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Gas Holdup and Axial Dispersion Coefficient in Gas–Liquid Cocurrent Spout-Fluid Beds

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ABSTRACT

An experimental investigation was carried out to determine the effect of gas velocity, liquid velocity, and column diameter on the gas holdup and axial dispersion coefficients in cocurrent spout-fluid beds in three columns with diameters of 74, 114, and 144 mm, all with a height of 1.20 m, at superficial gas and liquid velocities of 0.001–0.186 $m \cdot s^{-1}$ and 0.002–0.06 $m \cdot s^{-1}$, respectively. The axially dispersed plug flow model equation was solved by using the finite difference technique and compared with the analytical solution proposed by Uysal and Anabtawi. Gas holdup was found to increase with increases of both the gas velocity and column diameter. The effect of liquid velocity on the gas holdup was found to be insignificant. The axial dispersion coefficient was found to increase with increasing gas velocity, liquid velocity, and column diameter. New correlations for predicting the gas holdup and axial dispersion coefficient in a spout-fluid bed and based on large data for a two-phase system are presented with maximum deviations not exceeding 6 and 7%, respectively.

Key Words. Spout-fluid beds; Gas holdup; Dispersion coefficient

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INTRODUCTION

Spout-fluid bed has emerged in recent years as the most promising device which overcomes many limitations of fluidized beds and spouted beds by superimposing one system on another to achieve better solid–fluid contact (2). There are few studies on the axial dispersion of the liquid phase in two-phase flow in both bubble columns and fluidized beds where axial dispersion of liquid phase is expected to be quite large (3–7). It was reported by the present author (8) that in a spouted bed, the liquid phase is sheared by the gas phase injected in the spout, resulting in the formation of a large number of small bubbles which increase the gas holdup in a two-phase system and provide much better mixing than do rising bubbles in a bubble column. Gas holdup and axial dispersion coefficients are important parameters for the design and operation of gas–liquid contacting reactors. The gas holdup for a gas–liquid spout-fluid bed of square cross-sectional area operated continuously with respect to both gas and liquid flow was reported by the same authors (9). They showed that gas holdup increased with increasing gas velocity and decreased with increasing liquid velocity as was reported in bubble columns operating continuously with respect to both gas and liquid flow (6, 7, 10). The effect of column diameter on gas holdup has not yet been investigated in a continuous spout-fluid bed. Kim and Kim (4) investigated the liquid axial dispersion coefficient in a two-phase fluidized bed and have it increased with increasing liquid velocity and column diameter. Muroyama et al. (5) and Tomida et al. (7) studied the liquid axial dispersion coefficient in two-phase bubble columns and found it to increase with increasing gas velocity, liquid velocity, and column diameter. However, Wachi et al. (11), in contrast to those authors, reported a decrease in the axial dispersion coefficient with increasing liquid velocity. The effect of these variables has not yet been investigated in spout-fluid beds.

It is thus the purpose of this study to investigate the gas holdup and the axial dispersion coefficient in a gas–liquid spout-fluid bed. It is aimed to solve the general plug-flow model equation by the finite difference technique and to compare the solution with the analytical solution. It is also aimed to correlate the gas holdup and the axial dispersion coefficient as a function of gas velocity, liquid velocity, and column diameter.

APPARATUS AND EXPERIMENTAL PROCEDURES

Experiments were carried out in three different cylindrical Plexiglass spout-fluid columns of 6.0 mm thickness, 120 cm height, and internal diameters of 7.4, 11.4, and 14.4 cm. A schematic diagram of one of the columns is shown as Fig. 1. Each column was connected to a calming section of 150 mm height and of a cross-sectional area equivalent to each column, flanged together with the column diameters. The calming sections were packed with 10 mm spherical glass particles to give a uniform liquid distribution. Pressure taps, mounted flush with the wall of the column at 100 mm height intervals, were used for taking samples to be analyzed for concentration. The liquid used was deionized water. The liquid was pumped continuously via a calibrated rotameter through a calming section and then through distributors of 48, 164, and 376 holes for the three columns, respectively, each hole of 2.0 mm diameters and arranged in a triangular pitch into the column. The overflow water was drained outside the system. Compressed air was admitted to the bed through a single nozzle of 10 mm diameter into the column via another calibrated rotameter. The average gas holdup in the bed at different gas and liquid flow rates was determined using a bed expansion technique by a piezometer

