9 M tonnes of various ores by 1972 (Malinovskii et al., 1973). The principle of the froth separator (Figure 12.41) is that conditioned feed is discharged onto the top of a froth bed, the hydrophobic particles being retained while the hydrophilic species pass through and are thereby separated. This method is particularly suited to the separation of coarse particles. The slurry is introduced at the top of the machine and descends over sloping baffles before entering an aeration trough, where it is strongly aerated before floating horizontally onto the top of the froth bed. Water and solids which penetrate the froth bed pass between aerator pipes into the pyramidal tank. The aerators are rubber pipes with 40–60 fine holes per cubic centimetre, into which air is blown at 115 kPa. The machine, which has two froth discharge lips, each 1.6 m long, is capable of treating 50 t h⁻¹ of solids at slurry densities of between 50 and 70% solids. Although used little in the western world, they have great potential for treating coarse feeds at up to ten times the rate of mechanical machines. The upper size limit for flotation is increased to about 3 mm, but they are not suited to fines treatment, a typical feed size range being about 75 μm to 2 mm.

The role of flotation time is reversed, increasing flotation times reducing the recovery but increasing the concentrate grade.

Mechanical flotation machines are the most widely used, being characterised by a mechanically driven impeller which agitates the slurry and disperses the incoming air into small bubbles. The machines may be self-aerating, the depression created by the impeller inducing the air, or “supercharged” where air is introduced via an external blower. In a typical flotation bank, there are a number of such machines in series, “cell-to-cell” machines being separated by weirs between each impeller, whereas “open-flow” or “free-flow” machines allow virtually unrestricted flow of the slurry down the bank.

The most pronounced trend in recent years, particularly in the flotation of base metal ores, has been the move towards larger capacity flotation cells, with corresponding reduction in capital and operating costs, particularly where automatic control is incorporated. In the mid-1960s, flotation cells were commonly 5.7 m³ (200 ft³) in volume, or less (Figure 12.42) and in the 1970s and 1980s 8.5 m³ to 14.2 m³ cells were widely used (Figure 12.43), with 28.3 m³ cells, and larger, being increasingly adopted. Manufacturers in the forefront of this industry included Denver Equipment (36.1 m³), Galigher (42.5 m³), Wemco (85 m³), Outokumpu Oy (38 m³), and Sala (44 m³). Many
other machines were manufactured, however, and were comprehensively reviewed by Harris (1976) and Young (1982).

As described above, in the 1990s and 2000s flotation cell suppliers developed the tank cell designs (Figures 12.20(b), 12.44). These cells are currently 150 m³ in volume but bigger cells are on the drawing boards (Weber et al., 2005). The cells are circular in shape, fitted with froth crowders, multiple froth launders and discharge and designed with effective level control systems.

In the 1970s most of the flotation machines were of the "open flow" type, as they were much better suited to high throughputs and are easier to maintain than cell-to-cell types. The Denver "Sub-A" was perhaps the most well-known cell-to-cell machine, having been used widely in the past in small plants, and in multi-stage cleaning circuits, where the pumping action of the impellers permitted the transfer of intermediate flowed without external pumps. They were manufactured with cell sizes of up to 14.2 m³, and were used mostly as coal-cleaning devices, where the users reported a significant improvement in selectivity over open-flow designs.

The flotation mechanism is suspended in an individual square cell separated from the adjoining cell by an adjustable weir (Figure 12.45). A feed pipe conducts the flow of pulp from the weir of the preceding cell to the mechanism of the next cell, the flow being aided by the suction action of the impeller. The positive suction created by the impeller draws air down the hollow stand-pipe surrounding the shaft. This air stream is sheared into fine bubbles by the impeller action and is intimately mixed with the pulp which is drawn into the cell onto the rotating impeller. Directly above the impeller is a stationary hood, which prevented "sanding-up" of the impeller if the machine is shut down. Attached to this hood are four baffle vanes, which extend almost to the corners of the cell. These prevented agitation and swirling of the pulp above the impeller, thus producing a quiescent zone where bubbles can ascend with their mineral load without being subjected to scouring
which may cause them to drop it. The mineral-laden bubbles separate from the gangue in this zone and pass upward to form a froth. As the bubbles move to the pulp level, they are carried forward to the overflow lip by the crowding action of succeeding bubbles, and quick removal of froth is accomplished by froth paddles which aid the overflow.

Pulp from the cell flowed over the adjustable tailings weir, and was drawn on to the impeller of the next cell where it was again subjected to intense agitation and aeration. Particles which are too heavy to flow over the tailings weir are by-passed through sand relief ports, which prevented the build-up of coarse material in the cell.

The amount of air introduced into the pulp depends upon the impeller speed, which is normally in the range of 7–10 ms⁻¹ peripheral. More air may be obtained by increasing the impeller speed, but this may in certain circumstances overagitate the pulp as well as increase impeller wear and energy consumption. In such cases, supercharging may be applied by introducing additional air down the stand-pipe by means of an external blower.

Supercharging is required with the Denver D–R machine (Figures 12.43 and 12.46), which ranges in
Figure 12.44 160 m$^3$ Outokumpu tank cell (Courtesy JKMRC and JKTech Pty Ltd)

Figure 12.45 Denver sub-aeration cell

size from 2.8 m$^3$ to 36.1 m$^3$, and which was developed as a result of the need for a machine to handle larger tonnages in bulk-flotation circuits. These units are characterised by the absence of intermediate partitions and weirs between cells. Individual cell feed pipes have been eliminated, and pulp is free to flow through the machine without interference. The pulp level is controlled by a single tailings weir at the end of the trough. Flotation efficiency is high, operation is simple, and the need for operator attention is minimised. Most high-tonnage plants use a free-flow type of flotation machine and many are equipped with automatic control of pulp level and other variable factors.

A widely used flotation machine is the Wemco Fagergren (Figures 12.47 and 12.48) manufactured in sizes up to 85 m$^3$. The modern 1 + 1 design consists of a rotor-disperser assembly, rather than an impeller, and the unit usually comprises a long rectangular trough, divided into sections, each containing a rotor-disperser assembly. The feed enters below the first partition, and tails go over partitions from one section to the next, the pulp level being adjusted at the end tailings weir.

Pulp passing through each cell, or compartment, is drawn upwards into the rotor by the suction created by the rotation. The rotor also draws air down the standpipe, no external blower being needed. The air is thoroughly mixed with the pulp before being broken into small, firm bubbles by the disperser, a stationary, ribbed, perforated band encompassing the rotor, by abruptly diverting the whirling motion of the pulp.

Perhaps the most well known of the supercharged machines is the Galigher Agitair (Sorensen, 1982) (Figure 12.49). This system, again, offers a straight-line flow of pulp through a suitably proportioned row of cells, flow being produced by a gravity head. The Agitair machine is often used in large-capacity plants. In each compartment, which may be up to 42.5 m$^3$ in volume, is a separate impeller rotating in a stationary baffle system. Air is blown into the pulp through the hollow standpipe surrounding the impeller shaft, and is sheared into fine bubbles, the volume of air being controlled separately for each compartment. Pulp depth is controlled by means of weir bars or dart valves at the discharge end of the bank, while the depth of froth in each cell can be controlled by varying the number and size of froth weir bars provided for each cell. Agitair machines produce copious froths and have found favour in mills handling ores of poor floatability, which require large froth columns to help weakly aerophilic particles to overflow.

The Wemco cell has also experienced significant design change in the late 1990s and 2000s. The cell still uses the same rotor design as the Wemco Fagergren cell but the rotor is now inside a new Smart Cell tank design (Figures 12.50 and 12.51).

Outokumpu Oy has operated several base metal mines and concentrators in Finland and elsewhere, and is well known for its mineral processing equipment including its range of OK flotation cells.
The OK impeller differs markedly from that of most other machines. It consists of a number of vertical slots which taper downwards, the top of the impeller being closed by a horizontal disc (Figure 12.52). As it rotates, slurry is accelerated in the slots and expelled near the point of maximum diameter. Air is blown down the shaft and the slurry and air flows are brought into contact with each other in the rotor-stator clearance, the aerated slurry then leaving the mechanism to the surrounding cell volume. The slurry flow is replaced by fresh slurry which enters the slots near their base where the diameter and peripheral speed are less. Thus the impeller acts as a pump, drawing in slurry at the base of the cell, and expelling it outwards. The tank cell design and the rotor design minimise short-circuiting, as pulp flow is towards the bottom of the
cell and the new feed entering is directed towards the mechanism due to the suction action of the rotor. It is because of this that banks containing only two large cells are now in use in many of the world’s concentrators (Niitti and Tarvainen, 1982).

The Sala AS series of flotation cells, ranging in size from 1.2 to 44 m³, differs in design from the machines previously described. Most machines are designed to promote ideal mixing conditions, the vertical flows achieved maintaining solids in suspension. The Sala design (Figure 12.53) minimises vertical circulation, the manufacturers claiming that the natural stratification in the slurry is beneficial to the process. The impeller is positioned under a stationary hood which extends out to, and supports, the stationary diffuser. The impeller is a flat disc with vertical radial blades on both surfaces, the upper blades expelling air which is blown down the standpipe, and the lower blades expelling slurry from the central base area of the tank, all slurry flowing into the impeller being from below. The aerated slurry is then expelled through the conventional circular stator. Although the impeller has an unusually large diameter in relation to the rather shallow cell size, this preventing sanding in the corners, it is claimed that the air is dispersed into very closely sized fine bubbles, which are particularly suited to fines flotation. The machines are used to treat a variety of materials, including base metals, iron ore, coal, and non-metallic minerals.

**Comparison of flotation machines**

Selection of a particular type of flotation machine for a given circuit is usually the subject of great debate and controversy (Araujo et al., 2005; Lelinski et al., 2005).

The main criteria in assessing cell performance are:

1. metallurgical performance, i.e. product recovery and grade
2. capacity in tonnes treated per unit volume
3. economics, e.g. initial costs, operating and maintenance costs.

In addition to the above factors, less tangible factors, such as the ease of operation and previous experience of personnel with the equipment, may contribute.

Direct comparison of cells is by no means a simple matter. Although comparison of different cell types, such as mechanical against pneumatic, should be based on metallurgical performance
when testing the same pulp in parallel streams, even here results are suspect; much depends on the operator’s skill and prejudices, as an operator trained on one type of cell will prefer it to others.

In general, the differences between mechanical machines are small and selection depends a lot on personal preference. One of the basic problems that hampers comparison of flotation cells is that a cell is required to perform more than a selective collection operation; it is required to deliver the collected solids to a concentrate product with minimal entrainment of pulp. The observed rate of recovery from a cell can be dependent on the froth-removal rate, which in turn can be affected by such process variables as reagent additions, impeller speed, position of the pulp-froth interface and aeration. Research has shown that bubble size
Rotation
Impeller (cut away)

Figure 12.53 Sala flotation mechanism

in mechanical machines of all designs is between 0.1 and 1 mm, the size being controlled mainly by the frother. The machine stator does not change the bubble size, but only the flow pattern in the cell (Harris, 1976). Machine suppliers recommend impeller speeds that allow the machine to maintain the particles in suspension and disperse the bubbles throughout the cell.

Mechanical and conventional pneumatic machines have been used for many decades, whereas froth separators and columns are fairly recent developments. Mechanical machines have been the dominant type, pneumatic machines, apart from the Davcra, now rarely being seen, except in a few old concentrators. Columns now have a role in flotation but froth separators have, as yet, to gain wide acceptance.
Although there is very little information on pneumatic machines, Arbiter and Harris (1962) reported comparative tests on a number of mechanical and pneumatic cells showing the former to be generally superior. Gaudin (1957) suggested that mechanical machines are better suited to difficult separations, particularly where fines are present. The impellers tend to have a scouring effect which removes slimes from particle surfaces. An American survey of flotation columns was reported by Clingan and McGregor (1987). All the production columns surveyed were in use as cleaners or scavengers. All operators indicated that improved metallurgical performance was part of their justification for using flotation columns and most indicated operating-cost savings and ease of operation.

An analysis of the effectiveness of the various types of flotation machine has been made by Young (1982), who discusses machine performance with regard to the basic objectives of flotation, which are the recovery of the hydrophobic species into the froth product, and, at the same time, achieving a high selectivity by retaining as much of the hydrophilic species as possible in the slurry. Achievement of recovery is dependent on the mechanism of particle–bubble attachment, which may be by “coursing bubble” contact between an ascending bubble and a falling particle, by precipitation of dissolved gas onto a particle surface, or by contact between a particle and an unstable “nascent” bubble in a pressure gradient. Coursing bubble attachment requires non-turbulent conditions, which is not found in mechanical or Darcra cells. A mechanical impeller can be compared to a turbine operating in a cavitating mode, air bubbles forming on the trailing, low pressure side of the impeller blades, while slurry flows are concentrated mainly upon the leading, high pressure side. The air and slurry flows are therefore separated to some extent, and the possibility of air precipitation on particles and for contact between particles and nascent bubbles is low. Mechanical impellers, therefore, do not appear to be the ideal device for particle–bubble contact, and the nozzle in the Darcra cell would appear to be much more efficient, which may explain why the Darcra cell can give the same recovery as a short bank of mechanical cells.

The particle–bubble contact in column machines is by coursing bubble only, and these are ideal displacement machines, whereas the mechanical cells are ideal mixers. The more favourable conditions for particle–bubble attachment, together with a lower tendency to break established bonds, may account for the high recoveries reported from flotation columns.

As selectivity is reduced by slurry turbulence, it is clear that the column machines, which also improve selectivity by froth washing, have an advantage over the mechanical machines. The froth separators, however, are not well suited to achieving selectivity from fine feeds, as the fine hydrophilic particles must descend through the total froth bed to report to the tailings, and this is difficult to achieve.

As Young concludes, mechanical flotation machines dominate the Western industry, but the reasons for this may be more due to commercial realities than to design excellence. The major Western manufacturers make only this type, and many flotation engineers are familiar with no other. However, in the future, the mechanical machines will no doubt encounter the increasing challenge of other types, and there is no reason why a number of different units could not be installed in concentrators for specific duties.

**Electroflotation**

Industrial flotation is rarely applied to particles below 10 μm in size due to lack of control of air bubble size. With ultra-fine particles, extremely fine bubbles must be generated to improve attachment. Such bubbles can be generated by in situ electrolysis in a modified flotation cell. Electroflotation has been used for some time in waste-treatment applications to float solids from suspensions. Direct current is passed through the pulp within the cell by two electrodes, generating a stream of hydrogen and oxygen bubbles. Considerable work has been done on factors affecting the bubble size on detachment from the electrodes, such as electrode potential, pH, surface tension and contact angle of the bubble on the electrode. On detachment, the majority of bubbles are in the 10–60 μm range, and bubble density can be controlled by current density to yield optimum distribution of ultra-fine bubbles as well as adequate froth control. Conventional flotation processes produce bubbles ranging from 0.6 to 1 mm in diameter and there is considerable variation in bubble size.
Some other factors have also been noted in addition to the fine bubbles. For example, the flotation of cassiterite is improved when electrolytic hydrogen is used for flotation. This may be due to nascent hydrogen reducing the surface of the cassiterite to tin, allowing the bubbles to attach themselves.

Although the main applications of electroflotation are in sewage treatment, this technique is capable of selectively floating solids and has been used in the food industry. It may have a future role in the treatment of fine mineral particles (Venkatachalam, 1992).

Agglomeration-skin flotation

In agglomeration flotation, the hydrophobic mineral particles are loosely bonded with relatively smaller air bubbles to form agglomerates, which are denser than water but less dense than the particles wetted by the water. Separation of the agglomerated particles is achieved by flowing film gravity concentration. When the agglomerates reach a free water surface, they are replaced by skin-floating individual particles. In skin-flotation, surface tension forces result in the flotation of the hydrophobic particles, while the hydrophilic particles sink, effecting a separation.

In table flotation, the reagentised particles are fed onto a wet shaking table. The pulp is diluted to 30–35% solids and is aerated by air jets from a series of drilled pipes arranged above the deck, at right-angles to the riffles, in such a way that the holes are immediately above the material carried by the riffles. The hydrophobic particles form aggregates with the air bubbles and float to the water surface, from where they skin-float to the normal "tailings" side of the table. The wetted particles become caught in the riffles and discharge at the end of the table where the concentrate normally reports in most shaking table gravity-separation processes.

With table flotation, and other agglomeration processes, it is desirable to film and float the most abundant mineral, if possible, as the capacity of the table is limited to the amount of material that can be carried along the riffles. This is the reverse of the ideal conditions for froth flotation, where it is desirable to film and float the mineral that is least abundant, so as to reduce entrainment of unwanted material to the minimum. This difference renders table flotation most suitable for removing sulphide minerals from pyritic tin ore concentrate, or for the concentration of non-metallic minerals, such as fluorite, barite, and phosphate rock. Such minerals are often liberated at sizes too coarse for conventional flotation. Agglomeration separations are possible over a wide range of sizes, usually from a maximum of about 1.5 mm in diameter to 150 μm. Minerals with low specific gravities, such as fluorite, have been separated at sizes of up to 3 mm.

Table flotation was used until relatively recently in the treatment of coarse phosphate rock, but has been replaced by similar methods utilising pinched sluices and spirals as the flowing film devices (Moudgil and Barnett, 1979).

Flotation plant practice

The ore and pulp preparation

It is inevitable that there will be changes in the character of the ore being fed to a flotation circuit. There should therefore be means available for both observing and adjusting for such changes. Variations in the crystal structure and intergrowth may have an important effect on liberation and optimum grind size. Change in the proportion of associated minerals is a very common occurrence and one which may be largely overcome by blending the ore both before and after crushing has been completed. When the feed is high grade it is relatively easy to produce a highly mineralised froth and high-grade concentrate; when it is low grade it is harder to maintain a stable froth and it may be necessary to switch one of the final cleaning cells to a lower-grade section if cells and launders have suitable flexibility.

Fluctuation in the nature and proportion of minerals in the run-of-mine ore inevitably occurs when ores are drawn from more than one location, and the variation observed may be further accentuated by partial oxidation of the ore. This may occur from geological changes or from delayed transportation of broken ore from the stope in the mine to the processing plant. Oxidation also commonly occurs as a result of overlong storage in stockpiles or ore-bins and therefore it is necessary to determine how prone a particular ore is to oxidation and to ensure that the holding time is kept well below a
critical level. Oxidised ores are softened by lattice decomposition and become more inert to collector reagents and more prone to overgrinding.

Wet grinding is the most important factor contributing to the performance of the flotation circuit. It is therefore of vital importance that the grinding circuit shall provide a reliable means of control as a guarantee that the milled product will allow maximum liberation of the values. In the comminution section of the plant, poor operation in the crushing stage can be offset in grinding. There is, however, no way of offsetting poor grinding practice and it is wise to use experienced operators on this section.

The degree of grinding required is determined by testwork and removal of free mineral at the coarsest possible size is always desirable. Modern flotation takes this into account, as is evidenced by the general trend towards floating the mineral in stages: first coarse, then fine. The advantages of floating a mineral as coarse as possible include:

- lower grinding costs;
- increased recovery due to decreased slime losses;
- fewer overground particles;
- increased metallurgical efficiency;
- less flotation equipment;
- increased efficiency in thickening and filtration stages.

Laboratory control of grinding must be carried out on a routine basis, by screening and assaying the tailings in order to determine the losses in each size fraction and the reason for their occurrence. It is often found that the largest losses occur in the coarsest particles, due to inadequate liberation, and if grinding all the ore to a finer size improves recovery economically then it should be done. Often the losses occur in the very fine "sub-sieve" fractions, due to overgrinding of heavy mineral values. In this case it may be necessary to "scalp" the grinding circuit and remove heavy minerals, which are returned to the mill by the classifier at sizes below optimum grind. This can be done by adding flotation reagents to the mill discharge and removing fine liberated minerals by a unit flotation cell between this and the classifier (Figure 12.54). Apart from the advantages of reducing overgrinding, the density of the mill discharge can be controlled so as to give optimum flotation efficiency. In conventional circuits the mill discharge density is controlled according to the cyclone requirements, the cyclone overflows often requiring dewatering before feeding flotation.

Figure 12.54 Removal of fine heavy minerals from the grinding circuit by unit flotation cell

Flotation within the grinding circuit, particularly of heavy, coarse lead minerals, is performed at several concentrators and the aim of Outokumpu's flash flotation method is to recover such coarse valuable minerals which would normally be recycled to the grinding circuit via classification (Warder and McQuie, 2005). The concentrate produced is final concentrate, needing no further cleaning, and is produced in a specially designed flotation machine, _Skim-air_, which removes the coarse floatable particles while allowing the others to return to the mill for further grinding, thus reducing the amount of valuable mineral lost to fines and increasing the average particle size of the final concentrate. A number of these cells have been installed in Finnish concentrators, with considerable benefits (Anon., 1986b).

As was shown earlier, if the mineral is readily floatable, and is associated with a relatively non-floatable gangue, it may be more economical to produce a coarse final tailings and regrind the resulting low-grade rougher concentrate, which may then be considered almost as a middlings product (Figure 12.29). The secondary, or regrind operation, treating only a small percentage of the original fine ore feed, can therefore be carried out.
to a size fine enough to liberate. Subsequent flotation then produces the maximum recovery/grade results in the greatest economic return per tonne of ore milled. There is, of course, an upper limit on size at which flotation may be practised effectively, due to the physical limitations of the bubble in lifting coarse particles. Whilst it may be argued that factors such as shape, density, and aerophilic properties may be influential, the practical upper limit rarely exceeds 0.5 mm and is usually below 0.3 mm. For a wide variety of minerals, reagents, and flotation machines, recovery is greatest for particles in the size range 100–10 μm (Trahar and Warren, 1976; Trahar, 1981). Below about 10 μm recovery falls steadily. There is no evidence of a critical size below which particles become unfloatable. The reason for the difficulties experienced in selective flotation of fine particles is not fully understood and varies from ore to ore. Very fine particles have relatively high surface area in relation to mass and tend to oxidise readily, or be coated with slimes before reaching the conditioning section, which makes collector adsorption difficult. Particles of low mass tend to be repelled by the slip-stream which surrounds a fast-moving air bubble, and should therefore be offered small, slow-moving bubbles. On the other hand, if small particles are in suspension near the froth column, they tend to overflow with the froth column regardless of their composition, as the downward pull of gravity is offset by the upward force due to the drift of the bubbles. Fine hydrophilic particles can also be mechanically entrained in the interstices between bubbles or be entrained in the water overflowing with the froth (Kirjavainen and Laapas, 1988). Such entrainment can be reduced by froth washing (Kaya and Laplante, 1991), as is performed in flotation columns.

When the ore value is low, the slimes (the ultrafine fraction which may be detrimental to flotation) are often removed from the granular fraction by passing the feed through de-slimming cyclones, and discarding the overflow. Alternatively, de-slimming can be carried out between flotation stages; for instance, rougher flotation may be followed by a de-slimming operation, which improves recovery in the scavenging stage. If the slimes contain substantial values, they are sometimes treated separately, thus increasing overall recovery.

Figure 12.55 shows the flowsheet used by the White Pine Copper Co. of Michigan, USA, to treat an ore consisting of chalcocite and native copper finely disseminated in a shale gangue (Tveter and McQuiston, 1962). Rapid flotation of the fine chalcocite and native copper is followed by de-slimming of the primary flotation tailings. Elimination of these gangue slimes accelerates recovery of the middlings in the scavenging stage.

Kaolin clay has been beneficiated for many years by carrier flotation, in which −60 μm particles of calcite are added to the system with oleic acid as collector. During conditioning, the fine anatase particles in the raw clay coat the coarse calcite particles and are separated from the clay when the calcite is recovered by flotation (Sivamohan, 1990). Fuerstenau et al. (1991) have demonstrated that carrier flotation can be carried out autogenously i.e. using the same mineral, and a hematite ore has been classified into a coarse and a fine fraction, the coarse hematite particles being used as carrier particles for the fine hematite. This phenomenon is an important type of shear flocculation (see Chapter 1), and has been used successfully in a number of mines in China, for the flotation of hematite, copper oxides, lead–zinc slimes, and tin slimes. In all cases concentrate is returned to the slimes feed, the coarser particles acting not only as carriers, but also promoting aggregation of the fines (Wang et al., 1988). Fuerstenau (1988) has argued that the consideration of such multi-feed circuits is expected to become an integral part of circuit selection for the separation of refractory ores.

The operating density of the pulp is determined by testwork, and is influenced by the mean size of particles within the feed. Coarse particles will
settle in a flotation cell at a relatively rapid rate, which may be substantially reduced by increasing the volume of particles in the pulp. As a general rule higher density pulps are applied to coarser sizes. In treating heavy sulphide ores, low-grade rougher concentrates are obtained from pulps of between 30 and 50% solids, whilst reground cleaner concentrates are obtained from pulps of between 10 and 30% solids.

Reagents and conditioning

Each ore is a unique problem and reagent requirements must be carefully determined by testwork, although it may be possible to obtain guidelines for reagent selection from examples of similar operations. An enormous amount of experience and information is freely available from reagent manufacturers. One vital requirement of a collector or frother is that it becomes totally emulsified prior to usage. Suitable emulsifiers must be used if this condition is not apparent.

Selection of reagents must be followed by careful consideration of the points of addition in the circuit. It is essential that reagents are fed smoothly and uniformly to the pulp, which requires close control of reagent feeding and on the pulp flow rate. Frothers are always added last when possible; since they do not react chemically they only require dispersion in the pulp, and long conditioning times are unnecessary. Adding frothers early tends to produce a mineralised froth floating on the surface of the pulp during the conditioning stage. This is due to entrained air, which can cause uneven distribution of the collector.

In flotation, the amount of agitation and consequent dispersion are closely associated with the time required for physical and chemical reactions to take place. Conditioning prior to flotation is now considered standard practice and is an important factor in decreasing flotation time. This is perhaps the most economical way of increasing the capacity of a flotation circuit. The minerals are converted to a readily floatable form as a result of ideal conditioning, and therefore a greater volume can be treated. Although it is possible to condition in a flotation machine, it is generally not economical to do so, although it is currently common practice for stage addition to include booster dosage of collector into cell banks, particularly at the transition from rougher to scavenger collection. Machines in the flow-line are often used as conditioners. Agitators are often interposed between the grinding mills and the flotation circuit to smooth out surges in grade and flow rate from the mills. Reagents are often added to these storage reservoirs for conditioning. Alternatively, reagents may be added to the grinding circuit in order to ensure optimum dispersion. The ball mill is a good conditioner and is often used when the collectors are oily and need emulsifying and long conditioning times. The advantage of conditioning in the mill is that the collector is present at the time that new surface is being formed, before oxidation can take place. The disadvantage is that reagent rate control is difficult, as the feed to the mill may have continual minor grade fluctuations, and the mill may have a high circulating load, which can become overconditioned. Where very close control of conditioning time is essential such as in the selective flotation of complex ores, special conditioning tanks are incorporated into the flow-line (Figure 12.56). The pulp and reagent are fed down the open stand-pipe and fall on to the propeller, which forces the mixture downwards and outwards. The outlet at the side of the tank can be adjusted to give a height sufficient to give the pulp its desired residence time within the tank.

Stage addition of reagents often yields higher recoveries at substantially lower cost than if all reagents are added at the same point in the circuit.
prior to flotation. The first 75% of the values is normally readily floatable, providing optimum grind size is achieved. The remaining values may well be largely composite in nature and will therefore require more careful reagent conditioning, but perhaps 15% are sufficiently large or sufficiently rich in value to be recovered relatively easily. The remaining 10% can potentially affect the whole economic balance of the process, being both fine in size and low in values. Because this fraction is such a critical one it must be examined extremely carefully and regularly, and reagent addition must be carefully and quickly controlled.

When feasible to do so, it is usually more desirable to float in an alkaline or neutral circuit. Acid circuits usually require specially constructed equipment to withstand corrosion. It is a common finding that the effectiveness of a separation may occur within very narrow pH limits, in which case the key to success for the whole process lies in the pH control system. In selective flotation where more than one mineral is concentrated, the separation pH may well vary from one stage to the next. This, of course, makes it vitally important to regulate reagents to bring about these conditions and control them accurately.

The first stage of pH control is often undertaken by adding dry lime to the fine ore-bins, which tend to reduce oxidation of sulphide mineral surfaces. Final close pH control may be carried out on the classifier overflow, by the addition of lime as a slurry. The slurry is usually taken from a ring main, as lime settles out quickly if not kept moving, and forms a hard cement within the pipelines.

Solid flotation reagents can be fed by rotating disc, vibro, and belt feeders, but more commonly reagents are added in liquid form. Insoluble liquids such as pine oil are often fed at full strength, whereas water-soluble reagents are made up to fixed solution strengths, normally about 10%, before addition. Reagent mixing is performed on day shifts in most mills, under close supervision, to produce a 24 h supply. The aqueous reagents are usually pumped through ring mains, from where they are drawn off to feeders as required.

Modern flotation plants typically add reagents via either positive displacement metering pumps or automatically controlled valves, where reagents are added in frequent short bursts from a ring main or manifold. With the increased complexity of measurement and control these methods allow online reagent addition rates to be relayed to either remote monitors or computers within control rooms. For small quantities, peristaltic pumps can be used, where rollers squeeze a carrier tube seated in a curved track, thus displacing the reagent along the tube. In a number of older flotation plants reagents are still added via Clarkson feeders, which use small cups on a rotating wheel, and through flow rotameters.

Small amounts of frother can be injected directly into the pipeline carrying the flotation feed, by positive-displacement piston metering pumps.

**Control of flotation plants**

Automatic control is increasingly being used, the control strategies being almost as numerous as the number of plants involved. The key to effective control is online chemical analysis (Chapter 3), which produces real-time analysis of the metal composition of process streams. Control strategies are implemented in distributed control systems (DCS) or programmable logic controllers (PLC) and there are many vendor-supplied solutions.

However, although there are many reports of successful applications, in reality few if any plants can claim to be fully automatic in the sense of operating unattended overextended periods, despite the availability of robust instrumentation, a wide range of control algorithms, and powerful computing assets. McKee (1991) has reviewed some of the reasons. The main problems have been in first successfully stabilizing a complex process, and then developing process models which will define set-points and limits to accommodate changes in ore type, mineralogy, texture, chemical composition of the mine water, and contamination of the feed. Control systems have also been unsuccessful in some cases due to inadequate maintenance of instrumentation. It is essential, for instance, that pH probes are kept clean, and that all online instrumentation is regularly serviced and calibrated. Implementation of control strategies at the plant design stage have rarely been successful as the most significant control variables are often not identified until experience of the plant has been gained. Only then can control strategies based on these variables, and with specific objectives, be successfully attempted. Another limitation is the training of plant operating and metallurgical staff in the principles and
application of control, and the shortage of control engineers needed to keep control systems running. The most successful systems have been those which allow the control room operator to interact with the plant control system when necessary to adjust set-points and limits. In this respect it is doubtful whether automatic control can achieve better metallurgical efficiency than experienced, conscientious operators in the short term. Its great advantage, however, is that the DCS is constantly vigilant, not being affected by shift changeovers, tea breaks, and other interruptions which affect the human operator.

A flotation control system consists of various subsystems, some of which may be manually controlled, while others may have computer-controlled loops, but all contributing to the overall control objective (Paakkinen and Cooper, 1979; Lynch et al., 1981). The aim should be to improve the metallurgical efficiency, i.e. to produce the best possible grade–recovery curve, and to stabilise the process at the concentrate grade which will produce the most economic return from the throughput (Figure 12.57), despite disturbances entering the circuit. This has not, as yet, been achieved by automatic control alone.

![Figure 12.57 Flotation control objective](image)

Disturbances caused by variations in feed rate, pulp density, and particle size distribution should be minimal if grinding circuit control is effective, such that the prime function of flotation control is to compensate for variations in mineralogy and floatability. The variables which are manipulated, either manually or automatically, to effect this are mass flows, reagent and air addition rates, pulp and froth levels, pH, and circulating loads by the control of cell-splits on selected banks.

Best practice involves getting the basic control objectives established first, such as stabilizing control of pulp and sump levels, air flow, and reagent flows. More advanced stabilizing control can then be attempted, such as pH, reagent ratio control (based on plant input flows and assays), pulp flow, circulating load, concentrate grade, and recovery. Finally, true optimising control can be developed, such as maximum recovery at a target grade. Higher level optimising control is generally not possible until stable operation has been achieved (McKee, 1991).

The key variable to control is pulp level in the cell, because a constant pulp level is very important to ensure stable and efficient flotation performance. The pulp level can be measured by a number of different means. Ultrasonic devices measure the time sound waves take to reach the pulp level, or a "float" resting at the froth/pulp interface. Floats can also be connected to sensing devices which measure how far the float moves as the pulp level changes, either through a vertical motion sensor or a horizontal lever. Conductivity probes register the difference in electrical conductivity between the froth and pulp to determine the pulp level. Differential pressure cells are submerged in the flotation tank and measure the pressure exerted on them by the liquid above. Bubble tubes also determine the pulp level based on pressure of the pulp compressing air within the bubble tube though these are not much used today.

Control of pulp level is effected by dart valves or pinch valves. In older flotation plants movable weirs are also used. In general each bank of cells will have a level detection transducer (usually a float-based device) and the level is then controlled by a simple feed back PI loop which adjusts the valve on the bank tailings outlet based on a set-point either entered by the operator or determined by a higher-level control strategy responding to changes in grade, recovery, froth condition, or other criteria. Feed forward in combination with feedback control is often required to avoid damaging interactions between different flotation banks. Feed forward control is based on feed flow measurement or inference (e.g. from a variable speed pump or preceding level controllers).
Level control can either be simple, as outlined above, or involve more complex interactions (Kampjarvi and Jamsa-Jounela, 2002). “Float-Star™”, developed by Mintek in South Africa, is an integrated package providing level control throughout a flotation circuit, plus additional capabilities such as an algorithm to calculate optimum level set-points and/or aeration rates that aim to optimize the residence times, mass pulls, and circulating loads within a flotation circuit (Singh et al., 2003).

