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Advanced Gravity Concentration of Fine Particles: A Review

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ABSTRACT
The article makes an effort in consolidating the developments in the field of advanced gravity separation techniques for processing fine particles over the last few decades. The application potentials of various unit operations and the influence of process variables and feed characteristics on the process performance are identified. Fundamentals of advanced gravity separators are discussed from theoretical and applied perspectives. Comparative analyses of various related unit operations are incorporated including the advantages and short falls of different advanced gravity separation techniques. A general guideline is presented that will help designing new mineral processing flow sheets with better efficiency indices.

1. Introduction
Gravity concentration is one of the oldest industrial methods widely used for minerals and coal beneficiation. Ease of operation, low operating cost, and ecofriendly nature are major benefits associated with the use of gravity separation (Burt 1999). It was thought that with the advent of froth flotation, its relative importance would go down in twenty-first century, which is far from reality due mainly to the numerous advantages of gravity-based separation methods (Richards and Palmer 1997; Burt 1999).

Separating mineral particles on the basis of specific gravity is known as gravity concentration. It relies on differential settling of particles in a fluid medium. In mineral processing operations, the fluid is mostly water and sometimes air. The ease and efficiency of separation depend primarily on the concentration criterion which is defined as the ratio of the density differential between the heavy mineral and the fluid and the same between light mineral and fluid. The greater the quotient the easier it is to separate the two minerals by gravity concentration. The quotient, in essence, indicates the differential settling characteristics of the two minerals. When the resistance to motion is primarily due to the Newtonian drag (turbulent resistance) the application of concentration criterion is very convenient. This concept is effectively exploited in the conventional gravity separation unit operations such as jigs, spiral concentrators, Wilfley tables, and heavy medium cyclones (Wills 1997).

The problem with this simplistic description is that the settling of particles is also strongly dependent on particle size. When particles are too small the settling kinetics are very sluggish. Therefore, for such fine particles, even though the concentration criterion is favorable, the time required for effective separation is impractically long. Conventional gravity concentrations are not able to provide such long residence times and are, therefore, not employable for fine particles. On the other hand, the finer the particle size the greater is the mineral liberation. Thus, the concentration of fully liberated valuable minerals is significantly high in the fine fractions. It is imperative that such liberated valuable minerals be recovered economically at a faster rate. Advanced gravity separators/concentrators bridged this technological gap to provide economic and faster separation of fine minerals (Majumder and Barnwal 2006; Tripathy et al. 2017).

The recovery of the fine liberated particles in these concentrators is accomplished in mainly two ways: a) by subjecting the particles to a centrifugal force field and thereby enhancing the settling kinetics and b) by subjecting the settling particles to a counter-current flow and thereby effecting more intimate fluid–particle interaction to generate a hindered settling condition (Epstein 2005). The centrifugal force leads to enhanced g-force that significantly aids the stratification in the radial direction and concentration is achieved in a short time. On the other hand, examination of the hindered settling ratio clearly indicates that the density effect is significantly enhanced under hindered settling conditions (Wills 1997). This is effectively exploited in the teeter chambers which generate an autogenous heavy medium through counter current flow of solids settling under gravity and rising current of fluid for more effective gravity concentration (Sarkar et al. 2008a). Several useful applications have emerged by variations in the employment of these concepts and became integral parts of modern mineral processing. This work is a humble effort to review the process, application, science and engineering behind these developments.

2. Genesis of advanced gravity concentration
In gravity concentration, mineral/coal particles are separated into desired valuables and unwanted gangue depending on their density/specific gravity. In essence, the difference between settling velocities of two particles is the basis of separation in gravity concentration techniques: settling...
velocity of heavy particles will be higher than that of light particles. All gravity separation techniques work on a common principle in the sense that they employ gravity as the dominant force that drives different particles at different settling velocities. Separation performance of any gravity concentration depends on the difference between settling velocities of two separating particles which may be referred to as the potential for gravity separation.

Figure 1 presents the calculated particle settling velocities for different sizes and specific gravities in a fluid of density 1000 kg/m$^3$ (top). It can be seen that as the particle size decreased, difference in settling velocity between heavy and light particle decreased too. In other words, the potential for gravity separation in any conventional gravity concentrator decreased as particle size became smaller. Separation of particles based on density differences for smaller particles becomes difficult and as a result, performance of conventional gravity concentrators becomes poor for the fine particles. This is one of the major limitations of conventional gravity separation.

2.1. Hydraulic separation

In the last three to four decades significant research and developmental efforts are made in academia and industries to develop high throughput advanced gravity separators for fine minerals. These are developed based on the following principles: (1) enhancement of separation potential by application of centrifugal force (separators are known as EGS such as Knelson concentrator, Falcon concentrator, etc.), (2) generation of autogenous heavy media (separators are known as autogenous density separators such as teeter bed separators), and (3) application of a combination of centrifugal force and autogenous heavy media separation (such as multi-gravity separator, water-only cyclone, etc.).

Application of centrifugal force can increase the settling velocity and consequently, the separation potential. Figure 2 presents the calculated settling velocity of particles of different size and density under varied centrifugal force. Under normal gravity field, separation of particle smaller than 1 mm is difficult (Figure 2). However, as the effective acceleration increased from 1 g to 200 g, separation of particles smaller than 0.1 mm would be possible.

Similarly, separation potential can also be enhanced by increasing the density of separating liquid. Figure 1 shows that as separating liquid density is increased from 1000 kg/m$^3$ to 2000 kg/m$^3$, the separation potential increased. Density of separating liquid can be increased in two ways: (i) by mixing a third material with liquid; and (ii) by generating an autogenous heavy media. In the case of mixing a third material with separating liquid, maintaining stability of the mixture is always a difficult problem. Such stability problem of heavy media is not present in autogenous density separators where feed material itself participates in heavy media formation and help separating particles.

2.2. Pneumatic separation

Another issue with conventional gravity separation is the use of water as separating media that leaves the concentrate and tailing in wet condition. The problem is especially severe for fine coal cleaning where sometime clean coal contains more than 10% moisture (Houwelingen and Jong 2004). Removing ash from fine coal by using wet gravity separation becomes
non-beneficial due to high moisture contents. This problem can be overcome by introducing dry gravity separation techniques that eliminate the need of water. As a result, expensive dewatering operations, which need pumping, screening, filtering, and centrifuging, can be eliminated.

Several dry gravity separation techniques have been developed over the last two decades such as air jigs, air table, dry fluidized bed separators (FBS), Pardee spiral separator, FGX separator, pneumatic reflux classifier, and air dense media separation. Some of the more advanced dry gravity separators are presented in Figure 3. Air dense media separators provide several advantages in cleaning dry coarse coal (Chen and Yang 2003; Sahu et al. 2009). These dry gravity separators are different from conventional wet liquid-based gravity separators and their merits include them in the group of advanced gravity separators. We will discuss briefly the working principle of dry gravity separators.

The separation principle of particles in modern dry gravity concentration devices is similar to that of autogeneous density separators such as floatex density separator (FDS). Pneumatic density separators consist of a horizontal vibrating separating chamber/deck, vibrator, feeder, air chamber, and dust collector. Air passes through the perforated bottom of the separating chamber. Air flow rate is controlled in such a way that it fluidizes small heavy particles forming a fluidized bed that acts as the heavy medium. Sometimes sand is mixed with air.
and fluidized to achieve desired separation density. Low density particles float and heavy particles sink, and as a result, the particle bed stratifies into two fractions with heavy particles at the bottom and light particles on top. Pneumatic gravity separators can generate more than two products of different grades. The coarse coal separators can be used for coal beneficiation for feed size in the range of $-75 + 6 \, \text{mm}$. However, efficiency of separation for coal particles smaller than 6 mm becomes low and can not be used efficiently (Dwari and Rao 2007).

### 2.3. Features of pneumatic gravity separator

Separation of mineral particles in pneumatic density separators also follow similar principle as in conventional gravity separators: particle separation is based on density differences of constituent mineral particles. Differential settling velocities of particles determine the actual separation performance which can be modulated by varying air density. Magnetite or sand, suspended in air, is often used to form the heavy medium of target density. Separation of particles in pneumatic gravity concentration depends on the settling velocity under fluidized condition. As in wet gravity separators, particle settling velocity depends on its density, size and, shape, and principle of equal settling determines the separation performance for particles of various sizes and densities. The limiting ratio of size (diameter) of particles (known as settling ratio) that can be separated by gravity methods is $d_l/d_h$ (settling ratio) = $(\rho_l - \rho_f)/(\rho_c - \rho_f)$; $d_l$ and $d_h$ are diameters of light and heavy particles, respectively, and $\rho_l, \rho_h, \rho_f$ are the densities of light particles, heavy particles, and fluid, respectively. Thus, if we consider a system of particles of two different densities, $\rho_l = 2.0$ and $\rho_h = 5.0$, $\rho_{\text{air}} = 0.0 \, \text{g/cm}^3$, the settling ratio would be 2.5. This means, these particles can be separated by using air if $d_l/d_h$ is within 2:5:1. Now, if we can increase the effective density of air by mixing magnetite or sand to a higher value, say 1.0 g/cm³, the settling ratio increases to 4.0. Therefore, by simply increasing the density of air, limiting diameter ratio (settling ratio) and hence the performance can be enhanced significantly.

Apart from the density of separating medium, several machine and process parameters can influence the performance of the pneumatic density separators such as air flow rate, nature of feed material (density and size distribution, moisture in feed), machine type, vibration, size of the separator, type of sand (if sand is used), splitter position, etc. Air flow rate has direct influence on the separation density. As air flow rate increases, separation density decreases (Figure 4). With increasing air flow rate, fluidized bed expands, and as a result, effective density of the heavy medium decreases leading to the decrease in separation density.

Air dense medium gravity separation technique was used in coal beneficiation circuit by China University of Mining and Technology (CUMT) for cleaning $-50 + 6 \, \text{mm}$ coal particles. The beneficiation flow sheet designed by CUMT for 50 ton/hr capacity is presented in Figure 5 (Chen and Yang 2003). Application of dry pneumatic separation techniques have been increasing since then. Use of air-based gravity separator in different stages can be a useful strategy which can produce clean coal for industrial applications. For example, air table was used in different stages (Figure 6) to produce clean coal from high ash bearing (42% ash) feed coal for use in thermal power plant (Chalavadi et al. 2016a). Processing of high sulfur coal was studied by Yu et al. (2016) using a compound dry separator.

### 2.4. Optimal gravity circuit for valuable recovery

The optimum recovery of valuables from run-of-mine (ROM) ores can not be achieved by one single type of gravity concentration equipment since each equipment is designed to operate for a specific particle size range. A complementary set of gravity concentration devices should be used to achieve maximum valuables recovery from ROM ores. Approximate operating particle size range of different gravity concentration devices are presented in Table 1. If a combination of gravity concentrators is not adequate to achieve the maximum valuable recovery in a cost-effective manner, a combination of gravity concentrators, froth flotation, electrostatic, and magnetic separators/concentrators may need to be employed. We will briefly discuss how a combination of gravity concentration devices and other mineral processing devices can maximize valuable recovery.

Large amounts of fine minerals are produced during mining and comminution processes which go into the tailing. For example, about 50% (by weight) of total feed chromite ores become tailing and is rejected during beneficiation (Tripathy et al. 2013). Therefore, a lot of effort and resources are needed to recover valuables from the fine tailings where advanced gravity separators can play an important role. For example, the use of conventional gravity separation techniques such as Wilfley table, for chromite ore tailings remained unsuccessful to produce metallurgical grade chrome concentrate (Tripathy et al. 2013). The authors went on to show that the use of advanced gravity separation techniques such as FDS and multi-gravity separator, help improve chrome recovery and produce metallurgical grade chrome (45% mass yield with Cr/Fe ratio of 2.3) from the tailings.
3. Advanced gravity separators

Fluid based advanced gravity separators can be broadly divided into two categories:

a. EGS where centrifugal force field is employed to facilitate particle separation;

b. FBS where dense fluidized bed act as separation medium for fine particles often employing counter-current flow and hindered settling.

Centrifugal field is employed in Knelson concentrator, falcon concentrator, Kelsey jig, multi-gravity separator and water-only cyclone (also known as stub cyclone). The Knudsen bowl is also of the same generic type as the Knelson and falcon concentrator but less widespread and is not included in the review. On the other hand, the reflux classifier, crossflow separator (CFS), allflux separator, and the FDS work on counter-current flow principle. Size classifiers also operate on similar principles (Laleh et al. 2017; Tripathy et al. 2015) but are not within the purview of this review of advanced gravity concentration. In fact, many of the counter-current separators mentioned above are also used as size classifiers (Davis et al. 1989; Sarkar et al. 2008b).

3.1. Centrifugal separators

EGS have been used for fine mineral and coal concentrations for years. The separation principle of EGS is similar to that of any other conventional gravity separator, i.e.,
principle of differential settling velocities. However, unlike conventional gravity separators, a centrifugal force is applied to enhance the differential settling velocities between heavy and light particles. When particles are subjected to centrifugal force they are made to settle in the fluid in the radial direction. Depending on the centrifugal force and mass of the particle (which is related to particle size and density), each particle moves with a different settling velocity which helps separate particles from each other. The outward centrifugal force acting on the particle is quantified as: \( F_c = m\omega^2r \), where \( m \) is the mass of the particle, \( \omega \) is the angular velocity, and \( r \) is the radial location of the particle. In this case, \( \omega^2r \) is the centrifugal acceleration and is often 50 times or more of the gravitational acceleration. This enhanced radial acceleration is referred to as the enhanced gravity effect.

The buoyancy force, \( F_b \), is inward and is dependent on the particle volume and the density of the fluid. The third force acting on the particle is the fluid drag, \( F_d \), and for small particles it will be essentially the viscous resistance (Majumder and Barnwal 2006). The motion of the particle is governed by the influence of these three forces neglecting the interactive forces. Thus, the simplified force balance on the particle can be written as:

\[
\frac{du_p}{dt} = F_c - F_b - F_d
\]  
(1)

where, \( u_p \) is the particle settling velocity in the radial direction. Under steady state (terminal condition) particle settling in Stokes flow regime under centrifugal field could be expressed by Equation 2 which is equivalent to Stokes law under gravitational force field. In the formulation of this expression it was assumed that the drag force is essentially skin drag opposing particle motion.

\[
u_p = \frac{(\rho_p - \rho_f)d_p^2\omega^2r}{18\mu}
\]  
(2)

where, \( \rho_p \) and \( \rho_f \) are particle and fluid density respectively, \( d_p \) is the particle diameter and \( \mu \) is the fluid viscosity. Influence of centrifugal field on particle settling velocity can be seen from this equation. Luttrell et al. (1995), discussed the influence of centrifugal field on settling velocity of coal particles. Improving the settling kinetics of the fine particles under enhanced gravity is the main operating principle of the centrifugal separators. The salient features of these separators are discussed in the ensuing sections.

