

INCREASED RETENTION OF FINES IN A CIRCULATING SYSTEM OF FLUIDIZED PARTICLES

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Introduction

A circulation system of fluidized particles is considered to have a number of advantages over the conventional fluidized bed system:^{1,2-5,7-9)} 1) uniformity of bed temperature, 2) reduced segregation, 3) rapid dispersion of fed particles and 4) prevention of agglomeration. Application of the system to coal combustion and gasification,^{1,5)} biomass gasification,⁴⁾ drying⁷⁾ and chemical vapor deposition of silicon⁸⁾ have been reported. A new gasification process with a circulation system of fluidized particles has been developed to produce medium BTU gas without using oxygen where the gasification zone and combustion zone are separated.^{2,5)} In addition to the above advantages, Judd *et al.* suggested a reduction of elutriation of fine particles ($dp=20\mu\text{m}$) and an increase in holdup within the circulation system of particles.¹⁾

These findings by Judd *et al.* are worth noting, and further quantitative evaluation is needed. This paper reports the results of experimental work on the residence time distribution of fine particles introduced as an impulse into fluidized beds with and without internal circulation of particles. The effects of internal circulation and location of feed point on the elutriation rate and the amount of initial elutriation are clarified, and the effectiveness of the circulation system for CVD and coal conversion processes is evaluated.

1. Experimental

1.1 Apparatus and materials

Experiments were conducted in a two-dimensional transparent fluidized bed ($0.4\text{m} \times 0.1\text{m}$ in cross section) which was divided by a partition plate to form two beds with different gas velocities. The configuration of the bed is shown in Fig. 1. The bed was equipped with two taps ("A" and "B") of 28 mm i.d. for injection of tracer particles. The distributor was

divided into two parts, the left part with sixteen holes (0.8 mm in diameter) and the right part with sixteen holes (1.3 mm in diameter). The plate was removed for conventional fluidized bed experiments without circulation. A sampling tube (8 mm o.d. 6 mm i.d.) was located at the center of the duct. The tube was connected to an air pump through a sampling filter.

Air was employed as the fluidizing gas. The bed materials employed were micro-spherical particles with particle density of $1540\text{kg}\cdot\text{m}^{-3}$, shape factor of 1.0 and average diameter of 0.162 mm (0.05–0.25 mm). The void fraction and the velocity at minimum fluidization of the particles at 300 K were respectively 0.51 and $1.4\text{cm}\cdot\text{s}^{-1}$. The particles were charged in the bed to a static height of 450 mm. The void fraction in the packed bed was 0.49, and thus the total weight of the bed materials was 14.1 kg.

Fine alumina powders with particle density (true density) of $3980\text{kg}\cdot\text{m}^{-3}$ and diameter of $3.0\mu\text{m}$ were used as the tracer. The calculated terminal velocity of a single particle in the air at 300 K was $1.1\text{mm}\cdot\text{s}^{-1}$.

1.2 Procedure

In the present experiments, the gas velocity was much higher than the terminal velocity of a tracer particle, and the value of TDH (transport disengaging height) calculated by Zenz and Weil's equation¹⁰⁾ was smaller than the height of the free board. Therefore, it

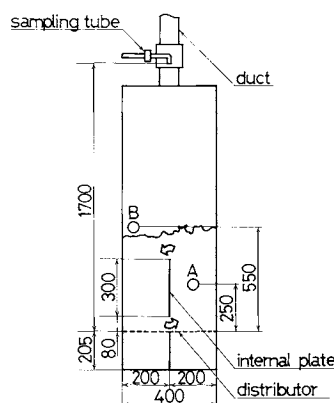


Fig. 1. Two-dimensional fluidized bed.

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was considered that the elutriation rate⁶⁾ of the tracer particles could be obtained by the present method. The suction velocity in the sampling tube was adjusted to be equal to the average velocity of air in the duct.

A sheet of filter paper completely dried in a desiccator (more than fifteen hours) was set in the sampling line. Under the experimental conditions, the bed was operated for more than 60 min, and the moisture content in the filter paper was equilibrated with the humidity of the air stream. An impulse of 3.0 g of tracer particles was introduced through an injection. The weight change of the filter which captured the elutriated tracer particles was measured intermittently. After approximately ten hours' operation, the bed was operated at a gas flow rate of more than 6 U_{mf} for more than six hours to purge the remaining tracer particles.

Supplementary experiments with circulation showed that tracer particles don't accumulate in the dead space located at the bottom of the bed of descending particles.

2. Results

The total amount of tracer particles collected on the filter paper up to time t , $M(t)$, is expressed as a fraction of $M(\infty)$, the corresponding quantity after infinite time. $M(\infty)$ can be expressed by

$$M(\infty) = Q_0 A_s / A_a \quad (1)$$

where Q_0 is the total amount of tracer particles introduced and A_s/A_a is the ratio of the cross-sectional area of the sampling tube to that of the air duct. The value of $M(t)/M(\infty)$ is plotted against time. Typical results are shown in Fig. 2. As shown in Fig. 2, the value of $M(t)/M(\infty)$ seems to approach the expected value of unity.

