Korean J. Chem. Eng., 19(3), 491-494 (2002)

SHORT COMMUNICATION

Axial Dispersion of Gas in a Circulating Fluidized Bed

Jimin Kim, Guiyoung Han† and Changkeun Yi*

Department of Chemical Engineering, Sungkyunkwan University, Suwon 440-746, Korea
*Korea Institute of Energy Research, Daejeon 305-343, Korea

(*Received 23 October 2001 • accepted 7 February 2002*)

Abstract−An axial dispersion of gas in a circulating fluidized bed was investigated in a fluidized bed of 4.0 cm I.D. and 279 cm in height. The axial dispersion coefficient of gas was determined by the stimulus-response method of trace gas of CO_2. The employed particles were 0.069 mm and 0.147 mm silica-sand. The results showed that axial dispersion coefficients were increased with gas velocity and solid circulation rates as well as suspension density. The experimentally determined axial dispersion coefficients in this study were in the range of 1.0-3.5 m^2/s.

Key words: Circulating Fluidized Bed, RTD, Peclet Number, Axial Dispersion Coefficient

INTRODUCTION

The circulating fluidized beds (CFB) have found wide application as catalytic and non-catalytic reactors in the chemical process industry. The performance of a CFB reactor is influenced by the mixing of gas and particles [Cho et al., 2000]. Therefore an adequate understanding of the mixing behavior is important and the knowledge on the mixing characteristics is also useful for validation of computer simulation of CFB risers [Sterneus et al., 2000; Kim et al., 2001]. Cankurt and Yerusalmi [1978] suggested that backmixing of gas was negligible compared with the convective flow in risers. However, Brereton et al. [1988] have shown that the CFB as a whole exhibits a considerable amount of backmixing in the gas phase. Recently, it was pointed out that the contradiction is due to the existence of different flow patterns inside the CFB, such as perfect mixing in the dense region and plug flow in the dilute region. Experimental results showed that a significant amount of backmixing and radial dispersion exists near the wall [Patience and Chaouk, 1993; Gayan et al., 1997; Namkung and Kim, 1999; Sterneous et al., 2000]. Li and Wu [1991] have shown that the extent of axial gas dispersion depends on average voidage, which is influenced by the gas velocity and solid circulation rate. These scattered and contradictory results might be due not only to the difficulties of the experimental measurements, but also to the different experimental units and operating conditions used. Thus it is clear with these limited data that the significance of axial dispersion of gas in a CFB as whole is still in question.

In this present investigation the axial dispersion of gas in a circulating fluidized bed was carried out to measure the value of axial dispersion coefficient (D_a) and its variation with the operating conditions.

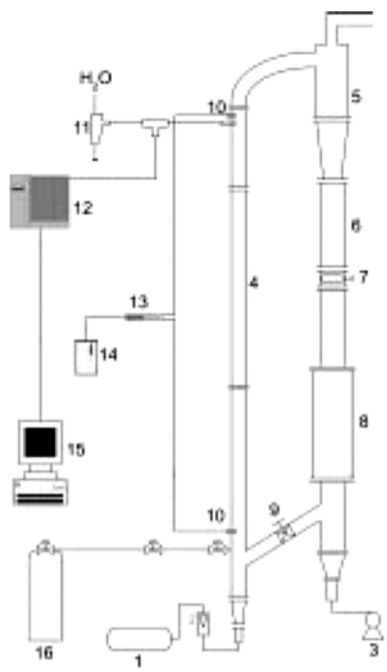

Fig. 1. Schematic diagram of circulating fluidized bed experimental facility.

