LATERAL SOLID PARTICLES MIXING IN A CENTRIFUGAL FLUIDIZED BED

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The lateral mixing of solid particles in a centrifugal fluidized bed in which the rotation axis is horizontal was investigated from the unsteady-state behavior of tracer particles. The lateral mixing characteristics were analyzed by a diffusion model identical to that for a conventional fluidized bed.

The lateral dispersion coefficient of particles was affected by variables such as gas velocity, rotational speed of the rotor, particle diameter, bed height and minimum fluidized velocity. From the experimental results an empirical equation for the coefficient was obtained. Its value for a centrifugal fluidized bed was smaller than that calculated from empirical equations for a conventional fluidized bed, because of smaller bubble diameter and bed height, centrifugal force, and the like.

Introduction

In a conventional fluidized bed air flows vertically upward through the distributor, fluidizing the bed against gravitational force acting on its particles. On the other hand, a centrifugal fluidized bed is cylindrical in shape and rotates about its axis of symmetry. As a consequence of the circular motion, the bed particles are forced into an annular region at the circumference of the container. Gas flows radially inward through the cylindrical distributor, fluidizing the bed particles against the centrifugal force generated by the rotation.

It is known that a centrifugal fluidized bed has a number of advantages.[1)] (1) It can be used in much wider ranges of operating conditions and permits larger flow rate per unit area of distributor than a conventional fluidized bed. (2) It is easier to control by manipulating the rotational speed. (3) Smaller particles can be used in it than in a conventional fluidized bed. (4) Its space requirement is much smaller.

However, sufficient studies of centrifugal fluidized beds have not been made. In particular, available data on solid particles mixing in such beds are very few. Kroger *et al.*[3)] reported the mixing characteristics of particles in a centrifugal fluidized bed apparatus having a vertical rotation axis. We experimentally investigated the lateral mixing of particles in a centrifugal fluidized bed using an apparatus rotated about the horizontal axis, and compared it with that in a conventional fluidized bed.

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1. Experimental Apparatus and Procedure

A schematic diagram of the experimental apparatus is shown in **Fig. 1**. The fluidized bed consisted of a cylindrical distributor made of sintered metal elements with mean pore size of 20 μm. The distributor of the cylinder was 14.4 cm I.D. and the width was 4.53 or 9.53 cm. The distributor was covered by a rotor of 28.8 cm diameter which rotated about the horizontal axis.

Air, used as the fluidizing medium, was supplied to the air chamber in the rotor through the hollow drive shaft. It flowed radially inward through the porous surface of the cylindrical distributor. Then it left the bed from an opening of 0.4 cm diameter set in side wall.

The annular partition was made up of interlocking cardboard sectors having dimensions such that they

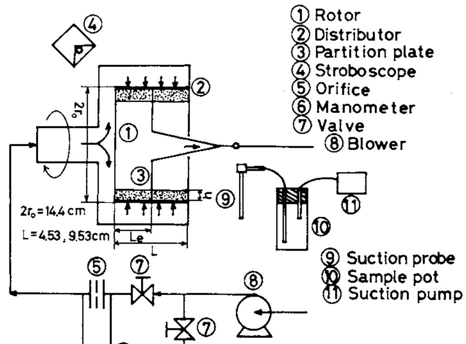

Fig. 1. Experimental apparatus

Table 1. Properties of fluidized particles

	$d_p \times 10^6$ [m]	$\rho_p \times 10^{-3}$ [kg/m^3]	u_{mf} (N=9 1/s)* [m/s]
G-1	63–78	2.40	0.043
G-2	105–125	2.42	0.220
G-3	210–250	2.43	0.632
G-4	250–297	2.44	0.894

* Experimental data for W=0.6 kg.

could be removed through the opening. The strings for removal itself and the pivot to prevent them from getting entangled by rotation were attached to it. At the beginning of the experiments, the partition was placed at a suitable position in the distributor. Then the particles were put in one side, the tracer particles in the other side.

The rotor was rotated and the particles were fluidized at high gas velocity to uniform bed thickness. After setting gas velocity by an orifice flow meter, the particles were fluidized for a few minutes. Then the strings attached to the partition were suddenly pulled and passed through the opening. After a suitable time the fluidization was suddenly stopped and the particles were taken at suitable bed position by the suction probe.