### 3.1.1. Knelsol concentrator

The Knelsol concentrator is one of the EGSs that works on a combined principle of centrifugal separation and fluidized bed separation. In a centrifugal field the particles form a bed around the inner periphery. The formation of the bed relies on the principle of hindered settling and interstitial trickling enhanced by the centrifugal force (Knelsol 1988). When fluidization water is introduced radially inward the particles will be in quicksand-like state. This concept is exploited in the Knelsol concentrator to achieve a stratification based on specific gravity on the bed of particles (Napier-Munn 1997). Knelsol concentrators mainly consist of a rotating conical inner shell (Figure 7) at the bottom of which feed slurry is introduced through a central vertical feed inlet. The inner shell has concentrating rings at the inner surface into which fluidization water is introduced uniformly through a series of fluidization holes. When the slurry is fed at the bottom of the cone, it is forced outward and driven up the cone wall toward the concentrate rings. Particles are trapped inside the concentrating rings and acted upon by the fluidization water. Under the centrifugal force, particle motion is determined by its size and specific gravity. Heavy particles experience a greater force and form the outer layer of the bed while the lighter particle form the inner (closer to the axis of rotation) layer. Fluidization water dilates the bed and forces the lighter particles out of the concentrating ring which flow out through the top of the inner shell into the overflow launder. Heavy particles trapped in the concentrating rings are taken out through pinch valves at controlled rates to maintain the stability of the fluidized bed.

**Figure 7.** Schematic diagram of Knelsol concentrator.

### Table 1. Effective operating range (particle size range) of different gravity concentration equipment (Chatterjee 1998).

<table>
<thead>
<tr>
<th>Concentration process</th>
<th>Minimum size (µm)</th>
<th>Maximum size (µm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Inline pressure jig</td>
<td>50</td>
<td>50,000</td>
</tr>
<tr>
<td>Conventional jig</td>
<td>120</td>
<td>10,000</td>
</tr>
<tr>
<td>Heavy media drum</td>
<td>400</td>
<td>50,000</td>
</tr>
<tr>
<td>Heavy media cyclone</td>
<td>180</td>
<td>20,000</td>
</tr>
<tr>
<td>Floatax density separator</td>
<td>45</td>
<td>500</td>
</tr>
<tr>
<td>Multi-gravity separator (MGS)</td>
<td>5</td>
<td>1000</td>
</tr>
<tr>
<td>Spiral</td>
<td>50</td>
<td>3000</td>
</tr>
<tr>
<td>Sieves</td>
<td>100</td>
<td>30,000</td>
</tr>
<tr>
<td>Shaking tables</td>
<td>18</td>
<td>10,000</td>
</tr>
<tr>
<td>Spinner</td>
<td>10</td>
<td>50,000</td>
</tr>
<tr>
<td>Falcon concentrator</td>
<td>9</td>
<td>80,000</td>
</tr>
<tr>
<td>Knelson concentrator</td>
<td>10</td>
<td>80,000</td>
</tr>
<tr>
<td>Kelsey centrifugal jig</td>
<td>4</td>
<td>3000</td>
</tr>
<tr>
<td>Reichert cone</td>
<td>50</td>
<td>9000</td>
</tr>
<tr>
<td>Continuous strake</td>
<td>52</td>
<td>7500</td>
</tr>
<tr>
<td>Plane table</td>
<td>51</td>
<td>2000</td>
</tr>
<tr>
<td>Hydrocyclone</td>
<td>6</td>
<td>200</td>
</tr>
<tr>
<td>Stub cyclone/water only cyclone</td>
<td>100</td>
<td>800</td>
</tr>
<tr>
<td>Pneumatic jig</td>
<td>400</td>
<td>40,000</td>
</tr>
<tr>
<td>Air table</td>
<td>80</td>
<td>800</td>
</tr>
<tr>
<td>Air fluidized bed density separator</td>
<td>6000</td>
<td>50,000</td>
</tr>
</tbody>
</table>
The Knelson concentrator uses a centrifugal field of about 60–100 g in most applications although they can go up to 180 g (Koppalkar 2009). It is widely used for recovering gold particles in a grinding circuit (Laplante and Shu 1992). It has also been shown to have the capability of beneficiating fine coal (Honaker and Das 2004). A combination of Knelson and spiral concentration was proposed as a possible substitute for flotation process, especially, at lower target ash levels. Honaker et al. (2005) also studied the effect of air injection to Knelson concentrator and established that recovery could be increased by 10–20% with air injection along with additional 2% ash reduction in the clean coal product. The effective separation density is reduced due to air injection which helps lower the ash level. Uslu et al. (2012) used Knelson concentrator for cleaning an oxidized coal with high sulfur content. They classified the −500 µm coal fines into three fractions, namely, −106 µm, −300 + 106 µm, and −500 + 300 µm fractions. They achieved a maximum combustible recovery of 99.1%, pyritic sulfur rejection of 91.6%, and ash rejection of 60.9%. They also reported a separation efficiency of 67.9% for pyritic sulfur and 39.5% for ash. Knelson concentrators have also been used with air as the fluid medium to investigate tungsten recovery from a synthetic mixture of tungsten and quartz which closely simulates gold ore (Greenwood et al. 2013; Kökkiliç et al. 2015; Zhou et al. 2016). Although the recovery of tungsten could be increased significantly in these studies under optimum conditions, the concentrate grades were much lower than what could be achieved under wet operating conditions (Greenwood et al. 2013). From the above discussion, it is evident that the Knelson concentrator is very useful for fine particle processing in a number of applications, especially where the specific gravity differential between the valuable and the gangue is high. Although use of pneumatic separation in the Knelson concentrator is feasible, it is still not attractive from efficiency perspective.

3.1.2. Falcon concentrator

In 1986 a commercial unit of falcon concentrator was first installed for fine gold particle concentration (Napier-Munn 1997). A falcon concentrator consists of a vertical cylindrical open-topped bowl that is mounted on a rotating shaft (Figure 8). The bottom portion of the bowl is tapered. The bowl may be classified into two zones. The lower section is called the migration zone while the upper section is referred to as the retention zone (Honaker et al. 1996). The g-force can be varied by changing the rotational speed of the bowl. Details of the separation principle of a falcon concentrator are discussed by Falconer (2003).

The feed slurry is introduced at the bottom of the bowl through a central pipe. The feed is then forced to the wall of the bowl by means of an impeller. Strong enhanced gravity force resolves into two components: (i) a force component normal to the wall and (ii) a force component parallel to the wall. Stratification starts at the migration zone by the action of impeller due to differential acceleration. Normal force is very strong and it induces radial velocity on the particle. Heavier and coarser particles have higher radial velocity while lighter and smaller particles have lower radial velocity. The heavier particles form a bed adjacent to the wall of the bowl and the lighter particle layer remains at the farthest location from the wall. Weak parallel force component helps in migration of layers in upward direction. In the retention zone the upward movement of the heavier layer is restricted so as to facilitate overflow discharge. The heavy layer clings to the wall and is discharged through pinch valves fitted on the wall of the bowl in the retention zone.

A centrifugal field of about 300 g (Honaker et al. 1996) may be achieved in a falcon concentrator. It is used widely in gold ore concentration (Laplante et al. 1996; Zhang 1998; Alp et al. 2008). A continuous falcon concentrator is also shown to be applicable in fine coal cleaning (Honaker et al. 1996; Honaker 1998; Tozsin et al. 2016; Zhu et al. 2017a, 2017b). The high recovery of valuables along with a high separation efficiency are added advantages of this device (Honaker et al. 1996). Honaker et al. (2000) established that the performance of the falcon concentrator nearly approaches the feed coal washability which corresponds to the maximum achievable yield at a target ash level. However, it was also shown that the falcon concentrator gives a poor performance in ultrafine tantalum recovery (Burt et al. 1995). The prospects of recovering heavy metals from a rare earth slime was investigated by Marion et al. (2017). The authors used heavy medium (lithium metatungstate) in a modified falcon concentrator. The grades of the heavy metals in the concentrate were comparable with those obtained using a laboratory heavy medium centrifuge. Excellent rejection of silicate gangue in the light fraction of the falcon concentrator (~91% Si recovery) was obtained. However, recovery of Zircon and iron oxides in the heavy fraction were observed to be low (~48% recovery of Zr and ~31% recovery of Fe). Thus, the falcon concentrator can be employed in a wide range of applications such as gold, coal, heavy minerals, etc.

3.1.3. Kelsey jig

Centrifugal jiggling was successfully employed by Chris Kelsey in a Kelsey jig (Napier-Munn 1997). Kelsey jigs can operate with an increased gravitational force of about 100 g to accomplish gravity separation under high centrifugal field (Singh and Das 2013). They are capable of separating particles down
to a few microns (Tucker 1995; Falconer 2003). A Kelsey jig consists of a spinning rotor which generates the centrifugal force. A parabolic inclined screen is spun co-axially together with the rotor (Figure 9). Ragging material spreads on the screen surface evenly. The feed slurry is introduced through a central pipe and is distributed evenly on the ragging material. Particles having specific gravity greater than that of the ragging material would pass through the ragging bed. The movement of particle through the ragging bed can be explained considering the differential initial acceleration, hindered settling and consolidation trickling of the particles (Wills 1997). Heavier materials pass through the ragging bed and then through the screen into the concentrate hutch. These materials then pass through a spigot and accumulate in a reservoir. Lighter particles are unable to pass through the ragging bed. They move upward and discharge over the top of the screen and ragging retention ring and constitute the tailing.

The ragging bed is pulsated by water using mechanical pulse arms which are connected to pads. These pads work against the jig’s flexible diaphragm. The diaphragm is sealed to an outer cut-away section of each of the four concentrate hutches. The ragging bed acts like a packed bed and prevents the free passage of the particles through the bed. The dilution effect introduced by the pulsating water allows heavy minerals to pass through the ragging bed. Pulsations also accentuate the different rates of acceleration between particles of differing specific gravities. The particles are accelerated at a rate proportional to their mass until they approach a critical velocity which is related with the surface area (Jones 2006). They are not allowed to reach their critical velocity where surface area determines the critical velocity. The shockwaves continually stop the particles to reach their critical velocity. The particles are separated at their initial settling regime throughout the process. The Kelsey jig was first applied in the recovery of fine gold particles from plant tailing of Bougainville Copper Ltd. (Napier-Munn 1997). Kelsey jig is used widely for beneficiation of beach sand (Jones 2006). It can also be used for recovery of tungsten from plant tailings (Tucker 1995). Evidently, the Kelsey jig has been successful in a number of mineral processing applications.

### 3.1.4. Multi-gravity separators

The difficulty of separation of fine particles in conventional tabling processes was largely overcome by introducing the multi-gravity separator (MGS). The principle involved in the operation of the MGS is discussed by Traore et al. (1995), Falconer (2003), and Goktepe (2005). The MGS consists of a horizontal drum fitted with a gentle slope (Figure 10). The cylindrical drum which is slightly tapered, rotates and shakes simultaneously. The forces in conventional tabling are combined with the centrifugal force and the net effect of these forces are exploited by the MGS to effect gravity-based separation. The drum is rotated in a clockwise direction at a rotational speed between 100 and 250 rpm. Sinusoidal shaking action is applied on the drum axially. The frequency of shaking may be varied between 4.0 and 6.0 cps whereas the amplitude may be varied from 10 to 20 mm. It is obvious that with such a low rpm, the g-forces are not really high in this case.

Feed slurry is introduced continuously at the mid-point on the internal surface of the rotating drum through a mesh ring so as to reduce the turbulence due to entry effect. Wash water is added near the open end through a similar type of mesh ring. A flowing film forms on the internal surface of the drum. The heavier particles settle on the inner surface and move in the upward direction by the shaking action and accumulate in the concentrate stream. In contrast, the lighter

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**Figure 9.** Mid-sectional view of Kelsey centrifugal jig.
particles do not settle and are carried to the tailing end by the wash water.

The MGS is reported to be successful in the concentration of even \(-5\,\mu m\) particles of rare earth ore (Ozbayoglu and Atalay 2000). Burt et al. (1995) proposed a strategy for ultra-fine tantalum recovery using the MGS. In their study it was found that the MGS offered better recovery than conventional high gradient magnetic separation and oil phase extraction. Recovery of valuables from fine particles was studied by Traore et al. (1995). It was reported that the MGS performs better than fines table for fine particles. Singh et al. (1997) reported processing of iron ore slimes (\(-105\,\mu m\)) and chromite slimes (\(-40\,\mu m\)) in MGS to recover valuables. They also observed that the MGS performed better than the conventional gravity concentrators for iron ore slimes. However, for the chromite slimes, the enrichment was found to be unsatisfactory. Recovery of chromite fines was also studied by Çiçek and Cöcen (2002) using the MGS and it was concluded that it is capable of reducing fines loss in gravity tailings. They concluded that the MGS can increase the chromite concentrate yield by up to 11%. Dependence of MGS performance on particle size and tilt angle was investigated by Goktepe (2005) for treating lead mine waste. It was established that about 12% of the lead can be recovered from mine waste. Celestite concentration using MGS was investigated by Aslan (2007). It was concluded that about 96% of the celestite can be recovered under optimized conditions. The MGS is widely useful especially in the final concentration stages where relatively lower volume of materials needs to be processed.

3.1.5. Water-only cyclone

While the hydrocyclone is essentially used for size classification, the water-only cyclone (WOC) is an enhanced gravity separator where particulate suspension forms an artificial heavy medium and particles are separated on the basis of their specific gravity. The structure of WOC and hydrocyclone have some similarity but significant differences are also notable along with the separation principle. The WOC consists of a vertical cylindrical body followed by a conical bottom (Figure 11). It has much wider cone angle (80–140°) than that of a hydrocyclone (10–70°). The vortex finder is much longer in a WOC than in a hydrocyclone. Feed slurry is introduced tangentially from the top of the unit. The particles slow down drastically at the conical bottom as its cone angle is very wide and they get accumulated. Thus a bed of particle is formed which acts like a heavy medium (Patil et al. 1997). The WOC has some advantages such as: (1) creation of autogenous heavy medium; (2) lack of any moving parts inside the unit, hence the maintenance and operating costs are low; (3) independent of particle surface properties, so oxidized ores/ coals can be treated effectively; (4) sharpness of separation is higher when compared to that of spirals, tables, and fine ore jigs; and (5) low cost of washing.

The application of WOC is mainly confined to coal cleaning, removal of graphite from lead ore etc., where the separation densities are relatively lower. Patil et al. (1997) used WOC for removal of graphite from lead rougher concentrate. Their study demonstrated that a concentrate could be produced with 39% lead and 2.6% graphite with 45% recovery of lead from a feed of 19.6% lead and 9.8% graphite. The application of WOC in coal washing circuit was also investigated by Suresh et al. (1996). They established that a Rosin-Rammler type of equation can be used to predict the performance of WOC for coal washing. The WOC has limited applications beyond coal.

3.2. Counter flow fluidized bed separators

The counter-current gravity separators rely essentially on fluidization and hindered settling. Bed teetering, hindered settling and formation of an autogenous heavy medium are common in these separators. The basic separation principle is
the same for all these separators which differ mainly in the 
feed and discharge designs and other innovative attachments 
that affect settling behavior for improved performances under 
different circumstances.

