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Maximizing Bubble Segregation at High Liquid Fluxes

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ABSTRACT

This study is concerned with a common class of problem involving two phase separation of a dispersed gas flow from a continuous liquid flow under extreme processing conditions. Relatively fine spherical bubbles of order 500 μm were generated in the presence of a surfactant under a high shear rate within a rectangular, multi-channelled, cuboidal downcomer. Liquid fluxes, as high as 176 cm/s through each channel of the downcomer, sheared bubbles from a sintered surface mounted flush to the channel wall before disengaging the downcomer flow into a vertical vessel. Both high feed fluxes, up to 15 cm/s, and high gas fluxes, up to 5.5 cm/s, ensured a high gas holdup beneath the downcomer and the hindered rising of the bubbles. Enhanced bubble-liquid segregation was achieved using an arrangement of parallel inclined channels incorporated below the main vertical chamber. This novel device, referred to as the Reflux Flotation Cell, prevented the entrainment of bubbles to the underflow, and significantly reduced the liquid flux to overflow, even in the absence of a conventional froth zone. Extreme upward bubble surface fluxes of up to 600 s^{-1} were achieved, while counter-current downward liquid fluxes reached 14.4 cm/s, arguably four times the bubble terminal rise velocity. Hence successful phase separation was achieved while operating well beyond the so-called flooding condition arising from extreme levels of gas and feed fluxes. This hydrodynamic arrangement should find application in increasing surfactant extraction rates in foam fractionation and ion flotation, gas absorption, and even particulate flotation.

Keywords: flotation, bubble surface flux, inclined channels, segregation, Reflux Flotation Cell

1. Introduction

This study investigates a generalised problem involving the two phase separation of finely sized gas bubbles dispersed in liquid, subjected to high gas and liquid fluxes. This hydrodynamic study is focussed on a gas-liquid system only. Although the findings of this work may have application in foam fractionation, ion flotation and froth flotation [1], or even in promoting efficient dissolution of fine gas bubbles in liquid [2], the issue of surfactant adsorption, gas absorption, or the kinetics of bubble-particle collisions remain outside the scope of this paper. Rather, our objective here is to achieve the maximum possible bubble surface flux at a given liquid feed flux, utilizing inclined channels to enhance bubble segregation.

The basic problem with all flotation processes is that it is not possible to attain suitable separation near the so-called flooding condition, and hence the allowed gas flux and feed flux is highly constrained. Consider the addition of non-coalescing bubbles rising in a quiescent fluid at a low gas flux in a vertical vessel. The population of bubbles experiences hindered rising, causing the bubble rise velocity to be less than the terminal velocity of a single isolated bubble. Upon reaching the upper surface of the liquid, a single layer of bubbles, or bubble raft is formed, preserved by a film of liquid at the bubble surface. The continual addition of bubbles results in a multi-layered bubble raft and the formation of rising foam. The liquid entrained into the foam tends to drain, resulting in the concentration of the surfactant and any particles attached to the bubble surface. Once the foam reaches the upper discharge, or overflow launder, a steady state develops, resulting in a reasonably constant gas hold-up in the foam and constant liquid flux to the overflow. An increase in gas flux leads to an even stronger hindered rising of the bubbles, thus increasing the gas hold-up in the liquid zone as well as decreasing the gas hold-up in the foam. Flooding is said to occur when the gas hold-up in the liquid zone equates to the gas hold-up in the foam zone. If a flow of liquid opposes the rise of the bubbles, such as the counter-current flow of liquid to the underflow, the bubble segregation relative to the vessel

declines even further. Hence the tendency to flood is directly related to the bubble segregation, and to poor separation performance.

In gas-liquid systems, the segregation rate can be the rate determining step of the system. Apart from the choice of surfactant and gas flux the only degree of freedom is the bubble size. Fine bubbles in particulate flotation can improve bubble-particle collision kinetics and flotation rates [3, 4, 5]. Fine bubbles also permit a higher bubble surface (area) flux for a given gas flux, critical for the high recovery of dissociated ions in an aqueous environment such as in foam fractionation or ion flotation [1]. However, the flooding condition is reached at a much lower gas flux using fine bubbles, and hence a higher bubble surface flux is not realised. Fine bubbles are also easily entrained into the tailings stream when a high feed flux is applied. Therefore, it is normal practice to use larger bubbles, typically 1 to 1.5 mm in diameter, in flotation processes. Plainly, there has been little scope for promoting more favourable hydrodynamics to accommodate a higher processing rate.