The particles used in the experiments were glass beads with the properties shown in **Table 1**. Particles dyed by red ink were used as tracer. Their characteristics such as repose angle were the same as those of regular particles.

The tracer concentration was measured as follows. For large particles G-3 and G-4, about 2000 particles randomly selected from the sample particles were visually discriminated between colored tracer particles and uncolored regular particles and the bed particles were counted. For G-1 and G-2, a weighted sample of particles was washed with a known volume of acetone and the absorbance of the colored acetone was measured by a spectrophotometer. The wavelength of the absorption peak for the red ink was 490 nm. The relationship between the absorbance and the concentration of tracer particles was linear, as shown in **Fig. 2**. By using this calibration curve, the concentration was measured.

2. Experimental Results

A one-dimensional diffusion model was used to characterize the lateral mixing of solid particles in a conventional fluidized bed.[2,6] The governing differential equation of the model is given as

$$\frac{\partial C}{\partial t}=D_c\left(\frac{\partial^2 C}{\partial x^2}\right) \qquad (1)$$

The initial and boundary conditions are

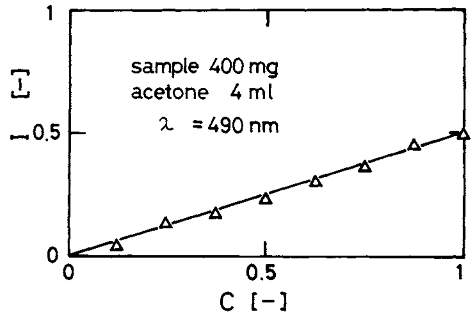

Fig. 2. Relationship between absorbance and tracer particle concentration

$$t=0\,,\quad 0\leqq x\leqq L_e:\quad C=1$$
$$t=0\,,\quad L_e<x\leqq L:\quad C=0$$
$$x=0\,,\quad x=L:\quad \partial C/\partial x=0$$

The analytical solution of Eq. (1) with the conditions above becomes

$$C=\frac{L_e}{L}+\frac{2}{\pi}\sum_{n=1}^{\infty}\frac{1}{n}\times\sin\left(\frac{n\pi L_e}{L}\right)\cos\left(\frac{n\pi x}{L}\right)\exp\left(-\frac{n^2\pi^2}{L^2}D_c t\right) \qquad (2)$$

The lateral dispersion coefficient of particles, D_c, is obtained from Eq. (2). Kroger *et al.*[3] represented this diffusion model as suitable for analysis of the lateral mixing of particles in a centrifugal fluidized bed in which the rotation axis is vertical.

Figure 3 shows the concentration variation of tracer particles at a bed position of 1 cm from the side end wall with elapsed time. The solid line in Fig. 3 represents calculated values from Eq. (2). The experimental data can be expressed by the diffusion model satisfactorly. Of course D_c is not affected by the position where the partition was set.

The lateral mixing of particles in a conventional gas-solid fluidized bed is induced by: (a) bubble movement through the bed, (b) bubble burst at the bed surface, (c) gross particle circulation in the bed.[2] Since the bed height in this centrifugal fluidized bed is very small, there is no gross particle circulation.

Shi *et al.*[6] showed that D_c in a conventional fluidized bed depends on the following factors: $u-u_{mf}$, h_{mf}, d_p, $\rho_p-\rho_g$, μ_g and ρ_g. Additionally, it depends on bed radius, r, and rotational speed, N, in a centrifugal fluidized bed.

Figure 4 shows logarithmic plots of D_c and $u-u_{mf}$. From this figure D_c is proportional to $(u-u_{mf})$. **Figure 5** shows the relationship between D_c and rotational speed N. The slope of the line in this figure is -4.

The pressure drop in a centrifugal fluidized bed is obtained by integrating centrifugal force acting on bed particles between the outer bed (the distributor)

