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Particle Transport and Separation in Inclined Channels Subject to Centrifugal Forces

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Abstract

A novel experimental system was established in order to investigate the effect of centrifugal forces on the hydrodynamic transport of particles through planar channels inclined at 20° to the direction of the centrifugal force. The device, consisting of a fluidized bed housing attached to a system of parallel inclined channels, was mounted inside a 2 m diameter centrifuge. Fluidization water entered the base of the housing, flowing in a direction towards the centre of the centrifuge, through an open chamber 150 mm long, and then through the system of planar channels, 160 mm long, with perpendicular channel spacing of $z=1\text{mm}$. The water then emerged via an overflow weir, 400 mm from the centre of the centrifuge. Batch elutriation experiments were conducted on a silica feed, 0 to 0.060 mm in diameter, achieving separations defined in terms of the particle diameter, d , and centrifugal acceleration gG , where g is the magnitude of the acceleration due to gravity. The throughput advantage achieved using $G=73$ over a conventional fluidized bed with $G=1$ ranged from 845 to 3094. The extraordinary results were in reasonable agreement with the theoretical throughput advantage given by $Gz/(3d)$. It was concluded that the hydrodynamic benefits of the inclined channels and the centrifugal force multiplied. The centrifugal force produced further synergy with the system hydrodynamics, resulting in the suppression of the effects of particle size,

and hence a powerful separation on the basis of the particle density. This physical arrangement offers the potential to separate ultrafine particles in coal and mineral processing on the basis of density, offering an alternative to flotation.

Keywords: hydrodynamics; sedimentation; particle processing; fluidization; inclined sedimentation; gravity separation; enhanced gravity

Introduction

In Australia there has been a reduction in the average yield from mining of about 40% over the past 30 years (Topp et al, 2008), reflecting a gradual decline in the quality of mining assets. With lower grade deposits it has been increasingly necessary to liberate the valuable mineral species from the gangue through a process of breakage and grinding. There has also been increasing interest in mining old tailings to recover valuable minerals. Clearly, the future of mining requires an increasing focus on fine particle beneficiation. While traditional methods such as flotation will continue to be used, new technologies need to be developed to address the difficulties experienced in many applications, in achieving the required grade of product at the maximum possible recovery.

Gravity separation is arguably the oldest approach to achieving beneficiation in mineral processing, but has been limited to particles typically in excess of 0.1 mm in diameter. This lower particle size limit significantly restricts the application of gravity separation. Here, particles are fed to spiral separators, jigs, or fluidized beds. Separations involving ultrafine particles less than 0.1 mm become physically constrained by the relatively low terminal

velocity, u_t , of the particles involved, and hence these particles are either discarded, or alternative methods of separation applied. It is a goal of this work to extend the particle size range of gravity separation to include the so-called ultrafines.

Centrifugal devices, such as the Knelson or Falcon, have provided larger capacity solutions to fine particle beneficiation in the mining industry (Honaker et al, 2000). This class of device, referred to as “*Enhanced Gravity*”, subject a fluidized bed arrangement to large “*G forces*”, allowing high rates of processing to be achieved. The feed suspension is transported as a thin film up through a rotating bowl, with fluidization water opposing the enhanced weight force of the faster settling particles. The faster settling particles are retained and then allowed to discharge as an underflow stream, while the slower settling particles pass upwards and exit as an overflow stream.

Some centrifugal systems have been designed to take advantage of inclined settling (Boycott, 1920), using settling tubes tilted relative to the direction of the centrifugal force. These devices simply achieve a faster batch settling of the feed suspension, in effect a solid-liquid separation, rather than a classification of the particles according to either their size or density. Other centrifugal devices have incorporated multiple conical surfaces separated by a narrow spacing (Gupta, 1981). This arrangement, designed to promote laminar flow of the fluid, also produces a superficial flow velocity that increases in the radial direction towards the central overflow exit. The solids are conveyed entirely by the entering feed suspension, with no independent fluidization. The system is useful in classifying particle species smaller than a given size to the overflow, but fails to prevent those fine particle species from reporting with the coarser particles to the underflow exit stream. The prospects for upgrading dense minerals by desliming low grade particles from the denser product are poor, while efficient particle

size classification requires the recycling of the underflow. In this device the particles and fluid are free to rotate relative to the inclined conical surfaces. It is believed that the Coriolis force compromises much of the benefit that arises from the inclined geometry (Schaflinger, 1990). Greenspan and Ungarish (1985), however, showed that the introduction of a compartment wall can lead to some dependence on the container geometry, and in turn improve enhancement in sedimentation rates.

