BUBBLY FLOWS AND THE PLUNGING JET FLOTATION COLUMN

Graeme J. Jameson
Department of Chemical Engineering
University of Newcastle
Newcastle, NSW
Australia

ABSTRACT

A new type of gas-liquid contacting device has been invented and is now in commercial use worldwide for the separation of minerals and fine coal particles. The device is a radical departure from previous technology, and its development has drawn heavily on classical theory of bubble mechanics, and the fluid mechanics of two-phase flows.

The phenomena which underlie the new flotation cell are described, together with research results regarding the behaviour of confined plunging jets.

INTRODUCTION

Bubble columns are used in the chemical and mineral processing industries, as a means of contacting gas and liquid phases. In the minerals industry, the separation of valuable particles from the host rock or gangue is carried out by flotation, in which the finely ground mineral is suspended in water, and the suspension is treated with reagents which render the particles to be floated non-wetting or hydrophobic, leaving the gangue in a wettable state. The suspension is then brought into contact with air bubbles which adhere to the non-wetting particles and carry them to the surface where they form a froth which can be removed. The gangue leaves the contact vessel as a tailings.

For the past eighty years, flotation has been carried out in stirred vessels, with air bubbles being generated by a rotating impeller. In recent years there has been a considerable growth in other forms of contacting devices, such as the bubble column. A sketch of a typical bubble column is shown in Figure 1. The liquid feed containing the particles enters at the top of the column, and makes contact with a stream of bubbles which are generated in the base of the column. Contact takes place in the "collection" zone. The froth which forms is usually quite deep, of the order of a metre or more, and as shown in the sketch, the froth is often washed with clean water, to remove gangue material which is entrained with the bubbles as they pass from the liquid phase into the froth. Clean concentrate is removed from the top of the column and the gangue leaves in the tails stream.

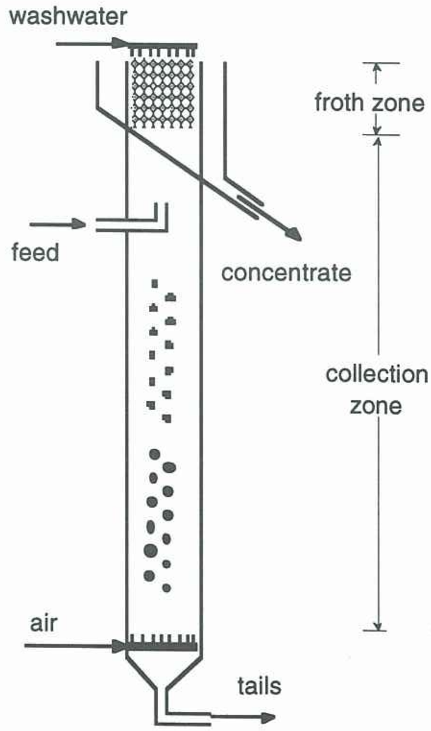

Figure 1. Diagrammatic sketch of a bubble flotation column. These columns are typically 10 to 15 metres in height and up to 4 metres in diameter.

Recently, another form of flotation column has been introduced, by the author, known as the Jameson Cell (Jameson, 1988). This cell works on a radically different principle to the bubble columns. It has now in use at over 100 sites around the world. In Australia, the value of coal recovered from tailing streams through its use is of order $75 million a year in exports, which would otherwise have gone to waste.

In this paper, some of the principles which led to the development of the Jameson Cell are discussed. Although a very simple device, it will be shown that its proper operation depends on a knowledge of classical theory for flow around bubbles, and of two-phase flows. Many of the phenomena on which it depends have been elucidated, but there are still some puzzling characteristics yet to be explained.

