Air-Lift Reactor Analysis: Interrelationships Between Riser, Downcomer, and Gas-Liquid Separator Behavior, Including Gas Recirculation Effects

An approach to air-lift reactor analysis is presented that examines the three parts of the air-lift—the riser, downcomer, and gas-liquid separator—as three interconnected and interrelated elements. Downcomer flow configurations are described and characterized by either straight or oscillating bubble flow patterns. The downcomer gas flow rate is determined using a two-sparger system, and a correlation is presented relating the downcomer superficial gas velocity to the downcomer gas holdup. Gas recirculation is calculated and related to the geometric and operating conditions of the gas-liquid separator. It is shown that an increase in residence time in the gas-liquid separator decreases the gas recirculation rate. Correlations are presented relating gas holdup and liquid velocities to superficial gas velocity.

A sparger was added near the entrance of the downcomer, in addition to the conventional placement near the entrance of the riser. Results of such a two-sparger system are discussed in relation to reactor stability and liquid velocities.

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SCOPE

The term "air-lift" has been used to denote gas-liquid or bubble reactors that exhibit a clear and defined cyclic flow pattern. An air-lift reactor is comprised of three connected parts: riser, gas-liquid separator, and downcomer. The air-lift has, until now, been studied as a unit. There has been a wide variation in the data published in the literature for gas holdup in air-lift reactors as a function of the riser superficial gas velocity. This discrepancy is due to the different gas holdup measuring methods and the previous lack of inclusion of the recirculated gas in defining the riser superficial gas velocity. Our research has shown that the air-lift should be evaluated as three interconnected and interrelated elements, including gas recirculation effects.

The approach of the present study was to examine the three different sections of the air-lift and how they operate in unison, for 40 different air-lift reactor geometric configurations and/or operating conditions. The "true" riser superficial gas velocity used in this method of analysis is defined as the sum of the gas input rate (from the gas spargers) plus the gas flow rate in the downcomer due to recirculation. In all experiments the gas holdup was measured in both the riser and the downcomer.

A two-sparger system was used to determine the gas flow rate in the downcomer as well as to examine

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the possibility of operating an air-lift with two spargers. A two-sparger system may be advantageous in aerobic fermentation processes with high oxygen demand.

A correlation was found for the superficial gas velocity in the downcomer as a function of the measured gas holdup. This enabled the determination of the gas flow rate in the downcomer, and the gas recirculation rate. With this, it was possible to determine the true gas flow rate and superficial gas velocity in the riser. Correlations are presented for gas holdup and liquid velocity as functions of the true superficial gas velocity.

The data were analyzed both by considering the device as a unit, independent of recirculation effects, and by considering the effect of recirculation. The importance of the gas-liquid separator design on reactor performance was also demonstrated. It was found that the recirculation rate is primarily determined by the gas-liquid separator configuration.

Conclusions and Significance

From the experiments described here it is clear that the gas-liquid separator of the air-lift has a major influence on the entire behavior of the reactor. This work shows that the effects of the separator must be considered in design and scale-up of air-lift reactors. For effective control of the process, the reactor operator must be able to manipulate the gas recirculation rate and thus the fluid residence time in the gas-liquid separator. This can be accomplished by adjusting the liquid level and/or the size of the separator.

The results and conclusions obtained were as follows:

1. A correlation relating the downcomer superficial gas velocity to the downcomer gas hold-up was determined.
2. The gas recirculation rate was calculated. It is the ratio between the downcomer gas flow rate and the true riser gas flow rate.
3. A correlation was found for the riser liquid velocity as a function of the true riser superficial gas velocity and the ratio of the characteristic diameters of the riser and downcomer.
4. Oscillating bubble flow operating conditions were described. These are characterized by swirls and waves of descending bubbles in the downcomer. This is coupled with liquid velocity oscillations.
5. In order to describe the performance of an air-lift it is necessary to be able to describe the behavior of each of its parts and the interactions between them. Expressing data of one section of the air-lift in terms of properties of another section does not provide useful information for reactor design and scale-up.
6. The gas recirculation rate is largely determined by the gas-liquid separator's geometric configuration and the liquid level in the separator. A shorter fluid residence time in the gas-liquid separator increases the gas recirculation rate at the expense of reactor stability.
7. The gas holdup in the riser was satisfactorily correlated with the true riser superficial gas velocity for a wide range of operating conditions and gas-liquid separator configurations. It appears that for the air-lift the dependence of the gas hold-up on the superficial gas velocity can be approximated by the proportionality, $\phi = b(JG)^{0.7}$, regardless of the direction of flow.
8. In the downcomer, during stable operation the liquid velocity has a minimal effect on the gas holdup; however, the effect is pronounced in the riser.
9. A minimum downcomer relative gas-liquid velocity (approximately 20 to 30 cm/s, the terminal free rise velocity for air bubbles in water) is required to maintain straight bubble flow operating conditions in the reactor.

Overview

The effects of various reactor conditions in the riser, downcomer, and gas-liquid separator, were examined. The overall behavior of the air-lift is determined by the sum of these three parts. The extent of gas recirculation has a strong influence on the entire process. The amount of recirculation essentially determines the gas holdup in the downcomer, unless a second gas sparger is used to introduce gas to the downcomer. However, no data have been reported in the literature on the gas recirculation rate from the gas-liquid separator, the influence of gas-liquid separator parameters, or the gas flow rate in the downcomer. Also, *few data have been* reported on the gas holdup in the downcomer of air-lift reactors (Chakravarty et al., 1974; Orazem et al., 1979; Hsu and Dudukovic, 1980; Barker and Worgan, 1981; Field and Slater, 1983).

It is postulated here that the gas recirculation rate and gas holdup in the downcomer are determined by the gas-liquid separator configuration. The extent of gas-liquid separation in the separator will determine the pseudodensity of the gas-liquid dispersion in the downcomer. This in turn will determine the liquid velocity for any given geometric configuration and input gas flow rate. The gas holdup in the riser increases with decreasing liquid velocity. The recirculation rate and flow configuration are then influenced by the downcomer liquid velocity, and the bubble size distribution.

