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Fluidized bed desliming in fine particle flotation

– Part III flotation of difficult to clean coal

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ABSTRACT

A novel flotation system was used to process fine coal feeds supplied from coal preparation plants. The system consisted of an inverted fluidized bed arranged above a system of inclined channels. High fluidization (wash water) fluxes were imposed through a distributor enclosing the free-surface, producing strong positive bias of up to 2.4 cm/s, ideal for desliming. High gas fluxes of up to 2.6 cm/s, in excess of the flooding condition, were also imposed. The presence of the inclined channels prevented the entrainment of gas bubbles into the tailings stream. This paper, which is the third in a series, examines, for the first time, the hydrodynamic performance of this system on two actual plant feeds, each known to be difficult to wash. The first feed was a poorly liberated coal with particle size $<260\ \mu\text{m}$ and 69% feed ash. The second was a well liberated coal with nominal size $<125\ \mu\text{m}$ and 83% less than $38\ \mu\text{m}$. The product ash was shown to decrease significantly with an increasing fluidization flux to gas flux ratio. The single stage flotation system demonstrated a performance capable of matching the Tree Flotation curve with some cases in fact surpassing this result.

Keywords: Coal, Fine particle processing, Flotation machines, Froth flotation

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1. INTRODUCTION

Froth flotation is used extensively in coal and mineral processing to recover and concentrate valuable fine particles. The valuable particles are made hydrophobic through the application of a collector, and their attachment and adhesion promoted using a high shear rate to effect both a collision, and then final adhesion with air bubbles. A frother is introduced to stabilize the air-water interface. The bubbles then rise up through the device to form a froth zone, allowing liquid drainage, releasing hydrophilic particles and some hydrophobic particles from the final froth product (Neethling and Cilliers, 2002). Additional water applied as drops at a sufficient rate from above the free-surface of the froth delivers a downwards positive bias flux, improving the product grade.

Conventional froth flotation suffers from incomplete recovery of the valuable particles due to (i) ineffective collisions between the ultrafine particles, typically below 10 microns, and the air bubbles (Miettinen *et al*, 2010), and (ii) the ease with which relatively coarse particles, several hundred microns, detach from the air bubbles (Goel and Jameson, 2012; Jameson, 2012). Bubble coalescence, which can arise for a multitude of reasons, can also lead to significant loss in product recovery. Moreover, the product retains fine hydrophilic gangue particles due to incomplete or non-uniform drainage and desliming of the froth product (Britan *et al*, 2009) or due to the entrainment of excessive liquid into the froth zone when a high gas flux is applied (Smith and Warren, 1989).

We have previously developed a fine particle gravity separator known as the Reflux Classifier (Galvin *et al*, 2009; Galvin *et al*, 2010; Galvin *et al*, 2012). This separator consists of a fluidized bed, with a system of parallel inclined channels mounted above. The technology is now deployed around the world in a range of fine coal and mineral processing

applications. Froth flotation can, in part, be thought of as a gravity separation process, in which rising air bubbles segregate from the liquid. We have therefore applied our recent advances in gravity separation to the field of flotation, in turn inverting the Reflux Classifier, to produce the Reflux Flotation Cell shown in Figure 1.

The Reflux Flotation Cell is fully enclosed at the top by a fluidized bed distributor, with a central port used to continuously discharge the bubbly overflow product. The inverted fluidization provides a basis for delivering uniform wash water, and hence counter-current washing of the product. A downcomer, consisting of a sleeved sparger, is used to introduce the feed particles and air bubbles at a high shear rate (Johnson and Gershey, 1991; Kracht *et al*, 2008), formed in the annulus around the sparger tube, thus providing efficient bubble particle interaction and collisions, and attachment and adhesion of the hydrophobic particles. The gas flux is run at levels that, ordinarily, would lead to flooding of the system. However, the incorporation of the system of parallel inclined channels below the main vertical cell prevents the transport of the gas bubbles to the tailings discharge at the base of the vessel. Thus, the upper system is permitted to flood and hence form spherical, extremely wet “foam” with a bubble volume fraction of order 0.5. In comparison, the volume fraction of bubbles in the pulp zone of a conventional cell typically lies within 0.02 to 0.33 (Gorain *et al*, 1995). Moreover, this new system is run in the absence of a froth zone, and thus there is relatively little tendency for coalescence. The system is thus more resilient to the effects of the (fluidization) wash water, providing a strong basis for driving efficient counter current washing of the bubbly mix, and hence efficient desliming.

