

GAS HOLDUP, LIQUID CIRCULATION AND MIXING BEHAVIOUR OF VISCOUS NEWTONIAN MEDIA IN A SPLIT-CYLINDER AIRLIFT BIOREACTOR

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A split-cylinder airlift reactor was used to characterize the effects of Newtonian viscosity on the induced liquid circulation rate, the overall gas holdup, the distribution of holdup between the riser and the downcomer zones, and the mixing time. Over the viscosity range $(1.54-19.50) \times 10^{-3}$ Pa s, the viscosity did not affect the liquid circulation rate because any viscosity-associated increase in frictional resistance to flow was compensated by a corresponding increase in the driving force for liquid circulation. The overall gas holdup declined with increasing viscosity, but the difference between the gas holdups in the riser and the downcomer zones increased. The mixing time was little affected by the viscosity. The observed mixing behaviour was satisfactorily correlated by a mechanistic equation that associated mixing to relative velocity between the phases and the length of the circulation loop. The ratio of the mixing time and the liquid circulation time was constant irrespective of the aeration rate or the viscosity of the liquid.

Keywords: airlift bioreactors; gas holdup; liquid circulation; mixing time; hydrodynamics

INTRODUCTION

Airlift bioreactors are elegantly simple devices that significantly outdo bubble columns in mixing, suspending solids, heat transfer, and other measures of performance¹. In comparison with mechanically agitated fermenters, airlift systems are almost always more productive in terms of the specific power demands¹ and commercial-scale effectiveness of airlift bioreactors has been proven in numerous applications, e.g., animal cell culture, microbial fermentations, and activated sludge wastewater treatment. Smaller-scale demonstration runs with plant cells and photosynthetic cultures have given good results in airlift bioreactors. The various aspects of operation, selection and design of airlift units for bioprocess and environmental use have been reviewed comprehensively elsewhere^{1,2}.

Substantial information exists on the hydrodynamics of airlift bioreactors¹⁻³, but much of it relates to concentric draft-tube types of systems and to the commercially less relevant external-loop configuration. Split-cylinder airlift devices have not been examined extensively even though they have the advantage of providing the smallest internal wall area in contact with the fluid relative to other types of airlift reactors. Reduced fluid-wall contact should lessen frictional resistance to the induced liquid circulation for a given power input. Consequently, split-cylinder devices are especially attractive for use with relatively viscous fluids. Viscous Newtonian media such as sugar solutions have been previously investigated in external-loop reactors^{4,5} which, because of the gas-liquid separating effect of the head zone⁶, behave very differently than internal-loop reactors. Effects of viscosity on liquid circulation and gas holdup have been reported in draft-tube type of reactors by Philip *et al.*⁷. Here

a report on liquid circulation, mixing, and gas holdup in a split-cylinder device operated with relatively viscous homogeneous Newtonian media is given.

EXPERIMENTAL

A split-cylinder internal-loop reactor shown schematically in Figure 1 was used. The reactor consisted of a glass tube partitioned into riser and downcomer by a Teflon[®] baffle. The baffle was 0.055 m wide; thickness of 0.004 m and a total height of 0.7 m. The riser-to-downcomer cross-sectional area ratio was unity. The clearance, h_c , of the baffle above the sparger (see Figure 1) was 0.05 m, whereas the clearance, h_b , between the static liquid surface and the upper edge of the baffle was 0.06 m. The vessel diameter was 0.056 m; its overall height was 0.91 m. The static liquid level was 0.81 m giving a working volume of 1.75×10^{-3} m³. Air was sparged into the reactor through a sintered glass disc (1.5×10^{-2} m sparger diameter; 110×10^{-6} m pore size) located at the base of the riser (see Figure 1). The gas flow was metered and controlled by a rotameter, pressure gauge and control valve arrangement. The air flow rates range was $0-0.039 \times 10^{-3}$ m³ s⁻¹, corresponding to the riser superficial gas velocities in the range $0-0.039$ m s⁻¹. The specific power input due to aeration varied between $0-234$ W m⁻³. Measurements were done at room temperature (ca. 25°C).