INITIAL CONSIDERATIONS

The Jameson Cell was the culmination of over 20 years of research into the physical mechanisms of flotation, and especially regarding the flotation of very fine particles. In the 1960's, it was recognised that the industry was moving towards finer and finer grinds, in order to liberate particles of values from the host rock, in so-called complex sulfide ore deposits. The flotation of fine particles, those below 20 µm in diameter, presented the conventional mechanical flotation cells with difficulties, in that the rate of flotation of such particles was known to be much slower than that of larger particles, so larger and larger cell volumes were required. Research by various authors, including Collins and Jameson (1976, 1977) showed that there was nothing mysterious about the behaviour of the fine particles, but that the diminished rate of flotation could be explained in hydrodynamic terms, including the effect of electrostatic forces arising from double-layer repulsion. Flotation was viewed simply as the result of a collision between a spherical bubble rising in a suspension of fine particles. These researches suggested that the flotation rate constant varied approximately as

$$k \approx d_p^2/d_b^3. \tag{1}$$

where d_p is the particle diameter and d_b is the bubble diameter. Thus the rate of flotation could be improved by reducing the size of the bubbles, which were customarily made by the shearing action at the tip of a rotation impeller.

The invention of the Jameson Cell was rather serendipitous. The author had seen flotation columns being installed at Mount Isa in Queensland, and it seemed that there was a lot of "dead" space. The column volume needed for contacting was very large and residence times are typically of order 20 to 30 minutes. It was felt that if the contacting process could be made more efficient, the volume of the device could be considerably reduced, with consequent savings in construction and operating costs, and in convenience of operation. The collection zone, which occupied a total height of 12m approximately, seemed inordinately large, and it would be advantageous if it could be reduced in height or even done away with, so the author began playing with new ideas on paper and in the laboratory, aimed at an improved design.

Based on previous work, two guiding principles were followed:

(a) The bubbles in the system should be quite small, in the range 250 to 500 µm in diameter. This would give a high surface-to-volume ratio of the air dispersion, while retaining a sufficient buoyancy force so that the bubbles could rise in the disengagement zone, and not be carried out in the tailings.

(b) The void fraction should be as high as possible, so as to give a high surface-to-volume ratio for the contactor as a whole. It was realised that a countercurrent downwards design would best achieve this situation.

In order to control the bubble size, it was necessary to be able to predict the bubble diameter as a function of the flow environment, and the paper of Lewis and Davidson (1982) came to mind. They had investigated the splitting of a bubble in a shear flow, and had produced a simple criterion from which it would be possible to predict the maximum possible stable bubble size. In their analysis they assumed that a bubble of diameter d_b existed in a uniform shear field ω . Relative to the centre of the bubble, the velocity at the bubble extremities would be higher by an amount $\Delta U = \omega d_b/2$ approximately, leading to a pressure deficit of $(\rho_L \Delta U^2)/2$ by Bernoulli. The pressure reduction on diametrically opposite sides of the bubble would tend to split it apart, but would be opposed by the restraining force of surface tension, which is of order σ/d_b. Thus when $\rho_L \omega^2 d_b^2 >> \sigma/d_b$, the bubble will split, and for a given bubble diameter, there is a critical ratio of the splitting force to surface tension restoring force, which when exceeded, will cause the bubble to split. Lewis and Davidson termed this dimensionless ratio the critical Weber number

$$We_{crit} = \rho_L \omega^2 d_b^3 / \sigma \tag{2}$$

and by a numerical analysis, found We_{crit} to be 4.7.

For various practical reason, the author decided to create a shear field by passing the liquid through an orifice. The shear rate is then of order

$$\omega \approx U_j/d_j \tag{3}$$

U_j, d_j being the velocity and diameter of the liquid in the orifice, and since these two parameters can be varied in engineering terms, it should be possible to provide the shear rate appropriate to the formation of bubbles of the desired diameter.

A number of configurations were tried based on these considerations of bubble size, and the arrangement shown in Figure 2 was eventually arrived at. The design is of the utmost simplicity. The liquid issues as a jet , so the appropriate terms in equation (3) are thus the jet velocity and the jet diameter.

