Greek Letters

ϵ = porosity
ρ = density, $g \cdot cm^{-3}$
τ = tortuosity factor
θ = calcination and sintering time, h

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Solids Mixing in a Gas-Liquid-Solid Fluidized Bed Containing a Binary Mixture of Particles

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In the practical application using a gas-liquid-solid fluidized bed for chemical or biochemical processing, a size distribution of the solid particles in the bed is commonly encountered. Particle stratification or segregation would occur in the bed, should adequate mixing among particles not be established. This note describes the experimental efforts on the investigation of the solids mixing in a three-phase fluidized bed containing a binary mixture of particles. Qualitative analysis of the mixing states including complete segregation, partial intermixing, and complete intermixing is conducted.

EXPERIMENTAL

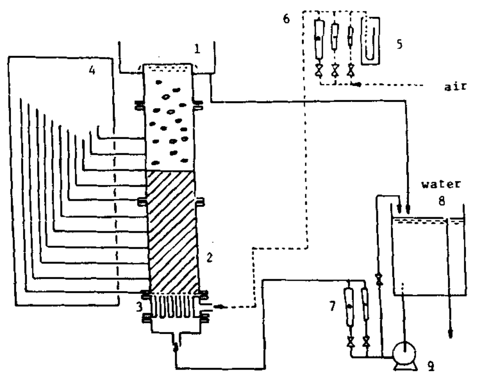
The schematic diagram of the experimental apparatus is shown in Figure 1. The vertical Plexiglas column in the figure has the dimension of 76.2 mm ID with a maximum height of 2.730 m. The column consists of three sections: the gas-liquid disengagement section, test section, and gas-liquid distsributor section. The gas-liquid distributor, which is located at the bottom of the test section, is designed in such a manner that uniform distributions of liquid and gas can be maintained in the column (Fan et al., 1982).

Water and air were used as the liquid and gas phases in the experiment. The gas-liquid flow is cocurrent and upward. Calibrated rotameters were used for the measurement of the gas and the liquid flow rates. Pressure taps are evenly spaced at 51 mm intervals axially on the wall of the test section. The pressure taps were connected to water manometers for the measurement of the static pressure gradient along the column.

In each experiment, two sizes of glass particles were used. The small particles were colored with black paint (Dean and Barry, A-23, wrought iron-flat paint) while the large particles were colored with white paint (Krylon, flat paint). The colored particles were well premixed before placing in the column. Two binary mixtures, including the mixture of 3 and 4 mm particles, and that of 3 and 6 mm particles, are considered in this study. The weight ratio of each size of the particles in the mixture is 1:1.

RESULTS AND DISCUSSION

The qualitative analysis of the solids mixing based on visual observation is presented. Three states of solids mixing are classified in the experiment. These states are complete segregation, partial intermixing, and complete intermixing. It should be noted that although the classification of the states may be to some extent



1	Gas-Liquid Disengagement Section	5	Pressure Gauge
2	Test Section	6,7	Rotameters
3	Gas-Liquid Distributor Section	8	Liquid Reservoir
4	Manometers	9	Liquid Pump

Figure 1. Schematic diagram of the experimental apparatus for the study of solids mixing in a gas-liquid-solid fluidized bed containing a binary mixture of particles.

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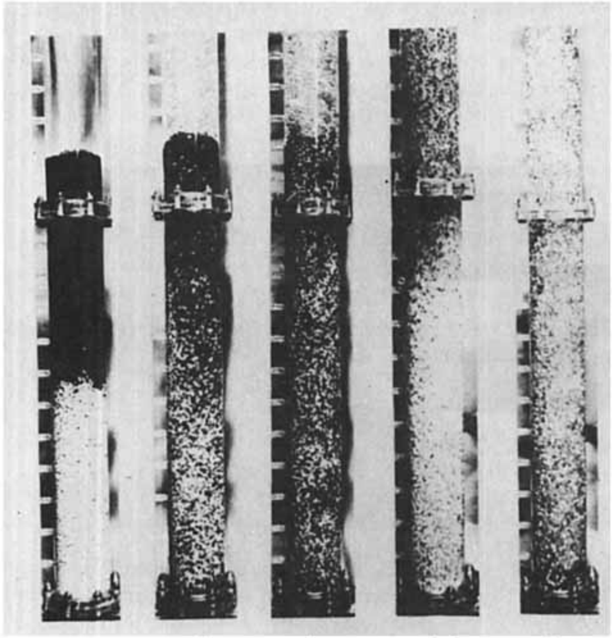

Figure 2. Potographical presentation of solid mixing states in a gas-liquid-solid fluidized bed containing mixed particles of sizes of 3 mm and 4 mm. The operating conditions are given in Table 1.

