Korean J. of Chem. Eng., 7(3) (1990) 182-187

AXIAL DISPERSION CHARACTERISTICS IN THREE PHASE FLUIDIZED BEDS

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(Received 16 February 1990 • accepted 31 May 1990)

Abstract—Axial dispersion coefficients in three-phase fluidized beds have been measured in a 0.152 m-ID × 1.8 m high column by the two points measuring technique with the axially dispersed plug flow model.

The effects of liquid velocity (0.05-0.13 m/s), gas velocity (0.02-0.16 m/s) and particle size (3-8 mm) on the axial dispersion coefficient at the different axial positions (0.06-0.46 m) in the bed have been determined.

The axial dispersion coefficient increases with increasing gas velocity but it decreases with an increase in particle size and exhibits a maximum value with an increase in the axial position from the distributor.

The axial dispersion coefficients in terms of the Peclet number have been correlated in terms of the ratio of fluid velocities, the ratio of the particle size to column diameter, and the dimensionless axial position in the bed based on the isotropic turbulence theory.

INTRODUCTION

Three-phase (liquid-gas-solid) fluidized bed operation can be achieved in a bed of solid particles which are fluidized by cocurrent upflow of gas and liquid phases.

Three phase fluidized beds as a chemical reactors have been widely used in petroleum, pulp and paper, coal liquefaction and biological processes since it has been generally known that three phase fluidized bed reactors are superior in their performance to that of three phase fixed bed operations.

Various aspects of three phase fluidized beds have been studied extensively as can be found in previous review articles [1-6]. However, information on the axial dispersion characteristics in three phase fluidized bed are very limited but these information are essentially required for designing three-phase fluidized bed reactors. Especially, the variation of axial dispersion coefficient (D_z) along the bed height has not been reported up to date.

Therefore, in this study, the axial dispersion coefficients in three phase fluidized beds have been determined by using the two points measuring technique with the axially dispersed plug flow model. The effects of liquid and gas velocities and paticle size on the axial dispersion coefficient (D_z) along the bed height have been determined. The D_z values in three-phase fluidized beds have been well represented by a correlation based on the isotropic turbulence theory.

EXPERIMENTAL

Experiments were carried out in a Plexiglas column 1.80 m high and 0.152 m in diameter as shown in Figure 1. Solid particles were supported on a perforated plate containing 156 evenly-spaced holes of 3 mm diameter which served as a liquid phase distributor. The distributor was situated between the main column section and a 0.3 m high stainless steel distributor box into which the liquid phase was introduced through a 50.8 mm pipe from the reservoir. The liquid flow rate was measured with a turbine flow meter (Hoffer Flow Controls) and regulated by means of globe valves on the feed and bypass lines. The gas was fed to the column through four evenly spaced 6.35 mm distributor pipes which contained 26 holes of 2.0 mm diameter. Eight pressure taps were mounted flush with the wall of the column at 0.16 m height intervals from the distributor plate. The static pressure at each of these points was measured with a liquid manometer.

Throughout this study, water was used as the liquid phase, air as the gas phase and either 3.0, 6.0 or

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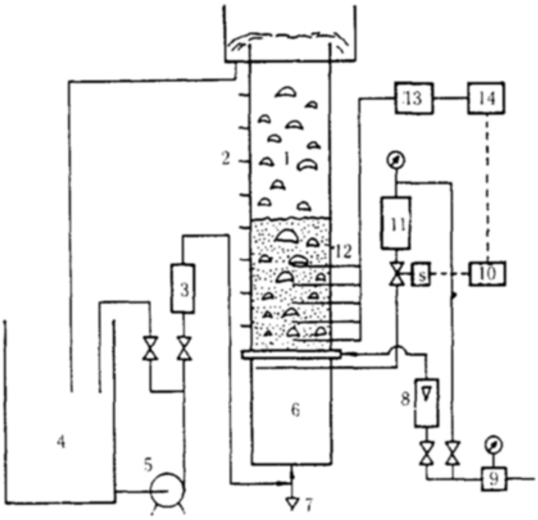

Fig. 1. Schematic diagram of the experimental apparatus.

1. Main column	2. Pressure tap
3. Turbine flow meter	4. Liquid reservoir
5. Pump	6. Distributor box
7. Drain	8. Gas flow meter
9. Filter and regulator	10. Timer
11. Tracer tank	12. Conductivity probes
13. Conductivity bridge	14. Recorder

8.0 mm glass beads, each with a density of 2500 kg/m^3 as the solid phase. Superficial velocities of liquid and gas phases ranged from 0.05 to 0.13 m/s and from 0.02 to 0.16 m/s, respectively.