Control of slurry pH is a very important requirement in many selective flotation circuits, the control loop often being independent of the others, although in some cases the set-point is varied according to changes in flotation characteristics. For automatic control of lime or acid it is important that time delays in the control loop are minimised, which requires reagent addition as close as possible to the point of pH measurement. Lime is often added to the grinding mills in order to minimise corrosion and to precipitate heavy metal ions from solution. In the circuit shown in Figure 12.58, lime addition is controlled by the ratio of the mass flow to the mill, and the ratio set-point is adjusted by a pH controller which measures pH early in the flotation process with an operator-determined pH set-point. Lags are sufficient to allow sufficient mixing in the mill.

Control of collector addition rate is sometimes performed by feed-forward ratio control based on a linear response to assays or tonnage of valuable metal in the flotation feed. Typically, increase in collector dosage increases mineral recovery until a plateau is reached, beyond which further addition may either have no practical effect, or a slight reduction in recovery may occur. The gangue recovery also increases with collector addition, such that beyond the plateau region selectivity is reduced (Figure 12.59). The operator can intervene to change the ratio set-point or bias to respond to the changing feed conditions.

The most common aim of collector control is to maintain the addition rate at the edge of the plateau, the main difficulty being in identifying the optimum point, especially when the response changes due to changes in ore type, or the interaction with other reagents. For this reason, automatic control using feed-forward loops has rarely been successful in the long term. There are many cases of successful semi-automatic control, however, where the operator adjusts the set-point to accommodate changes in ore type, and the computer controls the reagent addition over fairly narrow limits of feed grade. For example, feed-forward control of copper sulphate activator and xanthate to the zinc roughers has been used in the control strategy at Mattagami Lake Mines, Canada (Konigsman et al., 1976). The reagents are varied in proportion to changes in feed grade according to a simple ratio/bias algorithm, which is a standard algorithm supplied with all DCSs or PLCs:

\[ \text{Reagent flow rate} = A + (B \times \%Zn \text{ in feed}) \]

Where \( A \) and \( B \) vary for different reagents. The operator may change the base amount \( A \) as different

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**Figure 12.58** Control of pH in a flotation circuit
ore types are encountered. The logic of the control system is shown in Figure 12.60.

![Figure 12.59 Effect of collector addition](image)

Figure 12.59 Effect of collector addition

At Mount Isa in Australia, feed-forward control of xanthate addition to the copper roughers was unsatisfactory, as the optimum addition rate was not simply related to the mass of copper in the flotation feed (Fewings et al., 1979). The assay of concentrate produced in the first four cells of the bank was combined with the four-cell tailings and feed assays to compute the four-cell recovery. It was found that there was a linear response between this recovery and the collector dosage required to maintain the overall recovery at the edge of the plateau. The control strategy, although fairly successful in the short-term, eventually failed when changes in ore type occurred. Computation of unit process recovery in this way is also subject to error due to inherent inaccuracies of on-stream analysis data (see Example 3.13).

The amount of frother added to the flotation system is an important variable, but automatic control has been unsuccessful in many cases, as the nature of the froth is dependent on only very minor changes in frother addition and is much affected by intangible factors such as contamination of the feed, mine water chemistry, etc. At low addition rates, the froth is unstable and recovery of minerals is low, whereas increasing frother addition rate has a marked effect on the flotation rate, increasing the weight, and reducing the grade of concentrate produced. The usual approach is to manually adjust the frother set-point, or less commonly to ratio the frother to the feed rate of solids and water.

Flow rate of concentrate has been controlled in some systems by regulating the frother addition. The grade is not as sensitive to changes in frother addition, but there may be a good relationship between grade and flow rate. Cascade control can be used, where the concentrate grade controls the concentrate flow rate set-point, which in turn controls the frother addition set-point (Figure 12.61). Stabilising control of conventional column cells, which generally operate with deep froths, is relatively simple (Finch and Dobby, 1989).

Air input to the flotation process and froth depth are parameters which, like frother addition, affect the recovery of minerals into the concentrate, and can be used to control concentrate grade, tailings grade, or mass flow rate of concentrate. Aeration and froth depth do not, however, affect subsequent cleaning operations, as will residual frother
carried over from the roughers, and they are often used as primary control variables. Flotation generally responds faster to changes in aeration than to changes in froth depth, and because of this aeration is often a more effective control variable, especially where circulating loads have to be controlled. There is obviously interaction between frother addition, aeration and froth depth, and where computer-controlled loops are used it is necessary to control these variables such that only minor changes are made. This can be done by manipulating only one of these variables, maintaining the others constant at predetermined optimum levels unless the conditions deviate outside acceptable limits, which may vary with ore type. At Vihanti in Finland (Figure 12.71), the copper grade of the bulk copper–lead rougher concentrate has been used to control the rate of aeration and frother addition to the roughers and scavengers. Aeration has priority, being the cheaper “reagent” and leaving no residual concentration if used in excess. However, if the addition rate reaches a certain upper limit, then the frother rate is increased (Wills, 1983).

The importance of froth depth is mainly due to the effect that it has on the gangue content of the concentrate. Free gangue can be carried into the concentrate mainly by mechanical entrainment, and the deeper the froth layer the more the drainage of gangue into the cell occurs. Froth depth is very commonly used to control the concentrate grade, an increase in froth depth increasing the grade, but often at the expense of a slight reduction in recovery. Froth depth is often regarded as the difference between the pulp level and the level of the flotation cell overflow lip, and as such is controlled by changing the pulp level by the control and measurement methods mentioned earlier.

Froth level set-points can be cascaded to aeration or frother set-point controllers in order to maintain the required depth. Specification of the actual froth depth requires a knowledge of the level of the froth column surface, which may not coincide with the height of the cell overflow lip. Figure 12.62 shows a device developed at Mattagami Lakes for sensing the level of the froth column, this level controlling the frother dosage set-point (Kitzinger et al., 1979). The sensor consists of a set of stainless steel electrodes connected to an electronic circuit which senses the number in contact with the froth. The seven electrodes, one of which is always immersed in the pulp, are of gradually decreasing length, so that the number in contact with the froth is directly proportional to the depth of the froth column.

![Figure 12.61 Cascade control of frother addition](image1)

![Figure 12.62 Froth measuring device](image2)

The most common, recent, froth measurement device utilises ultrasonics. A ball float lies at the froth–pulp interface and is connected to a vertical shaft. A target plate is mounted on the upper end of the vertical shaft, above the top of the froth. An ultrasonic transmitter directs soundwaves to the target plate and the froth depth is calculated from the time taken for the sound waves to return to the source.

Froth level devices are used at Pyhasalmi in Finland (Figure 12.70). The addition of copper sulphate activator to the zinc circuit is controlled mainly by the on-stream analysis data but an excess tends to depress the froth level. The circuit contains several froth level measuring devices which indicate the improper addition of copper sulphate early enough to adjust the frother and sulphate addition to prevent a disturbance.
The ultimate aim of control is to increase the economic efficiency of the process by seeking to optimise performance, and there are several strategies which can be adopted to achieve this. Evolutionary optimisation (EVOP) methods (Chapter 3) (Oberg and Deming, 2000) have potential for flotation optimisation but have not been widely used. The control method involves periodically adjusting the set-points of the controlled variables according to a defined experimental design strategy such as a factorial or simplex search, the effect on economic efficiency being calculated and fed back to the operating system. The set-points are then shifted slightly to move in the direction of the optimum, and the process repeated until an optimum is encountered. Such methods cannot, however, be fully effective unless satisfactory stabilisation of plant performance can be achieved over long periods.

Herbst et al. (1986) discussed the use of advanced model-based control strategies in flotation, highlighting the advantages of these modern methods over classical control schemes. McKee (1991) also reviewed the progress in this area.

The Black Mountain concentrator in South Africa developed adaptive optimisation to control lead flotation (Twiddle et al., 1985). Optimising control calculated the combination of metal recovery and concentrate grade which would achieve the highest economic return per unit of ore treated under the prevailing conditions. The criterion used to evaluate plant performance was the concept of economic efficiency (Chapter 1), in this case defined as the ratio between the revenue derived per tonne of ore at the achieved concentrate grade and recovery, and that derived at the target grade and recovery. Target concentrate grade and recovery were calculated from the operating recovery-grade curve, which was continuously updated based on a 24 h data bank, to allow for changes in the nature of the ore, quality of grinding, etc. Many factors influence the optimum combination of recovery and grade, such as commodity prices, reagent and treatment costs, transport costs, etc. The fundamental principle of adaptive optimisation is that concentrate grade and recovery can be predicted by online multivariable linear regression models, the coefficients of the models being continuously updated from the 24 h data bank. Independent variables that determine grade and recovery can be reagent additions, grades of rougher concentrate, final concentrate and cleaner tailing, feed grade and throughputs. Some independent variables are controllable whilst others are not.

The Pyhasalmi concentrator developed optimisation control based on a multi-linear response model, to optimise copper and zinc recoveries, and the balance of these metal values in each concentrate, to provide the highest economic efficiency (smelter value of metal in concentrate/value of metal in feed) (Miettunen, 1983). This took into account factors such as the penalties caused by the presence of zinc in the copper concentrate and the increasing costs of transportation due to a low copper content in the concentrate. Cyanide addition was the most influential variable in the copper circuit, while copper sulphate dosage to the zinc rougher bank was adjusted to maximise the economic recovery of the total zinc flotation circuit. The effect of copper sulphate on the rougher concentrate assays and the scavenger tailing assay was determined, and the approach used was to apply multiple linear regression to a 3 hour history of data stored in the process control computer. With this procedure, the effect of copper sulphate changes on economic recovery could be determined, and therefore the requirement to either increase or decrease copper sulphate to improve the economic recovery was known. Copper sulphate changes, determined by the optimising control system, were usually made every 6–30 min.

In recent years adaptive control (Thornton, 1991), expert systems (Kittel et al., 2001) and neural networks (Cubillos and Lima, 1997) have all been applied to the flotation problem with varying degrees of success. The texture, velocity, and colour of flotation froths are diagnostic of the flotation condition and are used by skilled operators to adjust set-points, particularly air addition rates. This function has now been implemented in machine vision systems which measure these properties online (van Olst et al., 2000; Holtham and Nguyen, 2002), allowing control systems to make use of froth characteristics in optimising performance (Kittel et al., 2001).

A comprehensive control system for a flotation plant requires extensive instrumentation and involves a considerable capital outlay. Figure 12.63 shows the instrumentation requirement for a simple
A feed-forward system which could assist in control of a sulphide rougher bank, and Figure 12.64 shows the instrumentation used in the control of the Mount Isa copper flotation circuit in Queensland, Australia (Fewings et al., 1979). Although various cascade control loops have been attempted in this circuit, they have been unsuccessful in the long term due to changes in feed conditions, and set-points within the loops are mainly controlled by the operators.

Lynch et al. (1981) analysed the cost of such installations, which provide potentially significant economic and metallurgical benefits. The majority of plants which have installed instrumentation for manual or automatic control purposes have reported improved metal recoveries varying from 0.5 to 3.0%, sometimes with increased concentrate grades. Reduction in reagent consumptions of between 10 and 20% have also been reported.
**Typical flotation separations**

The expansion of flotation as a method of mineral concentration can be observed from the following data. According to surveys carried out by the US Bureau of Mines, the ore treated by flotation in the United States expressed as million tonnes was 180 in 1960, 368 in 1970, 440 in 1980, and 384 in 1985. Worldwide, froth flotation is used to treat 2000 Mt of material annually. In 1980, just before the recession in the American mineral industries, 55% of the total US tonnage was base metal sulphides, 27% phosphates, 9% iron ores, 6% industrial minerals, and 3% coal (Fuerstenau, 1988).

Although flotation is increasingly used for the non-metallic and oxidised minerals, the bulk of the world’s tonnage currently processed is sulphide from the ores of copper, lead, and zinc, often associated in complex ore deposits. The treatment of such ores serves as an introduction to the flowsheets encountered in plant practice. Comprehensive reviews of the complete range of sulphide, oxide, and non-metallic flotation separations can be found elsewhere (Jordan et al., 1986; Malhotra et al., 1986; Redeker and Bentzen, 1986; Crozier, 1990) and good reviews of the flotation of specific materials such as coal (Osborne, 1988; Firth, 1999; Meenan, 1999), phosphates (Lawver et al., 1984; Hsieh and Lehr, 1985; Anon., 1986a; Moudgil, 1986; Wiegel, 1999), iron ore (Houot, 1983; Iwasaki, 1983, 1999; Nummela and Iwasaki, 1986), cassiterite (Lepetic, 1986; Senior and Poling, 1986; Andrews, 1990), scheelite (Beyzavi, 1985), chromium and manganese minerals (Fuerstenau et al., 1986), and gold (O’Connor and Dunne, 1994) are also available.

**Flotation of copper ores**

Over 15 Mt of copper are produced annually in the world, and in 2003 about 35% was from Chile (Yianatos, 2003). Significant tonnages were also produced in Canada (11%), Zambia (7.4%), Zaire (4.9%), and Australia (4.5%) (Thompson, 1991). During the 1990s, low copper prices resulted in a considerable quantity of mine capacity being unused, particularly in the United States, where many mines were forced to shut down or cut back on production. Ore grades in US mines average only 0.6% Cu, compared with 2.2% Cu in Africa and 1.2% Cu in South America, and to convert such low-grade ores to saleable concentrates, especially with current low metal prices and high production costs, requires a high level of technology and control, and a careful balance between concentrate grades, recovery and milling costs. Substantial increases in copper prices since 2003 (Chapter 1) have resulted in the reopening of many mines and significant increases in copper flotation capacity.

Copper is characterised by having a number of economic ore minerals (Appendix 1), many of which may occur in the same deposit, and in various proportions according to depth. Copper sulphides in the upper part of an ore body often oxidise and dissolve in water percolating down the outcrops of the deposit. A typical reaction with chalcopyrite is:

$$2\text{CuFeS}_2 + 17\text{O} + 6\text{H}_2\text{O} + \text{CO}_2 \rightarrow 2\text{Fe(OH)}_3$$

$$+ 2\text{CuSO}_4 + 2\text{H}_2\text{SO}_4 + \text{H}_2\text{CO}_3$$

The residual ferric hydroxide left in this leached zone is called *gossan* and its presence has often been used to identify a copper ore body. As the water percolates through the zone of oxidation it may precipitate secondary minerals such as malachite and azurite to form an oxidised cap on the deeper primary ore.

The bulk of the dissolved copper, however, usually stays in solution until it passes below the water table into reducing conditions, where the dissolved metals may be precipitated from solution as secondary sulphides, e.g.:

$$\text{CuFeS}_2 + \text{CuSO}_4 \rightarrow \text{FeSO}_4 + 2\text{CuS(covellite)}$$

$$5\text{FeS}_2 + 14\text{CuSO}_4 + 12\text{H}_2\text{O} \rightarrow 5\text{FeSO}_4$$

$$+ 12\text{H}_2\text{SO}_4 + 7\text{CuS(chalcocite)}$$

As these secondary sulphide minerals contain relatively high amounts of copper, the grade of the ore in this *zone of supergene enrichment* is increased above that of the underlying primary mineralisation, and where supergene enrichment has been extensive, spectacularly rich copper “bonanzas” are formed.

The earliest copper miners worked the relatively small amount of metallic copper contained in the oxidised zone of the ore bodies. The discovery of smelting allowed high-grade oxidised copper minerals to be worked and processed. With improved developments in copper metallurgy, such as matte smelting and conversion, the
secondary sulphide supergene zones were mined and processed, these deposits often being shallow and containing 5% or more copper.

The development of froth flotation had an enormous impact on copper mining, enabling the most abundant primary mineral, chalcopyrite, and other sulphides to be efficiently separated from ores of relatively low grade and fine grain size. Another major development was the introduction of vast tonnage open-pit mining methods to the copper industry, allowing the excavation of tens of thousands of tonnes of ore per day. This made economical the processing of huge low-grade bulk copper deposits known as porphyries, the most important being found in the United States and South America.

The importance of froth flotation and high-tonnage mining can be seen by considering that until 1907 practically all the copper mined in the United States was from underground vein deposits, averaging 2.5% Cu, whereas at present ore grades in the United States average only 0.6% Cu and about 50% of the world's copper is produced from porphyry deposits, the rest mainly from vein-type and bedded deposits.

The exact definition of copper porphyry has been the subject of debate amongst geologists for a long time (Lacy, 1974). They are essentially very large oval or pipe-shaped deposits containing on average 140 Mt of ore, averaging about 0.8% Cu and 0.015% Mo, and a variable amount of pyrite (Sutolov, 1974). All porphyry copper deposits contain at least traces of molybdenite (MoS₂), and in many cases molybdenum is an important by-product. Porphyry copper mineralisation is often referred to as disseminated, and although on a large scale immense volumes of ore may contain disseminated values, on a small scale the occurrence of sulphides is controlled by fractures. Even apparently disseminated sulphide minerals are often aligned with quartz micro-veinlets, or lie in a chain-like fashion (see Figure 1.2b). The chains mark early fractures, which have been sealed and camouflaged by quartz and feldspar (Edwards and Atkinson, 1986).

The first deposits of this type to be mined on a large scale were in the southwestern states of the United States. It was apparent that the deposits could be economically mined in bulk by large-scale low-cost methods such as block-caving and open-pit methods. This is because the copper minerals are distributed uniformly through large blocks of the deposit so that the expensive selective mining methods which must be used with vein or bedded deposits are not needed. The extent of the ore body is usually determined by its copper content rather than by geological structure, the copper content tending to decrease away from the core of the mass. The cut-off grade, which determines the boundary between ore and waste, varies from mine to mine and according to the prevailing economic climate.

Porphyry copper operations are very much influenced by the geology of the ore deposition. Mining necessarily starts in the upper zones of the ore body where secondary alteration has enriched the ore grade, and where the mineralogy allows the production of concentrates often grading more than 40% Cu at high recovery. High levels of output can be achieved with fairly compact mills and smelters. As the operation matures, however, lower grade primary (hypogene) ore is encountered, in which the mineralogy limits concentrate grades to only around 25–30% Cu, and more ore needs to be produced to realise the same net copper output, the alternative being to maintain the current plant throughput while metal output declines. Reagent use and flowsheets often have to be adapted to accommodate these changes in mineralogy. A classic case is the El Teniente mine in Chile, the world's largest underground copper mine, which was developed in one of the largest known copper porphyry deposits on earth (estimated to contain 44 Mt of copper in ore grading 0.99% Cu or more). In 1979, the ore, of grade 1.54% Cu, was being mined and processed at the rate of 57,500 t d⁻¹ to produce a concentrate containing 40% Cu (Dayton, 1979). By 1984, with the secondary supergene zone approaching exhaustion, the ore grade had fallen to 1.4% Cu, and the mining rate had increased to 68,500 t d⁻¹, with a further expansion to 90,000 t d⁻¹ being undertaken. It was predicted that the mined grade would fall to 1.2% by the end of the 1980s and to 1.0% by the end of the twentieth century (Burger, 1984).

Although mining and processing of copper porphyries is on a vast scale, concentration of the ore is fairly straightforward, due to the high efficiency of froth flotation, and to the fact that breakage of the ore occurs preferentially at the
fracture zones containing the copper sulphides. This means that relatively coarse grinding produces composite particles with much of the valuable mineral exposed, facilitating rougher flotation.

Copper sulphide minerals are readily floatable and respond well to anionic collectors such as xanthates, notably amyl, iso-propyl and butyl. Alkaline circuits of pH 8.5–12 are generally used, with lime controlling the pH and depressing any pyrite present. Frother usage has changed significantly in recent years, away from the natural reagents such as pine oil and cresylic acids, to the synthetic frothers such as the higher alcohols (e.g., MIBC) and polyglycol esters. Cleaning of the rougher concentrates is usually necessary to reach an economic smelter grade (25–50% Cu depending on mineralogy), and rougher concentrates as well as middlings must often be reground for maximum recovery, which is usually between 80 and 90%. Primary grinding is normally to about 50–60% –75 microns, rougher concentrates being reground to 90–100% –75 microns to promote optimum liberation of values. Reagent consumption is generally in the range 0.002–0.3 kg t\(^{-1}\) of xanthate, and 0.02–0.15 kg t\(^{-1}\) of frother.

One of the largest copper concentrators in the world is at the Freeport mine in the Republic of Indonesia on the island of New Guinea. The plant was progressively expanded since initial start-up in 1972 from 7500 t d\(^{-1}\) to 200,000 t d\(^{-1}\) to compensate for the lower grade ore encountered as the open pit deepened. The principal copper mineral in the porphyry deposit is chalcopyrite. Gold and silver are also present in the primary ore, which in 1997 graded 1.3% Cu, 1.32 g t\(^{-1}\) Au and 2.82 g t\(^{-1}\) Ag (Coleman and Napitupulu, 1997). The gold content is the largest known reserve of gold in the world.

The flotation circuit is large (comprising four concentrators) but fairly simple. After primary grinding to produce a flotation feed grind size of 15% passing 212 microns, the ore is conditioned with lime, frother and collector, before being fed to the rougher flotation circuit. The rougher flotation circuit consists of four parallel banks of Wemco 127 m\(^3\) flotation cells with nine cells per bank. The cleaner circuit consists of fourteen column cells for primary and secondary cleaning and twelve 85 m\(^3\) mechanical scavenger flotation cells. The concentrate produced from the columns report to final concentrate while the concentrate from the scavengers are recycled back to the cleaner feed. In 1996, the Freeport operation produced 526,000 t of copper and 1,760,000 troy ounces of gold (Coleman and Napitupulu, 1997). Typical copper and gold recoveries are 86 and 76%, respectively.

By-products are important to the economics of copper porphyry operations, and the most important by-product of the North and South American porphyries is molybdenum. Molybdenum occurs as the highly floatable mineral, molybdenite, which is separated from the copper minerals after regrinding and cleaning of the copper rougher concentrates. Regrinding to promote optimum liberation needs careful control, as molybdenite is a soft mineral which slimes easily and whose floatability decreases as particles become finer. Rougher concentrates are therefore classified, only coarse cyclone underflows being reground in closed circuit. Cleaned copper concentrates are thickened, after which the copper minerals are depressed allowing molybdenite to be floated into a concentrate which is further cleaned, sometimes in up to twelve stages. Cleaning is important as molybdenite concentrates are heavily penalised by the smelter if they contain copper and other impurities, and the final copper content is often adjusted by leaching in sodium cyanide, which easily dissolves chalocite and covellite and some other secondary copper minerals. Chalcopyrite, however, does not dissolve in cyanide, and in some cases is leached with hot ferric chloride.

Copper depression is achieved by the use of a variety of reagents, sometimes in conjunction with prior heat treatment. Heat treatment is used to destroy residual flotation reagents, and is most commonly achieved by the use of steam injected into the slurry. Depression of chalcopyrite may be effectively accomplished by the use of sodium cyanide, but this reagent is not so effective when chalcocite and bornite are present, in which case depression can be completed by the use of ferro- and ferri-cyanides, or by using “Nokes Reagent”, a product of the reaction of sodium hydroxide and phosphorus pentasulphide. This reagent has an instantaneous depressing action on copper minerals and is rapidly consumed, so is added to the circuit in stages. It can be an expensive depressant because of its high (2–5 kg t\(^{-1}\) of concentrate)
consumption, and is sometimes used in combination with cyanide. Other copper depressants are arsenic Nokes (As$_2$O$_3$ dissolved in Na$_2$S), sodium sulphide, sodium hydrosulphide, and thioglycolic acid. Ye et al. (1990) have shown that ozone conditioning can also effectively depress copper minerals. The molybdenite is floated using a light fuel oil as collector.

Figure 12.65 shows the molybdenum recovery flowsheet at the Chuquicamata Mine in Chile, the world’s biggest copper producer (Sisselman, 1978). The copper concentrate, containing 0.8–3% MoS$_2$ is floated in the rougher circuit after depressing the copper minerals with sodium hydrosulphide (Shirley and Sutolov, 1985). The first cleaner concentrate is recleaned in four to seven stages using 2.5 kg t$^{-1}$ of arsenic Nokes reagent, and regrinding of first and fourth cleaner concentrates, to produce a concentrate containing 55% Mo and 1–2% Cu. This product is then leached with sodium cyanide to reduce the copper content, which is predominantly as chalcopyrite, to below 0.3%. Sodium cyanide is added also to the last two cleaner stages. All flotation cells in the molybdenum plant operate with nitrogen from the smelter oxygen plant, rather than air, the reducing potential considerably lowering the consumption of depressant (Crozier, 1986).

By-products play an important role in the economics of the Palabora Mining Co. in South Africa, which treats a complex carbonatite ore to recover copper, magnetite, uranium, and zirconium values. The ore assays about 0.5% Cu, the principal copper minerals being chalcopyrite and bornite, although chalcocite, cubanite (CuFe$_2$S$_3$), and other copper minerals are present in minor amounts. The flotation feed is coarse (80%–300 microns) due to the high grinding resistance of the magnetite in the ore which would increase grinding costs if ground to a finer size, and due to the fact that the flotation tailings are treated by low-intensity magnetic separation to recover magnetite, and Reichert cone gravity concentration to recover uranothorite and baddeleyite.

The flotation circuit consists of eight separate sections, the last two sections being fed from an

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**Figure 12.65** Molybdenum flotation at Chuquicamata (Sisselman, 1978)
autogenous grinding circuit. The first five parallel sections, the original Palabora flowsheet, are fed from conventional mills, each at the rate of 385 t h\(^{-1}\) (Figure 12.66). Flotation feed is conditioned with sodium isobutyl xanthate and frother before being fed to the rougher flotation banks. The more readily floatable minerals, mainly liberated chalcopyrite and bornite, float off in the first few cells, and more collector is added before the final scavenger cells, in order to float off the less floatable particles, such as cubanite, and in order to attempt to float the less responsive copper minerals, such as valleriite, a copper–iron sulphide containing Mg and Al groups in the crystal lattice. Valleriite occurs intergrown with other sulphide minerals (Figure 12.67), and due to the fact that it is a very soft mineral, it can lead to poor flotation recoveries. During comminution, breakage occurs along the soft and friable

Figure 12.66 Flowsheet of original section of Palabora flotation circuit

Figure 12.67 Palabora copper ore. Valleriite (V) and cubanite intergrown with chalcopyrite (Ch)
vallerite, leaving grains of other copper sulphides with a vallerite coating, preventing these grains from floating (Figure 12.68).

Rougher and scavenger concentrates are reground to 90%-45 microns, before being fed to the cleaner circuit at a pulp density of 14% solids, this dilution being possible due to the removal of magnetite and other heavy minerals into the tailings, and the fine particle size produced after grinding.

Oxidised copper ores

Due to the nature of copper deposits and mineralisation, it is sometimes possible to selectively mine and process the oxidised cap on the primary zone. Minerals such as malachite and azurite are soluble in dilute sulphuric acid and can be processed economically by acid leaching as a prelude to precipitation of the copper by electrolysis (electrowinning). Processing of such oxidised ores has become more attractive due to the availability of cheap sulphuric acid produced at smelters, as a means of reducing sulphur dioxide emissions into the atmosphere.

In Central Africa significant tonnages of oxidised ore were concentrated by flotation before being leached, ores containing a mixture of sulphide and oxidised minerals being treated by first floating off the sulphides to produce a concentrate for the smelter. A good example was the Chingola Division of Nchanga Consolidated Copper Mines in Zambia where sulphide tailings were floated in an oxide circuit using sodium hydrosulphide, sodium isopropyl xanthate and frother, with the concentrates being acid leached.

Much of the published work on oxide copper minerals is concerned with malachite, and chrysocolla, a copper silicate (Deng and Chen, 1991). The latter is one of the most widely distributed and least understood of all the major copper minerals, being a very difficult mineral to characterise and float (Laskowski et al., 1985). Malachite responded well to flotation techniques and in Central Africa flotation of malachite ores after sulphidisation is successfully practised (Fuerstenau and Raghavan, 1986). Xanthate collector coatings are loosely bound to oxide copper minerals and sulphidisation enhances the flotation process.

Today flotation is rarely used for copper oxide recovery. Such ores are generally leached with sulphuric acid, and the metal is recovered by solvent extraction and electrowinning. Low grade ores are often heap leached (Witt et al., 1999).

Flotation of lead–zinc ores

The bulk of the world’s lead and zinc is supplied from deposits which often occur as finely disseminated bands of galena and sphalerite, with varying
amounts of pyrite, as replacements in various rocks, typically limestone or dolomite. This banding sometimes allows dense medium preconcentration prior to grinding (Figure 11.13).

Although galena and sphalerite usually occur together in economical quantities, there are exceptions, such as the lead ore body in S.E. Missouri, of the United States, where the galena is associated with relatively minor amounts of zinc (Watson, 1988), and the zinc-rich Appalachian Mountain region, mined in Tennessee and Pennsylvania, where lead production is very small.

Feed grades are typically 1–5% Pb and 1–10% Zn, and although relatively fine grinding is usually needed (often to well below 75 µm), fairly high flotation concentrate grades and recoveries can be achieved. In an increasing number of cases, ultrafine grinding down to 10 µm is needed to produce acceptable flotation performance from very fine grained ores such as those at the Century mine in Australia. Typically, lead concentrates of 55–70% lead are produced, containing 2–7% Zn, and zinc concentrates of 50–60% Zn, containing 1–6% Pb. Although galena and sphalerite are the major ore minerals, cerussite \((\text{PbCO}_3)\), anglesite \((\text{PbSO}_4)\), marmatite \(((\text{Zn,Fe})_2\text{S})\) and smithsonite \((\text{ZnCO}_3)\) can also be significant. In some deposits the value of associated metals, such as silver, cadmium, gold, and bismuth, is almost as much as that of the lead and zinc, and lead–zinc ores are the largest sources of silver and cadmium.

Several processes have been developed for the separation of galena from zinc sulphides, but by far the most widely used method is that of two-stage selective flotation, where the zinc and iron minerals are depressed, allowing the galena to float, followed by the activation of the zinc minerals in the lead tailings to allow a zinc float.

Sphalerite (and to a lesser extent pyrite) can become activated by heavy metal ions in solution, which replace metallic zinc on the mineral surfaces by a process of ion exchange (e.g. Equation 12.11). This activated surface can adsorb xanthate and produce a very insoluble heavy metal xanthate which provides the surface with a water-repellent “envelope”. Clean sphalerite is not strongly hydrophobic in xanthate solutions, as zinc xanthate has a relatively high solubility, and hence a stable envelope is not formed.

Heavy metal ions are often present in the slurry water, especially if the ore is slightly oxidised. The addition of lime or soda ash to the slurry can precipitate them as relatively insoluble basic salts, thus “de-activating” the sphalerite to some extent. The alkali is usually added to the grinding mills as well as to the lead float conditioner, as it is in the grinding process that many heavy metal ions are released into solution.

Lead flotation is usually performed at a pH of between 9 and 11, lime, being cheap, often being used to control alkalinity. Not only does lime act as a strong depressant for pyrite, but it can also depress galena to some extent. Soda ash is sometimes preferred because of this, especially when the pyrite content is relatively low.

The effectiveness of alcalis as deactivators is dependent on the concentration of heavy metal ions in solution, as the basic salts which are precipitated, although of extremely limited solubility, can provide a source of heavy metal ions sufficient to cause sphalerite activation. In most cases, therefore, other depressants are required, the most widely used being sodium cyanide (up to 0.15 kg t\(^{-1}\)) and zinc sulphate (up to 0.2 kg t\(^{-1}\)), either alone or in combination. These reagents are commonly added to the grinding circuit, as well as to the lead float, and their effectiveness depends very much on pulp alkalinity.

Apart from the reactions with metal ions in solution, cyanide has long been used to dissolve surface copper from activated sphalerite, and can react with iron and zinc xanthates to form soluble complexes, eliminating xanthate from the surfaces of the minerals of these metals. Pyrite is thus depressed with the sphalerite, and cyanide is generally the preferred depressant where soda ash regulates alkalinity and pyrite presence is significant.

The effectiveness of depressants also depends on the concentration and selectivity of the collector. Xanthates are most widely used in lead–zinc flotation, and the longer the hydrocarbon chain, the greater the stability of the metal xanthate in cyanide solutions and the higher the concentration of cyanide required to depress the mineral. If the galena is readily floatable, potassium or sodium ethyl xanthate may be used, together with a “brittle” frother such as MIBC. Sodium isopropyl xanthate may be needed if the galena is tarnished, or if considerable amounts of lime are used to promote
pyrite depression. Powerful collectors such as amyl xanthate can be used if the sphalerite is clean and hydrophilic, and are needed where the galena is highly oxidised and floats poorly.

Although cyanides are widely used due to their high degree of selectivity, they do have certain disadvantages. They are toxic and expensive, and they depress and dissolve some of the gold and silver which are often present in economic amounts. For these reasons, zinc sulphate is used in many plants to supplement cyanide. This reduces cyanide consumption (usually to well below 0.1 kg t\(^{-1}\)), and a number of mines in the USA achieve depression by the use of zinc sulphate alone.

After flotation of the galena, the tailings are usually treated with between 0.3 and 1 kg t\(^{-1}\) of copper sulphate, which reactivates the surface of the zinc minerals (Equation 12.11), allowing them to be floated. Lime (0.5–2 kg t\(^{-1}\)) is used to depress pyrite, as it has no depressing effect on the activated zinc minerals, and a high pH (10–12) is used in the circuit. Isopropyl xanthate is perhaps the most commonly used collector, although ethyl, isobutyl, and amyl are also used, sometimes in conjunction with dithiophosphate (aerofloats), depending on conditions. As activated sphalerite behaves in a similar way to chalcopyrite, thionocarbamates such as Z-200 are also common collectors, selectively floating the zinc minerals from the pyrite.