In the present study we investigate a novel flotation device, the Reflux Flotation Cell (RFC), which offers, for the first time, the potential to increase the bubble surface flux to levels well beyond the normal flooding condition, while effecting an efficient separation of the gas and liquid phases. This device, shown in Fig. 1, consists of (i) a multi-channelled downcomer designed to generate very fine bubbles under a very high shear rate, and (ii) a system of parallel inclined channels below the main device. The system is run at a high gas flux, which in turn leads to a high bubble surface flux. Once the bubbly plume emerges from the downcomer, the bubbly zone expands to fill the upper system, and expands into the system of inclined channels. The inclined channels lead to enhanced bubble-liquid segregation rates due to the Boycott effect [6], and thus the bubbly zone remains contained and the liquid reporting to the underflow free of bubbles.

Previous investigations using the RFC have been concerned with the desliming of gangue particles using fluidization wash water and a modest feed flux [7, 8, 9]. While imposing downward fluidization water and operating beyond the flooding point using relatively high gas fluxes, the liquid overflow rate leaving the RFC was shown to be maintained to an equivalent rate of a conventional system operating under non-flooding conditions. However, it is most uncertain to what extent bubble-liquid segregation is possible under extreme processing conditions. This study advances the feed fluxes tested by more than an order of magnitude, while achieving extreme bubble surface area fluxes, an order of magnitude beyond conventional flotation standards, and a three-fold increase over previous RFC studies.

The Reflux Flotation Cell has the potential to transform the hydrodynamics of flotation processes. Thus a systematic series of experiments was conducted, the sole purpose being to subject the system to the processing and transport of extreme levels of gas and liquid fluxes, and in turn assess its performance. The goal of this work was to ensure that all of the entering gas transported upwards as bubbles, transferring a relatively small fraction of the entering feed liquid, ideally less than 10%, to overflow. Clearly it makes sense to increase the gas flux in response to an increase in the feed flux in order to preserve the recovery of surfactant. Thus experiments were conducted using specific values of the feed to gas flux ratio. Focus has been given to factors governing the bubble size, the bubble surface flux transported upwards to the overflow, and the liquid flux reporting to the overflow. Qualitative comparisons with conventional flotation cells are therefore made in order to place this work into context.

2. Material and methods

The flotation experiments were performed using the laboratory scale Reflux Flotation Cell arrangement shown schematically in Fig. 1. The device consists of three parts, a downcomer

that incorporates two parallel vertical channels, containing sintered stainless steel plates, a vertical chamber under the downcomer, and a system of parallel inclined channels below the vertical chamber. The nominal pore size of the sparger material was 2 μm , though the system generated considerably larger bubbles. The gap size (width) of each channel in the downcomer was 4.5 mm. The height, length and width of the downcomer were 1140 mm, 127 mm, and 56 mm respectively. The vertical chamber was 960 mm high with a cross-sectional area given by 146 mm $\times$ 146 mm. The cross-sectional area of the vertical chamber, less the area of the downcomer, was 130.5 cm². This value was used in the flux calculations. The inclined channels were 925 mm long, and consisted of 6 parallel channels inclined 70° to the horizontal. The perpendicular channel spacing was 20 mm.

The feed liquid was retained in a 500 L tank. A Mono-pump was used to supply the feed to the top of the downcomer. Underflow was discharged via a vented pipe set to a fixed height of 0.44 m above the entrance to the inclined channels. Overflow was collected in a launder at the top of the vertical chamber, and then returned to the feed tank. Likewise, the underflow stream was returned to the feed tank, hence the system was operated in closed-loop. A ball valve was used to control the underflow rate leaving the system. The steady state underflow rate was set by slowly increasing the underflow rate until the bubbles entered approximately half-way to three-quarters the length of the inclined channels. Any further substantial increase in this rate would have run the risk of bubbles being transported down the inclined channels and discharging as part of the underflow. It is noted that a small portion, indeed a negligible amount, of the exceedingly fine bubbles, most likely finer than 50 μm , were discharged with the tailings flow. The air flow rate was controlled and monitored using a gas rotameter. The feed liquid was prepared using mains water and the reagent cetyltrimethyl ammonium bromide (CTAB, purity > 98%), manufactured by Ajax Finechem Pty Ltd. The concentration of the CTAB in the initial

feed solution was 0.01 g/L. Digital photographs were taken to manually measure and determine the mean bubble diameter, given by $D_b = \sum d_i^3 / \sum d_i^2$. Before sample collection, the system was allowed to run for at least 10 minutes to ensure steady state.