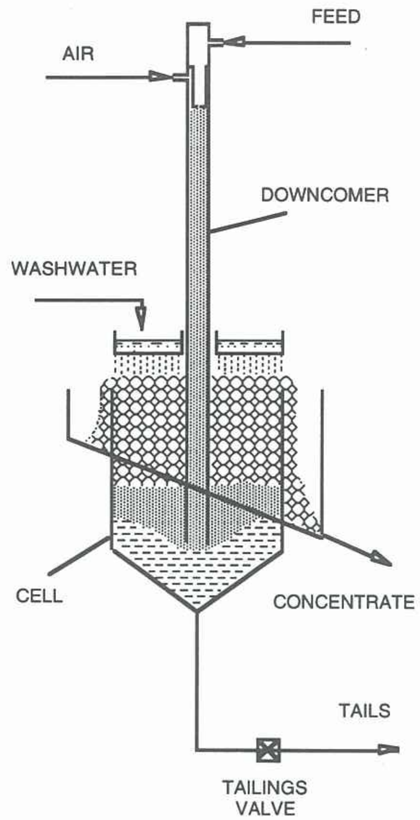

Figure 2. Diagrammatic sketch of a Jameson Cell. The downcomer is typically 3 m length and the cell itself is 1.5 m high. Largest cell diameter to date is 6 m. Downcomer diameter is up to 0.3 m.

In operation, the feed liquid is introduced in the form of a jet at the head of a "downcomer", which consists of a vertical tube which is sealed at the top except for a regulated air inlet and a vertical nozzle through which the liquid feed is introduced. The base of the downcomer is located below the pulp level inside the riser. On startup, the air inlet at the top of the downcomer is closed and the pulp feed is pumped through the nozzle. The air in the downcomer is entrained into the pulp which forms the seal in the bottom of the cell, and consequently, the pulp is drawn upwards from the cell into the downcomer. The pulp level reaches the tip of the nozzle quite quickly, and as a consequence of the hydrostatic suction developed by the head of this pulp, the pressure in the head of the downcomer is less than atmospheric. When the inlet is opened, air is drawn into the headspace at the top of the downcomer where it is entrained into the downcomer contents by the plunging jet.[1] The entrained air is broken up into fine bubbles which are quickly dispersed into the pulp and carried downward by the bulk fluid motion.

The three-phase mixture passes from the base of the downcomer into the cell proper, which has a much greater cross-sectional area than the downcomer. Consequently, the downward superficial velocity of the mixture is reduced, allowing the particle-laden bubbles to disengage from the liquid and rise to the surface and form a layer of froth. The froth then drains before overflowing into a collection launder, while the liquid phase and unrecovered particles leave through a valve at the base of the cell.

The fact that air can be drawn directly into the downcomer from the atmosphere is a strong practical advantage, because it obviates the need for a compressor. It should be emphasised that air induction is almost entirely due to hydrostatic effects, and dynamical effects such as are found in the conventional inductor or ejector are largely absent[2].

BUBBLE DYNAMICS IN THE DOWNCOMER

Once it had been demonstrated to work in the laboratory and in the field, it became necessary to provide design criteria and to carry out further studies of the behaviour of the plunging jet. Many of the design criteria were suggested by classical works in the field. For example, it is clear that the downwards velocity of the liquid should be such that it would exceed the rise velocity of the bubbles, including all bubbles which may result from coalescence. The rise velocity of air bubbles rising in water is shown in Figure 3, and it is very fortuitous that over a wide range of equivalent bubble diameters, the dependence is very weak. The reason for the flattening out of the rise velocity is that as the bubble increases in diameter, it changes shape. Small bubbles, less than about 0.3 mm in diameter, behave as solid spheres and their rise velocity can be calculated from Stokes law. As they approach 1 mm in diameter, bubbles start to flatten out like oblate spheroids, and Moore (1963) was able to relate the terminal velocity to the change in shape to the stresses acting over the surface of the flexible bubble. He found that in the vicinity of 2 mm in diameter, there was a maximum in the rise velocity.