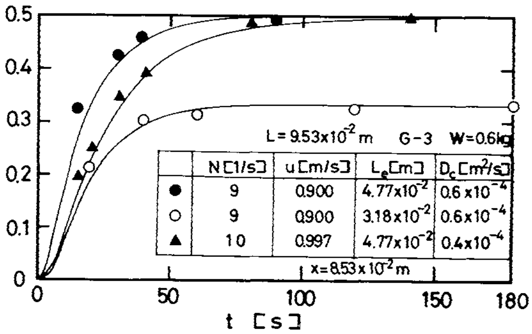

Fig. 3. Variation of tracer particle concentration

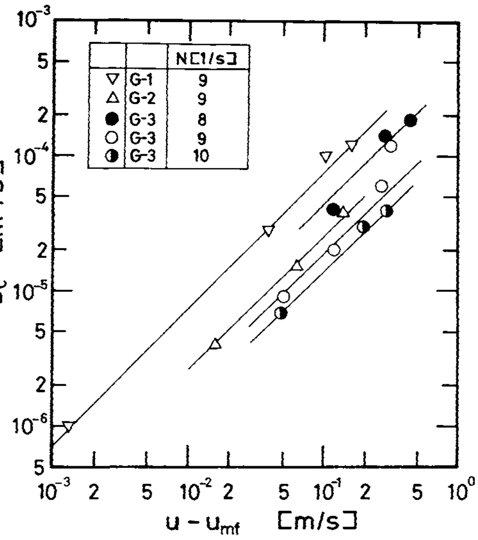

Fig. 4. Effect of excess gas velocity on D_c

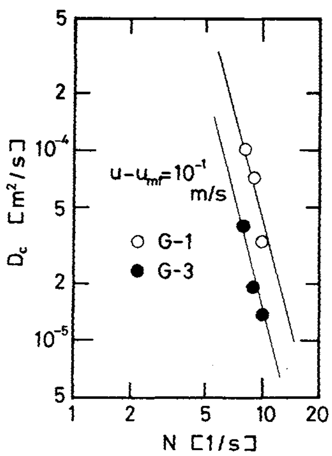

Fig. 5. Effect of rotational speed on D_c

radius r_o and the inner bed surface, r_i.[7)]

$$\Delta P_{\max} = \int_{r_i}^{r_o} 2\pi\rho_p(1-\varepsilon)r^2\omega_o^2 dr/(2\pi r_o)$$

$$= (1/3)\rho_p(1-\varepsilon)\omega_o^2(r_o^3 - r_i^3)/r_o \tag{3}$$

In the fixed-bed region the pressure drop based on the Kozeny-Carman equation is applied:

$$\Delta P = 5S_v^2\frac{(1-\varepsilon)^2}{\varepsilon^3}\mu_f\int_{r_i}^{r_o} u_r dr \tag{4}$$

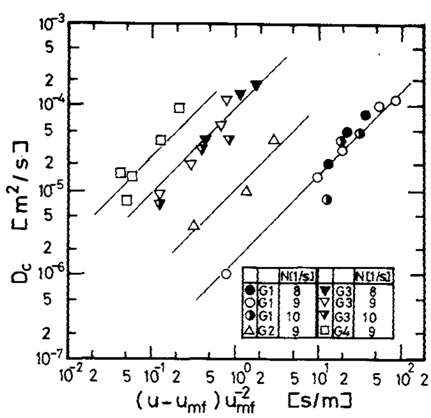

Fig. 6. Relationship between $(u-u_{mf})u_{mf}^{-2}$ and D_c

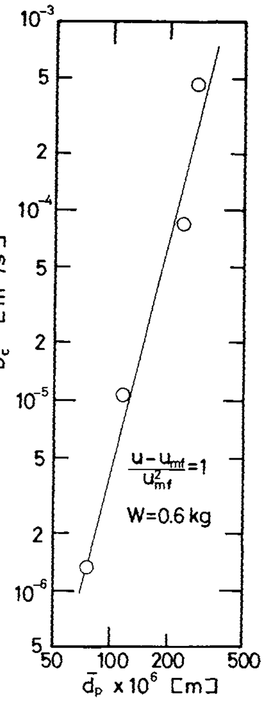

Fig. 7. Effect of particle diameter on D_c

The fluidizing gas velocity assumed to be uniformly distributed over the circumferential distributor on $u = u_o$. Superficial gas velocity at suitable radius u_r is given by

$$u_r = u_o(r/r_o) \tag{5}$$

The average minimum fluidized velocity at the distributor u_{mf} is obtained by equating Eqs. (3) and (4) and substituting Eq. (5) into Eq. (4).

$$u_{mf} = \Delta P_{\max}\alpha/\{\mu_f r_o \ln(r_o/r_i)\} \tag{6}$$

$$\alpha = \varepsilon^3/\{5S_v^2(1-\varepsilon)^2\} \tag{7}$$

Equation (6) shows that u_{mf} is proportional to the square of the rotational speed, ω_o^2. From this relationship and Figs. 4 and 5, we obtain logaritmic plots of D_c and $(u-u_{mf})/u_{mf}^2$ in **Fig. 6**.