Careful control of reagent feeding must be observed when copper sulphate is used in conjunction with xanthates, as xanthates react readily with copper ions. Ideally, the minerals should be conditioned with the activator separately, so that when the conditioned slurry enters the collector conditioner there is little residual copper sulphate in solution. Although the activation process is fairly rapid in acidic or neutral conditions, in practice it is usually carried out in an alkaline circuit in order to prevent pyrite activation, and a conditioning time of some 10–15 min is required to make full use of the reagent. This is because the alkali precipitates the copper sulphate as basic compounds which are sufficiently soluble to provide a reservoir of copper ions for the activation reaction.

The Sullivan concentrator of Cominco Ltd, British Columbia, operates an interesting flowsheet which includes de-zincing of the lead concentrates and de-leading of the zinc concentrates (Fairweather, 2005). The ore is essentially a replacement deposit in argillaceous quartzite, the ore bodies being massive, fine-grained mixtures of sulphides, sometimes interbanded with the country rock. The principal economic minerals are galena and marmatite (7ZnS:FeS), iron being present mainly as pyrrhotite, and to a lesser extent pyrite. Silver is closely associated with the galena and is an important by-product.

The flowsheet is shown in Figure 12.69. After primary grinding to 55%–74 μm with cyanide, xanthate, and lime, the ore is fed to a unit flotation cell, where a mixture of MIBC and pine oil frothers is added. The pH is maintained at 8.5, and a coarse lead concentrate is floated, and cleaned once. This concentrate, assaying about 65% lead, is used as medium in the DMS circuit preceding grinding. The tailing from the coarse lead flotation is ground to 87%–74 μm, and is conditioned with sodium isopropyl xanthate, cyanide, lime, and MIBC, before being fed to the lead roughers at a pH of 9.5. Further addition of cyanide and xanthate to the head of the scavenger cells produces a concentrate which is returned to secondary grinding. The lead rougher concentrate is cleaned, the tailings being reground and returned to the lead roughers. The pH in the cleaners is 10.0, and the cleaner concentrate is further cleaned at pH 10.5 to produce a concentrate containing 10–14% Zn. The final stage of lead flotation is the de-zincing of the second lead cleaner concentrate. After activating the zinc minerals with copper sulphate, the galena is depressed by raising the pH to 11.0 by the addition of lime, and by steam heating the slurry to 30–40°C. A rougher de-zincer concentrate is cleaned once in the first few cells of the bank, and the dezincer tailing is the final lead concentrate, assaying about 62% Pb and 4.5% Zn. The lead scavenger tailings are conditioned with about 0.7 kg t\(^{-1}\) of copper sulphate, prior to feeding to zinc rougher flotation where xanthate, lime and frother are added to the cells. A rougher concentrate is floated at pH 10.6, and is reground before being fed to the first stage of cleaning. The tailings from this stage, containing 2.5–4% Pb, are pumped back to the head of the lead circuit to allow a better recovery of lead in that concentrate. The cleaner concentrate is recleaned twice, the final concentrate being combined with the de-zincer concentrate to produce the final zinc concentrate containing 50% Zn and 4% Pb.
The increasing fine-grained nature and complexity of lead–zinc ores has led in some cases to the need for rougher concentrates to be ground extremely finely. Flotation of a material that had been ground to an ultra-fine size was achieved at the MacArthur River Mine in Australia, where rougher concentrates are ground to 12 μm to produce a bulk lead–zinc concentrate. At Mount Isa Mines rougher concentrates of lead and zinc are ground to 10 and 15 μm respectively prior to cleaner flotation (Young and Gao, 2000). At the Century Mine, zinc concentrates are ground to below 10 μm to effectively liberate fine-grained silicates (Burgess et al., 2003). Due to the high intensity of ultra-fine grinding, inert grinding media is often used to prevent oxidation of mineral surfaces. The production of ultra-fine concentrates usually results in very tenacious froths, with pulping and material handling problems being common.

Extremely fine intergrowth between galena and sphalerite inhibits selective flotation separation, and in some cases sphalerite is activated by copper ions in the ore to such an extent that depression of sphalerite fails, even when the most powerful combinations of reagents, such as zinc sulphate and cyanide, are used. Bulk flotation of lead and zinc minerals may in such cases have a number of economical advantages. Coarse primary grinding is often sufficient with bulk flotation, as the valuable minerals need be liberated only from the gangue, not from each other. The flotation circuit design is normally relatively simple. In contrast, selective flotation calls for finer primary grinding, in order to free the valuable minerals not only from the gangue, but also from each other. This increases mill size and energy requirements; the flotation volume will increase proportionally to the number of selective concentrates.

However, the production of bulk lead–zinc concentrates is only reasonable if there are smelters which are adequately equipped for such concentrates. The only smelting process available is the Imperial Smelting Process, which was developed at a time when most of the lead and zinc was recovered from low-pyrite ore deposits. In recent years, however, lead and zinc are increasingly being recovered from complex and highly pyritic ores.
Bulk concentrates for smelting in the ISP should be low in iron, as iron is recovered in the smelter slag. An increase in iron content increases slag production, and correspondingly increases zinc losses, as the slag carries about 5% zinc. Furthermore, a high iron content increases smelter energy consumption. When smelter revenues are compared, the highest revenues are achieved when selective concentrates are produced. Even mixing selective concentrates into a bulk concentrate will yield higher revenues than bulk concentrates produced by direct flotation. This is because better selectivity between non-ferrous minerals and pyrite is achieved by the optimal conditions adapted to the separation of galena and pyrite in the first, and sphalerite and pyrite in the second step. The chemical conditions in a bulk flotation cannot be adjusted to meet both conditions simultaneously if a high amount of pyrite is present. It has been shown that, although selective is more expensive than bulk flotation, the increase in revenues gained is often much higher than the additional operating costs (Bergmann and Haidlen, 1985).

Bulk flotation followed by separation can sometimes be used, although in most cases the activated sphalerite and pyrite in the bulk concentrate are covered with a layer of collector, and are difficult to depress unless extremely large amounts of reagent are used. This is especially the case if copper sulphate has been used to activate the sphalerite; cyanide will react with residual copper ions in solution. Every attempt is made at plants using bulk flotation to use the minimum collector feed for the bulk flotation step, which can lead to low recoveries. Bulk flotation is performed at Zinkgruvan, Sweden’s largest zinc mine (Anon., 1977). Grinding is autogenous and the lead ions released during grinding activate the sphalerite to such an extent that deactivation by alkali is not practical at this stage. The flotation plant consists of bulk flotation and lead flotation stages, each circuit consisting of rougher, scavenger, and cleaner steps. The galena and sphalerite are floated with 0.12 kg t$^{-1}$ of potassium ethyl xanthate, no activator being required. After five stages of cleaning, the concentrate is conditioned with 0.6 kg t$^{-1}$ of ZnSO$_4$ to depress the sphalerite, and the galena is floated at pH 10 with potassium ethyl xanthate. After six stages of cleaning, and further additions of ZnSO$_4$, a lead concentrate of 65% and a zinc concentrate of 55% are produced.

An interesting bulk-selective flowsheet is operated at the Tochibora mine, in Japan (Anon., 1984a), which has an annual output of 960,000 t of ore, grading 4.3% Zn, 0.3% Pb, and 22 g t$^{-1}$ Ag. Pyrite is not present to any extent in the ore, the principal gangue minerals being hedenbergite (CaFeSi$_2$O$_6$), quartz, calcite, and epidote. Crushed ore is ground to 80% passing 75 μm, and is conditioned with Na$_2$CO$_3$ and CuSO$_4$ before bulk flotation at pH 9.4. Sodium ethyl xanthate is used as collector and pine oil as frother. After cleaning the bulk flotation concentrates, the slurry is conditioned with NaCN and activated carbon, after which galena is floated, the tailings being the zinc concentrate. The lead concentrate is fed into trommels, the oversize forming a graphite by-product concentrate, while the undersize is fed to shaking tables. The table middlings and tailings are recycled to the differential flotation circuit, the cleaned concentrate being the final lead concentrate. Concentrates grading 60.7% Zn and 65.3% Pb are obtained at recoveries of 93.3% Zn and 80.2% Pb.

Flotation of copper–zinc and copper–lead–zinc ores

The production of separate concentrates from ores containing economic amounts of copper, lead, and zinc is complicated by the similar metallurgy of chalcopyrite and activated zinc minerals. The mineralogy of many of these ores is a complex assembly of finely disseminated and intimately associated chalcopyrite, galena, and sphalerite in a gangue consisting predominantly of pyrite or pyrrhotite (often 80–90%), quartz, and carbonates. Such massive sulphide ores of volcanosedimentary origin are also a valuable source of silver and gold.

The complex Cu–Pb–Zn ores represent 15% of total world production and 7.5% of the world copper reserves, these percentages being higher for zinc (Cases, 1980). Grades of ore mined 0.3–3% Cu, 0.3–3% Pb, 0.2–10% Zn, 3–100 g t$^{-1}$ silver, and 0–10 g t$^{-1}$ gold, on average.

The major processing problems encountered are related specifically to the mineralogy of the assemblies. Due to the extremely fine dissemination and interlocking of the minerals, extensive fine grinding is often needed, usually to well below...
There are notable exceptions to this such as at Bleikvassli in Norway where a primary grind of 80%-240 μm is adequate, with no regrinding (Anon., 1980). In the New Brunswick deposits in Canada, however, grinding to 80%-40 μm is required in certain areas, optimum mineral recoveries being in the range 10-25 μm. Such extensive fine grinding is extremely energy intensive (in the order of 50 kWh t⁻¹), and the large surface area produced leads to high reagent consumptions, the release of metal ions into solution, which reduces flotation selectivity, and a greater tendency for surface oxidation. Oxidation is particularly serious with galena, which is often overground in closed-circuit grinding, being the heaviest mineral in the complex ores.

In most cases, concentrates are produced at relatively poor grades and recoveries, typical grades being:

<table>
<thead>
<tr>
<th></th>
<th>%Cu</th>
<th>%Pb</th>
<th>%Zn</th>
</tr>
</thead>
<tbody>
<tr>
<td>Copper concentrates</td>
<td>20-30</td>
<td>1-10</td>
<td>2-10</td>
</tr>
<tr>
<td>Lead concentrates</td>
<td>0.8-5</td>
<td>35-65</td>
<td>2-20</td>
</tr>
<tr>
<td>Zinc concentrates</td>
<td>0.3-2</td>
<td>0.4-4</td>
<td>45-55</td>
</tr>
</tbody>
</table>

Recoveries of 40-60% for copper, 50-60% for lead, and 70-80% for zinc are reported for New Brunswick deposits (Stemerowicz and Leigh, 1978). Smelting charges become excessive with contaminated concentrates, as very rarely is a metal paid for when it is not in its proper concentrate and penalties are often imposed for the presence of zinc and lead in copper concentrates. Silver and gold are well paid for in copper and lead concentrates, whereas payment in zinc concentrates is often zero. Direct sale of the concentrates to custom smelters is necessary where the size of the ore body precludes the development of a specialised smelter complex, such as that at the Ronnskar works of Boliden, Sweden, where a collection of metallurgical plants facilitates the transfer, or recycling, of residues and by-products from one process stage to another for the recovery of all metal values (Barbery et al., 1980).

The overall revenue for a mine exploiting such deposits can be very low compared to the relatively high contained value of the ore. Gray (1984) has shown the economic limitations of processing complex ores by a standard route by comparing the concentrator performance at two Australian mines: North Broken Hill and Woodlawn. The former mine realised about 56% of the potential ore value in payments received, whereas Woodlawn realised only about 27% of the ore value in payments. The disparity in the two balances is almost solely due to the differences in recovery resulting from the much greater mineralogical complexity of the Woodlawn deposits. Deposits with such complex mineralogy are to be found in many parts of the world, whereas deposits with mineralogy comparable with North Broken Hill are now rare. The metallurgist's task is to characterise each deposit quantitatively and systematically and then to select the economically optimum combination of process steps to suit the characteristics. Imre and Castle (1984) have also comprehensively reviewed the exploitation strategies for complex Cu-Pb-Zn ore bodies, discussing the interaction and optimisation of the beneficiation and extractive metallurgical flowsheets and the options for extractive metallurgy in processing complex sulphides containing pyrite. Barbery (1986) has also discussed the many potential processing options available for treating complex sulphides, concluding that it is likely, for some years, that combined processes will be developed, linking physical separation processes with hydrometallurgy for maximum efficiency in recovering values into concentrates that are well paid by conventional existing smelters. In turning such integrated treatment concepts into reality, the fundamental question will be: is one flowsheet, involving one set of processes, capital and operating costs superior to another treatment approach with a different set of costs and metallurgical performance? Further, it is necessary to assess the impact of different product grades from the integrated process on subsequent downstream processes. The answer to this question, although critical, is likely to be very complex, and McKee (1986) has analysed the role of computer analysis in answering such questions.

Flotation is, at present, the only method that can be used to beneficiate the complex sulphide ores, and a wide variety of flowsheets are in use, some involving sequential flotation, others bulk flotation.
of copper and lead minerals followed by separation. Bulk flotation of all the economic sulphides from pyrite has also been studied. Although bulk flotation has certain advantages, it has been shown that the requirements for adequate galena flotation, as well as those for selective flotation of sphalerite from pyrite, are difficult to meet in a single bulk circuit, and better metallurgical efficiency can be obtained by floating, and then mixing, separate copper–lead and zinc concentrates. However, the main disadvantage is that a concentrate having no market is produced, for which new metallurgical processes have to be developed (Barbery, 1986).

In the flotation of copper–zinc ores, where lead is absent, or is not present in economic quantities, lime is almost universally used to control alkalinity at pH 8–12, and to deactivate the zinc minerals by precipitation of heavy metal ions. In a few cases, the addition of lime to the mills and flotation circuit is sufficient to prevent the flotation of zinc minerals, but in most cases supplementary depressants are required. Sodium cyanide is often added in small quantities (0.01–0.05 kg t⁻¹) to the mills and cleaners; if present in large amounts chalcopyrite is also depressed. Zinc sulphate is also used in conjunction with cyanide, and in some cases sodium sulphite (or bisulphite) or sulphur dioxide depressants are used. Work in the United States (Hoyack and Raghavan, 1987) has indicated that sulphite depresses pyrite, but only has a slight effect on the flotation of sphalerite. The depression of sphalerite is probably governed by electrochemical reactions that yield a hydrophilic surface product, Fe₂(SO₄)₃·Fe(OH)₃.

After conditioning, the copper minerals are floated using either xanthates, or if the mineralogy allows, a selective copper collector such as isopropyl thionocarbamate. Typically, copper concentrates contain 20–30% Cu and up to 5% Zn. Copper flotation tailings are activated with copper sulphate, and zinc minerals are floated as described in the previous section.

Due to the very close control of reagent additions required in copper–zinc separations, on-stream X-ray analysis of plant flow-streams is being increasingly used, together with some form of automatic control. A good example is the Pyhasalmi concentrator in Finland (Figure 12.70), which is highly automated, and involves sequential flotation of copper, zinc, and pyrite (Wills, 1983). The copper circuit consists of conventional roughing and scavenging, followed by three cleaning stages, the tailings passing to the zinc flotation circuit. Despite the use of cyanide (0.025 kg t⁻¹) and zinc sulphate (1.45 kg t⁻¹), a problem in the copper circuit is the natural activation of sphalerite by copper-bearing water; because of this a flotation time of about 20 min is required for satisfactory copper recovery (about 90%) and the copper concentrate contains about 25% Cu and 3.5% Zn. Reagent additions are controlled automatically according to set-points regulated by on-stream analysis of copper, zinc, and iron contents in various flowstreams. Due to the varying quality of the ore, caused by fluctuating quantities of activated zinc minerals, cyanide addition is the most important variable affecting the economic recovery and is controlled from the set-points to keep the zinc content of the copper concentrate at a minimum while maintaining optimum copper recovery.

The method most widely used to treat ores containing economic amounts of lead, copper, and zinc is to initially float a bulk lead–copper concentrate, while depressing the zinc and iron minerals. The zinc minerals are then activated and floated, while the bulk concentrate is treated by the depression of either the copper or lead minerals to produce separate concentrates.

The bulk float is performed in an alkaline circuit, usually at pH 7.5–9.5, lime, in conjunction with depressants such as cyanide and zinc sulphate, being added to the mills and bulk circuit. Depression of zinc and iron sulphides is sometimes supplemented by the addition of small amounts of sodium bisulphite or sulphur dioxide to the cleaning stages, although these reagents should be used sparingly as they can also depress galena.

The choice and dosage of collector used for bulk flotation are critical not only for the bulk flotation stage but also for the subsequent separation. Xanthates are commonly used, and while a short-chain collector such as ethyl xanthate gives high selectivity in floating galena and chalcopyrite and permits efficient copper–lead separation, it does not allow high recoveries into the bulk concentrate, particularly of the galena. Much of the lost galena subsequently floats in the zinc circuit, contaminating the concentrate, as well as representing an economic loss. Because of this, a powerful collector such as amyl or isobutyl xanthate is commonly
Figure 12.70 Pyhasalmi flotation circuit. F = flow rate; L = level; A = assay; FL = froth level; C = conductivity
used, and very close control of the dosage is required. Usually, fairly small amounts, of between 0.02 and 0.06 kg t\(^{-1}\), are used, as an excess makes copper–lead separation difficult, and large amounts of depressant are required which may depress the floating mineral, contaminating lead and copper concentrates.

Although the long-chain collectors improve bulk recovery, they are not as selective in rejecting zinc, and sometimes a compromise between selectivity and recovery is needed, and a collector such as sodium isopropyl xanthate is chosen. Dithiophosphates, either alone or in conjunction with xanthates, are also used as bulk float collectors, and small amounts of thionocarbamate may be used to increase copper recovery.

The choice of the method for separating copper from the lead minerals depends on the response of the minerals and the relative abundance of the copper and lead minerals. It is preferable to float the mineral present in least abundance, and galena depression is usually performed when the ratio of lead to copper in the bulk concentrate is greater than unity.

Lead depression is also undertaken if economic amounts of chalcocite or covellite are present, as these minerals do not respond to depression by cyanide, or if the galena is oxidised or tarnished and does not float readily. It may also be necessary to depress the lead minerals if the concentration of copper ions in solution is high, due to the presence of secondary copper minerals in the bulk concentrate. The standard copper depressant, sodium cyanide, combines with these ions to form complex cuprocyanides (Equation 12.23), thus reducing free cyanide ions available for copper depression. Increase in cyanide addition only serves to accelerate the dissolution of secondary copper minerals.

Depression of galena is achieved using sodium dichromate, sulphur dioxide, and starch in various combinations, whereas copper minerals are depressed using cyanide, or cyanide–zinc complexes. Methods of depression used at various concentrators can be found elsewhere (Wills, 1984).

Depression of galena by the addition of sodium dichromate at high pH is still used in many plants. The hydrophobic character of the xanthate layer on the galena surface is inhibited by the formation of hydrated lead chromate (Cecile et al., 1980). At Vihanti (Figure 12.71), the galena is depressed by the addition of 0.01 kg t\(^{-1}\) of sodium dichromate to the bulk concentrate. After copper flotation, the separation tailings are further floated to remove residual copper, the cleaner tailings producing the final lead concentrate. Although there is no automatic control of the separation circuit, the rate of addition of dichromate is critical, as an excess is returned to the rougher feed with the cleaner tailing, which depresses lead into the zinc circuit.

Although the amount of dichromate used is only small (0.01–0.2 kg t\(^{-1}\)), chromate ions can cause environmental pollution, and other methods of depression are sometimes preferred. Depression of galena by sulphite adsorption is the most widely used method, sulphur dioxide, either as liquid or gas, being added to the bulk concentrate; sodium sulphite is less commonly used. In many cases, causticised starch is added in small amounts as an auxiliary depressant, but tends to depress the copper if insufficient sulphur dioxide is used. The sulphur dioxide reduces the pH to between 4 and 5.5, the slightly acidic conditions cleaning the surfaces of the copper minerals, thus aiding their floatability. Small amounts of dichromate may be added to the circuit to supplement lead depression.

In some plants, galena depression is aided by heating the slurry to about 40°C by steam injection. Kubota et al. (1975) showed that galena can be completely depressed, with no reagent additions, by raising the slurry temperature above 60°C, and this method is being used by the Dowa Mining Company in Japan (Anon., 1984b, 1984c). The xanthate adsorbed on the galena is removed, but that on the chalcopyrite surface remains. It is thought that preferential oxidation of the galena surface at high temperature is the mechanism for depression. At Woodlawn in Australia, the lead concentrate originally assayed 30% Pb, 12% Zn, 4% Cu, 300 ppm Ag, and 20% Fe, and received very unfavourable smelter terms (Burns et al., 1982). Heat treatment of the concentrate at 85°C for 5 min, followed by reverse flotation, gave a product containing 35% Pb, 15% Zn, 2.5% Cu, 350 ppm Ag, and 15% Fe, with improved sales terms.

At the Brunswick Mining concentrator in Canada (McTavish, 1980) (Figure 12.72), the bulk copper–lead concentrate is conditioned for 20 min with 0.03 kg t\(^{-1}\) of a wheat dextrine–tannin extract
mixture to depress the galena, and 0.03 kg t\(^{-1}\) of activated carbon to absorb excess reagents and contaminants, and then the pH is lowered to 4.8 with liquid SO\(_2\). The slurry is further conditioned for 20 min at this low pH, then 0.005 kg t\(^{-1}\) of thionocarbamate is added to float the copper minerals. The rougher concentrate is heated by steam injection to 40°C, and is then cleaned three times to produce a copper concentrate containing 23% Cu, 6% Pb, and 2% Zn. The lead concentrate produced is further upgraded by regrinding the copper separation tails, and then heating the slurry with steam to 85°C, and conditioning for 40 minutes. Xanthate and dithiophosphate collectors are then added to float pyrite. The rougher concentrate produced is reheated to 70°C and is cleaned once. The hot slurry from the lead upgrading tailing contains about 32.5% Pb, 13% Zn, and 0.6% Cu, and, after cooling, is further treated to float a lead–zinc concentrate, leaving a final lead concentrate of 36% Pb and 8% Zn.

In general, where the ratio of lead to copper in the bulk concentrate is less than unity, depression of the copper minerals by sodium cyanide may be preferred. Where standard cyanide solution may cause unacceptable dissolution of precious metals and small amounts of secondary copper minerals, a cyanide–zinc complex can sometimes be used to reduce these losses. At Morococha in Peru (Pazour, 1979), a mixture of sodium cyanide, zinc oxide, and zinc sulphate has been used, allowing a recovery of 75% of the 120 g t\(^{-1}\) of silver in the ore.

Close alkalinity control is necessary when using cyanides, a pH of between 7.5 and 9.5 commonly being used, although the optimum value may be higher, dependent on the ore. Cyanide depression is not used if economic quantities of chalcocite or covellite are present in the bulk concentrate, since it has little depressing action on these minerals. As cyanide is a very effective sphalerite depressant, most of the zinc reporting to the bulk concentrate is depressed into the copper concentrate, which may incur smelter penalties. Cyanide, however, has little action on galena, allowing effective flotation of the galena from the chalcopyrite, and hence a low
lead–copper concentrate. Lead is never paid for in a copper concentrate, and is often penalised.

In a few cases, adequate metallurgical performance cannot be achieved by semi-bulk flotation, and sequential selective flotation must be performed. This necessarily increases capital and operating costs, as the bulk of the ore – the gangue minerals – is present at each stage in separation, but it allows selective use of reagents to suit the mineralogy at each stage. The general flowsheet for sequential flotation involves conditioning the slurry with SO₂ at low pH (5–7), and using a selective collector such as ethyl xanthate, dithiophosphate, or thionocarbamate, which allows a copper concentrate which is relatively low in lead to be floated. The copper tailings are conditioned with lime or soda ash, xanthate, sodium cyanide, and/or zinc sulphate, after which a lead concentrate is produced, the tailings being treated with copper sulphate prior to zinc flotation.

Sequential separation is required where there is a marked difference in floatability between the copper and lead minerals, which makes bulk rougher flotation and subsequent separation of the minerals in the bulk concentrate difficult, as at the Black Mountain concentrator in South Africa (Beck and Chamart, 1980). In Australia, sequential separation was performed at Cobar Mines Ltd (Seaton, 1980). Metallurgical development at Woodlawn in Australia was an ongoing process. The original circuit, designed to depress lead with dichromate, was never effective for various reasons, and a combination of bulk and sequential flotation was then used (Roberts et al., 1980; Burns et al., 1982). The feed, containing roughly 1.3% Cu, 5.5% Pb, and 13% Zn, was conditioned with SO₂, starch, sodium metabisulphite and a dithiophosphate collector, after which a copper concentrate was produced, which was cleaned twice. The copper tailings were conditioned with lime, NaCN, starch, and sodium secondary butyl xanthate, prior to flotation of a lead concentrate which contained the less floatable copper minerals. This concentrate was reverse cleaned by steam heating to 85°C prior
to flotation of the copper minerals with no further reagent addition. The floated copper minerals were pumped to the initial copper cleaning circuit. Lead rougher tailings fed the zinc roughing circuit.

**Flotation of nickel ores**

Nickel is produced from two main sources: sulphidic ores and lateritic ores. While 70% of the land-based nickel resources are contained in lateritic deposits, the majority of the world’s current production of nickel still comes from sulphidic sources (Bacon et al., 2000). The dominant nickel mineral in these deposits is pentlandite – (NiFe)$_9$S$_8$. However, many ores also have minor amounts of millerite (NiS) and violarite (Ni$_2$FeS$_4$). Nickel can also be found within the pyrrhotite (Fe$_8$S$_9$) lattice (substitute for iron). In some cases the nickel is in the pyrrhotite (Kerr, 2002). Depending on the downstream smelting requirements, nickel flotation can occur as two processes: bulk sulphide flotation (e.g. in Western Australia’s nickel operations) or separate pyrrhotite flotation (e.g. Canada’s Sudbury area). In addition to iron sulphides, nickel often occurs with economic concentrations of copper (Sudbury), cobalt (Western Australia), and precious metals such as gold, platinum, palladium, rhodium, ruthenium, iridium, and osmium (Noril’sk operation in north west Siberia, and in the Bushveld Complex in South Africa).

A good review of six of the current major nickel flotation operations is given by Kerr (2002), which covers typical Sudbury area operations as well as operations in Western Australia (Mt Keith) and Russia (Noril’sk).

**Flotation of platinum ores**

Platinum is one of the Platinum Group Metals (PGMs), which also include palladium, iridium, osmium, rhodium, and ruthenium. They are generally found together in economic ores, and 90% of PGM production comes from South Africa and Russia. In 2004, 44% of platinum was used in catalysts for motor vehicle emission control, and 33% in jewellery. The PGMs are classed with gold and silver as precious metals.

There are three main types of PGM deposit: PGM-dominant (e.g. the Bushveld Igneous Complex in South Africa), Ni–Cu dominant (e.g. Sudbury in Canada and Noril’sk in Russia) and miscellaneous. PGMs are usually recovered by flotation as a bulk low-grade sulphide concentrate, followed by smelting and refining.

There are over 100 known PGMs including sulphides, tellurides, antimonides, arsenides, and alloys. Each of these has a unique metallurgical behaviour, and the mode of occurrence and grain size considerably varies according to location (Corrans et al., 1982). The mineral association and gangue minerals present specific challenges to flotation that affect downstream processing, e.g. talc (Shortridge et al., 2000) and chromite (Wesseldijk, 1999). Typical reagent suites include thiol collectors (xanthate, in some cases with co-collectors dithiophosphate or dithiocarbamate); in some cases, copper sulphate is added as an activator; polymeric depressants such as guar or carboxymethyl cellulose are added to inhibit recovery of naturally floatable talcaceous gangue (Wiese et al., 2005).

The wide range of valuable mineral densities in PGM ores presents problems in conventional classification in grinding circuits, so the South African flotation concentrators sometimes employ combined milling and flotation circuits without classification (Snodgrass et al., 1994). Flash flotation and preconcentration by DMS or gravity are also used.

**Flotation of iron ore**

Iron ore minerals such as goethite and hematite are floated by collectors such as amines, oleates, sulphonates, or sulphates. Processing involves preconcentration by gravity or magnetic separation, followed by flotation. Iron ore flotation has increased in importance due to market requirements for higher grade products. This requires the flotation of silicate impurities from the iron ore. Amines are commercially used for the flotation of silica from magnetite ore at the Kudremukh Iron Ore Company Ltd, India, and in many other parts of the world (Das et al., 2005).

The requirement for higher grade product has seen an increase in the use of flotation columns in iron ore treatment. In Brazil, all new iron ore concentration circuits commissioned since
the 1990s have consisted of rougher-cleaner-scavenger column-only configurations (Araujo, et al. 2005).

Flotation of coal

Unlike metalliferous flotation, where all of the product is treated by flotation, in coal treatment only a portion is treated. This is typically 10 to 25% of the feed tonnage and represents the fines portion, usually below 250 μm in size, but sometimes up to 1 mm. Mining production methods, in particular the increased use of longwall mining, has resulted in an increase in fines production and made the flotation of coal fines more economically viable. In many countries environmental legislation has limited the amount of coal that can be sent to tailing ponds, with flotation being the only effective way to recover this coal.

Flotation circuits in coal processing are relatively simple with roughing and scavenging flotation used. Sometimes roughing alone is adequate. The mass recovery in coal flotation is high (up to 70%) and frother usage rates can be high to keep the froth mobile. Many flotation circuits use mechanical paddles to physically remove the heavy froth from the flotation cells. Petrochemical products are usually used as collectors with the most common used being diesel oil, liquid paraffin, and kerosene.

Coal operations can produce one of two products, depending on the quality of coal mined, these being either high value coking coal for pyrometallurgical industries or lower value thermal coal for power generation. Coking coal product demands few impurities and the ash content (non-combustible content) is typically between 5 and 8%. Often coking coals require washing and this has seen flotation machines such as the Jameson Cell and flotation columns increasingly used. Flotation concentrates for thermal coals range from 8 to 14%. Often this can be achieved without froth washing and mechanical flotation cells are still commonly used (Nicol, 2000).

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Magnetic and electrical separation

Introduction

Magnetic and electrical separators are being considered in the same chapter, as there is often a possibility of an overlap in the application of the two processes. For example, as can be seen later, there is often debate as to which form of separation is best suited at various stages to the treatment of heavy mineral sand deposits.

Magnetic separation

Magnetic separators exploit the difference in magnetic properties between the ore minerals and are used to separate either valuable minerals from non-magnetic gangue, e.g. magnetite from quartz, or magnetic contaminants or other valuable minerals from the non-magnetic values. An example of this is the tin-bearing mineral cassiterite, which is often associated with traces of magnetite or wolframite which can be removed by magnetic separators.

All materials are affected in some way when placed in a magnetic field, although with most substances the effect is too slight to be detected. Materials can be classified into two broad groups, according to whether they are attracted or repelled by a magnet:

(1) Diamagnetics are repelled along the lines of magnetic force to a point where the field intensity is smaller. The forces involved here are very small and diamagnetic substances cannot be concentrated magnetically.

(2) Paramagnetics are attracted along the lines of magnetic force to points of greater field intensity. Paramagnetic materials can be concentrated in high-intensity magnetic separators. Examples of paramagnetics which are separated in commercial magnetic separators are ilmenite (FeTiO₃), rutile (TiO₂), wolframite ((Fe, Mn)WO₄), monazite (rare earth phosphate), siderite (FeCO₃), pyrrhotite (FeS), chromite (FeCr₂O₄), hematite (Fe₂O₃), and manganese minerals.

Some elements are themselves paramagnetic, such as Ni, Co, Mn, Cr, Ce, Ti, O, and the Pt group metals, but in most cases the paramagnetic properties of minerals are due to the presence of iron in some ferromagnetic form.

Ferromagnetism can be regarded as a special case of paramagnetism, involving very high forces. Ferromagnetic materials have very high susceptibility to magnetic forces and retain some magnetism when removed from the field (remanence). They can be concentrated in low-intensity magnetic separators and the principal ferromagnetic mineral separated is magnetite (Fe₃O₄), although hematite (Fe₂O₃) and siderite (FeCO₃) can be roasted to produce magnetite and hence give good separation. The removal of “tramp” iron from ores can also be regarded as a form of low-intensity magnetic separation.

It is not intended to review the theory of magnetism in any depth, as this is amply covered elsewhere (Svoboda, 1987).
The unit of measurement of magnetic flux density or magnetic induction \( B \) (the number of lines of force passing through a unit area of material) is the tesla (T).

The magnetising force which induces the lines of force through a material is called the field intensity \( H \), and by convention has the units ampere per metre \((1 \text{ A m}^{-1} = 4\pi \times 10^{-7} \text{ T})\).

The intensity of magnetisation or the magnetisation \((\text{M A/m})\) of a material relates to the magnetisation induced in the material, and:

\[
B = \mu_0 (H + M) \tag{13.1}
\]
the constant of proportionality, \( \mu_0 \) being the permeability of free space, and having the value of \( 4\pi \times 10^{-7} \text{ T} \cdot \text{m/A} \). In vacuum, \( M = 0 \), and it is extremely low in air, such that Equation 13.1 becomes:

\[
B = \mu_0 H \tag{13.2}
\]
so that the value of the field intensity is virtually the same as that of flux density, and the term magnetic field intensity is then often loosely used. However, when dealing with the magnetic field inside materials, particularly ferromagnetics that concentrate the lines of force, the value of the induced flux density will be much higher than the field intensity, and it must be clearly specified which term is being referred to.

Magnetic susceptibility \((S)\) is the ratio of the intensity of magnetisation produced in the material to the magnetic field which produces the magnetisation, i.e.:

\[
S = M/H \tag{13.3}
\]
Combining Equations 13.1 and 13.3:

\[
B = \mu_0 H(1 + S)
\]
or

\[
B = \mu \mu_0 H \tag{13.4}
\]
where \( \mu = 1 + S \), and is a dimensionless number known as the relative permeability.

