CONTINUOUS DRYING OF A FINE PARTICLES-WATER SLURRY IN A POWDER-PARTICLE FLUIDIZED BED

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Key Words: Drying, Fine Particle, Powder, Slurry, Fluidization, Fluidized Bed

The powder-particle fluidized bed is a new type of fluidized bed developed to fluidize fine particles to prevent their agglomeration in the bed.

Continuous drying of a slurry of fine particles with water was carried out in a powder-particle fluidized bed, and its drying characteristics were examined. Moisture in the slurry evaporated almost fully in the bed, and the moisture content of the exiting fine particles was nearly zero under the range of operating conditions in this study. Temperature and relative humidity of the outlet gas agreed with those calculated by mass and heat balances across the bed without knowledge of the rates of heat and mass transfer.

Introduction

Recently, use of fine particles has been increasing in many industries such as ceramics, electronics and

medicine because of their attractive properties. Solid fine particles are often produced in a liquid solvent and separated from it by drying the fine particles solvent slurry. It is important to develop an efficient dryer of the fine particles.

The fluidized bed has been widely used in gas-

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Fig. 1. Schematic diagram of experimental apparatus

solid contact operations and has been applied to the drying of solid particles. As is well known, it is rather difficult to fluidize fine particles (under $20\text{ }\mu\text{m}$ in diameter) of group C in the Geldart classification¹⁾ in a normal fluidized bed, because the fine particles easily agglomerate.

The authors have developed a new type of fluidized bed to fluidize fine particles.^{2,3)} The principle of the new fluidized bed is as follows. A powder of fine particles is fed to a bed in which coarse particles are fluidized. An agglomerate of the fine particles is crushed by collisions with coarse particles, so the fine particles are uniformly dispersed in the bed and fluidized. Only the fine particles are elutriated from the column, based on the difference in the terminal velocity of the particles. The fine particles seem to move through the bed while adhering to the coarse particles. The average residence time of the fine particles in the bed was considerably longer than that of the gas.³⁾ We call this type of fluidized bed a powder-particle fluidized bed. It is expected to be applied as a dryer of fine particles.

There have been some reports about the drying characteristics of various fluidized bed dryers.⁴⁻⁸⁾ Toei *et al.*^{7,8)} showed that the falling drying rate period was proportional to the free moisture content in particles when the particles were uniformly dispersed in the dryer. Kato *et al.*⁴⁾ investigated the drying characteristics of a packed fluidized-bed dryers. However, the drying characteristics of the powder-particle fluidized bed as a dryer of fine particles have not yet been made clear.

In this study, drying of a slurry of alumina fine particles in water was carried out in a powder-particle fluidized bed. The slurry was continuously fed to a bed in which coarse particles were fluidized by dried hot air. Effects of operational conditions on the

relative humidity of outlet gas and on the drying efficiency were examined.

Experimental apparatus and procedure

Figure 1 shows a schematic diagram of the experimental apparatus.

The fluidized bed consisted of a cylindrical column of polycarbonate, 10.4 cm I.D. and 100 cm high. A perforated plate of polycarbonate with 3.0 percent open area (348 holes of 1 mm I.D. in 5 mm pitch) was used as a gas distributor. Air from a compressor was passed through an oil filter and dehydration towers filled with silica gel. The dried air was heated to a certain temperature by an electric heater and was fed to the bottom of the column. The flow rate of the air was measured with orifice meters. The column and the pipes were covered with glass wool to minimize heat loss. Silica particles of average diameter $225\text{ }\mu\text{m}$, $322\text{ }\mu\text{m}$ and $460\text{ }\mu\text{m}$ were used as coarse particles and were fluidized by dried hot air.

Alumina or silica particles of $1.8\text{ }\mu\text{m}$, $2.6\text{ }\mu\text{m}$ and $13.0\text{ }\mu\text{m}$ diameter were used as fine particles. These fine particles were not porous. The fine particles were mixed with water in a stirred tank. The fine particles-water slurry was fed to the fluidized bed of coarse particles through a silicon tube of 3 mm I.D. by the micro-tube pump shown in the figure. The nozzle for slurry feeding was 44 cm above the distributor and 2 cm inside the column wall. The wet fine particles were dried by contact with the dried hot air being dispersed uniformly in the bed. Only the fine particles were elutriated from the column, based on the difference in terminal velocity between the fine particles and the coarse particles as shown in Table 1. The fine particles elutriated were collected by a cyclone.