1. Compressor
2. Flow meter
3. Blower
4. Riser
5. Cyclone
6. Measuring column
7. Butterfly valve
8. Reservoir
9. Check valve
10. Pressure tap
11. Aspirator
12. Mass spectrometer
13. Pressure transducer
14. Scopemeter
15. Computer
16. Carbon-dioxide reservior

EXPERIMENT

A schematic diagram of the experimental apparatus employed in this study is illustrated in Fig. 1. The riser was made of a 0.04 m I.D.×2.8 m high acrylic column. Two pressure taps were mounted at the bottom and at the top of riser in order to measure the pressure drops in the riser. Pressure transducers were connected to pressure taps to convert the pressure drop data into electrical signals. The measured signals of pressure drop in the riser were stored through the data acquisition system. The measured pressure drop data were used to determine the suspension density of the riser. The entrained

†To whom correspondence should be addressed.
E-mail: gyhan@skku.ac.kr

solid particles in the riser were separated from the gas by cyclone connected at the top of the riser and fed back to the bottom of the riser through bubbling bed type particle reservoir. The particle feed rate to the riser was controlled by adjusting the opening of the gate valve and regulating the flow of fluidization air to the particle reservoir. The solid circulation rate (G_s) was measured by temporarily closing the butterfly valve located above the particle reservoir and timing the accumulation of a packed bed of solids along the return column. The solid circulation rates were measured between 20-180 kg/m^2s in the preliminary experiment. The suspension density (solid concentration in the circulating fluidized bed), which depends on the particle density, gas velocity and solid circulation rates, was determined from the measurement of pressure drop gradient and it was found to be in the range of 30-250 kg/m^3. The bed material was silica sand with the average particle diameter (d_p) of 0.069 mm and 0.146 mm. Physical properties of employed particles are shown in Table 1. The experimental ranges of superficial gas velocities (U_g) were 1.99 m/s, 2.65 m/s, and 3.32 m/s for smaller particles and U_g= 3.32 m/s, 3.98 m/s and 4.64 m/s for larger particles respectively. During the experiment, CO_2 was employed as the tracer and the trace gas was introduced through the injection probe at the bottom of the bed. The concentration of the tracer was measured at the top of the bed through a sampling probe and mass spectroscope gave the voltage corresponding to CO_2 concentration of the sampled gas.

In order to uniformly distribute the trace gas into the bed, the injection probe that was made of brass of 5.0 mm ID was horizontally immersed 0.25 m above the distributor and ten pin holes were drilled on the upper surface of the probe to make sure the upward flow of tracer.

Table 1. Physical properties of employed sand particle

Particle	d_p (μm)	ρ_b (kg/m^3)	ρ_s (kg/m^3)	U_{mf} (cm/sec)	U_t (cm/s)
Sand 1	69	1,590	2,800	0.72	30.84
Sand 2	147	1,880	2,780	2.08	106.49

d_p: Surface Median Diameter

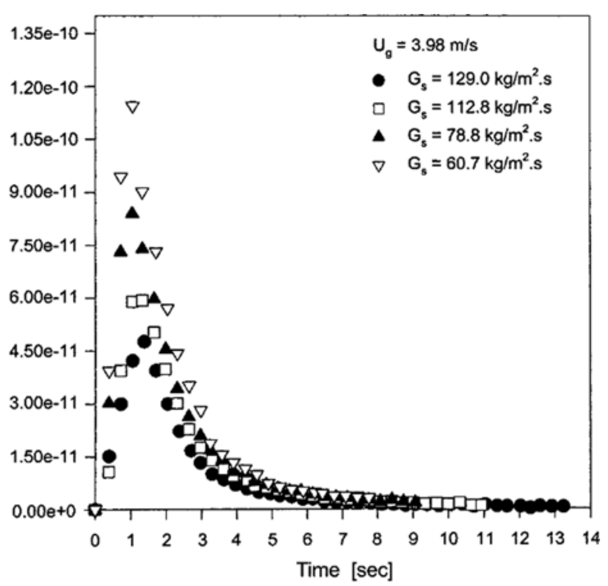

Fig. 2. RTD curves with solid circulation rate using 147 μm silica sand.