The present paper is concerned with the separations achieved using a new approach consisting of small scale Reflux Classifiers located inside a centrifuge. It is appropriate here to provide a brief description of the conventional Reflux Classifier, which operates in the normal gravitational field. The arrangement consists of a fluidized bed, with a system of parallel inclined channels located above, as shown in Figure 1. The system exploits the discovery by Boycott (1920) of increased rates of segregation when particles are exposed to inclined surfaces. Galvin et al (2009) applied closely spaced inclined channels with a gap of 1.77 mm to promote both a laminar flow and a high shear rate, and in turn inertial lift, selectively conveying the lower density particles to the overflow, while allowing the higher density particles to segregate and return to the fluidized bed (Galvin et al, 2009). Particles that segregated onto the inclined surfaces were either conveyed upwards towards the overflow exit or slid downwards, returning to the lower fluidized zone, discharging as underflow. The selectivity was promoted by the inertial lift of relatively coarse low density particles, resulting in remarkable separation performance over a wide size range (Galvin et al, 2010), surpassing that achieved using conventional fluidized beds (Honaker and Mondal, 1999).

Figure 2 shows the arrangement of the Reflux Classifiers within a centrifuge, as used in the present study. Each Reflux Classifier is aligned in a tilted radial direction, with fluidization water directed inwards through the fluidization zone, and then through to the system of inclined channels. The uniform geometry insures the suspended solids are fluidized at a fixed superficial velocity. Each Reflux Classifier operates independently thereby minimizing the effect of the Coriolis force (Schaflinger, 1987). Moreover, the overflow weir operates independently with respect to each channel, insuring a common channel velocity across the device. In the absence of this feature, variable channel velocities, and even full-flow reversible, can occur.

The objective of this study is to deliver a precise hydrodynamic condition, combining the full benefits of the centrifugal forces and hydrodynamic forces that arise in the inclined channels. Thus this study aims to show these two features multiply, producing a powerful separation. Moreover, the closely spaced inclined channels should promote a high shear rate and in turn inertial lift of relatively low density particles, resulting in a separation predominantly on the basis of the particle density. This effect is investigated for the first time in the present paper, using semi-batch elutriation to separate the particles on the basis of their size and also their density.

We already have a good understanding of the particle transport that occurs in the inclined channels of the Reflux Classifier subjected to the usual gravitational acceleration, $g=9.8 \text{ ms}^{-2}$ (Galvin and Liu, 2011). In principle, our existing model should also apply to separations conducted using a centrifugal force, with an effective acceleration of Gg , provided the effects of the Coriolis force are minimized. Thus, our aim here is to validate this model for the batch elutriation of particles, covering a broad range of G values. Our hypothesis is that the benefits

that arise from the combined application of centrifugal forces and inclined channels should multiply, making it possible to apply gravity separation down to exceedingly fine particles less than 0.010 mm. Centrifugal forces sufficient to produce a value of G up to 73 are used in this study. Although this value of G is not especially high, the value is more than sufficient to test the basic hypothesis, and in turn form the basis for a future study to cover even larger values of G . Thus this work is significant in providing the foundations for developing a separation technology suitable for solving a broad range of industrial problems.

2. Theoretical

Galvin and Liu (2011) developed a theoretical model to describe the tendency of a particle of a given diameter and density to convey up through an inclined channel. This model, which involved no adjustable parameters, produced very good agreement with the data of Galvin et al (2009). Here, very closely spaced inclined channels with a perpendicular spacing of 1.77 mm and length of 1 m were used, resulting in low flow Reynolds numbers, in fact laminar flow. With such a large channel aspect ratio, fully developed laminar flow profiles were assumed. The experiments involved very low solids concentrations, and hence hindered settling was neglected. This section reviews this work in the context of the present study concerned with the effect of centrifugal forces.

2.1 Elutriation Condition in an Inclined Channel

Figure 3 shows two particles located at the upward facing surface of an inclined channel, subjected to a parabolic velocity profile with superficial velocity, U' . The planar channel has a perpendicular spacing, z , inclined at an angle, $\theta=70^\circ$, to the vertical. In the absence of a

significant inertial lift force we can assume that a particle, of diameter d , will locate a distance $x=d/2$ from the upwards facing surface. According to Bird et al, (1976), the local fluid velocity, u , a distance x from the planar surface is,

$$u = \frac{6U'x}{z} \left(1 - \frac{x}{z}\right) \quad (1)$$

Assuming the upward lift force acting on the particle is less than the buoyant weight force of the particle, the particle resides at the planar surface with $x=d/2$. A simple elutriation condition based on a balance between the local fluid velocity and the particle terminal velocity in the tangent direction of the plane applies. Here it is assumed the contact friction is negligible, which is certainly reasonable in the presence of a shear induced lift force.

This elutriation condition suppresses the effects of particle size given that smaller particles, which have lower terminal velocities, u_{iT} , are exposed to proportionally lower local velocities, u . Assuming $x=d/2 \ll z$, the elutriation condition leads to,

$$u = u_{iT} = \frac{3U'd}{z} \quad (2)$$

It is noted that the Stokes' terminal velocity scales with the particle diameter squared, d^2 . According to equation 2, the superficial velocity, U' , then scales directly with the particle diameter, d^1 , hence the system geometry reduces the dependence on particle diameter by a full decade. In the Intermediate settling regime (Vance et al, 1965), $u_{iT} = kd$, where k is a constant dependent on the particle density. It follows that the superficial velocity

becomes, $U = kz/3$, which is independent of the particle diameter, d . This suppression of the effect of particle size led to enhanced separation performance, under continuous steady state conditions (Galvin et al, 2010).