Temperatures of the inlet gas T_{d1} , the outlet gas

Table 1. Properties of the fine and coarse particles

		$d_p[\mu\text{m}]$	$\rho_p[\text{kg/m}^3]$	$U_{mf}[\text{m/s}]^*$	$U_t[\text{m/s}]^*$
Fine particles	Activated alumina	1.8	3970	3.3×10^{-6}	3.0×10^{-4}
		13.0	3970	1.7×10^{-4}	1.6×10^{-2}
	SiO ₂ flux	2.6	2500	4.4×10^{-6}	4.0×10^{-4}
Coarse particles	Silica sand	225	2700	0.034	1.9
		322	2700	0.070	2.7
		460	2700	0.144	3.9

*Calculated.

T_{d2} and the fluidized bed T_b respectively were measured at 2 cm below, 50 cm above and 5 cm above the distributor with chromel-alumel thermocouples as shown in the figure. The humidity of the outlet gas was also measured. The relative humidity at the outlet was defined as follows.

$$H_2^* = H_2 / H_{w2} \quad (1)$$

where H_2 is the absolute humidity of the outlet gas and H_{w2} is that of saturated gas. At the gas exit of the cyclone, both wet- and dry-bulb temperatures were also measured. Fine particles collected in the sampling bottle were dried at 403 K for 2 hr in an electric dryer. The moisture content of the fine particles at the cyclone, w_c , was obtained from the change in weight by this procedure.

Properties of the fine and coarse particles and the operational conditions used in this study are shown in **Tables 1** and **2**, respectively.

Here we define a drying efficiency η_d as follows.

$$\eta_d = \frac{\bar{G}_{12} \bar{C}_{H12} (T_{d1} - T_{d2})}{\bar{G}_{1w} \bar{C}_{H1w} (T_{d1} - T_{w1})} \quad (2)$$

where $\bar{G}_{ij}$ and $\bar{C}_{Hij}$ are respectively the average of total mass flow rate of gas and that of humid heat between i and j , and T_{w1} is the wet-bulb temperature at the inlet. The drying efficiency is the ratio of heat used for drying to that of input. It corresponds to the thermal efficiency. We can evaluate the drying operation from an energy-saving point of view by the drying efficiency.

Effects of feed condition of the inlet gas, feed condition of the slurry and other operational conditions on the moisture content of exiting fine particles and relative humidity of exiting gas were examined.

Results and Discussion

Figure 2 shows the relationship between time elapsed after start-up and bed temperature T_b , outlet gas temperature T_{d2} , dry- and wet-bulb temperatures at the cyclone, T_{dc} and T_{wc} respectively, and moisture content of the fine particles collected by

Table 2. Operational conditions used in this study

$U[\text{m/s}]$	0.318, 0.397, 0.476, 0.555, 0.635, 0.715
$L_f[\text{m}]$	0.05, 0.10, 0.15
$d_{pf}[\mu\text{m}]$	225, 322, 460
$d_{ps}[\mu\text{m}]$	1.8, 2.6, 13.0
$T_1[\text{K}]$	363, 373, 383, 393, 403
$w_s[\text{g/s}]$	0.097~0.212

Fig. 2. Time progresses of temperatures at each point and moisture content of outlet fine particles

the cyclone w_c . T_b , T_{d2} and T_{dc} gradually decreased with time and become almost constant after 25 minutes. About 25 minutes was necessary to establish a steady state for this equipment. T_{d2} was almost equal to the bed temperature T_b . This means that the heat loss from the bed was negligibly small. On the other hand, the gas temperature measured at the cyclone, T_{dc} , was almost 10 K below that of the bed, T_b , suggesting that the heat loss between the bed and the cyclone was not negligible. T_{wc} became constant within a few minutes. The moisture content of the fine particles collected, w_c , was almost zero and was not affected by time.