Bubbles greater than about 4 mm equivalent diameter adopt a spherical-capped shape, and the rise velocity can be predicted by the equation of Davies and Taylor (1950):

$$U_t = 0.711(gD_e)^{0.5}, \tag{4}$$

where D_e is the equivalent bubble diameter. In the present case, any very large bubbles which form in the downcomer are broken and dispersed by the

[1] The operation described here is only possible if the rate of coalescence of bubbles in the downcomer is small. In mineral processing, bubble coalescence is inhibited by the presence of surface active materials, including frothers which are deliberately added to provide a stable foam, so that the foam layer containing the values can be removed.

[2] The novelty in this design owes more to Archimedes than to Daniel Bernoulli.

confined plunging jet, so a downwards velocity in excess of say 0.3 m/s is normally adequate.

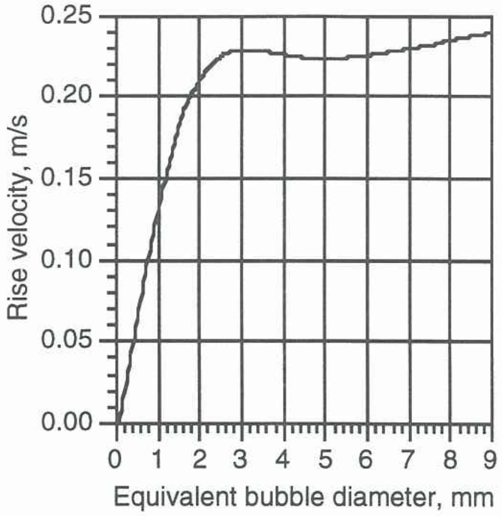

Figure 3. The terminal velocity of air bubbles rising in water contaminated with surface active agents, after Motarjemi and Jameson (1978).

Where the downcomer is very large, it is possible in principle for large air slugs to form, which could bridge the downcomer. Davies and Taylor (1950) showed that the rise velocity of such slugs was given by:

$$U_t = 0.46(ga)^{0.5}, \quad (5)$$

where a is the pipe radius. In practice, such slugs are not seen in the mixing zone in the downcomer, due to the disruptive action of the plunging jet. However, they do occur in the "pipe-flow" zone.

JET EXPANSION IN THE MIXING ZONE

The flow in the downcomer has been described by Evans (1990) and Evans et al. (1994). A number of flow regimes have been observed as depicted in Figure 4.

When the jet plunges into the confined bubbly mixture in the downcomer, air is entrained which is dispersed as fine bubbles in the shear layer adjacent to the jet. The jet itself loses momentum to the surrounding fluid, and expands with distance from the point of issue. At a certain point, the expanded jet touches the wall of the downcomer and from that point downwards, the gas-liquid mixture is essentially in a pipe-flow regime. Since the jet is confined, a recirculating eddy is established above the pipe-flow zone, and some bubbles escape from the eddy. The total air flow entrained into the jet is therefore the sum of air which is entering from the atmosphere and air which has escaped from the recirculating zone. The high velocity gradients between the submerged jet and the recirculating eddy result in high energy dissipation rates within the mixing zone, which are responsible for the breakup of the air once it is entrained by the plunging jet. The entrained air is broken into fine bubbles before being transported downward into the pipe-flow zone by the bulk fluid motion.