The hydraulic capacity of the Reflux Classifier system shown in Figure 1 is defined using the “throughput advantage”, $F = U/u_t$ (Laskovski et al, 2006). A given superficial flow velocity, U , through the vertical zone causes particles to elutriate from the vessel, with the separation defined by a particle exhibiting an equal tendency to elutriate to the overflow or be retained. The relevant terminal velocity, u_t , is based on the density, ρ_s , and diameter, d , of such a particle. In a conventional fluidized bed, the throughput advantage, F , has a maximum value of 1, hence F provides a useful reference.

It follows directly from equation 2 that for laminar flow, achieved using closely spaced inclined channels, the throughput advantage is approximately,

$$F' = \frac{U'}{u_{tT}} \sim \frac{z}{3d} \quad (3)$$

where U' and u_{tT} are the superficial and terminal velocities in the direction of the inclined channels. By way of example, for a channel spacing, $z = 1\text{ mm}$, and separation diameter, $d = 0.01\text{ mm}$, the throughput advantage within the channel is, $F' = 33$. Across a system of multiple channels containing plates of finite thickness the effective throughput advantage is lower in value as the plate thickness reduces the available flow area.

It is noted that the throughput advantage described by equation 3 is fundamentally different to the relationship first provided by Ponder (1925) and Nakamura and Kuroda (1937), the so-called PNK model. Laskovski et al (2006) outlined in detail the departure that occurs from the PNK model as the aspect ratio of the inclined channels increases, and the role of the particle Reynolds number in influencing the asymptotic result in the limit of an infinite aspect ratio. The model used here applies to laminar flow through inclined channels of high aspect ratio, sufficient for developing the full laminar profile.

2.2 Inertial Lift

King and Leighton (1997) used a rotating Couette device to examine the inertial lift of a moving spherical particle in a shear field. Galvin and Liu (2011) used their criterion, incorporating conditions applicable to the Intermediate settling regime (Vance et al, 1965), with particles having a Reynolds number between 2 and 500, to produce an explicit expression for the lift force. That is,

$$L_f = 0.0567 \frac{\rho^{1.8} \gamma^{2.8} d^{5.6}}{\mu^{0.8}} \quad (4)$$

The shear rate, γ , obtained by differentiating equation (1) is given by,

$$\gamma = \frac{6U'}{z} \left(1 - \frac{2x}{z}\right) \quad (5)$$

The buoyant weight force of the particle, in the direction normal to the plane, is,

$$F_n = \Delta \rho g (\pi / 6) d^3 \cos(\theta) \quad (6)$$

When the lift force exceeds the buoyant weight force of the particle in the fluid, the particle should physically lift from the surface and convey along the channel.

Figure 4 shows the lift force fraction, L/F_n , given by the ratio of the lift force to the buoyant weight force normal to the plane, versus the channel spacing. The theoretical analysis applies to a particle of one diameter, $d=0.310$ mm, and density, $\rho_s=2600$ kg/m³. The central curve, based on a superficial velocity of $U^*=0.05$ m/s, is applicable to the conditions used in the study by Galvin et al (2009). This curve reaches a fractional lift force of 1.0 when the channel spacing decreases to 1.77 mm, corresponding to a flow Reynolds number of $\rho U^* 2z/\mu=177$. Note the flow Reynolds number depends on the hydraulic diameter of the channel, $2z$. It is evident the fractional lift force decreases to 0.01 when the channel spacing increases to 10 mm, and falls to 0.0001 for a channel spacing of 60 mm. However, at $U^*=0.1$ m/s, lift is achieved at a larger channel spacing of 4 mm, corresponding to a flow Reynolds number of 800. In general, a lift force ratio of 1.0 is maintained when the flow velocity and channel spacing both double, given this scaling preserves the value of the shear rate at the wall. This scaling corresponds to a four-fold increase in the Reynolds number. Thus with a further modest increase in the velocity and spacing the required laminar flow condition is soon lost, giving way to a new mechanism of particle transport through the inclined channels (Laskovski et al, 2006). Thus, for particles of silica in water, the laminar-shear separation mechanism is limited to closely spaced channels less than about 8 mm.

Figure 5 shows the lift force fraction for a particle of diameter $d=0.310$ mm, subject to a superficial fluid velocity of $U^*=0.05$ m/s versus the channel spacing. The theoretical data cover particles of density 1400, 2600, and 4450 kg/m³. A particle of density 1400 kg/m³ experiences a lift force fraction of 1.0 when the channel spacing $z=3.3$ mm. Here the lift force fraction for particles of density 2600 kg/m³ is 0.23, while for particles of density 4450 kg/m³ is 0.10, hence these denser particles show no tendency to lift and convey. Selectivity is promoted by inertial lift.