When the moisture content of fine particles at the outlet is negligibly small, $w_2=0$, the following equations are obtained from mass balance of water and heat balance in the apparatus.

$$W_s w_1 = \bar{G}_{12} (H_2 - H_1) \quad (3)$$

$$W_s (C_{ps} + C_{pw} w_1) (T_{d2} - T_s) + W_s w_1 r_w = \bar{G}_{12} \bar{C}_{H12} (T_{d1} - T_{d2}) \quad (4)$$

where W_s is the feed rate of the fine particles, and w is the moisture content of fine particles. Subscripts 1 and 2 show inlet and outlet, respectively. From Eqs. (3) and (4), the absolute humidity at the outlet, H_2 , and the temperature of outlet gas, T_{d2} , respectively, can be calculated without knowledge of the rates of heat and mass transfer between the fluid and the particles. Therefore, the relative humidity defined by Eq. (1) can be calculated from Eq. (3), and the drying efficiency of Eq. (2) can be calculated from Eq. (4).

Figure 3 shows the axial distribution of the bed temperature when the temperature of the inlet gas was 403 K. As a reference, the distribution when the slurry had not been fed is also plotted in the figure. It is found that the bed temperature was uniform along the bed height and increased with gas velocity. The solid lines show the temperature of the exiting gas calculated from Eq. (4). The bed temperature as measured agreed with the calculated exiting gas temperature.

Figure 4 shows the effect of gas velocity on the temperature of the outlet gas, T_{d2} , at fine particles feed rates of 0.130 g/s and 0.212 g/s. T_{d2} increased with gas velocity. The solid lines show T_{d2} as calculated from Eq. (4). Calculated T_{d2} values agreed with those measured. This means that the moisture fed to the bed was completely evaporated in the bed and that the moisture content of the exiting fine particles was almost zero.

The dashed line shows wet-bulb temperature at the outlet, T_{w2} , obtained from the humidity chart by using T_{d1} , $H_1 = 0$. T_{d2} came close to T_{w2} with decreasing gas velocity. However, when the gas velocity was too small, such as $U < 0.3$ m/s at $W_s = 0.130$ g/s, the slurry was lumped with coarse particles and the lumps remained in the bed. Therefore, continuous operation could not be maintained at that gas velocity.

The experimental results shown in this paper were obtained in conditions under which such lumps were not accumulated.

Figure 5 shows the relationship between the relative humidity of the exiting gas and the gas velocity with the slurry feed rate as a parameter. The relative humidity decreased with increasing gas velocity and with decreasing feed rate of fine particles.

The solid lines show the relative humidity calculated by using Eqs. (1) and (3). The calculated values of the relative humidity agreed with the experimental ones. This means that $w_2 = 0$, assumed for Eq. (3), was also

Fig. 3. Axial distribution of the temperature in the fluidized bed

Fig. 4. Dependence of the outlet gas temperature, T_{d2} , on the gas velocity

Fig. 5. Relationship between the gas velocity and the relative humidity H_2^*

valid under this condition. The relative humidity was accurately calculated without the rates of heat and mass transfer.

Figure 6 shows the effect of superficial gas velocity on drying efficiency as defined by Eq. (2). Drying efficiency decreased with increasing gas velocity and with decreasing feed rate of fine particles. Under some conditions the drying efficiency was almost 0.9. This means that about 90% of heat input was used for the drying. An operation of such high drying efficiency is effective for saving energy. However, continuous operation at higher efficiency was not achieved, because the slurry was lumped with coarse particles and the lumps accumulated in the bed. The solid lines show the drying efficiency as calculated by Eqs. (2) and (4). Very good agreement between experimental and calculated values is seen. The drying efficiency was also evaluated without knowledge of the rates of heat and mass transfer.

Figure 7 shows the relationship between drying efficiency and gas velocity when the fine particles in the slurry were 1.8 μm and 2.6 μm in diameter. The

Fig. 6. Relationship between the gas velocity and the drying efficiency

drying efficiency obtained experimentally agreed with that calculated. The drying characteristics was not affected by the diameter of the fine particles.