For paramagnetic materials, \( S \) is a small positive constant, and is a negative constant for diamagnetic materials. Figure 13.1 shows plots of induced magnetisation \((M)\) versus the strength of the external field \((H)\), for paramagnetic (hematite) and diamagnetic (quartz) materials. Both plots show straight line relationships between \( M \) and \( H \), in each case the slope representing the magnetic susceptibility \((S)\) of the material, i.e. about 0.01 for hematite and around −0.001 for quartz.

The magnetic susceptibility of a ferromagnetic material is dependent on the magnetic field, decreasing with field strength as the material becomes saturated. Figure 13.2 shows a plot of \( M \) versus \( H \) for magnetite, showing that at an applied field of 1 T the magnetic susceptibility is about 0.35, and saturation occurs at about 1.5 T. Many high-intensity magnetic separators use iron cores and frames to produce the desired magnetic flux concentrations and field strengths. Iron saturates magnetically at about 2–2.5 T, and the non-linear ferromagnetic relationship between inducing field strength and magnetisation intensity necessitates the use of very large currents in the energising coils, sometimes up to hundreds of amperes.

The capacity of a magnet to lift a particular mineral is dependent not only on the value of the field intensity, but also on the field gradient, i.e. the rate at which the field intensity increases towards the magnet surface. Because paramagnetic minerals have higher magnetic permeabilities
than the surrounding media, usually air or water, they concentrate the lines of force of an external magnetic field. The higher the magnetic susceptibility, the higher is the field density in the particle and the greater is the attraction up the field gradient towards increasing field strength. Diamagnetic minerals have lower magnetic susceptibility than their surrounding medium and hence expel the lines of force of the external field. This causes their expulsion in the direction down the gradient of the field towards the decreasing field strength. This negative diamagnetic effect is usually orders of magnitude smaller than the positive paramagnetic attraction (Cohen, 1986).

It can be shown that

\[ F \propto H \frac{dH}{dl} \]  

where \( F \) is the force on the particle, \( H \) is the field intensity, and \( dH/dl \) is the field gradient.

Thus in order to generate a given lifting force, there are an infinite number of combinations of field and gradient which will give the same effect. Production of a high field gradient as well as high intensity is therefore an important aspect of separator design.

Magnetic separator design

Certain elements of design are incorporated in all magnetic separators, whether they are low or high intensity, wet or dry. The prime requirement, as has already been mentioned, is the provision of a high-intensity field in which there is a steep field strength gradient. In a field of uniform flux, such as in Figure 13.3(a), magnetic particles will orient themselves, but will not move along the lines of flux. The most straightforward method for producing a converging field is by providing a V-shaped pole above a flat pole, as in Figure 13.3(b). The tapering of the upper pole concentrates the magnetic flux into a very small area giving high intensity. The lower flat pole has the same total magnetic flux distributed over a larger area. Thus there is a steep field gradient across the gap by virtue of the different intensity levels.

Another method of producing a high field gradient is by using a pole which is constructed of alternate magnetic and non-magnetic laminations (Figure 13.4).

Provision must be incorporated in the separator for regulating the intensity of the magnetic field so as to deal with various types of material. This is easily achieved in electromagnetic separators by varying the current, while with permanent magnets the interpole distance can be varied.

Commercial magnetic separators are continuous-process machines and separation is carried out on a moving stream of particles passing into and through the magnetic field. Close control of the speed of passage of the particles through the field is essential, which rules out free fall as a means of feeding.
Belts or drums are very often used to transport the feed through the field.

The introduction into a magnetic field of particles which are highly susceptible concentrates the lines of force so that they pass through them (Figure 13.5).

Since the lines of force converge to the particles, a high field gradient is produced, which causes the particles themselves to behave as magnets, thus attracting each other. Flocculation, or agglomeration, of the particles can occur if they are small and highly susceptible and if the field is intense. This has great importance as these magnetic “flocs” can entrain gangue and can bridge the gaps between magnetic poles, reducing the efficiency of separation. Flocculation is especially serious with dry separating machines operating on fine material. If the ore can be fed through the field in a monolayer, this effect is much less serious, but, of course, the capacity of the machine is drastically reduced. Flocculation is often minimised by passing the material through consecutive magnetic fields, which are usually arranged with successive reversal of the polarity. This causes the particles to turn through 180°, each reversal tending to free the entrained gangue particles. The main disadvantage of this method is that flux tends to leak from pole to pole, reducing the effective field intensity.

Provision for collection of the magnetic and non-magnetic fractions must be incorporated into the design of the separator. Rather than allow the magnetics to contact the pole-pieces, which would cause problems of detachment, most separators are designed so that the magnetics are attracted to the pole-pieces, but come into contact with some form of conveying device, which carries them out of the influence of the field, into a bin or a belt. Non-magnetic disposal presents no problems, free fall from a conveyor into a bin often being used. Middlings are readily produced by using a more intense field after the removal of the highly magnetic fraction.

Types of magnetic separator

Magnetic separators can be classified into low- and high-intensity machines, which may be further classified into dry-feed and wet-feed separators.

Low-intensity separators are used to treat ferromagnetic materials and some highly paramagnetic minerals.

Low-intensity magnetic separation

Dry low-intensity magnetic separation is confined mainly to the concentration of coarse sands which are strongly magnetic, the process being known as cobbng, and often being carried out in drum separators. Below the 0.5 cm size range, dry separation tends to be replaced by wet methods, which produce much less dust loss and usually a cleaner product. Low-intensity wet separation is now widely used for purifying the magnetic medium in the dense medium separation process (see Chapter 11), as well as for the concentration of ferromagnetic sands.

Drum separators are the most common machines in current use for cleaning the medium in DMS
circuits and are widely used for concentrating finely ground iron ore. They consist essentially of a rotating non-magnetic drum (Figure 13.6) containing three to six stationary magnets of alternating polarity, although the Permos separator uses many small magnet blocks, whose direction of magnetisation changes in small steps. This is said to generate a very even magnetic field, requiring less magnetic material (Wasmuth and Unkelbach, 1991).

Although initially drum separators employed electromagnets, permanent magnets are used in modern devices, utilising ceramic or rare earth magnetic alloys, which retain their intensity for an indefinite period (Norrgran and Marin, 1994). Separation is by the “pick-up” principle. Magnetic particles are lifted by the magnets and pinned to the drum and are conveyed out of the field, leaving the gangue in the tailings compartment. Water is introduced into the machine to provide a current which keeps the pulp in suspension. Field intensities of up to 0.7 T at the pole surfaces can be obtained in this type of separator.

The drum separator shown in Figure 13.6 is of the concurrent type, whereby the concentrate is carried forward by the drum and passes through a gap, where it is compressed and dewatered before leaving the separator. This design is most effective for producing an extremely clean magnetic concentrate from relatively coarse materials and is widely used in dense medium recovery systems.

The separator shown in Figure 13.7 is of the counter-rotation type, where the feed flows in the opposite direction to the rotation. This type is used in roughing operations, where occasional surges in feed must be handled, where magnetic material losses are to be held to a minimum, while an extremely clean concentrate is not required, and when high solids loading is encountered.
Figure 13.7 Counter-rotation drum separator

Figure 13.8 shows a counter-current separator, where the tailings are forced to travel in the opposite direction to the drum rotation and are discharged into the tailings chute.

This type of separator is designed for finishing operations on relatively fine material, of particle size less than about 250 μm.

Drum separators are widely used to treat low-grade taconite ores, which contain 40–50% Fe, mainly as magnetite, but in some areas with hematite, finely disseminated in bands in hard siliceous rocks. Very fine grinding is necessary to free the iron minerals that produce a concentrate that requires pelletising before being fed to the blast furnaces.

In a typical flowsheet the ore is ground progressively finer, the primary grind usually being undertaken autogenously, or by rod milling, followed by magnetic separation in drum separators. The magnetic concentrate is reground and again treated in drum separators. This concentrate may be further reground, followed by a third stage of magnetic separation. The tailings from each stage of magnetic separation are either rejected or, in some cases, treated by spiral or cone concentrators to recover hematite.

At Palabora, the tailings from copper flotation (Figure 12.66) are deslimed, after which the +105 μm material is treated by Sala drum separators to recover 95% of the magnetite at a grade of 62% Fe.

The cross-belt separator (Figure 13.9) and disc separators once widely used in the mineral sands industry, particularly for recovering ilmenite from heavy mineral concentrates, are now considered obsolete. They are being replaced with rare earth roll magnetic separators and rare earth drum magnetic separators (Arvidson, 2001).

Rare earth roll separators use alternate magnetic and non-magnetic laminations as shown in Figure 13.4. Feed is carried onto the magnetic roll by means of a thin belt as shown in Figure 13.9, hence there is no bouncing or scattering of particles as they enter the magnetic zone, and they all enter the magnetic zone with the same horizontal velocity. These factors contribute to achieving a sharp separation. Roll speed can be adjusted over a wide range, allowing the product quality to be "dialled in".

Dry rare earth drum separators provide a fan of separated particles which can often be seen as distinct streams (Figure 13.10). The fan can be separated into various grades of magnetic product and a non-magnetic tailing. In some mineral sands applications, drum separators have been integrated with one or more rare earth rolls, arranged to treat the middlings particles from the drum as shown in Figure 13.10.

Figure 13.8 Counter-current separator
High-intensity separators

Very weakly paramagnetic minerals can only be effectively removed from an ore feed if high-intensity fields of 2 T and more can be produced (Svoboda, 1994).

Until the 1960s high-intensity separation was confined solely to dry ore, having been used commercially since about 1908.

Induced roll magnetic separators, IRMs (Figure 13.11), are widely used to treat beach sands, wolframite, tin ores, glass sands, and phosphate rock. They have also been used to treat feebly magnetic iron ores, principally in Europe. The roll, on to which the ore is fed, is composed of phosphated steel laminates compressed together on a non-magnetic stainless steel shaft. By using two sizes of lamination, differing slightly in outer diameter, the roll is given a serrated profile which promotes the high field intensity and gradient required. Field strengths of up to 2.2 T are attainable in the gap between feed pole and roll. Non-magnetic particles are thrown off the roll into the tailings compartment, whereas magnetics are gripped, carried out of the influence of the field.
and deposited into the magnetics compartment. The gap between the feed pole and rotor is adjustable and is usually decreased from pole to pole to take off successively more weakly magnetic products.

The setting of the splitter plates cutting into the trajectory of the discharged material is obviously of great importance.

In some cases IRMs are now being displaced by the new rare earth drum and roll separators.

Dry high-intensity separation is largely limited to ores containing little, if any, material finer than about 75 μm. The effectiveness of separation on such fine material is severely reduced by the effects of air currents, particle–particle adhesion, and particle–rotor adhesion.

Without doubt the greatest advance in the field of magnetic separation was the development of continuous wet high-intensity magnetic separators, WHIMS machines (Lawver and Hopstock, 1974). These reduce the minimum particle size for efficient separation allowing ores to be concentrated magnetically that cannot be concentrated effectively by dry high-intensity methods, because of the fine grinding necessary to ensure complete liberation of the magnetic fraction. In some flowsheets, expensive drying operations can be eliminated by using a wet concentration system.

Perhaps the most well-known WHIMS machine is the Jones separator, the design principle of which is utilised in many other types of wet separator used today.

The machine consists of a strong main frame (Figure 13.12) made of structural steel. The magnet yokes are welded to this frame, with the electromagnetic coils enclosed in air-cooled cases. The actual separation takes place in the plate boxes which are on the periphery of the one or two rotors attached to the central roller shaft. The feed, which is thoroughly mixed slurry, flows through the separator via fitted pipes and launders into the plate boxes (Figure 13.13), which are grooved to concentrate the magnetic field at the tip of the ridges. Feeding is continuous due to the rotation of the plate boxes on the rotors and the feed points are at the leading edges of the magnetic fields. Each rotor has two symmetrically disposed feed points.

The feebly magnetic particles are held by the plates, whereas the remaining non-magnetic slurry passes straight through the plate boxes and is...
collected in a launder. Before leaving the field any entrained non-magnetics are washed out by low-pressure water and are collected as a middlings product.

When the plate boxes reach a point midway between the two magnetic poles, where the magnetic field is essentially zero, the magnetic particles are washed out with high pressure scour water sprays operating at up to 5 bar (Figure 13.14). Field intensities of over 2 T can be produced in these machines. The production of a 1.5 T field requires an electric power consumption in the coils of 16 kW per pole. Of the 4 t of water used with every tonne of solids, approximately 90% is recycled.

Wet high-intensity magnetic separation has its greatest use in the concentration of low-grade iron ores containing hematite, where they frequently replace flotation methods, although the trend towards magnetic separation has been slow in North America, mainly due to the very high capital cost of such separators. It has been shown (White, 1978) that the capital cost of flotation equipment for concentrating weakly magnetic ore is about 20% that of a Jones separator installation, although flotation operating costs are about three times higher. Total cost depends on terms for capital depreciation; over 10 years or longer the high-intensity magnetic separator may be the most attractive process. Additional costs for water treatment may also boost the total for a flotation plant. Figure 13.15 shows a Jones separator in operation on a Brazilian iron ore plant.

Various other designs of wet high-intensity separator have been produced, a four-pole machine being manufactured by Boxmag-Rapid Ltd. The plate boxes in this design are an array of magnetic stainless steel "wedge-bars" similar to those used in fine screening (Figure 13.16).

In addition to their large-scale application for the recovery of hematite, wet high-intensity separators are now in operation for a wide range of duties, including removal of magnetic impurities from cassiterite concentrates, removal of fine magnetic from asbestos, removal of magnetic impurities from scheelite concentrates, purification of talc, the recovery of wolframite and non-sulphide molybdenum-bearing minerals from flotation tailings, and the treatment of heavy mineral beach sands. They have also been successfully used for the recovery of gold and uranium from cyanidation...
residues in South Africa (Corrans, 1984). These residues contain some free gold, while some of the fine gold is locked in sulphides, mainly pyrite, and in various silicate minerals. The free gold can be recovered by further cyanidation, while flotation can recover the pyritic gold. Magnetic separation can be used to recover some of the free gold, and much of the silicate-locked gold, due to the presence of iron impurities and coatings.

The paramagnetic properties of some sulphide minerals, such as chalcopyrite and marmatite, have been exploited by applying wet high-intensity magnetic separation to augment differential flotation processes commonly used to separate these minerals from less magnetic or non-magnetic sulphides (Tawil and Morales, 1985). Testwork showed that a Chilean copper concentrate could be upgraded from 23.8 to 30.2% Cu, at 87% recovery.
This was done by separating the chalcopyrite from pyrite in a field of 2 T. In Cu–Pb separation operations, it was found that chalcopyrite and galena could be effectively separated with field strengths as low as 0.8 T. When the process was applied to the de-coppering of a molybdenite concentrate, it was possible to reduce the copper content from 0.8 to 0.5% with over 97% Mo recovery.

**High-gradient magnetic separators**

In order to separate paramagnetic minerals of extremely low magnetic susceptibility, high magnetic forces must be generated. These forces can be produced by increasing the magnetic field strength, and in conventional high-intensity magnetic separators use is made of the ferromagnetic properties of iron to generate a high B-field (induced field) many hundreds of times greater than the applied H-field, with a minimum consumption of electrical energy. The working field occurs in air-gaps in the magnetic circuit, the disadvantage being that the volume of iron required is many times greater than the gap volume where separation takes place. The steel plates in a Jones separator, for example, occupy up to 60% of the process volume. Thus high-intensity magnetic separators using conventional iron circuits tend to be very massive and heavy in relation to their capacity. A large separator may contain over 200 t of iron to carry the flux, hence capital and installation costs are extremely high.

As iron saturates at around 2–2.5 T, conventional iron circuits are of little value for generating fields above about 2 T. Such fields can only be generated by the use of high H-fields produced in solenoids, but the energy consumption is extremely high and there are problems in cooling the solenoid.

An alternative is to increase the magnetic force by increasing the value of the magnetic field gradient. Instead of using one large convergent field in the gap of a magnetic circuit, the uniform field of a solenoid is used (Figure 13.17). The core, or working volume, is filled with a matrix of secondary poles, such as ball bearings, or wire wool, the latter filling only about 10% of the working volume. Each secondary pole, due to its high permeability, can produce a maximum field strength of the order of 2 T, but more importantly, each pole produces, in its immediate vicinity, high field gradients of up to 14 T mm\(^{-1}\). Thus a multitude of high gradients across numerous small gaps, centred around each of the secondary poles, is achieved.

The solenoid can be clad externally with an iron frame to form a continuous return path for the magnetic flux, thus reducing the energy consumption for driving the coil by a factor of about 2. The matrix is held in a canister into which the slurry is fed. As particles are captured, the ability of the matrix to extract particles is reduced. Periodically the magnetic field can be removed and the matrix flushed with water to remove the captured material.

An inherent disadvantage of high gradient separators is that an increase in field gradient necessarily reduces the working gap between secondary poles, the magnetic force having only a short reach, usually not more than 1 mm. It is therefore necessary to use gaps of only about 2 mm between poles, such that the matrix separators are best suited to the treatment of very fine particles. They are used mainly in the kaolin industry for removing micron-sized particles which contain iron. Several large separators, with the ferromagnetic matrix contained in baskets approximately 2 m in diameter.
are in commercial use in the United States and in Cornwall, England. They operate with fields of 2 T, and have capacities ranging between 10 and 80 t h⁻¹ depending on the final clay quality desired.

One of the most important factors which will affect coal preparation policy in the future is the environmental issue associated with acid rain and its link with sulphur emissions from fossil fuels. Sulphur occurs in coal in three forms. It is part of the coal substance (organic sulphur), or occurs as the minerals pyrite and marcasite, or as sulphates. The most important factor for the engineer is the pyritic sulphur content, as technology is not yet sufficiently developed to consider the removal of organic sulphur. If pyrite can be liberated by fine crushing to around 1 mm, then froth flotation or gravity methods can be used to remove it from the coal. However, if very fine crushing is necessary to liberate the pyrite, then high-gradient magnetic separation is a possibility. Increased international interest is at present being shown by coal preparation engineers in coal–liquid mixtures as a replacement for conventional hydrocarbon fuels such as diesel oil and natural gas. A typical coal–water mixture consists of pulverised coal of less than 50 microns particle size, and low ash content (2–6%) dispersed in an aqueous slurry, with a pulp density of between 50 and 80% solids. In order to produce these mixtures it is necessary to treat good quality coal by fine grinding and deep cleaning to remove ash and sulphur. High-gradient magnetic separation is capable of removing pyrite from pulverised coal, and much work is currently being performed on a variety of coal types (Lua and Boucher, 1990).

Superconducting separators

Undoubtedly the future developments and applications of magnetic separation in the mineral industry will lie in the use of high magnetic forces. Matrix separators with very high field gradients and multiple small working gaps can draw little advantage from field strengths above the saturation levels of the secondary poles. However, “open-gradient” separators, with large working volumes to deflect coarser particles at high capacity, rather than capture particles, as in high-gradient separators, need to use the highest possible field strengths in order to generate the high magnetic forces required to treat feebly paramagnetic particles. Field strengths in excess of 2 T can only be generated economically by the use of superconducting magnets (Kopp, 1991; Watson, 1994).

Certain alloys have the property of presenting no resistance to electric currents at extremely low temperatures. An example is niobium–titanium at 4.2 K, the temperature of liquid helium. Once a current commences to flow through a coil made from such a superconducting material, it will continue to flow without being connected to a power source, and the coil will become, in effect, a permanent magnet. Superconducting magnets can produce extremely intense and uniform magnetic fields, of up to 15 T. The main problem, of course, is in maintaining the extremely low temperatures, and in 1986 a Ba/La/Cu oxide composite was made superconductive at 35 K, promoting a race to prepare ceramic oxides with much higher superconducting temperatures (Malati, 1990). Unfortunately these materials are of a highly complex crystal structure, making them difficult to fabricate into wires. They also have a low current-carrying capacity, so it is likely that for the foreseeable future superconducting magnets will be made from ductile niobium alloys, embedded in a copper matrix.

In 1986 a superconducting high-gradient magnetic separator was designed and built by Eriez Magnetics to process kaolinite clay in the United States (Stefanides, 1986). This machine uses only about 0.007 kW in producing 5 T of flux, the ancillary equipment needed requiring another 20 kW. In comparison, a conventional 2 T high-gradient separator of similar throughput would need about 250 kW to produce the flux, and at least another 30 kW to cool the magnet windings.

The 5 T machine is an assembly of concentric components (Figure 13.18). A removable canister is installed in a processing chamber located at the centre of the assembly. This is surrounded by a double-walled, vacuum-insulated container that accommodates the superconductive magnet’s niobium/titanium-tantalum winding, and the liquid helium coolant. A thermal shield, cooled with liquid nitrogen to 77 K, limits radiation into the cryostat. In operation, the supply of slurry is periodically cut off, the magnetic field is shut down, and the canister backwashed with water to clear out accumulated magnetic contaminants.
An open-gradient drum magnetic separator with a superconducting magnet system has been operating commercially since the late 1980s (Unkelbach and Kellerwessel, 1985; Wasmuth and Unkelbach, 1991) (Figure 13.19). Although separation is identical to that in conventional drum separators, the magnetic flux density at the drum surface can reach over 4 T, generated by the superconductive magnet assembly within the drum.

**Electrical separation**

Electrical separation utilises the difference in electrical conductivity between the various minerals in the ore feed. Since almost all minerals show some difference in conductivity it would appear to represent the universal concentrating method. In practice, however, the method has fairly limited application, and its greatest use is in separating some of the minerals found in heavy sands from beach or stream placers (Dance and Morrison, 1992). The fact that the feed must be perfectly dry imposes limitations on the process, but it also suffers from the same great disadvantage as dry magnetic separation – the capacity is very small for finely divided material. For most efficient operation, the feed should be in a layer, one particle deep, which severely restricts the throughput if the particles are as small as, say, 75 μm.

---

**Figure 13.18** 5 T superconducting magnetic separator

**Figure 13.19** Superconducting drum separator:
1 – magnetic coils, 2 – radiation shield, 3 – vacuum tank, 4 – drum, 5 – plain bearing, 6 – helium supply, 7 – vacuum line, 8 – current supply
The first mineral separation processes utilising high voltage were virtually true electrostatic processes employing charged fields with little or no current flow. High-tension separation, however, makes use of a comparatively high rate of electrical discharge, with electron flow and gaseous ionisation having major importance. The theory of electrostatic and high-tension separations has been comprehensively reviewed by Kelly and Spottiswood (1989a–c) and Manouchehri et al. (2000).

The attraction of particles carrying one kind of charge towards an electrode of the opposite charge is known as the “lifting effect”, as such particles are lifted from the separating surface towards the electrode. Materials which have a tendency to become charged with a definite polarity may be separated from each other by the use of the lifting effect even though their conductivities may be very similar. As an example, quartz assumes a negative charge very readily and may be separated from other poor conductors by an electrode which carries a positive charge. Pure electrostatic separation is relatively inefficient, even with very clean mineral, and is sensitive to changes of humidity and temperature.

A large percentage of the commercial applications of high-tension separation has been made using the “pinning effect”, in which non-conducting mineral particles, having received a surface charge from the electrode, retain this charge and are pinned to the oppositely charged separator surface by positive–negative attraction. Figure 13.20 shows a laboratory high-tension separator, which makes use of the pinning effect to a high degree in combination with some lifting effect. Figure 13.21 shows the principle of separation diagrammatically.

The mixture of ore minerals, of varying susceptibilities to surface charge, is fed on to a rotating drum made from mild steel, or some other conducting material, which is earthed through its support bearings. An electrode assembly, comprising a brass tube in front of which is supported a length of fine wire, spans the complete length of the roll, and is supplied with a fully rectified DC supply of up to 50 kV, usually of negative polarity. The voltage supplied to the assembly should be such that ionisation of the air takes place. This can often be seen as a visible corona discharge. Arcing between the electrode and the roll must be avoided, as this destroys the ionisation. When ionisation occurs, the minerals receive a spray discharge of electricity which gives the poor conductors a high surface charge, causing them to be attracted to and pinned to the rotor surface. The particles of relatively high conductivity do not become charged as rapidly, as the charge rapidly dissipates through the particles to the earthed rotor. These particles of higher conductivity follow a path, when leaving the rotor, approximating to the one which they would assume if there were no charging effect at all.
The electrode assembly is designed to create a very dense high-voltage discharge. The fine wire of the assembly is placed adjacent to and parallel to the large diameter electrode and is mechanically and electrically in contact with it. This fine wire tends to discharge readily, whereas the large tube tends to have a short, dense, non-discharging field. This combination creates a very strong discharge pattern which may be "beamed" in a definite direction and concentrated to a very narrow arc. The effect on the minerals passing through the beam is very strong and is due largely to gaseous ions which are created due to the high-voltage gradient in the field of the corona.

A combination of the effects of pinning and lifting can be created by using a static electrode large enough to preclude corona discharge, following the electrode. The conducting particles, which are flung from the rotor, are attracted to this electrostatic electrode, and the compound process produces a very wide and distinct separation between the conducting and the non-conducting particles.

Table 13.1 shows typical minerals which are either pinned to or thrown from the rotor during high-tension separation.

<table>
<thead>
<tr>
<th>Minerals pinned to rotor</th>
<th>Minerals thrown from rotor</th>
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<tbody>
<tr>
<td>Apatite</td>
<td>Cassiterite</td>
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<tr>
<td>Barite</td>
<td>Chromite</td>
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<tr>
<td>Calcite</td>
<td>Diamond</td>
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<tr>
<td>Corundum</td>
<td>Fluorspar</td>
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<td>Garnet</td>
<td>Galena</td>
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<td>Gypsum</td>
<td>Gold</td>
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<td>Kyanite</td>
<td>Hematite</td>
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<td>Monazite</td>
<td>Ilmenite</td>
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<td>Quartz</td>
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<td>Scheelite</td>
<td>Magnetite</td>
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<td>Sillimanite</td>
<td>Pyrite</td>
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<tr>
<td>Spinel</td>
<td>Rutile</td>
</tr>
<tr>
<td>Tourmaline</td>
<td>Sphalerite</td>
</tr>
<tr>
<td>Zircon</td>
<td>Stibnite</td>
</tr>
<tr>
<td></td>
<td>Tantalite</td>
</tr>
<tr>
<td></td>
<td>Wolframite</td>
</tr>
</tbody>
</table>

To cater for such an extensive range of minerals, all the parameters influencing separation must be readily adjusted while the separator is performing. These variables include the roll speed, the position of the electrode wire with respect to the electrode tube, the position of the electrode assembly with respect to the roll, variation of the DC voltage and polarity, the splitter plate position, the feed rate, and heating of the feed. Heating the feed is important, since best results are generally obtained only with very dry material. This is particularly difficult in high humidity regions. It is not often that a single pass will sufficiently enrich an ore and Figure 13.22 shows a typical flowsheet, where the falling particles are deflected to lower sets of rollers and electrodes until the required separation has taken place.

High-tension separators operate on feeds containing particles of between 60 and 500\(\mu\)m in diameter. Particle size influences separation behaviour, as the surface charges on a coarse grain are lower in relation to its mass than on a fine grain. Thus a coarse grain is more readily thrown from the roll surface, and the conducting fraction often contains a small proportion of coarse non-conductors. Similarly, the finer particles are most influenced by the surface charge, and the non-conducting fraction often contains some fine conducting particles.
Final cleaning of these products is often carried out in purely electrostatic separators, which employ the “lifting effect” only. Modern electrostatic separators are of the plate or screen type, the former being used to clean small amounts of non-conductors from a predominantly conducting feed, while the screen separators remove small amounts of conductors from a mainly non-conducting feed. The principle of operation is the same for both types of separator. The feed particles gravitate down a sloping, grounded plate into an electrostatic field induced by a large, oval, high-voltage electrode (Figure 13.23).

This is the converse of the separation which takes place in the high-tension separators, where most effective separation of fine non-conductors from coarse conductors takes place; a combination of high-tension separators as primary roughers, followed by final cleaning in electrostatic separators, is therefore used in many flowsheets. Since the magnitude of the forces involved in electrostatic separation is very low, the separators are designed for multiple passes of the non-conductors (Figure 13.24).

The electrostatic field is effectively shorted through the conducting particles, which are lifted towards the charged electrode in order to decrease the energy of the system. Non-conductor grains are poorly affected by the field. The fine grains are most affected by the lifting force, and so fine conductors are preferentially lifted to the electrode, whereas coarse non-conductors are most efficiently rejected.

**Figure 13.23** (a) Plate and (b) screen electrostatic separators

**Figure 13.24** Plate electrostatic separator with two-start, ten electrodes
High tension roll (HTR) and electrostatic plate (ESP) separators have been the mainstay of the mineral sands industry for the last 50 years. Very little development of the machines has occurred in that period, their generally poor single pass separation has been tolerated, and overcome by using multiple machines and multiple recycle streams. However, in the last few years innovative new designs have started to appear, from new as well as established manufacturers. OreKinetics has introduced the new CoronaStat and UltraStat machines. These machines which are significant developments of existing HTR and ESP machines employ additional static electrodes which improve efficiency of separation. Unlike existing machines the static electrodes are not exposed, making the machines much safer to operate.

Existing manufacturers have also introduced new electrical separation machines. Roche Mining (MT) have developed the Carara HTR separator which incorporates an additional insulated plate static electrode (Germain et al., 2003). Outokumpu Technology have developed the eForce HTR separator, which also incorporates additional static electrodes, as well as an electrostatic feed classifier (Elder and Yan, 2003).

These new generation machines will change the way heavy minerals plants are designed. Their improved efficiencies will reduce the number of stages required, and hence the capital cost of the plant.

It was mentioned earlier that there is some possibility of an overlap in the application of magnetic and high-tension separators, particularly in the processing of heavy mineral sand deposits. Table 13.2. shows some of the common minerals present in such alluvial deposits, along with their properties, related to magnetic and high-tension separation. Mineral sands are commonly mined by floating dredges, feeding floating concentrators at up to 2000 t h⁻¹ or more. Figure 13.25

<table>
<thead>
<tr>
<th>Magnetics</th>
<th>Non-magnetics</th>
</tr>
</thead>
<tbody>
<tr>
<td>Magnetite – T</td>
<td>Rutile – T</td>
</tr>
<tr>
<td>Ilmenite – T</td>
<td>Zircon – P</td>
</tr>
<tr>
<td>Garnet – P</td>
<td>Quartz – P</td>
</tr>
<tr>
<td>Monazite – P</td>
<td></td>
</tr>
</tbody>
</table>

T = thrown from high-tension separator surface.  
P = pinned to high-tension separator surface.
shows a typical dredge and floating concentrator operating at Richards Bay in South Africa. Such concentrators, consisting of a complex circuit of sluices, spirals, or Reichert cones, upgrade the heavy mineral content to around 90%, the feed grades varying from less than 2%, up to 20% heavy mineral in some cases. The gravity concentrate is then transferred to the separation plant for recovery of the minerals by a combination of gravity, magnetic, and high-tension methods.

Flowsheets vary according to the properties of valuable minerals present, wet magnetic separation often preceding high-tension separation where magnetic ilmenite is the dominant mineral. A generalised flowsheet for such a separation is shown in Figure 13.26. Low-intensity drum separators remove any magnetite from the feed, after which high-intensity wet magnetic separators separate the monazite and ilmenite from the zircon and rutile. Drying of these two fractions is followed by high-tension separation to produce final separation, although further cleaning is sometimes carried out by electrostatic separators. For example, screen electrostatic separators may be used to clean the zircon and monazite concentrates, removing fine conducting particles from these fractions. Similarly, plate electrostatic separators could be used to reject coarse non-conducting particles from the rutile and ilmenite concentrates.

Figure 13.27 shows a simplified circuit used by Tiwest Joint Venture, on the west coast of Australia (Benson et al., 2001).

The heavy mineral concentrate is first separated into conductor and non-conductor streams using HTR separators. The conductors are treated using crossbelt and roll magnet separators to remove the ilmenite as a magnetic product. The non-magnetic stream is cleaned with high intensity roll and rare earth magnets to separate the weakly magnetic leucoxene from non-magnetic rutile. The non-conductors undergo another stage of wet gravity separation to remove quartz and other low density contaminants, before sizing and cleaning using HTR, ESP and Ultrastat separators to produce fine and coarse zircon products.

Similar flowsheets are used in South-East Asia for the treatment of alluvial cassiterite deposits, which are also sources of minerals such as ilmenite, monazite and zircon.

Magnetic separators are commonly used for upgrading low-grade iron ores, wet high-intensity separation often replacing the flotation of hematite. A combination of magnetic and high-tension separation has been used at the Scully Mine of Wabush Mines in Canada (Anon., 1974). The ore, grading about 35% Fe, is a quartz-specular hematite-magnetite schist, and after crushing and autogenous grinding to -1mm, is fed to banks of rougher and cleaner spiral concentrators (Figure 13.28).

The spiral concentrate is filtered and dried, and cleaned in high-tension roll separators. The spiral tailings are thickened, and further treated by magnetic drum separators to remove residual magnetite, followed by Jones wet high-intensity separators, which remove any remaining hematite. The magnetic concentrates are classified and dried, and blended with the high-tension product, to give a final concentrate of about 66% Fe. Cleaning of only the gravity tailings by magnetic separation is preferred, as relatively small amounts of magnetic concentrate have to be dealt with, the bulk of the material being unaffected by the magnetic field. Similarly, relatively little material is pinned to the rotor in the high-tension treatment of the gravity concentrate, the iron minerals being unaffected by the ionic field.
Figure 13.27  Simplified mineral sands circuit used by Tiwest Joint Venture (from Benson et al., 2001)
Figure 13.28 Flowsheet of Scully concentrator

References


Ore sorting

Introduction

Ore sorting is the original concentration process, having probably been used by the earliest metal workers several thousand years ago. It involves the appraisal of individual ore particles and the rejection of those particles that do not warrant further treatment.

Hand sorting has declined in importance due to the need to treat large quantities of low-grade ore which requires extremely fine grinding. Hand sorting of some kind, however, is still practised at some mines, even though it may only be the removal of large pieces of timber, tramp iron, etc. from the run-of-mine ore.