1. Measurements of phase holdups and axial dispersion coefficients

The liquid and gas phases were introduced into the bed of solids with the desired superficial velocities of both phases. When steady-state was reached, the pressure profile up to the entire height of the column was measured by using liquid manometers at different axial positions. The expanded bed height was taken as the point at which a change in the slope of the plot was observed [7]. Values of the expanded bed height so obtained agreed well with visual observation. The liquid and gas phase holdups were determined from the knowledge of pressure drop, bed height, fluids and solid properties [8].

Sodium chloride solution (0.3 N), as the tracer, was contained in a reservoir (0.1 m-ID × 0.4 m high) pressurized by compressed air. This reservoir was connected through a solenoid valve controlled by an automatic timer to the down faced tracer feed distributor located underneath of the distributor plate. In each experiment, a 0.25-0.5 second pulse of tracer was in-

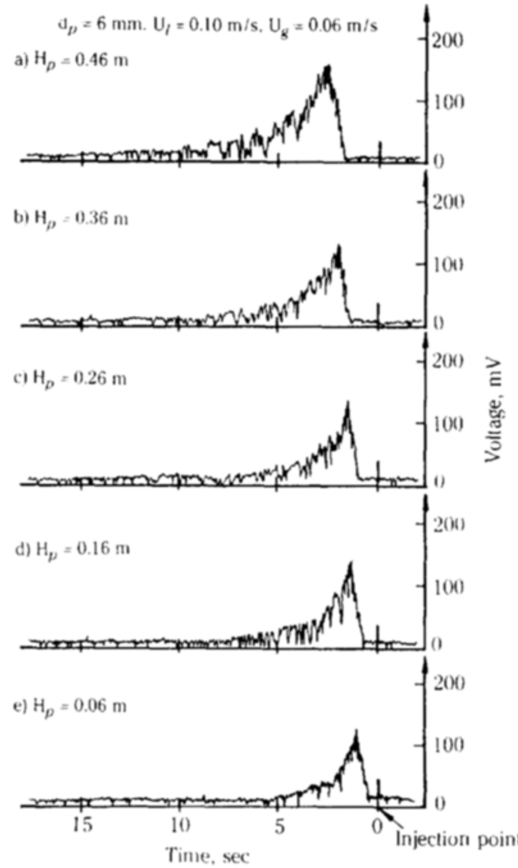

Fig. 2. A typical tracer response curve.

jected 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 five conductivity probes with the conductivity bridges. The probes (0.5 mm-OD platinum wire) were located at the center of the column 0.06, 0.16, 0.26, 0.36 and 0.46 m above the distributor plate. Typical tracer response curves are shown in Figure 2. The conductivity gain obtained from the response curve was converted into the concentration from the previously calibrated relationship between conductivity gain and the concentration of NaCl.

The dispersion data were interpreted by means of the axially dispersed plug flow model [9]. This model has been found to adequately describe axial mixing in liquid-solid [7], gas-liquid [10,11] and three phase fluidized beds [7,12-17].

To circumvent the difficulties of perfect pulse injection and tailing problem, two measuring points technique was employed [13].

The moment difference between the two measur-

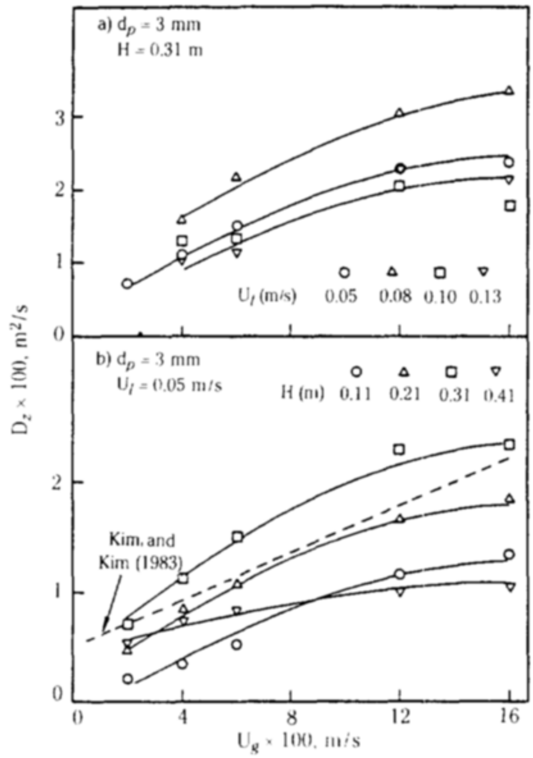

Fig. 3. Effect of gas velocity on D_z in three phase fluidized beds of 3 mm glass beads.

