HYDRODYNAMIC BEHAVIOUR IN A NEW GAS-LIQUID-SOLID INVERSE FLUIDIZATION AIRLIFT BIOREACTOR

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A gas-liquid-solid inverse fluidization airlift bioreactor was proposed by combining advantages of external-loop airlift reactors and inverse fluidized beds. This airlift bioreactor comprises a gas-liquid riser, a gas separator, a liquid-solid inverse fluidized downcomer and a bottom connector. The effects of the gas velocity and the particle loading on the gas holdup in the riser and the liquid circulation velocity in the loop were investigated. Liquid-solid inverse fluidization in the downcomer was also studied. The gas holdup in the riser was experimentally found to increase with the increase in the particle loading and in the gas velocity. The liquid circulation velocity decreased with the increase in the particle loading whereas it increased with the gas velocity. It was found that the Richardson and Zaki model fitted experimental data of bed expansion of liquid-solid inverse fluidization in the downcomer well. Based on an energy balance, two hydrodynamic models were proposed to predict the liquid circulation velocity in the present bioreactor. It was shown that the present models gave a good fit to experimental data in this bioreactor.

Keywords: airlift reactor; inverse fluidized bed; liquid circulation velocity; gas holdup; liquid-solid bed expansion.

INTRODUCTION

Three phase (gas, liquid and solid) inverse fluidization was first recognized by Page¹. Its application, especially to wastewater treatment, resulted in a patent by Shimodaira et al.². During the past decades, three phase inverse fluidization has gained attention, as this mode of operation is very promising for wastewater treatment, biochemical, and other industrial applications (Ibrahim et al.³, Farag et al.⁴). In three phase inverse fluidized bed bioreactors, the biofilm thickness can be controlled within a narrow range, and it has been experimentally shown that this kind of bioreactor is very efficient for biological aerobic wastewater treatment at both laboratory scale and pilot-plant scale (Nikolov and Karamanev⁵,⁶).

For conventional three phase inverse fluidization, the solid particles have a lower density than the liquid phase, and gas phase flows upwards while the liquid phase flows downwards. Operating in this kind of counter-current flow mode, the gas bubbles must have a rising motion. If the liquid velocity becomes too large, the bubbles are not able to rise, and they become entrained by the liquid. This maximum liquid velocity corresponds to the limiting velocity of rise of the bubbles (Legile et al.⁷). Experimental results (Fan et al.⁸,⁹, Chem et al.¹⁰) showed that three main flow regimes exist, namely, dispersed bubble regime, bubbling regime and inverse slugging regime in three phase inverse fluidization. At high gas velocities, bubbles tend to coalesce leading to increased agitation and shear, and less effective liquid-solid phase contact.

In three phase inverse fluidized bed bioreactors, the three phases (gas, liquid and solid) make direct contact. It is now well established that gas sparging can damage microorganisms. In addition, the damage to cells was found to result from the appearance of bubbles in the bioreactor (Chalmers and Bavarian¹¹, Cherry and Hulle¹², Trinh et al.¹³, Michaels et al.¹⁴).

Airlift reactors have been known to be efficient contactors for processes involving gas, liquid and solid phases. Their relatively simple mechanical design, low shear rate, high capacity, good mixing, absence of mechanical agitators and low cost make them a versatile type of bioreactor (Douek et al.¹⁵, Snape et al.¹⁶, Dhaoudi et al.¹⁷, Gavrilescu and Tudose¹⁸). Currently, there are mainly two types of airlift reactors, i.e., internal and external airlift reactors.

Nikolov et al.¹⁹ proposed an inverse fluidization airlift bioreactor which was composed of two concentric tubes with air sparged in the inner tube. Garnier et al.²⁰ experimentally investigated effects of the operating parameters on the hydrodynamic characteristics within such a reactor.

External-loop airlift reactors have been used as gas-liquid-solid contacting devices in biotechnology processes, preferentially at large scale, due mainly to their high and readily controllable liquid circulation velocity (Akita et al.²¹). It has also been shown that external-loop airlift reactors have a high efficiency of homogenization and introduce intense mixing, so that they can be applied to the industrial bio-process which involves highly viscous fluid (Saunder et al.²², Siegel and Robinson²³).