RESULTS AND DISCUSSION

1. RTD Curves

The experimentally obtained RTD curves for sand particles of 0.067 mm and 0.147 mm for different solid circulation rates (G_s) when the tracer was injected 0.25 m above the distributor and detected 2.27 m above the distributor are shown in Fig. 2. From the experimental results of Fig. 2, it can be qualitatively said that at the lower solid circulation rate, the gas flows showed close to the plug flow. This means that at the higher solid circulation rates, due to the downward movement of solid particles near the wall, the gas flows deviated from the plug flow. Thus these RTD data confirmed the known facts that solid circulation rates increase gas mixing in a fluidized bed. This may result from the fact that solid mixing closely

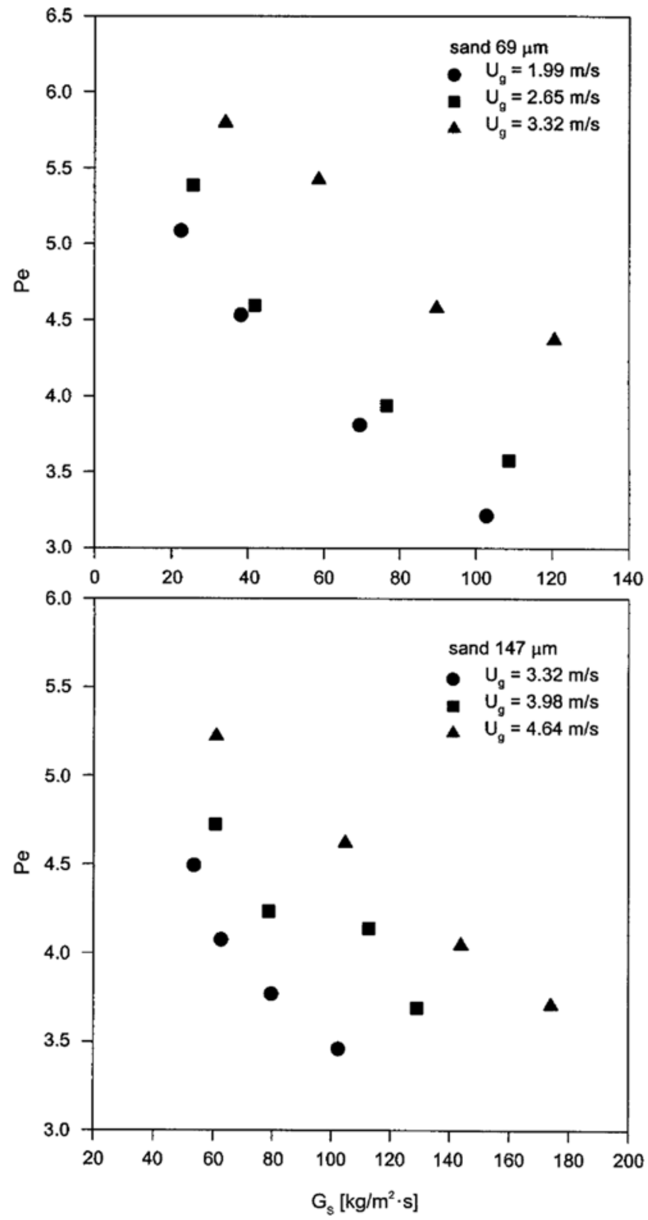

Fig. 3. Peclet number with solid circulation rate.

related to gas mixing and solid mixing cause gas mixing in the riser, thus increasing residence time of tracer gas.

2. Effect of Solid Circulation Rates and Suspension Density on Axial Dispersion Coefficient

The effect of solid circulation rates on the Peclet number is given in Fig. 3 at the different gas velocity for 0.067 mm and 0.147 mm sand particles, respectively. As can be seen in the figures, Peclet number decreased with the solid circulation rate but increased with gas velocity for the range of conditions studied. This trend was mainly due to the fact that Peclet number increased with gas velocity as the definition of Peclet number and decreased with gas turbulence by solid circulation rate. As shown in Fig. 3, the Peclet number with smaller sand particles was larger than that with larger sand particles at a gas velocity of 3.32 m/s and same solid circulation rate. This means that gas turbulence was increased with particle size.