The effect of particle diameter, d_p, on D_c is shown in **Fig. 7**. The slope of the line is 4. Hirama *et al.*[2)]

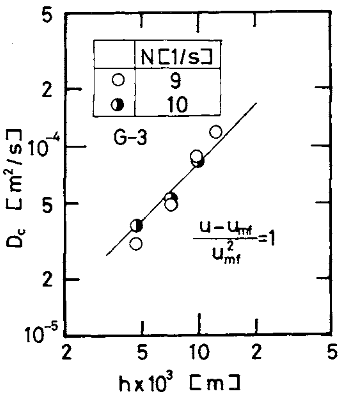

Fig. 8. Effect of bed height on D_c

showed that D_c increased with bubble size on the basis of the so-called bubble model[4] and their experimental data. Toei *et al.*[8] reported that bubble diameter increased with particle diameter, because the frequencies of disturbance and splittings of bubbles decreased with increasing particle diameter in a conventional fluidized bed. Also, in a centrifugal fluidized bed the bubble size increased with particle diameter. We observed them visually at the end wall of the rotor. In additon, the height of the gas jet generated on the distributor becomes larger with increasing particle diameter in a conventional fluidized bed.[5] Because the bed height is very small in a centrifugal fluidized bed, the gas jet height should affect D_c strongly.

Figure 8 shows the relationship between bed height, h and D_c. As expected from the results in a conventional fluidized bed, D_c increases with h in this experimental range, $h < 1.24 \times 10^{-2}$ m. Hirama *et al.*[2] found a bubble breaking up at the upper surface of the bed induced lateral dispersion of particles in a conventional fluidized bed. In a centrifugal fluidized bed bubble size at the bed's inner surface increases with bed height. Also, the radial gas velocity at the bed surface becomes larger with the bed expansion and the elutriation of particles there from increases. This elutriation of particles causes D_c to be large.

Based on the previous discussion, the relationship between D_c and $(u-u_{mf})u_{mf}^{-2}d_p^4h$ shown in **Fig. 9**, gives rise to the following expression.

$$D_c = K(u-u_{mf})u_{mf}^{-2}d_p^4h \qquad K = 3.8 \times 10^{12} \tag{8}$$

3. Discussion

Let us compare lateral dispersion coefficients of particles in a centrifugal fluidized bed with those in a conventional one.

The conventional fluidized bed is a rectangular column, shown in **Fig. 10(b)**. It has a shape which is obtained by expanding the centrifugal fluidized bed shown in **Fig. 10(a)** linearly. **Figure 11** shows experimental data obtained in the centrifugal fluidized bed.