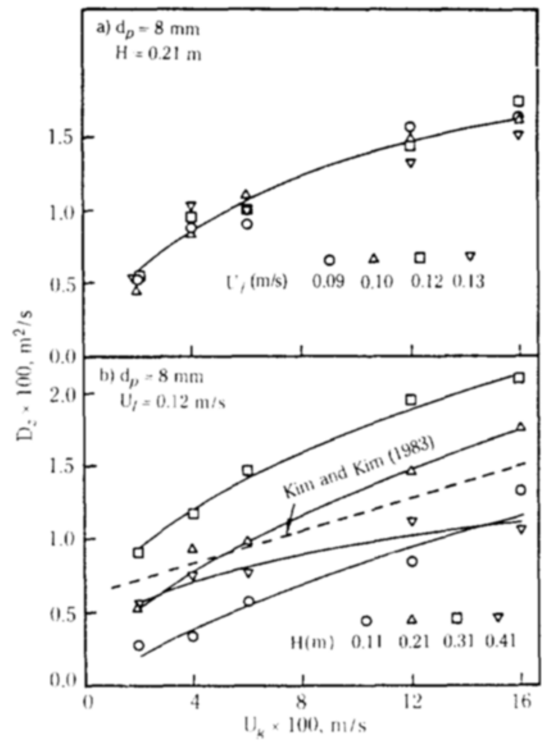

Fig. 4. Effect of gas velocity on D_z in three phase fluidized beds of 8 mm glass beads.

ing points can be expressed as:

$$\Delta\mu = 1.0 \tag{1}$$

$$\Delta\sigma^2 = 2/\mathrm{Pe} = \frac{2D_z\epsilon_1}{U_1 L}. \tag{2}$$

The axial dispersion coefficient, D_z, has been calculated from the difference of second moments of the response curves for each experimental condition from Equation (2).

RESULTS AND DISCUSSION

It has been accepted that the axial movements of gas bubbles and liquid wakes are the main cause of axial mixing in the direction of flow in liquid-gas [18,19] and three phase fluidized beds [7,12,14,16]. The rising bubbles are trailed by turbulent wakes which concurrently form and separate during their rise, these fast moving liquid elements will cause axial mixing in three fluidized beds [20].

2. Effect of gas velocity on D_z

The effect of gas velocity on D_z in three phase fluidized beds of different particle sizes is shown in Figures 3 and 4. It is well known that the increase of gas velocity causes the increase of D_z in three phase fluidized bed regardless of particle size and liquid velocity as obtained in this study. Also it can be seen in Figures 3b and 4b that D_z increases with gas velocity in all the axial positions of the bed. On the basis of the present and previous studies, it can be concluded that D_z increases with the gas velocity in all the axial positions of the beds of different particle sizes since bubble size, bubble phase holdup and wake volume behind bubbles may increase with an increase in gas velocity in the beds of different particle sizes along the axial positions.

3. Effect of liquid velocity on D_z

The effect of liquid velocity on D_z in the beds of different particle sizes is shown in Figures 5 and 6. The liquid velocity has a positive effect on D_z in the beds of 3.0 and 6.0 mm glass beads at higher gas velocities. However, the coefficient in the beds of 8.0 mm particles is nearly independent of liquid velocity in the range of 0.09-0.13 m/s.