In the present paper, the advantages of external airlift reactors and inverse fluidized beds are combined to construct a novel external-loop bioreactor, the so-called
gas-liquid-solid inverse fluidization airlift bioreactor (GLS-IFAB). By combining the turbulent gas-liquid contact to the riser, and liquid-solid contact to the downcomer, the potential problems such as cell death by bubbles can be eliminated in bioreactors. The hydrodynamic behaviour in GLS-IFABs has been studied experimentally. Mathematical models have also been developed for the relationship between liquid circulation velocity and superficial gas velocity and other operating parameters.

**LIQUID CIRCULATION VELOCITY**

A fundamental parameter characterizing GLS-IFAB operation is the velocity of bulk liquid circulation in the bioreactor, which can be quantified by the superficial liquid velocity in the riser. This parameter is a function of gas velocity, solids size, density and reactor geometry, etc. To develop the model of liquid circulation velocity in the bioreactor, the following main assumptions are made:

1. the airlift reactor is operated at a stable flow condition;
2. the volume change of the gas phase due to the static pressure variation in the riser follows the isothermal expansion or contraction of an ideal gas;
3. the amount of gas entrapped in the downcomer is negligible compared to that present in the riser;
4. compared to other dissipation terms, energy losses due to skin or wall friction in the riser and the downcomer are negligible for Newtonian low viscosity fluids (Chisti);
5. mass transfer between gas and liquid phases is negligible.

Based on the energy balance approach used by Chisti, the driving force for liquid circulation in the bioreactor is produced by the change in energy as gas bubbles rise and expand up the riser. As shown in Figure 1, the energy balance over the loop in the bioreactor is:

\[ E_{in} = E_r + E_i + E_d + E_b \]  

in which,

- \( E_{in} \) — energy input due to the isothermal gas expansion;
- \( E_r \) — energy loss in the riser;
- \( E_i \) — energy loss at the top section;
- \( E_d \) — energy loss in the downcomer;
- \( E_b \) — energy loss at the bottom section.

**In the Riser**

Using the liquid phase in the riser as the control volume, the energy balance for the riser is given by:

\[ E_{in} = E_r - (\dot{v}_l \rho_l + \dot{v}_g \rho_g)g \left( H_p U_p S_r + \rho_l g H_p U_p S_r \right) \]  

The second and third terms on the right-hand side of equation (2) represent the pressure energy loss and potential energy gain in the riser, respectively.

Compared with the density of liquid phase \( \rho_l \), the density of gas phase \( \rho_g \) is negligible, so equation (2) is rewritten as:

\[ E_{in} = E_r + \dot{v}_g \rho_g g H_p U_p S_r \]  

**In the Top Section**

The energy loss due to expansion, contracting and reversal of the flow in the top section is given:

\[ E_i = K_i \left( U_p \rho_l S_r \right)^{1/2} \frac{V^2}{2} \left( 1 - \frac{\rho_g}{\rho_l} \right) U_p S_r \]  

where \( K_i \) is the pressure drop through the downcomer.

**In the Downcomer**

By analogy with pipe flow, the energy loss due to particles in the downcomer can be written as:

\[ E_d = U_{id} S_d \Delta P \]  

where \( \Delta P \) is the pressure drop through the downcomer. Because the continuity relationship for incompressible flow governs the flow between the riser and the downcomer, we have:

\[ U_{id} S_d = U_p S_r \]  

Substituting the above equation into equation (5), we obtain:

\[ E_d = U_p S_r \Delta P \]  

In the downcomer, with the increase in the liquid velocity, there exist three flow regimes, namely, packed bed, inverse fluidized bed and transport regimes. In the present paper, particles were not circulated in the bioreactor. Therefore, we only discuss energy loss in the packed bed and inverse fluidized bed regimes, respectively.