Electronic ore-sorting equipment was first produced in the late 1940s, and although its application is fairly limited, it is an important technique for the processing of certain minerals (Sassos, 1985; Salter and Wyatt, 1991; Sivamohan and Forssberg, 1991; Collins and Bonney, 1998; Arvidson, 2002).

Electronic sorting principles

Sorting can be applied to pre-concentration, in which barren waste is eliminated to reduce the tonnage reporting to the downstream concentration processes, such as in uranium or gold ore sorting, or to the production of a final product, such as in limestone or diamond sorting. The ore must be sufficiently liberated at a coarse size (greater than 5–10 mm) to allow barren waste to be discarded without significant loss of value. Preconcentration by sorting is seen as a method of improving the sustainability of mineral processing operations by reducing the consumption of energy and water in grinding and concentration, and achieving more benign tailings disposal (Cutmore and Ebehardt, 2002).

Many rock properties have been used as the basis of electronic sorting, including reflectance and colour in visible light (magnesite, limestone, base metal and gold ores, phosphates, talc, coal), ultraviolet (scheelite), natural gamma radiation (uranium ore), magnetism (iron ore), conductivity (sulphides), and X-Ray luminescence (diamonds). Infrared, Raman, microwave attenuation, and other properties have also been tested.

Electronic sorters inspect the particles to determine the value of some property (e.g. light reflectance) and then eject those particles which meet some criterion (e.g. light vs dark rocks). Either valuables or waste may be selected for ejection. It is essential, therefore, that a distinct difference in the required physical property is apparent between the valuable minerals and the gangue.

The particle surfaces must be thoroughly washed before sorting, so that blurring of the signal does not occur and, as it is not practical to attempt to feed very wide rock size ranges to a single machine, the feed must undergo preliminary sizing. The ore must be fed in a monolayer, as display of individual particles to the sorting device must be effected.

Photometric sorting is the mechanised form of hand-sorting, in which the ore is divided into components of differing value by visual examination (Arvidson, 2002).

The basis of the photometric sorter (Figure 14.1) is a laser light source and sensitive photomultiplier, used in a scanning system to detect light reflected from the surfaces of rocks passing through the sorting zone (Figure 14.2). Electronic circuitry analyses the photomultiplier signal, which changes with the intensity of the reflected light and produces control signals to actuate the appropriate valves of an air-blast rejection device to remove certain particles selected by means of the analysing process. The sorter is fully automatic and can be attended by one operator on a part-time basis. Typical throughput per machine ranges from 25 t h⁻¹ for a -25 + 5 mm feed to 300 t h⁻¹ for a -300 + 80 mm material.
Examples

The Gunson’s Sortex MP80 machine was probably the first sorter to employ microprocessor technology (Anon., 1980). The sorter handled minerals in the size range 10–150 mm at feed rates of up to 150 t h⁻¹.

The RTZ Ore Sorters Model 16 photometric sorter has been used successfully in industry since 1976 on a wide range of ore types (Anon., 1981b).

An RTZ ore-sorting machine at the Doornfontein mine in South Africa was used for treating gold ores (Keys et al., 1974). Rocks having white or grey quartz pebbles in a darker matrix are accepted, while quartzite, ranging from light green, through olive green to black, is rejected. Most of the gold occurs in rocks in the “accept” category. Uniform distribution of the ore entering the sorter is achieved by the use of tandem vibrating feeders and the ore is washed on the second feeder to remove slimes which may affect the light-reflecting qualities.

The modern photometric sorter is typified by the UltraSort UFS120, which is used in the processing of magnesite, feldspar, limestone, and talc. Ore passes from a vibrating feeder to high pressure water sprays and counterweight feeder where water is removed and the rocks are accelerated to form a monolayer. They drop onto a short conveyor
moving at 2 m/s from where they pass via a high speed 5 m/s “slinger” conveyor into free fall, now well separated. The rock layer, 0.8–1.2 m wide, is scanned by a laser beam at up to 4000 times per second, and the reflection analysed in less than 0.25 μs by photomultiplier tubes and high speed parallel processors operating in excess of 80 MB/s. One or more of 120 air ejectors are fired to divert the value or waste past a cutter and into the accept/reject bins. As the position of the rock is accurately identified, and the ejector firing duration is less than 1 ms, the sorter can operate very selectively. A range of lasers of different wavelengths can be used, from ultraviolet to infrared, to achieve optimum discrimination.

Scanning video cameras can be used in place of the scanning laser and photomultiplier tubes, to provide more subtle discrimination of rock properties using image analysis techniques.

Electronic sorting has been employed in diamond recovery since the 1960s, initially using simple optical sorters and more recently machines based on the fact that diamonds luminesce when irradiated by X-Rays (Anon., 1971; Rylatt and Popplewell., 1999). X-Ray sorters are used in almost all diamond operations for the final stages of recovery after the ore has been concentrated by DMS (Chapter 11). They replace grease separation (Chapter 12) which is now used only in rare cases where the diamonds luminesce weakly or to audit the X-Ray tailings. Luminescence is a more consistent diamond property than oleophilicity, and sorters are more secure than grease belts or tables. Figure 14.3 shows an early dry X-Ray sorter, in which the DMS concentrates are exposed to a beam of X-Rays in free fall from a conveyor belt, the luminescence detected by photomultiplier tubes and the diamonds ejected by air ejectors. Both dry and wet X-Ray machines are now available, and the process is usually multi-stage to ensure efficient rejection of waste with very high diamond recoveries.

Radiometric sorting has been used to preconcentrate uranium ore in South Africa (Anon., 1981c), Namibia, Australia (Bibby, 1982), and Canada. A sorter installed at the Rossing Mine in Namibia (Gordon and Heuer, 2000) detects gamma radiation from the higher grade ore pieces using scintillation counters comprising NaI crystals and photomultiplier tubes mounted under the belt (Figure 14.4). Lead shielding is used to achieve improved resolution of detection. A laser-based optical system similar to that used in photometric sorters is used to determine rock position and size for ejection, and can be adapted to determine additional optical characteristics of the rocks.

Several other physical properties of ores and minerals have been exploited in a range of sorting machines.

Neutron absorption separation has been used for the sorting of boron minerals (Mokrousov et al., 1975). The ore is delivered by a conveyor belt between a slow neutron source and a scintillation neutron detector. The neutron flux attenuation by the ore particles is detected and used as the means of sorting. The method is most applicable in the size range 25–150 mm. Boron minerals are easy to sort by neutron absorption since the neutron capture cross-section of the boron atom is very large compared with those of common associated elements and thus the neutron absorption is almost proportional to the boron content of the particles.

Photoneutron separation is recommended for the sorting of beryllium ores, since when a beryllium isotope in the mineral is exposed to gamma radiation of a certain energy, a photoneutron is released and this may be detected by scintillation or by a gas counter.

The RTZ Ore Sorters Model 19 sorter measured conductivity and magnetic properties and had application to a wide variety of ores including sulphides, oxides, and native metals (Anon., 1981a). The machine treated 25–150 mm rocks at up to 120 t/h. Such systems employ a tuned coil under the belt which is influenced by the conductivity and/or magnetic susceptibility of the rocks in its proximity. Phase shift and amplitude are used to decide on acceptance or rejection. The radiometric sorter shown in Figure 14.4 can be adapted to conductivity and magnetic sorting by replacing the scintillometers with 40 electromagnetic sensors.

Outokumpu developed the “Precon” sorter and installed it at its Hammaslathi copper mine, now closed (Kennedy, 1985). It used gamma scattering analysis to evaluate the total metal content, and had a capacity of 7 t/h for 35 mm lumps rising to 40 t/h for 150 mm lumps. It preconcentrated primary crushed ore, rejecting about 25% as waste grading 0.2% copper, compared with an average feed grade of 1.2%.
Microwave attenuation has been used to sort diamond-bearing kimberlite from waste rock (Salter et al., 1989). The development was notable for the first use of high speed pulsed water ejectors.

Equipment to sort asbestos ore has been developed (Collier et al., 1973). The detection technique was based on the low thermal conductivity of asbestos fibres and used sequential heating and infra-red scanning to detect the asbestos seams. A similar machine was installed at King Island Scheelite in Tasmania, where the scheelite was sensed by its fluorescence under ultra-violet radiation.
References

Dewatering

Introduction

With few exceptions, most mineral-separation processes involve the use of substantial quantities of water and the final concentrate has to be separated from a pulp in which the water–solids ratio may be high.

Dewatering, or solid–liquid separation, produces a relatively dry concentrate for shipment. Partial dewatering is also performed at various stages in the treatment, so as to prepare the feed for subsequent processes.

Dewatering methods can be broadly classified into three groups:

1. sedimentation;
2. filtration;
3. thermal drying.

Sedimentation is most efficient when there is a large density difference between liquid and solid. This is always the case in mineral processing where the carrier liquid is water. Sedimentation cannot always be applied in hydrometallurgical processes, however, because in some cases the carrier liquid may be a high-grade leach liquor having a density approaching that of the solids. In some cases, filtration may be necessary.

Dewatering in mineral processing is normally a combination of the above methods. The bulk of the water is first removed by sedimentation, or thickening, which produces a thickened pulp of perhaps 55–65% solids by weight. Up to 80% of the water can be separated at this stage. Filtration of the thick pulp then produces a moist filter cake of between 80 and 90% solids, which may require thermal drying to produce a final product of about 95% solids by weight.

Sedimentation

Rapid settling of solid particles in a liquid produces a clarified liquid which can be decanted, leaving a thickened slurry, which may require further dewatering by filtration.

The settling rates of particles in a fluid are governed by Stokes’ or Newton’s laws, depending on the particle size (Chapter 9). Very fine particles, of only a few microns diameter, settle extremely slowly by gravity alone, and centrifugal sedimentation may have to be performed. Alternatively, the particles may be agglomerated, or flocculated, into relatively large lumps, called flocs, that settle out more rapidly.

Coagulation and flocculation

Coagulation causes extremely fine colloidal particles to adhere directly to each other. All particles exert mutual attraction forces, known as London–Van der Waals’ forces, which are effective only at very close range. Normally, the adhesion due to these forces is prevented by the presence around each particle of an electrically charged atmosphere, which generates repulsion forces between particles approaching each other. There is, therefore, in any given system a balance between the attractive forces and the electrical repulsion forces present at the solid–liquid interface (Figure 15.1).

In any given system the electrical charges on the particle surfaces will be of the same sign, aqueous suspensions of pH 4 and above generally being negative. Positively charged surfaces occur mainly in strong acid solutions.

The repulsion forces not only prevent coagulation of the particles, but also retard their settlement by keeping them in constant motion, this effect being more pronounced the smaller the particle.
Coagulants are electrolytes having an opposite charge to the particles, thus causing charge neutralisation when dispersed in the system, allowing the particles to come into contact and adhere as a result of molecular forces. Inorganic salts have long been used for this purpose, and as counterions in aqueous systems are most frequently positively charged, salts containing highly charged cations, such as Al\(+\+\), Fe\(+\+\), and Ca\(+\+\), are mainly used. Lime, or sulphuric acid, depending on the surface charge of the particles, can also be used to cause coagulation. Most pronounced coagulation occurs when the particles have zero charge in relation to the suspending medium, this occurring when the zeta potential is zero. The nature of the zeta potential can be seen from Figure 15.2, which shows a model of the electrical double layer at the surface of a particle (Moss, 1978). The surface shown has a negative charge, such that positive ions from solution will be attracted to it, forming a bound layer of positive ions, known as the Stern layer and a diffuse layer of counter ions decaying in concentration with increasing distance until the solution equilibrium concentration is attained. These layers of ions close to the surface constitute the electrical double layer. When a particle moves in the liquid, shear takes place between the bound layer, which moves with the particle, and the diffuse layer, the potential at the plane of shear being known as the zeta potential. The magnitude of the zeta potential depends on the surface potential and the concentration and charge of the counter-ions. In general, the greater the counter-ion charge and counter-ion concentration, the lower is the zeta potential, although ions of high charge may cause complete charge reversal; therefore optimum doses of electrolyte are critical.

Flocculation involves the formation of much more open agglomerates than those resulting from coagulation and relies upon molecules of reagent acting as bridges between separate suspended particles (Hunter and Pearse, 1982; Pearse, 1984; Hogg, 2000). The reagents used to form the “bridges” are long-chain organic polymers, which were formerly natural minerals, such as starch, glue, gelatine, and guar gum, but which are now increasingly synthetic materials, loosely termed polyelectrolytes. The majority of these are anionic in character but some of them are non-ionic, and some cationic, but these form a minor proportion of the commercially available products of today’s flocculant market. Inorganic salts are not able to perform this bridging function, but they are sometimes used in conjunction with an organic reagent as a cheaper means of charge neutralisation, although an ionic polyelectrolyte can and often does perform both functions.

The polyacrylamides, which vary widely in molecular weight and charge density, are extensively used as flocculants (Mortimer, 1991). The charge density refers to the percentage of the acrylic
monomer segments which carry a charge. For instance, if the polymer is uncharged it comprises $n$ similar segments of the acrylic monomer. The polymer is thus a homopolymer polyacrylamide.

**Chemical formula**

If the acrylic monomer is completely hydrolysed with NaOH, the product comprises $n$ segments of sodium acrylate – an anionic polyelectrolyte, having a charge density of 100%.

Charge density may be controlled in manufacture between the limits 0 and 100%, to produce a polyacrylamide of anionic character, weak or strong, depending on the degree of hydrolysis.

By similar chemical reactions, polymers of cationic character can be produced. Much of the development, to date, of the polyacrylamide family of products has been directed towards providing products of increasingly higher molecular weight, whilst maintaining the high degree of water solubility required for use in solid–liquid separation. It is now possible to obtain water soluble products with a wide range of ionic character varying from 100% cationic content through non-ionic to 100% anionic content and with molecular weights from several thousand to over 10 million (Moody, 1992).

It would be expected that, since most suspensions encountered in the minerals industry contain negatively charged particles, cationic polyelectrolytes, where the cation adsorbs to the particles, would be most suitable. Although this is true for charge neutralisation purposes, and attraction of the polymer to the particle surface, it is not necessarily true for the “bridging” role of the flocculant. For bridging, the polymer must be strongly adsorbed, and this is promoted by chemical groups having good adsorption characteristics, such as amide groups. The majority of commercially available polyelectrolytes are anionic, since these tend to be of higher molecular weight than the cationics, and are less expensive.

The mode of action of the anionic polyacrylamide depends on a segment of the very long molecule being adsorbed on the surface of a particle, leaving a large proportion of the molecule free to be adsorbed on another particle, so forming an actual molecular linkage, or bridge, between particles (Figure 15.3).

While only one linkage is shown in Figure 15.3, in practice many such interparticle bridges are formed, linking a number of particles together. The factors influencing the degree of flocculation are the efficiency or strength of adsorption of the polymer on the surface, the degree of agitation during flocculation and the subsequent agitation, which can result in breakdown of flocs (Lightfoot, 1981; Owen et al., 2002).

Although the addition of flocculants to a slurry can lead to significant improvements in sedimentation characteristics, it also affects dewatering behaviour, flocculation generally being detrimental to final consolidation of the sediment. It is usually beneficial, however, to filtration processes, and flocculants are widely used as filter aids. However, the specific requirements of a flocculant used to promote sedimentation are not necessarily the same as for one used as a filter aid. The behaviour of the flocculated suspension and the performance of solid–liquid separations are determined by the size of the flocs and by their structure. Large flocs promote settling and are desirable for clarification and thickening. Floc density is of secondary importance in these processes. Conversely, dense flocs

![Figure 15.3 Action of an anionic polyelectrolyte](image-url)
are most appropriate for consolidation of the sediment, and size is of lesser importance in this stage. Therefore the optimisation of solid–liquid separation processes requires careful control of floc size and structure. The maximum effect of a flocculant is achieved at an optimum dosage rate and pH; excess polymer can cause dispersion of the particles due to floc breakdown. Physical factors are also of great importance, growth and development of the flocs being affected by particle–particle collisions and hydrodynamic interactions (Hogg et al., 1987). Laboratory batch cylinder tests are commonly used to assess the effectiveness of flocculants to enhance the settling rate of suspensions. Reproducibility of such tests is often poor, depending on factors such as number of cylinder inversions and cylinder diameter. A new method using vertically mounted concentric rotating cylinders (Couette geometry) has been found to overcome these problems and give highly reproducible results (Farrow and Swift, 1996).

Due to the fragile nature of the flocs, flocculating agents are not successful with hydrocyclones, while success with centrifuges can only be achieved with special techniques for a limited range of applications. Even pumping of the flocculated slurry may destroy the flocs due to rupture of the long-chain molecules.

Polyelectrolytes are normally made up of stock solutions of about 0.5–1%, which are diluted to about 0.01% before adding to the slurry. The diluted solution must be added at enough points in the stream to ensure its contact with every portion of the system. A shower pipe is frequently used for this purpose (Figure 15.4). Recent work has shown that the age of the stock solution can have a significant effect on flocculant performance. Dosage decreased with solution age, the optimum age being 72 h (Owen et al., 2002).

Mild agitation is essential at the addition points, and shortly thereafter, to assist in flocculant dispersion in the process stream. Care should be taken to avoid severe agitation after the flocs have been formed.

Selective flocculation

The treatment of finely disseminated ores often results in the production of ultra-fine particles, or slimes, which respond poorly to conventional separation techniques, and are often lost in the process tailings. Selective flocculation of the desired minerals in the pulp, followed by separation of the aggregates from the dispersed material, is a potentially important technique, although plant applications are at present rare (Attia, 1992). Although attempts have been made to apply selective flocculation to a wide range of ore types, the bulk of the work has been concerned with its application to the treatment of clays, iron, phosphate, and potash ores. A prerequisite for the process is that the mineral mixture must be stably dispersed prior to the addition of a high molecular weight polymer, which selectively adsors on only one of the constituents of the mixture. Selective flocculation is then followed by removal of the flocs of one component from the dispersion.

The greatest amount of work on selective flocculation has been concerned with the treatment of fine-grained non-magnetic oxidised taconites, which has led to the development of Cleveland Cliffs Iron Company’s 10 Mt/yr operation in the United States. The finely intergrown ore is autogenously ground to 85%–25 μm with caustic soda and sodium silicate, which act as dispersants for the fine silica. The ground pulp is then conditioned with a corn-starch flocculant which selectively flocculates the hematite. About one-third of the fine silica is removed in a de-slime thickener, together with a loss of about 10% of the iron values. Most of the remaining coarse silica is removed from the flocculated underflow by reverse flotation, using an amine collector (Paananen and Turcotte, 1980; Siirak and Hancock, 1988).

Gravity sedimentation

Gravity sedimentation or thickening is the most widely applied dewatering technique in mineral
processing, and it is a relatively cheap, high-
capacity process, which involves very low shear
forces, thus providing good conditions for floccu-
lation of fine particles.

The thickener is used to increase the concentra-
tion of the suspension by sedimentation, accompa-
nied by the formation of a clear liquid. In most
cases the concentration of the suspension is high
and hindered settling takes place. Thickeners may
be batch or continuous units, and consist of rela-
tively shallow tanks from which the clear liquid
is taken off at the top, and the thickened suspen-
sion at the bottom (Suttill, 1991; Schoenbrunn and
Laros, 2002). The clarifier is similar in design,
but is less robust, handling suspensions of much
lower solid content than the thickener (Seifert and

The continuous thickener consists of a cylin-
drical tank, the diameter ranging from about 2 to
200 m in diameter, and of depth 1–7 m. Pulp is
fed into the centre via a feed-well placed up to
1 m below the surface, in order to cause as little
disturbance as possible (Figure 15.5). The clari-
fied liquid overflows a peripheral launder, while
the solids which settle over the entire bottom of
the tank are withdrawn as a thickened pulp from
an outlet at the centre. Within the tank are one or
more rotating radial arms, from each of which are
suspended a series of blades, shaped so as to rake
the settled solids towards the central outlet. On most
modern thickeners these arms rise automatically if
the torque exceeds a certain value, thus preventing
damage due to overloading. The blades also assist
the compaction of the settled particles and produce
a thicker underflow than can be achieved by simple
settling. The solids in the thickener move contin-
uously downwards, and then inwards towards the
thickened underflow outlet, while the liquid moves
upwards and radially outwards. In general, there is
no region of constant composition in the thickener.

Thickener tanks are constructed of steel,
concrete, or a combination of both, steel being
most economical in sizes of less than 25 m in
diameter. The tank bottom is often flat, while the
mechanism arms are sloped towards the central
discharge. With this design, settled solids must
“bed-in” to form a false sloping floor. Steel floors
are rarely sloped to conform with the arms because
of expense. Concrete bases and sides become more
common in the larger-sized tanks. In many cases
the settled solids, because of particle size, tend to
slump and will not form a false bottom. In these
cases the floor should be concrete and poured to
match the slope of the arms. Tanks may also be
constructed with sloping concrete floors and steel
sides, and earth bottom thickeners are in use, which
are generally considered to be the lowest cost solu-
tion for thickener bottom construction (Hsia and
Reinmiller, 1977).

The method of supporting the mechanism
depends primarily on the tank diameter. In rela-
tively small thickeners, of diameter less than about
45 m, the drive head is usually supported on a
superstructure spanning the tank, with the arms
being attached to the drive shaft. Such machines
are referred to as bridge or beam thickeners
(Figure 15.6). The underflow is usually drawn from
the apex of a cone located at the centre of the
sloping bottom.

A common arrangement for larger thickeners, of
up to about 180 m in diameter, is to support the
drive mechanism on a stationary steel or concrete
centre column. In most cases, the rake arms are
attached to a drive cage, surrounding the central
column, which is connected to the drive mecha-
nism. The thickened solids are discharged through
an annular trench encircling the centre column
(Figure 15.7). Figure 15.8 shows an 80 m diameter
thickener of this type.

In the traction thickener, a single long arm
is mounted with one end on the central support
column while to the other are fixed traction wheels
that run on a rail on top of the tank wall. The wheels
are driven by motors which are mounted on the
end of the arm and which therefore travel around
with it. This is an efficient and economical design.

Figure 15.5 Flow in a continuous thickener
since the torque is transmitted through a long lever arm by a simple drive. They are manufactured in sizes ranging from 60 m to approximately 120 m in diameter.

*Cable thickeners* have a hinged rake arm fastened to the bottom of the drive cage or centre shaft. The hinge is designed to give simultaneous vertical and horizontal movement of the rake arm. The rake arm is *pulled* by cables connected to a torque or drive arm structure, which is rigidly connected to the centre shaft at a point just below the liquid level. The rake is designed to automatically lift when the torque developed due to its motion through the sludge rises. This design allows the rake arm to find its own efficient working level in the sludge, where the torque balances the rake weight.

In all thickeners the speed of the raking mechanism is normally about $8 \text{ m min}^{-1}$ at the perimeter, which corresponds to about $10 \text{ rev h}^{-1}$ for a 15 m diameter thickener. Energy consumption is thus extremely low, such that even a 60 m unit may require only a 10 kW motor. Wear and maintenance costs are correspondingly low.

The underflow is usually withdrawn from the central discharge by pumping, although in clarifiers the material may be discharged under the hydrostatic head in the tank. The underflow is usually collected in a sludge-well in the centre of the tank bottom, from where it is removed via piping through an underflow tunnel. The underflow lines should be as short and as straight as possible to reduce the risk of choking, and this can be achieved, with large tanks, by taking them up
Figure 15.7 Thickener with column supported by centre column

Figure 15.8 80 m diameter centre-column-supported thickener
from the sludge-well through the centre column to pumps placed on top, or by placing the pumps in the base of the column and pumping up from the bottom. This has the advantage of dispensing with the expensive underflow tunnel. A development of this is the caisson thickener, in which the centre column is enlarged sufficiently to house a central control room; the pumps are located in the bottom of the column, which also contains the mechanism drive heads, motors, control panel, underflow suction, and discharge lines. The interior of the caisson can be a large heated room. The caisson concept has lifted the possible ceiling on thickener sizes; at present they are manufactured in sizes up to 180 m in diameter.

Underflow pumps are often of the diaphragm type (Anon., 1978). These are positive action pumps for medium heads and volumes, and are suited to the handling of thick viscous fluids. They can be driven by electric motor through a crank mechanism, or by directly acting compressed air. A flexible diaphragm is oscillated to provide suction and discharge through non-return valves, and variable speed can be achieved by changing either the oscillating frequency or the stroke. In some plants, variable-speed pumps are connected to nucleonic density gauges on the thickener underflow lines, which control the rate of pumping to maintain a constant underflow density. The thickened underflow is pumped to filters for further dewatering.

Thickeners often incorporate substantial storage capacity so that, for instance, if the filtration section is shut down for maintenance, the concentrator can continue to feed material to the dewatering section. During such periods the thickened underflow should be recirculated into the thickener feedwell. At no time should the underflow cease to be pumped, as chokage of the discharge cone rapidly occurs.

Since capital is the major cost of thickening, selection of the correct size of thickener for a particular application is important.

The two primary functions of the thickener are the production of a clarified overflow and a thickened underflow of the required concentration.

For a given throughput the clarifying capacity is determined by the thickener diameter, since the surface area must be large enough so that the upward velocity of liquid is at all times lower than the settling velocity of the slowest-settling particle which is to be recovered. The degree of thickening produced is controlled by the residence time of the particles and hence by the thickener depth.

The solids concentration in a thickener varies from that of the clear overflow to that of the thickened underflow being discharged. Although the variation in concentration is continuous, the concentrations at various depths may be grouped into four zones, as shown in Figure 15.9.

![Figure 15.9 Concentration zones in a thickener](From Chemical Engineers' Handbook by J.H. Perry., McGraw-Hill, 1963)
When materials settle with a definite interface between the suspension and the clear liquid, as is the case with most flocculated mineral pulps, the solids-handling capacity determines the surface area. Solids-handling capacity is defined as the capacity of a material of given dilution to reach a condition such that the mass rate of solids leaving a region is equal to or greater than the mass rate of solids entering the region. The attainment of this condition with a specific dilution depends on the mass subsidence rate being equal to or greater than the corresponding rise rate of displaced liquid. A properly sized thickener containing material of many different dilutions, ranging from the feed to the underflow solids contents, has adequate area such that the rise rate of displaced liquid at any region never exceeds the subsidence rate.

The satisfactory operation of the thickener as a clarifier depends upon the existence of a clear-liquid overflow at the top. If the clarification zone is too shallow, some of the smaller particles may escape in the overflow. The volumetric rate of flow upwards is equal to the difference between the rate of feed of liquid and the rate of removal in the underflow. Hence the required concentration of solids in the underflow, as well as the throughput, determines the conditions in the clarification zone.

The method developed by Coe and Clevenger (1916) is commonly employed to determine surface area when the material settles with a definite interface.

If \( F \) is the liquid-to-solids ratio by weight at any region within the thickener, \( D \) is the liquid-to-solids ratio of the thickener discharge, and \( W \) t h\(^{-1}\) of dry solids are fed to the thickener, then \( (F - D)W \) t h\(^{-1}\) of liquid moves upwards to the region from the discharge.

The velocity of this liquid current is thus

\[
\frac{(F - D)W}{A - S} \tag{15.1}
\]

where \( A \) is the thickener area (m\(^2\)) and \( S \) is the specific gravity of the liquid (kg l\(^{-1}\)).

Because this upward velocity must not exceed the settling rate of the solids in this region, at equilibrium

\[
\frac{(F - D)W}{A - S} = R \tag{15.2}
\]

where \( R \) is the settling rate (m h\(^{-1}\)).

The required thickener area is therefore

\[
A = \frac{(F - D)W}{RS} \tag{15.3}
\]

From a complete set of \( R \) and \( F \) values the area required for various dilutions may be found by recording the initial settling rate of materials with dilutions ranging from that of the feed to the discharge. The dilution corresponding to the maximum value of \( A \) represents the minimum solids-handling capacity and is the critical dilution.

In using this method the initial constant sedimentation rate is found through tests in graduated cylinders using dilutions ranging from the feed dilution to the underflow dilution, the rate of fall of the interface between the thickened pulp and clarified solution being timed.

Once the required surface area is established, it is necessary to apply a safety factor to the calculated area. This should be at least two.

The Coe and Clevenger method requires multiple batch tests at different arbitrary pulp densities before an acceptable unit area can be selected. The Kynch model (1952) offers a way of obtaining the required area from a single batch-settling curve, and is the basis of several thickening theories, which have been comprehensively reviewed by Pearse (1977).

The Talmage and Fitch method (1955) applies Kynch’s mathematical model to the problem of thickener design. The results of a batch-settling test are plotted linearly as mudline (interface between settled pulp and clear water) height against time (Figure 15.10).

![Figure 15.10 Batch-settling curve](image)
Talmage and Fitch showed that by constructing a tangent to the curve at any point, then if \( H \) is the intercept of the tangent on the ordinate,

\[
CH = C_0 H_0
\]

(15.4)

where \( C_0 \) kg l\(^{-1} \) is the original feed solids concentration, \( H_0 \) cm is the original mudline height, and \( H \) is the mudline height corresponding to a uniform slurry of concentration \( C \) kg l\(^{-1} \), at the point where the tangent was taken. Therefore, for any selected point on the settling curve, the local concentration can be obtained from Equation 15.4, and the settling rate from the gradient of the tangent at that point. Thus a set of data of concentration against settling rate can be obtained from the single batch-settling curve.

For a pulp of solids concentration \( C \) kg l\(^{-1} \), the volume occupied by the solids in 1 litre of pulp is \( C/d \), where \( d \) kg l\(^{-1} \) is the specific gravity of dry solids.

Therefore the weight of water in 1 litre of pulp

\[
= 1 - \frac{C}{d} = \frac{d - C}{d}
\]

Therefore the water–solids ratio by weight

\[
= \frac{d - C}{dC}
\]

For pulps of concentrations \( C \) kg l\(^{-1} \) of solids, and \( C_u \) kg l\(^{-1} \) of solids, the difference in water–solids ratio

\[
= \frac{d - C}{dC} - \frac{d - C_u}{dC_u}
\]

\[
= \frac{1}{C} - \frac{1}{C_u}
\]

Therefore the values of concentration obtained, \( C \), and the settling rates, \( R \), can be substituted in the Coe and Clevenger Equation 15.3, i.e.

\[
A = \left( \frac{1}{C} - \frac{1}{C_u} \right) \frac{W}{RS}
\]

(15.5)

where \( C_u \) is the underflow solids concentration.

A simplified version of the Talmage and Fitch method is offered by determining the point on the settling curve where the solids go into compression. This point corresponds to the limiting settling conditions and controls the area of thickener required. In Figure 15.11, \( C \) is the compression point, and a tangent is drawn to the curve at this point, intersecting the ordinate at \( H \). A line is drawn parallel to the abscissa, cutting the ordinate of a tangent from a point \( C_u \) on the curve, where \( C_u \) is the solids concentration of the thickener underflow. The tangent from \( C \) intersects this line at a time corresponding to \( t_u \). \( H_u \) can be calculated from Equation 15.4.

The required thickener area (Equation 15.5).

\[
= \frac{W \left( 1/C - 1/C_u \right)}{(H - H_u)/t_u}
\]

where \( (H - H_u)/t_u \) is the gradient of the tangent at point \( C \), i.e. the settling rate of the particles at the compression point concentration. Since \( CH = C_0 H_0 \),

\[
A = \frac{W \left[ (H/C_0 H_0) - (H_u/C_0 H_0) \right]}{(H - H_u)/t_u}
\]

\[
= W \frac{t_u}{C_0 H_0}
\]

(15.6)

In most cases, the compression point concentration will be less than that of the underflow concentration. In cases where this is not so, then the tangent construction is not necessary, and \( t_u \) is the point where the underflow line crosses the settling curve. In many cases, the point of compression on the curve is clear, but when this is not so, a variety of methods have been suggested for its determination (Fitch, 1977; Pearse, 1978).

The Coe and Clevenger and modified Talmage and Fitch methods are the most widely used
in the metallurgical industry to predict thickener area requirements. Both methods have limitations (Waters and Galvin, 1991), the Talmage and Fitch technique relying critically on identifying a compression point, and both must be used in conjunction with empirical safety factors. Generally, the Coe and Clevenger method tends to underestimate the thickener area requirement, whilst the Talmage and Fitch method tends to overestimate. It is usually better to overestimate in design to allow for feed fluctuations and increase in production, and because of this, and its relative experimental simplicity, the Talmage and Fitch method is often preferred, providing that a compression point is readily identifiable.

Recent work has described software for the prediction of continuous thickener area based on a phenomenological model of particle settling. The model is similar in form to the equation of Coe and Clevenger (Garridon et al., 2003). The development of thickener models over the last 100 years is reviewed by Concha and Burger (2003).

The mechanism of thickening has been far less well expressed in mathematical terms than the corresponding clarifying mechanisms. The depth of the thickener is therefore usually determined by experience. The diameter is usually large compared with the depth, and therefore a large ground area is required. Tray thickeners (Figure 15.12) are sometimes installed to save space. In essence, a tray thickener is a series of unit thickeners mounted vertically above one another. They operate as separate units, but a common central shaft is utilised to drive the sets of rakes.

**High-capacity thickeners**

Conventional thickeners suffer from the disadvantage that large floor areas are required, since the throughput depends above all on the area, while depth is of minor importance.

In recent years, machines known as “high-capacity” or “high rate” thickeners have been introduced by various manufacturers. Many varieties exist, and the machines are typified by a reduction in unit area requirement from conventional installations (Keane, 1982; Green, 1995).

The “Enviro-Clear” thickener developed by Envirotech Corporation is typical (Emmett and Klepper, 1980) (Figure 15.13).

The feed enters via a hollow drive shaft where flocculant is added and is rapidly dispersed by staged mechanical mixing. This staged mixing action is said to improve thickening since it makes most effective use of the flocculant. The flocculated feed leaves the mixing chambers and is injected into a blanket of slurry where the feed solids are further flocculated by contacting previously flocculated material. Direct contact between rising fluid and settling solids, which is common to most thickeners, is averted with slurry blanket injection. Radially mounted inclined plates are partially submerged in the slurry blanket; the settling solids in the slurry blanket slide downwards along the inclined plates, producing faster and more effective thickening than vertical descent. The height of the slurry blanket is automated through the use of a level sensor.