3.2.1. Floatex density separator
Teeter bed separators are hindered settling-based hydraulic 
separators which are widely used for mineral beneficiation. 
The FDS is one such counter-current, autogenous teeter bed 
separator. Particles are separated based on their differential 
specific gravity. The FDS produces two products, namely, an 
underflow enriched in heavier particles and an overflow rich 
in low density particles. High density particles settle against 
the rising teeter water (TW) through a bed acting as an 
artificial heavy medium. Settled particles are accumulated at 
the bottom forming a teeter bed. Since the size and density of 
the particles are not uniform, the particles are segregated 
according to their mass (Luttrell et al. 2006). In general, the 
coarser-heavier particles form a layer at the bottom of the bed 
and the coarser-lighter particles form the top layer in the bed. 
Other particles are distributed throughout the bed depending 
on their density and size. Detailed description of the FDS is 
available elsewhere (Honaker and Mondal 2000; Luttrell et al. 
2006; Sarkar et al. 2008a). A schematic diagram of the FDS is 
shown in Figure 12. It consists of a vertical square tank 
attached to a conical bottom. Feed slurry, along with feed 
water, is introduced through a centralized feed well. TW is 
introduced uniformly from the bottom of the square tank 
through a bank of perforated tubes. A pressure sensor is 
mounted at the bottom of the square tank. An underflow 
pinch valve is operated automatically using a PID controller 
and an actuator. The conical bottom helps in dewatering the 
underflow product which is discharged at the bottom.

Galvin et al. (1999a) predicted the FDS performance using 
an empirical relative velocity equation and concluded that the 
performance improved with increasing suspension density, 
and that the use of synthetic heavy media was beneficial in 
this regard. Their study also showed that the cut density 
increased as the particle size was decreased. However, this 
study was limited to only particles of size ~2.0 + 0.375 mm. 
Efficacy of hindered settling separation in combustible recov-
ery from Korean anthracite was studied by Cho and Kim 
(2004). They concluded that with an increase in both the 
TW flow rate and the bed pressure (BP), the recovery 
increased without much deterioration in the product quality. 
A dynamic population balance model of FDS was developed 
by Honaker and Mondal (2000), and extensive simulation 
studies were carried out. The authors recommended 3:1 size 
ratio between the coarsest and the finest fractions of a pre-
sized feed for effective concentration in a hindered bed clas-
sifier. Luttrell et al. (2006) observed that the teeter beds 
performed better as a concentrator with pre-classified feeds 
having a size ratio of 4:1 between the coarsest and the finest 
fractions. A fluid dynamics-based model was developed to 
study the effect of particle size distribution in the FBS (Wei 
and Sun 2016). They found that the finer the feed the greater 
is the separation density and the Ep value. Both these para-
eters decrease with a smaller variation in the particle size. 
Sarkar et al. (2008b) investigated fine coal processing in the 
FDS. They concluded that below a threshold BP the FDS acts 
like a size classifier rather than a concentrator. A narrowly 
sized feed was shown to result in improved performance by 
reducing the size effect on gravity separation. Therefore, the 
FDS may be result in a more effective concentration in con-
junction with a classifier.

3.2.2. Reflux classifier
The reflux classifier is also a teeter bed separator in which the 
基本 operating principle is the same as what is described 
earlier for the FDS. The faster settling particles form the 
bottom of the particle column and the slower settling particles
are hydraulically carried to the overflow. However, the reflux classifiers have a set of inclined plates that aids the settling of finer and lighter particles and enhances the throughput significantly (Nguyentrinlam and Galvin 2001). These classifiers can be used for size classification as well as gravity concentration. The feed is introduced below the inclined plates while the TW is introduced at the bottom of the chamber. Slow settling particles are carried by the rising water and are separated from the fast settling particles. As the upward flow passes through the inclined plates, the particles settle onto these plates and slide down (Figure 13) to give different products. Essentially, the fine heavy particles are separated from the fine light particles by the plates. The former are made to accumulate back into the system. Faster settling particles are removed from the chamber at the bottom through a valve. The reject suspension generates the autogenous heavy medium which forms the basis of gravity concentration. Heavy particles are able to penetrate this heavy medium and report to the underflow while light particles having a low settling velocity are carried to the overflow.

Pilot scale testing of the reflux classifier was carried out by Galvin et al. (2002) for the beneficiation of ~2 mm coal fines. Excellent gravity concentration was reported with a throughput of 47 t/(m² h) which is much higher than what can be achieved in a conventional teeter bed separator. The reflux classifier was also shown to offer less variation in separation density with particle size compared to the conventional teeter bed separator. Pilot scale testing of ~150 micron iron ore fines was also carried out by Amariei et al. (2014) to separate hematite from siliceous gangue. A spiral feed and a spiral tailings were tested in this investigation. It was shown that iron minerals lost in the spiral tailings could be efficiently recovered by the reflux classifier with a final tailings nearly completely depleted of iron values. Orupold et al. (2014), studied the Lamella High Shear Rate reflux classifier for ~2 mm coal application. It was established that this classifier can process coals at lower cut points with a high efficiency. Thus, low ash clean coals can be easily produced by these classifiers at improved yields. They also reported that promising test results for ~1 + 0.106 mm iron ore sample were achieved using the reflux classifier. Galvin et al. (2016) employed reflux classifier with closely spaced channels for the separation of heavy minerals from minerals sand. The closely spaced channels accomplish improved gravity-based separation through laminar-shear mechanism in single stage operation. The feed with about 5% heavy minerals was upgraded to nearly 80% heavy minerals in the concentrate at about 85% recovery level for feed rates up to about 15 t/(m² h). The recovery remained very similar up to about 18 t/(m² h) and dropped to 75% around 25 t/(m² h) feed rate. Notably, the concentrate grade achieved at high throughput, ~25 t/(m² h), was ~88% heavy minerals, which was attributed to the enhanced conveying of the low-density coarse sands upward through the channels. At higher throughputs, more than 95% recovery levels of ZrO₂ were achieved. Both classification and concentration may be efficiently accomplished in the reflux classifier.

3.2.3. Crossflow separator
The CFS differs from a conventional teeter bed separator essentially in the feed entry system. The former employs a tangential and low velocity feed entry system through a feed well located outside the main chamber but attached to it (Figure 14). The feed water is prevented from causing any disturbance to the flow behavior in the main teeter chamber. The feed solids in such a feeding system has only axial velocity and thus closely mimics the one-dimensional flow system very closely (Kohmuench et al. 2003). The feed solids overflow from the stilling well into the main chamber and settle against the rising TW. The feed water also enters the main chamber but travels across and reports to the overflow. Thus, the feed water is not allowed to cause any hindrance to the internal flow in the separation chamber. The resulting quiescent flow situation helps in improving the separation efficiency.

Tangential low velocity feed delivery system indeed leads to a more efficient separation for coal fines and also enhances the capacity (Dunn et al. 2000). These advantages are primarily due to a more quiescent flow regime in the teeter chamber. Plant-level testing of a CFS was carried out by Westerfield (2004) for coal fines. It was demonstrated that for ~1.0 + 0.2 mm size particles the separation efficiencies were higher than existing classifiers. The CFS resulted in a higher clean coal yield than that of a spiral concentrator at the target ash or sulfur levels. It was also established that the CFS could accommodate the entire flow of multiple spirals in a single-stage circuit.

3.2.4. Allflux separator
The allflux separator accomplishes two-stage concentration/classification in a single unit. The central chamber is

Figure 13. Schematic of the reflux classifier.
somewhat similar to the conventional teeter columns but with significant differences (Figure 15). The feed slurry is introduced in a swirl flow so as to distribute the feed toward the periphery. Also some stratification is obtained while feeding in this manner with heavier/coarser particles thrown further than the lighter/finer particles due to the centrifugal force. The fluidizing water helps generate the autogenous heavy medium through which heavier/coarser particles with a faster settling kinetics settle. They are collected through the peripheral annular opening into a densifying cone and are discharged through the underflow valve. The slow settling finer and lighter particles are hydraulically transported into a peripheral ring at the top. This section is the fines section where fine heavy particles are separated from the fine light particles. The fines section also works on similar fluidization and hindered settling principles. Of course, the rising water velocities are much lower than that in the main chamber. The overflow of the fines section contains essentially the light ultrafine particles and the heavy fraction at the bottom constitutes the third product of the separator.

The allflux separator has been successfully used in iron ore and coal beneficiation as well as chromite beneficiation, heavy mineral sand concentration, and slag classification. Studies on allflux separator was reported for the treatment of South African coal (Piennar 2000). A feed coal with 30.2% ash was enriched to 14.5% ash in the clean coal with a mass yield of 50%. The allflux separator was employed in the beneficiation of iron ore also (Williams et al. 2012). In this case, the classifier was used as a pre-concentrator. A hematite ore was classified into different fractions for further downstream processing. In another investigation, iron ore concentration was carried out in allflux separator for particles finer than 1.4 mm
3.2.5. Air dense medium separator (ADMS)

In dry air-dense medium fluidized bed separator, a medium is created by suspending solid particles in an upward direction of air flow (He et al. 2016). Two types of material for dense medium are used for the system, namely, magnetic pearl for low density medium and fine magnetite powder for high density medium. This separator is mainly used for coal preparation. In some cases, fine coal powder is added to the magnetite medium for creating a stable fluidizing bed. By means of a two phase gas-solid pseudo-fluid separating medium, the light and heavy particles stratify in the fluidized bed according to their individual densities. The bed density is more or less the same throughout the fluidized region.

Fraser and Yancey (1926) first described the fluidized bed process which used river sand as the fluidized medium with a bulk density of 1.45 g/cm³ for the beneficiation of coal. Using fine magnetite and sand particles Lohn (1971) developed a fluidized bed device with continuous flow conditions generating densities in the range of 1.7–2.2 g/cm³ and 1.2–1.4 g/cm³, respectively. Chen and Yang (2003) reported that Mineral Processing Research Centre of China University of Mining and Technology (CUMT) developed the ADMS in which magnetite particle suspension formed the medium for treating −50 + 6 mm coal. It was claimed that the construction and operational costs were less than half of those of a hydraulic cleaning plant. Zhenfu et al. (2002) studied the mechanism and separation efficiency of coal particles in the ADMS having a cross-sectional area of 150 mm × 200 mm and with the size of ROM coal feed of −50 + 6 mm with an ash content of 21.48%. The experimental results showed that the ADMS provided a good separation performance for this coal with a clean coal ash content of 11.80% and a refuse ash content of 85.75% with an Ep value of 0.03. In another study using the ADMS, it was shown that under optimum conditions the ash content of the coal (−25 + 6 mm) was reduced from 40% to 31% with 70% mass recovery in the clean coal (Sahu et al. 2005). Understandably, the ADMS is confined to coal applications only.

3.2.6. Air tables

Air tables employ fluidization in a shallow bed through which particle settling occurs. The table has a wire mesh screen deck to support the material while air is introduced from underneath the deck. The table has both longitudinal and transverse inclinations with mechanism for its oscillation (Figure 16). A complex interaction of gravity forces, fluidization forces as well as oscillatory forces enforce stratification and movement of the particles on the deck surface. The trajectories of the stratified heavy and light particles are differentiated and collected at separate locations of the table. Air tables are used mainly for coal preparation although plastics separation have also been investigated (Dodbiba et al. 2003).

The air tables at the Florence Mining Co., Pennsylvania processed 1200 tph coal having a top size of 19 mm (Wright 1985). In order to obtain reliable performance data a series of tests with air tables in two different preparation plants were performed (Killmeyer and Deurbrouck 1979). The feed top size was kept at 50 mm and the feed rate was varied from 90 to 150 tph. The air tables provided high-separation density cuts between 1.78 and 2.67 with an Ep value of 0.3. The poor performance was attributed to significant quantity of clay in the feed coal. Stotts et al. (1987) conducted tests at the Federal mines, Elkhorn city, Kentucky to evaluate the effect of feed moisture and the performance of air tables for −19 mm feed coal. The tests results showed that Ep values were 0.24–0.26 for particle sizes above 6.3 mm. Below 6.3 mm size, the efficiency was so poor that an Ep could not be calculated. Pilot-scale tests conducted at two coal mine sites in Utah showed that the dry separation of coarse coal −50 + 6.3 mm is feasible with a negligible loss of coal to the reject stream (Honaker et al. 2006). Data indicated that when treating −50 + 6.3 mm run-of-mine bituminous coal 70%–90% of the rock with greater than 2.0 relative density could be rejected. The potential of dry cleaning of coal of varying ranks using the air table was also evaluated by Honaker et al. (2008). Results showed that rock removal into the reject stream was achieved with little loss of coal, regardless of the mineral matter type. As a result, a saleable product was generated from several coal sources including lignite, sub-bituminous, and bituminous coals. Recently, Chalavadi et al. (2016b) has shown that nearly 10% (absolute) reduction in the ash level is possible at over 40% yield of the clean coal while treating −1.0 mm coal in an air table in single stage operation. In view of the low density of air, the air table finds applications are limited to coal and light materials such as plastics only.

![Figure 16. Schematic diagram of air table (adapted from Dodbiba et al. 2003).](image-url)
4. Theoretical formulations

Hindered settling is at the core of separation in all the advanced gravity separation methods discussed. In many instances an autogenous heavy medium develops. This artificial heavy medium determines the density of separation. Particles heavier than this medium settle through while the lighter ones stay on top of the medium. The lighter fraction is often removed by hydraulic/pneumatic transport. Of course, for specific systems additional mechanisms are employed to aid the separation process. It is also well known that particle size influences gravity separation significantly. Therefore, many of the advanced gravity separators are also used as size classifiers under special circumstances in which specific gravity also plays significant role. However, in this article efforts were made to clearly distinguish between classification (separation based on size) and concentration (separation based on specific gravity) with very little or no emphasis on the former. The introduction to size classification is mainly for illustration.

4.1 Fundamentals of centrifugal gravity separators

Settling takes place in the radial direction in these cases due to the action of the centrifugal force. Quite high centrifugal forces are usually applied in order to enhance the settling kinetics substantially. Therefore, most of these operate with a high g-force. The g-force employed in the multi-gravity separator is the lowest (around 25 g) with a maximum achievable g-force of around 300g in a falcon concentrator. A brief description of the features of various centrifugal gravity concentrators are given in Table 2.

The fundamentals of centrifugal gravity concentrators lie in the estimation of particle settling velocity. Settling velocity of an individual particle is determined from its physical properties such as size, specific gravity, and shape. Interaction between the fluid and the particle as well as inter-particle interactions play significant roles in the determination of settling velocity. In fact, there are a large number of interactive forces acting on these systems. Thus, the settling velocity is not quite observed to be given by Equation 2.

Form drag

The first complicacy that can be added is that when the skin drag changes to form drag. Under these conditions, the settling velocity is expressed as:

$$u_{ps} = \left[ \frac{3(\rho_p - \rho_f) d^2 \omega^2 r}{\rho_f} \right]^{1/2}$$

This is the equivalent of Newton’s law in a centrifugal force field.