The data published in the literature for gas holdup in air-lift reactors have varied greatly, as shown in Figure 1. The lines rep-

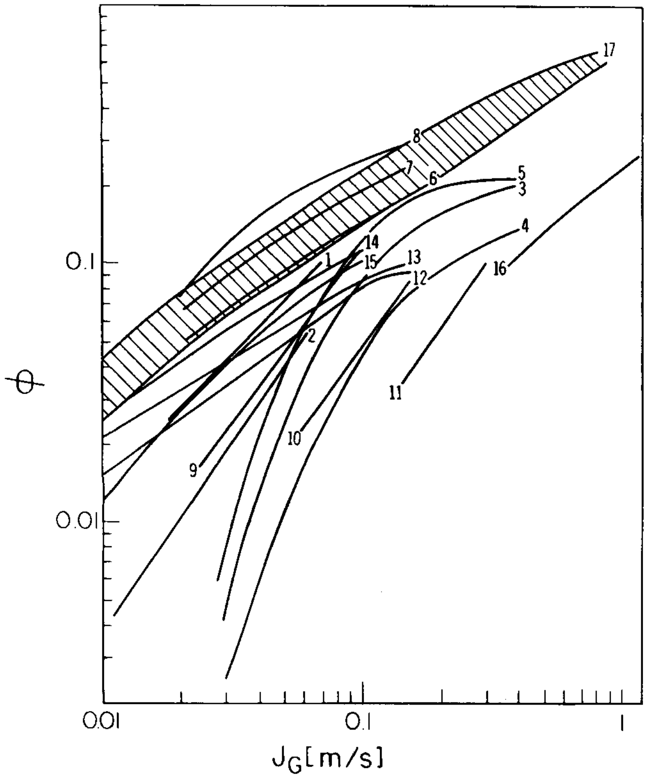


Figure 1. Experimental data on gas holdup in air-lift reactors.

	Air-Lift Type*	Reactor Dia., m	Draft Tube Dia., m	System	Reference
1.	CT	0.1	0.07	starch sol.	Barker et al. (1981)
2.	CT	0.3	0.21	starch sol.	Barker et al. (1981)
3.	CT	0.152	0.095	water	Field & Slater (1983)
4.	CT	0.152	0.095	1% ethanol	Field & Slater (1983)
5.	CT	0.152	0.095	antifoam agt.	Field & Slater (1983)
6.	CT	0.140	0.094	water	Koide et al. (1983)
7.	CT	0.140	0.082	270 mol/m^3 BaCl	Koide et al. (1983)
8.	CT	0.140	0.082	50% glycerol	Koide et al. (1983)
9.	CT	0.1	0.074	water (annulus)	Chakravarty et al. (1973)
10.	CT	0.1	0.059	water (annulus)	Chakravarty et al. (1973)
11.	CT	0.1	0.045	water (annulus)	Chakravarty et al. (1973)
	Air-Lift Type*	**Riser**	**Downcomer**	**System**	**Reference**
12.	EL	0.1	0.05	water	Onken & Weiland (1980)
13.	EL	0.1	0.05	51% sacchrose	Onken & Weiland (1980)
14.	EL	0.33	0.052	water (riser)	Akita & Kawasaki (1983)
15.	EL	0.33	0.052	water (downcomer)	Akita & Kawasaki (1983)
16.	EL	0.149	0.149	water	Hills (1976)
17.	Correlations for bubble columns, various authors				Shah et al. (1982)

*CT, concentric tubes; EL, external loop

resent gas holdup as a function of the riser superficial gas velocity, as reported by different authors. A wide range of operating conditions and reactor configurations are represented in these works. Some of the reactors had flow regimes that maximized the gas recirculation to the downcomer (generally, concentric tubes), while others had bubble-free liquid flowing in the downcomer (generally, internal loops). Also, most of the reported data on gas holdup do not differentiate between the different sections of the air-lift. This is particularly true when the gas holdup is determined by measuring the difference between the ungassed and gassed liquid levels, rather than from local pressure measurements. The shaded area in Figure 1 represents correlations for gas holdup in bubble columns (Shah et al., 1982). The reported data for bubble columns do not exhibit the wide variation of data from air-lift reactor experiments. The variation in the reported data for air-lifts is probably at least in part due to the exclusion of gas recirculation in the calculation of the superficial gas velocity in the riser, and/or to presenting data in terms of overall reactor gas holdup.

Superficial gas velocity has previously been defined in the literature as the gas input rate divided by the riser cross-sectional area. In the work presented here we differentiate between the superficial gas velocity and the "true" superficial gas velocity. The true superficial gas velocity is defined as the sum of the gas input rate from the gas spargers plus the gas flow rate in the downcomer due to recirculation. Excluding the gas recirculation term when calculating the superficial velocity will give higher values of gas holdup for lower values of superficial gas velocity. Thus, reporting data for gas holdup as a function of superficial gas velocity, instead of the true superficial gas velocity, will give different results depending on the amount of gas recirculation in the air-lift. This helps explain the lack of agreement in the reported data in the literature on gas holdup as a function of superficial gas velocity.

The data presented here were analyzed by three methods:

1. Considering the air-lift as a unit, independent of recirculation effects.
2. Considering the air-lift as a unit, including recirculation effects.
3. Considering the independent and interrelated effects of the riser, gas-liquid separator, and downcomer.

It will be shown that only the last approach can satisfactorily represent the reactor's behavior. Forty different geometric configurations and/or reactor operating conditions were examined. Each experiment was conducted over a range of input gas flow rates and repeated at least twice. Each change was examined separately, and in combination with others. The following conditions were varied:

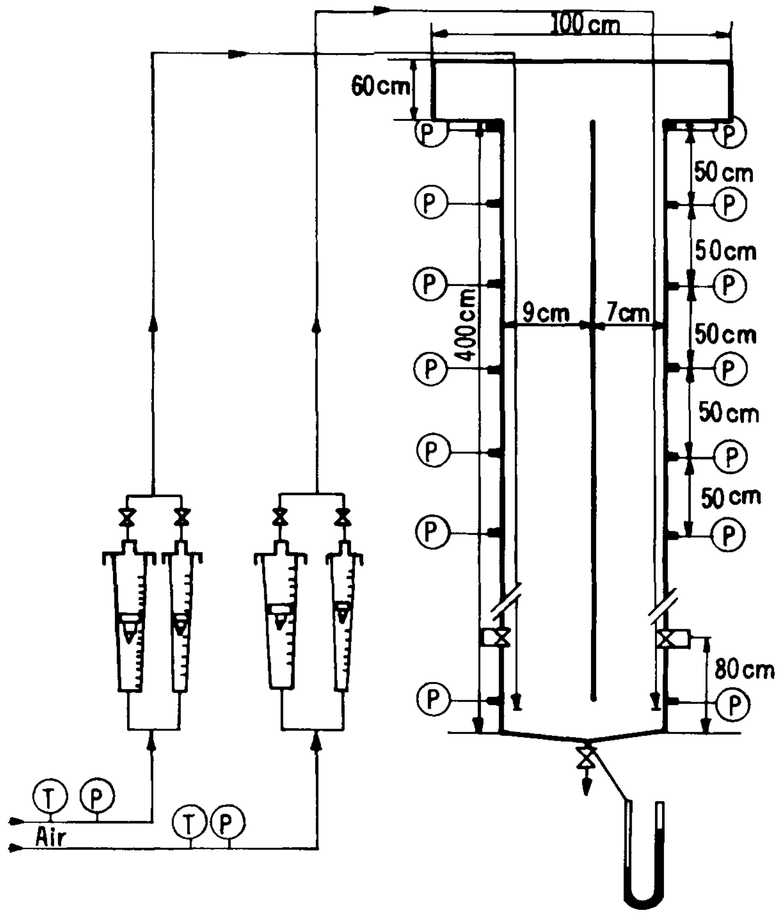
- Riser and downcomer width
- Sparger location
- Gas-liquid separator size and configuration
- Liquid level in the gas-liquid separator.