High density thickeners (or high compression thickeners) are an extension of high capacity thickening utilising a deeper mud bed to increase capacity and underflow density. High rate rakeless thickeners use a deep tank and a steep bottom cone to maximise underflow density while eliminating the rake and rake drive. In some applications underflow with the consistency of paste can be produced from high density and rakeless thickeners. However for consistent paste underflow several manufacturers offer deep cone thickeners in applications where surface tailings

![Figure 15.12 Tray thickener](image-url)
Figure 15.13  Enviro-Clear high-capacity thickener. 1 – mixer drive; 2 – feed pipe; 3 – overflow launder; 4 – inclined settling plates; 5 – rake arm; 6 – level sensor; 7 – flocculant feed pipe; 8 – drive unit with overload control; 9 – sludge discharge; 10 – mixing chamber

disposal by wet stacking or underground paste backfill is required. The tank height to diameter ratio is often 1:1 or greater (Schoenbrunn and Laros, 2002).

Centrifugal sedimentation

Centrifugal separation can be regarded as an extension of gravity separation, as the settling rates of particles are increased under the influence of centrifugal force. It can, however, be used to separate emulsions which are normally stable in a gravity field.

Centrifugal separation can be performed either by hydrocyclones or centrifuges. The simplicity and cheapness of the hydrocyclone (Chapter 9) make it very attractive, although it suffers from restrictions with respect to the solids concentration which can be achieved and the relative proportions of overflow and underflow into which the feed may be split. Generally the efficiency of even a small-diameter cyclone falls off rapidly at very fine particle sizes and particles smaller than about 10 μm in diameter will invariably appear in the overflow, unless they are very heavy. Flocculation of such particles is not possible, since the high shear forces within a cyclone rapidly break up any agglomerates. The cyclone is therefore inherently better suited to classification rather than thickening.

By comparison, centrifuges are much more costly and complex, but have a much greater clarifying power and are generally more flexible. Much greater solids concentrations can be obtained than with the cyclone.

Various types of centrifuge are used industrially (Bragg, 1983; Bershad et al., 1990; Leung, 2002), the solid bowl scroll centrifuge having widest use in the minerals industry due to its ability to discharge the solids continuously.

The basic principles of a typical machine are shown in Figure 15.14. It consists essentially of a horizontal revolving shell or bowl, cylinbroconical in form, inside which a screw conveyor of similar section rotates in the same direction at a slightly higher or lower speed. The feed pulp is admitted to the bowl through the centre tube of
the revolving-screw conveyor. On leaving the feed pipe the slurry is immediately subjected to a high centrifugal force causing the solids to settle on the inner surface of the bowl at a rate which depends on the rotational speed employed, this normally being between 1600 and 8500 rev min$^{-1}$. The separated solids are conveyed by the scroll out of the liquid and discharged through outlets at the smaller end of the bowl. The solids are continuously dewatered by centrifugal force as they proceed from the liquid zone to the discharge. Excess entrained liquor drains away to the pond circumferentially through the particle bed.

When the liquid reaches a predetermined level it overflows through the discharge ports at the larger end of the bowl.

The actual size and geometry of these centrifuges vary according to the throughput required and the application. The length of the cylindrical section largely determines the clarifying power and is thus made a maximum where overflow clarity is of prime importance. The length of the conical section, or "beach", decides the residual moisture content of the solids, so that a long shallow cone is used where maximum dryness is required.

Centrifuges are manufactured with bowl diameters ranging from 15 to 150 cm, the length generally being about twice the diameter. Throughputs vary from about 0.5 to 50 m$^3$ h$^{-1}$ of liquid and from about 0.25 to 100 t h$^{-1}$ of solids depending on the feed concentration, which may vary widely from 0.5 to 70% solids, and on the particle size, which may range from about 12 mm to as fine as 2 µm, or even less when flocculation is used. The wide application of flocculation is limited by the tendency of the scroll action to damage the flocs and thus redisperse the fine particles. The moisture content in the product varies widely, typically being in the range 5–20%.

**Filtration**

Filtration is the process of separating solids from liquid by means of a porous medium which retains the solid but allows the liquid to pass. The theory of filtration has been comprehensively reviewed mathematically elsewhere (Coulson and Richardson, 1968; Cain, 1990) and will not be covered here.

The conditions under which filtration are carried out are many and varied and the choice of the most suitable type of equipment will depend on a large number of factors. Whatever type of equipment is used, a filter cake gradually builds up on the medium and the resistance to flow progressively increases throughout the operation. Factors affecting the rate of filtration include:

(a) The pressure drop from the feed to the far side of the filter medium. This is achieved in pressure filters by applying a positive pressure at the feed end and in vacuum filters by applying a vacuum to the far side of the medium, the feed side being at atmospheric pressure.
(b) The area of the filtering surface.
(c) The viscosity of the filtrate.
(d) The resistance of the filter cake.
(e) The resistance of the filter medium and initial layers of cake.

Filtration in mineral processing applications normally follows thickening. The thickened pulp may be fed to storage agitators from where it is drawn off at uniform rate to the filters. Flocculants are sometimes added to the agitators in order to aid filtration. Slimes have an adverse effect on filtration, as they tend to "blind" the filter medium; flocculation reduces this and increases the voidage between particles, making filtrate flow easier. The lower molecular weight flocculants tend to be used in filtration, as the flocs formed by high molecular weight products are relatively large, and entrain water within the structure, increasing the moisture content of the cake, even with the lower molecular weight flocculants, which have a higher shear-resistance, and the resultant filter cake is a uniform porous structure which allows rapid dewatering, yet prevents migration of the finer particles through the cake to the medium (Moss, 1978). Other filter aids are used to reduce the liquid surface tension, thus assisting flow through the medium.

The filter medium

The choice of the filter medium is often the most important consideration in assuring efficient operation of a filter. Its function is generally to act as a support for the filter cake, while the initial layers of cake provide the true filter. The filter medium should be selected primarily for its ability to retain solids without blinding. It should be mechanically strong, corrosion resistant, and offer as little resistance to flow of filtrate as possible. Relatively coarse materials are normally used and clear filtrate is not obtained until the initial layers of cake are formed, the initial cloudy filtrate being recycled.

Filter media are manufactured from cotton, wool, linen, jute, nylon, silk, glass fibre, porous carbon, metals, rayon and other synthetics, and miscellaneous materials such as porous rubber. Cotton fabrics are by far the most common type of medium, primarily because of their low initial cost and availability in a wide variety of weaves. They can be used to filter solids as fine as 10 μm.

Filtration tests

It is not normally possible to forecast what may be accomplished in the filtration of an untested product, therefore preliminary tests have to be made on representative samples of the pulp before the large-scale plant is designed. Tests are also commonly carried out on pulps from existing plants, to assess the effect of changing operating conditions, filter aids, etc. A simple vacuum filter leaf test circuit is shown in Figure 15.15.

![Figure 15.15 Laboratory test filter](image)

The filter leaf, consisting of a section of the industrial filter medium, is connected to a filtrate receiver equipped with a vacuum gauge. The receiver is connected to a vacuum pump. If the industrial filter is to be a continuous vacuum filter, this operation must be simulated in the test. The cycle is divided into three sections – cake formation (or "pick-up"), drying, and discharge. Sometimes pick-up is followed by a period of washing and the cake may also be subjected to compression during drying. While under vacuum, the test leaf is submerged for the pick-up period in the agitated pulp to be tested. The leaf is then removed and held with the drainpipe down for the allotted drying time. The cake can then be removed, weighed, and dried. The daily filter capacity can then be determined by the dry weight of cake per unit area of test leaf multiplied by the daily number of cycles and the filter area.

Bench scale testing of samples for specification of filtration equipment is described by Smith and Townsend (2002).
Types of filter

Cake filters are the type most frequently used in mineral processing, where the recovery of large amounts of solids from fairly concentrated slurries is the main requirement. Those where the main requirement is the removal of small amounts of solid from relatively dilute suspensions are known as screening or clarification filters.

Cake filters may be pressure, vacuum, batch, or continuous types. The various types are reviewed by Cox and Traczyk (2002).

Pressure filters

Because of the virtual incompressibility of solids, filtration under pressure has certain advantages over vacuum. Higher flow rates and better washing and drying may result from the higher pressures that can be used. However, the continuous removal of solids from the pressure-filter chamber can be extremely difficult and consequently, although continuous pressure filters do exist, the vast majority operate as batch units.

Filter presses are the most frequently used type of pressure filter. They are made in two forms—the plate and frame press and the recessed plate or chamber press.

The plate and frame press (Figure 15.16) consists of plates and frames arranged alternately. The hollow frame is separated from the plate by the filter cloth. The filter press is closed by means of a screw or hydraulic piston device and compression of the filter cloth between plates and frames helps to prevent leakages. A tight chamber is therefore formed between each pair of plates. The slurry is introduced to the empty frames of the press through a continuous channel formed by the holes in the corners of the plates and frames. The filtrate passes through the cloth and runs down the grooved surfaces of the plates and is removed through a continuous channel. The cake remains in the frame and, when the frame is full, the filter cake can be washed, after which the pressure is released and the plates and frames separated one by one. The filter cake in the frames can then be discharged, the filter press closed again and the cycle repeated.

The chamber press (Figure 15.17) is similar to the plate and frame type except for the fact that the filter elements consist solely of the recessed filter plates. The individual filter chambers are therefore formed between successive plates. All the chambers are connected by means of a comparatively large hole in the centre of each plate. The filter cloth with a central hole covers the plate and slurry is led through the inlet channel. The clear filtrate passing through the cloth is removed by means of smaller holes in the plate, the cake gradually depositing in the chambers.

Automatic pressure filters are now widely used in most new flotation plants. Automatic means a
filter in which plate pack opening, pump and ancillary equipment starting, and valve operation as well as cake discharge are all automatically controlled (Townsend, 2003). Modern pressure filters can process up to 150 t/h dry solids per unit of copper concentrate in filters with filtration areas of up to 144 m². Even higher throughputs can be achieved in iron ore applications. Residual cake moisture depends on the material being filtered but values in the range 7.5–12.5% are typical.

Vacuum filters
There are many different types of vacuum filter, but they all incorporate filter media suitably supported on a drainage system, beneath which the pressure is reduced by connection to a vacuum system. Vacuum filters may be batch or continuous (Keleghan, 1986a,b).

Batch vacuum filters The leaf filter has a number of leaves, each consisting of a metal framework or a grooved plate over which the filter cloth is fixed (Figure 15.18).

Numerous holes are drilled in the pipe framework, so that when a vacuum is applied, a filter cake builds up on both sides of the leaf. A number of leaves are generally connected and are first immersed in slurry held in a filter feed tank and then to a cake-receiving vessel where the cake is removed by replacing the vacuum by air pressure (Figure 15.19).

Although simple to operate, these filters require considerable floor space and suffer from the possibility of sections of cake dropping from the leaves during transport from tank to tank. They are now used only for clarification, i.e. the removal of small amounts of suspended solids from liquors.

Horizontal leaf, or tray filters, work in much the same manner as a laboratory Buchner filter and consist of rectangular pans having a false bottom of filter medium. They are filled with pulp, the vacuum is applied until the cake is dry, when the pan is inverted, being supported on pivots, the vacuum is disconnected and low-pressure air is introduced under the filter medium to remove the cake.

Continuous vacuum filters These are the most widely used filters in mineral processing applications and fall into three classes – drums, discs, and horizontal filters.

The rotary-drum filter (Figure 15.20) is the most widely used type in industry, finding application both where cake washing is required and where it is unnecessary.

The drum is mounted horizontally and is partially submerged in the filter trough, into which the feed slurry is fed and maintained in suspension by agitators. The periphery of the drum is divided into compartments, each of which is provided with a number of drain lines, which pass through the inside of the drum, terminating at one end as a ring of
ports which are covered by a rotary valve to which vacuum is applied. The filter medium is wrapped tightly around the drum surface which is rotated at low speed, usually in the range 0.1–0.3 rev min\(^{-1}\), but up to 3 rev min\(^{-1}\) for very free-filtering materials.

As the drum rotates, each compartment goes through the same cycle of operations, the duration of each being determined by the drum speed, the depth of submergence of the drum, and the arrangement of the valve. The normal cycle of operations consists of filtration, drying, and discharge, but it is possible to introduce other operations into the basic cycle, such as cake washing and cloth cleaning.

Various methods are used for discharging the solids from the drum, depending on the material being filtered. The most common form makes use of a reversed blast of air, which lifts the cake so that it can be removed by a knife, without the latter actually contacting the medium. Another method is string discharge, where the filter cake is formed on an open conveyor – the strings – which are in contact with the filter cloth in the filtration, washing, and drying zones. A further advance on this method is belt discharge, as shown in Figure 15.20, where the filter medium itself leaves the filter and passes over the external roller, before returning to the drum. This has a number of advantages in that very much thinner cakes can be handled, with consequently increased filtration and draining rates and hence better washing and dryer products. At the same time, the cloth can be washed on both sides by means of sprays before it returns to the drum (Figure 15.21), thus minimising the extent of blinding. Cake washing is usually carried out by means of sprays or weirs, which cover a fairly limited area at the top of the drum.

The capacity of the vacuum pump will be determined mainly by the amount of air sucked through the cake during the washing and drying periods when, in most cases, there will be a simultaneous flow of both liquid and air. A typical layout is shown in Figure 15.22, from which it is seen that the air and liquid are removed separately.

The barometric leg should be at least 10 m high to prevent liquid being sucked into the vacuum pump.

Variations on standard drum filters to enable them to handle coarse, free-draining, quick-settling

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**Figure 15.18** Cross-section of typical leaf filter

**Figure 15.19** Typical leaf filter circuit (From *Chemical Engineers' Handbook* by J.H. Perry, McGraw-Hill, 1963)
Figure 15.20  Rotary-drum filter with belt discharge

Figure 15.21  Belt discharge filter with cloth washing

Figure 15.22  Typical rotary-drum filter system
materials include top feed units where the material is distributed at between 90 and 180° from the feed point. Hyperbaric filters have been developed to satisfy the need for pressure filtration (to give high filtration rates and dry cakes) and continuous operation. Some of these contain a conventional drum filter operating inside a large pressure vessel (Anlauf, 1991; Bott et al., 2003).

The principle of operation of disc filters (Figure 15.23), is similar to that of rotary drum filters. The solids cake is formed on both sides of the circular discs, which are connected to the horizontal shaft of the machine. The discs rotate and lift the cake above the level of the slurry in the trough, whereupon the cake is suction-dried and is then removed by a pulsating air blow with the assistance of a scraper. The discs can be located along the shaft at about 30 cm centres and consequently a large filtration area can be accommodated in a small floor space. Cost per unit area is thus lower than for drum filters, but cake washing is virtually impossible and the disc filter is not as adaptable as a drum filter.

The horizontal belt filter (Figure 15.24), consists of an endless perforated rubber drainage deck supporting a separate belt made from a suitable filter cloth. At the start of the horizontal travel, slurry flows by gravity on to the belt. Filtration immediately commences, due partly to gravity and partly to the vacuum applied to the suction boxes which are in contact with the underside of the drainage deck during the coarse of its travel.

The cake which forms is dewatered, dried by drawing air through it, and then discharged as the belt reverses over a small-diameter roller. If required, one or more washes can be incorporated.

The applications for horizontal belt filters are increasing. They are particularly suited to hydrometallurgical circuits where metal values are dissolved in alkali or acid. These values can be recovered from waste solids by filtration of the leached slurry and countercurrent washing (Bragg, 1983). Large belt filters are in operation on cyanide-leached gold ore and acid-leached uranium ore. Belt filters are also suited for concentrated slurries of fast settling products, where efficient washing is required. In addition to their low installed capital cost when compared with disc, drum, and press-type filters, relatively low operating costs mean that these filters offer a particularly cost-effective and reliable solution to filtration problems, especially with low-value material such as mine tailings. Work on coal slurries has shown that horizontal belt vacuum filtration should produce lower cake
moistures than those from rotary vacuum filtration and at a reduced cost per tonne (Vickers et al., 1985).

Drying

The drying of concentrates prior to shipping is the last operation performed in the mineral-processing plant. It reduces the cost of transport and is usually aimed at reducing the moisture content to about 5% by weight. Dust losses are often a problem if the moisture content is lower.

*Rotary thermal dryers* are often used. These consist of a relatively long cylindrical shell mounted on rollers and driven at a speed of up to 25 rev min$^{-1}$. The shell is at a slight slope, so that material moves from the feed to discharge end under gravity. Hot gases, or air, are fed in either at the feed end to give parallel flow or at the discharge to give counter-current flow.

The method of heating may be either direct, in which case the hot gases pass through the material in the dryer, or indirect, where the material is in an inner shell, heated externally by hot gases. The direct-fired is the dryer most commonly used in the minerals industry, the indirect-fired type being used when the material must not contact the hot combustion gases. Parallel flow dryers (Figure 15.25) are used in the majority of current operations because they are more fuel efficient and have greater capacity than counterflow types (Kram, 1980). Since heat is applied at the feed end,
build-up of wet feed is avoided, and in general these units are designed to dry material to not less than 1% moisture. Since counter-flow dryers apply heat at the discharge end, a completely dry product can be achieved, but its use with heat-sensitive materials is limited because the dried material comes into direct contact with the heating medium at its highest temperature.

Prokesch (2002) reviews the various types of drying equipment available and describes dryer selection based on the required duty.

An alternative to direct-fired drying of slurries is the tube press, which uses hydraulic pressure at 100 bars to squeeze water from the slurry that enters the annular space between the filter tube and an outer tube (Figure 15.26). The outer tube contains the filtration pressure that is applied hydraulically by a tubular membrane and squeezes the water from the slurry through perforations in the filter tube. This is a perforated steel tube covered with a fine wire mesh backing and a filter cloth, known as a candle. The filtrate which collects in the central well of the candle is discharged from the press cloth by compressed air. It is reported that the tube press can save up to 80% of the energy required by comparable capacity thermal dryers (Anon., 1987).

The product from the dryers is often stock-piled, before being loaded on to trucks or rail-cars as required for shipment. The containers may be closed, or the surface of the contents sprayed with a skin-forming solution, in order to eliminate dust losses (Kolthammer, 1978).

Figure 15.26 Operation of tube press

References


Introduction

The disposal of mill tailings is a major environmental problem, which is becoming more serious with the increasing exploration for metals and the working of lower-grade deposits. Apart from the visual effect on the landscape of tailings disposal, the major ecological effect is usually water pollution, arising from the discharge of water contaminated with solids, heavy metals, mill reagents, sulphur compounds, etc. (Chalkley et al., 1989). Waste must therefore be disposed of in both an environmentally acceptable and, if possible, economically viable manner (Sofrás and Boger, 2002). Disposal is governed by legislation and may involve long-term rehabilitation of the site.

The nature of tailings varies widely; they are usually transported and disposed of as a slurry of high water content, but they may be composed of very coarse dry material, such as the float fraction from dense medium plants. Due to the lower costs of mining from open pits, ore from such locations is often of very low grade, resulting in the production of large amounts of very fine tailings.

Methods of disposal of tailings

The methods used to dispose of tailings have developed due to environmental pressures, changing milling practice, and realisation of profitable applications. Early methods included discharge of tailings into rivers and streams, which is still practised at some mines, and the dumping of coarse dewatered tailings on to land. The many nineteenth-century tips seen in Cornwall and other parts of Britain are evidence of this method. Due to the damage caused by such methods, and the much finer grinding necessary on most modern ores, other techniques have been developed. The most satisfactory way of dealing with tailings is to make positive use of them, such as reprocessing in order to recover additional values (see Chapter 1), or to use them as a useful product in their own right, e.g. the use of coarse (20–30 mm) DMS float as railway ballast and aggregate.

It is common practice in underground mines, in which the method of working requires the filling of mined-out areas, to return the coarser fraction of the mill tailings underground. This method has been used since the beginning of the century in South Africa’s gold mines (Stradling, 1988). Backfilling worked-out stopes reduces the volume of tailings which must be impounded on the surface, but not all tailings are suited as back-fill material. It is invariably necessary to de-slime the tailings, the resultant slimes, which may account for up to 50% of the total weight, requiring surface disposal (Down and Stocks, 1977a). Some tailings swell or shrink after the fill has been placed, and some have the useful property of being self-cementing, which removes the necessity of adding cement to the backfill, which is common practice prior to placement underground. The use of back-fill can cause surface disposal problems, in that borrowed fill may have to be used to construct the tailings impoundment, as the coarse fraction of the tailings, which is often used for construction, has been removed.

Back-fill methods have not been applied to the large amounts of tailings produced by open-pit mining methods, as this would entail temporary storage during the life of the mine prior to disposal in the worked-out pit and the most widely used method is to contain the tailings within a purpose built dam. The impoundment must provide safe and
Economical storage for the required volume of tailings and permit the construction and operation of pollution control facilities.

For operations that are close to the sea, submarine tailings disposal is an alternative to conventional tailings disposal provided the governmental regulations permit disposal in such a manner. The basic submarine tailings disposal design comprises a tailings line to a de-aeration/mixing chamber, with a seawater intake line, and discharge to location and depth allowing gravity flow of a coherent density to the final sedimentation area. Such systems can place mine tailings at locations and depths constraining environmental impact to restricted areas of the seabed and deep water turbidity (Ellis et al., 1995). This form of tailings disposal attracts considerable attention from environmental groups as the final disposal of the tailings is not in a controlled impoundment but is released directly into the lower levels of the ocean and can therefore affect the deep sea ecosystem. The process is increasingly used in the Asia-Pacific region where on-land disposal options are problematic. In comparison to tailing retentions on land, the mining industry has argued that submarine tailings disposal in the Asia-Pacific region is safer for the local people and the environment as the land is unsuited to the construction of tailings dams due to the natural topography, regular seismic activity, and high rainfall (McKinnon, 2002). Due to the complexity of the decision-making process for the viability of submarine tailings disposal, tools such as an expert system have been developed to assist mining project planners explore the feasibility of this method of tailings disposal (Ganguli et al., 2002).

**Tailings DAMS**

The design, construction, and operation of tailings dams is a major consideration for most new mining developments, as well as for many existing operations (Klohn, 1981; Vick, 1981).

It is economically advantageous to site the impoundment close to the mine, but this imposes limits on site selection. The type of tailings embankment is generally determined by the local seismic activity, water clarification, tailings properties and stability, tailings distribution, foundation and hydrological conditions, and environmental conditions (Mohd. Azizli et al., 1995). The ground underlying the dam must be structurally sound and able to bear the weight of the impoundment. If such a site cannot be found close to the mine, it may be necessary to pump the tailings, at a high slurry density, to a suitable location.

Tailings dams may be built across river valleys, or as curved or multi-sided dam walls on valley sides, this latter design facilitating drainage. On flat, or gently sloping ground, lagoons are built with walls on all sides of the impoundment.

The disposal of tailings adds to the production costs, so it is essential to make disposal as cheap as possible. This requirement led initially to the development of the once commonly used upstream method of tailings-dam construction, so named because the centre line of the dam moves upstream into the pond.

In this method, a small starter dam is placed at the extreme downstream point (Figure 16.1) and the dam wall is progressively raised on the upstream side. The tailings are discharged by spigoting off the top of the starter dyke and, when the initial pond is nearly filled, the dyke is raised and the cycle

---

**Figure 16.1** Upstream tailings dam
repeated. Various methods are used to raise the dam; material may be taken from the dried surface of the previously deposited tailings and the cycle repeated, or more commonly the wall may be built from the coarse fraction of the tailings, separated out by cyclones, or spigots, the fines being directed into the pond (Figures 16.2 and 16.3).

The main advantages of the upstream construction are the low cost and the speed with which the dam can be raised by each successive dyke increment.

The method suffers from the disadvantage that the dam wall is built on the top of previously deposited unconsolidated slimes retained behind the wall. There is a limiting height to which this type of dam can be built before failure occurs and the tailings flow out and, because of this, the upstream method of construction is now less commonly used.

The downstream method has evolved as a result of efforts to devise methods for constructing larger and safer tailings dams. This method produces safer
dams both in terms of static and seismic loading (Mohd. Azizli et al., 1995). It is essentially the reverse of the upstream method, in that as the dam wall is raised, the centreline shifts downstream and the dam remains founded on coarse tailings (Figure 16.4). Most procedures involve the use of cyclones to produce sand for the dam construction.

Downstream dam building is the only method that permits design and construction of tailings dams to acceptable engineering standards. All tailings dams in seismic areas, and all major dams, regardless of their location, should be constructed using some form of the downstream method. The major disadvantage of the technique is the large amount of sand required to raise the dam wall. It may not be possible, especially in the early stages of operation, to produce sufficient sand volumes to maintain the crest of the tailings dam above the rising pond levels. In such cases, either a higher starter dam is required or the sand supply must be augmented with borrowed fill, such procedures increasing the cost of tailings disposal.

The *centre-line method* (Figure 16.5) is a variation of that used to construct the downstream dam and the crest remains in the same horizontal position as the dam wall is raised. It has the advantage of requiring smaller volumes of sand-fill to raise the crest to any given height. The dam can thus be raised more quickly and there is less trouble keeping it ahead of the tailings pond during the early stages of construction. Care, however, must be exercised in raising the upstream face of the dam to ensure that unstable slopes do not develop temporarily.

Very stable tailings dams can be constructed from open-pit over-burden, or waste rock, according to the local circumstances. An example is shown in Figure 16.6. Since the tailings are not required for the dam construction, they may be fed into the pool without separation of the sands from the slimes. In some cases the output of over-burden may not be sufficient to keep the dam crest above the tailings pond, and it may be necessary to combine waste rock and tailings sand-fills to produce a safe economical dam.

An interesting method of disposal has been used at the Ecstall (Kidd Creek) operation at Texasgulf Canada Ltd. (Amsden, 1974). The tailings disposal area consists of 3000 acres enclosed by a gravel dyke. Mill tailings are thickened and pumped to a central spigoting location inside the dam. The system is designed to build a mountain of tailings in the central area and thus keep the height of the perimeter dyke to a minimum.

Erosion of dams due to wind and rain can affect the stability and produce environmental problems. Many methods are used to combat this, such as vegetation of the dam banks (Hill and Nothard,
There is little doubt that tailings dams have a visual impact on the environment due to their regular geometric shape. Perhaps the most conspicuous is the downstream type, whose outer wall is continually being extended, and cannot be re-vegetated until closure. There are, however, few reasons why dam walls should not be landscaped at some stage in their life, and many dams have been designed to permit early visual integration with the environment (Down and Stocks, 1977b). An example is the impoundment at Flambeau, North Wisconsin, USA (Shilling and May, 1977), where a rock-fill dam wall 18 m high, 24 m wide at the crest, and 111 m wide at the base was designed to minimise both visual and pollution effects (Figure 16.7). The wall consists of a clay core, with the downstream side faced with non-pyrite rock and covered with top-soil, permitting re-vegetation and consequently reduced visual impact.

The most serious problem associated with the disposal of tailings is the release of polluted water, and this has been extensively investigated (Anon., 1980). The main effects of pollution are due to the effluent pH, which may cause ecological changes; dissolved heavy metals, such as copper, lead, zinc, etc., which can be lethal to fish-life if allowed to enter local water-courses; mill reagents, which are usually present in only very small quantities, but, nevertheless, may be harmful; and suspended solids, which should be minimal if the tailings have spent long residence times in the dam, thus allowing the solids to settle and produce a clear decant. The potential effect of submarine tailings on fish-life and their prey either from altered physical habitat or from possible exposure to contaminants such as heavy metals or milling reagents is of major concern (Johnson et al., 1998). In these cases the environment is exposed to all of the tailings, not just the clear decant.

Figure 16.6  Dam constructed from overburden

Figure 16.7  Flambeau impoundment
Figure 16.8 shows a generalised representation of water gain and loss at a tailings impoundment (Down and Stocks, 1977b). With the exception of precipitation and evaporation, the rates and volumes of the water can be controlled to a large extent. It is more satisfactory to attempt to prevent the contamination of natural waters rather than to purify them afterwards, and if surface run-off to the dam is substantial, then interception ditches should be installed. It is difficult to quantify the amount of water lost to groundwater, but this can be minimised by selecting a site with impervious foundations, or by sealing with an artificial layer of clay. Seepage through the dam wall is often minimised by an impervious slimes layer on the upstream face of the dam, but this is expensive, and many mines prefer to encourage free-drainage of the dam through pervious, chemically barren material. In the case of upstream dams, this can be a barren starter dyke, while with downstream and centre-line constructions, a free-draining gravel blanket can be used. A small seepage pond with impervious walls and floors situated below the main dam can collect this water, from where it can be pumped back into the tailings pond. If the dam wall is composed of metal-bearing rock, or sulphide tailings, the seepage is often highly contaminated due to its contact with the solid tailings, and may have to be treated separately.

The tailings are often treated with lime in order to neutralise acids and precipitate heavy metals as insoluble hydroxides before pumping to the dam. Such treated tailings may be thickened and the overflow, free of heavy metals, returned to the mill (Figure 16.9), thus reducing the water and pollutant input to the tailings dam.

Assuming good control of the above inputs and outputs of dam water, the most important factor in achieving pollution control is the method used to remove surplus water from the dam. Decant facilities are required on all dams, to allow excess free water to be removed. Inadequate decant design has caused many major dam failures. Many older dams used decant towers with discharge lines running
through the base of the dam to a downstream pump-house. Failures of such structures were common due to the high pressures exerted on the pipelines, leading to uncontrolled losses of fluids and tailings downstream. Floating, or movable, pump-houses situated close to the tailings pond are now in common use.

Recycling of water from the decant is becoming more important due to pressures from governments and environmentalists. As much water as possible must be reclaimed from the tailings pond for re-use in the mill and the volume of fresh make-up water used must be kept to a minimum. The difference between the total volume of water entering the tailings pond and the volume of water reclaimed plus evaporation losses must be stored with the tailings in the dam. If that difference exceeds the volume of the voids in the stored tailings, there becomes a surplus of free water that can build up to tremendous quantities over the life of a mine. A typical dam-reclaim system is shown in Figure 16.10.

The main disadvantage of water reclamation is the recirculation of pollutants to the mill, which can interfere with processes such as flotation. Water treatment may overcome this, at little or no extra cost, as similar treatment would be required for the effluent discharge in any case. A number of wastewater treatment techniques are available, such as physical adsorption methods using active carbon, coal or bentonite clay or mineral slimes, biological oxidation of organics, removal of ionic species by ion exchange resins, and relatively new techniques such as reverse osmosis and atmospheric freezing (Rao and Finch, 1989).

Advances in the disposal of tailings using semi-dry or dry techniques offer a number of advantages over the wet disposal techniques. Dry disposal techniques require that tailings be thickened or de-watered prior to disposal. The dried tailings can then be disposed by dry stacking, thickened tailings disposal or paste fill for back-filling underground mines. These are all schemes that improve water and reagent recovery and decrease tailings volumes and footprint, which greatly assists site rehabilitation (Sofrário and Boger, 2002). Although semi-dry or dry disposal of tailings has benefits these techniques are not as capital cost-effective as the more traditional wet disposal of tailings and require a detailed understanding of the rheology and transport of the dried tailings (Nguyen and Boger, 1998).

Complexes of metals with cyanide and ammonia are especially prone to stabilisation and solubilisation in caustic solution and may require special treatment other than straightforward neutralisation by lime. Although natural degradation occurs to some extent, this is of little value in many cases.

![Figure 16.10 Water-reclamation system](image)

Figure 16.10 Water-reclamation system
during the winter months, when the tailings ponds may be ice-covered, and several processes have been developed to treat cyanide-bearing effluent (Scott and Ingles, 1981). Alkaline chlorination, whereby cyanide is oxidised to cyanate, has perhaps received the greatest attention (Eccles, 1977), but cyanides can also be effectively destroyed by oxidation with ozone (Jeffries and Tczap, 1978) or hydrogen peroxide, by reactions with sulphur dioxide and air (Lewis, 1984), and by electrochemical treatment, ion-exchange, and volatilisation of hydrogen cyanide. In the latter method, which has been proved full-scale in the mining industry, the tailings are acidified to produce hydrogen cyanide. This is volatilised by intensive air-sparging, while simultaneously recovering the evolved gas in a lime solution for recycling. The aerated, acidified barren solution is then reneutralised to precipitate the metal ions.

The mineralogical nature of the tailings often provides natural pollution control. For instance, the presence of alkaline gangue minerals such as limestone can render metals less soluble and neutralise oxidation products. Such ores thus present less problems than sulphide ores associated with neutral-acid gangues, which oxidise to produce sulphuric acid, and apart from acidifying the water, consume dissolved oxygen as well (Down and Stocks, 1977c). Chemical treatment of such acid effluents is essential, neutralisation by lime usually being performed, which precipitates the heavy metals, and promotes flocculation as well as reducing acidity.

There is a continuing need for the development of new, more economical methods for the removal of heavy metals from dilute acid effluents, and much research is being carried out worldwide by environmental and minerals engineers. Apart from chemical techniques such as oxidation and reduction, ion-exchange and electrochemical treatment, biological methods are also being researched and developed. For instance, it has been established that various fresh water and marine microalgae species are able to abstract heavy metal ions from aqueous solutions, thus making it possible not only to solve some industrial environmental problems, but also to recover a currently wasted product (Golab and Smith, 1992).

It has been shown by Rao et al. (1992) that acid mine drainage has potential as a coagulant for municipal waste-water, although the resultant heavy metal contamination of the discharge precludes its general use without pretreatment. It is evident that there is much research potential in these areas and that the methods used by the minerals engineer are set to play an increasingly important role in reducing the environmental impact of modern industry. Particular attention is being given to the modification of mineral processing operations to mitigate environmental impact (Feasby et al., 1995), and work has been done on incorporating the management of acid mine drainage into the block model of the mine for production planning purposes (Bennett et al., 1997). The ultimate way of avoiding water-based environmental impact is to operate dry mineral processes and consideration is being given to such options, particularly in arid areas (Napier-Munn and Morrison, 2002).