This is the third in a series of papers concerned with the hydrodynamics of the Reflux Flotation Cell. In Part I (Dickinson and Galvin, 2013) the device was investigated using a

gas-liquid feed system, and then using feed slurry consisting of fine hydrophilic silica of particle size range between 5 and 22 μm . This work demonstrated the potential to operate at extreme gas flux levels, while maintaining a relatively low liquid split to overflow. For example, in the presence of a fluidization wash water flux of 1.1 cm/s it was possible to operate with a gas flux of 0.8 cm/s, generating a bubble surface flux of 144 $\text{m}^2/\text{m}^2\cdot\text{s}$ using bubbles 0.340 mm in diameter. Under these conditions the system achieved better than 97% desliming of the ultrafine silica, and better than 99% when the gas flux was lowered to 0.5 cm/s and fluidization flux increased to 2.7 cm/s.

In Part II (Galvin and Dickinson, 2013) the device was fed with a model flotation feed, consisting of equal portions of fine hydrophobic coal ($< 260 \mu\text{m}$) previously recovered by flotation, and ultrafine hydrophilic particles of silica ($<< 63 \mu\text{m}$). Again the system was shown to withstand extreme conditions in the level of the applied fluidization wash water flux and the imposed gas flux. A regime map was presented in order to illustrate the vast operating zone. At relatively low wash water fluxes the combustible recovery was close to 100% across the full range of conditions, and the product grade, defined by the ash %, was about 8.1%, significantly better than the 10.1% ash in the original flotation product. At an extreme wash water flux, and low gas flux, selective recovery of the coal particles was achieved, with the combustible recovery reduced to about 75%, while the product ash % was reduced to 4.0%. By comparison, the lowest product ash obtained from the Tree Flotation method was 5.4% with the combustible recovery at a low of 23.6%.

The present work was concerned with the application of this novel flotation device to actual industrial feeds. Clearly, the system performed remarkably well on very well defined feeds consisting of fine hydrophilic silica. However, it was still unclear whether this performance

could be translated to cover the presence of more realistic conditions involving ultrafine clays, and poorly liberated feed particles. Thus the feeds used in this study were selected because of the significant challenge they presented. The first feed contained very high levels of fine clays and minerals, with the overall feed ash at 69%, and 79% ash in the $-38\ \mu\text{m}$ particle size fraction. The fine coal, having a top size of $260\ \mu\text{m}$, also exhibited poor liberation. The second feed also contained very high levels of clays and minerals, with 55% ash. In this case the feed, which had a top size of $\sim 125\ \mu\text{m}$, was well liberated, however the portion of particles less than $38\ \mu\text{m}$ was in excess of 83%.

These feeds contained hydrophobic particles that were capable of attaching to air bubbles. Thus they were clearly floatable, and hence there was no fundamental issue associated with the system chemistry. This was most appropriate because the benefits of the Reflux Flotation Cell are fundamentally hydrodynamic in nature. However, these coals were deemed hard to wash due to the extreme levels of clays and mineral matter, and their ultrafine sizes. It was also known that conventional flotation had failed to deliver acceptable product ash levels.

2. THEORETICAL

It is important to realise there is a subtle but significant theoretical difference between applying wash water at the boundary of the foam compared to injecting the wash water just beneath the upper boundary (Dickinson and Galvin, 2013). Thus the boundary condition for wash water addition is a complex matter. Typical flotation practice involves the spraying of wash water from a distance above the free-surface of the foam. However, in the Reflux Flotation Cell fluidizing wash water is added via a distributor enclosing the top vertical section of the cell, and has previously been shown to function as if it were injected deep beneath the free-surface of the rising foam (Dickinson and Galvin, 2013). This permits

effective desliming by imparting the downward flow of the wash water, counter current to the rising foam, leading to positive bias.