Experimental

Gas holdup experiments

The air-lift of rectangular cross section used in the experiments was made of transparent polyvinylchloride (PVC). A sketch of the reactor can be seen in Figure 2. The total volume of the reactor is approximately 0.3 m^3. The air-lift has three major sections, riser, downcomer, and gas-liquid separator. The riser and downcomer have a shared center wall. The reactor is designed so that either side can be used as riser or downcomer. The dimensions of the air-lift are given in Table 1.

The gas-liquid separator is connected at the top of the air-lift. The separator is designed in such a way that baffles can be added to modify its size and configuration. The width of the gas-liquid separator was changed from a maximum width of 100 cm



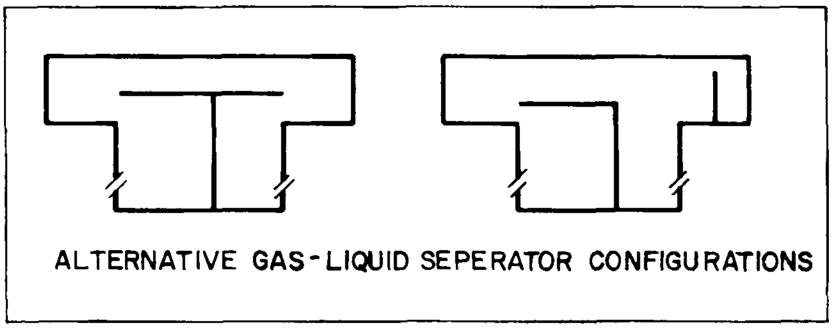


Figure 2. Air-lift experimental set-up.

($C = 0$) to a width of 64 cm ($C = 1$) or 32 cm ($C = 2$). The separator configuration was also modified by using either a straight baffle between the riser and downcomer ($B = |$) or a T-baffle ($B = T$). The various gas-liquid separator configurations are shown in Figure 3. The ungassed liquid level above the baffle between the riser and downcomer was also changed from flush with the top of the baffle ($WL = 0$) to 20 cm above the baffle ($WL = 20$).

Table 1. Main Dimensions of Air-Lift Reactor

Riser/downcomer 1	0.09 m × 0.25 m × 4.0 m
Riser/downcomer 2	0.07 m × 0.25 m × 4.0 m
Gas-liquid separator	1.00 m × 0.25 m × 0.6 m

The local gas holdup was measured by a system of eighteen 1.5 m inverted differential manometers. The manometer fluid was air, and measurements were made in mm of water. The absolute pressure in the air-lift was measured by two mercury manometers open to the atmosphere. One end of the manometer tubes was connected to a manifold, which was connected to a reference point in the air-lift. The other end of each manometer was attached to a sampling port of the air-lift by flexible polyethylene tubing. Starting at the connection zone at the bottom of the riser and downcomer chambers, sampling ports are located at 0.5 m intervals. The sampling ports were 8 mm in diameter.

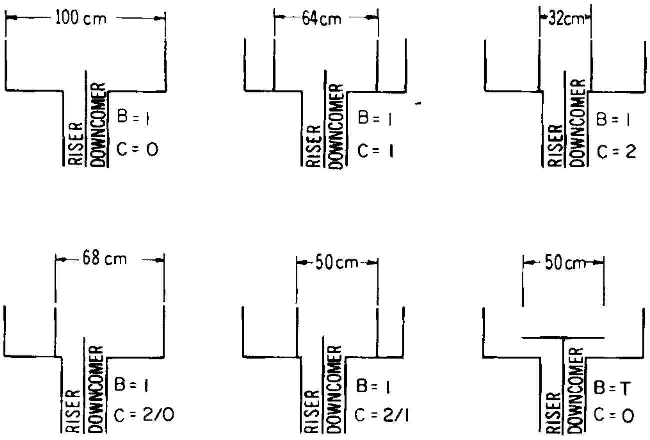


Figure 3. Gas-liquid separator configurations.

They were located on each wall of the riser and downcomer, for a total of six ports per level. The ports at each level were connected so that an average pressure across the cross section was read. There were two sample ports at the base of the air-lift to measure the absolute pressure.

To lessen the effects of pressure fluctuations, inherent to the bubble flow regime, capillary tubes were inserted in the lines. This was at the expense of a delay in the response time of the manometers. To eliminate gas entrapment in the lines the air-lift was always filled with water via the manometer system. Air was then injected into the manometers via a tap at the top of each manometer.

The sparger system allowed air introduction at any level in either the riser and/or the downcomer. The air supply was set by pressure regulators at either two or six atmospheres. Flow rate was controlled by Fisher-Porter rotameters. Three configurations of sparger were used, and orifice size varied from 0.5 mm to 2.5 mm. The spargers were designed so that the gas would be introduced in the direction of the liquid flow.

A Signet "Paddlewheel Flosenser" was installed in the air-lift to enable liquid velocity measurement. Fittings were placed on both the 7 and 9 cm sides of the reactor. The meter was used on the side that was operated as the downcomer during a particular experiment. The measured values were for an average liquid velocity over the cross section.

Salt tracer experiments were conducted to calibrate the velocity meter. A saturated solution of KCl was injected in the gas-liquid separator, at the entrance to the downcomer. Conductivity probes were placed in the downcomer, one 0.5 m from the entrance, and the second 0.5 m from the exit. The experiments were conducted for a wide variety of flow rates and conditions. A calibration curve was thus generated for the meter.

Temperatures were measured both in the liquid and the gas. A thermometer was installed in the air line just upstream of the rotameters. The liquid temperature was measured in the gas-liquid separator.

Local Gas Holdup Calculations

One of the difficulties in developing a model for the air-lift is describing gas holdup and its relationship to gas and liquid velocities. In the air-lift, these velocities are not independent. The gas flow rate determines the liquid flow rate and the gas holdup. Also, because of the different direction and velocity of the liquid flow in the various sections of the air-lift, it is necessary to distinguish each section when writing the momentum balances for the air-lift, in accordance with the orientation of the axial coordinates.

Assuming steady state, one-directional isothermal flow, constant cross-sectional area along the riser or downcomer, constant thermodynamic and transport properties of the pseudofluid across a cross-sectional area, and neglecting the mass transfer effects between the gas and liquid phase, a momentum balance for each section can be written using the pseudohomogeneous model for pressure drop in a two-phase flow regime (Wallis, 1969). The momentum balance on a differential element of the pseudofluid gives an equation that expresses the three components responsible for the pressure drop in the riser and downcomer as friction, acceleration, and gravitation.