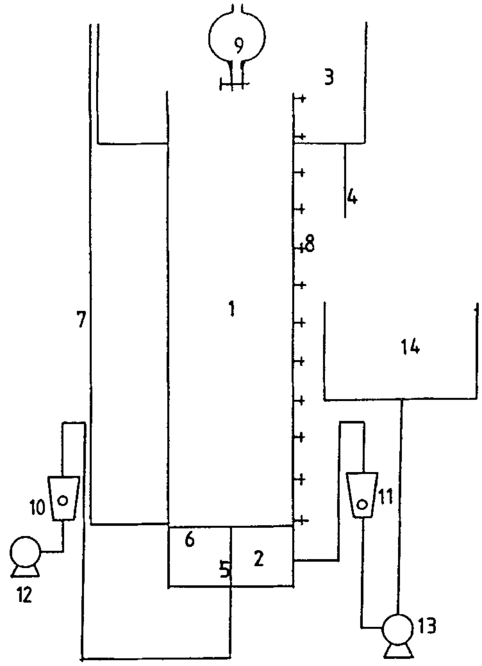

FIG. 1 Schematic diagram of the equipment: (1) column, (2) calming section, (3) water collector, (4) discharge pipe, (5) nozzle, (6) distributor, (7) piezometer, (8) sample taps, (9) tracer funnel, (10) calibrated gas rotameter, (11) calibrated liquid rotameter, (12) compressor, (13) pump, (14) water tank.

attached to the column. Potassium permanganate as a 1.0 M solution was used as a tracer material. The $KMnO_4$ solution was introduced through a funnel with manual control. The funnel stem was positioned so that its end just touched the liquid surface at the discharge opening. The time taken for the tracer material color edge to pass from the funnel stem end to a distance 0.90 m from the top of the edge of the stem was recorded. At the same time the tracer supply was closed, two samples of solution, one next to the stem and another at 0.90 m distance, were taken and analyzed for the concentration of the $KMnO_4$ by titration using an automatic Karl Fisher titrator. A third sample was taken at 1.0 m depth from the stem and analyzed for the presence of $KMnO_4$.

DATA ANALYSIS

The one-dimensional axially dispersed plug flow model of constant coefficient is best described by

$$\partial c/\partial t = E_Z \partial^2 c/\partial z^2 + U_1' \partial c/\partial z \qquad (1)$$

This equation was solved by using the finite difference Crank–Nicolson scheme with the following boundary conditions:

$$c(0, t) = c_0 \qquad \text{at } z = 0$$

$$c(\infty, t) = 0 \qquad \text{at } z \to \infty \quad \text{and} \quad t \geqslant 0$$

Initial conditions

$$c(0, 0) = c_0 \qquad \text{at } z = 0 \quad \text{and} \quad t = 0$$

By putting dimensionless groups

$$\mathrm{Pe} = U_1' h/E_Z$$

$$\tau = tU_1'/(\mathrm{Pe}h) = tE_Z/h^2$$

$$Z = z/h$$

$$C = c/c_0$$

Eq. (1) can be rewritten as:

$$\partial C/\partial \tau = \partial^2 C/\partial Z^2 + \mathrm{Pe}\partial C/\partial Z \qquad (2)$$

with boundary condition

$$C(0, \tau) = 1.0 \qquad \text{at } Z = 0$$

$$C(\infty, \tau) = 0 \qquad \text{at } Z \to \infty \quad \text{and} \quad \tau \geqslant 0$$

and initial conditions

$$C(0, 0) = 1.0 \qquad \text{at } Z = 0 \quad \text{and} \quad \tau = 0$$

Equation (2) was solved using the finite difference Crank–Nicolson scheme with a mesh size of $\Delta Z = 0.025$ and $\Delta\tau = \Delta Z/2\text{Pe}$. Since the solution requires prior knowledge of the axial dispersion coefficient E_Z in order to calculate the dispersion coefficient E_z for a known concentration at a given penetration distance, the Newton–Raphson's method was applied.