Consider the steady state flow of bubbles and liquid in a vertical flotation column. We will define all system inputs and outputs as positive in value, and all vector quantities as positive in the upward direction, apart from the bias flux, j_b , which is, by convention, positive in the downward direction. The effect of hindered settling on the bubble velocity can be described by the equation given by Richardson and Zaki (1954),

$$V_s = V_t(1 - \theta)^n \quad (1)$$

where V_t is the terminal velocity of a rising bubble of a given diameter, d_b , θ the volume fraction of bubbles, and n a scaling constant. Using Eq. 1, together with one-dimensional Drift Flux theory (Wallis, 1969), Dickinson and Galvin (2013) derived the well-known relationship between the gas flux, j_g , and the volume fraction of bubbles, θ_b , in the foam rising above the level of water injection,

$$\frac{V_{sb}}{j_g} = \frac{(1 - \theta_b)}{n\theta_b^2} \quad (2)$$

and the liquid flux, j_f , rising up with the foam,

$$\frac{j_f}{j_g} = \frac{(1 - \theta_b)}{\theta_b} - \frac{(1 - \theta_b)}{\theta_b^2 n} \quad (3)$$

Using a simple flux balance, they derived an equation for describing the zone below the wash water injection point, where the bubble volume fraction is θ_w . That is,

$$j_g(1 - \theta_w) = V_{sw}\theta_w - j_b\theta_w \quad (4)$$

Here the bubble velocity, V_{sw} , is evaluated at the lower bubble volume fraction, θ_w . Further, the bias flux is $j_b = j_w - j_f$, where j_w is the wash water flux injected into the system and j_f the liquid flux associated with the foam, rising from the injection point.

Figure 2 shows the bubble volume fraction versus the imposed gas flux derived using the above equations with $V_t = 15$ cm/s and $n = 2.5$. Example calculations have previously been given by Galvin and Dickinson (2013). The bold continuous curve represents the case when $j_w = 0$. When $j_w = 0$, two conjugate volume fractions are present in the system for a given gas flux below the flooding condition, θ_b in the foam zone, and θ_{bl} in the dilute bubbly zone beneath the foam. Increasing the gas flux results in the system flooding, with a single volume fraction of bubbles throughout the system. Included in Figure 2 are two wash water curves derived using $j_w = 0.1$ cm/s and $j_w = 2.5$ cm/s. These two j_w values represent the minimum and maximum level of fluidization wash water used in this study. Also included is a dashed line from $j_g = 2.6$ cm/s, which was the higher of two gas fluxes used in this study. Two observations are evident. Firstly, the addition of wash water causes the flooding point to shift from right to left to a lower value in the imposed gas flux. Secondly, three conjugate values are now present in the system. Here the volume fraction of bubbles, θ_b , above the wash water inlet remains unchanged, the volume fraction of bubbles, θ_w , in the wetted foam below the wash water inlet is reduced, and the volume fraction of bubbles, θ_{wl} , in the dilute bubbly zone is increased. Clearly the operating parameters used in this study reach well beyond the flooding condition, which is avoided in conventional flotation processes by limiting the imposed gas flux to much lower values.

In conventional flotation, flooding occurs at a bubble surface flux, $S_b = 6j_g/d_b$, of about 100 m²/m².s (Wace *et al*, 1968; Yianatos and Henríquez, 2007), with industrial cells limited to about 30-60 m²/m².s (Fuerstenau *et al*, 2007; Massinaei *et al*, 2009). This lower flux ensures a clear interface forms between the bubbly zone and the foam above, which is essential for process control. The Reflux Flotation Cell permits the use of extreme levels in the bubble surface flux. The system of inclined channels below the vertical cell amplifies the segregation

of the bubbles from the liquid. Laskovski *et al* (2006) developed an empirical asymptotic relationship to describe this effect, expressed in terms of the ratio of the superficial fluid velocity, U , in the vertical section, to the terminal velocity of the particle, V_t , that just fails to be entrained through the inclined channels. That is

