AXIAL DISPERSION CHARACTERISTICS OF THREE PHASE FLUIDIZED BEDS

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The axial dispersion characteristics of liquid phase in two (gas-liquid, liquid-solid) and three (gas-liquid-solid) phase fluidized beds were studied in a 14.5 cm-I.D. column.

The effects of liquid velocity (2-13 cm/s), gas velocity (0-12 cm/s), liquid viscosity (1-27 cP), surface tension (38.5-76 dyn/cm) and particle size (1.7-6.0 mm) on axial dispersion of liquid phase were examined.

Liquid-phase axial dispersion in terms of Peclet numbers were correlated empirically by equations involving the ratio of fluid velocities and the ratio of particle-to-column diameters. The dispersion coefficients increased with gas flow rate, liquid surface tension and viscosity. However, liquid viscosity generally reduced the coefficient in beds of smaller particles at higher gas rates.

Introduction

In three-phase fluidization, particles are fluidized by the cocurrent flow of liquid and gas. The gas forms a discrete bubble phase and the liquid a continuous phase containing solid particles.

Three-phase fluidized beds can be used in reaction systems in which the liquid is not readily vaporized, the gas is not easily liquefied, and a catalyst is required to enhance the rate of reaction. Present industrial applications include catalytic hydrogenation and desulphurization of petroleum products, production of ammonium bisulfite, biochemical processes and a coal liquefaction (H-coal) process. To date, studies on bed expansion, liquid and gas hold-ups, gas bubble characteristics and fluid mixing have been reported. Recent literature has been reviewed by Østergaard,18,20 Kim and Kim,9 and Epstein.6

Afschar and Schügerl1 reported the intensity of mixing in liquid and gas phases of 13.5 cm-diameter fluidized beds of five different kinds of particles having mean diameters in the range of 0.04-0.5 mm. The intensity of liquid-phase mixing increased with liquid and gas velocities. The presence of solid particles increases the relative intensity of mixing in liquid phase.

Michelson and Østergaard15 investigated liquid- and gas-phase mixing in two- and three-phase fluidized beds. They employed beds of 15.24 cm and 21.6 cm diameter using air, water and four different sizes of glass beads ranging from 0.25 to 6 mm in diameter. They observed that particle size played an
important role in governing the intensity of axial mixing in three-phase fluidized beds. While beds of 6-mm glass beads were characterized by a very low degree of liquid-phase mixing, 0.25- and 1.0-mm particle beds were characterized by a high degree of mixing. However, in beds of 3-mm glass beads, the degree of mixing in the liquid phase was very low and decreased with increasing gas velocity. At increased gas velocity bubble coalescence became important and a sharp increase in mixing was observed.

Vail et al.24) studied turbulent mixing of the liquid phase in two- and three-phase fluidized beds of 114.3 cm diameter. Air was used as the gas phase, tap water as the liquid phase, and spheres having a diameter of 0.87 mm as the solid phase. They found that an increase in gas velocity intensified mixing whereas an increase in liquid velocity reduced it.

Kim et al.10) studied liquid-phase mixing in three-phase fluidized beds of 2.5-mm gravel and 6-mm glass beads. Axial mixing increased with either gas or liquid velocity.

Muroyama et al.16) measured axial mixing of liquid phase in three-phase beds of six different particle sizes (0.216–6.9 mm). Correlations were derived from the experimental data for each flow regime.

El-Temtamy et al.5) determined the axial dispersion of liquid phase in three-phase fluidized beds of 5 cm internal diameter. Glass beds of 0.45, 0.96, 2 and 3 mm diameter were fluidized with tap water and air. The particle Peclet number was correlated in terms of particle Reynolds number and column diameter. Axial mixing was found to increase with increasing gas rate. Axial dispersion did not vary greatly with particle size, and was higher for three-phase fluidized beds than for liquid-gas systems.