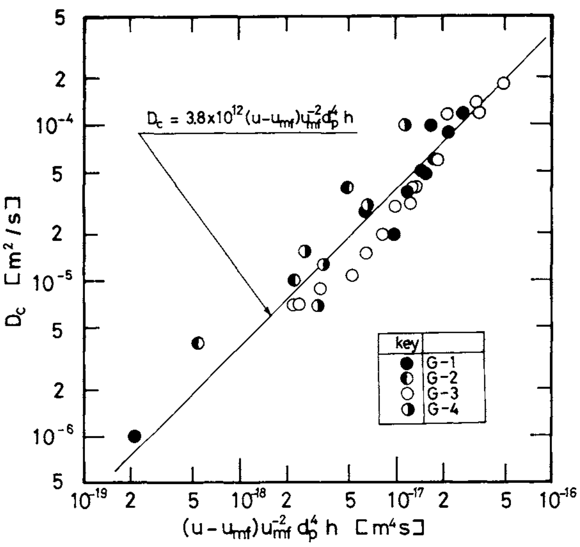

Fig. 9. Relationship between $(u-u_{mf})u_{mf}^2d_p^4h$ and D_c

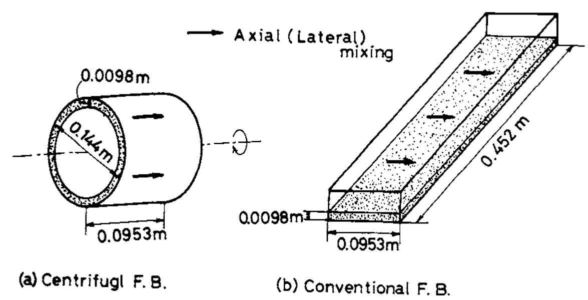

Fig. 10. Centrifugal and conventional fluidized-bed apparatus

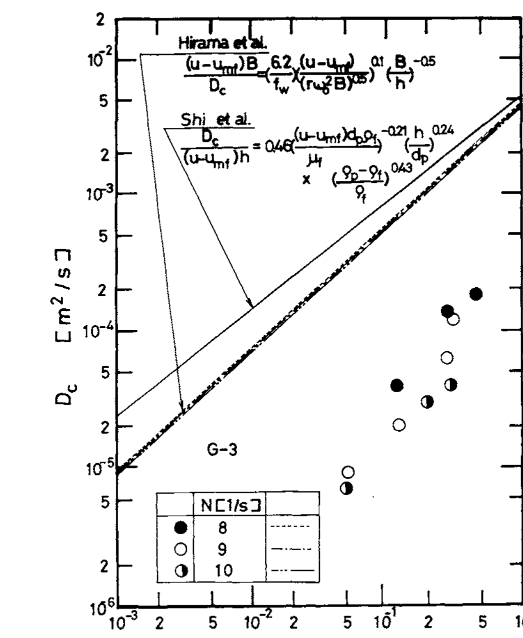

Fig. 11. Comparison of D_c in a centrifugal bed with that in a conventional fluidized bed

Solid lines show the values calculated for the rectangular bed in Fig. 10(b) by applying the empirical equations of Hirama *et al.*[2] and Shi *et al.*[6] In these calculations we changed gravitational acceleration, g, to centrifugal acceleration at the outer bed radius, $r_o\omega_o^2$. The lateral diffusion coefficients in the centrifugal fluidized bed are found to be smaller than the calculated values in the conventional one.

This shows that bubbles in a centrifugal fluidized bed are smaller than in a conventional one. The size of bubbles generated at the distributor are small due to centrifugal force. Also, the frequency of bubble coalescence is low. This is evident by visual observation of bubbles at the end wall of the rotor. Further, lateral dispersion of particles by breaking up at the upper surface of the bed decreases because of centrifugal force on the particles.

Conclusion

The lateral mixing of solid particles in a centrifugal fluidized bed was measured. The data are interpreted by a one-dimensional diffusion model.

The lateral dispersion coefficients of particles are affected by superficial gas velocity, minimum fluidized velocity, rotational speed of the rotor, particle diameter and bed height. An empirical equation, Eq. (8), is obtained.

The lateral solid particle mixing in a centrifugal fluidized bed is less than in a conventional one. This may be caused by smaller bubble size, smaller bed height and reduced elutriation of particles due to centrifugal force.

Nomenclature

C	= tracer particle concentration	[—]
D_c	= lateral dispersion coefficient of particles	[m²/s]
d_p	= particle diameter	[m]
g	= gravitational acceleration	[m/s²]
h	= bed height of fluidized bed above the distributor	[m]
I	= absorbance	[—]
K	= constant in Eq. (7)	[1/(m²·s²)]
L	= distributor width	[m]
L_e	= lateral length from end-wall to partition plate	[m]
N	= rotational speed	[1/s]
ΔP_{max}	= maximum pressure drop	[Pa]
ΔP	= pressure drop	[Pa]
r	= radius	[m]
r_i	= inner-bed height	[m]
r_o	= outer-bed height	[m]
S_v	= specific surface area	[m⁻¹]
t	= elapsed time	[s]
u	= gas velocity	[m/s]
u_{mf}	= minimum fluidized velocity	[m/s]
u_o	= gas velocity through the distributor	[m/s]
W	= weight of packed particles	[kg]
x	= distance along the dispersion direction	[m]
α	= constant defined by Eq. (7)	[1/m⁴]
ε	= void fraction of bed	[—]
λ	= wave length	[nm]
μ_g	= gas viscosity	[Pa·s]
ρ_p	= particle density	[kg/m³]
ρ_g	= gas density	[kg/m³]
ω_0	= rotational speed	[rad/s]

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