An increase in liquid velocity will result in higher bed porosity or lower particle concentration, finer bubble dispersion at lower gas velocities and higher particle oscillation in the vertical direction of three phase

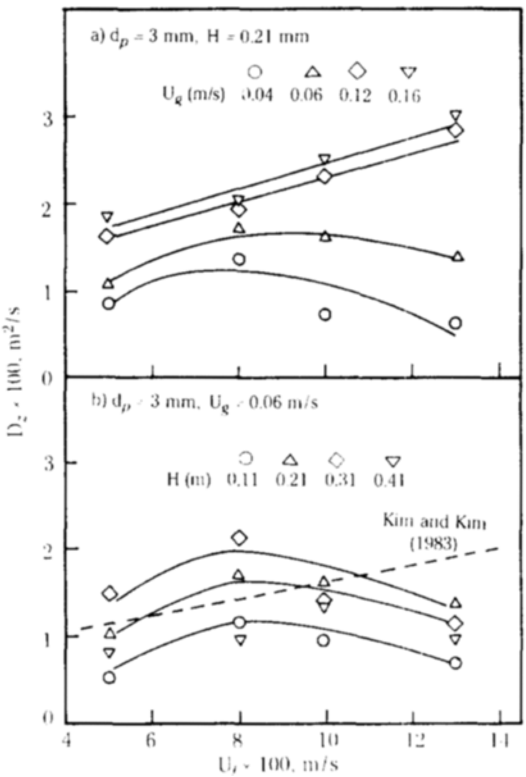

Fig. 5. Effect of liquid velocity on D_z in three phase fluidized beds of 3 mm glass beads.

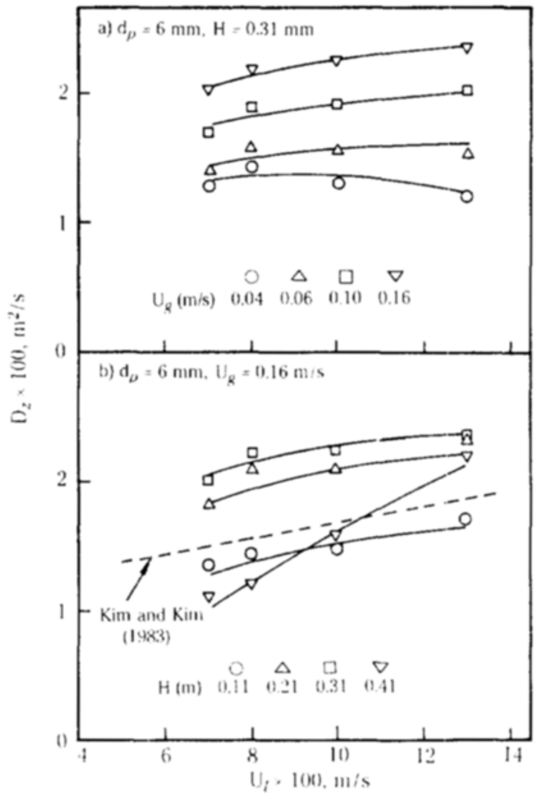

Fig. 6. Effect of liquid velocity on D_z in three phase fluidized beds of 6 mm glass beads.

fluidized beds [12]. The increasing inertia force of solid particles with an increase in fluid velocity may contribute to the non-uniformity of three phase fluidized bed by generating the form turbulence [21] and may oscillate the particle movement and disintegrate the rising bubbles [22]. These two conflicting effects may cause a maximum increase in liquid phase axial mixing. As can be seen in Figure 5a, this phenomena is well represented by the variation of D_z in the beds of 3.0 mm glass beads in which D_z exhibits a maximum value at a given liquid velocity at the lower gas velocities. However, at higher gas velocities, D_z increases with increasing liquid velocity since the beds of 3.0 mm glass beads is characterized by the bubble coalescence flow regime at higher gas velocities [8,13].

In the beds of 6.0 mm glass beads, it can be anticipated from previous studies [23-25] that the flow pattern of solid phase would be in random motion with higher solid concentration and the larger particles may generate larger momentum which may easily disintegrate the gas bubbles. Therefore, as can be seen in Figure 5b, the rate of increase in D_z is relatively low in the gas velocity range 0.04-0.16 m/s.

In Figure 6b, the rate of increase in D_z at the axial position of 0.41 m from the distributor is higher than that at the other axial positions. This may reflect that the rate of increase in bubble size from the bubble coalescence at the higher axial positions is higher than those at the lower axial positions in the bed at higher bed porosities with increasing liquid velocity.

4. Effect of the dimensionless axial position in the bed on D_z

As can be seen in Fig. 7, D_z exhibits a maximum value with increasing the dimensionless axial positions in the bed (H/H_o) in the range of 0.6-0.7 regardless of particle size. It can be attributed to the bubble rising velocity which appears a maximum value in the dimensionless axial positions of 0.65-0.75 [26] due to the end effect.