Most studies of fluid mixing have been carried out using water as the liquid phase and air as the gas phase in small-diameter beds. The present work was undertaken to obtain information about the effect of liquid properties on liquid dispersion or mixing in three-phase fluidized beds since no other such information has been published so far. In the present study, the effects of liquid and gas velocity, particle size and the properties of liquid on axial liquid mixing have been determined in order to provide prerequisite knowledge in designing a multiphase flow system.

1. Experimental

Experiments were carried out in a relatively large plexiglass column 250 cm high and 14.5 cm in diameter, as shown schematically in Fig. 1. The main section of the column was constructed from two pieces of 14.5 cm-diameter × 125 cm-high × 0.5 cm-thick plexiglass cylinder flanged together. An approximately constant dynamic liquid level was maintained in the column by means of a concentrically mounted outlet header which acted as a liquid weir. A stainless steel screen was attached in the top of the column to prevent loss of particles.

The solid particles were supported on a perforated plate which contained 237 evenly spaced holes of 3 mm-diameter and which served as the liquid distributor. The grid was situated between the main column section and a 35.5 cm-high × 14.5 cm-diameter stainless steel distributor box into which liquid was introduced through a 2.54-cm pipe. The liquid was pumped from a 600-liter reservoir through a 2.54-cm pipe. Its flow rate was measured with one of two calibrated rotameters and regulated by means of globe valves on the feed and bypass lines.

Oil-free compressor air was fed to the column through a pressure regulator, filter and calibrated rotameter. It was admitted to the bed through four 6.35-mm perforated feed pipes drilled horizontally in the grid. The pipes were evenly spaced across the distributor plate.

Pressure taps were mounted flush with the wall of the column at 180-mm height intervals. The static pressure at each of these points was measured with a liquid manometer.

The solids used were either 1.7-mm, 3.0-mm or 6.0-mm glass beads with a density of 2.5 g/cm³. The unfluidized bed height in the liquid-solid and three-phase experiments was 270 mm (6.0 kg).

Details of the fluidizing liquid employed are given in Table 1. To determine the effect of liquid viscosity on liquid phase dispersion, solutions of commercial-grade glycerol and of carboxymethyl cellulose (CMC) in tap water were employed. Technical-grade methanol–water and Triton X-100 water solutions were
Table 1. Properties of liquids (at 19.5°C)

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Density [g/ml]</th>
<th>Viscosity (mNs/m²) or consistency index [mNs/m²]</th>
<th>Behaviour index [n]</th>
<th>Surface tension [dyn/cm]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>1.00</td>
<td>1.00</td>
<td>1.00</td>
<td>73.0</td>
</tr>
<tr>
<td>Water-CMC (0.2% wt%)</td>
<td>1.001</td>
<td>7.85*</td>
<td>0.946</td>
<td>73.6</td>
</tr>
<tr>
<td>Water-CMC (0.4 wt%)</td>
<td>1.002</td>
<td>27.12*</td>
<td>0.836</td>
<td>76.01</td>
</tr>
<tr>
<td>Water-glycerol (6.38 vol%)</td>
<td>1.012</td>
<td>3.042</td>
<td>1.00</td>
<td>71.85</td>
</tr>
<tr>
<td>Water-methanol (18.9 vol%)</td>
<td>0.971</td>
<td>1.770</td>
<td>1.00</td>
<td>56.71</td>
</tr>
<tr>
<td>Water-Triton X-100 (0.045 vol%)</td>
<td>1.003</td>
<td>1.069</td>
<td>1.00</td>
<td>38.50</td>
</tr>
</tbody>
</table>

CMC, carboxymethyl cellulose; Triton X-100, alkyl polyester sulfonate, water-soluble; *, consistency index.

used to examine the influence of surface tension on axial dispersion.