Added mass

When a particle moves in a fluid, some amount of fluid must move around it. When the particle accelerates, the fluid also does so. Thus, more force is required to accelerate the particle in the fluid than in vacuum. Since force equals mass times acceleration, it may be said that this additional force is arising out of an imaginary added mass of the particle in the fluid. The added mass will depend on the shape of the particle. For a spherical particle the added mass is expressed as:

$$m_a = \frac{1}{12} \rho_f \pi d_p^3$$

### Table 2. Salient features of various centrifugal gravity concentrators.

<table>
<thead>
<tr>
<th>EGS</th>
<th>Salient features</th>
<th>Disadvantages</th>
<th>Applications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Knelson concentrator</td>
<td>• About 60g is achievable</td>
<td>• Large amount of fluidization water is required</td>
<td>• Alluvial gold, coal, etc.</td>
</tr>
<tr>
<td></td>
<td>• Misplacement is minimized using wash water</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Both semi-batch and continuous units are available</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Falcon concentrator</td>
<td>• About 300g is achievable</td>
<td>• Requires feed to be screened to less than opening size of concentrate orifices to prevent blinding</td>
<td>• Coal, iron ore, lead, zinc, copper etc.</td>
</tr>
<tr>
<td></td>
<td>• High capacity</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• No wash water</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Both semi-batch and continuous units are available</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Good metallurgical performance</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Able to treat particle size down to 15–20 micron</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Mechanically simple and robust</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Low operator attention required</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• About 60g achievable</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• High capacity</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Much finer particle could be separated</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Narrow specific gravity difference required</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Kelsey jig</td>
<td>• About 60g is achievable</td>
<td>• Complex system</td>
<td>• Tin, chromites, iron ore, gold etc.</td>
</tr>
<tr>
<td></td>
<td>• High capacity</td>
<td>• Relatively high capital and operating cost</td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Much finer particle could be separated</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Narrow specific gravity difference required</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Multi-gravity separator (MGS)</td>
<td>• About 25 g is achievable</td>
<td>• Low capacity</td>
<td>• Zinc, lead, copper, tin, gold etc., iron ore slime</td>
</tr>
<tr>
<td></td>
<td>• Able to handle very fine particles (–75 + 10 micron)</td>
<td>• Mechanically complex</td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Excellent metallurgical performance</td>
<td>• Unsuitable for treating coarse materials</td>
<td></td>
</tr>
<tr>
<td></td>
<td>• Excellent metallurgical performance</td>
<td>• Require operator attention</td>
<td></td>
</tr>
<tr>
<td>Water only cyclone</td>
<td>• About 40 g is achievable</td>
<td>• Larger diameter units are inherently inefficient.</td>
<td>• Fine coal cleaning</td>
</tr>
<tr>
<td></td>
<td>• Very simple operation</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>• High capacity</td>
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</tr>
</tbody>
</table>

The first complicacy that can be added is that when the skin drag changes to form drag. Under these conditions, the settling velocity is expressed as:

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This is the equivalent of Newton’s law in a centrifugal force field.

### Added mass

When a particle moves in a fluid, some amount of fluid must move around it. When the particle accelerates, the fluid also does so. Thus, more force is required to accelerate the particle in the fluid than in vacuum. Since force equals mass times acceleration, it may be said that this additional force is arising out of an imaginary added mass of the particle in the fluid. The added mass will depend on the shape of the particle. For a spherical particle the added mass is expressed as:

$$m_a = \frac{1}{12} \rho_f \pi d_p^3$$
This added mass when multiplied by the acceleration of the particle gives the additional force.

**Shear lift**

Small particles in a shear field experience a lift force perpendicular to the direction of flow. The shear lift originates from the inertia effects in the viscous flow around the particle. This lift force is also known as Saffman force (Saffman 1965) and is given as:

\[
F_l = 1.615 \left( \rho \mu \right)^{0.5} d_p^2 (u_f - u_p) \left[ \frac{du_f}{dy} \right]^{0.5}
\]

(5)

**Magnus effect**

This effect is associated with the spinning of the particle which is quite relevant in the case of centrifugal separation. For a particle spinning about an axis perpendicular to its direction of settling, the velocity of the particle, relative to the fluid, is different on opposite sides. Since drag is dependent on the velocity, the opposite sides experience differential drag. The unequal drag introduces a net side force acting on the particle which is known as Magnus Effect. This force is estimated as:

\[
F_M = \frac{\pi}{8} \rho_i d_p^3 \omega (u_f - u_p)
\]

(6)

where, \( \omega \) is the rotational speed of the particle.

**Basset force**

When a particle moves in a fluid boundary layer development lags due to changing relative velocities with time. The viscous effects due to the temporal delay in boundary layer formation are accounted for by the use of Basset force. Because of the time factor in this force it is also known as the 'history' term. It becomes very large if the particles are accelerated at high rate. Otherwise, this force is often neglected due to its low magnitude and difficulty in implementation. For an accelerating spherical particle, this force is expressed as:

\[
F_B = \frac{3}{2} d_p^2 \pi \rho_f \mu \left[ \frac{du_f}{dt} - \frac{du_p}{dt} \right] \int_0^t \left( \frac{du_f}{dt} - \frac{du_p}{dt} \right) dt
\]

(7)

**Bagnold forces**

When a fluid film containing suspended particles is subjected to shear, pressure develops across the plane of shear perpendicular to the surface. The pressure is proportional to the square of the particle diameter and rate of shear (Hunt et al. 2002). The Bagnold force leads to a reverse classification in which coarse light particles appear at the top while fine heavy particles report to the bottom of the film. In order to maintain a high shear rate by fluid motion, the flow must be fast and move down an inclined surface (Burt 1984). Thus, a large tonnage can be treated through exploitation of the Bagnold forces which is a shear-induced dispersive force (Holtham 1992).

**Interactive forces**

Fluid-particle and particle-particle interactive forces are common in the systems under consideration. However, these forces are difficult to account for numerically. Usually, in systems containing particles of more than one size and density, these interactions are incorporated in terms of an interaction coefficient, \( \beta \), with subscripts to denote different interacting phases. This coefficient accounts for collisions and contact between particles of different size by transferring momentum between phases. The product of the interaction coefficient and the slip velocity between the corresponding phases is included in the momentum balance equation to account for the interactive forces. A model for the interaction coefficient was proposed by Syamlal (1985). Bell et al. (1997) suggested modifications of the model to offer a new particle-particle interaction coefficient that can be successfully used in fluid dynamic calculations. Bagnold force, as discussed earlier, is essentially a kind of particle-particle interactive force.

**4.1.1. Knelson and falcon concentrators**

These operate in a very similar manner. Density-based stratification in the radial direction is accomplished with the heavier layer close to the periphery and the lighter layer closer to the axis. In fact, fluidization is also employed radially inward to loosen the bed and release any light material entrapped in the bed. In the Knelson concentrator, the fluidization water is introduced tangentially. It is counter-current to the bowl rotation to prevent any mass transfer between the material layer already recovered in the inner portion of the ring and the material layer recovered subsequently (Koppalkar 2009). The axial flow of the fluid recovers the lighter layer in the overflow. Gravity force is often negligible in these separators.

The concentration mechanism in a rotating vessel was explained by Ferrara (1960). He derived an expression for the motion of the particle in a cylindrical tube. It was assumed that only three forces are active; centrifugal force, fluid drag and frictional force between the particles and the tube wall. A laminar flow condition was also assumed.

\[
\frac{4}{3} \pi \left( \frac{d_p}{2} \right)^3 \rho_p \frac{du_p}{dt} = - \frac{4}{3} \pi \left( \frac{d_p}{2} \right)^3 \rho_p - \rho_f \Psi \omega^2 \left( \frac{d_i}{2} - \frac{d_p}{2} \right)^3 - 18 \mu \kappa \left( \frac{d_p}{2} \right)^3 Q \left( \frac{d_p}{2} \right) - 6 \pi \kappa \mu \left( \frac{d_p}{2} \right)
\]

(8)

where, \( \omega \) is the angular speed of the tube, \( \Psi \) is the coefficient of friction between the particles and the tube wall, \( \kappa \) is the coefficient of non-sphericity of the particle, and \( Q \) is the flow rate of the fluid through the tube.

Buonvino (1993) noted that the major setback to this formulation is that it ignores particle concentration and inter-particle interactions. Also, it is difficult to track particle trajectory experimentally at higher particle concentrations. He suggested two recovery mechanisms in the Falcon concentrator; the coarse (and heavy) particles are recovered in the underflow as they get buried deep into the bed while fine (and light) particles are lodged in the interstices at the bed surface and recovered in the overflow stream.
A mechanistic model for the Knelson concentrator was suggested by Coulter and Subasinghe (2005). They considered the three major forces determining the stratification are centrifugal ($F_c$), fluid drag ($F_d$), and Bagnold dispersive force ($F_b$). The centrifugal and Bagnold forces are both dependent on the angular speed and therefore can be clubbed together. This net centrifugal force was termed $P_c$, and it was postulated that the ratio of the drag force and this net force was the factor ($X$) that determined the particle retention by the concentrate rings. A high value of $X$ indicates a relatively larger drag due to greater fluidization which aids in the ejection of the particle in the overflow. On the contrary, a low value of this factor indicates that fluidization is less significant and the bowl rotation plays a more dominant role which aids in the retention of particles in the concentrate rings. The authors used a synthetic mixture of silica and magnetite in a laboratory Knelson concentrator to develop an expression for volumetric material retention, $V_i$, as shown:

$$V_i = V_{0i} \exp \left[ - \frac{X_i}{X^*} \right]$$  \hspace{0.5cm} (9)$$

where, $V_{0i}$ is the maximum volume of material retained under a given set of conditions, $f_i$ is the volume fraction of mineral in the feed, $X^*$ is the critical value of $X$ at the transition between the two regions, and $n$ is an exponent. Based on this approach, Ghaffari and Farzanegan (2017a) proposed a very similar model for the retained mass of the $i$th material of $j$th size class as:

$$R_{ij} = \rho_{ai} \left[ R_{e\,\text{max}} - \omega \exp \left[ -c \left( \frac{3P_w}{2(\rho_{ai} - \rho_w)D_{ij}r_{ai}} \right)^d \right] \right]$$  \hspace{0.5cm} (10)$$

where, $R_{e\,\text{max}}$, $b$, $c$, and $d$ are model parameters, $P_w$ is the fluidization water pressure, $\rho_{ai}$ and $\rho_w$ are the densities of ith solids and water, respectively, $D_{ij}$ is the size of $i$th material in $j$th size class, $r$ is the radial location, and $\omega$ is the angular speed of the bowl. The authors also applied this model for two-component feed separation (Ghaffari and Farzanegan 2017b).

The velocity profile of the fluid as well as the solids determine the stratification in these concentrators, both of which use a rotating cone. Significant amount of fluid dynamic computation on the distribution profiles have been made. Pressure gradient in the radial direction was neglected by Bruint (1969) to simplify the Navier-Stokes equations. The velocity profiles were obtained using a complex function method. The resulting equations were much simpler and less complex radial, tangential and axial velocity profiles were obtained by solving them. The authors reported that the simplification did not lead to any significant loss in accuracy. Makarytchev et al. (1997) subsequently pointed out the errors in the derivation and suggested modified expressions for the velocity profiles which could be successfully used in fluid dynamic computations. The physical process inside a spinning cone was subsequently explained in details by the same group (Langrish et al. 2003). They considered the flowing film to consist of two layers, namely, a laminar sublayer at the bottom and a wavy layer at the top. The combined film thickness and the average radial velocity were given by:

$$\delta^+ = 0.91 \eta^{-2/3} + 1.95 \eta^{-3}$$  \hspace{0.5cm} (11)$$

$$u_{av} = \frac{0.28 \eta^{-4/3}}{1 + 2 \eta^{-7/3}}$$  \hspace{0.5cm} (12)$$

where, $\delta^+$ is the dimensionless total film thickness, $u_{av}$ is the average radial velocity, and $\eta$ is the dimensionless radial coordinate.

A fluid dynamic model of the falcon concentrator has been provided by Kroll-Rabotin et al. (2010). The authors considered the buoyancy force, pressure gradient effect, added mass force, and the drag force to describe the particle motion. Assuming that the flow has rotational symmetry and no slip between the particle and the fluid in the azimuthal direction, the force balance equations were solved in a fixed azimuthal plane as:

$$\left( \rho_p + \rho f C_m \right) \frac{d\bar{u}_{i,2D}}{dt} = \rho_f \left( 1 + C_m \right) \frac{D\bar{u}_{i,2D}}{Dt} + \left( \rho_p - \rho_f \right) \left( \frac{\bar{u}_{ij}^2}{r^2} \vec{r} + \vec{g} \right) - \frac{3}{4d_f} \rho_f C_d (\bar{u}_{i,2D} - \bar{u}_{ij,2D}) (\bar{u}_{i,2D} - \bar{u}_{ij,2D})$$  \hspace{0.5cm} (13)$$

where, $C_m$ is the added mass coefficient (taken as 0.5) and the subscript 2D indicates two-dimensional velocity in the $r-z$ plane and $C_d$ is the drag coefficient. The drag coefficient given by Clift et al. (1978) was employed by the authors:

$$C_d = \frac{24}{R_{ep}} \left( 1 + 0.15 R_{ep}^{0.687} \right)$$  \hspace{0.5cm} (14)$$

where, $R_{ep}$ is the particle Reynolds’s number.

The authors used a direct solution of the Navier-Stokes equation for incompressible fluid to arrive at the fluid velocity profile and solved Equation 13 numerically. An expression for the partition function, $C_p$, was derived as:

$$C_p = \min \left[ \frac{1}{9} \lambda (1 - 1.6 \phi) Q^{-1} (\rho_p - \rho_s) d_f^2 \mu^{-1} R_0^{-2 - a} \cos \frac{\beta}{2} L_{bowl,1}^{1+a} \right]$$  \hspace{0.5cm} (15)$$

where, $\lambda$ is the calibration constant, $\phi$ is the solids volume concentration, $R_0$ is the base radius, $\beta$ is the opening angle of the bowl, and $L_{bowl}$ is the bowl length. For a particle travel length of $L$, $a$ is defined as:

$$a = \ln \left( \frac{1 + \frac{1}{2} \sin \frac{\beta}{2}} {\ln \frac{1}{R_0}} \right)$$  \hspace{0.5cm} (16)$$

The model was validated using a laboratory falcon concentrator (Kroll-Rabotin et al. 2012). Through a systematic experimental program the authors were able to calibrate the model which could be used to make quantitative predictions for the performance of the falcon concentrator.
4.1.2. Kelsey centrifugal jig (KCJ)

It utilizes the principles of a conventional jig along with the additional feature of employing centrifugal force field which enhances the recovery of fine particles by improving their settling characteristics. The feed slurry is distributed at the bottom of the central bowl which flows upwards over the surface of the ragging bed supported by a cylindrical screen. The screen spins coaxially with the rotor and pressurized water is introduced into the bowl behind the screen. Water pulsates through the ragging bed and stratifies the particles. The ones heavier than the ragging material pass through the ragging bed. The principles of differential acceleration, hindered settling and interstitial trickling hold as in conventional jigging. The differential acceleration rates are substantially enhanced by the higher apparent gravitational forces arising out of the rotation. The denser particles pass through the internal screen to underflow hutch and then through spigots to an underflow launder. The lighter particles are swept away by the rising flow and are discharged over a ragging retention ring into the overflow launder. Beniuk et al. (1994) provided a detailed description of the theory and development on the KCJ. Through laboratory and pilot scale trials, they showed that the KCJ worked well for tin recovery. It was established that the KCJ gave superior performance compared to that of the shaking table in terms of both concentrate grade and recovery.