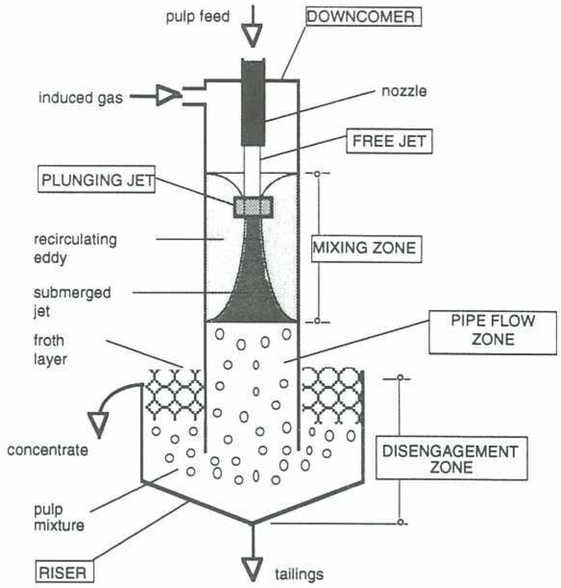

Figure 4. Flow regimes in the vicinity of a confined plunging jet.

The size of the bubbles produced inside the mixing zone – the volume confined between the surface of the froth and the point at which the jet touches the wall – is related to the forces acting on the bubble surface. In low viscosity fluids, like those normally encountered in flotation circuits, the bubbles are deformed by forces arising from liquid velocity fluctuations acting over distances of the order of the bubble diameter. The restoring force resisting the deformation of the bubble is due to surface tension acting at the air-liquid interface. The ratio of these forces can be written as a Weber number We, defined by

$$We = \rho_L \overline{u^2}\, d/\sigma \quad (6)$$

where $\overline{u^2}$ is the average value of the squares of the velocity differences acting over a distance of the order of the bubble diameter d ; and ρ_L and σ are the liquid (pulp) density and surface tension respectively[3]. If the bubbles are small compared to the turbulent macroscale but large compared to the microscale, the velocity fluctuations, following Hinze (1955), can be related to the energy dissipation rate per unit volume, $\dot{E}$, by the relationship:

$$\overline{u^2} = C_1 \left(\dot{E}\, d/\rho_L\right)^{2/3} \quad (7)$$

where $C_1 \approx 2$ according to Batchelor (1951).

[3] The derivation of equation (6) follows lines similar to those underlying equation (2), except that the pressure difference at the extremities of the bubble which lead to breakup is due to turbulent fluctuations rather than to a steady mean shearing flow.

For a given bubble size, and at low Weber numbers, the deformation forces are dominated by the restoring forces, and the bubble is stable. However, as *We* increases, the shape of the bubble becomes increasingly distorted. Eventually a critical Weber number, We_c, is reached where the bubble becomes unstable and breaks up into smaller bubbles. The maximum bubble size, d_m, corresponding to the critical Weber number is

$$d_m = (We_c/2)^{3/5}(\rho_L)^{-1/5}\dot{E}^{-2/5} \tag{8}$$

Equation (5) has been used by Evans *et al.* (1992) to predict the maximum stable bubble diameter inside the mixing zone, using estimates of the critical Weber number and the energy dissipation rate per unit volume.

The volume of the mixing zone is determined by the point of contact of the expanding jet and the wall of the enclosing downcomer. By considering the radial flow of momentum from the incoming jet to the recirculating eddies in the mixing zone, Evans (1990) derived the following expression for the half angle β subtended by the expanding jet at the virtual origin:

$$\tan\beta = D_d/L_{MZ}$$

$$= \eta\left(\frac{P_1}{\rho_L U_j^2}\right)\left(\frac{\rho_{MZ}}{\rho_L}\right)\left\{\frac{U_j - U_{e,max}}{U_j}\right\} \times$$

$$\left\{\frac{\left(0.37(R_c^2 - 0.5\,R_j^2)\right)^{0.5} - 0.64R_j}{R_j}\right\} \tag{9}$$

where η is a jet energy transfer efficiency, found by experiment to be 0.089; ρ_{MZ} is the density of the two-phase mixture in the mixing zone; R_c and R_j are respectively the downcomer and jet radius; P_1 is the pressure in the headspace at the top of the downcomer, U_j is the jet velocity and $U_{e,max}$ is the maximum return velocity in the eddy, which has been found by experiment to be $0.085U_j$. In operation, the jet half-angle is typically about 7^o.