2. Tracer Response Measurement Technique

Sodium chloride solution (1.0 N), the tracer, was contained in a reservoir pressurized by air. This reservoir was connected by a solenoid valve to a feed distributor located inside the column, 1.5 cm above the liquid distributor. In each experiment, a half-second pulse of tracer was injected into the column. Its volume was measured by means of a level gauge attached to the tracer reservoir. The concentration of tracer was monitored by a conductivity probe connected to a conductivity bridge. The probe was made of 0.5-mm diameter platinum wires. Seven probes were installed horizontally in a stainless steel tube 14 cm long and 5.0 mm in diameter. The probes were 2 cm apart and each pair of electrodes was connected in parallel with the high and low lines of the bridge. The supporting pipe of the probes could be fixed at the top of the rack so that the height of the measuring points could be varied with experimental conditions. The conductivity gain obtained from the response curve was converted into concentration from the previously calibrated relationship between conductivity gain and concentration of NaCl.

The dispersion data was interpreted by use of the axially dispersed plug flow model.\(^3\) This model has been found to adequately describe axial mixing in liquid-solid,\(^4\) liquid-gas,\(^4,17\) and three-phase fluidized beds.\(^4,5,10,15\)

The viscosity of glycerol solution was measured with an Ostwald capillary viscometer. A Brookfield Synchроlectric rotational viscometer was used to characterize the pseudoplastic CMC solutions. Within the range of shear rates studied (7.5–75 s\(^{-1}\)), power law behaviour was observed. The parameters \(n\) and \(k\) in the power law equation are also given in Table 1. A capillary rise method was used to determine the surface tension of all the solutions.

3. Liquid Holdup and Bed Expansion Measurements

Air and liquid were introduced into the bed of solid at the desired superficial velocities. When steady state was reached, the pressure profile up the entire height of the column was measured, using liquid manometers. The bed height, \(H_e\), was taken as the point at which a change in slope of the plot was observed.\(^10,11\)

Values so obtained agree well with those determined by visual observation. Liquid and gas holdups were calculated from the relations:\(^10\):

\[ e_l + e_g + e_s = 1.0 \] (1)
\[ P_b = H_e(e_l\rho_l + e_g\rho_g + e_s\rho_s)g \] (2)

in which \(e_l\) and \(e_g\) are the only unknown quantities since \(e_s\) can be calculated from a knowledge of the weight and density of the solids in the bed.

The second moment, \(\sigma^2\), about the mean, \(\bar{r}\), for the tracer response curves was computed by use of the following relationship:

\[
\bar{r} = \int_0^\infty tC(t)dt / \int_0^\infty C(t)dt
\] (3)
\[
\sigma^2 = \int_0^\infty (t - \bar{r})^2C(t)dt / \int_0^\infty C(t)dt
\] (4)

To circumvent the difficulties of perfect pulse injection and tailing problem the numerical Laplace transform method was employed.\(^15\) Peclet numbers, computed from the second moment, \(\sigma^2\), of the corrected tracer response curves with the closed-open vessel assumption,\(^10,25\) were used to calculate the axial dispersion coefficient.

\[
\sigma^2 = \frac{1}{Pe^2} (2Pe + 3)
\] (5)
\[
D_x = \frac{V_iH_e}{Pe}
\] (6)

in which expanded bed height, \(H_e\), was determined from pressure profile measurement. The axial dispersion coefficient values quoted below were obtained by averaging the results of more than five tracer response experiments. The deviation of individual values from the mean was around 10%.

4. Results and Discussion

Axial mixing parameters were presented\(^5,10,15\) pre-
viously in two ways, namely, axial dispersion coefficient, $D_z$, and the height of a mixing unit (HMU). Since the two terms contain the same experimental variables, the effects of fluid velocities and the properties of liquids and solids on axial dispersion coefficient, $D_z$, are discussed in the present report.