The series of experiments performed using the Reflux Flotation Cell were repeated using a vertical column, that was constructed by replacing the inclined channels with a vertical chamber (no channels) of the same vertical height. All other variables remained the same between the two column arrangements, including the use of the two-channelled downcomer and the extreme gas and feed fluxes. Hence, the vertical column was operated under conventional flooding conditions. A ball valve was used to control the underflow rate leaving the system. This permitted the liquid flux to the underflow to be maximized without entraining bubbles to the underflow. It is noted some exceedingly fine bubbles, arguably finer than 50 μm , were entrained into the underflow, however, their volume fraction in the underflow was almost negligible, and of little concern.

3. Theoretical

3.1 Use of drift flux theory to obtain the maximum bubble surface flux

Drift Flux theory provides a conceptually useful model linking the bubble volume fraction, ϕ_1 , within the bubbly liquid to the bubble volume fraction, ϕ_2 , within the foam. Fig. 2 shows a typical flux curve formed from a plot of the bubble volume flux, Ψ_s , versus the volume fraction, ϕ . The bubble volume flux is given by the product of the bubble velocity V_s and the corresponding volume fraction of the bubbles, ϕ . An operating line is applied to the flux curve, extending from the gas flux, j_g , through the flux curve at the volume fraction, ϕ_1 , and through the tangent at the volume fraction ϕ_2 , intersecting along the horizontal axis at the overflow discharge concentration, ϕ_3 . The slope of the operating line, V_p , is equal to the sum of the gas

flux, j_g , and liquid overflow flux, j_f . This model takes full account of the so-called hindered-settling effect, which slows the rate of bubble rise as the volume fraction of the bubbles increases. At low gas flux the steady state volume fraction of bubbles in the liquid, ϕ_1 , is relatively low, while the volume fraction of the bubbles in the foam, ϕ_2 , is very high, approaching 1. Hence there exists a clear interface between the two zones.

High rates of extraction require a high gas flux. However, as the gas flux increases the value of ϕ_1 increases, and the value of ϕ_2 decreases. This becomes obvious when a new operating line is drawn in Fig. 2. Ultimately, the interface between the bubbly liquid zone and the foam becomes “blurred”, leading to the flooding condition. At this point there is no difference between the volume fraction of the bubbles in the foam and in the bubbly liquid, hence it is no longer possible to produce a separation. The flooding condition is depicted in Fig. 3 with the operating line intersecting the inflection point of the flux curve as a tangent. Moreover, as the gas flux increases, the liquid flux, j_f , reporting to the overflow increases almost exponentially. This invariably means that a high proportion of the feed liquid reports to the overflow, entraining a concomitant level of unwanted material, either molecules or particles. Hence the targeted upgrade is lost in a flotation process. In particular, the gas bubbles will tend to entrain with the underflow, resulting in a loss of product.

Foam fractionation, ion flotation and froth flotation are highly constrained processes. Clearly, the bubble size is a critical parameter. Without loss of generality, we describe the bubble velocity using the Richardson and Zaki hindered settling model,

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

In Equation 1, V_t is the terminal rise velocity of an isolated bubble, n is a constant, and V_s is the bubble velocity relative to the container under batch conditions. Thus the bubble flux is,

$$\Psi_s = V_s \phi.$$

According to Drift Flux theory, based on the use of Equation 1, the flooding condition is given by [10],

$$\phi_f = \frac{2}{n+1}. \quad (2)$$

The graphical construction shown in Fig. 3 consists of propagation flux and a “settling” flux defined by the tangent [7],

$$j_g = |V_p| \phi + V_s \phi = (|V_p| + V_s) \phi. \quad (3)$$

In accordance with continuity considerations the propagation velocity, V_p , of a given concentration, or propagation wave is given by [11],

$$V_p = \frac{-\partial V_s \phi}{\partial \phi}. \quad (4)$$

Combing Equations (1), (3) and (4), the corresponding gas flux is given by,

$$\frac{V_s}{j_g} = \frac{1-\phi}{\phi^2 n}. \quad (5)$$

The bubble slip velocity is,

$$V_{slip} = \frac{V_s}{1-\phi} = \frac{j_g}{\phi} - \frac{j_f}{1-\phi}. \quad (6)$$