In order to verify the computational algorithm, it was compared with an analytical solution derived by Uysal and Anabtawi (1) as follows:

$$C = 0.5(\exp(-hU_1'/E_z)\ \text{erfc}(h/(2\sqrt{E_z t}) - U_1'/2\sqrt{t/E_z}) + \text{erfc}(h/(2\sqrt{E_z t}) + U_1'/2\sqrt{t/E_z}))$$

The comparison is shown in Fig. 8. The results were obtained for dimensionless concentration $C = 0.001$, a penetration distance of 0.90 m, and a mesh size of 0.025. In general, the agreement was satisfactory although the numerical solution does overpredict E_z slightly. This overprediction was approximately 7%.

RESULTS AND DISCUSSION

The gas flow rate was varied over the range from 0.001 to 0.186 $\text{m}\cdot\text{s}^{-1}$ and the liquid flow rate from 0.002 to 0.06 $\text{m}\cdot\text{s}^{-1}$ in three columns with internal diameters 74, 114, and 144 mm of 1200 mm height. Small, nearly uniform bubbles were observed at low flow rates. Larger bubbles were observed in the annular region at large flow rates. At the interface between the spout and the annulus there were small bubbles, mainly due to the larger shear force acting on the gas at its surfaces. The gas holdup increased with increasing gas velocity, because as the gas velocity increased, a large number of small and large bubbles resulted from the coalescence of smaller bubbles in the annulus. They were responsible for expansion of the bed surface and therefore resulted in a higher gas holdup value. The purple color of $KMnO_4$ faded as the tracer traveled from the top to the bottom of the bed. Below a distance of 0.90 m from the top of the bed, the water was clear. For further confirmation of the absence of $KMnO_4$, samples were taken at 1.0 m from the top and analyzed.

Gas Holdup

Effect of Gas Velocity

The variation of gas holdup with gas velocity at different liquid velocities and column diameters is shown in Fig. 2. The gas holdup was found to increase with increasing gas velocity as was reported by the same authors (9) who worked a in rectangular spout-fluid bed operated continuously

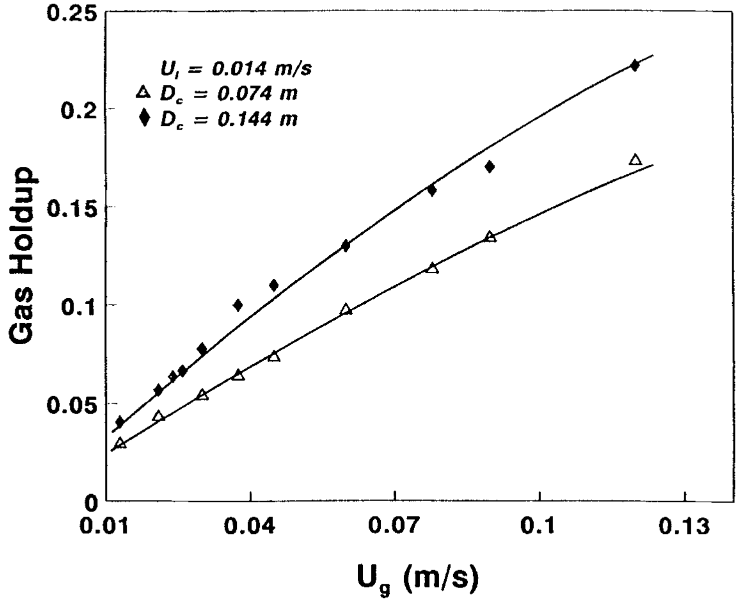

FIG. 2 Effect of gas velocity on gas holdup.

with respect to both liquid and gas flow. It was also in agreement with Nishikawa et al. (10) and Fan et al. (12) who worked in spouted beds operated continuously with respect to both gas and liquid flow. It was also in agreement with those who worked in a continuous bubble column (3, 11, 13–15). The increase in gas velocity caused an increase in the number of small bubbles due to the large shear force acting on the surface of the bubbles at the interface between the spout and the annulus. In the presence of larger bubbles in the annulus, the bed expanded and the gas holdup increased.