$$\frac{U}{V_t} = [\sin \emptyset] 7.5 \text{Re}_t^{-1/3} \quad (5)$$

where $\emptyset$ is the angle of channel inclination with respect to the horizontal and Re_t the bubble Reynolds number. Applied to the present work, Eq. 5 describes the increase in the bubble surface flux at the flooding condition. The increase is relatively low for bubbles larger than 1.0 mm in diameter, but increasingly significant as the bubble size decreases. Using Eq. 5 along with Drift Flux theory to calculate the gas flux and liquid overflow flux at the flooding condition with $n = 3$, Jiang *et al* (2013) predicted a maximum bubble surface flux of 518 $\text{m}^2/\text{m}^2.\text{s}$ for a bubble diameter, d_b , of 0.02 mm in the absence of any fluidization water. In a series of hydrodynamic experiments, with no fluidization water, they achieved a bubble surface flux of 600 $\text{m}^2/\text{m}^2.\text{s}$ using a mean bubble diameter of 0.73 mm. In the presence of significant fluidization water the bubble surface flux is limited to lower levels in the range of 100-200 $\text{m}^2/\text{m}^2.\text{s}$ (Dickinson and Galvin, 2013; Galvin and Dickinson, 2013). It is clear that the hydrodynamics of this system are very different to conventional flotation.

3. EXPERIMENTAL

The laboratory scale Reflux Flotation Cell used in this investigation is shown schematically in Figure 1. It featured seven 1 m long and 0.7 mm thick stainless steel plates set at an angle of inclination of 70° with respect to the horizontal, creating eight evenly spaced channels with a perpendicular gap of 9.3 mm. The vertical chamber, above the inclined channels, was 1 m in length, with a cross-sectional area of 72 cm^2 , excluding the area of the downcomer. All fluxes reported are based on this area. At the top of the vessel a truncated rectangular

pyramid formed a plenum chamber for fluidization water to enter the cell through a grid of 1.5 mm diameter holes via a four-way manifold. Located centrally to the fluidization distributor was the product discharge port, with the downcomer passing through the middle. Thus the overflow rose up through an annulus surrounding the downcomer before exiting through a discharge port. The sleeve and sparger downcomer, 300 mm long and 25.4 mm outer diameter, had a media grade of 10. A further 230 mm length of smooth, non-porous tube, sealed at both ends, was connected to the base of the sparger to form a 530 mm long, high shear mixing zone consisting of a 4.8 mm annulus gap. The outlet of the downcomer was located 0.33 m above the entrance of the inclined channels.

The two different metallurgical coal feeds were tested. The first, reported in Section 4.1, was poorly liberated and wet screened to a top size of 260 μm . The slurry was conditioned with diesel in a 120 L mixing tank for 5 minutes at a pulp density of 10 wt%. A secondary feed tank, 120 L in volume, containing the frother MIBC, was pumped in unison with the coal slurry to the downcomer to produce a combined volumetric flux of $0.70 \text{ cm}^3/\text{cm}^2.\text{s}$ and feed pulp density of 4.9 wt%. Sufficient mixing time between the two feed streams was achieved by pumping through a 3 m coiled hose before entering the system.

The fluidization wash water was prepared in a 700 L tank and dosed with MIBC to the same concentration as the feed. The tailings discharge rate was controlled using two pumps; one of high volumetric capacity for removal of the majority of the underflow, the other, a low rate pump to provide finer control over the volumetric split between the overflow and underflow to obtain steady state. The overflow product stream was open to atmosphere and discharged at a rate controlled by the tailings pumps; increasing the tailings pump decreased the overflow rate. The gas rate to the system was controlled using a rotameter.

Fifteen minutes of operation was sufficient to reach steady state, before overflow and underflow samples were simultaneously taken over several minutes. This sampling procedure was repeated to ensure the mass fractions were consistent between samples and the system was at steady state. A feed sample was then taken. Samples were wet screened to remove the -38 μm particles, and the remaining +38 μm particles were dry sieved. Sufficient settling time was allowed for the -38 μm particles to be collected in full. Experiments were conducted at a slurry temperature of $\sim 10^{\circ}\text{C}$.