Combing Equations (5) and (6), the corresponding liquid flux, j_f , reporting to the overflow is given by [7],

$$\frac{j_f}{j_g} = \frac{(1-\phi)}{\phi} - \frac{(1-\phi)}{\phi^2 n} \quad (7)$$

Fig. 4 shows the bubble surface flux as a function of the bubble diameter, calculated at the flooding condition determined using Equation 2 by assuming a fixed value of $n = 3$. For bubble sizes ranging between 1 and 5000 μm , the gas flux and liquid flux at the flooding condition were determined using Equations 5 and 7 respectively. The empirical equation of Zigrang and Sylvester [12], applicable to rigid spheres, was used to obtain the bubble terminal velocity, V_t . That is,

$$Re_t = \left[\left(14.51 + 1.83 \frac{d_{50}^{1.5}}{\mu} (g\rho(\rho - \rho_s))^{0.5} \right)^{0.5} - 3.81 \right]^2 \quad (8)$$

where g is the acceleration due to gravity, d_{50} the bubble separation size, ρ the density of the fluid, ρ_s the density of the bubble and μ the fluid viscosity. The particle Reynolds number is given by $Re_t = \rho V_t d_{50} / \mu$. However, to avoid empirical error which arises at the finer bubble diameters, Stokes' law was used to calculate V_t for bubble diameters 10 μm and smaller. For larger sized bubbles, bubble deformation renders the rise velocity independent of further increases in diameter [13], thus for bubble diameters exceeding 2 mm the velocity was assumed to be constant at 0.20 m/s. It is evident the bubble surface flux is limited by the flooding condition to levels below about 118 s^{-1} .

3.2 Hydrodynamic enhancement of flotation

Laskovski *et al.* [14] investigated the capacity advantage that arises when a system of inclined channels is used to increase segregation rates of solid particles from liquid in the Reflux Classifier. They defined the capacity advantage as the ratio of the superficial liquid velocity through an inclined channel, U' , to the terminal velocity of the particle, V_t , that just fails to be entrained. They showed that for high aspect ratio channels,

$$\frac{U}{V_t} = [\sin \theta] 7.5 Re_t^{-1/3} \quad (9)$$

where $U = U' \sin \theta$ is the liquid superficial velocity in the vertical section. Here, the angle of channel inclination, θ , was 70° . The ratio, U/V_t , in Equation 9 was applied to the value of the gas flux, j_g , in Equation 5, and the liquid flux, given by Equation 7, in order to estimate the bubble surface flux, and the corresponding overflow liquid flux, achieved using the Reflux Flotation Cell. Fig. 4 shows the bubble surface flux as a function of the bubble diameter for the Reflux Flotation Cell. It is evident that the bubble surface flux greatly exceeds the level previously possible, especially for relatively fine bubbles. Furthermore, high gas fluxes and high feed fluxes should be achievable under the flooding condition using the Reflux Flotation Cell without greatly impinging on the level of liquid reporting to the overflow. This is particularly important in foam fractionation and ion flotation applications in which product enrichment is highly dependent on liquid entrainment.

It is noted that even more favourable predictions would be produced from Equation 9 if a smaller value of n was applied in Equation 1. At $n = 3$ flooding occurs at slightly lower bubble volume fraction than for larger n . However, the new system provides an additional degree of freedom, permitting much higher bubble volume fractions to be produced at the flooding point simply by increasing the underflow rate. Thus, we have assumed a constant volume fraction of 0.5 at the flooding condition to produce the upper curve shown in Fig. 4. This analysis suggests it may be possible to produce a bubble surface flux greater than 500 s^{-1} . This means the Reflux Flotation Cell may, for the first time, offer the potential to increase extraction rates by more than an order of magnitude under stable operating conditions.