The volume of the mixing zone is then assumed to be the volume enclosed by the cylinder formed by the downcomer of internal diameter D_d and the length L_{MZ} of the mixing zone.

Evans *et al.* (1992) have compared the sizes of bubbles predicted by equation (5), with measurements of the sizes of bubbles formed in an air water system, and found agreement within ± 20%. Two downcomers were used, of diameters 44 mm and 74 mm, and the jet velocity varied from 7.8 to 11.53 m/s. The bubble sizes were generally in the range 200 to 400 μm.

TWO-PHASE FLOW IN THE PIPE ZONE

The void fraction in the downcomer, and especially in the pipe flow regime is regarded as important in providing an environment capable of providing the maximum probability of capture of particles by the bubbles. Measurements have shown that the void fraction can be as high as 0.5, which compares very favourably with values of 0.05 to 0.1 reported for mechanical flotation cells, and 0.1 to 0.15 for flotation columns. However, without further examination, it is not possible to say that the void fraction is entirely due to a dispersion of fine bubbles. For a given liquid flowrate, when the air to liquid ratio is small, less than 0.1, all the air introduced into the downcomer is dispersed into fine bubbles. As the ratio is increased however, it is possible that there will be a maximum in the air to liquid ratio at which it is possible for an assemblage of fine bubbles to be created which do not coalesce to any significant extent. It is quite possible for example, that as the air flow is increased, a point will be reached where the small bubbles are so tightly packed that there is no more room for them in the liquid, and any increase in the gas to liquid ratio will lead to the formation of a different flow regime, consisting of large bubbles or slugs moving together with a dispersion of fine bubbles. In the literature, two flow regimes have been identified which appear to coincide with the two cases discussed here: bubbly flow at low void fractions, and churn-turbulent flow at high void fractions.

Evans and Jameson (1994) have recently reported experiments in which the air void fraction, and the volumetric flowrates of the air and water were analysed using the drift flux model of Wallis (1962). The aim was to discover if there was any critical void fraction which would mark a transition from the bubbly flow to the churn-turbulent regime.

Wallis defines a drift velocity, v', for each phase which is the difference between the component linear velocity, v, and the total volumetric flux, J, for the two-phase mixture. For example, the gas drift velocity is equal to

$$v_g = j_g / \varepsilon, \tag{10}$$

where the total volumetric flux is the sum of the gas, j_g, and liquid, j_l, volumetric flux components, and is given by

$$J = j_g + j_\ell = (Q_g + Q_\ell)/A. \tag{11}$$

In eq. (11) Q_g and Q_ℓ are the gas and liquid volumetric flow rates respectively, and A is the cross-sectional flow area of the column.

Wallis also defines a drift flux, j', for the gas and liquid phases as being equal to the drift velocity multiplied by the fractional flow area occupied by the phase, i.e.

$$j_g' = \varepsilon v_g' \quad , \tag{12a}$$

$$j_l' = (1-\varepsilon)v_l' \quad , \tag{12b}$$

For a given bubble size, and at low Weber numbers, the deformation forces are dominated by the restoring forces, and the bubble is stable. However, as We increases, the shape of the bubble becomes increasingly distorted. Eventually a critical Weber number, We_c, is reached where the bubble becomes unstable and breaks up into smaller bubbles. The maximum bubble size, d_m, corresponding to the critical Weber number is

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$$j'_g = \varepsilon v'_g \quad , \tag{12a}$$

$$j'_l = (1-\varepsilon)v'_l \quad , \tag{12b}$$

where ε is the gas void fraction. By assuming that there is no net drift through the plane moving at the superficial velocity of the gas-liquid mixture then the system can be described by analysing the flow equations for either the gas or liquid phases.