References


### Appendix I

#### Metallic ore minerals

<table>
<thead>
<tr>
<th>Metal</th>
<th>Main applications</th>
<th>Ore minerals</th>
<th>Formula</th>
<th>% metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Aluminium</td>
<td>Where requirements are lightness, high electrical and thermal conductivity, corrosion resistance, ease of fabrication. Forms high tensile strength alloys.</td>
<td>Bauxite</td>
<td>A1O(OH)</td>
<td>-</td>
<td>3.2-3.5</td>
<td>Bauxite, which occurs massive, is a mixture of minerals such as diaspore, gibbsite, and boehmite with iron oxides and silica. Occurs as residual earth from weathering and leaching of rocks in tropical climates</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diaspore</td>
<td>A1(OH)</td>
<td></td>
<td>2.3-2.4</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Gibbsite</td>
<td>A1O(OH)</td>
<td></td>
<td>3.0-3.1</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Boehmite</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Antimony</td>
<td>Flame-resistant properties of oxide used in textiles, fibres, and other materials. Alloyned with lead to increase strength for accumulator plates, sheet, and pipe. Important alloying element for bearing and type metals</td>
<td>Stibnite</td>
<td>Sb₂S₃</td>
<td>71.8</td>
<td>4.5-4.6</td>
<td>Main ore mineral. Commonly in quartz grains and in limestone replacements. Associates with galena, pyrite, realgar, orpiment, and cinnabar</td>
</tr>
<tr>
<td>Arsenic</td>
<td>Limited use in industry. Small amounts alloyed with copper and lead to toughen the metals. In oxide form, used as insecticide</td>
<td>Arsenopyrite</td>
<td>FeAsS</td>
<td>46.0</td>
<td>5.9-6.2</td>
<td>Widely distributed in mineral veins, with tin ores, tungsten, gold, and silver, sphalerite and pyrite. Since production of metal is in excess of demand, it is commonly regarded as gangue</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Realgar</td>
<td>AsS</td>
<td>70.1</td>
<td>3.5</td>
<td>Often associate in mineral veins in minor amounts</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Orpiment</td>
<td>As₂S₃</td>
<td>61</td>
<td>3.4-3.5</td>
<td></td>
</tr>
<tr>
<td>Beryllium</td>
<td>Up to 4% Be alloyed with copper to produce high tensile alloys with high fatigue, wear, and corrosion resistance, which are used to make springs, bearings, and valves, and spark-proof tools. Used for neutron absorption in nuclear industry. Used in electronics for speakers and styluses</td>
<td>Beryl</td>
<td>Be₃Al₂Si₆O₁₈</td>
<td>5</td>
<td>2.6-2.8</td>
<td>Only source of the metal. Often mined as gemstone – emerald, aquamarine. Commonly occurs as accessory mineral in coarse-grained granites (pegmatites) and other similar rocks. Also in calcite veins and mica schists. As similar density to gangue minerals; difficult to separate other than by hand-sorting</td>
</tr>
</tbody>
</table>

(continued)
<table>
<thead>
<tr>
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<th>Main applications</th>
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<th>Formula</th>
<th>%metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>BISMUTH</td>
<td>Pharmaceuticals; low melting point alloys for automatic safety devices, such as fire-sprinklers. Improves casting properties when alloyed with tin and lead</td>
<td>Native</td>
<td>Bi</td>
<td>100</td>
<td>9.7-9.8</td>
<td>Minor amounts in veins associated with silver, lead, zinc, and tin ores.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Bismuthinite</td>
<td>Bi₂S₃</td>
<td>81.2</td>
<td>6.8</td>
<td>Occurs in association with magnetite, pyrite, chalcopyrite, galena, and sphalerite, and with tin and tungsten ores. Majority of bismuth produced as by-product from smelting and refining of lead and copper</td>
</tr>
<tr>
<td>CADMIUM</td>
<td>Rust-proofing of steel, copper, and brass by electroplating and spraying; production of pigments; negative plate in alkali accumulators; plastic stabilisers</td>
<td>Greenockite</td>
<td>CdS</td>
<td>77.7</td>
<td>4.9-5.0</td>
<td>Found in association with lead and zinc ores, and in very small quantities with many other minerals. Due to volatility of the metal, mainly produced during smelting and refining of zinc, as a by-product</td>
</tr>
<tr>
<td>CAESIUM</td>
<td>Low ionisation potential utilised in photoelectric cells, photomultiplier tubes, spectro-photometers, infra-red detectors. Minor pharmaceutical use</td>
<td>Pollucite</td>
<td>Cs₄Al₅Si₉O₂₆·H₂O</td>
<td>10.0</td>
<td>2.9</td>
<td>Occurs in pegmatites of complex mineralogical character. Rare mineral.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Lepidolite</td>
<td>K(Li, Al)₃(Si, Al)₄O₁₀(OH, F)₂</td>
<td>–</td>
<td>2.8-2.9</td>
<td>Occurs in pegmatites, often in association with tourmaline and spodumene. Often carries traces of rubidium and caesium</td>
</tr>
<tr>
<td>CHROMIUM</td>
<td>Used mainly as alloying element in steels to give resistance to wear, corrosion, heat, and to increase hardness and toughness. Used for electroplating iron and steel. Chromite used as refractory with neutral characteristics. Used in production of bichromates and other salts in tanning, dyeing, and pigments</td>
<td>Chromite</td>
<td>FeCr₂O₄</td>
<td>46.2</td>
<td>4.1-5.1</td>
<td>Occurs in olivine and serpentine rocks, often concentrated sufficiently into layers or lenses to be worked. Due to its durability, it is sometimes found in alluvial sands and gravels</td>
</tr>
<tr>
<td>COBALT</td>
<td>Used as alloying element for production of high-temperature steels and magnetic alloys. Used as catalyst in chemical industry. Cobalt powder used as cement in sintered carbide cutting tools</td>
<td>Smaltite</td>
<td>CoAs₂</td>
<td>28.2</td>
<td>5.7-6.8</td>
<td>Smaltite and cobaltite occur in veins, often together with arsenopyrite, silver, calcite, and nickel minerals</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Cobaltite</td>
<td>CoAsS</td>
<td>35.5</td>
<td>6.0-6.3</td>
<td>Carrolite and linnaeite sometimes occur in small amounts in copper ores. Cobalt is usually only a minor constituent in ores such as lead, copper, and nickel and extracted as by-product</td>
</tr>
</tbody>
</table>
**Copper**

Used where high electrical or thermal conductivity is important. High corrosion resistance and easy to fabricate. Used in variety of alloys - brasses, bronzes, aluminium bronzes, etc.

| **Chalcopyrite** | CuFeS₂ | 34.6 | 4.1-4.3 | Main ore mineral. Most often in veins with other sulphides, such as galena, sphalerite, pyrrhotite, pyrite, and also cassiterite. Common gangue minerals quartz, calcite, dolomite. Disseminated with bornite and pyrite in porphyry copper deposits |
| **Chalcocite** | Cu₂S | 79.8 | 5.5-5.8 | Often associated with cuprite and native copper |
| **Bornite** | Cu₂FeS₄ | 63.3 | 4.9-5.4 | Associates with chalcopyrite and chalcocite in veins |
| **Covellite** | CuS | 66.5 | 4.6 | Sometimes as primary sulphide in veins, but more commonly as secondary sulphide with chalcopyrite, chalcocite, and bornite |
| **Cuprite** | Cu₂O | 88.8 | 5.9-6.2 | Found in oxidised zone of deposits, with malachite, azurite, and chalcocite |
| **Malachite** | Cu₃Co₃·Cu(OH)₂ | 57.5 | 4.0 | Frequently associated with azurite, native copper, and cuprite in oxidised zone |
| **Native** | Cu | 100 | 8.9 | Occurs in small amounts with other copper minerals |
| **Tennantite** | Cu₆As₂S₇ (variable) | 57.5 (variable) | 4.4-4.5 | Tennantite and tetrahedrite found in veins with silver, copper, lead, and zinc minerals. |
| **Tetrahedrite** | 4Cu₂S·Sb₂S₃ | 52.1 | 4.4-5.1 | Tetrahedrite more widespread and common in lead-silver veins |
| **Azurite** | 2CuCO₃·Cu(OH)₂ | 55 | 3.8-3.9 | Occurs in oxidised zone. Not as widespread as malachite |
| **Enargite** | Cu₃As₄S₄ | 48.4 | 4.4 | Associates with chalcocite, bornite, covellite, pyrite, sphalerite, tetrahedrite, baryte, and quartz in near-surface deposits |

(continued)
<table>
<thead>
<tr>
<th>Metal</th>
<th>Main applications</th>
<th>Ore minerals</th>
<th>Formula</th>
<th>% metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gallium</td>
<td>Electronics industry for production of light-emitting diodes. Used in electronic memories for computers</td>
<td>Occurs in some zinc ores, but no important ore minerals</td>
<td>–</td>
<td>–</td>
<td>–</td>
<td>About 90% of production is a direct by-product of alumina output. Also found in coal ash and flue dusts</td>
</tr>
<tr>
<td>Germanium</td>
<td>Electronics industry</td>
<td>Argyrodite</td>
<td>3Ag₂S·GeS₂</td>
<td>8.3</td>
<td>6.1</td>
<td>Occurs with sphalerite, siderite, and marcasite. No important ore minerals. Chief source is cadmium fume from sintering zinc concentrates</td>
</tr>
<tr>
<td>Gold</td>
<td>Jewellery, monetary use, electronics, dentistry, decorative plating</td>
<td>Native</td>
<td>Au</td>
<td>85–100</td>
<td>12–20</td>
<td>Disseminated in quartz grains, often with pyrite, chalcopyrite, galena, stibnite, and arsenopyrite. Also found alluvially in stream or other sediments. South African “banket” is consolidated alluvial deposit</td>
</tr>
<tr>
<td>Hafnium</td>
<td>Naval nuclear reactors, flashbulbs, ceramics, refractory alloys, and enamels</td>
<td>No ore minerals</td>
<td>–</td>
<td>–</td>
<td>–</td>
<td>Produced as co-product of zirconium sponge</td>
</tr>
<tr>
<td>Indium</td>
<td>Electronics, component of low-melting-point alloys and solders, protective coating on silverware and jewellery</td>
<td>Occurs as trace element in many ores</td>
<td>–</td>
<td>–</td>
<td>–</td>
<td>Recovered from residues and flue dusts from some zinc smelters</td>
</tr>
<tr>
<td>Iron</td>
<td>Iron and steel industry</td>
<td>HEMATITE</td>
<td>Fe₂O₃</td>
<td>70</td>
<td>5–6</td>
<td>Most important iron ore. Occurs in igneous rocks and veins. Also as ooliths or cementing material in sedimentary rocks</td>
</tr>
<tr>
<td></td>
<td></td>
<td>MAGNETITE</td>
<td>Fe₃O₄</td>
<td>72.4</td>
<td>5.5–6.5</td>
<td>The only ferromagnetic mineral. Widely distributed in several environments, including igneous and metamorphic rocks; and beach-sand deposits</td>
</tr>
</tbody>
</table>
| Lead Batteries, corrosion resistant pipes and linings, alloys, pigments, radiation shielding | Galena PbS 86.6 7.4–7.6 | Very widely distributed, and most important lead ore. Occurs in veins, often with sphalerite, pyrite, chalcopyrite, tetrahedrite, and gangue minerals such as quartz, calcite, dolomite, baryte, and fluoride. Also in pegmatites, and as replacement bodies in limestone and dolomite rocks, with garnets, feldspar, diopside, rhodonite, and biotite. Often contains up to 0.5% Ag, and is important source of this metal
| Goethite Fe₂O₃ · H₂O 62.9 4.0–4.4 | Widespread occurrence, associated with hematite and limonite
| Limonite Hydrous ferric oxides Variable 3.6–4.0 | Natural rust, chief constituent being goethite. Often associates with hematite in weathered deposits
| Siderite FeCO₃ 48.3 3.7–3.9 | Occurs massive in sedimentary rocks and as gangue mineral in veins carrying pyrite, chalcopyrite, galena
| Pyrrhotite FeS 61.5 (variable) 4.6 | The only magnetic sulphide mineral. Occurs disseminated in igneous rocks, commonly with pyrite, chalcopyrite, and pentlandite
| Pyrite FeS₂ 46.7 4.9–5.2 | One of most widely distributed sulphide minerals. Used for production of sulphuric acid, but often regarded as gangue
| Cerussite PbCO₃ 77.5 6.5–6.6 | In oxidised zone of lead veins, associated with galena, anglesite, smithsonite, and sphalerite
| Anglesite PbSO₄ 68.3 6.1–6.4 | Occurs in oxidation zone of lead veins
| Jamesonite Pb₄FeSb₆S₁₄ 50.8 5.5–6.0 | Occurs in veins with galena, sphalerite, pyrite, stibnite
| Spodumene LiAlSi₂O₆ 3.7 3.0–3.2 | Occurs in pegmatites with lepidolite, tourmaline, and beryl
| Lithium Lightest metal. Lithium carbide used in production of aluminium. Used as base in multipurpose greases; used in manufacture of lithium batteries. Large application in ceramics industry. Very little use in metallic form | Amblygonite 2LiF · Al₂O₃ · P₂O₅ 4.7 3.0–3.1 | Rare mineral occurring in pegmatites with other lithium minerals
| Lepidolite LiF · KF · Al₄O₁ · 3SiO₂ 1.9 2.8–3.3 | Mica occurring in pegmatites with other lithium minerals
| Tourmaline Complex borosilicate of Al, Na, Mg, Fe, Li, Mn 3.0–3.2 | Not a commercial source of metal. Some crystals used as gems. Occurs in granite pegmatites, schists, and gneisses
<p>| (continued) |</p>
<table>
<thead>
<tr>
<th>Metal</th>
<th>Main applications</th>
<th>Ore minerals</th>
<th>Formula</th>
<th>% metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>MAGNESIUM</td>
<td>Small amounts used in aluminium alloys to increase strength and corrosion resistance. Used to desulphur blast-furnace iron. Added to cast-iron to produce nodular iron. Used in cathodic protection, as a reagent in petrol processing and as reducing agent in titanium, and zirconium production. Structural uses where lightness required – magnesium die castings</td>
<td>Dolomite</td>
<td>MgCa(CO₃)₂</td>
<td>13</td>
<td>2.8–2.9</td>
<td>Most magnesium extracted from brine, rather than ore minerals</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Magnesite</td>
<td>MgCO₃</td>
<td>29</td>
<td>3.0–3.2</td>
<td>Mineral used in manufacture of refractories. Occurs as gangue mineral in veins with galena and sphalerite. Also occurs widely as rock-forming mineral.</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Carnallite</td>
<td>KMgCl₃·6H₂O</td>
<td>9</td>
<td>1.6</td>
<td>Used mainly for cement and refractory bricks. Often associates with serpentinite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Brucite</td>
<td>Mg(OH)₂</td>
<td>42</td>
<td>2.4</td>
<td>Occurs with halite and sylvine</td>
</tr>
<tr>
<td>MANGANESE</td>
<td>Very important ferro-alloy. About 95% of output used in steel and foundry industry. Balance mainly in manufacture of dry cells and chemicals</td>
<td>Pyrolusite</td>
<td>MnO₂</td>
<td>63.2</td>
<td>4.5–5.0</td>
<td>Often found in oxidised zone of ore deposits containing manganese. Also in quartz veins and manganese nodules</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Manganite</td>
<td>Mn₂O₃</td>
<td>62.5</td>
<td>4.2–4.4</td>
<td>Occurs in association with baryte, pyrolusite, and goethite and in veins in granite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Braunite</td>
<td>3Mn₂O₃·MnSiO₃</td>
<td>78.3</td>
<td>4.7–4.8</td>
<td>Occurs in veins with other manganese minerals</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Psilomelane</td>
<td>Mixture of Mn oxides</td>
<td>–</td>
<td>3.3–4.7</td>
<td>Found with pyrolusite oxides and limonite in sediments or quartz veins</td>
</tr>
<tr>
<td>MERCURY</td>
<td>Electrical apparatus, scientific instruments, manufacture of paint, electrolytic cells, solvent for gold, manufacture of drugs and chemicals</td>
<td>Cinnabar</td>
<td>HgS</td>
<td>86.2</td>
<td>8.0–8.2</td>
<td>Only important mercury mineral. Occurs in fractures in sedimentary rocks with pyrite, stibnite, and realgar. Common gangue minerals are quartz, calcite, baryte, and chalcedony</td>
</tr>
<tr>
<td>MOLYBDENUM</td>
<td>Main use as ferro-alloy. Metal also used in manufacture of electrodes and furnace parts. Also used as catalyst corrosion inhibitor, additive to lubricants</td>
<td>Molybdenite</td>
<td>MoS₂</td>
<td>60</td>
<td>4.7–4.8</td>
<td>Widely distributed in small quantities. Occurs in granites and pegmatites with wolfram and cassiterite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Wulfenite</td>
<td>PbMoO₄</td>
<td>26.2</td>
<td>6.5–7.0</td>
<td>Found in oxidised zone of lead and molybdenum ores. Commonly with anglesite, cerrusite, and vanadinite</td>
</tr>
</tbody>
</table>
### Nickel

**Very important ferro-alloy due to its high corrosion resistance (stainless steels).** Also alloyed with many non-ferrous metals – chromium, aluminium, manganese. Used for electroplating steels, as base for chromium plate. Pure metal corrosion resistant, and resists alkali attack. Is non-toxic and used for food handling and pharmaceutical equipment.

<table>
<thead>
<tr>
<th>Mineral</th>
<th>Formula</th>
<th>Density</th>
<th>Melting Point</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pentlandite</td>
<td>(FeNi)S</td>
<td>22.0</td>
<td>4.6–5.0</td>
</tr>
<tr>
<td>Garnierite</td>
<td>Hydrated Ni-Mg</td>
<td>25–30</td>
<td>2.4</td>
</tr>
<tr>
<td></td>
<td>silicate</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Niccolite</td>
<td>NiAs</td>
<td>44.1</td>
<td>7.3–7.7</td>
</tr>
<tr>
<td>Millerite</td>
<td>NiS</td>
<td>64.8</td>
<td>5.3–5.7</td>
</tr>
</tbody>
</table>

### Niobium (Columbium)

**Important ferro-alloy.** Added to austenitic stainless steels to inhibit intergranular corrosion at high temperatures.

<table>
<thead>
<tr>
<th>Mineral</th>
<th>Formula</th>
<th>Density</th>
<th>Melting Point</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pyrochlore</td>
<td>(Ca, Na)₂(Nb, Ta)₂O₆(O, OH, F)</td>
<td>–</td>
<td>4.2–6.4</td>
</tr>
<tr>
<td>Columbite</td>
<td>(Fe, Mn)(Nb, Ta)₂O₆</td>
<td>–</td>
<td>5.0–8.0</td>
</tr>
</tbody>
</table>

### Platinum Group

Platinum and palladium have wide use in jewellery and dentistry. Platinum, due to its high melting point and corrosion resistance, is widely used for electrical contact material and in manufacture of chemical.

<table>
<thead>
<tr>
<th>Mineral</th>
<th>Symbol</th>
<th>Density</th>
<th>Melting Point</th>
</tr>
</thead>
<tbody>
<tr>
<td>Native</td>
<td>Pt</td>
<td>45–86</td>
<td>21.5 (pure)</td>
</tr>
</tbody>
</table>

Platinum group metals occur together in nature as native metals or alloys. Platinum alloyed with other platinum group metals, iron, and copper. Occurs disseminated in igneous rocks, associates with chromite and copper ores. Found in lode and alluvial deposits.
<table>
<thead>
<tr>
<th>Metal</th>
<th>Main applications</th>
<th>Ore minerals</th>
<th>Formula</th>
<th>% metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>Radium</td>
<td>Industrial radiography, treatment of cancer, and production of luminous paint</td>
<td>See Uranium Minerals</td>
<td>–</td>
<td>–</td>
<td>–</td>
<td>Constituent of uranium minerals</td>
</tr>
<tr>
<td>Rare Earths</td>
<td>The cerium subgroup is the most important industrially. Rare earths used as catalysts in petroleum refining, iron–cerium alloys used as cigarette-lighter flints. Used in ceramics and glass industry and in production of colour televisions</td>
<td>Monazite</td>
<td>Rare earth and thorium phosphate</td>
<td>–</td>
<td>4.9–5.2</td>
<td>See Thorium</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Bastnaesite</td>
<td>(Ce, La)(CO₃)F</td>
<td>–</td>
<td>4.9–5.2</td>
<td>Often found in pegmatites, veins, and carbonatite plutons</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Xenotime</td>
<td>YPO₄</td>
<td>48.4 (Y)</td>
<td>4.4–5.1</td>
<td>Source of yttrium. Wide occurrence as accessory mineral, often in pegmatites, and alluvial deposits, associated with monazite, zircon, rutile, ilmenite, and feldspars</td>
</tr>
<tr>
<td></td>
<td>Osmium, the heaviest metal known, with a melting point of 2200°C, and ruthenium have little commercial importance</td>
<td>Osmiridium</td>
<td>Alloy of Os–Ir</td>
<td>–</td>
<td>19.3–21.1</td>
<td>Found in small amounts in some gold and platinum ores, where it is recovered as by-product</td>
</tr>
<tr>
<td></td>
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<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Element</td>
<td>Uses</td>
<td>Formula</td>
<td>Density</td>
<td>Strength</td>
<td>Remarks</td>
<td></td>
</tr>
<tr>
<td>-----------</td>
<td>----------------------------------------------------------------------</td>
<td>---------</td>
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<td>-------------------------------------------------------------------------------------------------</td>
<td></td>
</tr>
<tr>
<td><strong>Rhenium</strong></td>
<td>Used as catalyst in production of low-lead petrol. Used as catalyst with platinum. Used extensively in thermocouples, temperature controls, and heating elements. Also used as filaments in electronic apparatus.</td>
<td>Molybdenite MoS₂</td>
<td>–</td>
<td>4.7-4.8</td>
<td>Rhenium occurs associated with molybdenite in porphyry copper deposits, and is recovered as by-product.</td>
<td></td>
</tr>
<tr>
<td><strong>Rubidium</strong></td>
<td>Rubidium and caesium largely interchangeable in properties and uses, although latter usually preferred to meet present small industrial demand.</td>
<td>See Caesium</td>
<td></td>
<td></td>
<td>Rubidium widely dispersed as minor constituent in major caesium minerals.</td>
<td></td>
</tr>
<tr>
<td><strong>Silicon</strong></td>
<td>Used in steel industry and as heavy medium alloy as ferro-silicon. Also used to de-oxidise steels. Metal used as semi-conductor.</td>
<td>Quartz SiO₂</td>
<td>46.9</td>
<td>2.65</td>
<td>Commonest mineral, forming 12% of earth’s crust. Essential constituent of many rocks, such as granite and sandstone, and virtually sole constituent of quartzite rock.</td>
<td></td>
</tr>
<tr>
<td><strong>Selenium</strong></td>
<td>Used in manufacture of fade-resistant pigments, photo-electric apparatus, in glass production, and various chemical applications. Alloyed with copper and steel to improve machineability.</td>
<td>Naumanite Ag₅Se</td>
<td>26.8</td>
<td>8.0</td>
<td>Selenides occur associated with sulphides, and bulk of selenium recovered as by-product from copper sulphide ores.</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Claushalite PbSe 27.6 8.0</td>
<td>Claushalite PbSe</td>
<td>27.6</td>
<td>8.0</td>
<td>Closely associated with lead, zinc, and copper ores, and bulk of silver produced as by-product from smelting such ores.</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Eurcairite (AgCu)₂Se 18.7 7.5</td>
<td>Eurcairite (AgCu)₂Se</td>
<td>18.7</td>
<td>7.5</td>
<td>Usually alloyed with copper, gold, etc., and occurs in upper part of silver sulphide deposits.</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Berzelianite Cu₃Se 38.3 6.7</td>
<td>Berzelianite Cu₃Se</td>
<td>38.3</td>
<td>6.7</td>
<td>Occurs in upper parts of silver veins together with native silver and cerussite.</td>
<td></td>
</tr>
<tr>
<td><strong>Silver</strong></td>
<td>Sterling ware, jewellery, coinage, photographic and electronic products, mirrors, electroplate, and batteries.</td>
<td>Argentite Ag₂S</td>
<td>87.1</td>
<td>7.2-7.4</td>
<td>As well as ore minerals, certain tin slags are becoming important source of tantalum.</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Native Ag 100 (max.) 10.1-11.1</td>
<td>Native Ag</td>
<td>100 (max.)</td>
<td>10.1-11.1</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Cerargyrite AgCl 75.3 5.8</td>
<td>Cerargyrite AgCl</td>
<td>75.3</td>
<td>5.8</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Tantalum</strong></td>
<td>Used in certain chemical and electrical processes due to extremely high corrosion resistance. Used in production of special steels used for medical instruments. Used for electrodes, and tantalum carbide used for cutting tools. Used in manufacture of capacitors.</td>
<td>Pyrochlorite See Niobium</td>
<td>See Niobium</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Metal</td>
<td>Main applications</td>
<td>Ore minerals</td>
<td>Formula</td>
<td>% metal</td>
<td>Sp. gr.</td>
<td>Occurrence/associations</td>
</tr>
<tr>
<td>---------</td>
<td>-----------------------------------------------------------------------------------</td>
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<td>----------------------------------------------------------------------------------------</td>
</tr>
<tr>
<td><strong>TELLURIUM</strong></td>
<td>Used in production of free machining steels, in copper alloys, rubber production, and as catalyst in synthetic fibre production</td>
<td>Sylvanite</td>
<td>(\text{See GOLD})</td>
<td>–</td>
<td>9.3</td>
<td>Produced with selenium as by-product of copper refining. These metal tellurides, which are important gold ores, and other tellurides of bismuth and lead, are most important sources of tellurium.</td>
</tr>
<tr>
<td><strong>THALLIUM</strong></td>
<td>Very poisonous, and finds limited outlet as fungicide and rat poison. Thallium salts used in Clerici solution, an important heavy liquid</td>
<td>Occurs in some zinc ores, but no ore minerals</td>
<td>–</td>
<td>–</td>
<td>By-product of zinc refining</td>
<td></td>
</tr>
<tr>
<td><strong>THORIUM</strong></td>
<td>Radioactive metal. Used in electrical apparatus, and in magnesium–thorium and other thorium alloys. Oxide of importance in manufacture of gas-mantles, and used in medicine</td>
<td><strong>MONAZITE</strong></td>
<td>((\text{Ce, La, Th})\text{PO}_4)</td>
<td>4.9–5.4</td>
<td>9.3</td>
<td>Although occurring in lode deposits in igneous rocks such as granites, the main granites, deposits are alluvial, beach-sand deposits being most prolific source. Occurs associated with ilmenite, rutile, zircon, garnets, etc. Occurs in some beach-sand deposits</td>
</tr>
<tr>
<td><strong>TIN</strong></td>
<td>Main use in manufacture of tin-plate, for production of cans, etc. Important alloy in production of solders, bearing-metals, bronze, type-metal, pewter, etc.</td>
<td><strong>CASSITERITE</strong></td>
<td>(\text{SnO}_2)</td>
<td>78.6</td>
<td>6.8–7.1</td>
<td>Found in lode and alluvial deposits. Lode deposits in association with wolfram, arsenopyrite, copper, and iron minerals. Alluvially, often associated with ilmenite, monazite, zircon, etc.</td>
</tr>
<tr>
<td><strong>TITANIUM</strong></td>
<td>Due to its high strength and corrosion resistance, about 80% of titanium produced were used in aircraft and aerospace industries. Also used in power-station heat-exchanger tubing and in chemical and desalination plants</td>
<td><strong>ILMENITE</strong></td>
<td>(\text{FeTiO}_3)</td>
<td>31.6</td>
<td>4.5–5.0</td>
<td>Accessory mineral in igneous rocks especially gabbros and norites. Economically concentrated into alluvial sands, together with rutile, monazite zircon.</td>
</tr>
<tr>
<td></td>
<td></td>
<td><strong>RUTILE</strong></td>
<td>(\text{TiO}_2)</td>
<td>60</td>
<td>4.2</td>
<td>Accessory mineral in igneous rocks, but economic deposits found in alluvial beach-sand deposits</td>
</tr>
</tbody>
</table>
### Tungsten
Production of tungsten carbide for cutting, drilling, and wear-resistant applications. Used in lamp filaments, electronic parts, electrical contacts, etc. Important ferro-alloy, producing tool and high-speed steels

<table>
<thead>
<tr>
<th>Wolfram</th>
<th>(Fe, Mn)WO₄</th>
<th>50</th>
<th>7.1–7.9</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Scheelite</th>
<th>CaWO₄</th>
<th>63.9</th>
<th>5.9–6.1</th>
</tr>
</thead>
</table>

### Uranium
Nuclear fuel

<table>
<thead>
<tr>
<th>Pitchblende (Uraninite)</th>
<th>UO₂ (variable – partly oxidised to U₃O₈)</th>
<th>80–90</th>
<th>8–10</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Carnotite</th>
<th>K₂(UO₂)₂(VO₄)₂·3H₂O (approx.)</th>
<th>Variable</th>
<th>4–5</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Autunite</th>
<th>Ca(UO₂)₂(Po₄)₂·10–12H₂O</th>
<th>49</th>
<th>3.1</th>
</tr>
</thead>
</table>

### Uranium

<table>
<thead>
<tr>
<th>Torbernite</th>
<th>Cu(UO₂)₂(Po₄)₂·12H₂O</th>
<th>48</th>
<th>3.5</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Patronite</th>
<th>VS₄ (approx.)</th>
<th>28.5</th>
<th>–</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Carnotite</th>
<th>See Uranium</th>
<th>Variable</th>
<th>4–5</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Roscoelite (Vanadium mica)</th>
<th>H₈K(MgFe)(AlV)₄(SiO₅)₁₂</th>
<th>Variable</th>
<th>2.9</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th>Vanadinite</th>
<th>(PbCl)Pb₂(Po₄)₃</th>
<th>Variable</th>
<th>6.6–7.1</th>
</tr>
</thead>
</table>

**Uranium**
Most important uranium and radium ore. Occurs in veins with tin, copper, lead, and arsenic sulphides, and radium

Secondary mineral found in sedimentary rocks, also in pitchblende deposits. Source of radium

Occur together in oxidised zones as secondary products from other uranium minerals

**Vanadium**
Important ferro-alloy. Vanadium used in manufacture of special steels, such as high-speed tool steels. Increases strength of structural steels – used for oil and gas pipelines. Vanadium – aluminium master alloys used in preparation of some titanium-based alloys. Vanadium compounds used in chemical and oil industries as catalysts. Also used as glass-colouring agent and in ceramics

Ocurs with nickel and molybdenum sulphides and asphaltic material

Frequently with carnitite

Occurs in oxidation zone of lead, lead–zinc deposits. Also with other vanadium minerals in sediments

(continued)
<table>
<thead>
<tr>
<th>Metal</th>
<th>Main applications</th>
<th>Ore minerals</th>
<th>Formula</th>
<th>% metal</th>
<th>Sp. gr.</th>
<th>Occurrence/associations</th>
</tr>
</thead>
<tbody>
<tr>
<td>ZINC</td>
<td>Corrosion protective coatings on iron and steel (&quot;galvanising&quot;). Important alloying metal in brasses and zinc die-castings. Used to manufacture corrosion-resistant paints, pigments, fillers, etc.</td>
<td>SPHALERITE</td>
<td>ZnS</td>
<td>67.1</td>
<td>3.9-4.1</td>
<td>Most common zinc ore mineral, frequently associated with galena, and copper sulphides in vein deposits. Also occurs in limestone replacements, with pyrite, pyrrhotite, and magnetite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Smithsonite</td>
<td>ZnCO₃</td>
<td>52</td>
<td>4.3-4.5</td>
<td>Mainly occurs in oxidised zone of ore deposits carrying zinc minerals. Commonly associated with sphalerite, galena, and calcite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Hemimorphite</td>
<td>Zn₄Si₂O₇(OH)₂·H₂O</td>
<td>54.3</td>
<td>3.4-3.5</td>
<td>Found associated with smithsonite accompanying the sulphides of zinc, iron, and lead</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Marmatite</td>
<td>(Zn, Fe)S</td>
<td>46.5-56.9</td>
<td>3.9-4.2</td>
<td>Found in close association with galena</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Franklinite</td>
<td>Oxide of Fe, Zn, Mn</td>
<td>Variable</td>
<td>5.0-5.2</td>
<td>Franklinite, zincite, and willemite occur together in a contact metamorphic deposit at Franklin, New Jersey, in a crystallite limestone, where the deposit is worked for zinc and manganese</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Zincite</td>
<td>ZnO</td>
<td>80.3</td>
<td>5.4-5.7</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>Willemite</td>
<td>Zn₂SiO₄</td>
<td>58.5</td>
<td>4.0-4.1</td>
<td></td>
</tr>
<tr>
<td>ZIRCONIUM</td>
<td>Used, alloyed with iron, silicon, and tungsten, in nuclear reactors, and for removing oxides and nitrides from steel. Used in corrosion-resistant equipment in chemical plants</td>
<td>ZIRCON</td>
<td>ZrSiO₄</td>
<td>49.8</td>
<td>4.6-4.7</td>
<td>Widely distributed in igneous rocks, such as granites. Common constituent of residues of various sedimentary rocks, and occurs in beach sands associated with ilmenite, rutile, and monazite</td>
</tr>
</tbody>
</table>
## Common non-metallic ores