Sucrose solutions in deionized water were used for the liquid phase in various experiments. The sucrose concentration ranged over $162-500$ kg m⁻³, providing a Newtonian viscosity variation of $(1.54-19.5) \times 10^{-3}$ Pa s as noted in Table 1. The liquid velocity and mixing time measurements

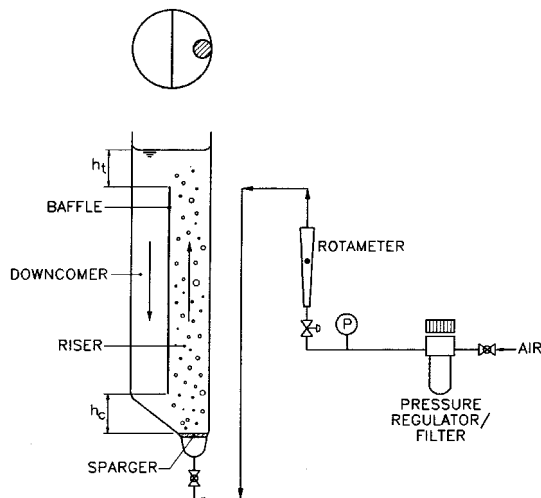


Figure 1. Schematic of the split-cylinder airlift bioreactor.

were based on an acid–base decolourization reaction, a commonly used technique for these types of measurements^{2,8}. Hydrochloric acid (1 M) or sodium hydroxide (1 M) were added to the sucrose solution as required and phenolphthalein was the acid–base indicator. The liquid velocity in the riser and the downcomer was determined by measuring the time taken by the purple colour front, developed on adding 0.5 mL of sodium hydroxide solution to the reactor, to travel a certain distance (0.4 m) within the riser or the downcomer. The solution was added to the surface of the liquid, in the centre of the reactor. The mixing time was defined as the time taken for complete decolourization of the reactor, following the addition of 0.5 mL of the hydrochloric acid solution to the reactor which had previously been made alkaline with 0.5 mL sodium hydroxide⁹. The liquid circulation time, t_c , was determined as the time taken by the colour front to complete one full circulation through the riser–downcomer loop.

The gas holdup was determined by measuring the increase in height of the dispersion upon aeration; thus

$$\varepsilon = \frac{h_D - h_L}{h_D}, \quad (1)$$

where h_D is the gas–liquid dispersion height and h_L is the height of gasfree liquid.

RESULTS AND DISCUSSION

Gas Holdup

The variation of the overall gas holdup with the riser gas velocity is shown in Figure 2 for the various sugar solutions. The figure reveals three different flow regimes: (i) a

Table 1. Liquid phase physical properties.

| Sucrose concentration (wt/wt, %) | Viscosity (Pa s) | Density (kg m ⁻³) |
|----------------------------------|------------------|-------------------------------|
| 16.16 | 0.00154 | 1057 |
| 28.57 | 0.00299 | 1106 |
| 37.50 | 0.00584 | 1163 |
| 50.00 | 0.01950 | 1223 |

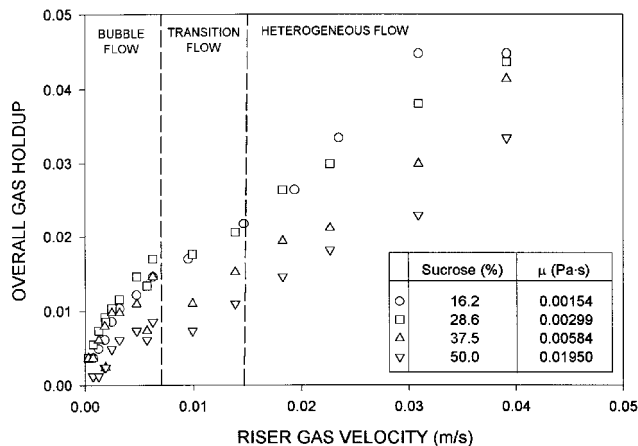


Figure 2. Overall gas holdup versus superficial gas velocity in the riser.