Figures 8 and 9 show the relationship between feed rate of moisture, $W_s w_1$, and drying efficiency. The drying efficiency increased proportionally with the feed rate of moisture, and the efficiency obtained from the experiment agreed with that calculated. However, drying efficiency was hardly affected by feed rate of fine particles W_s , as shown in Fig. 9.

Figure 10 shows the relationship between temperature of the inlet gas and relative humidity. The relative humidity gradually decreased with temperature as expected from Eqs. (1) and (3).

It was also confirmed that drying efficiency was not affected by bed height or diameter in the range of this study. However, it was supposed that there is a minimum bed height for maintaining continuous operation.

Fig. 7. Effect of diameter of the fine particles on the drying efficiency

Fig. 8. Relationship between the feed rate of moisture, $W_s w_1$, and the drying efficiency

Fig. 9. Effect of feed rate of solid, W_s , on the drying efficiency

Fig. 10. Relationship between the inlet gas temperature, T_{d1} , and the relative humidity H_2^*

Fig. 11. Correlation between H_2^* calculated and that of observed

Figures 11 and 12 respectively show the correlation between experimental values and those calculated for H_2^* and η_d . Those calculated agreed with

Fig. 12. Correlation between the drying efficiency calculated and that of observed

those obtained experimentally. These figures show that the moisture content of the exiting fine particles was nearly zero and that the humidity and temperature of the exiting gas can be calculated from Eqs. (3) and (4) by using only the knowledge of the slurry and the gas at the inlet. This is a great merit in designing and operating the dryer.

The drying characteristics are thought to depend on the porosity of the fine particles. The alumina fine particles used in this study were not porous. This may be a major reason why the moisture content of the exiting fine particles was nearly zero regardless of operational conditions. The drying characteristics of porous fine particles should be examined.

As mentioned above, when the drying efficiency was too high, the slurry lumped with coarse particles in the bed and continuous operation could not be achieved. It is important for practical application to

make clear the range of operational conditions under which continuous operation can be maintained, and also to make clear the drying mechanism in the powder-particle fluidized bed to aid in the design and scale-up of the dryer.

Conclusion

Drying of a slurry of alumina fine particles with water by a powder-particle fluidized bed was carried out, and the drying characteristics were examined. The following conclusions were obtained.

1) Continuous drying of the fine particles slurry could be achieved by the powder-particle fluidized bed.

2) The moisture content of the outlet fine particles was nearly zero under these operational conditions.

3) Humidity and temperature of the outlet gas were accurately calculated by using Eqs. (3) and (4) respectively.

4) The relative humidity of the outlet gas and the drying efficiency defined by Eq. (2) were affected by the slurry feed rate, the gas velocity, the moisture content of inlet slurry and the temperature of inlet gas, but were not affected by the height of the fluidized bed or the diameter of the coarse particles.

Acknowledgement

The authors would like to thank S. Aoshika, an undergraduate student in our university, for his assistance in part of the experimental work of this study.

Nomenclature

C_H	humid heat	[kJ/kg-dry gas K]
C_{ps}	specific heat of fine particle	[kJ/kg·K]
C_{pw}	specific heat of water	[kJ/kg·K]
d_{pf}	diameter of coarse particle	[m]

d_{ps}	diameter of fine particle	[m]
G	total mass flow rate of gas	[kg/s]
H	humidity of gas	[kg-water/kg-dry gas]
H^*	relative humidity	[—]
L_f	static height of fluidized bed	[m]
r_w	latent heat of water	[kJ/kg]
T_b	temperature of fluidized bed	[K]
T_d	dry-bulb temperature of gas	[K]
T_s	temperature of inlet slurry	[K]
T_w	wet-bulb temperature	[K]
U	superficial gas velocity	[m/s]
U_{mf}	minimum fluidization velocity	[m/s]
U_t	terminal velocity	[m/s]
w_i	moisture content of fine particles	[kg-water/kg-dry solid]
W_s	feed rate of solid	[kg/s]
Z	height from the distributor	[m]
η_d	drying efficiency defined by Eq. (2)	[—]
<Subscript>		
1	inlet	
2	outlet	
c	cyclone	
w	saturated	

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