Effect of Liquid Velocity

The effect of liquid velocity on gas holdup in a gas–liquid spout-fluid bed is shown in Fig. 3. The gas holdup was found to decrease slightly with increasing liquid velocity as was reported by the same authors (9) who worked in a continuous rectangular spout-fluid bed. This variation was also in agreement with that reported by Wachi et al. (11) and others (16) who worked in continuous bubble columns. However, this variation was in contrast with the results reported in continuous spouted beds reported by Nishikawa et al. (10) and Fan et al. (12) who found the gas holdup to increase with increasing liquid velocity. This finding indicates

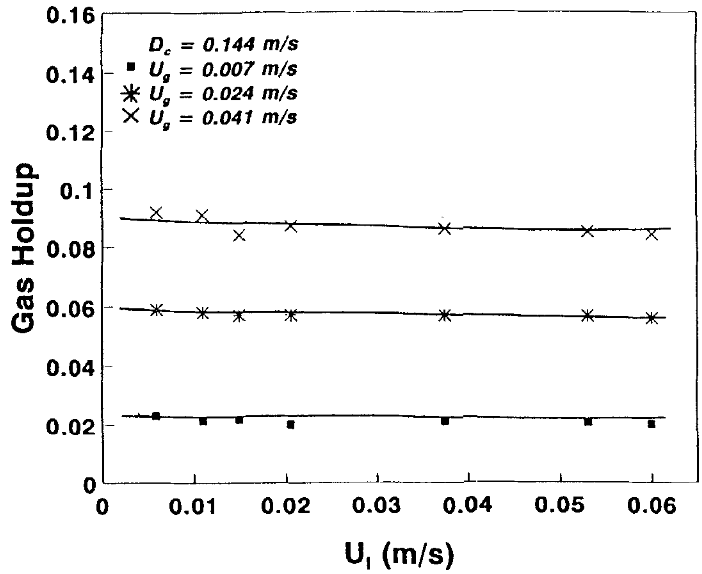

FIG. 3 Effect of liquid velocity on gas holdup.

that the two-phase spout-fluid bed is more similar in behavior to the bubble columns than to spouted beds.

Effect of Column Diameter

Effect of column diameter on gas holdup is shown in Fig. 4. Gas holdup was found to increase with increasing column diameter. This is in contrast to the variation reported in a spout-fluid bed operated batchwise with respect to liquid flow as was reported by the same author (17). It is also in contrast with the findings reported in bubble columns (6).

Axial Dispersion Coefficient

Effect of Gas Velocity

The variation of axial dispersion coefficient, E_z, with gas velocity is shown in Fig. 5. It was found that the axial dispersion coefficient increases with increasing gas velocity as was reported by Hikita and Kikukawa (18), Muroyama et al. (5), Sekizawa and Kubota (19) and Tomida et al. (7) who worked in two-phase bubble columns. Usually the axial movements of gas bubbles and wakes are the main cause of axial mixing in the direction of the flow in gas–liquid systems. In spout-fluid beds an increase in gas velocity leads to an increase in the number of small bubbles as a result