Applying the pseudohomogeneous model to the air-lift, and assuming that the gas-wall friction is negligible with respect to the liquid-wall friction, yields the following equations:

Continuity Equation

$$\frac{d}{dz}(\rho_L V_L) = 0 \tag{1}$$

Momentum Equation (relating the gas hold-up and liquid flux to the energy losses in the flow chamber)

$$\frac{dP}{dz} = -\rho_L \frac{d}{dz}[\rho_L V_L^2(1-\phi)] - \frac{(PR)}{A}\tau_W - (1-\phi)\rho_L g \tag{2}$$

writing the momentum equation in terms of the superficial liquid velocity:

$$-\frac{dP}{dz} = \frac{(PR)}{A}\tau_W + \rho_L JL^2 \frac{d}{dz}(1-\phi)^{-1} + \rho_L g(1-\phi) \tag{3}$$

In order to find an implicit expression of the gas holdup, the measured height in the differential manometers, h, is related to the pressure by:

$$(\rho_L - \rho_m) g \frac{dh}{dz} = \frac{dP}{dz} + \rho_L g \tag{4}$$

The following term (Wallis, 1969) was used to express the frictional term:

$$\tau_{WL} = 0.5 C_{fm} \rho_L (JL)(JM) \tag{5}$$

The coefficient C_{fm} is given by Nassos and Bankoff (1967) as:

$$C_{fm} = 0.046(Re)^{-0.2} \tag{6}$$

where the Reynolds number, Re, is given by:

$$Re = \frac{\rho_L JLDe}{(1-\phi)\mu_L} \tag{7}$$

The equivalent diameter is calculated as:

$$De = \frac{2(A)(B)}{(A + B)} \tag{8}$$

where A and B are the sides of the cross section of the chamber.

Combining Eqs. 3, 4, 5, and 6 gives the implicit expression for local gas holdup:

$$\phi = \frac{0.092(JL)^{0.8}(JM)(1 - \phi)^{0.2}}{g(\nu_L)^{-0.2}(De)^{1.2}} + \frac{(JL)^2}{g(1 - \phi)^2}\frac{d\phi}{dz} + \frac{(\rho_m - \rho_L)}{\rho_L}\frac{dh}{dz} \tag{9}$$

In Eq. 9, the term dh/dz is the variation of the readings of the manometers, h, along the height of the column. It is the derivative of the second-order polynomial fitting of the experimental data. Thus, the variation of the gas holdup along the height of the riser or downcomer is evaluated from the second derivative of the experimental data, given by:

$$\frac{d\phi}{dz} = \frac{(\rho_m - \rho_L)}{\rho_L}\frac{d^2h}{dz^2} \tag{10}$$

Equation 10 is an approximation used to evaluate $d\phi/dz$ in Eq. 9. Stein (1979) showed that the contribution of the variations of the frictional and acceleration terms to the changes in holdup along the column were negligible. The terms in Eq. 9, from left to right, represent friction, acceleration, and gravitation, with typical contributions of 24, 1, and 75% of ϕ, respectively. At very high gas flow rates, the acceleration term will rise to 2 or 3% of ϕ (Stein).

The superficial mixture velocity, JM, is defined as the sum of JG plus JL. For any given operational condition, JL, the superficial liquid velocity, is constant along the height of the column. However, JG, the superficial gas velocity, increases from the bottom of the column to the top. JG was calculated, according to Stein, from the measurements in the rotameters and the pressure and temperature for each height.

The mean, or average, gas holdup over the length, L, of the riser or downcomer was calculated from the local gas holdup profile as:

$$\phi = \frac{1}{L}\int_0^L \phi(z)\,dz \tag{11}$$

Superficial Gas Velocity Based on Gas Input Rate Only

The results of gas holdup as a function of the superficial gas velocity for five different gas-liquid separator configurations are presented in Figure 4. The gas recirculation rate was not considered in the calculation of the superficial gas velocity. In this case, the superficial gas velocity is calculated using the influent gas flow rate only. A wide range for gas holdup is seen for any particular superficial gas velocity. The higher gas holdup values

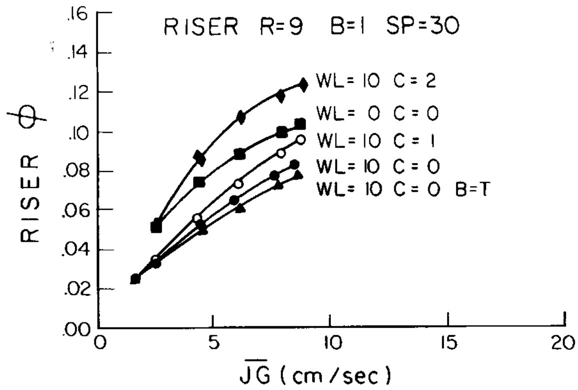


Figure 4. Riser gas holdup as a function of superficial gas velocity.

represent operating conditions with higher gas recirculation rates.

A similar pattern was obtained in all experimental runs, regardless of the side of the reactor used as the riser. This pattern of a wide range of gas holdups for any particular superficial gas velocity cannot be explained by differences in the liquid velocity.

It can be seen that expressing the riser gas holdup as a function of the superficial velocity without considering the gas recirculation does not adequately describe the behavior of the riser. A correlation based on such a superficial gas velocity would only be valid for that particular experimental reactor, with its particular gas-liquid separator design. This helps explain the lack of agreement in the literature of the data for gas holdup experiments.

Calculation of Gas Recirculation Rate

The data were analyzed including the effects of gas recirculation, in order to calculate the true superficial gas velocity. To do this it was necessary to calculate the gas flow rate in the downcomer. This was done by using a two-sparger system to develop a correlation for the superficial gas velocity as a function of the gas holdup in the downcomer.

It was observed that the gas recirculation and gas holdup in the downcomer approached zero with a long fluid residence time in the gas-liquid separator. This was accomplished using a T-baffle between the riser and the downcomer, and maintaining a liquid level 15 cm above the top of the T-baffle. The T-baffle lengthened the fluid flow path in the gas-liquid separator. A second gas sparger was placed in the downcomer, 10 cm from the entrance (top). The sparger in the riser was located 10 cm from the bottom. Each sparger was attached to its own rotameter to allow independent gas flow rate measurement. The sparger used in the downcomer was placed adjacent and parallel to the shared wall between the riser and downcomer. The holes in the sparger faced downward, in the direction of the liquid flow. Gas was first introduced in the riser. After the liquid circulation was fully established, gas was then injected via the downcomer sparger. Experiments were conducted at different downcomer gas flow rates. These experiments will be discussed in more detail below.