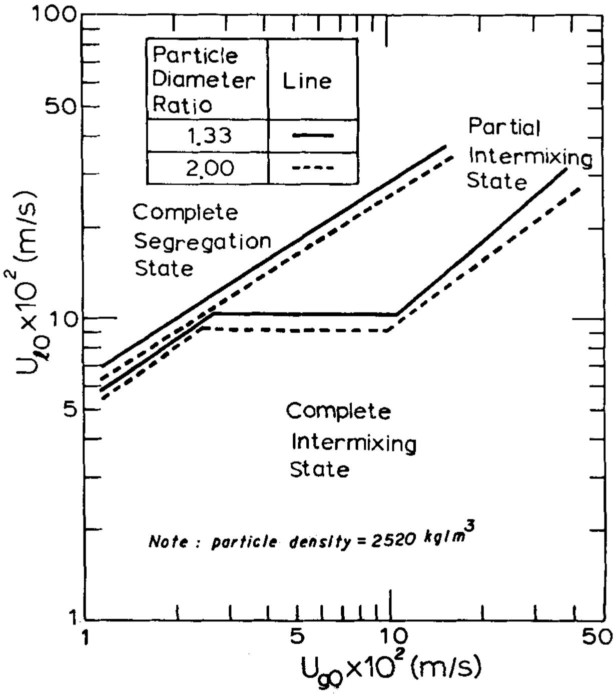

Figure 3. Solids mixing map for binary mixtures in fluidized states with particle diameter ratios of 1.33 and 2.0.

subjective, the macroscopic mixing effects in the system are well represented.

The photographic presentation of various solids mixing states in a gas-liquid-solid fluidized bed containing a binary mixture of the 3 mm and 4 mm particles is given in Figure 2. The operating conditions for the figure are summarized in Table 1. It is seen that when the bed is fluidized solely by a liquid with a velocity of 0.0828 m/s, the bed exhibits a state of complete segregation in which the 3 mm particles and 4 mm particles are well separated in the upper region and the lower region of the bed respectively, Figure 2a. The state of segregation observed under this operating condition is consistent with the criterion for segregation proposed by Wen and Yu (1966). The criterion states that the complete segregation occurs at the particle size ratios exceeding 1.3 for the liquid-solid fluidized bed.

The intermixing between two sizes of particles was, however, observed upon the introduction of the gas. As shown in Figure 2b, the particle intermixing occurs at the gas velocity of 0.0176 m/s. Substantial intermixing takes place only around the interphase of two regions of 3 and 4 mm particles leaving the remainder of the bed with mainly the black particles at the top and the white particles at the bottom. This state of mixing is denoted as partial in-*termixing*. As the gas velocity increases to 0.0352 m/s, both sizes of the particles are well distributed in the entire bed, Figure 2c. This state of mixing is characterized as complete intermixing.

The liquid velocity has a positive effect on the bed segregation. The higher the liquid velocity, the more complete the bed segregation becomes. As shown in Figure 2d, as the liquid velocity increased to as high as 0.169 m/s, the bed returns to the intermixing state even with the increase of the gas velocity from 0.0352 to 0.075

TABLE 1 OPERATING CONDITIONS FOR FIGURE 2

Figure	W_1 (kg)	W_2 (kg)	$U_{l0} \times 10^2$ (m/sec)	$U_{g0} \times 10^2$ (m/sec)
2-a	1.548	1.548	8.28	0
2-b	1.548	1.548	8.28	1.76
2-c	1.548	1.548	8.28	3.52
2-d	0.700	0.700	16.9	7.51
2-e	0.700	0.700	16.9	12.4

m/s. As the gas velocity increases to as high as 0.124 m/s and the liquid velocity remains constant at 0.169 m/s, the complete intermixing state is regained, Figure 2e.

Figure 3 shows the solids mixing map for binary mixtures of the particles with diameter ratios of 1.33 and 2. The map illustrates the effect of the gas and liquid flow rates on the three solids mixing states including complete intermixing, partial intermixing, and complete segregation. For the particle diameters ratio of 1.33, the boundary between the complete segregation and partial intermixing occurs at the liquid velocity of 0.07 m/s and gas velocity of 0.015 m/s. With the increase of the liquid velocity, the boundary moves responding with an approximately linear increase of the gas velocity. The boundary between the partial intermixing state and complete intermixing state occurs at the gas velocity of 0.015 m/s and the liquid velocity of 0.058 m/s. As the liquid velocity increases to 0.105 m/s the boundary moves responding with an approximately linear increase of the gas velocity to 0.027 m/s. It is interesting to note that at the liquid velocity of 0.105 m/s, an increase in the liquid velocity would require an approximately linear increase of the gas velocity for the bed to be maintained in a complete intermixing state.

For the particle diameter ratio of 2, a similar relationship between the liquid flow rate and gas flow rate defining the boundary of the mixing states is observed. However, it is seen that for a given gas velocity the required liquid velocity to reach the boundary for diameter ratio of 2 is lower than that for the diameter ratio of 1.33.