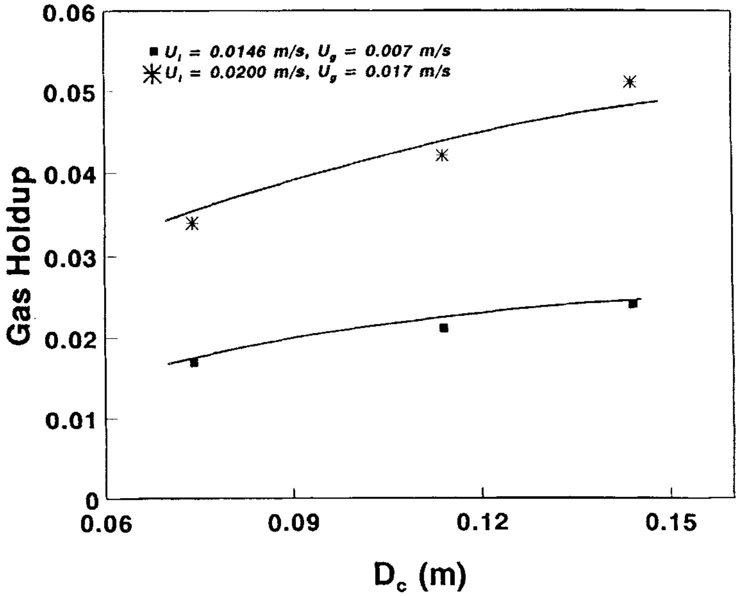

FIG. 4 Effect of column diameter on gas holdup.

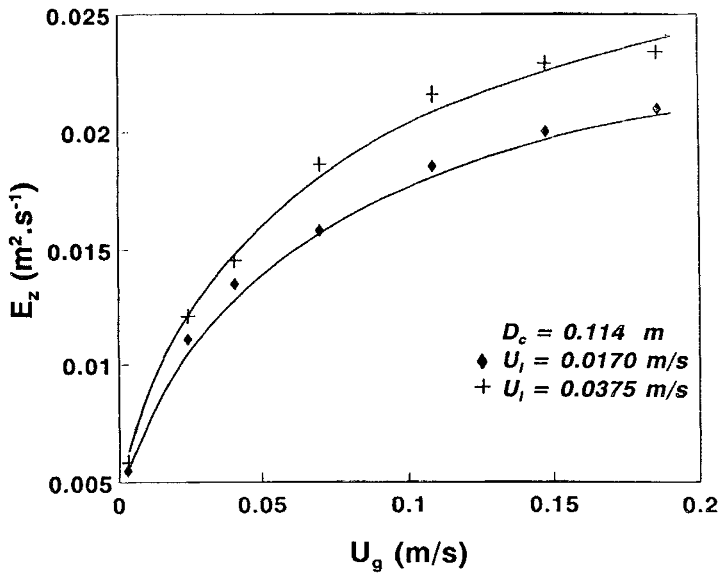

FIG. 5 Effect of gas velocity on axial dispersion coefficient E_Z.

of the large shear force acting by the gas on the bubble surfaces at the interface between the spout and the annulus. Also, large bubbles are formed by the coalescence of smaller bubbles in the annulus away from the spout. The combination of the different types of bubbles is responsible for the increase in axial mixing.

Effect of Liquid Velocity

The variation of axial dispersion coefficient with liquid velocity is shown in Fig. 6. The axial dispersion coefficient was found to increase with increasing liquid velocity as was reported by other investigators (4, 5, 7) who worked in both fluidized beds and bubble columns. In the present work, E_Z was found to be proportional to about the 0.093 power of the liquid velocity. This is qualitatively in agreement with Kim and Kim (4) who worked with a fluidized bed and also in agreement with Muroyama et al. (5) and Tomida et al. (7) who worked with bubble columns and have reported powers of 1.0, 0.262 and 1.68, respectively.

Effect of Column Diameter

The variation of the axial dispersion coefficient with column diameter is shown in Fig. 7. The axial dispersion coefficient was found to increase

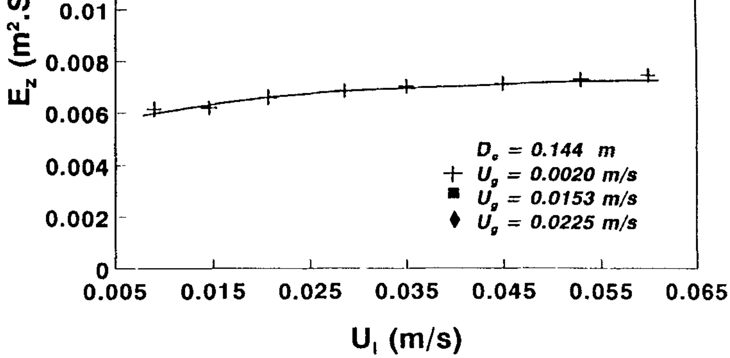

FIG. 6 Effect of liquid velocity on axial dispersion coefficient, E_Z.