2.3 Effect of the Centrifugal Force

The present paper examines a very specific proposition. If a system of parallel inclined channels is subjected to a centrifugal force and hence a significant value of G , what is the overall throughput advantage? At a given location within the vessel the acceleration governing the terminal settling velocities of the particles will be higher than the gravitational acceleration by the factor G , and hence the overall throughput advantage for particles settling in the Stokes regime should be GF' . Thus a remarkably high throughput advantage should be achievable, beyond the limits of any existing fluidized bed technology. With $G=100$, and $F'=33$, the overall throughput advantage becomes 3300! The experimental demonstration of this result has not been reported previously, and is far from trivial when a large centrifuge must first be developed incorporating the necessary system geometry and arrangement for delivering the fluid and particles.

It is further noted that as the value of G increases, the inertial force, and hence the particle Reynolds number, increases. In turn, the settling regime of ultrafine particles less than 0.1 mm in diameter shifts from the Stokes' regime to the Intermediate settling regime. Thus, the

dependence of the terminal velocity on the particle diameter shifts from a d^2 to d^1 dependence. For this reason gravity separation, achieved using centrifugal devices in mineral processing, is referred to as “*Enhanced Gravity*”, given there is increased performance evident over a wider particle size range. In particular, particles which now settle in the Intermediate settling regime experience a local fluid velocity at the wall that increases in direct proportion to their diameter, while the terminal velocity also increases in direct proportion to their diameter. Thus, particles of vastly different diameter should respond to a common elutriation condition at the one superficial channel velocity (Galvin et al, 2009; Galvin and Liu, 2011). Hence, ultrafine particles that would normally settle with a d^2 dependence could, in principle, be separated on the basis of their density with a d^0 dependence.

The theoretical model was used to predict the effects of different centrifugal forces, as defined by the value of G , on the elutriation of particles of different diameter and density. The results are shown in Figure 6 for a fixed superficial fluid velocity. The elutriation condition, $u=U_{IT}$, and inertial lift of the relatively large, low density, particles contribute to the findings. At a value of $G=1$ the separation density varies more strongly with the particle diameter, increasing to relatively high values at particle diameters less than about 0.1 mm. The present study is limited to a maximum centrifugal force corresponding to $G=73$. Here it is evident the breakaway point occurs at much lower particles diameters of about 0.030 mm, while for $G=300$ the breakaway point occurs at a remarkably small particle diameter of 0.010 mm. These predictions, which show a reasonably constant separation density over a broad range of particle diameter, demonstrate the potential for gravity separation to cover down to a near zero size, and compete with flotation. The potential benefits of hindered settling are assessed in Figure 7 for $G=73$ using a fluid of density 1150 kg/m^3 . Here it is evident a more

uniform separation density is produced across the particle size range. The required superficial fluid velocity is lower, further reducing the breakaway size.

Experimental

Figure 3 shows a schematic representation of the planar channel device, inclined at 20° to the direction of the centrifugal force. A variable speed drive was installed to control the strength of the centrifugal force. The initial work was conducted at 177 rpm, corresponding to an acceleration field of about 137 ms^{-2} at the overflow weir, located a radial distance of 400 mm from the centre line of the vertical shaft. Hence the value of G was 14. Subsequent work was conducted with G equal to 28, 55, and 73. Semi-batch elutriation experiments were conducted using a continuous supply of fluidization water to convey the slower settling particles through the inclined channels to the overflow.

It was essential the internal system be mechanically balanced hence two rotating arms were attached to the shaft, each pointing in opposite directions. Identical inclined channel devices were attached to the rotating arms, with the inclined channels aligned at 70° to the vertical axis. In order to simplify the design, the fluidization zone was aligned at the same angle as the inclined channels. In principle eight large units could be incorporated within the centrifuge, however, for this first study it was decided to use much smaller vessels, and use two units on either side of the shaft to insure mechanical stability. The internal cross-sectional area was 40 mm x 40 mm for both the fluidization zone and inclined channel zone, while half of the flow area of the inclined channel section was occupied by the 1 mm thick plates. The perpendicular distance between the adjacent plate surfaces was 1mm. The fluidization zone was 150 mm long, while the length of the plates to the overflow weir was

160 mm. The plates were extended a further 40 mm to create independent overflow weirs and hence drive a uniform flow through each channel. In the absence of this provision flow communication between adjacent channels can occur producing reduced, or even negative, flows in one channel and higher flows in the adjoining channel. The provision is therefore essential. A box was created around the system to capture the water and particles discharged from the overflow weirs. The overflow was then discharged towards the floor of the centrifuge, preventing the material from being flung around the unit.

A double annulus, aligned with the vertical shaft, and located above the rotating arms, was used to supply the fluidization water at a fixed rate to the experimental system. This fluidization water was added to the outer annulus at a fixed rate, ranging from 2 to more than 30 L/min, depending on the conditions required. This flow was then split into the two units. A slurry of the fine particles was also added, but at a very much lower volumetric rate. A reflux action then developed inside the separators, with the coarser or denser particles driven back along the inclined channels to the fluidization zone, and the finer or less dense particles continuing on through the device towards the overflow exit.