**Energy loss in the packed bed regime**

Pressure drop through beds of uniform particles can be satisfactorily correlated by the Carman-Kozeny equation applicable to laminar as well as turbulent flow:

\[ \Delta P = \frac{\alpha \rho_l (1 - \varepsilon_{id}) H_p U_{id}^2}{\varepsilon_{id}^3} \left( 5 Re^{-1} + 0.4 Re^{-0.1} \right) \]  

where the definition of the Reynolds number \( (Re) \) employed by Coulson and Richardson in equation (8) can be adapted for the airlift:

\[ Re = \frac{\rho_l U_{id}}{\alpha (1 - \varepsilon_{id}) \mu_l} \]  

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Combining equations (6), (7) and (8) gives energy dissipation $E_d$ in the packed bed regime in the downcomer:

$$E_d = \frac{\alpha \rho_l (1 - e_{ld}) H_p U_{lr}^3 S_r^3}{\rho_l S_d^2} \times (5 Re^{-1} + 0.4 Re^{-0.1})$$

(10)

Energy loss in the inverse fluidized bed regime

The inverse fluidized bed regime in the downcomer can be divided into two regions, i.e., liquid-solid inverse fluidized region and freeboard region. In the liquid-solid inverse fluidized region, the energy balance in the downcomer is given by:

$$E_d = \rho_l g H_{ld} U_{ld} S_d - (e_{ld} \rho_l + e_s \rho_s) g H_{ld} U_{ld} S_d$$

(11)

The two terms on the right-hand side of the above equation represent the potential energy loss and the pressure energy gain in the downcomer, respectively.

Based on the definition of $e_s$:

$$e_s H_{ld} = e_{s0} H_p$$

(12)

Substituting equations (6) and (12) into equation (11) gives:

$$E_d = (\rho_l - \rho_s) e_{s0} g H_p U_{lr} S_r$$

(13)

In the Bottom Section

Similar to the flow in the top section, the energy loss in the bottom section is given as:

$$E_s = K_b (U_{ld} \rho_l S_d) \frac{U_{lr}^2}{2} = \rho_l K_b \frac{S^3 U_{lr}^3}{2 S_d}$$

(14)

Finally, when the downcomer is operated in the packed bed regime, combining equations (1), (3), (4), (10) and (14) yields:

$$U_{lr}^2 = \frac{2 e_s g H_p}{K_b \left(\frac{S}{S_d}\right)^2 + \left(\frac{S}{S_d}\right)} (10 Re^{-1} + 0.8 Re^{-0.1})$$

(15)

Combining equations (1), (3), (4), (13) and (14) gives the following equation in the condition that the downcomer is operated in the fluidized bed regime:

$$U_{lr}^2 = \frac{2 g \left[e_s H_p - \left(1 - \frac{e_{s0}}{\rho_l}ight) e_{s0} H_p\right]}{K_b \left(\frac{S}{S_d}\right)^2}$$

(16)

Equations (15) and (16) can be applied to predict the liquid circulation velocity in GLS-IFABs. The undetermined parameters ($K_b$ and $K_t$) will be discussed in the section on 'liquid circulation velocity'.

**EXPERIMENTAL**

The gas-liquid-solid inverse fluidized airlift bioreactor used in the present work is shown in Figure 2. The reactor was made of Plexiglass materials with an approximate working volume of 15 l. The major dimensions of the reactor are summarized in Table 1. All experiments were carried out at room temperature (around 25°C) and atmospheric pressure.

Air was introduced through a circular perforated plate sparger containing 30 holes of 1 mm diameter. The plate was mounted at one end of a pipe centrally located along the riser axis. A stainless steel screen was located between the gas-liquid separator and the downcomer. This screen prevented the solid particles from rising from the downcomer to the gas-liquid separator.

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Tap water and air were used as the liquid and gas phases in the experiments. Perfectly spherical and hollow polyethylene particles with a diameter of 10 mm and a density of 388 kg m$^{-3}$ were used as the solid phase. The inverse fluidized bed porosity was determined from the direct experimental measurement of the fluidized bed height. The gas flow rate was measured by a calibrated rotameter. The gas holdup was obtained by inverted U-type manometers connected to pressure taps located at different axial positions in the riser. To make it easier to understand this measurement, the calculation of gas holdup is explained in the Appendix.