4. Results and discussion

4.1 Bubble size

Three series of experiments were conducted using a fixed liquid feed to gas flux ratio, j_o/j_g , thus preserving the intrinsic state of the system within the downcomer, and within the flotation device. The feed to gas flux ratios examined were 2.3, 4.5, and 9.1. It is noted that due to the rapid mixing and flow velocities within the system, producing clear photographs of the bubbles in motion proved difficult. Hence, clear images of the bubbles in the dispersion beneath the outlet of the downcomer were obtained by simultaneously stopping the feed and gas flux. Photographs of the bubbly zone were taken immediately after switching the feed and gas flows off in order to obtain a representative mean bubble diameter and avoid the effects of coalescence. In the absence of these high fluxes, the bubbly zone formed a packed bed of bubbles, and hence the rise velocity rapidly decreased, in turn allowing good representative images to be obtained. Granted that smaller sized bubbles have lower rise velocities, and given those less than 100 μm were entrained the furthest into the inclined channels, it is likely the true portion of these bubbles in the system were absent from the photographs of the consolidating bed of bubbles. These finer sized bubbles were observed to be only a small portion of the overall population of bubbles. Hence, the reported diameters are most certainly a conservative overestimate of the mean bubble diameters, and the true mean diameters were slightly smaller. Fig. 5 shows side-by-side images of the consolidating bed of bubbles representative of each j_o/j_g ratio. The finer bubble diameters produced using $j_o/j_g = 9.1$ are evident in Fig. 5(a), which have a top size around 0.5 mm. The top size of the bubbles for the lower two j_o/j_g ratios was around 0.74 mm.

Figure 6 shows for each j_o/j_g value the effect of the liquid feed flux, j_o , on the mean bubble diameter. Note that these feed fluxes are an order of magnitude higher than typically used in flotation processes. It is evident that the bubble diameter decreased as the feed liquid flux

increased, a result indicative of the role of the shear rate in the narrow channels of the downcomer. The shear rates are shown in Fig. 7 to range from 1450 to 2350 s⁻¹. These values are based on the liquid flow alone through each channel, and assume Newtonian fluid behaviour. The shear rate, γ , through the rectangular channel was given by $\gamma = 6U'/z$, where z is the perpendicular spacing of the inclined channel [15]. The liquid flux passing through each channel in the downcomer ranged from 1.09 to 1.76 m/s, and the residence time of the bubble-liquid mixture was around 0.3 to 0.4 s. The liquid residence time in the system ranged between a low 16 to 26 s.

The results in Fig. 7 indicate that the shear rate produced from the flow of liquid was the most effective in minimizing the bubble size at the highest feed to gas flux ratio of 9.1, which corresponded to using relatively low gas fluxes and high feed liquid fluxes. Kracht *et al.* [16] obtained similar results using a cylindrical frit and sleeve sparging arrangement with gas fluxes, based on the vessel area, of up to 1 cm/s. Here, gas fluxes reached up to 5.5 cm/s. It is evident that for the two lower feed to gas flux ratios, the variation in the bubble size with the feed flux was similar. At the highest feed to gas flux ratio of 9.1 the bubble diameters were clearly much smaller, approaching a mean value of 0.37 mm at a feed flux of 15.4 cm/s.

4.2 Superficial overflow liquid velocity

In conventional flotation of fine particles rapid froth drainage is essential in rejecting water and hence the entrainment of slimes. In ion flotation it is also essential to limit the flux of liquid reporting to the overflow given the importance of concentrating the valuable components of the feed into the smallest volume of product that is possible. Flotation processes are constrained by this need to limit the liquid overflow flux. Thus, it is important to ensure relatively low overflow liquid fluxes are obtainable when operating with high feed and gas fluxes. To place

the performance of the Reflux Flotation Cell into context, experiments were repeated using a vertical flotation column.

Figure 8 shows a remarkably lower overflow liquid flux, j_f , leaving the Reflux Flotation Cell compared to that leaving the vertical column when operating under flooding conditions, using gas fluxes up to 5.5 cm/s. This result is explained by the inclined channels in the Reflux Flotation Cell providing a further degree of freedom in the segregation of the bubbles and liquid. Consequently, the underflow rate was increased to levels much higher than were possible in the vertical column, drawing the bubbly zone down further into the system of inclined channels. Increasing the underflow rate in the Reflux Flotation Cell reduced the liquid flux to overflow to levels well below that of the conventional vertical arrangement. Any further increase in the underflow rate of the vertical column would have resulted in significant bubble entrainment to the underflow, which must be avoided to maximize product recovery.

The overflow liquid flux increased strongly, and linearly, in both systems with increasing gas flux under flooding conditions. The linear increase is inconsistent with the typical quadratic-type dependency observed in froth flotation under normal non-flooding operation [17], thus reflecting the difference in the hydrodynamics of the Reflux Flotation Cell compared with conventional systems. Interestingly, for the Reflux Flotation Cell, the rate of increase of the liquid overflow flux versus the gas flux remained reasonably constant at around 2.0, with varying feed to gas flux ratio, j_o/j_g . In comparison, this rate increased from 2.0 to 4.2 to 12.3 with j_o/j_g increasing from 2.3 to 4.5 to 9.1 for the vertical flotation column.