At the start of an experiment the flow rate of water to the system was set and measured. Then the centrifuge was switched on and the rotational speed increased to the required level. The feed slurry was then suspended in a beaker and added very gradually to the outer annulus. The fluidization water was also discharged into the outer annulus at a point ahead of the slurry discharge in order to help wash the slurry into the device. Once all of the slurry had been added, the water fluidization addition was maintained for an extended period. This period was increased when experiments were conducted at relatively low flow rates. The objective was to insure that every particle capable of being elutriated from the device would

be. Thus the experiments were conducted in order to achieve equilibrium separations, similar to the approach used by Laskovski et al (2006). All of the solids elutriated from the device were collected from the base of the centrifuge. Additional water was used to wash the particles into a small tank adjacent to the centrifuge. The suspension was then pumped to a further tank and allowed to settle.

The final part of the experimental method was concerned with the recovery of the particles retained within the Reflux Classifier units, referred to as the underflow. Ultimately the goal will be to regulate the withdrawal of an underflow stream and achieve continuous steady state separations. However, here, the focus is on the batch elutriation of a system of particles. Once the elutriation phase was completed, the centrifuge was switched off, allowing the rotating arms to slow down and come to rest. The fluidization water addition was continued in order to wash the remaining particles from the device. The water addition rate was increased to insure all of the particles were removed. Again water was added directly to the lower section of the centrifuge to wash all of the particles from the unit. The suspended particles were pumped to a second tank and allowed to settle.

The particles collected in a given tank were allowed to settle and the water removed by careful decanting. The remaining slurry was removed and dried in order to obtain the mass of particles. Recovery of the underflow was always more effective than the recovery of the overflow, especially when the overflow contained significant portions less than 0.010 mm in diameter. Thus the mass split of the solids reporting to the underflow was often based on the underflow and feed mass values. For relatively fine feeds dominated by particles finer than 0.038 mm, the finest sieve size, a laser based particle size analyser was used to obtain the size distribution data. The partition curve, which denotes the probability of a particle of a given

diameter reporting to a given flow-stream, was determined using the size distribution and mass split data.

The partition curve data versus the particle diameter was assessed by fitting the Whitten equation (Rao et al, 2003). That is,

$$P = \frac{\exp(\alpha x) - 1}{\exp(\alpha x) + \exp(\alpha) - 2} \quad (7)$$

where P is the partition number based on the fraction of the particles of a given diameter retained in the system, $x = d/S_{50}$ is the normalised particle diameter, d is the particle diameter, S_{50} is the separation diameter with equal probability of either being retained or elutriated, and α is a measure of the sharpness of separation.

For experiments focussed on the separation of the particles on the basis of density the following function was fitted to the experimental data. That is,

$$P = \frac{1}{1 + \exp\left(-\frac{\ln 3(D_{50} - \rho_s)}{\rho E_p}\right)} \quad (8)$$

P is the partition number, which denotes the probability of a particle with a density, ρ_s , reporting to the overflow. The D_{50} is the density of the particle that has an equal tendency to report to the overflow or be retained, referred to as the separation density, and ρ the density

of the fluid. The E_p is the Ecart probable (Wills, and Napier-Munn, 2006), which denotes the error in the separation density. That is,

$$E_p = \frac{D_{75} - D_{25}}{2\rho} \quad (9)$$

where D_{75} and D_{25} are the densities of the particles with a 0.75 and 0.25 probability of reporting to the overflow respectively and ρ is the density of the fluid.

The silica particles used in the first series of experiments had a specification, 400G. The particle size distribution, which is shown in Table 1, covers the range from 0 to approximately 0.060 mm. In the initial program of experiments the ultrafine silica feed was elutriated using different combinations of elutriation velocities and centrifugal forces. Partition curves were used to quantify the separation diameter. The experimental results were compared to the data predicted using the theoretical model of Galvin and Liu (2011), replacing the acceleration due to gravity with the centrifugal acceleration in the calculation of the particle terminal velocity (Zygrang and Sylvester, 1981).

Experiments were also conducted using particles of fine coal of lower density. The particles covered the size range of 0.038 to 0.260 mm, and were analysed using a series of sieves. The coal was prepared as samples covering a series of narrow density fractions, firstly 1250 to 1300 kg/m³, 1300 to 1350 kg/m³, 1350 to 1400 kg/m³, 1400 to 1450 kg/m³, 1450 to 1500 kg/m³ and 1500 to 1600 kg/m³. The initial series of experiments were conducted using the 1300 to 1350 kg/m³ sample, elutriated at different flow rates. Then separate experiments were conducted using the density fractions, using a fixed fluidization flow rate in each experiment of 33 L/min to the two units. The size distributions of the feed, overflow, and portion retained

were determined, and the partition curve versus particle diameter produced. The data from all of the experiments were also combined in order to generate partition curves based around the particle density. These analyses allowed the variation in the separation density versus the particle size to be measured.