4.1 Effect of gas phase velocity

Liquid-phase axial mixing data for beds of three different size of solids are shown in Fig. 2. The axial dispersion or mixing increased with increasing gas velocity in liquid-gas and three-phase fluidized beds. The intensity of mixing at the given fluid velocities in liquid-gas beds was considerably greater than that in three-phase beds, which is in accord with the results of previous studies of beds of large particles\(^5,10,24\) ($D_p > 1.0\ mm$). However, in contrast, Østergaard\(^19\) reported that gas velocity had no effect on axial mixing in beds of 6-mm glass beads in the lower range of gas velocity. On the basis of the present study and those of other workers\(^1,5,10,15,16,19,21,23,24\) it can be concluded that liquid-phase axial mixing increased with gas velocity in liquid-gas and three-phase fluidized beds. It may be seen in the figure that the agreement between the present data and the model of Baird and Rice\(^2\) was within a reasonable range. However, the present data show values of $D_z$ three to four times higher than those of El-Temtamy et al.\(^5\). It is in accord with the concepts of the isotropic turbulence model\(^2\) and the correlation of Joshi\(^8\) in which the axial dispersion coefficient is directly proportional to column diameter. It is usually accepted that the axial movements of gas bubbles and wakes are the main cause of axial mixing in the direction of flow in liquid-gas and three-phase fluidized beds.\(^2,5,10,14\) In this regard, an increase in gas velocity would lead to an increase in axial mixing parameters since bubble size and rising velocity increase with gas velocity in liquid-gas and three-phase fluidized beds.\(^10,21\) In addition, previous studies\(^10,12\) showed that bubble size and its rising velocity decreased with increasing particle size and that the bubbles were generally smaller and slower moving in three-phase beds than in liquid-gas beds. This finding may explain why the intensity of axial mixing in liquid-gas beds is significantly higher than that in three-phase beds in the present study.

4.2 Effect of liquid phase velocity

The flow rate of the liquid phase had a positive effect on the axial dispersion coefficient in the beds of 1.7- and 3.0-mm glass beads in liquid-solid and three-phase systems. However, axial mixing in the beds of 6-mm particles is nearly independent of liquid velocity in the range of 8-13 cm/s. It may be seen in Fig. 3 that axial mixing increased with liquid velocity until bed porosity reaches about 0.62-0.65 in liquid-solid beds and three-phase beds at low gas rates. However, a reverse trend was observed\(^5,15\) in beds in which bed porosity exceeded 0.65. In contrast, axial mixing increased with liquid velocity at high gas rate in both systems. This behaviour is not clearly exhibited for beds of 6-mm particles. In addition, the rate of increase in $D_z$ in beds of 3-mm particles was higher than that in beds of 1.7-mm particles.

An increase in liquid flow rate will result in higher bed porosity or lower particle concentration, finer bubble dispersion at lower gas rates and increased particle oscillation in the vertical direction in three-phase fluidized beds.\(^5\) The former two consequences contribute to decreasing the intensity of axial mixing, while the third acts to enhance the axial dispersion coefficient. In beds of 1.7-mm particles, axial dispersion increased sharply with liquid velocity up to 6 cm/s. However, the rate of increase is marginal at liquid velocities above 6.0 cm/s, where the solid concentration is relatively low and may lead to a decrease of axial mixing. The same trend was also exhibited in
Fig. 3. Effect of liquid velocity on axial dispersion in three-phase beds. A: air-water-3.0 mm glass beads; B: air-water-1.7 mm glass beads. $V_L$ (cm/s): □, 0.0; ▬, 1.0; ▼, 4.0; ◆, 6.0; △, 8.0; ○, 12.0.

Fig. 4. Effect of liquid surface tension on axial dispersion in three-phase beds. A: air-liquid-1.7 mm glass beads; $V_g$ = 4.0 cm/s; B: air-liquid-3.0 mm glass beads; $V_g$ = 6.0 cm/s; C: air-liquid-6.0 mm glass beads; $V_g$ = 12.0 cm/s. $V_L$ (cm/s): □, 0.0; ▬, 1.0; ▼, 4.0; △, 8.0.

beds of 3-mm particles at lower gas rates. A bed of 3-mm particles has been characterized by the bubble breakup regime at low liquid and gas velocities and the bubble coalescence regime at high gas rates.\(^{11,15}\) This characterization is well represented by the present findings, i.e., axial mixing is relatively low in the bubble breakup regime and is high in the bubble coalescence regime, as shown in Fig. 3.