The second coal feed, reported in Section 4.2, was well-liberated and finer in particle size. Here a similar procedure to that outlined above was followed with some variation. Firstly the feed, with a feed pulp density of 3.6 wt%, was used as received from the coal preparation plant. Thus, combined conditioning of diesel and MIBC was achieved in one mixing tank and then pumped directly to the vessel at a volumetric flux of $0.62 \text{ cm}^3/\text{cm}^2.\text{s}$. Secondly, only one underflow pump was used. The run time required to obtain steady state ranged from 8 to 14 minutes. Experiments were conducted at a slurry temperature of $\sim 20^{\circ}\text{C}$.

Samples of the two different feeds were sent for external Tree Flotation analysis. The Tree Flotation method produces an ultimate flotation response curve for a given floatable material through a series of successive bench scale flotation experiments (Pratten and Nicol, 1989; Nicol, 1994). Here, the initial feed slurry is re-floated over 3 minute intervals, using further collector and frother additions, until the mass of the concentrate and/or the tailings is estimated to be less than 5% of the initial mass of solids in the slurry suspension. This procedure is then repeated for each concentrate produced. No wash water is applied however an attempt has previously been made to incorporate wash water into the analysis (Jameson *et*

al, 2003). A comparable method, referred to as the Release Flotation analysis, has been found to provide more data in the lower ash region, whereas Tree Flotation produces more data points in the medium ash-higher yield region (Brown and Hall, 1999).

4. RESULTS AND DISCUSSION

4.1 Poorly liberated coal feed at high ash %

The feed coal used here was of very poor quality. Figure 3 shows the Tree Flotation curve obtained by an external, independent laboratory, in accordance with the Australian Standard, AS4156.2.2. Each point along the tree curve represents the cumulative combustible recovery, ordered by rank, of the seventeen concentrate and tails samples separated using the Tree Flotation method. The product ash rises at approximately 8% ash, reaching a combustible recovery in the vicinity of 35 to 45%, and then rises steadily to a combustible recovery of 100% at 68% ash. The results obtained using the Reflux Flotation Cell are also shown for a fixed gas flux of 2.6 cm/s and varying fluidization flux. Three scenarios are presented. The first is shallow bed fluidization (triangle symbols), in which the fluidization flux was relatively low. Here, the bed could be maintained at a fixed position (set above the outlet of the downcomer) without the need for the inclined channels. The second scenario involves channel fluidization (square symbols), in which a moderate to extreme level of wash water is used and the fluidized bed now enters the inclined channels under a high downward flow of liquid. For both of these scenarios the MIBC and diesel dosage was 20 ppm and 0.5 kg/t dry solids respectively. The third scenario is channel fluidization repeated with increased reagent dosages (circle symbols). Here the MIBC and diesel dosage was 40 ppm and 2.0 kg/t dry solids respectively.

In the extreme fluidization case involving the higher reagent dose, the Reflux Flotation Cell achieved an overall product ash of 4.9% at a combustible recovery of 40.6%, based on the assays. The corresponding yield, based on the assays was 13.5%, while the yield based on the actual mass values in the feed, product, and tailings was 11.2%. Hence, the raw data sets were very consistent. Indeed, the square of the relative error of the solids mass balance for each run closed well below 0.22% apart from one run, which closed at an acceptable 0.98%. Thus this separation is considered to be reliable. The detailed results obtained as a function of the particle size ranges are shown in Table 1. Here it is evident the product ash values in each of the size fractions are in the range 3 to 5% above 38 microns, and 7% ash in the -38 micron size range. These combine to produce the 4.9% ash product, well below the level of 8.1% at 12.9 % combustible recovery obtained using the Tree Flotation Method. This performance, beyond the Tree Flotation Curve, was also observed for the model flotation feed used by Galvin and Dickinson (2013).