Results and Discussion

4.1 Influence of the Elutriation Condition (with $L_f < F_n$)

In the first series of experiments the objective was to assess the validity of the elutriation condition, described earlier, in which the local fluid velocity, one particle radius from the upward facing planar surface, is equated with the tangential component of the terminal velocity of the particle. A basic assumption here is that the particles have sufficient settling time within the inclined channel to reside at the upward facing inclined surface. The conditions applied in these experiments involved a relative lift force less than 1.0, across the full particle size range. The model predicts that a particle that meets the elutriation condition will have an equal probability of being conveyed to overflow or of being retained within the system, and hence its diameter will be the separation size, the S_{50} . A series of elutriation experiments was therefore conducted, with G equal to 14, 28, 55, and 73 at the overflow weir of the device. A single feed of silica, with density 2600 kg/m^3 , was used in the work and hence the feed size distribution was fixed as shown in Table 1. Given the system geometry was also fixed, with channel spacing, z , of 1 mm and length $L=160$ mm, there was a unique separation produced for each flow rate. The data generated from this work are reported in Table 2.

Figure 8 shows partition curves produced from the experiments using $G=28$ and flow rates of 4, 8.2, and 12 L/min to the two units. The experimental data were calculated directly using the experimental value of the mass fraction retained in the device, and the size distributions of the feed and of the retained particles. The curve fits, based on the Whitten equation, were reasonable, with some deviation evident at the finest of particle sizes. The partition curves shows that the probability of the relatively small particles being retained within the device is very low, while the probability of the larger particles being retained is about 1.0. The separation diameter, the S_{50} , which corresponds to a partition number of 0.50, was determined for each case.

Figure 9 shows the value of the S_{50} versus the mass fraction of the feed retained within the device, covering the full range of G values and flow rates. It is evident there is very good consistency between the different experiments. Figure 10 shows the variation in the value of α used in the Whitten equation with the particle Reynolds number. The particle Reynolds number used here is based on the tangent component of the particle terminal velocity, calculated using the applicable value of G and the value of the S_{50} . It is evident that the value of α rises with the particle Reynolds number, and hence the particle inertia, approaching a steady value of about 4.5.

Figure 11 shows the variation in the particle separation diameter versus the superficial fluid velocity within the inclined channels. Each experimental data series corresponds to a different value of G (14, 28, 55, and 73) and is compared with the theoretical model prediction. The model predicts a linear relationship. The terminal velocity is calculated using an acceleration field equal to Gg . The tangential component is $u_{iT}=u_t \sin \theta$ where $\theta=70^\circ$ is the angle between the alignment of the channels and the vertical shaft. The agreement at

relatively low values of $G=14$ and 28 is very good, while there is evidence of departure at larger values of $G=55$ and 73 . Figure 12 shows a correlation between the actual superficial channel velocity required to achieve a separation at a given particle diameter, d , versus the value of the superficial channel velocity predicted using the elutriation model. The actual velocities are on average 81% of the theoretical values, a reasonable level of agreement given that the value of G varies along the inclined channels. The value of G governing a separation is assumed to apply at the overflow weir of the separator. The value of the Reynolds number increases from about 40 to 575 over the full range of experimental conditions. The laminar flow profile is not fully developed at the larger values of the flow Reynolds number, and hence the local fluid velocity at the wall is higher than predicted for laminar flow.

The actual throughput factor, U^*/u_{Tt} , based on the ratio of the superficial velocity of the fluid in the channels, U^* , and the tangential terminal velocity, u_{Tt} , of the S_{50} particle at $G=1$, is shown in Figure 13. This throughput factor is shown versus the inverse of the particle separation diameter, d , expressed as $\frac{Gz}{3d}$. This quantity is the product of $z/(3d)$ obtained from equation 2 and the value of G . It is evident there is reasonable agreement between the values of U/u_t and $\frac{Gz}{3d}$ indicating that the system benefits from the inclined geometry and the centrifugal force field, with the two benefits multiplying. The data also suggest some divergence from parity as the throughput factor increases. This trend may be due to there being insufficient opportunity to fully establish the parabolic laminar flow profile within the channels. The divergence might also be due to the Coriolis force, quantified in terms of the increasing Taylor number, which scales directly with the angular velocity and particle diameter squared, and inversely with the kinematic viscosity (Ungarish, 1994).

It is evident from Figure 13 that with $G=73$ it is possible to achieve a throughput factor of about 845 to 3094 fold, an extraordinary result. A more conventional fluidized bed subjected to the same value of G would be limited to a throughput factor of, at best, only 73 with respect to that possible at $G=1$. It should be noted that in this system the plates that defined the channels were 1 mm thick, and hence half of the overall flow area was occupied by the plates. Thus the effective throughput factor is half of the level reported here. It is further noted that at higher solids concentrations, required for processing solids at a satisfactory rate, hindered settling would lead to a further reduction in the velocity, both for conventional fluidized beds and for this system. Nevertheless the results reported here suggest a high capacity separator could be produced using relatively low values of G , or that extraordinary separation should be achievable at high values of G . Industrial scale separations involving particles of diameter smaller than 0.001 mm would then be possible.