The tracer response technique was applied to obtain the liquid circulation velocity in the bioreactor. Saturated NaCl solution was used as the tracer. The injection point was located in the downcomer and the two conductivity probes were introduced at different locations along the downcomer. The liquid circulation velocity was determined by using the distance over the difference in response time between the two probes. In the experiments, in order to obtain good experimental signals, the injection point was kept from the liquid-solid fluidized section. The output signals from two conductivity probes were processed by a computer.

RESULTS AND DISCUSSION
Gas Holdup

Figure 3 shows the effects of the superficial gas velocity and particle loading on the gas holdup in the riser. It can be seen that for any particle loading, the gas holdup in the riser increases with increasing superficial gas velocities, which is in accordance with the experimental results obtained by other researchers.

It can be seen from Figure 3 that for a given particle loading, the increase of the gas holdup in the riser becomes less steep with the increase in the superficial gas velocity at high gas velocity. This may be explained by the fact that gas bubbles tend to coalesce with the increase in the superficial gas velocity. Therefore, the appearance of larger bubbles results in the decrease in the bubble residence time in the riser. Moreover, as shown later in Figure 7, the liquid circulation velocity slightly increases with the increase in the superficial gas velocity due to the flow resistance through the packed bed region in the downcomer, then sharply increases with the gas velocity when solid particles are inversely fluidized. The increase in the liquid circulation velocity in the riser will enhance the rising of bubbles, resulting in the reduction of the gas holdup in the riser. Therefore, the change of the liquid circulation velocity with the superficial gas velocity in GLS-IFABs contributes to the change of the gas holdup in the riser.

From Figure 3, it can be seen that at a fixed superficial gas velocity, the gas holdup in the riser increases with increasing particle loading. The experimental data shown later in Figure 7 indicate that the liquid circulation velocity decreases with the increase in particle loadings due to higher flow resistance in the downcomer. When the superficial gas velocity is kept unchanged, the reduction of the liquid circulation velocity results in the increase in the gas holdup in the riser. Moreover, Figure 3 indicates that compared with the airlift reactor without solid particles, the gas holdup in the riser has a higher value at the same operating conditions, which means that GLS-IFABs provide better gas-liquid contact.

The axial distributions of gas holdup in the riser were measured in the three sections and shown in Figure 4. It was found that the gas holdup had very uniform distributions along the axial direction for the whole experiments in the present study. The average deviation is less than 5% in each operating condition.

From the experiments, it was observed that the separator in the present study effectively disengaged the gas from the liquid, leaving almost bubble-free liquid flowing into the downcomer.

<table>
<thead>
<tr>
<th>Item</th>
<th>Dimensions, m</th>
</tr>
</thead>
<tbody>
<tr>
<td>Height of the riser</td>
<td>1.3</td>
</tr>
<tr>
<td>Internal diameter of the riser</td>
<td>0.06</td>
</tr>
<tr>
<td>Height of the downcomer</td>
<td>1.3</td>
</tr>
<tr>
<td>Internal diameter of the downcomer</td>
<td>0.06</td>
</tr>
<tr>
<td>Gas separator</td>
<td>0.5 × 0.25 × 0.2</td>
</tr>
<tr>
<td>Distance between the riser and the downcomer</td>
<td>0.33</td>
</tr>
</tbody>
</table>

Figure 3. Effects of the superficial gas velocity and particle loading on the gas holdup in the riser.

Figure 4. Axial distributions of the gas holdup in the riser for different particle loadings at $U_g = 0.03$ m$^{-1}$.
The combined effect of the superficial gas and liquid velocities on the gas holdup in the riser can be taken into account by using the Zuber and Findlay model which has been widely used in studies of gas-liquid two-phase flow to determine gas holdup in the airlift reactor. It is based on the relation between the gas linear velocity and the mean velocities of gas and liquid phases:

\[ V_g = C(U_g + U_l) + U_{bt} \]  

where \( C \) and \( U_{bt} \) are coefficients taking into the non-uniform velocity and gas holdup profiles. \( C \) varies between 1.0 for a flat profile and 1.5 for a laminar flow profile.