Table 2 shows the recovery performance for the data set in Table 1. Clearly despite the extreme level of fluidization water applied, recovery was well maintained at an average value of 46% across all size fractions greater than 38 μm . Hence there was no tendency for the coarser size particles to preferentially detach from the bubble surface due to the extreme level of downward fluidization.

In general, as the fluidization wash water flux increased, the product ash decreased. This is observed in Figure 4. However, at the lower wash fluxes of between 0.1 and 0.5 cm/s, which are more typical of values used in conventional flotation, there was little, or no significant reduction in product ash, with values persisting around 20%. Increasing the fluidization to a moderate flux of 0.8 to 1.1 cm/s resulted in a significant drop in product ash to about 12%

ash, with the combustible recovery preserved in the range of 27 to 37%. Using an extreme level of wash water, of 2.1 cm/s, both the product ash and combustible recovery reduced to levels matching the Tree Flotation curve. This demonstrates the effective desliming and selective stripping of particles achieved by the downwards flow of fluidization water (Galvin and Dickinson, 2013). Increasing the dose of reagent is observed in Figure 5 to have shifted the combustible recoveries significantly higher.

The fluidization wash water flux was increased to 2.1 cm/s for a gas flux of 2.6 cm/s, levels much higher than used in conventional flotation. The bubble surface flux was in the order of $74 \text{ m}^2/\text{m}^2.\text{s}$ for a mean bubble diameter, $d_b = \Sigma d_i^3 / \Sigma d_i^2$, of 2.1 mm, measured when the feed MIBC concentration was 40 ppm. Note this bubble surface flux is less than half the bubble surface flux obtained in previous Reflux Flotation Cell studies due to the larger bubble sizes generated with this slurry. Furthermore, it is much lower than the extreme result of $600 \text{ m}^2/\text{m}^2.\text{s}$ obtained by Jiang *et al* (2013), while using a system of inclined channels, with no fluidization wash water. Thus, there remained significant potential for increasing the gas flux further.

4.2 Very fine well liberated coal feed at high ash %

Figure 6 shows the Tree Flotation Curve obtained for a second feed. In this case the combustible recovery rises sharply to the knee of the curve, at which point the curve rises almost linearly at a low rate. The sudden change in slope reflects a strong degree of liberation between coal of relatively low ash, and mineral matter of very high ash. The cluster of crosses corresponds to the combustible recovery versus product ash levels obtained by conventional flotation at the plant. These single stage flotation results failed to meet the requirements of the plant. The feed, which was 55% ash on average, consisted of 83% less

than 38 microns, and hence required extensive desliming. For each run, the square of the relative error of the solids mass balance closed well below 0.5% apart from two runs, which closed at 1.2% and 2%.

A total of six continuous steady state runs was performed using the Reflux Flotation Cell, operating the system as a single stage. All of the results obtained were found to be very close to the Tree Flotation Curve as shown in Figure 6. Interestingly, with the collector and frother level fixed at 2.0 kg diesel/t dry solids and 40 ppm MIBC, there was relatively little scatter in the results with respect to the Tree Flotation Curve. In practice a separation condition close to the knee of the curve should be targeted in order to maximize the combustible recovery while achieving an acceptable product ash %. Thus, although much of the data were well aligned with the Tree Flotation Curve, the combustible recoveries extended to low levels. Clearly, it would be important to ensure the recovery did not fall away to these low levels due to random disturbances.

Tables 3 and 4 detail the separation performance obtained when using a low fluidization flux of 0.7 cm/s and high gas flux of 2.6 cm/s. Despite the relatively low level of fluidization water applied, the high permeability of the fluidized bubbly bed clearly permitted a uniform downward flow of the wash water, with 98.8% of mass fraction reporting to the reject of particle size less than 38 μm , and with an ash of 87.1%. The overall recovery was 83.6% with the product ash at a respectable 9.8%. Note however, the combustible recovery was an impressive 98.6% for the particles greater than 38 μm , implying the loss in combustibles to tailings was in the <38 μm size fraction. This drop in recovery in the finest size fraction is thought to be a consequence of approaching the kinetic limit of the very finest of the particles (perhaps less than 5 μm) in the downcomer. This could be improved by increasing the shear

rate in the downcomer by increasing the feed rate and/or by narrowing the annulus width. Using a finer media grade frit to produce finer sized bubbles, and extending the length of the downcomer to increase the residence time is also recommended.