### 4.3 Effect of liquid phase surface tension

Typical plots of liquid-phase axial mixing coefficient against surface tension are shown in Fig. 4. The effect of surface tension on axial mixing in liquid-solid beds was minimal due to the turbulence motion of particles in the given range of surface tension. However, there did appear to be a small though significant increase in $D_z$ with increasing surface tension in beds of 3-mm glass beads at high gas rates.

A change in the surface tension of the liquid can be expected to have a less dramatic effect on $D_z$ since an increase of bubble size and rising velocity with surface tension is found to be small in liquid-gas and three-phase fluidized beds.\(^{12}\) However, in general, liquid axial mixing increased somewhat with surface tension due to the reduction of bed porosity\(^{11}\) in three-phase fluidized beds. The effect is more pronounced in beds of 3-mm particles in the bubble coalescence regime. However, this increase may be attributed to high gas velocity rather than surface tension effect. Moreover, these findings are somewhat inconclusive because of the rather narrow range of surface tension studied and the possible presence of surface-active contaminants. Also, these results indicated that the effect of surface tension on axial dispersion was rather small as those of bubble columns.\(^{7}\)

### 4.4 Effect of liquid phase viscosity

As noted previously,\(^{10,11}\) liquid holdup increased with liquid velocity and viscosity in liquid-solid and three-phase beds. However, the effect of viscosity on axial dispersion is mainly dependent on the flow rates of gas phase at a given liquid velocity, as shown in Fig. 5. Axial mixing increased with liquid viscosity in liquid-solid beds, three-phase beds of 6-mm particles, and beds of 1.7- and 3.0-mm glass beads at lower gas rates. In contrast, axial mixing decreased with viscosity in beds of 1.7- and 3.0-mm particles at higher gas rates.

Momentum transfer from gas bubbles to the liquid phase presumably increases with bubble size and velocity and leads to more vigorous mixing. Since the bubble rising velocity would decrease with increasing viscosity, it is logical that axial mixing decreased with liquid viscosity in the bubble coalescence regime. However, a smaller bubble size and rising velocity in highly viscous solutions did not play a significant role in axial mixing in the bubble disintegrating regime. In general, twenty-seven fold increase in viscosity increased or decreased $D_z$ by only 20%, confirming the essentially turbulent nature of the flow regimes in three-phase fluidized beds. Therefore, the effect of liquid viscosity on $D_z$ was not as significant in the
5. Correlation of the Data

The liquid-phase axial mixing data, Peclet numbers based on particle diameter, of the present and previous studies\textsuperscript{5, 15, 16, 21} of three-phase fluidized beds were correlated in terms of the ratio of fluid velocities and of particle-to-column diameters based on the concept of the isotropic turbulence model\textsuperscript{2} since the effects of liquid viscosity and surface tension on $D_z$ were marginal.

Using the isotropic turbulence model for three-phase fluidized beds, the dispersion coefficient, $D_z$, may be written as

$$D_z \propto aD_T \left[ g(V_l + V_g) \right]^b$$

(7)

in which $a$, $b$, and $c$ are constants.