CORRELATION OF THE DATA

The axial and radial dispersion coefficients have been well represented by the isotropic turbulence theory in two and three phase fluidized beds since the

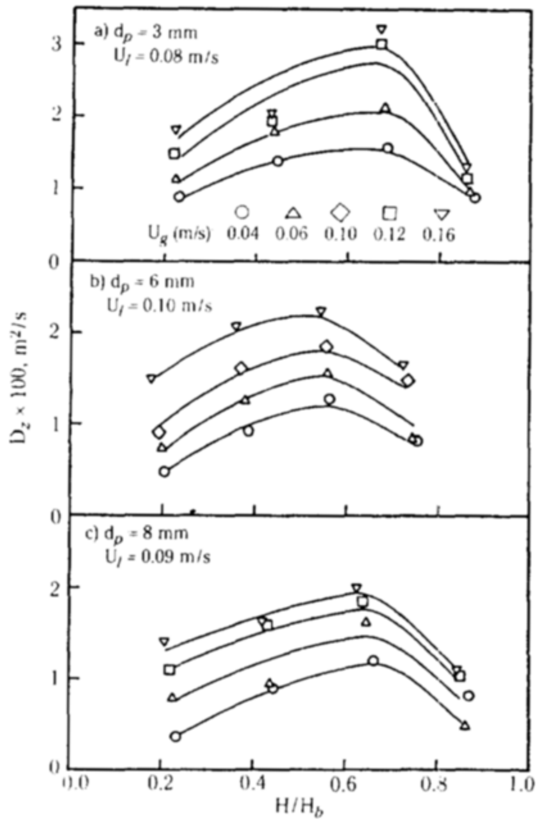

Fig. 7. Effect of axial bed height on D_z in three phase fluidized beds.

dispersion can be attributed to the turbulent eddy motions in the bed [14-18].

The eddy diffusivity can be written as:

$$D_z \propto D_c^a [g(U_l + U_g)]^b \quad (3)$$

where a and b are constants.

The coefficient in terms of the particle Peclet number in three phase fluidized beds based on the isotropic turbulence theory has been proposed by Kim and Kim [14] as shown in Eq. (4) which has been known as the best representation of D_z values in three phase fluidized beds [17].

$$Pe_z' = \left(\frac{d_p U_l}{D_z'}\right) = 20.19 \left(\frac{d_p}{D_c}\right)^{1.66} \left(\frac{U_l}{U_l + U_g}\right)^{1.03} \quad (4)$$

However, Equation (4) does not take account of the variation of D_z along the bed height of three-phase fluidized beds. Therefore, the present experimental data of the Peclet number (Pe_z) has been correlated in terms of the Peclet number ratio (Pe_z/Pe_z') in order to accommodate the effect of bed height (H/H_b) on D_z with the assumption that D_z in Eq. (4) are the mean values in three phase fluidized beds. The expanded bed height (H_b) has been correlated with the present experimental variables as:

$$H_b = 1.12 U_l^{0.46} U_g^{0.05} d_p^{-0.09} \quad (5)$$

with a correlation coefficient of 0.92. The Peclet number along the bed height can be determined from the following relation as:

$$Pe_z/Pe_z' = 2.94 - 7.01(H/H_b) + 5.87(H/H_b)^2 \quad (6)$$

in which H_b from Eq. (5) has been used in Eq. (6) and the correlation coefficient is found to be 0.90. The correlation covers the range of variables $0.24 \leq (H/H_b) \leq 0.97$, $0.0197 \leq d_p/D_c \leq 0.0526$, $0.238 \leq U_l/(U_l + U_g) \leq 0.867$ and its agreement with the data is shown in Figure 7.

CONCLUSIONS

The axial dispersion coefficient increases with an increase in gas velocity but it decreases with increasing particle size and exhibits a maximum value with increasing bed height from the distributor. Also, the coefficient exhibits a maximum value with an increase in liquid velocity at the lower gas velocities, but it increases with liquid velocity at higher gas velocities in the bed of 3 mm glass beads.

The axial dispersion coefficients in terms of the Peclet number have been correlated in terms of the ratio of fluid velocities, the ratio of the particle size to column diameter and the dimensionless bed height based on the isotropic turbulence theory.

ACKNOWLEDGEMENT

The authors wish to acknowledge a grant-in-aid for research from the Korea Science and Engineering Foundation.