The following correlation was obtained by fitting the Zuber and Findlay model with experimental data in the present study:

\[ V_g = \frac{U_g}{v_g} = 1.04(U_g + U_l) + 0.32 \]  

Table 2 shows some values for the parameters in Zuber and Findlay model, which were found by other researchers for external loop airlift reactors. The value of the distribution coefficient (1.04) obtained in the present study demonstrates that there should be relatively uniform radial distributions of flow and gas holdup profiles in GSL-IFABs. Abashar et al.\(^{25}\) reported three distinct flow regimes, namely, homogeneous, transition and heterogeneous regimes in external-loop airlift bioreactors. From the experimental data shown above, the gas-liquid flow in the riser of the GLS-IFAB was operated in the homogeneous regime in the present study.

Table 2. Parameters in Zuber and Findlay model.

<table>
<thead>
<tr>
<th>Reference</th>
<th>Riser diameter</th>
<th>C</th>
<th>( U_{bt} ) ms(^{-1} )</th>
<th>System</th>
</tr>
</thead>
<tbody>
<tr>
<td>Merchuk and Stein(^{30})</td>
<td>0.14</td>
<td>1.03</td>
<td>0.33</td>
<td>Air-water</td>
</tr>
<tr>
<td>Verlaan(^{31})</td>
<td>0.20</td>
<td>1.20</td>
<td>0.26</td>
<td>Air-water</td>
</tr>
<tr>
<td>Nicol and Davidson(^{32})</td>
<td>0.24</td>
<td>1.13</td>
<td>0.28</td>
<td>Air-water</td>
</tr>
<tr>
<td>Young et al.(^{33})</td>
<td>0.19</td>
<td>1.02–1.13</td>
<td>–</td>
<td>Air-water</td>
</tr>
<tr>
<td>Abashar et al.(^{25})</td>
<td>0.225</td>
<td>1.07</td>
<td>0.538</td>
<td>Air-water</td>
</tr>
</tbody>
</table>

Bed Expansion of Inverse Fluidization

The bed expansion in the downcomer was investigated in the present paper. The experimental data of bed expansion for liquid-solid inverse fluidization are presented in Figure 6.

There are different models for the correlation of bed expansion with the superficial liquid velocity. Among all these correlations, the model proposed by Richardson and Zaki\(^{34}\) has widely been used since it is simple and in good agreement with experimental data:

\[ \frac{U_l}{U_i} = e^{-n} \]  

Figure 5. Comparison between experimental and calculated gas holdup in the riser.

where \( n \) is the Richardson–Zaki index, and can be respectively determined from the following relations:

\[ n = \left( 4.4 + 18 \frac{d_p}{D} \right) Re_t^{-0.1} \quad 1 < Re_t < 200 \]  

\[ n = 4.4 Re_t^{-0.1} \quad 200 < Re_t < 500 \]  

\[ n = 2.4 \quad Re_t > 500 \]  

and \( U_i \) is the superficial fluid velocity at \( v_i = 1 \), which can be calculated from the following equations\(^{35}\):

\[ \log U_i = \log U_{pt} - \frac{d_p}{D} \]  

where \( U_{pt} \) is the particle terminal velocity.

Fan et al.\(^{8}\) first proposed two correlations for bed expansion of liquid-solid inverse fluidization. Based on the Richardson and Zaki model, they obtained the following equations to determine \( n \) in the model:

\[ n = 15 Re_t^{-0.35} \exp \left( 3.9 \frac{d_p}{D} \right) \quad 350 < Re_t < 1250 \]  

\[ n = 8.6 Re_t^{-0.2} \exp \left( -0.75 \frac{d_p}{D} \right) \quad Re_t > 1250 \]  