Tucker (1995) developed an empirical model for the partitioning of the feed to the concentrate stream. The g-force, ragging characteristics and pulsation stroke length were considered to describe the partitioning along with the feed characteristics. The recovery was estimated as:

$$T_ij = R \cdot x_i^b \cdot (1 - x_i)^b$$

(17)

where $T_ij$ is the recovery in the concentrate, $x_i$ is the normalized particle size of the $i$th class and $j$ denotes the specific gravity (SG) class, $R$ is the parameter controlling overall recovery, $a$ and $b$ are the parameters that control the recovery curve. These two parameters define the size at which maximum recovery ($T_{max}$) occurs:

$$x_{max} = a/(a + b)$$

(18)

The author defines the size $x_{50}$ at which the recovery drops to 50% of the maximum. Thus,

$$T_{50}/T_{max} = (x_{50}/x_{max})^a \cdot ((1 - x_{50})/(1 - x_{max}))^b = 0.5$$

(19)

$$x_{max} = (x_{0_{max}}) \cdot \left(1 - \frac{g}{C1} + \frac{L}{C2}\right) - SG_j$$

(20)

$$x_{50} = (x_{0_{50}}) \cdot \left(1 - \frac{g}{C1}\right)$$

(21)

where, $g$ is the g-force, $L$ is the stroke length, $SG_j$ is the specific gravity of the $j$th density class, $x_{0_{max}}$, $c1$ and $c2$ are scaling constants.

The dependence of recovery on specific gravity is introduced through the term $R$ as:

$$R = \left(\frac{1}{T_{max}}\right) \cdot \tanh(P1 (SG_j/SG_{ragging})^{P2})$$

(22)

where, $P1$ and $P2$ are lumped parameters that include all other influences not covered explicitly.

The concentrate assays are then computed by multiplying the feed assay with the recovery value of the corresponding size and density class. The model was validated for wolframite tailings treatment operation.

Singh and Das (2013) showed that particle momentum played an important role with regard to the performance of the KCJ in coal cleaning. They described the passage of the particles through the screen by identifying the particle settling characteristics considering the forces acting on the particle in centrifugal sedimentation. They estimated particle settling velocity by using the slip velocity model proposed by Concha and Almendra (1979):

$$v_s = \frac{20.52 \mu_f}{d \rho_b} f_1 \left(1 + 0.0921 \left[\frac{d^3 (\rho_s - \rho_f) \phi_s \omega^2}{0.75 \rho_f^2}\right] \frac{1}{2} \right)^{1/2} - 1$$

(23)

with,

$$f_1 = \frac{(1 - \phi)^2 (1 + 0.75 \phi^{1/3})}{\left(1 - \phi + 1.2 \phi^{2/3}(1 - 1.45 \phi)^{1.85}\right)}$$

(24)

$$f_2 = \frac{(1 - \phi)^{1.83} (1 - 1.45 \phi)^{1.85}}{(1 - \phi)(1 + 0.75 \phi^{1/3})}$$

(25)

where, $\mu_j$ is the viscosity of fluid, $d$ is the particle size, $\rho_b$ is the bulk density of the solids, $\rho_s$ is the density of solid particles, $\rho_f$ is the density of fluid, $\omega$ is the angular velocity, $\mu_j$ is the viscosity of fluid and $\phi$ is the solids concentration.

They computed $f_1$ and $f_2$ for different solids concentration to estimate the slip velocity and the particle momentum is then calculated as:

$$m \cdot v_s = \frac{\pi d^2 \rho_s \cdot v_s}{6}$$

(26)

Through these computations it was shown that particles having intermediate momentum values (intermediate density) played the decisive role in determining the process performance. In the stratified particle bed, the particles with highest momentum remained close to the ragging bed and reported to the tailing stream. The lightest particles (having lowest momentum) remained farthest from the ragging bed and reported in the overflow stream. The particles with intermediate density values were the ones which determined the grade and the mass yield of the products as the process conditions changed. If the momentum of these intermediate density particles were sufficiently high to penetrate the ragging bed, they would pass through the screen and would reduce the density of the tailings stream reducing the clean coal yield but enhancing its grade. It was also shown that bed porosity and particle momentum have contrasting impact on the process performance of the KCJ.
4.1.3. Multi-gravity separator
The MGS can be thought of as a shaking table rolled into a cylinder. While the centrifugal force improves the settling rate, the oscillatory motion advances the particle bed. The principles of flowing film concentration hold as the fluid flow effects the separation of the lighter and the heavier layers of the bed. While a lot of work dealing with the application of MGS in gravity concentration of various minerals have been reported, literature on the hydrodynamic fundamentals of the MGS has not been reported. Most of the studies focused on experimental design and statistical analysis and optimization (Aslan 2007, 2008; Chaurasia and Suresh 2017a, 2017b). In view of these, plenty of empirical models applicable to specific cases have been reported.

Concentration of strontium mineral, celestite, was studied by Aslan (2007). The author studied the effect of process parameters on the recovery and grade of the concentrate through response surface methodology using an experimental design. Empirical models were developed for the grade and recovery of SrSO₄ in the concentrate with drum speed, tilt angle and shaking amplitude as the independent variables. Second order models were utilized in this method for the grade (%SrSO₄) and recovery of SrSO₄ for the specific ore and experimental set up. The optimum conditions for obtaining nearly 97% grade of the concentrate as well as for over 98% recovery were also identified.

Ozbakir et al. (2017) studied the processing of lignite coal preparation plant tailings by desliming the tailings in a hydrocyclone and treating the deslimed material in the multi-gravity separator. The study was aimed at recovering the combustibles from the tailings stream. The authors considered the drum rotation, wash water rate, drum inclination, and solid ratio as the major operating variables and developed regression models for the ash content and clean coal yield in terms of these process variables. The authors indicated that the original tailings stream (feed to the hydrocyclone) had an ash level of 54.82% with a lower calorific value of 2279 kcal/kg. After concentrating it using the MGS, the ash content of the clean coal stream was observed to be 24.21% with a base calorific value of 3226 kcal/kg. The authors claimed that the technology could be useful in recovering the combustible from plant tailings as well as address environmental issues.

Concentration of chromite ore was studied by Aslan (2008) using the MGS. The authors used Taguchi quality loss function to optimize the product grade and chromite recovery with drum speed, inclination, wash water, and shaking amplitude as the independent process variables. It was reported that the feed material with 29.12% Cr₂O₃ could be enriched to 47.74% Cr₂O₃ in the concentrate with a recovery of 73.71% under optimum conditions. The most favorable operating conditions were also identified.

4.1.4. Water-only cyclone
In the WOC, the conical region is small and has a large cone angle. A sudden drop in the available cross sectional area causes pulp velocities to rise and a huge turbulence is generated. This helps generate an autogenous heavy medium allowing only heavier particles to penetrate the medium and report to underflow while the inner upward moving vortex captures the lighter particles to be recovered in the overflow. The WOCs are mainly used in fines cleaning circuit. They can be the first stage of cleaning followed by concentration in the tables or spirals or a second stage cleaning by another WOC. WOCs are known for easy set up and operational simplicity. They eliminate the need for chemical addition as required in flotation and addition of heavy medium as required in a heavy medium cyclone. However, their separation efficiencies are generally low which limit their application. They have found most applications in fine coal cleaning. WOCs have reportedly been used to clean coal up to 50 mm top size. However, most commonly, they are used for cleaning −0.5 mm coal. A feed suspension with approximately 15% solids is fed to recover about 40% solids in the underflow reject. A WOC with a size of 4 inches can handle about 50 tph of coal (dry basis) while a 20 inches WOC can handle over 800 tph of fine coal. Often multiple WOCs are used in a cluster to handle high tonnages instead of a large one.

In view of its structural similarity with a hydrocyclone or a heavy medium cyclone, the flow patterns are quite similar. There are a lot of literature describing the separation and flow regimes in the latter two. However, there is a clear distinction in the separation principle of the WOC and a fundamental approach to quantify the separation efficiency have not really been reported in literature. Most of the work with quantification of separation features relied heavily on empirical modeling of the WOC. Some of these are discussed here.

Majumder and Barnwal (2011) studied processing of −0.5 mm coal fines in a lab model WOC. The authors studied three design variables, namely, vortex finder length, vortex finder diameter and cone angle. They came up with regression models for clean coal yield and ash values through investigations according to a factorial design of experiments. The authors indicated that the particle residence time decreases with increase in vortex finder diameter. Also, the vortex finder length determines the volumetric slurry split between the overflow and underflow to influence the clean coal yield and ash levels. It was also observed that the separation efficiency of −0.075 mm coal fines was very poor. Evidently, the WOC may not be appropriate for such fine material.

Kim and Klima (1998) studied the separation of low density material from its high density counterparts using a 25.4 mm diameter WOC. They used silica to represent the low density material and magnetite, ferrosilicon and galena particles to represent the high density particles. Fine particles with a nominal size of −150 + 25 μm were used in this study. Apart from the effect of process variables such as solids concentration and particle density, design variables such as cone angle and vortex finder length were also investigated. The recovery of heavy minerals in the +25 μm size range was estimated as:

\[
R_m = \frac{u_m}{(o_m O + u_m U)}
\]

where, \(O\) and \(U\) are the mass flow rates of the solids in the overflow and underflow streams, respectively, \(o\) and \(u\) are the mass fractions of +25 μm particles in the overflow and underflow streams, respectively, \(o_m\) and \(u_m\) are the mass fractions of...
heavy minerals in the +25 µm material in the overflow and underflow streams, respectively.

The rejection of quartz from the +25 µm material was estimated as:

\[ r_q = 1 - \left( \frac{u_q U}{(\alpha_q 0 + u_q u_U)} \right) \]  

(28)

where, \( \alpha_q \) and \( u_q \) are the mass fractions of quartz in the +25 µm material in the overflow and underflow streams, respectively.

The separation efficiency was estimated as:

\[ E = R_m + r_q - 1 \]  

(29)

They observed that best separation efficiency was achieved when the cone angle was between 45° and 90°. At a very low cone angle of 10°, the WOC acted like a size classifier sending all the magnetite to the underflow and rejecting only 5% of the quartz. On the contrary, with 180° cone angle, only 15% magnetite reported to the underflow with 90% rejection of quartz. A longer vortex finder led to more solids reporting to the overflow accompanied by a decrease in magnetite recovery and an increase in quartz rejection. The separation efficiency dropped slightly with an increase in the vortex finder length initially and then dropped significantly.

4.2. Fundamentals of fluidized bed counter flow gravity separators

Much of the fundamentals discussed for the centrifugal concentration are also valid for the counter-current gravity separation with the centrifugal acceleration substituted by the gravitational acceleration. Of course, in this type of counter flow separators the gravity force plays the dominant role and settling takes place in the vertical direction.

4.2.1. Floatex density separator and crossflow separator

These two separators operate on the same principles of liquid fluidization, hindered settling and artificial heavy medium generation. The only major difference between them is the feeding arrangement. While a feed distributor immersed in the top central region of the FDS, a feed well outside the main chamber is used in the CFS. The flow patterns are somewhat different in the two cases. However, since the main governing principles are the same, they could be discussed technically from the same perspective.

Particle dynamics is of paramount importance in defining the separation. The fluid and the autogenous heavy medium resist the motion of particles. Those having sufficient mass overcome the resistance and report to the underflow. Others are hydraulically transported to the overflow. Richardson and Zaki (1954) showed that the relative motion between the fluid and the particle, the slip velocity, can be expressed as:

\[ V_{ij} = U_{t,ij} \left( 1 - \varphi_j \right)^{n_j - 1} \]  

(30)

where, \( n_j \) is Richardson-Zaki index which is a function of Reynolds number, \( V_{ij} \) is particle slip velocity, \( U_{t,ij} \) is particle terminal settling velocity, \( \varphi_j \) is particle volume fraction, \( i \) and \( j \) denote the size and density fractions, respectively.

The Richardson and Zaki index has been correlated with Reynolds number by several investigators (Richardson and Zaki 1954; Garside and Al-Dibouni 1979; Rowe 1987). The most accurate description of the coefficient was perhaps given by Rowe (1987):

\[ n_j = \frac{2(2.35 + 0.175Re^{0.75}_t)}{(1 + 0.175Re^{0.75}_t)} \]  

(31)

Galvin et al. (1999b) showed that under fluidized condition the slip velocity of a particle can be described as a function of the dissipative pressure gradient, \( dP/dH \), as:

\[ V_{ij} = U_{t,ij} \left( 1 - \frac{dP}{dH \rho_{ij} \rho_f g} \right)^{n_j - 1} \]  

(32)

\[ \frac{dP}{dH} = \varphi_j \left( \rho_{ij} - \rho_f \right) g \]  

(33)

A more convenient form for the slip velocity is:

\[ V_{ij} = U_{t,ij} \left( \frac{\rho_{ij} - \rho_{sus}}{\rho_{ij} - \rho_f} \right)^{n_j - 1} \]  

(34)

where, \( \rho_{sus} \) is the suspension density. The motion of fine particles can be tracked using the above slip velocity correlation. Concentration of solid particles and hence, the suspension density, decreases with height from bottom to top. It is assumed that the effective suspension density is the average suspension density inside the separator, and it is expressed as:

\[ \rho_{sus} = \sum \varphi_j \rho_{ij} + \left( 1 - \sum \varphi_j \right) \rho_f \]  

(35)

Therefore, the average suspension density and the BP are correlated as:

\[ \rho_{sus} = \frac{P/H}{g} \]  

(36)

The interstitial water velocity is calculated by estimating the bed voidage, which is averaged over some property of particles. The volume average density of the particles may be used for voidage determination:

\[ \varepsilon_{avg} = \frac{\rho_{avg, particle} - \rho_{sus}}{\rho_{avg, particle} - \rho_{water}} \]  

(37)

where \( \rho_{avg,particle} \) is particle average density and \( \varepsilon_{avg} \) is average bed voidage.

The interstitial TW velocity is then computed by dividing the superficial TW velocity by the voidage. Terminal settling velocity is computed by considering a simple force balance under gravitation or centrifugal field. However, accurate prediction of the drag coefficient, which is strongly dependent on the settling regime, is a key issue. Reynolds number (\( Re_t \)) of a particle at its terminal settling velocity (\( U_{t,ij} \)) is given by:

\[ Re_t = \frac{\rho U_{t,ij} d}{\mu} \]  

(38)

Terminal Reynolds number was estimated by a number of researchers (Galvin et al. 1999a; Zigrang and Sylvester 1981). Hartman et al. (1989) proposed an accurate correlation for the determination of the Reynolds number as a function of Archimedes number that predicts terminal settling velocity...
within an estimated error of ±1%. The correlation proposed by them is as follows:

\[
\log_{10} Re_t = \left[ (c_1 A - c_2) A + c_3 (A - c_4) \right] + \log_{10} \left( c_5 + c_6 \sin (c_7 A - c_8) \right) \tag{39}
\]

where, \( A = \log_{10} Ar \) and \( Ar = \frac{\rho g d^2 (\rho_d - \rho)}{\mu^2} \), the latter is the dimensionless Archimedes number.

The above equation can be used to compute the terminal settling velocity since this correlation is independent of settling regime and gives a more accurate estimation of the terminal settling velocity. Using this description of terminal settling velocity, Sarkar et al. (2008b) studied the performance of the FDS. They concluded that separation in FDS is well-described in terms of the terminal settling velocity to account for the size and density simultaneously rather than the size or density alone. Below a threshold BP setting, the FDS acts as a classifier rather than a concentrator. However, a narrowly sized feed enhances the gravity separation performance because with the widely distributed feed, the size effect is dominant. Good control over the product grade and recovery is achieved using proper setting of operating variables.

Efficient control of the BP and TW rate opens up the avenue for utilization of the FDS as a concentrator or a classifier. Fine particles can be classified using the FDS at a low BP with low TW rate. Splitting the feed into different closely sized fractions and then treating them separately in FDS will reduce the size effect and enhance the separation performance of the FDS (Ozcan and Ergun 2017).