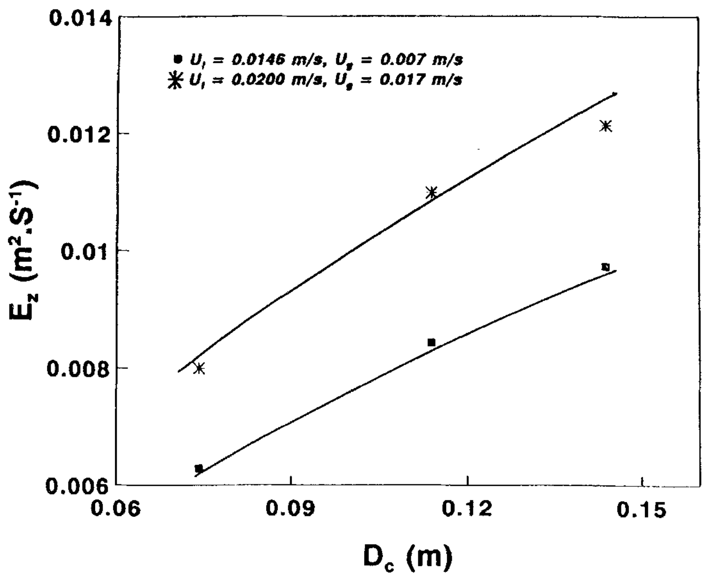

FIG. 7 Effect of column diameter on axial dispersion coefficient, E_Z.

with increasing column diameter. The effect of column diameter on axial dispersion in a fluidized bed and in bubble columns was studied by several investigators (4, 5, 7, 18). All these authors showed that the axial dispersion coefficient increased with increasing column diameter. The effect of column diameter was not investigated for a spout-fluid bed. In the present work, E_z was found to be proportional to about the 0.64 power of the column diameter. This dependency is qualitatively in agreement with the work reported by Kim and Kim (4) who worked in a fluidized bed, and by Hikita and Kikukawa (18), Muroyama et al. (5), and Tomida et al. (7) who worked in bubble columns and reported powers of 1.66, 1.25, 1.58, and 2.62, respectively.

Correlation of the Data and Comparison with Other Authors

The gas holdup data were used to develop the following correlation:

$$\epsilon_g = 2.6U_g^{0.777}U_l^{-0.019}D_c^{0.456} \tag{3}$$

with a correlation coefficient $r = 0.97$ and an average standard error of 2%. The maximum deviation of experimental data from prediction for this correlation did not exceed 6%. However, as Eq. (3) and Fig. 3 shows, the effect of liquid velocity is very small. The average liquid velocity can therefore be incorporated within the constant in Eq. (3) without losing much accuracy. Equation (3) can be rewritten as

$$\epsilon_g = 2.81 U_g^{0.776} D_c^{0.461} \tag{4}$$

A statistical check on this correlation using the original data confirms that Eq. (4) provides a satisfactory correlation coefficient of $r = 0.97$ and an average standard error of 2.1% with the same maximum deviation as for Eq. (3).

The superficial gas velocity U_g was varied from 0.001 to 0.186 m·s^{-1} and the liquid velocity was varied from 0.002 to 0.06 m·s^{-1}. It should be noted that all the curves drawn in Figs. 2–4 represent Eq. (3).