Another important parameter in understanding the flow pattern of circulating fluidized bed flow is the suspension density. It gives information of bed average solid fraction and can be used to predict the heat and mass transfer coefficients. As mentioned before, suspension density depends strongly on the solid circulation rate, gas velocity, particle density and particle size. Fig. 4 shows the effect of suspension density on the axial dispersion coefficient for the sand particles of 0.069 mm and 0.147 mm, respectively. From Fig. 4, it is clear that axial dispersion coefficient increased with suspension density at the given gas velocity. Since at the higher suspension density, the number of particles per unit volume increased and the solid flux to the downward near the wall increased due to the characteristics of dilute upward core flow and dense downward annulus flow in the circulating fluidized bed. Thus, this downward movement of particle cluster retained the upward flow of gas, and the residence time of gas in the circulating fluidized bed near the wall became

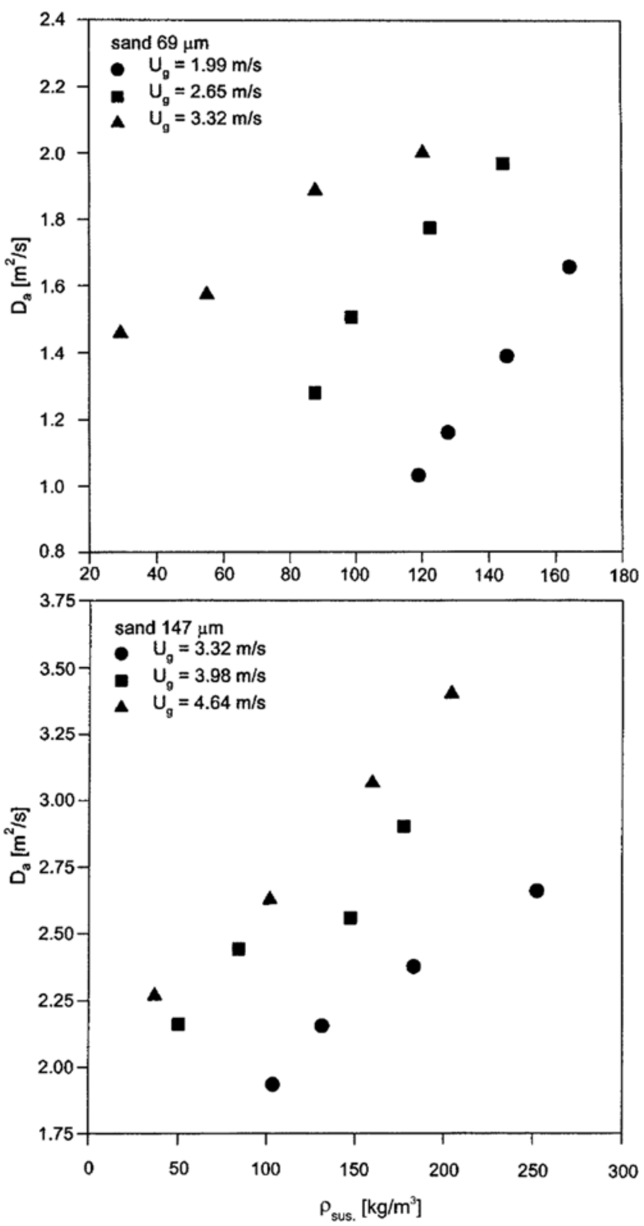

Fig. 4. Axial dispersion coefficient with suspension density.