Das et al. (2009a) combined the tracking of particles with the component-wise and overall mass balances over the FDS, to offer a complete description of the separation process. Component-wise mass balance and the overall mass balance over the FDS are expressed by the following relationships:

\[
F_{ij} = U_{ij} + O_{ij} \tag{40}
\]

\[
F = U + O \tag{41}
\]

where, \( F \) is feed mass flow rate, \( U \) is underflow mass flow rate, \( O \) is overflow mass flow rate, \( F_{ij} \), \( U_{ij} \), and \( O_{ij} \) are mass compositions of the feed, underflow, and overflow, respectively.

The authors performed detailed simulation studies to predict the performance of the FDS. They observed that a closely sized feed enhances the concentration performance by reducing the size effect. Presence of ultrafines in the widely sized feed results in a poor performance with a high misplacement of the heavier particles in the overflow stream. Misplacement of the larger and lighter materials in the underflow was also discussed. Simulation studies established that any set of conditions is suitable for density separation of particles of certain size range while the other size classes hamper the overall separation performance under those conditions. There is an optimum level of the BP and TW rate for a given feed material. A very low BP results in an inadequate formation of the artificial heavy medium while too high a BP leads to the misplacement of small heavy particles in the overflow. The misplacement phenomenon in liquid fluidized bed was also investigated by Tripathy et al. (2017) using binary mixture of quartz and iron ore. It was found that the mean particle size ratio between the quartz and iron ore affected the misplacement significantly. They also concluded that at a critical value of this ratio bed inversion starts where the misplacement is maximum.

The slip velocity approach is most predominant in describing the separation mechanism in this type of teeter bed separators. The slip velocity model proposed by Galvin et al. (1999b) was used to predict the performance of the FDS by Kari et al. (2006) and Kapure et al. (2007). Kari et al. (2006) compared the applicability of Galvin’s slip velocity model with reference to the slip velocity model of Lockett and Al-Habbooby (1974) for predicting the separation performance of the FDS in chromite ore beneficiation. They concluded that both these models are good in predicting the performance at a low TW flow rate and a low suspension density. However, in all these studies, the effect of the BP on the separation performance was not considered. Galvin’s slip velocity correlation was compared with other slip velocity models proposed by Masliyah (1979), Patwardhan and Tien (1985), and Van der Wielen et al. (1996), which are used for the prediction of performance of FDS by Sarkar and Das (2010).

Zhang et al. (2016) made use of jet theory to compute the velocity profile in the teeter bed separator. The authors related the velocity profile with the axial height of the separator. They also observed that the velocity profile is dependent on the fractional open area. An expression for the bed height was derived as:

\[
H_{sb} = \exp \left( \frac{1.0164 - \varphi}{0.2089} \right) \cdot d_0 \tag{42}
\]

where, \( \varphi \) is the fractional open area and \( d_0 \) is the hole diameter.

**4.2.2. Reflux classifier**

Particle settling on an inclined plane is the key factor in this classifier. Research work on single inclined plane settling of
particles have been reported by MacTaggart et al. (1988) and Davis et al. (1989). Since the single inclined planes are difficult to accomplish industrially, the multiple parallel inclined planes form the basis of the industrial reflux classifier. The underlying principle for reflux classifier is the Boycott Effect. The sedimentation rate of solid particles under the action of gravity can be significantly enhanced if the walls of the settling vessel are inclined. This is demonstrated in Figure 17 in which the enhancement occurs according to the following:

\[
Enhancement \text{ in settling rate } = \left( \frac{H}{b} \right) \sin(\theta) \quad (43)
\]

where, \(H\) is the height of the settling domain, \(b\) is the width of the channel, and \(\theta\) is the angle of inclination with the vertical. The settling rate indeed can be enhanced by several orders of magnitude with proper design of the vessel and the walls.

Davis et al. (1989) proposed a comprehensive theory of classification of a dilute suspension into a fine and a coarse fraction using a single inclined plane. The authors developed expressions for the probability density function for the distribution of particle settling velocities in the overflow and underflow under steady-state conditions:

\[
P_v(v) = \frac{(Q_o - S(v))P_f(v)}{\int_0^{v_0} (Q_o - S(v))P_f(v)dv} \quad \text{for } v < v_0 \quad (44)
\]

\[
P_0(v) = 0 \quad \text{for } v \geq v_0 \quad (45)
\]

\[
P_o(v) = \frac{Q_oP_f(v) - P_o(v)\int_0^{v_0} (Q_o - S(v))P_f(v)dv}{Q_o - \int_0^{v_0} (Q_o - S(v))P_f(v)dv} \quad (46)
\]

where, \(Q_o\) and \(Q_f\) are feed and overflow volumetric flow rates, respectively, \(S(v)\) is the volumetric rate of production of clarified fluid due to particle sedimentation, \(v\) is the particle settling velocity, \(v_o\) is the cut off sedimentation velocity and is defined by \(S(v_o) = Q_o\). \(P_f(v)\), \(P_o(v)\), \(P_v(v)\) are probability density functions for the distribution of particle settling velocities in the feed, overflow, and underflow, respectively. \(P_f(v)\) is assumed to be known in this analysis, which is found from the size distribution of the feed particles and relating the size with Stokes law to obtain the settling velocity distribution. The steady-state particle classification was described by the authors with the above equations by plotting the cumulative size distributions of the feed, overflow, and underflow.

MacTaggart et al. (1988) focused on gravity concentration in inclined plane settlers. They used two different systems to investigate the concentration process. The first one was a mixture of polystyrene (PS) and polymethyl methacrylate (PMMA) beads having densities of 1.05 and 1.186 g/cm³, respectively, in an aqueous solution of sodium chloride having a density of 1.12 g/cm³. The other system consisted of a suspension of PS and glass (density 2.835 g/cm³) beads in aqueous sucrose solution having a density of 1.23 g/cm³.

They studied the Boycott Effect and Fingering Phenomena to see if the effects are additive. In a bidisperse suspension finger-like structures form due to lateral inhomogeneity at high particle concentrations. This is known as fingering phenomena which enhances the settling and rising rates of heavy and light particles in the suspension, respectively. They observed that both these effects significantly influence the sedimentation kinetics in the systems investigated by them. It was also found that an optimum combination of solids concentration in the suspension and angle of wall inclination gives the maximum settling enhancement by the two phenomena in an additive manner.

Nguyentranlam and Galvin (2001) reported a simple kinetic analysis of particles to arrive at the particle trajectories, cut point and throughput in the reflux classifier. They considered the slip velocity of the particle to be related to its terminal settling velocity and a slip hindered settling function:

\[
u' = u_s G(\phi) \quad (47)
\]

where, \(u_s\) is the particle terminal velocity and \(G(\phi)\) is the slip hindered settling function for a solid volume fraction of \(\phi\).

Assuming that creeping flow conditions exist at low dilution and Stokes’ law is valid, they used the slip hindered settling function proposed by Lockett and Al-Habbooby (1973):

\[
G(\phi) = (1 - \phi)^{n-1} \quad (48)
\]

where, \(n\) is the Richardson-Zaki exponent.

With the above formulations, they developed expressions for the transients of particle motion as:

\[
\frac{dy}{dt} = u_0 - u'_s(1 - \phi) = u_0 - u'(1 - \phi)\sin\theta \quad (49)
\]

\[
\frac{dx}{dt} = u'_s(1 - \phi) = u'(1 - \phi)\cos\theta \quad (50)
\]

where, \(\Theta\) is the inclination of the channel and \(u_0\) is the superficial fluid velocity.

For a suspension of particles with same density but different sizes, the particle trajectory was found from the above as:

\[
y = \frac{u_0 - u_s(1 - \phi)^n \sin\theta}{u_s(1 - \phi)^n \cos\theta} \quad (51)
\]

An expression for the critical terminal velocity was derived at which particles will have an equal chance of reporting to the mixing zone above or slide down. This critical terminal velocity could be used to estimate the cut size and is given as:

\[
u^* = \frac{u_0}{(1 - \phi)^n} \left( \frac{L\cos\theta}{k} + \sin\theta \right) \quad (52)
\]

where the critical trajectory is traced by the particle entering the domain at a coordinate \((0, 0)\) and leaving at the coordinate \((k, L)\).

The authors also formulated a throughput advantage factor, \(f\), defined as the ratio of fluidization rate in reflux classifier to the conventional fluidization rate. Thus,

\[
f = \frac{u_0}{u_s(1 - \phi)^n} = \frac{L\cos\theta}{k} + \sin\theta \quad (53)
\]

Through a set of experiments with glass spheres, it was demonstrated that a sharp separation performance can be achieved by the reflux classifier.
4.2.3. Air table

Fluidization is on a shallow bed in air tables. Density-based stratification takes place against the rising air current through a perforated inclined deck. The deck is also inclined sideways to assist in the separation of the heavy and light particles which is assisted by the deck vibration. The motion of particles is substantially complex. Chalavadi and Das (2015) modeled the separation features in the air table. They divided the separation of light and heavy particles into vertical stratification and horizontal (deck plane) segregation.

Through a force balance in the vertical direction they formulated the vertical (z direction) stratification as:

\[
H' = \left( \left( \frac{h}{d} \frac{k}{\varepsilon^2} - Fr \right) \frac{D}{\rho_p} + Fr \right) \frac{T^2}{2} + \frac{A \sin(\omega T)}{h} \left( 1 - \frac{\rho_f}{\rho_p} \right) \]

where, \( H' = z/h \) and \( h \) is the bed thickness. \( D \) is particle diameter, \( \varepsilon \) is the void fraction, \( \rho_f \) and \( \rho_p \) are fluid and particle densities, respectively. \( K \) is defined as:

\[
k = \frac{150(1 - \varepsilon)}{Re} + 1.75 \frac{1}{\varepsilon} \]

where, \( Re \) is the Reynolds number.

\[
Froude number = Fr = \frac{g_{eff} h}{U_0^2} \]

where, \( g_{eff} \) is effective gravitational acceleration and \( U_0 \) is the superficial velocity. \( T \) is the dimensionless time and is defined as

\[
T = t/(h/U_0) \]

where, \( t \) is the time. The angular frequency is denoted as \( \omega \) for the simple harmonic motion of the table and \( A \) is the amplitude of vibration. Thus, Equation 54 describes the position of the particle in the vertical direction of the bed.

The horizontal segregation was modeled by considering longitudinal and transverse motion of the particles. The relevant equations are described as (Chalavadi and Das 2015):

\[
d^2X/dT^2 = \sin a - (1 - H')^5 \times 0.5 \times (\cos a) - \xi \sin(\tau \times T) \sin(\varepsilon - \beta) \times \cos y \]

\[
d^2Y/dT^2 = \xi \sin(\tau \times T) \cos(\varepsilon) - (1 - H')^5 \times 0.5 \times (\cos a) - \xi \sin(\tau \times T) \sin(\varepsilon - \beta) \times \sin y \]

where, \( a \) is the table inclination, \( y \) is the side tilt, and \( \beta \) is the angle that the vibration force makes with the deck surface. \( X \) and \( Y \) are dimensionless positions on the deck surface in the x and y directions.

Using the above formulation, the separation was modeled. The motion of particles on the deck surface is depicted in Figure 18.

5. Comparative applied considerations

Advanced gravity separators primarily cater to the requirements of recovery of fine particles. Many of them can handle larger particles too. However, for a given particulate system they do not respond similarly in terms of separation efficiency and performance. Therefore, one needs to carefully evaluate the available options before making a decision. The selection criteria should include applicability regime, efficiency level, maintenance issues, product quality requirements, economic considerations, etc. Some of these are discussed in this section.

Both Knelson and falcon concentrators are capable of generating high g-forces and thereby ensure efficient
processing of fine particles. The falcon generates 300g while a Knelson generates at most around 180g of g-force. Laplante (1993) compared the effectiveness of the two concentrators with reference to fine gold particle recovery. The author observed that Knelson is very effective in the recovery of free gold over a wide particle size range. However, it is not so effective in the recovery of fine or flaky gold that is finer than 15 µm. Grinding circuit products are ideally suited for Knelson concentrator. The batch falcon concentrator is best used as a pre-concentration unit. However, the continuous falcon is versatile and can recover fine heavy particles having a low density from coarse gangue minerals (Laplante 1993).

Ancia et al. (1997) also reported a comparative study of the Knelson and the falcon concentrators. They used quartz-tungsten, quartz-ilmenite-tungsten, and quartz-galena mixtures to investigate the separation performance. While the specific gravity of the four minerals are largely different, the authors also used various particle sizes of these minerals to study the impact of particle size as well. They observed that some segregation of the heavy particles do take place in the bowl of the falcon concentrator due to fluidization of the light particles before the concentration step and this phenomenon aids in the recovery process. The higher recoveries (~100%) in falcon are attributed to this pre-segregation and the higher g-forces. However, the recoveries obtained in the Knelson also were very high and in the range of 90%-95%. The water counter-pressure was found to play an important role. When the counter pressure is low, the bed is fixed and heavy particles infiltrate through this bed to effect concentration. When the counter pressure is intermediate, concentration takes place by substitution. Under these circumstances, the heavy and light particles exchange positions. At very high counter pressure, concentration is accomplished by elutriation of light particles from the bed.

The Knelson concentrator was originally developed for fine gold particle recovery and a numerous investigations on gold recovery are available (Laplante 1993; Meza S et al., 1994; Huang 1996). It was also found to be useful in chromite beneficiation (Sen 2016) and fine coal recovery (Honaker and Das 2004; Honaker et al. 2005; Ush et al. 2012). Gold recovery using falcon concentrator was reported by several authors (Buonvino 1993; Laplante 1993) while its fine coal applications were reported by Honaker et al. (1996). Fine cassiterite recovery from the Rio Kemptville concentrator using the falcon was reported by Morley (1992) while the Knelson was observed to perform poorly for the Uljin tin ore (Angadi et al. 2017).

In case of ultrafines recovery the falcon concentrator is more suitable due to the higher g-force. If a purer product is required the Knelson should be operated at higher counter pressure. On the other hand, if higher yield of the concentrate is more desirable then a lower counter pressure is adequate. However, the concentrate grade will be poorer in the latter case. Both falcon and Knelson are capable of recovering valuables from fines although the performance of these are strongly dependent on the raw material characteristics.

The allflux separator has been successfully used in iron ore and coal beneficiation as well as chromite beneficiation, heavy mineral sand concentration, and slag classification. Studies on allflux separator was reported for the treatment of South African coal (Piennar 2000). A feed coal with 30.2% ash was enriched to 14.5% ash in the clean coal with a mass yield of 50%. The allflux separator was also employed in the beneficiation of iron ore (Williams et al. 2012). In this case, the classifier was used as a pre-concentrator. A hematite ore was classified into different fractions for further downstream processing. In another investigation, iron ore processing was carried out in allflux separator for particles finer than 1.4 mm (Horn and Wellsted 2011). A feed with 57.5% Fe was upgraded to 60.1% Fe at around 20% yield against a theoretical maximum of 40% yield in this investigation. The allflux classifier works efficiently in the pre-concentration stage where it classifies the feed for subsequent concentration stages.