Karamanev and Nikolov\(^{36}\) experimentally confirmed that the expressions for the predictions of the exponent, \( n \), in the Richardson and Zaki model was effective to describe bed
expansion of liquid-solid inverse fluidization, but proposed the following two equations to predict the superficial velocity, \( U_i \), in the model:

\[
\log Re_i = -1.814 + 1.347\log N_D - 0.1243(\log N_D)^2 \\
+ 0.00634(\log N_D)^3 \quad 12.2 < Re_i < 130
\]

\[(23a)\]

\[
Re_i = \frac{N_D}{0.95} \quad Re_i > 1250
\]

\[(23b)\]

where \( N_D \), the Best number, is defined as:

\[
N_D = \frac{4\rho_g(\rho_i - \rho_g)d_p^3}{3\mu_l^2}
\]

\[(24)\]

As mentioned above, it can be inferred that experimental data plotted as a function of \( \ln \epsilon_{id} - \ln U_i \) can be approximated by a straight line if equation (19) is suitable for describing bed expansion of liquid-solid inverse fluidization. When the Richardson and Zaki correction is used to simulate experimental data, the slope of the line \( \ln \epsilon_{id} - \ln U_i \) is equal to \( 1/n \) and the intercept of the line with the horizontal axis at \( \epsilon_{id} = 1 \) gives \( \ln U_i \). The values of \( n \) and \( U_i \) can be obtained by linear regression of the experimental data shown in Figure 6. Table 3 shows the values of \( n \) and \( U_i \) obtained from the experiments (\( N_D = 8.1 \times 10^6 \) in the present system) and different models. From \( U_i \) and equation (21), the terminal rising velocity of particles, \( U_{pt} \), is equal to 0.54 ms\(^{-1}\). Surprisingly, it was found, that the Richardson and Zaki model predicted the experimental data better.

The Minimum Fluidization Velocity in the Downcomer

In the downcomer, both equations (10) and (13) can be applied for predicting the liquid circulation velocity at the initially fluidized point. Therefore, the following equation can be obtained from equations (10) and (13) to predict the minimum fluidization velocity, \( U_{mf} \), in the liquid-solid inverse fluidized bed:

\[
U_{mf}^2(5Re_{mf}^{-1} + 0.4Re_{mf}^{-0.1}) = \frac{(\rho_i - \rho_g)(1 - \epsilon_{id})^3 g}{\alpha\rho_l}
\]

\[(25)\]

where:

\[
Re_{mf} = \frac{U_{mf}\rho_l}{\alpha \rho_{g}\mu_l}
\]

| Table 3. Values of \( n \) and \( U_i \) obtained from the present experiments and different models. |
|------------------------------------------|-----------------|
| Model 1 \[(Richardson and Zaki)^{34}\] | 2.4 | 0.3 |
| Model 2 \[(Fan et al.\]^{35}\] | 1.54 | 0.3 |
| Model 3 \[(Karamanev and Nikolov\]^{36}\] | 2.4 | 0.2 |
| Exp. values in the present study | 2.54 | 0.37 |

Figure 7. Comparison of experimental and calculated minimum fluidization velocities.

The experimental data for minimum fluidization velocities under different operating conditions are shown in Figure 7. Figure 7 also shows a comparison between experimental and predicted values of the minimum fluidization velocities, and a good agreement was obtained.

Liquid Circulation Velocity

Figure 8 shows the effect of the gas velocity on the liquid circulation velocity with various solid particle loadings. With the increase in the gas velocity, the liquid circulation velocity increases, due to a larger driving force between the riser and the downcomer. The authors investigated the change of the liquid circulation velocity both in the packed bed regime and in the inverse fluidized bed regime. At a fixed loading, with the increase in the gas velocity, the downcomer was initially operated in the packed bed regime,

![Figure 8](image-url)
then entered the inverse fluidized bed regime. From Figure 8, it can be seen that initially, the liquid circulation velocity slightly increases with the increase in the gas velocity in the packed bed regime, then sharply increases in the fluidized bed regime. This experimental phenomena can be explained by the fact that equation (10) shows that the energy dissipation has a power law relationship with the liquid circulation velocity in the packed bed regime, but the energy dissipation expressed in equation (13) has a linear relationship with the liquid circulation velocity in the inverse fluidized bed regime.