homogeneous bubbly flow, where the holdup increases linearly when gas velocity is increased. In this regime the bubbles rise independently without interacting with one another. This regime exists over the approximate gas velocity range $0 \leq U_{Gr} \leq 0.007 \text{ m s}^{-1}$. (ii) A transition flow regime occurs over the gas velocity range where the bubbly flow develops into heterogeneous flow; and finally, (iii) the fully developed heterogeneous flow ($U_{Gr} > 0.015 \text{ m s}^{-1}$). During the flow transition the concentration of bubbles increases and coalescence begins. In the transition regime, the rate of increase in holdup with increasing gas flow velocity is much lower than in the bubbly flow regime. When the coalescence is well established, the holdup again increases relatively strongly with increasing gas flow rate because of the formation of large ‘spherical cap’ bubbles. As can be seen in Figure 2, the demarcation between flow regimes is influenced by the sugar concentration. Similar behaviour has been observed previously in bubble columns, as well as in airlift bioreactors with various media^{10–14}.

The sugar concentration, i.e. the viscosity of fluid, has the greatest effect on gas holdup at higher air flow rates corresponding to the heterogeneous flow regime. The overall gas holdup declines with increasing viscosity of solution. In the past, small amounts of sugar in water have been reported to slightly enhance gas holdups relative to values in water^{5,14,15}; however, higher sugar concentrations have always led to a decline in holdup. The lowest value of sugar concentration used in this work (16.16%) was higher than the values up to which the sugar had a holdup enhancing effect in previous studies; hence, an increase in holdup relative to water was not observed in this work.

The overall holdup was correlated with an equation similar to the one proposed by Godbole *et al.*¹⁶ for bubble columns; thus,

$$\varepsilon = 0.095 \mu^{-0.211} U_{Gr}^{0.635}. \quad (2)$$

In this equation the exponents on the viscosity and the gas velocity terms are quite close to those reported by Godbole *et al.*¹⁶: $\mu^{-0.19}$ and $U_{Gr}^{0.6}$. Equation (2) correlated the data within $\pm 20\%$ as shown in Figure 3. Unlike in bubble columns, a circulating liquid flow exists in airlift reactors and the ratio of the gas and liquid velocities in the riser has been proposed¹⁷ for correlating gas holdup. Following this

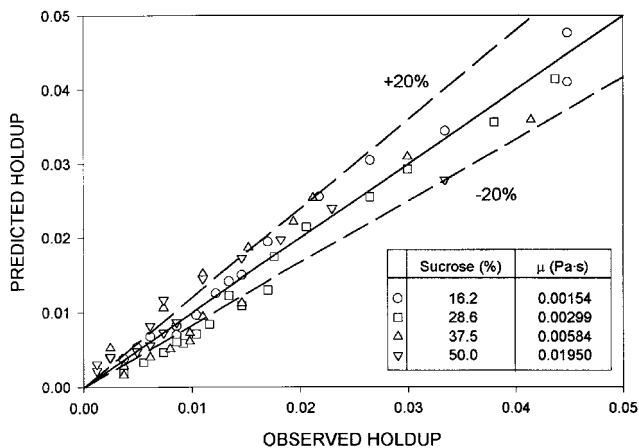


Figure 3. Overall gas holdup predicted with equation (2) versus the measured holdup.

approach another suitable equation for gas holdup is:

$$\varepsilon = 0.139\mu^{-0.158} \left(\frac{U_{Gr}}{U_{Lr}} \right)^{1.1} \quad (3)$$

Equation (3) suggests a lower influence of viscosity on gas holdup than does equation (2). The exponent on the velocity ratio term (i.e., 1.1) is almost the same as the values reported previously, but the parameter 0.139 is approximately half of that reported for salt solutions in a concentric tube airlift bioreactor¹³. As shown in Figure 4, equation (3) correlates the data only marginally better than does equation (2).