The reflux classifier, FDS, and the allflux classifier can be used for size classification as well as density-based concentration processes. Allflux classifier has been successfully used in coal (Piennar 2000) and iron ore applications (Williams et al. 2012; Horn and Wellsted 2011). Application of floatex for fine particle processing has been reported in various fields such as iron ore (Sarkar et al., 2008a; Murthy and Basavaraj 2012), coal (Honaker and Mondal 1999; Drummond et al. 2002; Sarkar et al. 2008b; Bu et al. 2017), chromite (Kari et al. 2006; Kapure et al. 2007; Tripathy et al. 2013), electronic waste (Das et al.

---

**Figure 19.** Partition curves (underflow) based on terminal settling velocity for processing coal fines (a) T1 for low and T2 for high bed pressure (b) T5 for low and T6 for high teeter water flow rate (Sarkar et al. 2008b).
It was suggested that below a threshold BP, the floatex will act as a size classifier and above which it will act as a concentrator. A narrow size distribution would help eliminate the size effect in gravity-based concentration (Sarkar et al. 2008b). A terminal settling velocity-based partition curve was proposed to incorporate the size and density effect of the particulate mass. Figure 19 shows typical partition curves based on terminal settling velocity of coal particles in the FDS.

Figure 19a shows that a significant increase in the cut point terminal velocity \(u_{s50}\) can be accomplished at higher BP levels. This relates to an increase in the effective separation density and size. At low BP only very light and fine particles are separated since they have low terminal settling velocities. With an increase in BP, heavier and coarser particles also subjected to effective separation which can be attributed to their higher terminal settling velocities. Examination of such partition curves can lead to applied guidelines. Figure 19b shows how the \(u_{s50}\) changes with TW flow. Evidently the \(u_{s50}\) increases as the TW flow increases. This increase is attributed to the greater hydraulic transport by the rising water. Therefore, if a better grade of the product is required then a low TW is recommended and a high yield of the product is favored by a higher TW rate.

The floatex does not offer a very sharp separation. The Ep (Ecart Probable) values obtained in fine coal preparation are in the range of 0.16–0.25 under reasonably good operating conditions (Sah et al. 2017). The authors also indicated that cut point densities below 1.7 are difficult to obtain for \(-1.18 + 0.15\) mm coal particles. However, within these limitations, floatex can perform quite well under optimal conditions as revealed in Figure 20. It is evident from this figure that with a good understanding of the separation mechanism, the floatex could be operated under optimum conditions. Such operations may achieve a performance level close to what is dictated by the washability curve as the theoretical maximum.

Beneficiation of spiral classifier feed, \(-1.0\) mm iron ore fines, using floatex was reported by Sarkar et al. (2008a). Good rejection of gangue minerals (silica and alumina) was observed along with reasonably high recovery of iron values. Figure 21a shows a summary of test results reported by the authors. The spiral classifier feed was deslimed in a hydrocyclone and the underflow was fed to the floatex. The deslimed material had 61.9% Fe with 3.06% silica and 3.39% alumina. Iron recovery values in floatex were in the range of 61%–67% with Fe content in the heavy product varying between 64.6% and 66.3%. Alumina rejection varied from 60% to 72% while silica rejection was in the range 52%–71%. The deslimed material still had over 26% material in the \(-45\) \(\mu\)m size range. The authors attributed the iron loss to the misplacement of these fines in the underflow (heavy) stream. The misplacement of these fine materials was correlated with the moisture content in the underflow (Figure 21b).

It was also observed that the desliming operation resulted in 45% rejection of alumina and 49% rejection of silica with 71% iron value recovery. Recovery of iron values from stockpiles iron ore tailings was investigated by Ozcan and Celik (2016) using the teeter bed separator in various flowsheet configurations. About 69% iron recovery was achieved when teeter bed was followed by gravity concentration. However, nearly 71% iron recovery could be achieved when teeter bed was followed by magnetic separation. The teeter bed separators are indeed useful in a wide range of applications. Although not known to

![Figure 20](image-url) 

**Figure 20.** Experimental recovery of coal fines in the clean product in terms of grade and yield vis-à-vis the washability and the optimized process (Sah et al. 2017).

![Figure 21](image-url) 

**Figure 21.** (a) Iron recovery and gangue rejection in various tests (b) Ultrafine content as a function of underflow moisture (Sarkar et al. 2008a).
offer very sharp separation, the efficiency may be improved by ensuring a thicker underflow and thus minimizing the misplacement.

Laboratory and pilot-scale application of reflux classifier was reported by Galvin et al. (2010) with reference to beneficiation of relatively coarse coal, up to 8 mm. With −8 + 0.5 mm coal feed in the laboratory scale reflux classifier, yield values obtained at various product ash levels were close to what is indicated by the washability curve. Figure 22 shows that the partition curves were sharp indicating a good efficiency of separation.

Pilot scale studies with reflux classifier for nominally −4.0 mm coal at two different sites corroborated the laboratory finding that high efficiency of separation could be achieved. Figure 23a showed that an overall cut point of 1.55 could be achieved with an Ep value of 0.071. Partition curves for different particle sizes (Figure 23b) reveal that the cut point increased as the particle size got lower. Also, the efficiency of separation decreased with a decrease in particle size as indicated by the shape of the partition curves. However, the efficiency levels were still quite good.

A combination of reflux classifier and heavy medium cyclone can be complementary to each other, and is more efficient for mineral beneficiation such as coal cleaning. For example, −13 + 4 mm fraction can be treated in the heavy medium cyclone and the −4 + 0.5 mm fraction may be processed in the reflux classifier. A convenient empirical expression for segregation efficiency of the reflux classifier was developed by Laskovski et al. (2006) as shown in the following equation:

\[
\eta = \frac{1}{1 + 0.133 \cos \theta \frac{Re}{13} \left( \frac{L}{z} \right)}
\]

where \( \theta \) is the angle of channel inclination, \( Re \) is the particle terminal Reynolds number and \( (L/z) \) is the aspect ratio of the channel. It was shown that the empirical and experimental segregation efficiency values were very close. This segregation efficiency could be used to estimate the throughput advantage of the reflux classifier over a conventional fluidized bed. According to this relationship, at high particle Reynolds number, the separation of particles is independent of particle size. Therefore, under these conditions a purely density based separation can be achieved. Thus, this relationship assumes immense practical importance in terms of deciding the flow pattern for different separation regimes.

Orupold et al. (2014) reported application of reflux classifier (RCTM 2020) for fine coal cleaning in the range −2.0 + 0.025 mm. It was observed that the cut-point achieved for this target material was RD 1.60 with an Ep value of 0.07. Apart from coal, the reflux classifier also found applications in iron ore, chromite, manganese ore and mineral sand processing (Orupold et al. 2014). Test results on −1 + 0.106 mm Australian iron ore fines using RCTM 300 reflux classifier indicated that a heavy underflow stream with 60% iron can be generated along with less than 10% combined silicates/aluminates in it with a mass yield of 19%. The iron recovery and grades from this study are shown in Figure 24 against the heavy liquid separation data.

The reflux classifier (RCTM 300) was also tested on a Canadian iron ore for −0.106 mm and a spiral reject stream.
The fines (+0.106 mm) were enriched to 69.8% iron and 1.13% combined silica at 54.8% mass yield from a feed containing 42.6% iron and 30.7% silica. On the other hand, for the coarser feed (+0.106 mm spiral reject) the reflux classifier underflow contained 67.1% iron with 3.0% combined silica. The feed assayed 30.0% iron and 55.6% silica and the yield achieved was 39.5%. These data indicate that even fine iron ore can be successfully enriched by a reflux classifier.

Mineral sand testing using a reflux classifier also was found to be successful (Orupold et al. 2014). Using a RC™2000 unit beneficiation of two different head grades were investigated. The results are shown in Table 3. The data indicate that over 90% recovery of the heavy minerals (HM) can be achieved by the classifier. As the head grade becomes better the recovery also improves significantly. It is noteworthy that the reflux classifier is able to accomplish lower Ep values in comparison to other purely teeter bed-based separators. It can generate a high grade product efficiently. The improved efficiency is attributed to the enhanced sedimentation in the inclined channels.

The two major setbacks of a CFS are that it requires a narrow size distribution (maximum top to bottom size ratio 6:1) and a moderately large difference in the densities of the light and heavy minerals (Westerfield 2004). Within these constraints, the CFS is capable of performing efficiently. Application of CFS in fine coal processing was investigated by Westerfield (2004). On site pilot scale studies were carried out at a number of different locations. Results from one such site are shown in Figure 25. The feed to the crossflow was nominally −2.0 mm with an overall ash content of 13.83%.

The feed contained 12.23% material less than 0.015 mm and the ash content of this size class was 28.28%. Since the −0.015 mm material is expected to be treated by flotation and not by the crossflow, this fraction was taken out mathematically from the test results and are shown along with the as-tested results in Figure 25. Evidently, excellent recovery of combustibles was possible. The recovery was averaging over 95% at 6%–8% ash in the product (with desliming at 0.015 mm). The sulfur content of the product was about 1.75%. Efficient separation is attributed to the quiescent flow accomplished by the feed well outside the main body of the separator.

The Kelsey centrifugal jig has been successfully employed in the concentration of tin ore (Beniuk et al. 1994). The authors investigated the processing of the gravity concentrate and tailings of the existing circuit at Renison Limited using the Kelsey centrifugal jig. The pilot scale test results for the combined gravity concentrate is shown in Table 4. It may be seen from the table that even for a feed with as low as 18.7% Sn, a concentrate grade of over 50% could be accomplished at 72.3% recovery. The study was also extended to treat the existing gravity circuit tailings using the pilot scale Kelsey jig. In roughing operation for the tailings (1.62% Sn), a concentrate with 7.14% Sn was generated at over 52% recovery. The authors identified that the choice of screen size was quite critical as near size material was responsible for blinding the screen. Based upon the pilot scale results, the plant installed and commissioned commercial Kelsey jig (Model J650) for treatment of combined gravity concentrate.

Table 3. Beneficiation results in a reflux classifier for two different head grades of mineral sands (Orupold et al. 2014).

<table>
<thead>
<tr>
<th>Head grade</th>
<th>Feed (wt% HM)</th>
<th>Underflow (wt% HM)</th>
<th>Overflow (wt% HM)</th>
<th>HM recovery (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Low</td>
<td>6.6</td>
<td>44.4</td>
<td>0.8</td>
<td>92.0</td>
</tr>
<tr>
<td>Medium</td>
<td>22.0</td>
<td>53.6</td>
<td>1.0</td>
<td>96.5</td>
</tr>
</tbody>
</table>

Figure 24. Processing results of −1 + 0.106 mm iron ore fines with RC™300 reflux classifier against the heavy liquid separation data. (adapted and replotted from Orupold et al. 2014).

Figure 25. On site processing results of −2 mm coal fines with 9 × 16 inch pilot scale crossflow separator (Top) Combustible recovery (Bottom) sulfur recovery. (adapted and replotted from Westerfield 2004).

Wolframite is very friable and in the comminution process most of it ends up being in the slimes fraction; often below 25 µm. Thus, in spite of being heavy, much of it gets lost into...
the tailings. The tailings of yesteryears often contained as high as 0.3% WO$_3$. The Kelsey jig has been proven to be very useful in recovering the WO$_3$ (Clemente et al. 1993). Ragging parameters were shown to be of utmost importance. The authors were able to achieve a concentrate grade of 1.93% at 76.6% overall recovery. Interestingly, in the $-40 + 16$ µm range, the recovery of wolframite was around 92% while for the $-12$ µm fraction the recovery was only 39%. Evidently, while the ultrafine slimes ($-12$ µm) may form part of the feed, it cannot be effectively recovered by the Kelsey jig.

Efficacy of Kelsey jig in fine coal cleaning was studied by Singh and Das (2013). A mono-sized feed ($-300 + 150$ µm) was used to eliminate the effect of particle size in this study. The concentration processed was optimized with respect to bowl rotation, pulsation rate and ragging weight (depth). Under optimized conditions, the feed coal ash of 27.5% was reduced to 21.1% in the clean coal at 74% mass yield in single stage operation. The authors used response surface methodology to optimize performance for the target ash and the surfaces are shown in Figure 26.

The response surfaces (Figure 26) clearly indicate that a high yield of the clean coal can be achieved at high pulsation rate while a moderate pulsation favors low product ash. The bowl rotation defines the centrifugal force field and hence, the porosity of the bed which in turn affects the separation behavior. The ragging bed provides the obstruction to particle motion and a higher bed depth favors high clean coal yield albeit with a poorer grade. It was indicated that the Kelsey jig performance was comparable to typical flotation performances for fine coal. However, Kelsey jig may not be attractive in the ultrafine size range.

Multi-gravity separators (MGS) have been studied extensively for concentration of various minerals. Some of these studies are reviewed here that may provide guidelines for future processing of similar materials. Although, the MGS cannot generate a high g-force, the oscillatory motion coupled with the gravity and centrifugal forces become very effective in gravity concentration of various ores. The investigation on chromite ore processing by Aslan (2008) is one such work reviewed here. The author used a ground ($-150$ µm) ore containing 29.12% Cr$_2$O$_3$. Taguchi method was used for the design of experiments with the parameters of drum speed, tilt angle, wash water flow and shaking amplitude. The parameters were optimized to obtain the most favorable operational regime for the two dependent variables: yield and grade. Figure 27 summarizes the results of this work. It is seen that the peak grade of the concentrate is around 53% Cr$_2$O$_3$ while the peak recovery is a little over 55%. Using the experimental data further optimization was performed and under optimum conditions 47.74% grade of the concentrate at 73.71% recovery was achieved. It may be recalled that about 48% Cr$_2$O$_3$ level in the concentrate is considered reasonably good for ferrochrome production. However, the authors did not report the Cr/Fe ratio in the concentrate which is often considered critical.

**Figure 26.** The response surfaces for (a) yield (b) ash at optimum bed depth (Singh and Das 2013).
Due to stricter environmental norms, waste processing has picked up a lot of attention over the past few decades. Wastes of yesteryears not only present a serious environmental concern but also contain various valuables which have become good resources today. Removal or recovery of lead from lead-zinc mines waste is an example that has great application potential. The multi-gravity separator was employed to achieve this separation (Goktepe 2005). The material was taken from a site that contained flotation tailings and gravity (jig) tailings. The tailings mainly contained anglesite, hematite minerals while the gangue materials were quartz, calcite and gypsum. It also contained some zinc in the form of sphalerite and zincite. Metal content of the feed material was 3.68% Pb, 0.60% Zn, 0.14% Cu, and 11.05% Fe with 15.58% S. The investigation was aimed at finding the influence of tilt angle of the multi-gravity separator on the lead grade and lead recovery in the concentrate keeping all other parameters constant. Lead being the heaviest of the metals in the feed would respond sharply to a variation in the tilt angle as this determines the effective gravity force acting on the particles. The results are summarized in Figure 28 for various particle size ranges after grinding. The general trend of increasing lead assay and decreasing lead recovery with increasing tilt angle was observed in all cases. To summarize, the assay and recovery improved as the grind size was lowered from 500 µm to 106 µm. However, with further reduction in particle size no further improvement was observed. With 106 µm particle size, about 12% Pb content in the concentrate was achieved at 12% recovery when the tilt angle was 8°.