At the highest gas flux of 5.5 cm/s the liquid flux reporting to the overflow remained relatively low, at less than 3 cm/s, using the Reflux Flotation Cell. This gas flux was tested using the lowest feed to gas flux ratio of 2.3. For this j_o/j_g ratio the largest feed flux tested was 12.4 cm/s. Therefore the portion of liquid reporting to the overflow was generally well below 24 %,

compared to the vertical flotation column where between 50 and 80 % of the liquid feed flux reported to the overflow. This is observed in Fig. 9. Hence, a moderate target of say 10 % liquid split is achievable at very high feed and gas flux. Fig. 9 also reveals that it is possible to maintain a low liquid flux to overflow with increasing gas flux. This means that the amount of liquid to the overflow can be controlled independently of the gas flux input to the flotation system.

4.3 Bubble surface area flux

The bubble surface flux, S_b , passing through the Reflux Flotation Cell is shown in Fig. 10 as a function of the underflow flux. The bubble surface flux, given by $S_b = 6j_g/D_b$, was calculated using the mean bubble diameters shown in Fig. 6. As shown in Fig. 6, the feed to gas flux ratio had an impact on the bubble size. Hence each data set in Fig. 10 is reported for a fixed feed to gas flux ratio.

The experimental results in Fig. 10 are remarkable given the liquid split to the overflow was maintained to low and acceptable levels under extreme flooding conditions and high feed fluxes. In all three cases, the bubble surface flux exceeded 178 s^{-1} , with the underflow flux no less than 9 cm/s. In particular, when the gas flux reached 5.5 cm/s, the bubble surface flux increased to an extraordinary 600 s^{-1} , based on a mean bubble diameter of 0.550 mm, while the underflow liquid flux was 9.5 cm/s. Equation 8 suggests that the terminal rise velocity for a bubble of this diameter is of order 5.7 cm/s, indicating that the underflow liquid flux was nearly twice the terminal rise velocity. For the higher j_o/j_g ratio of 9.1, the underflow flux is shown to have reached over 14 cm/s, four times the calculated terminal rise velocity of 3.6 cm/s for a bubble of diameter 0.370 mm. Here the bubble surface flux was a modest 275 s^{-1} . These results contrast with the relatively low bubble surface flux values used in mechanical flotation cells of 27 to

36 s⁻¹ [18], although fluxes as high as 80 s⁻¹ have been reported [19]. Massinaei *et al.* [20] reported a higher range of 30 to 60 s⁻¹ for industrial flotation columns deployed in rougher circuits. Whereas, in laboratory flotation columns the bubble surface flux typically lies between 12 and 120 s⁻¹ [21].

Arguably, the bubble surface fluxes achieved in the vertical column experiments presented in Fig. 8 and 9 were the same, or similar in value to those given in Fig. 10. The difference in the present study is that it is possible to operate at, and indeed well above, the flooding condition, while in conventional flotation operation must occur well below the flooding condition in order to maintain a reasonably low liquid split to the overflow. Indeed, the results in Fig. 8 and 9 were all achieved under flooding conditions with the underflow fluxes much lower in value, ranging between 3 and 5 cm/s. Yianatos and Henríquez [22] noted flooding to occur at a gas flux of 2 cm/s when the bubble diameter, D_b , was about 1.5 mm. In this study, the bubbles are nearly an order of magnitude smaller, and therefore the onset of flooding was reached at lower gas fluxes. Thus it appears the Reflux Flotation Cell offers the potential to deliver a ten-fold throughput advantage over conventional flotation cells.

Such high bubble surface fluxes were possible in the Reflux Flotation Cell using such extreme processing conditions due to the parallel inclined channels applied below the vertical zone, which ensured the retention of the gas bubbles. Bubbles entering each channel segregated towards the downwards facing inclined surface of the channel then rose up the inclined surface, returning to the upper vertical zone. It is acknowledged that some exceedingly fine bubbles, arguably finer than 50 µm, were entrained into the underflow at the highest gas fluxes. By passing a light into the inclined channel zone it was possible to see individual tracers of these fine bubbles sweep through. What was very clear however, was that their volume fraction in the underflow was almost negligible, and of little concern.