Liquid Circulation Time

The effect of superficial gas velocity on liquid circulation times for the different sugar concentrations is shown in Figure 5. At low gas flow rates, the circulation time decreased sharply with increasing superficial gas velocity; however, at a gas velocity of 0.01 m s⁻¹ or greater, the circulation time was almost independent of the gas flow rate. The viscosity had no significant influence on circulation time. Normally, for a given available driving force, an increase in viscosity of a fluid should reduce its flow rate because of a greater frictional resistance to flow. Absence of

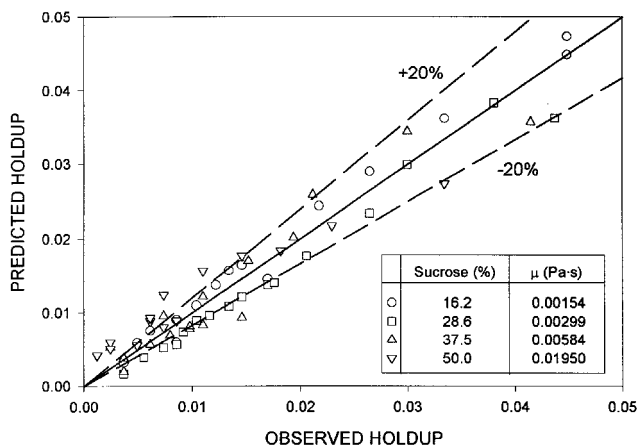


Figure 4. Overall gas holdup predicted with equation (3) versus the measured holdup.

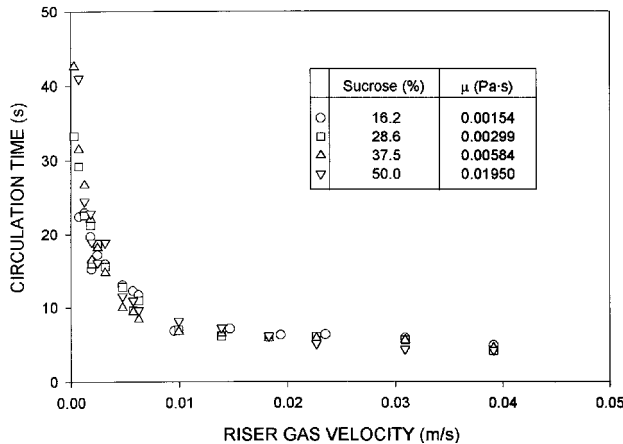


Figure 5. Liquid circulation time versus the superficial gas velocity in riser.

a decline in the liquid circulation velocity suggests that the driving force for liquid circulation—i.e., the difference between the gas holdups in the riser and the downcomer—increased with increasing viscosity for any given aeration rate. This was confirmed by visual observations and by an analysis of gas holdup distribution discussed later in this section.

Figure 5 shows another characteristic trend: over the gas velocity range 0.01 $\leq U_{Gr} \leq 0.025$ m s⁻¹, the circulation time is almost constant, then it declines slightly but significantly as the gas velocity increases further. This pattern has been observed and explained previously by Wenge *et al.*¹⁸ in split-cylinder airlift reactors that were almost 50-fold larger than the one used here. The exact aeration rate in the flat region (see Figure 5) at which the rate of decline in circulation time increases (i.e., $U_{Gr} \approx 0.025$ m s⁻¹ in this work) apparently depends on the scale and geometry of the reactor.