4.2 Influence of Inertial Forces

Initial experiments were conducted using particles of coal ranging in density from 1300 to 1350 kg/m³ (nominal density of 1325 kg/m³) and ranging in particle diameter from 0.038 to 0.260 mm. Each experiment was conducted at a different flow rate, and in each case the S_{50} value was obtained by interpolation or extrapolation using the Whitten equation. Figure 14 shows the particle size classification achieved in these experiments. Figure 15 shows the variation in the separation diameter with the velocity of the water through the channels for both the coal and the silica with $G=73$. The data shown for the silica are the same as that presented in Figure 11, with a linear relationship evident between the superficial velocity and the S_{50} separation diameter, despite the fact that Stokes' law applies, even at the high value of G . For the low density coal, there is a similar linear relationship however, the separation

diameter then increases rapidly with increases in the superficial channel velocity, with velocity scaling with $\sim d^{0.3}$. Thus, the system hydrodynamics has virtually eliminated all of the dependence on the particle diameter. This result is similar to the rapid increase in the separation size with superficial velocity reported by Galvin et al (2009) for $G=1$, which was later attributed to inertial lift by Galvin and Liu (2011). Thus, with $G=73$, it should be possible to generate a low density coal product covering the wide size range of 0 to 0.26 mm, incorporating very little contamination from the high density silica. At higher values of G it should be possible to reject even more of the silica from the coal product.

A second series of experiments was conducted using fine coal, each with a different specific (narrow) density range, with the particle size range from 0.038 to 0.260 mm. In every experiment, the elutriation was conducted using a flow rate of 33 L/min, supplied equally to the two devices. This flow rate corresponds to a channel velocity of 0.34 m/s. Figure 16 shows the partition curves, based on particle diameter, obtained from different experiments. It is evident that in each case the finest particles had a low partition number, and hence a low tendency to remain within the separator, while the coarser particles had a high tendency to be retained. Each curve, which corresponds to a different nominal particle density, provides a corresponding separation diameter referred to as the S_{50} . Thus the curves provide a basis for obtaining the relationship between the separation density and the particle size.

The data generated from this series of experiments were also used to generate the partition curves shown in Figure 17, based on the particle density. In gravity separation these density-based partition curves are applied to the known density distribution of the feed to predict the separation achievable from a given device. Here, each of the partition curves corresponds to a narrow particle size range, characterised by the mean diameter of the particles from that size

range. The partition curves are remarkably steep given the low solids concentrations, and hence the lack of an autogenous dense medium. The consistency in the data set is remarkable given that the data used to form a given curve were generated from six different experiments. On average the fluid elements spent less than 0.5 seconds in the channels, hence the speed of the separation was very high.

Figure 18 shows the variation in the separation particle density with the particle size for this channel velocity of 0.34 m/s and $G=73$. There is a gradual increase in the separation density as the particle diameter decreases. This quite limited dependence of the separation density on the particle size is highly desirable in gravity separation, where the objective is to separate particles on the basis of density. By contrast, the theoretical curve produced for a conventional fluidized bed at $G=1$ shows a very strong variation in the separation density with the particle diameter. A strong dependence is also evident for inclined channels at $G=1$. All data sets are applicable to a low solids concentration. It is clearly evident that the combination of the inclined channels and large value of $G=73$ reduces the variation in the separation density with particle diameter.

Figure 19 shows the variation in the Ecarte Probable, E_p , with the particle diameter for a channel velocity of 0.34 m/s and $G=73$. These data show that the separation performance is maintained over a wide particle size range with the $E_p \sim 0.07$, though there is evidence of increasing E_p as the particle diameter decreases to 0.050 mm.

Concluding Remarks

Centrifugal forces, with an acceleration field of Gg , ultimately produce a shift in the settling regime of ultrafine particles from the Stokes' to the Intermediate regime, making gravity separation an option for separating such particles on the basis of density. The dependence of the terminal settling velocity on particle diameter, d , changes by a full decade from d^2 to d^1 . Closely spaced inclined channels, which promote laminar flow, produce a further reduction in the dependence on particle diameter by up to a full decade, leading to a separation that becomes increasingly independent of the particle diameter. Thus a powerful synergy arises when the hydrodynamic benefits of centrifugal forces and inclined channels are combined. Some dependence on particle diameter clearly remains, within a modest range. A breakaway effect at exceedingly small particle diameters must of course exist, which shifts towards ever finer particle sizes as the value of G increases.

Ultrafine particles of silica up to 0.060 mm in diameter were fed to the system, with different fluidization superficial velocities applied, while using different values of G . In these experiments the superficial velocities were insufficient to produce physical particle lift, hence the particles were assumed to transport while in contact with the planar surfaces. Good agreement was obtained between the experimental data and the model of Galvin and Liu (2011) for $G=14$ and $G=28$, with less agreement for $G=55$ and $G=73$. In general the superficial fluidization velocities required to classify the silica at a given diameter were about 81% of the theoretical values. The throughput factor, U/u_{iT} , was found to be consistent with the theoretical value, $Gz/(3d)$, confirming the powerful hydrodynamic synergy between the effects of the centrifugal forces and the inclined channels. Although the maximum value of G used was 73 the corresponding through factors ranged from 845 to 3094. This level of

performance from a conventional centrifuge would require values of G more than 10 to 40 times higher.