4.4 Considerations for future application

The recovery of ions or hydrophobic particles from low grade, dilute streams frequently requires a significant number of large flotation cells. This approach is often considered uneconomical due to the substantial capital investment and energy consumption. Hence, such material is subsequently discharged as tailings. In order to process such dilute streams economically, a high throughput rate per unit vessel area and low residence time is desirable to reduce cell footprint. However, fast flotation will only be feasible if high kinetic and high segregation rates are accommodated. At high gas fluxes the benefit in the recovery of valuables is only realised if a satisfactory liquid split of the feed to the overflow is maintained.

In order to promote improved particle flotation kinetics or increased rates of surfactant adsorption it is desirable to operate with very fine bubbles. Yoon and Luttrell [23] reported on the increase in bubble-particle collision efficiency with decreasing bubble diameter. For bubble diameters, D_b , smaller than 100 μm , the collision efficiency varies inversely with D_b^2 . In a downcomer the kinetics is improved by using a very high liquid shear rate and increased bubble number concentration [2]. However, flotation processes such as mechanical cells, column flotation or pneumatic flotation are constrained to operate with relatively large bubbles, given fine bubbles induce flooding at a lower gas flux and are transported to the underflow when using very high feed liquid flux. The Reflux Flotation Cell is not constrained to the same extent in this way. It is expected that the benefit of processing with fine bubbles and a high shear rate in the Reflux Flotation Cell should multiply, producing an increase in flotation kinetics. This view will be investigated in a forth coming study. It should be noted that the segregation advantage offered by the inclined channels to achieve fast flotation is not exclusive to a pneumatic downcomer design, and could equally be applicable to applications such as

mechanical and column flotation, or other bubble-liquid segregating processes, including the dissolution of fine gas bubbles in liquid.

5. Conclusion

The Reflux Flotation Cell has been operated with moderate to extreme levels in both the liquid flux and the gas flux, with the ability to control the overflow liquid flux independently of the gas flux. The novel feature of the inclined channels located below the main vertical bubbly bed permitted enhanced segregation rates between the bubbles and liquid, thus allowing the bubbles entrained by the high downward liquid flux to underflow to be refluxed back to the bubbly-bed above, and then report to the overflow as product. The capacity advantage provided by the inclined channels allowed for short residence times and bubble surface fluxes, generated under high shear rates, to reach as high as 600 s^{-1} . This value is unlikely to have been previously reached in flotation under normal gravitational force. It is anticipated that the Reflux Flotation Cell will achieve throughput capacities far greater than previously possible using conventional flotation processes.

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Figure Captions

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

Fig. 2. Flux curve versus bubble volume fraction, given by $V_t = 1$ cm/s and $n = 3$.

Fig. 3. Flux curve with the operating line set at the flooding condition. Here $V_t = 1$ cm/s and $n = 3$.

Fig. 4. Drift Flux prediction of the bubble surface flux versus bubble diameter at the flooding condition for conventional flotation and the Reflux Flotation Cell for $n = 3$.

Fig. 5. Photograph of the bubbles in the Reflux Flotation Cell taken for each j_o/j_g ratio. (a) $j_o/j_g = 9.1$, where $j_o = 15.4$ cm/s and $j_g = 1.7$ cm/s; (b) $j_o/j_g = 4.5$, where $j_o = 11.3$ cm/s and $j_g = 2.5$ cm/s; (c) $j_o/j_g = 2.3$, where $j_o = 11.3$ cm/s and $j_g = 5.0$ cm/s. The resolution of each image was 450×437 , 437×438 , and 436×438 pixels, respectively.

Fig. 6. The effect of the feed flux on the bubble diameter for three different j_o/j_g ratios. The error bars denote 95% student t confidence intervals.

Fig. 7. The relationship between shear rate within each channel of the downcomer and the bubble diameter for three feed to gas flux ratios, j_o/j_g .

Fig. 8. The effect of the gas flux on the overflow liquid flux using the Reflux Flotation Cell (RFC) and the vertical flotation column (CF).

Fig. 9. Split of the feed liquid flux to the overflow as a function of the gas flux using the Reflux Flotation Cell (RFC) and the vertical flotation column (CF).

Fig. 10. Bubble surface area flux versus the liquid underflow flux in the Reflux Flotation Cell.

Fig. 1

Fig. 2

Fig. 3

Fig. 4

Fig. 5

Fig. 6

Fig. 7

Fig. 8

Fig. 9

Fig. 10