Polynomial curve fitting was used on the results of the two-sparger experiments, for stable operating conditions. A second-order polynomial gave an excellent fit of the data. The correlation found was:

$$JGD = 350.79(\phi_D)^2 + 22.02(\phi_D) + 0.011 \qquad (12)$$

Statistical analysis of the curve gave a R^2 value of 0.975 and a standard deviation of 0.144.

The downcomer gas flow rate can be determined in one-sparger and/or two-sparger systems with gas recirculation by measuring the gas holdup and calculating the downcomer superficial gas velocity using the correlation in Eq. 12 where:

$$\text{Downcomer gas flow rate} = QG_D = JGD/A_D \qquad (13)$$

$$\text{True riser gas flow rate} = QG_R = QG_D + QG_S \qquad (14)$$

$$\text{True riser superficial gas velocity} = JGR = QG_R/A_R \qquad (15)$$

$$\text{Gas recirculation rate} = QG_D/QG_R \qquad (16)$$

The downcomer gas flow rate comprises both recycled and freshly entrained air; the latter is relatively much smaller than the former.

Results Using True Riser Superficial Gas Velocity

The correlation expressed in Eq. 12 was used to calculate the true riser superficial gas velocity in the one-sparger systems. Figure 5, which represents data from experiments using the 9 and 7 cm sides of the reactor as the riser, shows the riser gas holdup as a function of the true riser superficial gas velocity for various reactor configurations and operating conditions. A comparison of Figures 4 and 5 clearly shows the difference in expressing gas holdup by excluding or including gas recirculation effects. This illustrates the need to calculate the true riser superficial gas velocity when evaluating the gas holdup.

The deviation of the points from the curves in Figure 5 is accounted for by the difference in the liquid velocities under the different experimental conditions. The points above the curve have a liquid velocity that is 15 to 20% less than the points below the curve.

By comparing the curves in Figure 5, it can be seen that results from the experiments using the 9 cm side of the reactor as the riser and those using the 7 cm side do not give the same curve, even when using the true superficial gas velocity in the calculations. This is due to the difference in the ratio of riser to downcomer cross-sectional areas. This ratio strongly influences the liquid velocities in the air-lift, and the behavior of its sections (Weiland, 1984; Bello et al., 1984).

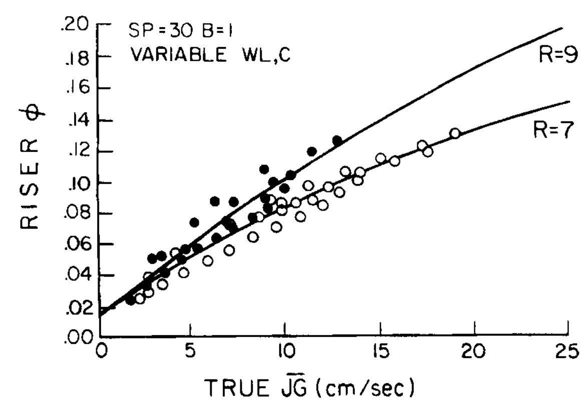


Figure 5. Riser gas holdup as a function of true riser superficial gas velocity.

Correlations for Gas Holdup in the Riser and Downcomer

Correlations were developed that described the gas holdup in the riser and the downcomer in terms of the true riser superficial gas velocity and the downcomer superficial gas velocity, respectively, for the various operating conditions and geometric configurations. A very close fit was obtained using exponential multiple regression on the data points from the gas holdup experiments. The independent variables found significant were the true superficial gas velocity and the ratio of riser equivalent diameter to downcomer equivalent diameter (De_R/De_D). The correlations, Eqs. 17, 18, and 19, are summarized in Table 2.

Correlation 17 illustrates that using the 9 cm side of the air-lift gave larger values of gas holdup then the 7 cm side, and shows the importance of riser/downcomer cross-sectional area ratio. One should bear in mind that difference in the overall pressure drop due to friction in the 9 and 7 cm sides is negligible. Thus the liquid velocities were proportionally lower on the 9 cm side. This concurs with the works of Merchuk and Stein (1981) and Onken and Weiland (1980), which showed that physically reducing the liquid velocity in the riser increases the riser gas holdup. This is also in agreement with the work of Weiland (1984), which showed the influence of riser to downcomer cross-sectional area ratio on gas holdup.

Comparing the above correlations for the riser and downcomer it can be seen that the rate of change of the gas holdup as a function of the superficial gas velocity is similar in the riser and downcomer. Thus, it appears that for the rectangular air-lift the dependence of the gas holdup on the true superficial gas velocity can be approximated by the proportionality, $\phi = b(JG)^{0.7}$, regardless of the direction of flow. The constant, b, depends on the physical properties of the system, direction of flow, and geometric design of the reactor. These correlations can be used in design and scale-up to predict the behavior of the gas holdup in the individual sections (the riser and downcomer). However, these correlations only describe the gas holdup for individual sections. The true superficial gas velocity in the riser cannot be used to describe the gas hold-up in the downcomer, or vice-versa.

Table 2. Gas Holdup Correlations

Reactor Configuration	Experiment Variables	Correlation	
		Riser	
$SP = 30$ $B = \mid$	R, WL, C	$\phi = 0.0176(JGR)^{0.71} (De_R/De_D)^{0.32}$ ($R^2 = 0.94$)	(17)
$R = 9$ $B = T$ $WL = 15$ $C = 0$	SP, GS	$\phi = 0.0136(JGR)^{0.69}$ ($R^2 = 0.99$)	(18)
		Downcomer (using two-sparger system, stable state)	
$R = 9$ $B = T$ $WL = 15$ $C = 0$	JGD	$\phi = 0.0311(JGD)^{0.68}$ ($R^2 = 0.94$)	(19)

Gas-Liquid Separator Parameters and Bubble Flow Configuration in the Downcomer

Visual and photographic observations, as well as the increase in gas holdup in the downcomer, showed that decreasing the residence time of the gas-liquid mixture in the gas-liquid separator increased the gas recirculation rate. The residence time was changed by varying the size and shape of the gas-liquid separator, as well as the liquid level in the separator. Roughly, two major flow configurations could be distinguished in the downcomer. At high liquid velocity, straight bubble flow was observed. Straight bubble flow is determined by bubble circulation where most of the bubbles exhibit a downward flow pattern that is clear and defined, free of flow oscillations. On the other hand, oscillating bubble flow develops mainly at low liquid velocities and short residence time in the gas-liquid separator.