All axial dispersion coefficient data, E_Z, consisting of 275 sets, were used to develop the following dimensionless correlation:

$$(\mathrm{Pe})_{D_c} = U_l D_c/(E_Z) = 0.525 U_l^{0.91} U_g^{-0.31} D_c^{-0.30} g^{-0.30} \tag{5}$$

with correlation coefficient $r = 0.963$ and an average standard error not exceeding 2.5%. The maximum deviation of experimental data from pre-

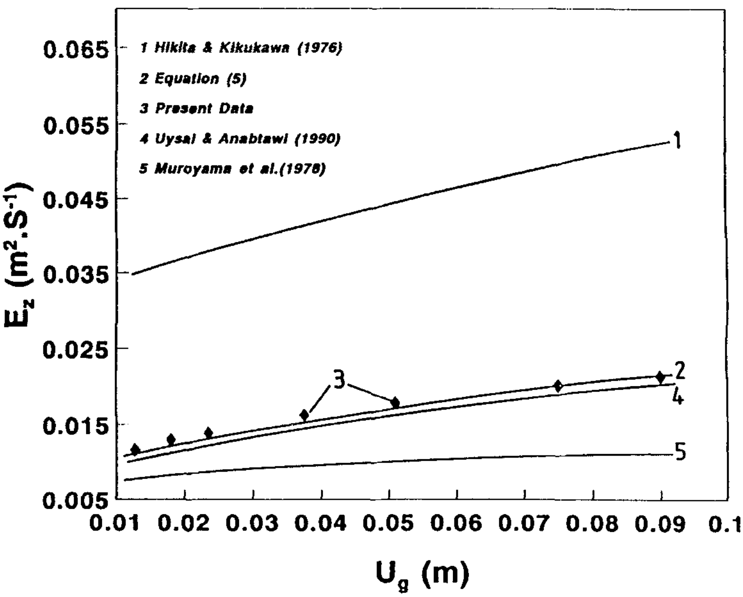

FIG. 8 Comparison of the axial dispersion coefficient in spout fluid bed using finite difference scheme with the analytical solution and with previous data of E_Z in bubble columns.

diction of this correlation was 7%. It should be noted that all the curves drawn in Figs. 5–7 represent Eq. (5).

Figure 8 shows a comparison between E_Z, proposed in Eq. (2), and the analytical solution proposed by Uysal and Anabtawi (1), the correlation predicted by Muroyama et al. (5) in bubble columns, and the correlation predicted by Hikita and Kikakawa (18) in bubble columns. The numerical solution was in agreement with the analytical solution, although overpredicting it by 7% on average. The Muroyama et al. correlation (5) underestimate the present data by 40%, and Hikita and Kikukawa's correlation (18) overestimated the present data by 62%.

CONCLUSIONS

For a gas–liquid cocurrent spout-fluid bed, the gas holdup and axial dispersion coefficients were measured by varying the values of superficial gas velocity from 0.001 to 0.186 $m \cdot s^{-1}$, superficial liquid velocity from 0.002 to 0.06 $m \cdot s^{-1}$, and in three different column diameters of 74, 114, and 144 mm. Gas holdup increased with gas velocity and column diameter. The effect of liquid velocity on gas holdup was found to be insignificant. The axial dispersion coefficient increased with increasing gas velocity, liquid velocity, and column diameter. The axial dispersion coefficient in terms of Peclet number have been correlated as a function of column diameter, gas velocity, and liquid velocity.

NOMENCLATURE

c	tracer concentration ($kmol \cdot m^{-3}$)
c_0	tracer concentration at top of the column ($kmol \cdot m^{-3}$)
C	dimensionless concentration
D_c	column diameter (m)
E_Z	axial dispersion coefficient ($m^2 \cdot s^{-1}$)
g	gravitational acceleration (m/s^2)
h	penetration distance (m)
Pe	Peclet number ($U_l h / E_Z$)
$(\mathrm{Pe})_{D_c}$	Peclet number based on column diameter ($U_l D_c / E_Z$)
t	time(s)
U_g	superficial gas velocity ($m \cdot s^{-1}$)
U_l	superficial liquid velocity ($m \cdot s^{-1}$)
U_l'	interstitial liquid velocity (U_l/ϵ_l) ($m \cdot s^{-1}$)
z	axial distance (m)
Z	dimensionless distance (z/h)
ϵ_g	gas holdup
ϵ_l	liquid holdup ($1 - \epsilon_g$)

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