For a given gas velocity, as shown in Figure 8, the liquid circulation velocity decreases with the increase in solid particle loading. This is due to the fact that heavier solid particle loading results in higher pressure drop in the downcomer, hence lower liquid circulation velocity.

Equations (15) and (16) were used to predict the liquid circulation velocity at different operating conditions. The values of $K_b$ and $K_t$ in the models are dependent on the reactor geometry. In the present study the energy dissipation coefficients in the bottom and top sections of the bioreactor were estimated as $K_t = K_b = 1.8$ (Streeter and Wylie), which have been confirmed by Veraan and Abashar in external-loop airlift reactors. In the present study the internal diameter of the riser is the same as that of the downcomer, i.e., $S_r = S_d$. Therefore, liquid circulation velocities under different operating conditions can be predicted using equations (15), (16) and (18). Figure 9 shows a comparison between experimental and predicted values of the liquid circulation velocity in the bioreactor. It has clearly been shown in Figure 9 that the present models can predict the liquid circulation velocity in GLS-IFABs with satisfactory accuracy (the deviation is $\pm 25\%$).

CONCLUSIONS

A gas-liquid-solid inverse fluidization airlift bioreactor was proposed by combining advantages of external-loop airlift reactors and inverse fluidized beds. Experiments were conducted to investigate the gas holdup in the riser, bed expansion in the downcomer and the liquid circulation velocity in the bioreactor. The following results were obtained:

1. The gas holdup in the riser increases with the increase in the superficial gas velocity and particle loading, and the two-phase drift model of Zuber and Findlay satisfactorily fits the gas holdup in the GLS-IFAB.

2. It was found that the Richardson and Zaki model gave a good fit to experimental data of bed expansion of liquid-solid inverse fluidization in the downcomer. A model was obtained to predict the minimum fluidization velocity for liquid-solid inverse fluidization, and fitted experimental data well.

3. The liquid circulation velocity increases with the increase in the gas velocity whereas it decreases with particle loading. Based on energy balance, two hydrodynamic models were proposed to predict the liquid circulation velocity in the bioreactor. It was shown that the present models fitted the experimental data well.

APPENDIX

The Calculation of Gas Holdup in the Riser

The experimental method of measuring gas holdups is illustrated in Figure A1. At any point in the riser, $C$, one can obtain:

$$H_1 \rho_g = H_0 (e_l + e_g \rho_g)g + H_2 \rho_g$$

Since $\rho_l \gg \rho_g$ and $e_g + e_l = 1$, the following equation can be obtained:

$$e_g = \frac{AH}{H_0}$$

where $AH$ represents the level difference in the inverse U-type manometer and $H_0$ the distance between two measured points. These two parameters can be measured in the experiments.

NOMENCLATURE

- $C$: constant in equation (17)
- $D$: diameter of the column, m
- $d_p$: particle diameter, m
- $E$: energy loss in the reactor, W
- $E_m$: energy input due to isothermal gas expansion, W
- $g$: gravitational acceleration, $ms^{-2}$
- $H$: height, m
- $H_p$: packed bed height, m
- $K$: energy loss coefficient
- $n$: exponent in equation (19)
- $N_D$: Best number defined by equation (24)
- $AP$: pressure drop, Pa
- $Re$: Reynolds number defined by equation (8)
- $Re_{mf}$: minimum fluidization Reynolds number defined by equation (25)
- $Re_p$: particle Reynolds number defined by $\frac{\rho \mu}{\rho_p \mu_p}$
- $S$: cross-sectional area, $m^2$
- $U$: superficial velocity, $ms^{-1}$
- $U_{bt}$: bubble terminal velocity, $ms^{-1}$
- $U_{sf}$: superficial fluid velocity at $\eta = 1$ in liquid-solid two phase, $ms^{-1}$
- $U_{mf}$: minimum fluidization velocity, $ms^{-1}$
- $U_{pt}$: particle terminal velocity, $ms^{-1}$
- $V$: linear velocity, $ms^{-1}$

Greek letters

- $\varepsilon$: holdup
- $\alpha$: the surface area per unit volume of the particles, $m^2 m^{-3}$
**REFERENCES**


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