<table>
<thead>
<tr>
<th>Material</th>
<th>Uses</th>
<th>Main ore minerals</th>
<th>Formula</th>
<th>Sp. gr.</th>
<th>Occurrence</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>ANHYDRITE</strong></td>
<td>Increasing importance as a fertiliser, and in manufacture of plasters, cements, sulphates, and sulphuric acid</td>
<td><strong>ANHYDRITE</strong></td>
<td>CaSO₄</td>
<td>2.95</td>
<td>Occurs with gypsum and halite as a saline residue. Occurs also in “cap rock” above salt domes, and as minor gangue mineral in hydrothermal metallic ore veins</td>
</tr>
<tr>
<td><strong>APATITE</strong></td>
<td>See PHOSPHATES</td>
<td></td>
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</tr>
<tr>
<td><strong>ASBESTOS</strong></td>
<td>Heat-resistant materials, such as fire-proof fabrics and brake-linings. Also asbestos cement products, sheets for roofing and cladding, fire-proof paints, etc.</td>
<td><strong>CHRYSOTILE</strong> (Serpentised asbestos)</td>
<td>Mg₃Si₂O₅(OH)₄</td>
<td>2.5–2.6</td>
<td>Fibrous serpentine occurring as small veins in massive serpentine</td>
</tr>
<tr>
<td></td>
<td></td>
<td><strong>CROCIDOLITE</strong></td>
<td>Na₂(Mg, Fe, Al)₅Si₈O₂₀(OH)₄</td>
<td>3.4</td>
<td>Fibrous riebeckite, or blue asbestos, occurring as veins in bedded ironstones</td>
</tr>
<tr>
<td></td>
<td></td>
<td><strong>AMOSITE</strong></td>
<td>(Mg, Fe)₇Si₈O₂₂(OH)₂</td>
<td>3.2</td>
<td>Fibrous anthophyllite, occurring as long fibres in certain metamorphic rocks</td>
</tr>
<tr>
<td></td>
<td></td>
<td><strong>ACTINOLITE</strong></td>
<td>Ca₂(Mg, Fe)₅Si₈O₂₀(OH)₂</td>
<td>3.0–3.4</td>
<td>True asbestos, occurring in schists and in some igneous rocks as alteration product of pyroxene</td>
</tr>
<tr>
<td><strong>BADDELEYITE</strong></td>
<td>Ceramics, abrasives, refractories, polishing powders, and manufacture of zirconium chemicals</td>
<td><strong>BADDELEYITE</strong></td>
<td>ZrO₂</td>
<td>5.4–6.0</td>
<td>Mainly found in gravels with zircon, tourmaline, corundum, ilmenite, and rare-earth minerals</td>
</tr>
</tbody>
</table>

(continued)
<table>
<thead>
<tr>
<th>Material</th>
<th>Uses</th>
<th>Main ore minerals</th>
<th>Formula</th>
<th>Sp. gr.</th>
<th>Occurrence</th>
</tr>
</thead>
<tbody>
<tr>
<td>BARYTES</td>
<td>Main use in oil- and gas-well drilling industry in finely ground state as weighting agent in drilling muds. Also in manufacture of barium chemicals, and as filler and extender in paint and rubber industries</td>
<td>BARYTE</td>
<td>BaSO₄</td>
<td>4.5</td>
<td>Most common barium mineral, occurring in vein deposits as gangue mineral with ores of lead, copper, zinc, together with fluorite, calcite, and quartz. Also as replacement deposit of limestone and in sedimentary deposits</td>
</tr>
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<tr>
<td>BORATES</td>
<td>Used in manufacture of insulating fibreglass, as fluxes for manufacture of glasses and enamels. Borax used in soap and glue industries, in cloth manufacture and tanning. Also used as preservatives, antiseptics, and in paint driers</td>
<td>BORAX</td>
<td>Na₂B₄O₇ ⋅ 10H₂O</td>
<td>1.7</td>
<td>An evaporate mineral, precipitated by the evaporation of water in saline lakes, together with halite, sulphates, carbonates, and other borates in arid regions</td>
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<td>KERNITE</td>
<td>Na₂B₄O₇ ⋅ 4H₂O</td>
<td>1.95</td>
<td>Very important source of borates. Occurrence as borax</td>
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<td></td>
<td>COLEMANITE</td>
<td>Ca₃B₆O₁₁ ⋅ 5H₂O</td>
<td>2.4</td>
<td>In association with borax, but principally as a lining to cavities in sedimentary rocks</td>
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<td></td>
<td>ULEXITE</td>
<td>NaCaB₆O₉ ⋅ 8H₂O</td>
<td>1.9</td>
<td>Occurs with borax in lake deposits. Also with gypsum and rock salt</td>
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<tr>
<td></td>
<td></td>
<td>SASSOLINE</td>
<td>H₃BO₃</td>
<td>1.48</td>
<td>Occurs with sulphur in volcanoes and in hot lakes and lagoons</td>
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<tr>
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<td></td>
<td>BORACITE</td>
<td>Mg₃B₁₂O₁₃Cl</td>
<td>2.95</td>
<td>Occurs in saline deposits with rock-salt, gypsum, and anhydrite</td>
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<tr>
<td>CALCITE</td>
<td>Many uses according to purity and character. Clayey variety used for cement, purer variety for lime. Marble for building and ornamental stones. Used as smelting flux, and in printing processes. Chalk and lime applied to soil as dressing. Transparent calcite (Iceland spar), used in construction of optical apparatus</td>
<td>CALCITE</td>
<td>CaCO₃</td>
<td>2.7</td>
<td>Calcite is a common and widely distributed mineral, often occurring in veins, either as main constituent, or as gangue mineral with metallic ores. It is a rock-forming mineral, which is mainly quarried as the sedimentary rocks limestone and chalk, and metamorphic rock marble</td>
</tr>
<tr>
<td><strong>CHINA CLAY</strong></td>
<td>Manufacture of porcelain and china. Used as filler in manufacture of paper, rubber, and paint</td>
<td></td>
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<tr>
<td><strong>CHROMITE</strong></td>
<td>Used as refractory in steel-making furnaces</td>
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<tr>
<td><strong>CORUNDUM</strong></td>
<td>Abrasive. Next to diamond, is hardest known mineral. Coloured variety used as gemstones</td>
<td></td>
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<tr>
<td><strong>CRYOLITE</strong></td>
<td>Used as flux in manufacture of aluminium by electrolysis</td>
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<tr>
<td><strong>DIAMOND</strong></td>
<td>Gemstone. Used extensively in industry for abrasive and cutting purposes - hardest known mineral. Used for tipping drills in mining and oil industry</td>
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</tr>
<tr>
<td><strong>DOLOMITE</strong></td>
<td>Important building material. Also used for furnace linings and as flux in steel-making</td>
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<tr>
<td><strong>EMERY</strong></td>
<td>See <strong>CORUNDUM</strong></td>
<td></td>
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</tr>
<tr>
<td><strong>EPSOM SALTS</strong></td>
<td>Medicine and tanning</td>
<td></td>
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</tr>
</tbody>
</table>

| **Kaolinite** | \( \text{Al}_2\text{Si}_2\text{O}_5(\text{OH})_4 \) | 2.6 | A secondary mineral produced by the alteration of aluminous silicates, and particularly of alkali feldspars |
| **Chromite** | See **Chromium Minerals** (Appendix 1) |
| **Corundum (Emery)** | \( \text{Al}_2\text{O}_3 \) | 3.9-4.1 | Occurs in several ways. Original constituent of various igneous rocks, such as syenite. Also in metamorphic rocks such as marble, gneiss, and schist. Occurs also in pegmatites and in alluvial deposits. Impure form is emery, containing much magnetic and hematite |
| **Cryolite** | \( \text{Na}_3\text{AlF}_6 \) | 3.0 | Occurs in pegmatite veins in granite with siderite, quartz, galena, sphalerite, chalcopyrite, fluorite cassiterite, and other minerals. Only real deposit in Greenland |
| **Diamond (Bort)** | C | 3.5 | Distributed sporadically in kimberlite pipes. Also in alluvial beach and river deposits. Bort is grey to black and opaque, and is used industrially |
| **Dolomite** | \( \text{CaMg(CO}_3\text{)}_2 \) | 2.8-2.9 | Rock-forming mineral. Occurs as gangue mineral in veins containing galena and sphalerite |
| **Epsomite** | \( \text{MgSO}_4 \cdot 7\text{H}_2\text{O} \) | 1.7 | Usually as encrusting masses on walls of caves or mine workings. Also in oxidised zone of pyrite deposits in arid regions |

*continued*
<table>
<thead>
<tr>
<th>Material</th>
<th>Uses</th>
<th>Main ore minerals</th>
<th>Formula</th>
<th>Sp. gr.</th>
<th>Occurrence</th>
</tr>
</thead>
<tbody>
<tr>
<td>FELDSPAR</td>
<td>Used in manufacture of porcelain, pottery, and glass. Used in production of glazes on earthenware, etc., and as mild abrasive</td>
<td>ORTHOCLASITE</td>
<td>KAISi$_3$O$_8$</td>
<td>2.6</td>
<td>Most abundant of all minerals, and most important rock-forming mineral. Widely distributed, mainly in igneous, but also in metamorphic and sedimentary rocks</td>
</tr>
<tr>
<td></td>
<td></td>
<td>(Isomorphous forms – Microline, Sanidine, and Adularia – the potassic feldspars)</td>
<td></td>
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</tr>
<tr>
<td></td>
<td></td>
<td>ALBITE</td>
<td>NaAlSi$_3$O$_8$</td>
<td>2.6</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>ANORTHITE</td>
<td>CaAl$_2$Si$_2$O$_8$</td>
<td>2.74</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>(Plagioclase feldspars form series having formulae ranging from NaAlSi$_3$O$_8$ to CaAl$_2$Si$_2$O$_8$, changing progressively from albite, through oligoclase, andesine, labradorite, and bytownite to anorthite)</td>
<td></td>
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</tr>
<tr>
<td>FLUORSPAR</td>
<td>Mainly as flux in steelmaking. Also for manufacture of specialised optical equipment, production of hydrofluoric acid, and fluorocarbons for aerosols. Colour-banded variety known as Blue-John used as semi-precious stone</td>
<td>FLUORITE</td>
<td>CaF$_2$</td>
<td>3.2</td>
<td>Widely distributed, hydrothermal veins and replacement deposits, either alone, or with galena, sphalerite, barytes, calcite, and other minerals</td>
</tr>
<tr>
<td>GARNET</td>
<td>Mainly as abrasive for sandblasting of aircraft components, and for wood polishing. Also certain varieties used as gemstones</td>
<td>PYROPE</td>
<td>Mg$_3$Al$_2$(SiO$_4$)$_3$</td>
<td>3.7</td>
<td>Widely distributed in metamorphic and some igneous rocks. Also commonly found as constituent of beach and river deposits</td>
</tr>
<tr>
<td></td>
<td></td>
<td>ALMANDINE</td>
<td>Fe$_3$Al$_2$(SiO$_4$)$_3$</td>
<td>4.0</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>GROSSULAR</td>
<td>Ca$_3$Al$_2$(SiO$_4$)$_3$</td>
<td>3.5</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>ANDRADITE</td>
<td>Ca$_3$Fe$_2$(SiO$_4$)$_3$</td>
<td>3.8</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>SPESSARTITE</td>
<td>Mn$_3$Al$_2$(SiO$_4$)$_3$</td>
<td>4.2</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>UVAROVITE</td>
<td>Ca$_3$Cr$_2$(SiO$_4$)$_3$</td>
<td>3.4</td>
<td></td>
</tr>
<tr>
<td>GRAPHITE (Plumbago)</td>
<td>Manufacture of foundry moulds, crucibles, and paint; used as lubricant and as electric furnace electrodes</td>
<td>GRAPHITE</td>
<td>C</td>
<td>2.1–2.3</td>
<td>Occurs as disseminated flakes in metamorphic rocks derived from rocks with appreciable carbon content. Also as veins in igneous rocks and pegmatites</td>
</tr>
<tr>
<td>Gypsum</td>
<td>Used in cement manufacture, as a fertiliser, and as filler in various materials such as paper, rubber, etc. Used to produce <em>plaster of Paris</em></td>
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</tr>
<tr>
<td>Ilmenite</td>
<td>About 90% of ilmenite produced is used for manufacture of titanium dioxide, a pigment used in pottery manufacture</td>
<td></td>
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</tr>
<tr>
<td>Magnesite</td>
<td>Used as refractory for steel furnace linings, and in production of carbon dioxide and magnesium salts</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mica</td>
<td>Used for insulating purposes in electrical apparatus. Ground mica used in production of roofing material, and in lubricants, wall-finishes artificial stone, etc. Powdered mica gives &quot;frost&quot; effect on Christmas cards and decorations</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Phosphates</td>
<td>Main use as fertilisers. Small amounts used in production of phosphorous chemicals</td>
<td></td>
<td></td>
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<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Phosphate Rock</th>
<th>Complex phosphates of Ca, Fe, Al</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gypsum</td>
<td>CaSO₄ · 2H₂O</td>
</tr>
<tr>
<td>Evaporate mineral, occurring with halite and anhydrite in bedded deposits</td>
<td></td>
</tr>
<tr>
<td>Ilmenite</td>
<td>See Titanium Minerals (Appendix 1)</td>
</tr>
<tr>
<td>Magnesite</td>
<td>See Magnesium Minerals (Appendix 1)</td>
</tr>
<tr>
<td>Muscovite</td>
<td>KAl₂(AlSi₃O₁₀)(OH, F)₂</td>
</tr>
<tr>
<td>2.8–2.9</td>
<td></td>
</tr>
<tr>
<td>Phlogopite</td>
<td>KMg₃(AlSi₃O₁₀)(OH, F)₂</td>
</tr>
<tr>
<td>2.8–2.85</td>
<td>Widely distributed in igneous rocks, such as granite and pegmatites. Also in metamorphic rocks – gneisses and schists. Also in sedimentary sandstones, clays, etc.</td>
</tr>
<tr>
<td>Biotite</td>
<td>K(Mg, Fe)₃(AlSi₃O₁₀)(OH, F)₂</td>
</tr>
<tr>
<td>2.7–3.3</td>
<td>Most commonly in metamorphosed limestones, also in igneous rocks rich in magnesia</td>
</tr>
<tr>
<td>Phosphates</td>
<td>Ca₅(PO₄)₃(F, Cl, OH)</td>
</tr>
<tr>
<td>3.1–3.3</td>
<td>Widely distributed in granite, syenite and diorite. Common constituent of schists and gneisses and of contact metamorphic rocks</td>
</tr>
</tbody>
</table>

Guano is an accumulation of excrement of sea-birds, found mainly on oceanic islands.
<table>
<thead>
<tr>
<th>Material</th>
<th>Uses</th>
<th>Main ore minerals</th>
<th>Formula</th>
<th>Sp. gr.</th>
<th>Occurrence</th>
</tr>
</thead>
<tbody>
<tr>
<td>POTASH</td>
<td>Used as fertilisers, and source of potassium salts. Nitre also used in explosives manufacture (saltpetre)</td>
<td>SYLVINE</td>
<td>KCl</td>
<td>2.0</td>
<td>Occurs in bedded evaporate deposits with halite and carnallite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>CARNALLITE</td>
<td>KMgCl₃·6H₂O</td>
<td>1.6</td>
<td>In evaporate deposits with sylvine and halite</td>
</tr>
<tr>
<td></td>
<td></td>
<td>ALUNITE</td>
<td>KAl₃(SO₄)₂(OH)₆</td>
<td>2.6</td>
<td>Secondary mineral found in areas where volcanic rocks containing potassic feldspars have been altered by acid solutions</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Nitre</td>
<td>KNO₃</td>
<td>2.1</td>
<td>Occurs in soils in arid regions, associated with gypsum, halite, and nitratine</td>
</tr>
<tr>
<td>QUARTZ</td>
<td>Building materials, glass making, pottery, silica bricks, ferro-silicon, etc. Used as abrasive in scouring soaps, sandpaper, toothpaste, etc. Due to its piezo-electric properties, quartz crystals widely used in electronics</td>
<td>See SILICON MINERALS (Appendix 1)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>ROCK SALT</td>
<td>Culinary and preserving uses. Wide use in chemical manufacturing processes</td>
<td>HALITE</td>
<td>NaCl</td>
<td>2.2</td>
<td>Occurs in extensive stratified evaporate deposits, formed by evaporation of land-locked seas in geological past. Associates with other water soluble minerals, such as sylvine, gypsum, and anhydrite</td>
</tr>
<tr>
<td>RUTILE</td>
<td>Production of welding rod coatings, and titanium dioxide, a pigment used in pottery manufacture</td>
<td>See TITANIUM MINERALS (Appendix 1)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>SERPENTINE</td>
<td>Used as building stone and other ornamental work. Fibrous varieties source of asbestos (See ASBESTOS)</td>
<td>SERPENTINE</td>
<td>Mg₂Si₂O₅(OH)₄</td>
<td>2.5–2.6</td>
<td>Secondary mineral formed from minerals such as olivine and orthopyroxene. Occurs in igneous rocks containing these minerals, but typically in serpentines, formed by alteration of olivine-bearing rocks</td>
</tr>
<tr>
<td>Mineral</td>
<td>Description</td>
<td>Chemical Formula</td>
<td>Density</td>
<td>Notes</td>
<td></td>
</tr>
<tr>
<td>-----------</td>
<td>-------------------------------------------------------------------------------------------------------</td>
<td>------------------</td>
<td>---------</td>
<td>------------------------------------------------------------------------------------------------</td>
<td></td>
</tr>
<tr>
<td>Sillimanite</td>
<td>Raw material for high-alumina refractories, for iron and steel industry, and other metal smelters. Also used in glass industry, and as insulating porcelains for spark-plugs, etc.</td>
<td>Al₂SiO₅ (Disthene)</td>
<td>3.2-3.3</td>
<td>Typically in schists and gneisses produced by high-grade regional metamorphism.</td>
<td></td>
</tr>
<tr>
<td>Kyanite</td>
<td></td>
<td>Al₂SiO₅</td>
<td>3.5-3.7</td>
<td>Typically in regionally metamorphosed schists and gneisses, together with garnet, mica, and quartz. Also in pegmatites and quartz veins associated with schists and gneisses.</td>
<td></td>
</tr>
<tr>
<td>Andalusite</td>
<td></td>
<td>Al₂SiO₅</td>
<td>3.1-3.2</td>
<td>In metamorphosed rocks of clayey composition. Also as accessory mineral in some pegmatites, with corundum, tourmaline, and topaz.</td>
<td></td>
</tr>
<tr>
<td>Sillimanite</td>
<td></td>
<td>Al₂SiO₅</td>
<td>3.2-3.3</td>
<td>Similarly to kyanite.</td>
<td></td>
</tr>
<tr>
<td>Mullite</td>
<td></td>
<td>Al₆Si₂O₁₃</td>
<td>3.2</td>
<td>Rarely found in nature, but synthetic mullite produced in many countries.</td>
<td></td>
</tr>
<tr>
<td>Talc</td>
<td>As filler for paints, paper, rubber, etc. Used in plasters, lubricants, toilet powder, French chalk. Massive varieties used for sinks, laboratory tabletops, acid tanks, etc.</td>
<td>Mg₃Si₄O₁₀(OH)₂</td>
<td>2.6-2.8</td>
<td>Secondary mineral formed by alteration of olivine, pyroxene, and amphibole, and occurs along faults in magnesium rich rocks. Also occurs in schists, in association with actinolite. Massive talc known as soapstone or stealite.</td>
<td></td>
</tr>
<tr>
<td>Vermiculite</td>
<td>Outstanding thermal and sound insulating properties, light, fire-resistant, and inert – used principally in building industry</td>
<td>Mg₃(Al, Si)₄O₁₀(OH)₂ · 4H₂O</td>
<td>2.3-2.4</td>
<td>Occurs as an alteration product of magnesian micas, in association with carbonatites.</td>
<td></td>
</tr>
<tr>
<td>Witherite</td>
<td>Source of barium salts. Small quantities used in pottery industry</td>
<td>BaCO₃</td>
<td>4.3</td>
<td>Not of wide occurrence. Sometimes accompanies galena in hydrothermal veins, together with anglesite and baryte.</td>
<td></td>
</tr>
<tr>
<td>Zircon Sand</td>
<td>Used in foundries, refractories, ceramics, and abrasives, and in chemical production</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Excel Spreadsheets for formulae in chapter 3

These spreadsheets are accessible by going to the Minerals Engineering International website at http://www.min-eng.com and following the prompts. The following notes describe the functionality of each spreadsheet. The spreadsheet names are the same as those used for the equivalent basic computer programs in previous editions of the book.

**Gy: Sample size by Gy Formula**
The function GYMass() calculates the minimum practical sampling weights required at each stage of sampling. The mass given is that obtained by Gy’s formula multiplied by a safety factor of 2. For routine sampling, a confidence interval of 95% in the results would be acceptable; but for research purposes, or where greater sampling accuracy is required, 99% level of confidence would be required.

**Gy: Sample error by Gy Formula**
The function GYError() will calculate the maximum relative error for a sample mass from each stage of sampling, i.e. the fundamental errors incurred after a sample has been taken. The calculated relative error is that obtained by Gy’s formula.

**RecVar: Estimation of errors in recovery calculations**
RecVar calculates the error associated with the two-product recovery formula for the assay recovery.

**MassVar: Estimation of errors in two-product mass flow rate**
MassVar calculates the error associated with the two-product recovery formula for the mass recovery. Enter the Feed, Concentrate, and Tail assays and the relative error for the Feed, Concentrate, and Tail assays in the highlighted cells. The spreadsheet returns the calculated assay recovery, and the variance and standard deviation of the calculated assay recovery.

**Lagran: Reconciliation of excess data by non-weighted least squares**
Lagran uses a simple node adjustment by least squares followed by Lagrangian multipliers. Enter the assay names into column B. Enter the Feed, Concentrate, and Tail assay values for each assay. The spreadsheet returns the balanced feed, concentrate, and tail assays and the balanced assay and mass recoveries.
WeightRe: Reconciliation of excess data by weighted least squares
WeightRe estimates the best mass rate by using weighted residuals least squares followed by Lagrangian multipliers. Enter the assay names and the Feed, Concentrate, and Tail assay values for each assay. Enter the relative standard deviations associated with the Feed, Concentrate and Tail assay values for each assay. The spreadsheet returns the balanced assays for the Feed, Concentrate, and Tail and the balanced assay and mass recoveries.

Wilman: Reconciliation of excess data by variances in mass equations
Wilman estimates the best mass rate by using variances in the component equations. Data adjustment is by Lagrangian multipliers. Enter the assay names and the Feed, Concentrate, and Tail assay values for each assay. Enter the relative standard deviations associated with the Feed, Concentrate, and Tail assay values for each assay. The spreadsheet returns the balanced assays for the Feed, Concentrate, and Tail and the balanced assay and mass recoveries.
**Wilman : Reconciliation of excess data by variances in mass equations**

B.A. Wilks (1985)
Updated to MS Excel, JK Tech Pty Ltd (2005)

**Assay Data Ranges:**
- **Feed/Conc/Tail**: C16:E25

**SDs Data Ranges:**
- **Feed/Conc/Tail**: F16:H25

**Balance Data Ranges:**
- **Feed/Conc/Tail**: I16:L25

---

**Click to Calculate the Weighted Recovery Balance**

**Mass Recovery:** 42.2 %

<table>
<thead>
<tr>
<th></th>
<th>Measured Data</th>
<th>SDs Data (Relative %)</th>
<th>Balanced Data</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Feed</td>
<td>Conc</td>
<td>Tail</td>
</tr>
<tr>
<td>Ti</td>
<td>21.90</td>
<td>43.00</td>
<td>6.77</td>
</tr>
<tr>
<td>Fe</td>
<td>3.46</td>
<td>5.50</td>
<td>1.76</td>
</tr>
<tr>
<td>Si</td>
<td>58.00</td>
<td>25.10</td>
<td>75.30</td>
</tr>
<tr>
<td>S</td>
<td>0.11</td>
<td>0.13</td>
<td>0.69</td>
</tr>
<tr>
<td>As</td>
<td>0.56</td>
<td>0.38</td>
<td>0.34</td>
</tr>
<tr>
<td>TiO2</td>
<td>4.91</td>
<td>9.24</td>
<td>2.07</td>
</tr>
</tbody>
</table>

---

Wilman estimates the best mass rate by using variances in the component equations. Data adjustment is by Lagrangian multipliers.

Enter the assay names into column B.

Enter the Feed, Concentrate, and Tail assay values for each assay (columns C to E).

Enter the relative standard deviations associated with the Feed, Concentrate, and Tail assay values for each assay (columns F to H).

The balanced assays for the Feed, Concentrate, and Tail are reported in columns I to K.

The balanced assay recovery is reported in column L.

The balanced mass recovery is reported in cell K12.

The number of assays can be extended beyond 10 by adjusting the maximum data ranges in cells C6, C9, and C12.
Reconciliation of excess data by weighted least squares

(1885)

to MS Excel, JKTech Pty Ltd (2005)

Function: 0.001 5.1E-04 actual convergence
Iterations: 100 2 number of iterations required

Data Ranges: C19:E28

Click to Calculate the Weighted Recovery Balance

<table>
<thead>
<tr>
<th>Measured Data</th>
<th>SDs Data (Relative %)</th>
<th>Balanced Data</th>
<th>Mass Recovery: 41.3 %</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Feed</td>
<td>Conc</td>
<td>Tail</td>
</tr>
<tr>
<td></td>
<td>21.90</td>
<td>43.00</td>
<td>6.77</td>
</tr>
<tr>
<td>n</td>
<td>3.46</td>
<td>5.50</td>
<td>1.76</td>
</tr>
<tr>
<td>ca</td>
<td>58.00</td>
<td>25.10</td>
<td>75.30</td>
</tr>
<tr>
<td>phur</td>
<td>0.11</td>
<td>0.12</td>
<td>0.09</td>
</tr>
<tr>
<td>Generic</td>
<td>0.36</td>
<td>0.38</td>
<td>0.34</td>
</tr>
<tr>
<td>02</td>
<td>4.91</td>
<td>9.24</td>
<td>2.07</td>
</tr>
</tbody>
</table>

WeightRe estimates the best mass rate by using weighted residuals least squares followed by Lagrangian multipliers.

Enter the assay names into column B.

Enter the Feed, Concentrate, and Tail assay values for each assay (columns C to E).

Enter the relative standard deviations associated with the Feed, Concentrate, and Tail assay values for each assay (columns F to H).

The balanced assays for the Feed, Concentrate, and Tail are reported in columns I to K.

The balanced assay recovery is reported in column L.

The balanced mass recovery is reported in cell K15.

The number of assays can be extended beyond 16 by adjusting the maximum data ranges in columns C, D, E, and F.
Lagran: Reconciliation of excess data by non-weighted least squares
B.A. Willis (1984)
Updated to MS Excel, JKTech Pty Ltd (2005)

Click to Calculate the Non-Weighted Recovery Balance

<table>
<thead>
<tr>
<th>Measured Data</th>
<th>Balanced Data</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Mass Recovery: 37.0%</td>
</tr>
<tr>
<td>Feed</td>
<td>Conc</td>
</tr>
<tr>
<td>Tin</td>
<td>21.90</td>
</tr>
<tr>
<td>Iron</td>
<td>3.46</td>
</tr>
<tr>
<td>Silica</td>
<td>58.00</td>
</tr>
<tr>
<td>Sulphur</td>
<td>0.11</td>
</tr>
<tr>
<td>Arsenic</td>
<td>0.36</td>
</tr>
<tr>
<td>TiO2</td>
<td>4.91</td>
</tr>
</tbody>
</table>

Lagran uses a simple node adjustment by least squares followed by Lagrangian multipliers.

Enter the assay names into column B.

Enter the Feed, Concentrate, and Tail assay values for each assay (columns C to E).

The balanced assays for the Feed, Concentrate, and Tail are reported in columns F to H.

The balanced assay recovery is reported in column I.

The balanced mass recovery is reported in cell I9.

The number of assays can be extended beyond 10 by adjusting the maximum data ranges in cells C6 and C9.
RecVar : Estimation of errors in recovery calculations

B.A.Wills (1984)
Updated to MS Excel, JKTech Pty Ltd (2005)

<table>
<thead>
<tr>
<th>Assay</th>
<th>Feed</th>
<th>Conc</th>
<th>Tail</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>3.5%</td>
<td>18%</td>
<td>1%</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>SDs:</th>
<th>Relative %</th>
<th>Absolute</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed</td>
<td>4%</td>
<td>0.14</td>
</tr>
<tr>
<td>Conc</td>
<td>2%</td>
<td>0.36</td>
</tr>
<tr>
<td>Tail</td>
<td>8%</td>
<td>0.08</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Calculation</th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>D</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>3</td>
<td>26</td>
<td>0</td>
<td>236</td>
</tr>
</tbody>
</table>

Recovery: 75.6%
Variance: 5.7
SDs: 2.4%

RecVar calculates the error associated with the two-product recovery formula for the *assay recovery*

Enter the Feed, Concentrate, and Tail assays in the highlighted cells.

Enter the relative error for the Feed, Concentrate, and Tail assays in the highlighted cells.

Calculated assay recovery
Variance of the assay recovery
Standard deviation of the assay recovery
MassVar: Estimation of errors in two-product mass flowrate

B.A. Wills (1984)  
Updated to MS Excel, JKTech Pty Ltd (2005)

**Assay:**

<table>
<thead>
<tr>
<th></th>
<th>Feed</th>
<th>Conc</th>
<th>Tail</th>
</tr>
</thead>
<tbody>
<tr>
<td>%</td>
<td>0.92</td>
<td>0.99</td>
<td>0.69</td>
</tr>
</tbody>
</table>

**SDs:**

<table>
<thead>
<tr>
<th></th>
<th>Relative %</th>
<th>Absolute</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed</td>
<td>1 %</td>
<td>0.0092</td>
</tr>
<tr>
<td>Conc</td>
<td>1 %</td>
<td>0.0099</td>
</tr>
<tr>
<td>Tail</td>
<td>1.5 %</td>
<td>0.01035</td>
</tr>
</tbody>
</table>

**Calculation**

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>333</td>
</tr>
<tr>
<td>B</td>
<td>256</td>
</tr>
<tr>
<td>C</td>
<td>78</td>
</tr>
</tbody>
</table>

**Yield:** 76.7 %

**Variance:** 16.5

**SDs:** 4.1 %

MassVar calculates the error associated with the two-product recovery formula for the *mass recovery*

Enter the Feed, Concentrate, and Tail assays in the highlighted cells.

Enter the relative error for the Feed, Concentrate, and Tail assays in the highlighted cells.

Calculated mass recovery

Variance of the recovery

Standard deviation of the recovery
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Confidence Level</td>
<td>%</td>
<td>95%</td>
</tr>
<tr>
<td>Elemental Assay of Ore</td>
<td>%</td>
<td>6%</td>
</tr>
<tr>
<td>Elemental Assay of Mineral</td>
<td>%</td>
<td>86.6%</td>
</tr>
<tr>
<td>Liberation Size</td>
<td>cm</td>
<td>0.015</td>
</tr>
<tr>
<td>Alluvial Gold</td>
<td>y/n</td>
<td>N</td>
</tr>
<tr>
<td>Mineral Density</td>
<td>t/m³</td>
<td>7.5</td>
</tr>
<tr>
<td>Gangue Density</td>
<td>t/m³</td>
<td>2.05</td>
</tr>
<tr>
<td>Number of Sampling Stages</td>
<td></td>
<td>4</td>
</tr>
<tr>
<td>Top Size Of Ore</td>
<td>cm</td>
<td>2.5 0.5 0.1 0.004</td>
</tr>
<tr>
<td>F95 / F5 Ratio</td>
<td></td>
<td>5 5 5 5</td>
</tr>
<tr>
<td>Sample Mass</td>
<td>g</td>
<td>5000 500 100 10.0</td>
</tr>
<tr>
<td>GYError</td>
<td>%</td>
<td>23.4% 9.9% 3.0% 0.1%</td>
</tr>
</tbody>
</table>

The function GYError() will calculate the maximum relative error for a sample mass from each stage of sampling. The calculated relative error is that obtained by Gy's formula.
**Gy : Sample size by Gy Formula**

B.Wills (1982)

Updated to MS Excel, JKTech Pty Ltd (2005)

<table>
<thead>
<tr>
<th><strong>Confidence Level</strong></th>
<th>%</th>
<th>95%</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Elemental Assay of Ore</strong></td>
<td>%</td>
<td>5%</td>
</tr>
<tr>
<td><strong>Elemental Assay of Mineral</strong></td>
<td>%</td>
<td>86.6%</td>
</tr>
<tr>
<td><strong>Assay Relative Error</strong></td>
<td>%</td>
<td>2%</td>
</tr>
<tr>
<td><strong>Liberation Size</strong></td>
<td>cm</td>
<td>0.015</td>
</tr>
<tr>
<td><strong>Alluvial Gold</strong></td>
<td>y/n</td>
<td>N</td>
</tr>
<tr>
<td><strong>Mineral Density</strong></td>
<td>t/m³</td>
<td>7.5</td>
</tr>
<tr>
<td><strong>Gangue Density</strong></td>
<td>t/m³</td>
<td>2.65</td>
</tr>
<tr>
<td><strong>Number of Sampling Stages</strong></td>
<td></td>
<td>4</td>
</tr>
<tr>
<td><strong>Top Size Of Ore</strong></td>
<td>cm</td>
<td>2.5, 0.5, 0.1, 0.004</td>
</tr>
<tr>
<td><strong>F95 / F5 Ratio</strong></td>
<td></td>
<td>5, 5, 5, 5</td>
</tr>
</tbody>
</table>

The function GYMass() will calculate the minimum practical sampling weights required at each stage of sampling. The mass given is that obtained by Gy's formula multiplied by a safety factor of 2. For routine sampling, a confidence interval of 95% in the results would be acceptable, but for research purposes, or where greater sampling accuracy is required, 99% level of confidence would be required.

The function GYError() will calculate the maximum relative error for a sample mass from each stage of sampling. The calculated relative error is that obtained by Gy's formula.
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