The next series of experiments were conducted using particles of fine coal ranging in diameter from 0.038 to 0.260 mm. These particles, which were much lower in density, and coarser, were predicted to exhibit inertial lift at the superficial fluid velocities used. The variation in the separation density with the particle diameter was found to be very low across this size range, far lower than expected from a conventional fluidized bed operated at $G=1$ or even a Reflux Classifier operated with $G=1$. This work provided further evidence that the benefits from centrifugal forces and inclined channels multiply. It is anticipated that the application of this system on a continuous steady state basis will lead to hindered settling, and hence the need to operate at lower superficial velocities. The tendency of the lower density particles to remain suspended and convey should be further enhanced, while the tendency of higher density particles to be retained and in turn report as part of the underflow should be stronger, as these denser particles, which remain at the planar surface, are shielded from the effects of the hindered settling. Thus a more powerful gravity separation performance on the basis of particle density should be achieved.

The development of the full laminar flow profile within the channels is clearly an important consideration. Thus we intend to extend the batch elutriation studies, using much narrower channel spacing, and different channel lengths in order to better understand the potential effects of the boundary layer formation on the moderate departure from the model, observed in this study. Our future work will also focus on establishing the controlled release of particles to an underflow stream, allowing the system to operate on a continuous steady state basis.

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Figure 14 Size classification of low density coal, of nominal density 1325 kg/m^3 achieved using flow rates of 15, 25, 30, 33, and 35 L/min of water to the two units, subjected to $G=73$.

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Figure 16 Size partition curves for fine coal achieved using a flow rate of 33 L/min to the two units operated at $G=73$. From left to right the data corresponds to particles of density 1550, 1475, 1425, 1375, 1325, 1275 kg/m^3 . In the curve fits the corresponding S_{50} values are 0.052, 0.071, 0.095, 0.125, 0.152, and 0.170 mm, and α values are 1.7, 2.4, 2.9, 3.4, 4.0, and 4.5.

Figure 17 Density partition curves for fine coal achieved using a combined flow rate of 33 L/min to the two units operated at $G=73$. From left to right the data corresponds to particles of diameter of 0.178 mm, 0.116 mm, 0.075 mm, and 0.0489 mm. In the curve fits the corresponding D_{50} values are 1270, 1380, 1465, 1570, and the corresponding E_p values are 0.075, 0.073, 0.074, and 0.090.

Figure 18 Variation in the particle separation density with the particle diameter, produced at 33 L/min to the two units ($U^*=0.34 \text{ m/s}$), at $G=73$. The separation density exhibits relatively little variation with the particle size. The theoretical result for conventional fluidization with $U=u_t=0.00405 \text{ m/s}$ and $G=1$, however, shows a very strong variation in separation density with particle size. Using the theory of Galvin and Liu (2011), there is also strong dependence

evident for inclined channels at $U^*=0.0079$ m/s and $G=1$. All data sets are applicable to low solids concentrations, with no hindered settling.

Figure 19 Variation in the separation E_p with the particle diameter, produced at 33 L/min of flow to the two units, at $G=73$. The E_p exhibits very little variation with the particle diameter.

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Table 1 Size Distribution of the G400 feed.

Lower (mm)	Upper (mm)	Mass (%)
0.040	0.060	5.7
0.020	0.040	25.7
0.010	0.020	29.7
0.005	0.010	18.1
0.025	0.005	13.2
0.0	0.025	7.6

Table 2

Table 2 Experimental data obtained using the G400 silica particles subjected to elutriation experiments at different values of G and flow rates, Q . The superficial channel velocity, the mass split from the separation, the separation diameter, and sharpness index are listed.

G	Q (L/min)	U' (m/s)	Mass Split	Diameter (μm)	α
14	2.0	0.021	0.74	7.5	2.0
14	4.0	0.042	0.62	11.0	3.5
14	4.0	0.042	0.59	11.8	3.5
14	6.0	0.063	0.46	16.0	4.4
14	8.0	0.083	0.35	20.0	4.5
14	9.9	0.103	0.27	23.3	4.2
28	4.0	0.042	0.71	7.5	2.8
28	8.2	0.085	0.56	11.9	4.3
28	8.0	0.083	0.55	12.5	4.2
28	12.0	0.125	0.41	17.0	4.1
55	8.1	0.084	0.67	8.5	3.2
55	16.8	0.175	0.47	14.2	4.7
55	16.8	0.175	0.47	14.5	4.7
55	24.0	0.250	0.30	22.0	4.3
73	6.0	0.063	0.78	5.1	2.0
73	9.9	0.103	0.66	8.6	3.0
73	12.0	0.125	0.64	10.0	3.3
73	21.6	0.225	0.40	17.5	4.5
73	27.6	0.288	0.34	19.5	4.8