Oscillating flow exhibited essentially three different stages of flow patterns as the system conditions changed toward operating conditions that produced straight flow. The various flow patterns are shown schematically in Figure 6. During operation with short residence time in the gas-liquid separator, it was observed that at very low gas input flow rates (less than 300 cm^3/s, superficial gas velocity less than 2.0 cm/s) there was almost complete separation of the gas and liquid phases in the gas-liquid separator. As the gas flow rate was increased, causing a parallel increase in the liquid velocity, bubbles were entrained in the flow stream to the downcomer. Gas-liquid mixture flow was erratic in the downcomer. A "front" of bubbles moves down the downcomer with increasing gas flow rate. The bubbles exhibit a swirling flow pattern in the front, similar to that seen in bubble columns. However, in contrast to bubble columns, the larger bubbles escape upward, medium size bubbles (approximately 2 to 4 mm dia.) appear at mid-level and stagnate, while very small bubbles (approximately, less than 2 mm dia.) are carried down the column. This produces a statification and stagnation of bubbles according to size along the length of the downcomer, with decreasing bubble size from top to bottom of the column. Under these conditions, calculations show that the downcomer has a low relative gas-liquid velocity (downcomer gas velocity is slightly less than the downcomer liquid velocity). Also, the gas holdup in the downcomer is very high.

As the gas flow rate is further increased the front will reach the bottom of the downcomer, and gas will be recycled to the riser. However, the state of bubble oscillation can persist even though gas is being recycled to the riser. As the liquid velocity increases, the swirling flow pattern will be replaced by intermittent straight and wavy flow patterns in the downcomer. This behavior is due to short, sporadic flow oscillations. Stability improves with increasing influent gas flow rates.

Liquid velocity also increased with increasing influent gas flow rates. It was observed that a minimum downcomer relative gas-liquid velocity (20 to 30 cm/s) was required for the system to shift to straight bubble flow operating conditions. In other words, the downcomer relative gas-liquid velocity had to exceed the terminal rise velocity of a bubble of like size (2 to 8 mm dia.) in stagnant liquid for the bubble to flow down the downcomer. This is in agreement with data presented by other researchers (Wallis, 1969; Weiland, 1984). During straight bubble flow operating conditions, bubble size was rather uniform, with bubble size approximately 4 to 6 mm in diameter.

In the case where the gas-liquid separator is closed to approximately the outer walls of the riser and downcomer, and the unaerated liquid level is even to the top of the baffle between the chambers, there is very high (almost 40%) recirculation of the gas. However, the reactor is very unstable, with sporadic swirling of bubbles in the downcomer for all gas flow rates examined. It was observed that bubble size was less uniform in the downcomer during oscillating bubble flow as opposed to straight bubble flow operating conditions. Also, in the former condition bubbles were generally larger (5 to 8 mm dia.) at higher gas flow rates. It is assumed that there was greater bubble coalescence within the bubble swirls during oscillating bubble flow operation, due to bubble collisions within the bubble swirls.

Figure 7 shows the gas recirculation rate as a function of the riser gas flow rate. It can be seen that recirculation rate remains fairly constant for changing gas flow rates in the riser. However,

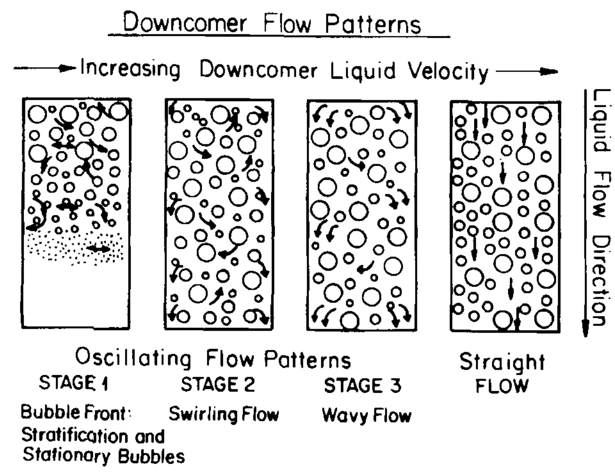


Figure 6. Downcomer flow patterns.

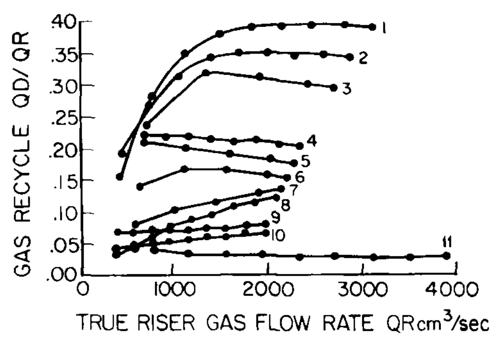


Figure 7. Gas recycle rate as a function of true riser gas flow rate (true superficial gas velocity times riser cross-sectional area).

Reactor Configuration

1.	$R = 7$	$WL = 0$	$C = 2$	$B = \mathrm{I}$
2.	$R = 7$	$WL = 10$	$C = 2$	$B = \mathrm{I}$
3.	$R = 9$	$WL = 10$	$C = 2$	$B = \mathrm{I}$
4.	$R = 7$	$WL = 0$	$C = 1$	$B = \mathrm{I}$
5.	$R = 7$	$WL = 0$	$C = 0$	$B = \mathrm{I}$
6.	$R = 9$	$WL = 0$	$C = 0$	$B = \mathrm{I}$
7.	$R = 9$	$WL = 10$	$C = 1$	$B = \mathrm{I}$
8.	$R = 7$	$WL = 10$	$C = 1$	$B = \mathrm{I}$
9.	$R = 9$	$WL = 10$	$C = 0$	$B = \mathrm{I}$
10.	$R = 7$	$WL = 10$	$C = 0$	$B = \mathrm{I}$
11.	$R = 9$	$WL = 15$	$C = 0$	$B = T$

R, side of air-lift used as riser, cm
WL, liquid level in separator above baffle, cm
C, sections closed in gas-liquid separator
B, baffle configuration

different rates are obtained using different gas-liquid separator configurations and liquid levels. This indicates that the recirculation rate is largely determined by the geometric configuration of the gas-liquid separator and the liquid level in the separator.

Three zones are demarcated in Figure 7. The zones represent operating conditions of oscillating, borderline, and straight bubble flow. The oscillating zone is defined as bubble flow that passed through the three stages of oscillating flow shown in Figure 6, but did not reach straight flow conditions at the gas flow rates of this study. The borderline condition is defined as oscillatory bubble flow at low gas flow rate that shifts to straight bubble flow with increasing input gas flow rate. The straight flow operation zone is distinguished by bubble flow in a straight, well-defined flow pattern for all input gas flow rates studied. A clear division is seen in Figure 7 between the zones of oscillating and borderline conditions, with a jump of approximately 10% in gas recirculation rate. A jump is also seen from the zones of straight flow to borderline flow conditions for low riser gas flow rates. Also, Figure 7 shows that for an oscillatory state, low gas input rates, and short residence time in the separator (curves 1, 2, and 3 with $C = 2$), the gas recirculation rate is not constant. Under these conditions, gas is present in the downcomer, but the liquid velocity is not sufficient to carry any but the very small bubbles (approx. less than 2 mm dia.) down the downcomer to the riser. These are the conditions shown in Figure 6 as stage 1 and the onset of stage 2 during oscillating flow conditions. Figure 7 shows that gas recirculation rates greater than 22% consistently gave oscillating operating conditions.