Emphasizing the waste processing aspect, recovery of combustibles from lignite tailings (Ozbakir et al. 2017) is also reviewed here. The −0.5 mm tailings sample contained 54.82% ash, 0.10% combustible sulfur, 0.48% total sulfur, 28.91% volatile matter and a calorific value of 2279 kCal/kg. The tailings was first classified in a hydrocyclone and then concentrated in multi-gravity separator. The investigators optimized both these unit operations for the tailings and came up with the best possible combination of process parameters. They proposed a flowsheet (Figure 29) for treating the lignite tailings to recover the combustibles. The essence of this study is summarized in this figure. It may be seen from Figure 29 that it is possible to process the lignite tailings with 54.82% ash to obtain 24.21% ash clean coal using the multi-gravity separator with significant enhancement of the calorific value. The clean coal yield was observed to be 36.16%. Upon decantation, the ash content came down to 21.84% with further increase in the calorific value. From the above discussion it is evident that the MGS does not generate high enough g-force and therefore, it is not capable of processing ultrafine particles. However, the combined effect of small centrifugal force and vibratory motion can accomplish efficient separation in various particulate systems of industrial relevance.

A water-only cyclone (WOC) is essentially a hydrocyclone with a wider cone angle and a longer vortex finder. The density effect is enhanced by the wider cone angle which facilitates particle build up at the bottom and generation of an autogenous heavy medium. On the other hand, the longer vortex finder reduces the vertical distance the larger low density particles have to travel in order to be engulfed by the rising overflow stream into the vortex finder. Before these particles achieve terminal settling velocity and migrate to the cyclone wall, they are trapped into the vortex finder. Thus, particles are separated based on differences in particle density rather than size. The WOCs have found applications in coal cleaning (Weyher and Lovell 1969), iron ore processing (Visman 1966) as well as gold recovery (Walsh 1985). Some
of these are reviewed here. WOCs are predominantly used as pre-concentrators to reduce the load on the final concentrations stages. However, for fine (−0.5 mm) coal cleaning the WOC may sometimes be employed for final cleaning depending on the clean coal ash requirements (Majumder and Barnwal 2011). The authors reported 53.68%–79.88% mass yield of clean coal for ash content range of 22.03%–26.89% while processing −0.5 mm feed coal (~36% feed ash) using WOC. Size wise ash reduction levels are shown in Table 5. Notably, the ash reduction was very poor under all experimental conditions for the −0.075 mm fraction. This is essentially due to the fact that the terminal velocities of this ultrafine material is very low even for the high density particles. Therefore, they achieve this velocity very quickly and before they could migrate toward the wall, they are trapped into the vortex finder regardless of their densities. As a result, the ash reduction is poor for this size fraction. It may be concluded that when using the WOC in the final cleaning stage, a desliming operation to remove the ultrafine fraction would be more effective.

Rejection of pyrite (heavier) from coal (lighter) was also effectively demonstrated for −6.3 mm feed material using the WOC (Weyher and Lovell 1969). The authors noted that the presence of fine pyrite (fine-heavy particle) in the feed was the most critical issue. This demanded a careful selection of the density cut-point for the WOC. While rejecting the fine pyrite into the underflow product, some loss of the coarse coal (coarse-light particle) will be incurred. However, fine coal recovery will not be affected by the selection and would remain high. The coarse size material reporting to the overflow stream will have very low ash content and will be of final product quality. The finer fraction of this stream may be treated by froth flotation to recover the combustibles. In view of the loss of coarse coal in the underflow stream of the WOC, it may be treated in a shaking table to recover the combustibles. These considerations led them to develop a flowsheet for processing the pyrite-bearing coal. In this flowsheet, the overflow of the WOC is sent to a sieve bend, the oversize of which is the clean coal. The undersize of the sieve bend is treated by flotation. The underflow of the WOC is concentrated in the Wilfley table. The tabling concentrate again passes through a sieve bend and the oversize becomes another final product stream.

Gold recovery from placer material could be an efficient separation process using the WOC since the densities of the valuable and the gangue are so widely different (Walsh 1985). While gold has a density of 19.3 g/cm³ the placer material will have about 2.6 g/cm³. The author tested a 101.6 mm WOC to process −4.7 mm run-of-pit placer material for the investigation. In this extensive study, the author systematically varied different parameters and quantified gold recovery. Through this research it was shown that recovering gold from placer material could be quite effective. It was concluded that the feed pulp density does not affect the recovery of gold while gold particle size, shape, and concentration ratio influences gold recovery significantly. Recovery of −0.425 + 0.3 mm gold particles is not influenced by the presence or absence of heavy minerals in the feed. The feed size does affect gold recovery. Smaller top size of the feed favors higher gold recovery while that of coarse gold increases when coarse heavy minerals are present in the feed. The lower the concentration ratio the

| Table 5. Ash reduction in % for various size fractions (Majumder and Barnwal 2011). |
|---|---|---|---|---|---|
| Expt. no. | −0.5 + 0.25 | −0.25 + 0.15 | −0.15 + 0.075 | −0.075 |
| 1 | 48.5 | 54.2 | 79.3 | 17.80 |
| 2 | 52.7 | 49.4 | 43.6 | 14.04 |
| 3 | 39.8 | 40.5 | 33.4 | 7.15 |
| 4 | 34.4 | 33.3 | 47.1 | 3.98 |
| 5 | 50.6 | 54.8 | 39.2 | 9.17 |
| 6 | 41.2 | 39.4 | 33.3 | 8.87 |
| 7 | 30.2 | 30.8 | 22.8 | 1.95 |
| 8 | 23.7 | 25.8 | 19.1 | 0.29 |
better is the coarse gold recovery while fine gold recovery is not much influenced by the concentration ratio.

The dry advanced gravity separations have been applied predominantly in coal preparation due to the low density of air. The air velocities required to fluidize the higher density metalliferous ores will be very high and may complicate the dust generation and collection issues along with loss of selectivity in separation. However, satisfactory performance has been achieved in coal preparation in many investigations and only two of them, one for coarse coal and the other for fine coal, are reviewed here.

Air dense-medium FBS is capable of treating large tonnage of coarse coal. Air fluidizes an externally supplied heavy medium through which heavier particles (shale) settle and are removed from the bottom. The lighter particles (coal) float above the heavy medium bed and are removed at the top. Investigation on the separation performance of the air dense-medium separator for $-50 + 6$ mm coal was carried out by Zhenfu and Qingru (2001). Two different conditions were investigated—a low separation density of 1440 kg/m$^3$ and a high separation density of 1760 kg/m$^3$. In the case of low separation density, clean coal at 41% mass yield with 18.21% ash could be obtained from a feed with 45.57% ash. On the other hand, for the high separation density case, the clean coal had 16.35% ash and a mass yield of 55.51% for a feed ash content of 39.11%. The product partition curves are shown in Figure 30. This figure indicates that good Ep values can be obtained in both cases which is evident from the sharp partition curves observed in both cases.

A low ash feed coal $(-50 + 6$ mm) with 21.48% ash was also studied (Zhenfu et al. 2002). Clean coal with 11.8% ash and the reject containing 85.75% ash could be obtained. As shown in Figure 31, excellent Ep values were obtained for all particle sizes. The overall Ep value was observed to be 0.03.

Relatively smaller sized coal $(-25 + 6$ mm) cleaning in air dense medium fluidized bed was also investigated (Sahu et al. 2011) in pilot-scale operation. The feed ash for this coal was 40.49%. The authors corrected the product partition curve for both the short circuit flows—to the sink as well as to the float product. The overall partition curve is shown in Figure 32. The shape of the partition curve is not so sharp and the authors ended up obtaining an Ep value of 0.12. The poor Ep value was attributed to the presence of large amount of near-gravity material. In this study the ash content in the clean coal could be reduced to 32%–35% with yield values ranging from 60% to 72%.

While good efficiency is achieved in coarse coal, dry fine coal processing is more difficult and is discussed here. Depending upon the feed coal characteristics air table can generate a product of variable quality with reasonably good performance for fine coal. The ultrafine fraction (<0.1 mm) should be removed while treating high ash fine coal in air table in order to achieve better performance (Chalavadi et al. 2015). A very high ash (49.2%) coal ground to $-1.0$ mm and deslimed at 0.1 mm was investigated in this study. It was shown that when 10% (absolute) reduction in the ash content was targeted, about 55% mass yield could be achieved (clean
coal ash ~39%). But further lower target of the clean coal ash resulted in sharp drop in the mass yield. For example, 35% ash in the clean coal resulted in only 33% mass yield (Figure 33).

Evidently, the air table fails to achieve the theoretical limit of yield as dictated by the washability curve for this high ash coal fines. It was concluded that in one stage, there is a limit of ash reduction if the yield has to be maintained at a reasonably high level. However, much greater ash reduction could still be possible with acceptable yield and recovery values if multiple stages of operation are employed.

Processing of low ash coal fines in air table was also reported (Chalavadi et al. 2016b). In this case no desliming was done to the −1.0 mm coal which had an ash content of 25.8% ash. The authors reported a mass yield of 60.6% for a clean coal ash of 17% under optimum conditions. The partition curve for the light product is shown in Figure 34. As can be seen from the relatively flat partition curve the separation efficiency is poor. In fact, the washability indicated a mass yield of 75% for 17% ash product. Consequently, the Ep value accomplished was 0.18 which is reasonable but certainly not good. However, this could be significantly lowered if the feed coal was deslimed at 0.1 mm.

The separation efficiency of the air table improved significantly for coarser coal (Patil and Parekh 2011). The studies were conducted with −6.3 + 3.35 mm coal (29.17% ash) and −3.35 + 1.4 mm coal (24.41% ash). Typical product grade-recovery curve obtained for the −6.3 + 3.35 mm coal is shown in Figure 35. It is evident from the figure that the product ash remains constant at around 7%–8% ash level with cumulative yield of about 70% indicating good separation performance. Overall, ash reduction from 27% in the feed to 10%–12% in the product could be achieved with clean coal yield ranging from 75% to 80%. Ash rejection of 77%-80% has been reported with nearly 95% combustible recovery. Thus, it is observed that the air table performed much better with relatively coarser coal particles.

While efficient dry separation of coarse coal (+3.35 mm) is readily accomplished, performances in fine coal cleaning (−1.0 mm) using dry operation are not so good. A narrow sized feed improves the efficiency but it is still not quite high. In view of the low efficiency, it may be concluded that the selectivity is lost significantly for fine coal in the −1.0 mm size range.

6. Summary and outlook

Mechanized mining of ores generates enormous amounts of fine particles. In addition, the comminution process contributes heavily to the fines generation. Therefore, the mining and mineral processing industry is faced with a challenge of treating these fines to recover the valuables. It is also undeniable that the reserves of rich ores have been dwindling steadily over the years. This fact led to the need for utilization of lean ores. More often, the lean ores require finer grinding sizes in order to achieve liberation of the valuables from the gangue.
Consequently, processes had to be developed for the treatment of these fines for values recovery. Secondary resources and tailings of yesteryears also became valuable resources in course of time. They also require processing at much finer size ranges to recover the valuables. In essence, the quantum of materials to be treated at a fine particle size range increased many-folds over time. Froth flotation is one of the widespread methods of treating fine particles. However, the increased load of fines may be difficult to be handled by flotation alone. Also, use of chemicals and consequent environmental issues are associated with flotation. Conventional gravity concentrators were unable to treat the fine particles efficiently which led to the development of the new genre of equipment known as advanced gravity concentrators. These concentrators have been hugely successful in various applications in mineral processing. They have undergone significant changes in design as well as operational features over the last few decades that resulted in improved efficiency indices. The need for a consolidated account of the developments was felt for a long time. In this review, the recent developments have been compiled and analyzed for the benefit of the mineral processing fraternity.

The developments have been reviewed from theoretical as well as applied perspectives. The advanced gravity concentrators performed remarkably well in processing various fine materials including wastes. Sometimes in a stand-alone mode and sometimes in combination with other unit operations they have been employed successfully in the recovery of valuables. In certain instances they have been used as pre-concentrators and in other instances they were used in the final stages of concentration. Sometimes, the final product was generated in a single stage operation while multiple stages of operation were required to generate the concentrate at times.

The separators based on centrifugal force field were particularly noteworthy. The Knelson and the falcon concentrators were quite complementary to each other. The falcon is able to treat even the ultrafines (<15 µm) due to the higher g-force. The Knelson is not so effective in that size range however the continuous units are quite versatile. With relatively narrow feed size distribution the difference in the densities between the valuable and the gangue need not be very high for these concentrators. The fluidization water is a useful tool for the plant engineer to shift operating regime from high grade product to high yield product. With limited g-force, the multi-gravity separators found wide applications in view of the coupled vibratory motion. Not able to achieve high throughputs is a setback for the MGS. Therefore, it may be employed in the final concentration stages where lower volume of material is treated. The Kelsey jig leverages the advantages of conventional jiggling in a centrifugal field. Also, the physical barrier due to the ragging material enforces a sharper cut and thus making the Kelsey jig suitable for final stages in the flowsheet. Water-only cyclone has limited scope outside coal although it has been put to good use in a few other situations.

The counter-current hindered settling-based separators have the added advantage that they can be used as classifiers as well. They usually do not offer a very low E_r but with a narrowly distributed feed they can achieve sharp separations. They give superior performance when operated in conjunction with a classification unit. The reflux classifier probably stands out with sharper separation due to the enhanced settling on the inclined channels. They can generate the final concentrate in a single stage operation. A more quiescent flow is the hallmark of the CFS that enhances the separation performance. Both crossflow and floatex demonstrated superior performance with a deslimed feed. In these cases, the heavy medium is autogenously developed requiring no external solids which is a great advantage.

From gold ore to coal preparation plant tailings and waste processing, the advanced gravity separators are able to recover the valuables quite efficiently. Industry personnel and researchers around the world have employed these separators successfully in almost all possible mineral processing applications. Variable efficiency levels were observed depending upon the nature of the raw material being treated. But that is true for any processing. The quality of the raw material will have a strong influence on the separation performance. Researchers have demonstrated the scope of optimizing the process and identify the most favorable operating regime to get the best possible results.

Wide range of literature reveals that there is industry-wide acceptance of different types of advanced gravity separators such as EGS, autogenous density separators, and dry gravity separators. However, there is need to carry out further detailed study (both experimental and as well as theoretical) to understand complex fluid-particle interactions which will enable further developments of the advanced gravity separators to improve performance for poor quality feed materials. Present literature review reveals that two main types of features directly influence the performance of any type of advanced gravity separators: (i) feed quality and (ii) process parameters. Feed quality includes size distribution, density distribution, and the presence of near gravity materials. On the other hand, process variable depends on the type of gravity separator. For example, in the case of EGS key performance enhancer is the g-force. On the other hand, for teeter bed separators, it is the density of separation (cur density) as dictated by the autogenous heavy media which is determined by the operating BP, fluidization water, etc. Accomplishing efficient separation calls for good understanding of the process features and stricter control of the process conditions.

Gravity-based concentration process is relatively inexpensive and has a hassle-free operation associated with it. In some instances there are moving parts involved which may call for resolution of some maintenance issues. However, the design and operation of these instruments are sturdy and stable and they have been able to survive the grueling test of prolonged plant operation. Considering all these and only part of the success story that could be reviewed in this article, the authors firmly believe that the advanced gravity concentrators are not only going to be there for a long time to come, but are likely to grow substantially. It is very likely that we will see the advent of more such new separators or advancements of the existing ones in the next few decades.

Disclosure statement

The authors report no conflicts of interest. The authors alone are responsible for the content and writing of the article.
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