Figure 7A shows the product ash versus the fluidization wash water flux for both a relatively low and a high gas flux. The same data are presented in Figure 7B versus the ratio of the wash water flux to gas flux. This ratio is indicative of the bubble washing achieved. It is evident that there is strong continuity in the data, and hence the results obtained are governed strongly by the precise conditions applied. The conditions close to the knee of the Tree Flotation Curve correspond to a low wash water flux to gas flux ratio. Figure 8 shows the combustible recovery versus the ratio of the wash water flux to gas flux. Here there is much more scatter, reflecting the steep nature of the combustible recovery versus the product ash %, and the selective stripping of particles under different levels of fluidization water. In general, for a given level of fluidization water, combustible recovery was found to be improved across all size fractions when employing the higher gas flux of 2.6 cm/s.

It is evident that the Reflux Flotation Cell achieves remarkably strong desliming with the data consistently at the limit of the Tree Flotation Curve. This performance is achieved in a single flotation stage. In principle, the Tree Flotation Curve represents the limit of flotation performance. This suggests the technology could be effective in producing ultra-low ash coal, using a ball mill to grind and liberate the valuable particles, releasing the mineral matter, and then finally rejecting the mineral matter through the strong counter current washing.

5. CONCLUSIONS

Difficult to clean coals were fed to the Reflux Flotation Cell. These were readily floatable, but difficult to deslime. The separation performances achieved were remarkably close to the Tree Flotation Curve. In some cases the separation performance was in fact better than the result obtained using the Tree Flotation Curve. This effect, which was observed in a previous study (Galvin and Dickinson, 2013), is associated with the higher reagent doses and higher fluidization wash water fluxes. The intense hydrodynamics in the Reflux Flotation Cell appears to selectively strip the less hydrophobic particles from the bubbles.

6. ACKNOWLEDGEMENTS

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7. DISCLOSURE STATEMENT

The University of Newcastle holds international patents and patent applications on the Reflux Classifier and related technologies and has a Research and Development Agreement with FLSmidth Ludowici to develop these technologies.

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Table 1 Separation performance obtained using the high reagent dose and fluidization wash water flux. The feed ash was 68.4%, product ash 4.9%, and reject ash 78.3%. This result was better than the result obtained using the Tree Flotation method, which had a lowest ash of 8.1%.

Size Range (μm)	FEED				OVERFLOW (Product)				UNDERFLOW (Reject)			
	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %
+260+212	2.9	2.9	35.3	35.3	8.1	8.1	4.1	4.1	2.2	2.2	54.4	54.4
-212+150	7.0	9.9	46.9	43.5	14.9	23.0	3.0	3.4	5.9	8.1	62.4	60.2
-150+90	11.4	21.4	50.2	47.1	23.0	46.0	4.1	3.7	10.0	18.1	65.3	63.0
-90+63	8.7	30.1	54.1	49.1	13.3	59.3	4.4	3.9	7.4	25.5	65.6	63.8
-63+45	4.1	34.2	55.2	49.8	6.4	65.7	4.7	4.0	3.8	29.2	69.5	64.5
-45+38	3.1	37.3	58.0	50.5	4.2	69.9	4.9	4.0	2.5	31.8	69.1	64.9
-38	62.7	100.0	79.1	68.4	30.1	100.0	7.0	4.9	68.2	100.0	84.6	78.3

Table 2 Separation performance obtained in terms of the yield and combustible recovery of coal, using the high reagent dose and fluidization wash water flux. This data corresponds to the data given in Table 1.