Figure 4 shows the map of the flow regimes determined based on visual observation for binary mixture of the particles with diameter ratios of 1.33 and 2. The flow regimes are defined as functions of the gas and liquid velocities. Four distinct flow regimes are classified according to Muroyama et al. (1978). They are the coalesced bubble regime, dispersed bubble regime, transition regime (between dispersed bubble regime and slugging regime) and slugging regime. Figure 4 shows that the relationship between the velocities of the gas and liquid, and the flow regimes is similar for the particle diameter ratios of 1.33 and 2. Furthermore, this relationship is similar to that reported by Muroyama et al. (1978) for the three-phase fluidized bed operating with monosize glass particles ranging from 2.6 to 6.9 mm in diameter. However, it is noted that, for a given gas velocity, the liquid velocity required to reach the boundary betweeen the dispersed bubble regime and coalesced

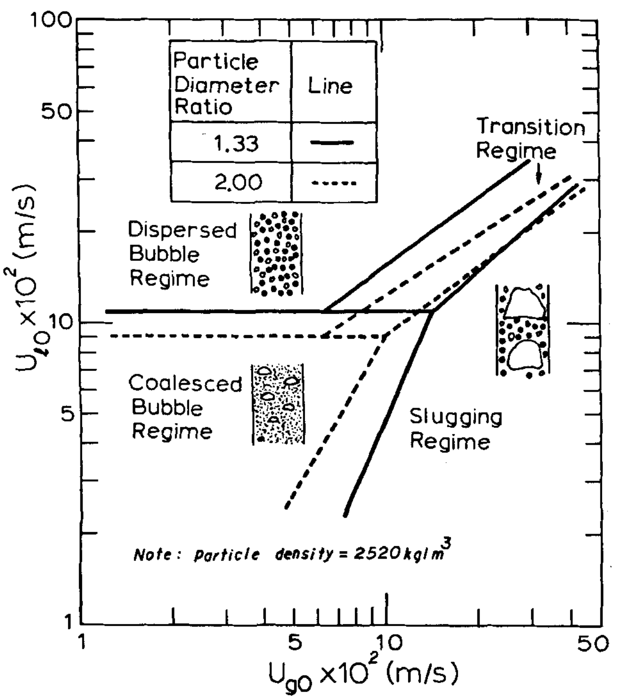

Figure 4. Flow regime map for binary mixtures in fluidized states with particle diameter ratios of 1.33 and 2.0.

bubble regime for the particle diameter ratio of 1.33 is higher than that for the particle diameter ratio of 2. Likewise, for a given liquid velocity, the gas velocity required to reach the boundary between the coalesced bubble regime and slugging regime for the particle diameter ratio of 1.33 is higher than that for the particle diameter ratio of 2.

Comparing Figure 4 with Figure 3, it is important to note that there is a strong correlation existing between the flow regime map and the solids mixing map for the binary mixtures considered. Evidently, the complete mixing state occurs largely in the coalesced bubble regime and the slugging regime and slightly in the transition regime. The partial intermixing state occurs largely in the dispersed bubble regime and the transition regime and slightly in the coalesced bubble regime. The complete segregation states occur solely in the dispersed bubble regime.

ACKNOWLEDGMENT

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NOTATION

W_1 = weight of particle 1, kg
W_2 = weight of particle 2, kg
U_{l0} = superficial liquid velocity, m/s
U_{g0} = superficial gas velocity, m/s

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Effect of Distributor to Bed Resistance Ratio on Uniformity of Fluidization

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The effect of the distributor-to-bed resistance ratio on the uniformity of fluidization is analyzed. A new model for the cause of channeling through a fluidized bed is proposed, and a criterion for uniform fluidization, giving rise to the condition of full fluidization, is established.

The performance of a fluidized bed depends largely on the satisfactory design of its distributor. Apart from supporting the weight of the bed during downtime and preventing the back flow of particles into the plenum section, the major function of the distributor is to distribute the fluidizing medium across the base of the bed so that the fluidized condition is maintained over the entire cross section. It has been known that to maintain a stable operation of the bed, the pressure drop of a certain magnitude needs to be established through the distributor.

Numerous papers have been published on the determination of the required distributor pressure drop or on its relation to the bed pressure drop (Hiby, 1964; Siegel, 1976; Mori and Moriyama, 1978; Sathiyamoorthy and Rao, 1979; Qureshi and Creasy, 1979). Among approaches presented in these papers, that based on Siegel's model is considered to be the simplest for deriving a stability criterion for a fluidized bed (Qureshi and Creasy, 1979); it has been used to calculate the relative pressure drop through a porous or perforated distributor for stable fluidization (Qureshi and Creasy, 1979; Sirotkin, 1979). The objectives of this work are to propose a new or improved model for the cause of channeling in a fluidized bed through critical analysis of Siegel's model and to establish a new criterion for uniform fluidization, giving rise to the condition of full fluidization.

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