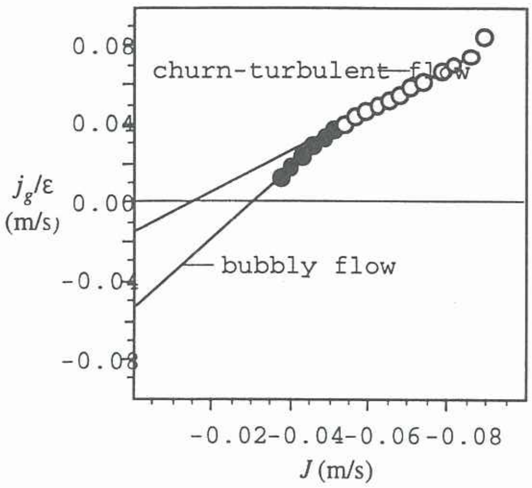

Figure 5. Drift flux plot for a particular experimental run.

Considering the gas phase only, equation (10) can be substituted into equation (12a), to give

$$j_g' = \varepsilon\left(v_g - J\right) \quad . \tag{13}$$

Expressing the total volumetric flux as the sum of the gas and liquid volumetric fluxes, equation (13) becomes

$$j_g' = \varepsilon\left(v_g - \left(j_g + j_l\right)\right) \quad , \tag{14}$$

and realising that

$$v_g = j_g/\varepsilon \quad , \tag{15}$$

the gas drift flux can be rewritten using only the volumetric flux terms and the gas void fraction, i.e.

$$j_g' = j_g(1-\varepsilon) - j_l\varepsilon \quad . \tag{16}$$

Equation (16) is a general expression based on the principle of continuity for the gas and liquid phases.

Wallis (1969) extended his drift-flux analysis and obtained an alternative expression for the gas drift flux, based on the terminal rise velocity of a single bubble in an infinite fluid, v_∞, and the gas void fraction, i.e.

$$j_g' = v_\infty g(\varepsilon)(1-\varepsilon)\varepsilon \quad , \tag{17}$$

where $g(\varepsilon)$ is a function of the gas void fraction. Ishii and Zuber (1979) developed expressions for $g(\varepsilon)$.

It will be appreciated that of the above variables, j_g, j_ℓ and ε are the only ones which can be measured directly, while the various drift velocities v_g, v_∞ etc are useful in providing insight into the phenomena. The value of the drift flux analysis can be seen from a number of representative plots of the data.

Figure 5 shows values of the drift velocity $v_g = j_g/\varepsilon$ at various values of the total superficial velocity J, following the analysis of Zuber and Findlay (1965). It is clear that in the experiment described, there is a transition in the slope of the data points at a particular value of J. The relevance of this point can be seen if the same data are plotted in the form $j_g' = j_g(1-\varepsilon) - j_l\varepsilon$ against the gas void fraction ε, as in Figure 6. Also shown on this plot is the volumetric drift flux curve predicted by the model of Ishii and Zuber (1979).

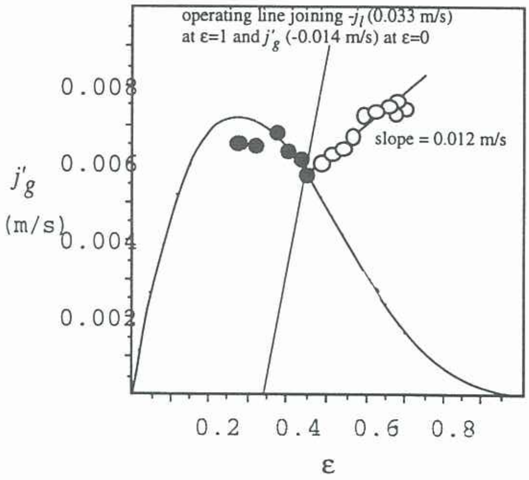

Figure 6. Values of the gas volumetric drift flux as a function of void fraction for a particular experiment. Conditions the same as in Figure 5.