The Peclet number based on particle diameter is

$$Pe_i = \frac{d_p V_l}{D_z} \propto \frac{d_p V_l}{D_T \left[ g(V_l + V_g) \right]} \propto a \left( \frac{d_p}{D_T} \right)^b \left( \frac{V_l}{V_l + V_g} \right)^c$$

(8)

where $d$, $e$ and $f$ are correlation constants. A least-square fit of the aforementioned sets of data on axial mixing to the above Eq. (8) results in

$$Pe_i = 20.19 \left( \frac{d_p}{D_T} \right)^{1.66} \left( \frac{V_l}{V_l + V_g} \right)^{1.03}$$

(9)

correlation coefficient = 0.93

This correlation covers the range of variables $0.143 < \frac{V_l}{(V_l + V_g)} < 0.962$ and $0.0022 < (d_p/D_T) < 0.08$. The goodness of fit between experimental and calculated values of Peclet number in the range of 0.0012-0.62 is shown in Fig. 6 for 424 data points.

Recently, Joshi\textsuperscript{8} proposed a unified correlation of liquid mixing in two- and three-phase fluidized beds. The present data have been tested with his correlation. However, agreement between the present data and his correlation was very poor since his correlation is only valid for the ideal bubble flow regime.

In this study, liquid-phase axial mixing characteristics have been studied at higher gas velocities and over a wide range of liquid viscosity and surface tension. It is of particular importance to note that the above correlation covers a wide range of operating and fixed variables and is not restricted to ideal bubble and slug flow, but is also applicable to industrially realistic intermediate flow regimes.

Nomenclature

\begin{tabular}{ll}
$C$ & = concentration [mol/cm$^3$] \\
$D_T$ & = column diameter [cm] \\
$D_z$ & = axial dispersion coefficient [cm$^2$/s] \\
$d_p$ & = particle diameter [cm] \\
g & = acceleration due to gravity [cm/s$^2$] \\
$H_e$ & = expanded bed height [cm] \\
\end{tabular}
\[ K = \text{fluid consistency index} \quad [\text{mN s}/\text{m}^2] \]
\[ n = \text{fluid behaviour index} \quad [\text{dimensionless}] \]
\[ P_e = \text{Peclet number based on expanded bed height;} \quad V_i H / D_e \]
\[ P_{el} = \text{Peclet number based on particle diameter;} \quad V_d D_p / D_e \]
\[ P_b = \text{bed pressure drop} \quad [\text{Pa}] \]
\[ t = \text{first moment of tracer response curve about origin} \]
\[ V = \text{superficial fluid velocity} \quad [\text{cm/s}] \]
\[ \theta = \text{holdup of individual phase} \quad [\text{dimensionless}] \]
\[ \sigma = \text{surface tension} \quad [\text{dyn/cm}] \]
\[ \sigma^2 = \text{second moment of tracer response curve about the mean} \quad [\text{cm}^2] \]
\[ \rho = \text{density} \quad [\text{g/cm}^3] \]
\[ \mu = \text{generalized viscosity constant}; K_8^{-1} \quad [\text{cP}] \]

\(<\text{Subscripts}>\)
\[ g = \text{gas phase} \]
\[ l = \text{liquid phase} \]
\[ s = \text{solid phase} \]

**Literature Cited**


**GAS-PARTICLE HEAT TRANSFER IN A DISPERSED BED**

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Wet activated alumina particles were dried in the constant drying rate period in a dispersed bed. The heat transfer coefficients between the particles and gas were measured.

The Nusselt number for gas-particle heat transfer in the dispersed bed was affected by the particle Reynolds number and the void fraction. The empirical equation for gas-particle heat transfer in the dispersed bed was obtained.

**Introduction**

The estimation of gas-particle heat transfer coefficient is very important in the design and analysis of various kinds of processes accompanied by heat transfer, such as drying, adsorption, combustion and catalytic reaction.

Ranz and Marshall\(^1\) obtained the empirical equation for heat or mass transfer coefficient from a single particle in fluid. The Nusselt number converges to 2 as the particle Reynolds number becomes very small. Ranz and Marshall's equation is usually used to estimate the gas-particle heat transfer coefficient in the pneumatic conveying dryer, in the case of high Reynolds number (\(Re \geq 100\)). However, there is no suitable empirical equation to estimate the heat transfer coefficient in the case of low particle Reynolds number.

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