Size Range (μm)	SEPARATION PERFORMANCE			
	Yield %	Cumulative Yield %	Combustible Recovery %	Cumulative Combustible Recovery %
+260+212	38.0	38.0	56.3	56.3
-212+150	26.1	29.4	47.7	50.3
-150+90	24.7	26.9	47.5	48.9
-90+63	18.8	24.5	39.1	46.3
-63+45	22.1	24.2	46.9	46.4
-45+38	17.3	23.6	39.1	45.8
-38	7.1	13.5	31.5	40.6

Table 3 Separation performance obtained for exceedingly fine and well liberated feed using a relatively low fluidization water flux of 0.7 cm/s and high gas flux of 2.6 cm/s. Note that for this run reject samples within the size ranges of +125 to +63 μm , and -63 to +38 μm , were combined to provide an ash value due to the small mass of sample in these size fractions.

Size Range (μm)	FEED				OVERFLOW (Product)				UNDERFLOW (Reject)			
	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %	Mass Fraction %	Cumulative Mass %	Ash %	Cumulative Ash %
+125	0.9	0.9	4.4	4.4	2.1	2.1	3.5	3.5	0.11	0.11	32.0	32.0
-125+90	2.4	3.3	4.4	4.4	5.6	7.8	4.4	4.2	0.11	0.22	32.0	32.0
-90+63	5.5	8.8	6.8	5.9	12.9	20.7	6.5	5.6	0.11	0.33	32.0	32.0
-63+45	4.8	13.6	11.6	7.9	11.9	32.6	9.3	7.0	0.33	0.65	71.1	51.6
-45+38	3.3	16.9	21.6	10.6	7.0	39.6	11.0	7.7	0.54	1.19	71.1	60.4
-38	83.1	100.0	62.6	53.8	60.4	100.0	11.2	9.8	98.8	100.0	87.1	86.8

Table 4 Separation performance obtained in terms of the yield and combustible recovery obtained for exceedingly fine and well liberated feed. The square of the relative error of the solids mass balance was 0.15%.

Size Range (μm)	SEPARATION PERFORMANCE			
	Yield %	Cumulative Yield %	Combustible Recovery %	Cumulative Combustible Recovery %
+125	96.8	96.8	97.8	97.8
-125+90	100.0	99.1	100.0	99.4
-90+63	98.8	98.9	99.1	99.2
-63+45	96.3	97.9	98.8	98.9
-45+38	82.4	94.5	93.5	97.6
-38	32.3	42.8	76.6	83.6

Fig. 1. Schematic representation of the Reflux Flotation Cell used in the study.

Fig. 2. Drift flux calculations showing the bubble volume fraction in a vertical flotation column versus the implied gas flux. Wash water injected into foam produces three conjugate volume fractions in three zones under non-flooding conditions. Here $V_t = 15$ cm/s and the hindered settling exponent is $n = 2.5$.

Fig. 3. Results obtained using a poorly liberated high ash feed. The continuous curve describes the Tree Flotation Curve. A broad range of reagent doses and fluidization wash water fluxes were used, resulting in the scatter. The best result was obtained using the high reagent dose and wash water flux. The gas flux was fixed at 2.6 cm/s.

Fig. 4. In general, the product ash decreased with increasing fluidization wash water flux using a poorly liberated coal feed. The gas flux was fixed at 2.6 cm/s.

Fig. 5. Combustible recovery versus the fluidization flux to gas flux ratio, using a fixed gas flux of 2.6 cm/s.

Fig. 6. Results obtained using the very fine feed at high ash %. The Tree Flotation curve is given by the curve through the small diamonds. Data from the Reflux Flotation Cell lie on the Tree Curve. Conventional flotation led to the scatter of crosses well off the curve.

Fig. 7A. The product ash correlated strongly with the fluidization wash water flux applied down through the system, and with the gas flux used. It is noted that a gas flux of 2.9 cm/s is considered to be approximately double the usual limit in flotation.

Fig. 7B. Product ash versus the fluidization to gas flux ratio. It is evident there is little scatter in the data, thus the conditions needed for achieving a given separation are very precise.

Fig. 8. Combustible recovery versus the fluidization to gas flux ratio. The scatter here reflects the steepness of the Tree Flotation Curve, and the selective stripping of the less hydrophobic particles from the bubble surface under varying levels of fluidization wash water.