From approximately 15 to 20% gas recirculation rate, the reactor is in a borderline state. If a straight bubble flow pattern is desired, the reactor should only be operated in this range of recirculation rate at high riser gas flow rates, where the reactor will shift toward stable operation.

Several experiments were conducted to determine the influence of shortening the riser side of the gas-liquid separator only. It was found that the riser side could be shortened so that the separator wall is approximately even with the wall of the riser, without any significant change in the recirculation rate or the holdup in the reactor. This indicates that this is more or less a dead region within the separator.

Shortening the downcomer side of the gas-liquid separator did influence the air-lift behavior. This influence was more pronounced with increasing true riser superficial gas velocity. It can be seen from Figures 7 and 8 that closing the downcomer side of the separator from $C = 0$ to $C = 1$ increases the gas recirculation rate and gas holdup in the air-lift with increasing true riser superficial gas velocity. This is illustrated by the increasing slope of curves 7 and 8 in Figure 7. At low true riser superficial gas velocity there is no significant difference between $C = 0$ and $C = 1$. Further shortening the separator on the downcomer side from $C = 1$ to $C = 2$ strongly increases the gas recirculation rate. Similar results were seen using either side of the reactor as the riser.

Experiments were also conducted changing the liquid level in the gas-liquid separator (from $WL = 0$, to $WL = 20$). The changes of liquid level were examined at various operating conditions and gas-liquid separator configurations. Figures 7 and 8 show that decreasing the liquid level in the separator, thus decreasing the fluid residence time, increases the gas recirculation rate and the gas holdup. Curves 4, 5, and 6 in Figure 7, which fall in the borderline stability zone all represent experiments with $WL = 0$. Also, these curves show that as the riser gas flow rate increases, the gas recirculation rate decreases slightly. This is due to the actual fluid level rising with the increase in the gas fraction in the fluid increasing the fluid residence time in the gas-liquid separator.

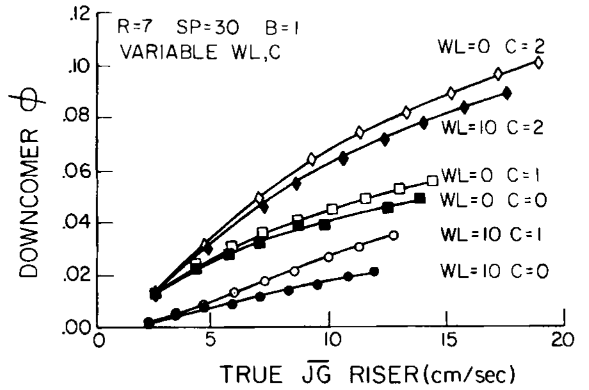


Figure 8. Downcomer gas holdup as a function of true riser superficial gas velocity.

Figure 8 illustrates trends found by combining liquid level and gas-liquid separator configuration changes. It is seen that the data essentially fall in three groups $WL = 10$, $WL = 0$, and $C = 2$. This indicates that until a critical gas-liquid separator size is reached the liquid level will play the predominant role in controlling gas recirculation rate and gas holdup. After the critical size is reached the separator size will have the dominant effect. When the gas-liquid separator size begins to have the dominant effect on reactor operation, the reactor does not attain a straight bubble flow state of operation. This indicates that there is a critical fluid residence time in the gas-liquid separator that can be manipulated to control the gas recirculation rate, gas holdup, and flow configuration for a given input gas flow rate and riser/downcomer configuration.

Effects of the Sparger Height and Configuration

The effect of riser sparger height on gas holdup in the riser was examined. It was found that raising the level of the gas sparger causes a decrease in the gas holdup in the riser. It was also found that the gas holdup and gas recirculation rate were essentially unchanged by further lowering the gas sparger from 30 to 5 cm from the bottom of the riser. The reason for lower gas holdup with higher sparger level is that the only gas present is the gas due to recirculation.

Three sparger configurations and two different sparger orifice sizes were tested. They did not cause a change in the gas holdup or the gas recirculation rate.

Correlation of Riser Liquid Velocity Data

The riser liquid velocity was correlated to the true riser superficial gas velocity. A very close fit ($R^2 = 0.97$) was obtained using exponential multiple regression on all the experimental data. Both sides of the reactor were used as the riser. The independent variables found significant were the true superficial gas velocity and the ratio of riser equivalent diameter to downcomer equivalent diameter. The correlation found for the riser liquid velocity as a function of the true riser superficial gas velocity is:

$$U_L = 33.868\,(JGR)^{0.40}(De_R/De_D)^{-0.41} \qquad (20)$$

This correlation is the same as that found by Merchuk and Stein (1981) for an external-loop air-lift reactor of the same height operated under conditions without gas recirculation. Thus their superficial gas velocity can be seen as the true superficial gas velocity. The correlation differs from that of Bello et al. (1984). This is primarily due to two reasons: the differences in superficial gas velocity calculation, and the fact that our reactor was much taller than the reactors used in their work. We have observed in our laboratories that air-lift reactors of significantly different heights exhibit different behaviors. This is currently under study.

Gas-Liquid Relative Velocity Correlation

The Zuber and Findley (1965) equation for gas velocity, V_G, as a function of the superficial mixture velocity, J_M, was applied to our data. The expression for two-phase, one-directional flow is:

$$V_G = JG/\phi = (C_o)(J_M) + V_d \qquad (21)$$

The drift velocity, V_d, is the gas velocity relative to the superficial mixture velocity, and C_o, the distribution parameter, is a constant dependent on the radial profiles of velocity and holdup in the column. Values of C_o closer to unity represent flatter radial profiles.

Figure 9 shows the riser gas velocity as a function of the riser superficial mixture velocity for the two types of gas-liquid separator baffle configurations used in the experiments. The values represent the gas and mixture velocities for the different operating conditions used with the two baffle configurations. The data for each configuration give a very close fit to a single line. The C_o values obtained for the straight and T-baffles are 1.11 and 1.44, respectively. This indicates flatter radial velocity and gas holdup profiles with the straight baffle. The value for the straight baffle is in agreement to that reported by Hills (1976) for a circular tube, and indicates a slightly parabolic profile, while the value for the T-baffle indicates a sharp parabolic profile. According to Zuber and Findley (1965) the profile will resemble an inverted cone (apex up) when C_o = 1.50. Menzel et al. (1985) also reported the presence of parabolic radial liquid velocity profiles in air-lift reactors. The difference between the two gas-liquid separator configurations is probably related to differences in bubble coalescence in the riser for the two geometries or differences in flow patterns and profiles in the downcomer, which would subsequently influence the riser profiles.