The riser and downcomer gas holdups were estimated following the method of Ganzeveld *et al.*⁹. Thus, the equation developed by Chisti *et al.*¹⁹ and repeatedly validated by numerous investigators^{20–25} was expected to apply also to the fluids used in this study. The equation, as recommended for a split-cylinder airlift reactor^{9,19}, is:

$$U_{Lr} = \left[\frac{2gh_D(\varepsilon_r - \varepsilon_d)}{K_B \left(\frac{A_r}{A_d} \right)^2 \frac{1}{(1-\varepsilon_d)^2}} \right]^{0.5} \quad (4)$$

where U_{Lr} is the superficial liquid velocity in the riser, g is the gravitational acceleration, h_D is the height of the gas-liquid dispersion, ε_r and ε_d are the gas holdups in the riser and the downcomer, respectively, K_B is the frictional loss coefficient for the bottom zone of the reactor, and A_r and A_d are the cross-sectional areas of the riser and the downcomer, respectively.

The superficial liquid velocity in the riser is related with the liquid circulation time by the following equation:

$$Q_L = U_{Lr}A_r = V_{Lr}A_r(1 - \varepsilon_r) = \frac{V}{t_c} \quad (5)$$

where V is the volume of the liquid in the reactor. Equation (5) assumes that all the liquid in the vessel circulates through the entire flow path in the interval corresponding to the circulation time t_c ; i.e., there is no dead zone as was the

case in this work. Hence, the mean circulation time and the volume of liquid can be used to calculate the liquid velocity in the riser:

$$U_{Lr} = \frac{V}{A_r t_c} \quad (6)$$

For split-cylinder reactors, the parameter K_B is related to the configuration of the bottom zone as follow¹⁹:

$$K_B = 11.402 \left(\frac{A_d}{A_b} \right)^{0.789} \quad (7)$$

In equation (7), A_b is the free area for liquid flow under the baffle. As ε_{db} the gas holdup in the downcomer, is much less than unity, equation (4) reduces to:

$$\frac{U_{Lr}^2 K_B (A_r/A_d)^2}{2gh_D} = \varepsilon_r - \varepsilon_d \quad (8)$$

In addition to equation (8), a second equation was necessary to determine the values for riser and downcomer holdups. This latter was the analytical relationship between the overall holdup (ε) and the holdups in the individual zones of the reactor²:

$$\varepsilon = \frac{\varepsilon_r A_r + \varepsilon_d A_d}{A_r + A_d} \quad (9)$$

As the geometric parameters (i.e., K_B from equation (7), h_L , A_r , A_d , and A_b) of the reactor were known and the overall gas holdup and liquid velocity (circulation time) were measured, the riser and the downcomer gas holdups could be calculated using equations (8) and (9).

Figure 6 shows the downcomer gas holdup plotted as a function of the riser holdup for the various fluids. For data in the bubbly flow regime, the relationship between the riser and downcomer holdups is independent of the sugar concentration; the relationship is:

$$\varepsilon_d = 0.815\varepsilon_r \quad (10)$$

This equation is similar to others reported in the literature^{2,11,17,26}.

In the heterogeneous flow regime, the sugar concentration strongly influences the distribution of gas holdup. The constant in equation (10) now depends on viscosity. This constant diminishes as the sugar concentration is increased. As shown in Figure 7, in heterogeneous flow, the relationship

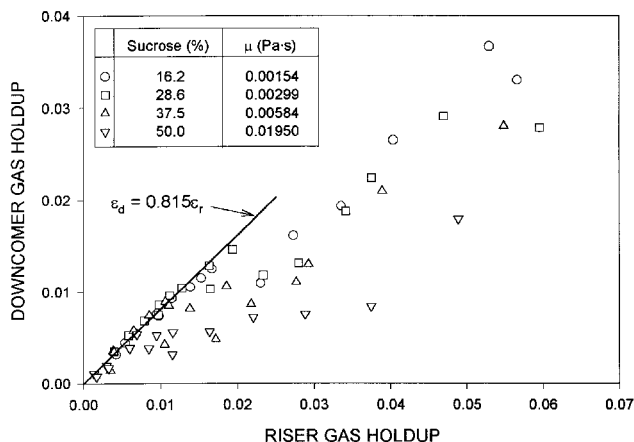


Figure 6. Relationship between the riser and the downcomer gas holdups.