The initial weight fraction of tracer particles in the bed is at most 0.02%. Thus the amount of tracer particles left in the system is expected to be formulated as⁶⁾

$$Q/Q_0 = \exp(-kt) \quad (2)$$

where Q is the amount of the tracer particles left in the system, given as

$$Q = Q_0 - M(t) A_a / A_s \quad (3)$$

Typical examples of the $\log(Q/Q_0)$ vs. time curve are shown in Fig. 3. As shown in Fig. 3, the data points fall on a straight line after almost three hours following the tracer injection. The slope of the line is dependent on the superficial gas velocity and is affected by the particle circulation. The difference in the injection point does not affect the slope but does affect the absolute value.

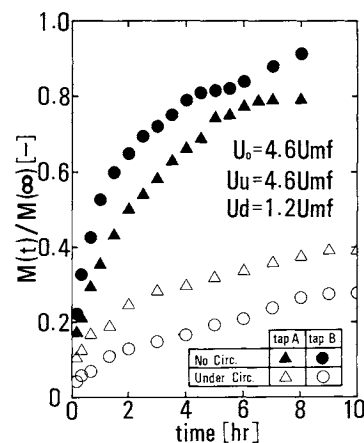


Fig. 2. Time variations of cumulative amount of tracer particles.

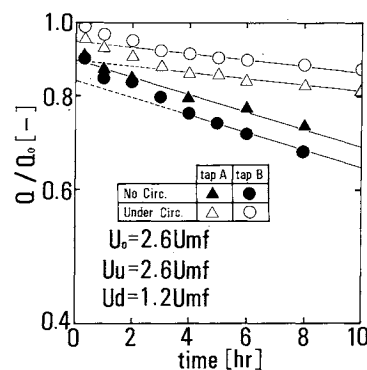


Fig. 3. Typical examples of $\log(Q/Q_0)$ vs. time curve.

3. Comparison between Presence and Absence of Circulation for Same Gas Velocity

In some experimental runs, it was confirmed that the relation between the mass circulation rate and the gas velocity was almost linear. Therefore, in the following analyses, the latter was chosen as an analytical parameter.

The data sets are compared for the same gas velocity in the main bed: the superficial gas velocity without circulation, U_0 , and that of gas introduced into the bed of ascending particles under the circulation operation, U_u . The superficial velocity of gas introduced into the bed of descending particles, U_d , was kept at 1.2 U_{mf} .

3.1 Amount of initial elutriation

The fraction of tracer particles left in the bed after initial elutriation, Q^*/Q_0 , was defined as the value at the intersection point between the straight line and the ordinate in Fig. 3. The initially elutriated fraction of the injected tracer particles, $1 - Q^*/Q_0$, defined above is plotted in Fig. 4 against the normalized superficial velocity, U_0/U_{mf} , for operation without circulation and against the normalized velocity of gas introduced into the main bed of ascending particles through the

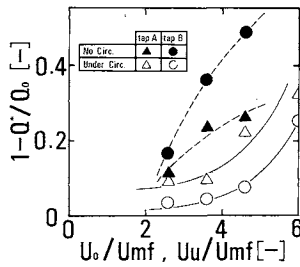


Fig. 4. Initially entrained fraction of injected particles under same gas flow rate in main bed.

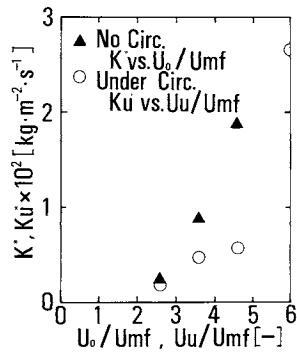


Fig. 5. Effect of circulation of particles on elutriation rate constant of main bed under same gas flow rate in main bed.

right distributor, Uu/Umf , for circulation operation. It is clearly suggested that injection of the particles above the surface of the bed of descending particles (tap B) reduces the initial elutriation in circulation operation, while injection from tap B increases the initial elutriation in operation without circulation. The initial elutriation of the particles injected from tap B in circulation operation was the least among all conditions employed.

3.2 Elutriation rate constant

The elutriation rate constant, K^* , is defined as follows⁶⁾ for the conventional fluidized bed.

$$K^* = kWp/At \quad (4)$$

where Wp is the total weight of the bed materials. The values of the constant, K^* , under the present experimental condition are plotted against the normalized superficial gas velocity, Uo/Umf , in Fig. 5.

The elutriation rate of tracer particles from the main bed of ascending particles in circulation operation was evaluated as follows. The rate constant, defined as

$$Ku^* = kWp/(At/2) \quad (5)$$

is also plotted against the normalized velocity of gas introduced into the main bed, Uu/Umf , in Fig. 5. In the present evaluation, the value of Ku^* may be overestimated, because the elutriation from the descending particles part, where the actual gas flow rate is less than $1.2 Umf$, is included in the value of Ku^* .