MOMENCLATURE

- a : constant
- b : constant
- d_p : particle diameter [m]
- D_c : column diameter [m]
- D_z : axial dispersion coefficient of liquid phase [m²/s]
- D_z' : mean axial dispersion coefficient of liquid phase [m²/s]
- g : gravitational acceleration [m/s²]
- H : height from the distributor [m]
- H_b : expanded bed height [m]
- H_p : measuring height at the probe location [m]

L : distance between two probes [m]

Pe : Peclet number based on axial dispersion coefficient and interstitial velocity of the continuous phase $\left(\frac{U_l d_p}{\epsilon_l D_z}\right)$

Pe_z : Peclet number based on axial dispersion coefficient and superficial liquid velocity $\left(\frac{U_l d_p}{D_z}\right)$

Pe_z' : Peclet number defined in Eq. (4)

U : superficial velocity [m/s]

Greek Letters

ϵ : phase holdup

μ : dimensionless first moment

σ^2 : dimensionless second moment

Subscripts

g : gas phase

l : liquid phase

REFERENCES

1. Epstein, N.: *Can. J. Chem. Eng.*, **59,** 649 (1981).
2. Wild, G., Saberian-Broudjenni, M., Schwartz, J.L. and Charpentier, J.C.: *Int. Chem. Eng.*, **24,** 639 (1984).
3. Muroyama, K. and Fan, L.S.: *AIChE J.*, **31,** 1 (1985).
4. Yu, Y.H. and Kim, S.D.: *Chem. Ind. Technol.*, **4,** 14 (1986).
5. Pandit, A.B. and Joshi, J.B.: *Chem. Eng. Res. Des.*, **64,** 125 (1986).
6. Fan, L.S.: "Gas-Liquid-Solid Fluidization Engineering", Butterworths, Stoneham (1989).
7. Kim, S.D., Baker, C.G.J. and Bergougnou, M.A.: *Can. J. Cherm. Eng.*, **50,** 695 (1972).
8. Kim, S.D., Baker, C.G.J. and Bergougnou, M.A.: *Can. J. Chem. Eng.*, **53,** 134 (1975)
9. Bischoff, K.B. and Levenspiel, O.: *Chem. Eng. Sci.*, **17,** 245 (1962).
10. Eissa, S.H. and Schugerl, K.: *Chem. Eng. Sci.*, **30,** 1251 (1975).
11. Oki, Y. and Inoue, H.: *Chem. Eng. Sci.*, **25,** 1 (1970).
12. El-Temtamy, S.A., El-Sharnoubi, Y.D. and El-Halwagi, M.M.: *Chem. Eng. J.*, **18,** 159 (1979).
13. Michelsen, M.L. and Ostergaard, K.: *Chem. Eng. J.*, **1,** 37 (1970).
14. Kim, S.D. and Kim, C.H.: *J. Chem. Eng. Japan*, **16,** 172 (1983).
15. Kang, Y. and Kim, S.D.: *Ind. Eng. Chem. Proc. Des. Dev.*, **25,** 717 (1986).
16. Kim, S.D., Lee, M.J. and Han, J.H.: *Can. J. Chem. Eng.*, **66,** 276 (1989).
17. Saberian-Broudjenni, M., Wild, G. and Kim, S.D.: *Chem. Eng. J.*, **40,** 83 (1989).
18. Baird, M.H.I. and Rice, R.G.: *Chem. Eng. J.*, **9,** 171 (1975).
19. Letan, R. and Elgin, J.C.: *Chem. Eng. J.*, **3,** 136 (1972).
20. Kang, Y. and Kim, S.D.: *Hwahak Konghak*, **25,** 394 (1987).
21. Schlichting, H.: "Boundary Layer Theory", McGraw-Hill, New York (1968).
22. Kim, J.O. and Kim, S.D.: *Particulate Sci. & Tech.*, **5,** 309 (1987).
23. Morooka, S., Kusakabe, K. and Kato, Y.: *Int. Chem. Eng.*, **20,** 433 (1980).
24. Chang, S.K., Kang, Y. and Kim, S.D.: *J. Chem. Eng. Japan*, **19,** 524 (1986).
25. Cairns, E.J. and Prausnitz, J.M.: *AIChE J.*, **6,** 554 (1960).
26. Jin, G.T.: Ph.D. Thesis, Korea Advanced Institute of Science and Technology, Seoul (1985).