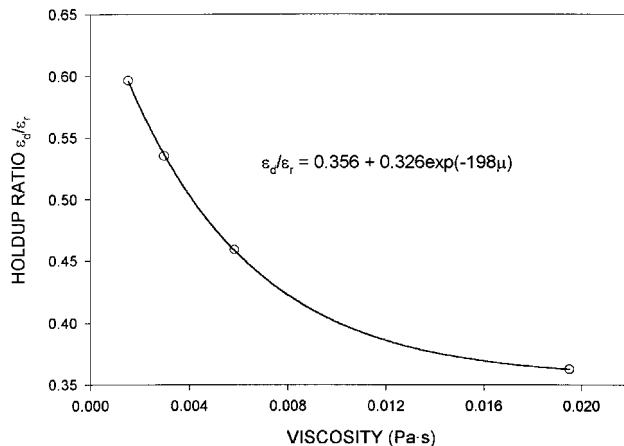


Figure 7. The riser-to-downcomer gas holdup ratio as a function of viscosity of the fluid (heterogeneous flow regime).

among the riser and the downcomer gas holdups and the viscosity becomes:

$$\frac{\varepsilon_d}{\varepsilon_r} = 0.356 + 0.326 \exp(-198\mu) \quad (11)$$

Equation (11) suggests that for a given holdup in the riser, the gas holdup in the downcomer declines as the viscosity of the fluid increases. This concurs with visual observations. The reasons for this effect are as follows: as viscosity increases, the mean bubble size in the riser increases. Many larger bubbles coexist with smaller ones in the heterogeneous flow regime in the riser section. Of these bubbles, only the smaller ones are dragged into downcomer zone near the top of the reactor. Most of the larger bubbles issuing from the riser leave the reactor because their rise velocity far exceeds the downflow velocity of the liquid. Hence, with increasing viscosity and a consequently bigger bubble size, fewer and fewer bubbles move into the downcomer.

Mixing Time

The measured mixing time is shown in Figure 8 as a function of the riser superficial gas velocity for the various sugar concentrations. The pattern of variation shown is similar to that previously seen for the circulation time

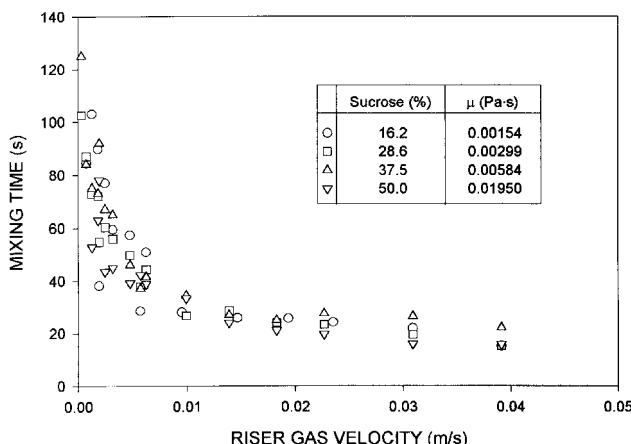


Figure 8. Variation of mixing time with the superficial gas velocity in the riser.

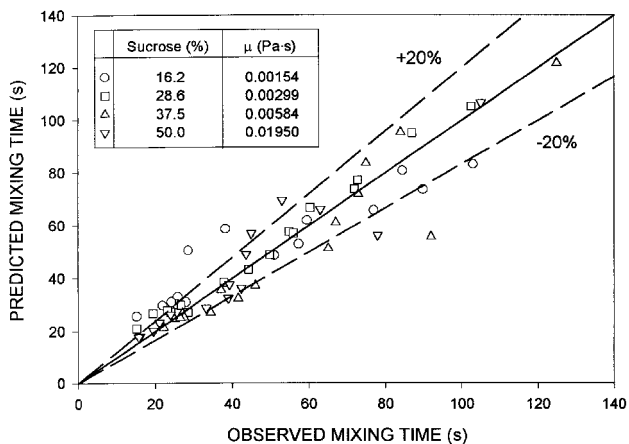


Figure 9. Comparison between the predicted (equation 12) and the measured mixing times.