However, the error will be negligibly small as is shown in Fig. 5. Furthermore, the error causes overestimation of the elutriation rate constant. Thus it is concluded that the constant in circulation operation, Ku^* , is much smaller than the value of K^* . The reason why the elutriation from a bed is suppressed by particle circulation is not fully elucidated, but it is presumed that the lateral motion of the bed materials (coarse particles) above the free board may be accompanied by entrained fine particles into the bed of descending particles.

Conclusion

The elutriation rate constant of the bed of ascending particles with circulation was much smaller than that of the conventional fluidized bed without circulation for the same local gas velocity. The results suggested that the increased retention of fine particles can be realized by employment of the bed of descending particles. Under the same local gas velocity, initial elutriation is also drastically reduced by injection above the surface of the bed of descending particles. The increased retention of fines will be advantageous for CVD and coal gasification processes, in which a relatively large gas flow rate in the main bed is necessary to prevent clinker or clogging trouble.

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Nomenclature

| | | |
|-------------|---|--|
| Aa | = cross-sectional area of air duct | [m ²] |
| As | = cross-sectional area of sampling tube | [m ²] |
| At | = cross-sectional area of fluidized bed | [m ²] |
| K^* | = elutriation rate constant | [kg·m ⁻² ·s ⁻¹] |
| Ku^* | = elutriation rate constant for part of ascending particles | [kg·m ⁻² ·s ⁻¹] |
| $M(t)$ | = cumulative amount of collected tracer particles | [mg] |
| $M(\infty)$ | = $Q_0 As/Aa$ | [mg] |
| Q | = amount of tracer particles left in system | [mg] |
| Q^* | = amount of tracer particles left in bed after initial elutriation | [mg] |
| Q_0 | = amount of tracer particles introduced | [mg] |
| Ud | = superficial gas velocity introduced through left part of gas distributor | [m·s] |
| Umf | = superficial gas velocity at minimum fluidization | [m·s] |
| Uu | = superficial gas velocity introduced through right part of gas distributor | [m·s] |
| Uo | = superficial gas velocity in conventional bed operation | [m·s] |
| Wp | = total weight of bed materials | [kg] |

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STYRENE SLURRY POLYMERIZATION WITHIN SUSPENDED DROPLETS IN WATER USING 2,2'-AZOBIS(2,4-DIMETHYL-VALERONITRILE) AS INITIATOR

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Key Words: Slurry Polymerization, Initiator Efficiency, Styrene, Microcapsule, Polymer Phase, Solvent Phase

In a previous paper,¹⁾ a slurry polymerization of styrene within isooctane droplets suspended in water considered to be a useful method for manufacturing microcapsules by *in situ* polymerization, was kinetically examined. However, a critical defect in its practical application was the extremely slow reaction rate, which came from the use of 2,2'-azobisisobutyronitrile (AIBN) as initiator. To improve this point, 2,2'-azobis(dimethyl-valeronitrile), the decomposition rate of which is 4.5 times that of AIBN, was used as initiator for the polymerization. The applicability of the previously proposed kinetic equations¹⁾ for this polymerization system was also studied.

1. Polymerization Kinetics

This polymerization could be described by a two-phase separation model.¹⁾ In this model, the system* is assumed to be separated into two homogeneous phases: a solvent phase containing no polymer, and a polymer-rich phase. Concentration equilibria of monomer, solvent and initiator between the two phases are reached. The gel effect would be predominant only in the polymer-rich phase. The following

rate equation for monomer has been derived.

$$R_p = \frac{1}{\delta} \frac{(2k_d f)^{1/2} A_0 (1-x) I^{1/2} / \Phi}{\{\chi_v(1-Q_A) + Q_A\} \{\chi_v(1-Q_I) + Q_I\}^{1/2}} \times \{\chi_v + \gamma(1-\chi_v)Q_A Q_I^{1/2}\} \quad (1)$$

The rate for dead polymer having r -monomer units ($=dM_r/dt$) was also derived in the previous paper.¹⁾ The number and weight average degrees of polymerization are written as Eqs. (2) and (3), respectively.

$$\bar{P}_N = \int R_p dt / \int \Sigma \left(\frac{dM_r}{dt} \right) dt \quad (2)$$

$$\bar{P}_w = \int \Sigma r^2 \left(\frac{dM_r}{dt} \right) dt / \int R_p dt \quad (3)$$

Kinetic parameters used in the above equations are constants or variables depending on initial monomer concentration and monomer conversion.

Since the following Eq. (4) is obvious in the initial stage of the polymerization, the plotting of $1/\bar{P}_N$ versus I_0/R_{p0} at various initiator concentrations at the same monomer concentration would show a linear relationship.¹⁾

$$\frac{1}{\bar{P}_N} = T_{fA} + T_{fA} \frac{S_0}{A_0} + k_d f \frac{I_0}{R_{p0}} \quad (4)$$

($T_{fA} + T_{fA} S_0/A_0$) and $k_d f$ are obtained from the ordinate value at $I_0/R_{p0} = 0$ and the slope, respectively.

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* No effect of water on the polymerization kinetics is considered. Hence, the organic phase alone is assumed to form the system.