It can be seen that for low void fractions, the experimental data follow the equation of Ishii and Zuber quite well, but at a value of the void fraction of 0.45 approx., there is a dramatic shift away from the model curve, indicating that there has been a change in the flow regime. The data suggest that above values of the void fraction of ε = 0.45, the bubbly flow is unable to accommodate further additions of air. Above this value, the bubbles are evidently so close-packed that any further air bubbles cannot be fitted in between the existing ones. Above ε = 0.45 for this particular flow, the two-phase mixture evidently behaves as if it were a dispersion of very large slugs or cavities moving in a dispersion of fine bubbles of void fraction 0.45.

STABILITY OF THE JET ENTRAINMENT SYSTEM

An effect which was not anticipated is the entrainment behaviour of the liquid jet as the air rate is increased. Referring to Figure 7, at low gas rates the downcomer is filled with a dense foam, but as the gas rate increases, it is seen that the height of the foam interface begins to drop, with corresponding increase in the headspace volume and the length of the exposed free jet.

It appears that at the higher gas rate, the system is able to come to a new equilibrium, in that for some reason, the rate of entrainment of gas into the dense foam is higher when the jet has been allowed to fall some distance from the point of formation, than when the foam interface is immediately adjacent to the jet exit point.

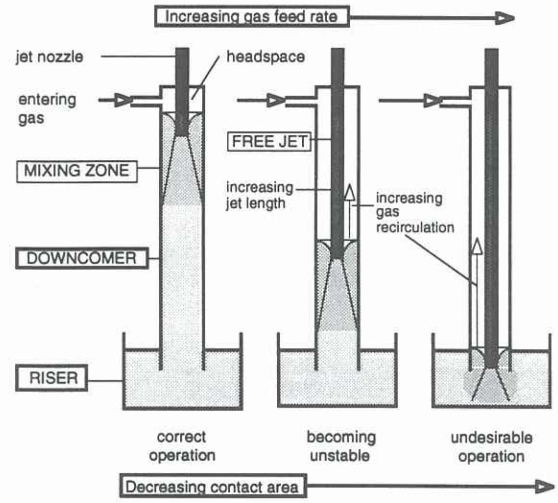

Figure 7. The effect of increased gas flow rate at constant liquid feed rate. At low rates the downcomer is filled with a dense two-phase foam. If the air flow is increased, the level of the top of the downcomer contents drops to a new equilibrium height at which the jet appears to be able to entrain the incremental air. When the air rate exceeds the maximum entrainment capacity of the jet, the foam bed collapses.

The explanation for this observation is not clear. The downcomer is typically 3 metres in length, and it can be shown that the increase in velocity due to free fall under gravity over the initial part of the downcomer is not very significant. It is possible that it is the relaxation of the velocity field inside the jet after departure from the nozzle which is responsible. Relative to the stationary downcomer, the surface velocity of the jet is initially zero, but momentum will transfer to the free jet from the interior, thereby tending to increase the surface velocity which must of course eventually become the same as the mean jet velocity. This phenomenon is still to be elucidated. If the air rate is increased further, a point is reached when the foam bed in the downcomer collapses totally, as shown in Figure 7.

It should be pointed out that the headspace in the downcomer is normally maintained at a pressure less than the ambient, so that air can be drawn into the downcomer from the atmosphere, obviating the need for a compressor or blower. The reason that the pressure is below atmospheric can be found by simple hydrostatic considerations, the pressure deficit due to changes in the momentum flux being of second order.

CONCLUSIONS

The Jameson Cell is an example of a practical fluid dynamical device which grew out of academic research relating to the behaviour of bubbles and bubbly flows. Cells based on this design are now in use at over 100 sites around the world. Many of the design considerations are based on classical results in bubble mechanics. The mechanisms of entrainment of air into the plunging jet are still not fully understood, and the unexpected stability of the system in response to changes in the gas flowrate is still be explained.

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