(Figure 5). Again, the sugar concentration or viscosity has little effect on the mixing behaviour. Intuitively, the mixing time in an airlift reactor is expected to be affected by the relative velocity between the gas and the liquid phases because the gas bubbles rising in a liquid stream produce axial mixing by transporting liquid elements in their wakes. Also, the distance over which this mechanism operates—i.e., twice the height of dispersion—should affect the mixing time. Based on this envisaged mechanism, the following expression was used to correlate the mixing time:

$$t_m = 5.03 \left(\frac{U_{Gr} - U_{Lr}}{2h_D} \right)^{-0.797} \quad (12)$$

As shown in Figure 9, the values of mixing time predicted with equation (12) agree remarkably well with the experimental data.

Another mixing parameter in airlift reactors is the dimensionless mixing time¹⁷ which is defined as:

$$\theta = \frac{t_m}{t_c} \quad (13)$$

where t_m is the mixing time and t_c is the circulation time. The dimensionless mixing time in airlift reactors has been shown to be virtually independent of the gas velocity¹⁷; however, the dimensionless time depends on the A_d/A_r ratio:

$$\theta = \alpha \left(\frac{A_d}{A_r} \right)^{0.5} \quad (14)$$

According to Bello¹⁷, the α value is 5.2 for external-loop airlift devices and α is 3.5 for concentric-tube type of airlift reactors. The α value obtained in this study was 3.76 which is close to Bello's¹⁷ observations for internal-loop reactors. Note that Bello's value was obtained in draft-tube type of airlift system whereas this work employed a split-cylinder device.

CONCLUSIONS

Based on the hydrodynamic characterization studies discussed above, the following specific observations can be made:

1. Over the viscosity range $(1.54-19.50) \times 10^{-3}$ Pa s the

viscosity of Newtonian media has no significant effect on the induced liquid circulation rate in split-cylinder airlift reactors.

- The driving force for liquid circulation, i.e., $\varepsilon_r - \varepsilon_d$ increases with increasing viscosity of the fluid; this increase compensates for any viscosity associated increase in frictional resistance to flow.
- The overall gas holdup declines with increasing viscosity of the fluid; i.e., $\varepsilon \propto \mu^{-0.2}$.
- The mixing time is not affected by viscosity of the fluid over the range examined. This suggests that mixing is being influenced principally by the velocity difference between the gas and the liquid phases. An equation based on this mechanistic explanation correlates well with the observed mixing behaviour.
- The ratio of the mixing-to-circulation times appears to be a constant independent of the fluid viscosity in a reactor of a given geometry.

NOMENCLATURE

| | |
|----------|--|
| A_b | free area for liquid flow under the baffle (m^2) |
| A_d | cross-sectional area of the downcomer (m^2) |
| A_r | cross-sectional area of the riser (m^2) |
| g | gravitational acceleration ($m\ s^{-2}$) |
| K_B | frictional loss coefficient for the bottom zone of the reactor |
| h_c | clearance of the baffle above the sparger (m) |
| h_D | height of the dispersion (m) |
| h_L | height of gas-free liquid (m) |
| h_t | clearance between the static liquid surface and the upper edge of the baffle (m) |
| Q_L | liquid flow rate ($m^3\ s^{-1}$) |
| t_c | circulation time (s) |
| t_m | mixing time (s) |
| U_{Gr} | superficial gas velocity in the riser ($m\ s^{-1}$) |
| U_{Lr} | superficial liquid velocity in the riser ($m\ s^{-1}$) |
| V | volume (m^3) |

Greek symbols

| | |
|-----------------|--|
| α | parameter in equation (14) |
| ε | overall gas holdup |
| ε_d | fractional gas holdup in the downcomer |
| ε_r | fractional gas holdup in the riser |
| μ | viscosity (Pa s) |
| θ | dimensionless mixing time |

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