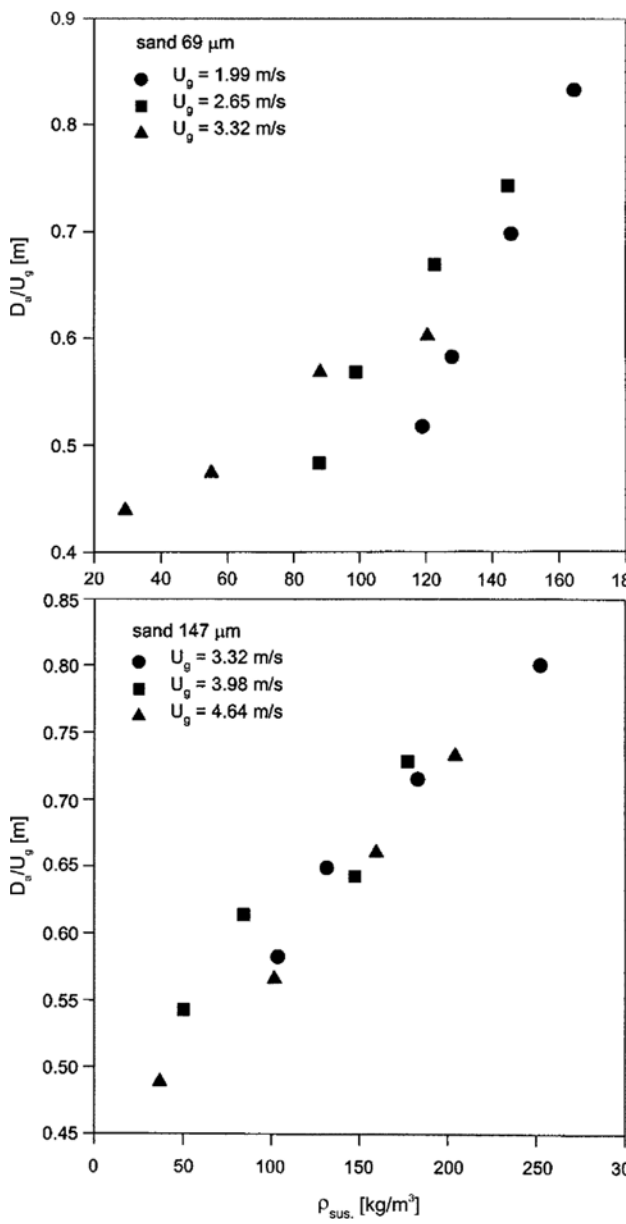

Fig. 5. Normalized axial dispersion coefficient with suspension density.

longer; this caused the wide distribution of residence time of gas in the bed. The axial dispersion coefficient of gas also depended on the gas velocity according to the equation derived from the axial dispersion model. As shown the Fig. 4, axial dispersion coefficient increased with gas velocity at the given suspension density. Li and Weinstein [1989] reported that the axial dispersion coefficient increased with gas velocity at constant ratio.

The experimentally obtained axial dispersion coefficients in this study as shown in Fig. 4 were the result of the coupled effect of gas velocity and suspension density. In order to get the suspension density effect only on the axial dispersion coefficient, the obtained axial dispersion coefficients were normalized by dividing the axial dispersion coefficients to gas velocity. Fig. 5 shows the effect of suspension density on the D_a/U_g. Since the gas velocity effect on the axial dispersion coefficient was eliminated by introducing the normalized coefficient, it was expected that the scattered data points would fall into one line irrespective of gas velocity. As can be seen in Fig. 5, the measured D_a/U_g values showed the linear dependence on the suspension density regardless of operating gas velocity. As shown in Fig. 5, the measured D_a/U_g values were in the range of 0.4-0.85 in the covered operating conditions.

Bai et al. [1992] reported that axial gas mixing in a CFB riser differs substantially from that in simple plug flow. From the analysis of RTD curves of experimental data, the results of this study confirmed that the axial dispersion of gas was considerable and should not be neglected especially at the higher solid circulation rates and gas velocity.

CONCLUSION

An experimental investigation of RTD characteristics and axial dispersion coefficient was performed in a circulating fluidized bed with different operating conditions. In a circulating fluidized bed, mean residence time and variance were increased with solid circulation rate, but decreased with gas velocity. Axial dispersion coefficient was increased with gas velocity and solid circulation rate. The experimentally determined axial dispersion coefficients were in the range of 1.0-3.5 m^2/s. From the analysis of RTD curves of experimental data, it was found that axial dispersion of gas was considerable and should not be neglected, especially at higher solid circulation rates and gas velocity.

NOMENCLATURE

C(t) : concentration of tracer gas [voltage]
D_a : axial dispersion coefficient [m^2/sec]
d_p : particle diameter [mm]
G_s : solid circulation rate [kg/m^2·sec]
Pe : Peclet number [-]
U_g : fluidizing gas velocity [m/sec]

Greek Letter

ρ_{sus} : suspension density [kg/m^3]

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