The values for V_d for the straight and T-baffle configurations are 23.8 and 33.5 cm/s, respectively. These drift velocities are within the range reported by Wallis (1969).

Two-Sparger Experiments

Addition of a second sparger, located near the entrance to the downcomer, was studied. This was done to calculate the gas flow rate in the downcomer, as well as to examine the possibility of operating an air-lift with two spargers. A two-sparger system may be advantageous in aerobic fermentation processes with high oxygen demand. In such processes gas recirculation could be controlled by previously described methods, and gas bubbles with depleted oxygen concentrations and relatively high CO_2 could be separated from the liquid in the gas-liquid separator. Fresh gas could then be introduced near the downcomer entrance. Introducing gas near the downcomer entrance, rather than increasing the gas flow rate in a single-sparger system, has the additional advantage of introducing gas against a lower pressure head, thus requiring less energy for gas injection.

Some of the results of the two-sparger experiments using the T-baffle in the gas-liquid separator to minimize the gas recirculation can be seen in Figure 10. The arrows on the figure indicate the points where the reactor switched from unstable to stable operating conditions. Unstable conditions were present whenever the difference between the liquid and gas velocities in the downcomer was less than 25 to 30 cm/s.

During unstable operating conditions, an air pocket was formed at the downcomer sparger. The air-pocket would grow to as much as 15 cm below the sparger, burst, and air would shoot out the top of the downcomer. This would cause a transitory inversion in the flow direction. As the riser sparger gas flow rate increases so does the size of the air pocket at the downcomer sparger. At lower riser sparger gas flow rates the air pocket bursts very rapidly, almost sending a constant stream of very

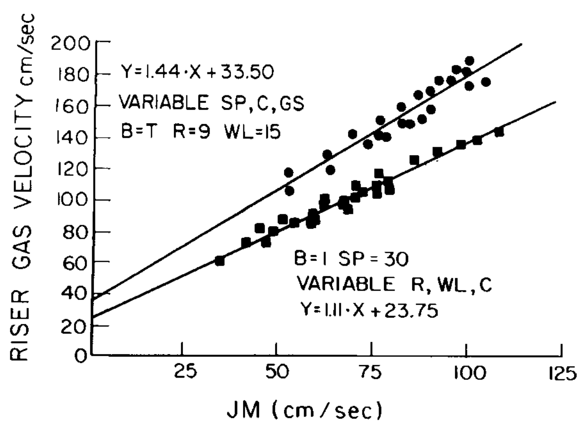


Figure 9. True riser gas velocity as a function of the true riser superficial mixture velocity.
Correlation of Zuber and Findley (1965).

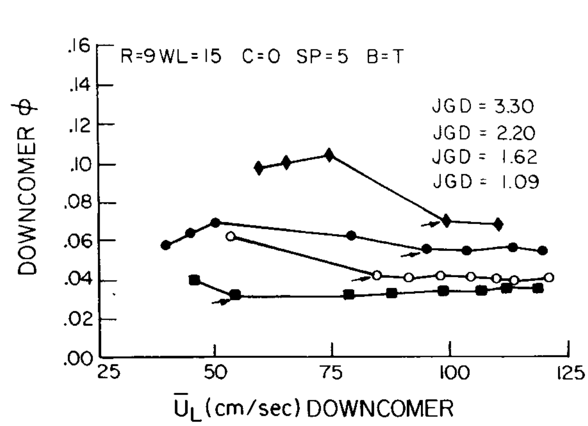


Figure 10. Downcomer gas holdup for various downcomer liquid velocities.
Two-sparger system, variable *JGD*.

large bubbles shooting out of the top of the downcomer. During unstable operating conditions no defined flow pattern was evident in the downcomer. The flow pattern would vary and pulsate in accordance with the rapidity of the bursting of the air-pocket and subsequent flow inversion. The flow pattern was in constant change between sporadic straight flow, bubble swirls and waves, and bubbles momentarily standing still. The sparger served as a center for coalescence by being a disturbance in the fluid streamlines of the downcomer flow path. The gas holdup is higher in the downcomer during unstable operation. However, the erratic behavior of the reactor under these conditions makes it an undesirable operating condition.

During conditions of stable operation, the bubbles form a straight flow pattern that is carried down the downcomer. Figure 10 shows that during stable operation the liquid velocity in the downcomer has a minimal effect on the gas holdup, with almost constant downcomer gas holdup with increasing downcomer liquid velocity. There is a sharp increase in the liquid velocity as stable operating conditions begin. Also, in contrast to the riser, there is a relative uniformity of bubble size and gas velocity in the downcomer.

Several experiments were also conducted that combined gas recirculation and gas introduced in the downcomer. The liquid velocities were sharply reduced throughout the reactor; this caused a subsequent increase in the gas holdup in the reactor. By not degassing the fluid in the gas-liquid separator, the fluid in the downcomer had a lower bulk density, since it included both recirculated gas as well as freshly injected gas. It was found that higher riser sparger gas flow rates (which caused a subsequent increase in the liquid velocity) were necessary to reach stable operation.

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Notation

A = cross-sectional area, cm^2
b = constant
C_{fm} = friction coefficient
C_o = distribution parameter
De = equivalent diameter
g = gravitational acceleration, cm/s^2
h = manometric differential reading, cm
J = superficial velocity, cm/s
JG = superficial gas velocity, cm/s
JGD = downcomer superficial gas velocity, cm/s
JGR = riser superficial gas velocity, cm/s
JL = superficial liquid velocity, cm/s
J_M = superficial mixture velocity, cm/s
L = chamber height, cm
m = mass flow rate, kg/s
P = pressure, kg/cm^2
Pr = chamber perimeter, cm
Q = volumetric flow rate, cm^3/s
Re = mixture Reynolds number
R^2 = R-square value describing closeness of curve fit
U = velocity, cm/s
V = velocity, cm/s
z = axial coordinate, cm

Subscripts

D = downcomer
d = drift
G = gas
L = liquid
M = mixture
m = manometer
R = riser
W = chamber wall
WL = wall-liquid

Greek letters

μ = viscosity, poise
ν = kinematic viscosity, stoke
ρ = density, kg/m^3
τ = shear stress, Pa
ϕ = gas holdup

Symbols for reactor configuration and operating conditions

B = baffle configuration: | straight baffle; T, T-baffle
C = sections closed in gas-liquid separator, max. width 1 m: $C = 0$ = 1 m width; $C = 1$ = 64 cm width; $C = 2$ = 32 cm width
GS = sparger orifice size, mm
R = side of air-lift used as riser, cm
SP = height of sparger from bottom of riser, cm
WL = liquid level in gas-liquid separator above baffle, cm

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