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Effect of Air Distributor on the Fluidization Characteristics in Conical Gas Fluidized Beds

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Abstract−The effect of an air distributor on the fluidization characteristics of 1 mm glass beads has been determined in a conical gas fluidized bed (0.1 m - inlet diameter and 0.6 m in height) with an apex angle of 20°. To determine the effect of distributor geometry, five different perforated distributors were employed (the opening fraction of 0.009-0.037, different hole size, and number). The differential bed pressure drop increases with increasing gas velocity, and it goes from zero to a maximum value with increasing or decreasing gas velocity. From the differential bed pressure drop profiles with the distributors having different opening fractions, demarcation velocities of the minimum and maximum velocities of the partial fluidization, full fluidization, partial defluidization and the full defluidization are determined. Also, bubble frequencies in the conical gas fluidized beds were measured by an optical probe. In the conical bed, the gas velocity at which the maximum bed pressure drop attained increases with increasing the opening fraction of distributors.

Key words: Conical Beds, Flow Regime, Minimum Fluidization Velocity, Optical Probe

INTRODUCTION

Conical beds can be operated successfully at fixed bed states of a gas-solid or liquid-solid system, but conventional fluidized beds often perform badly owing to solids properties. Conical fluidized beds provide low segregation of mixtures of solids with different particle size [Tanfara et al., 2002], such as wood chips [Olazar et al., 1995; San Jose et al., 1995; Inami et al., 2000]. This type of reactor can be used in catalytic polymerization of high density polyethylene (HDPE) or linear low density polyethylene (L-LDPE) where the catalyst is progressively crashed and coated with sticky polymers [Bilbao et al., 1987]. Conical fluidized beds have also been widely used in industrial processes such as gasification of bituminous coal, roasting sulfide ores, coating nuclear fuel particles, and granulation of pharmaceutical materials [Boldt, 1967; Tanfara et al., 2002; Tsuji et al., 1989; Uemaki and Tsuji et al., 1986]. The settling cone in the ore dressing process is another example of a conical liquid/solid fluidized bed [Kwauk, 1992].

Kwauk [1992] has reported that a decreasing fluid velocity gradient in the direction of fluid flow in a conical vessel has the following advantages: (1) For polydisperse solids, a higher velocity at the lower section of a cone provides adequate fluidization of the coarser particles, while a lower velocity at the top section prevents excessive carry-over of the fines. (2) The highly agitated coarse particles in the lower zone serve as a normal gas distributor to disperse the fluidizing medium to the upper zone of the finer solids in the conical bed. (3) For highly exothermic reactions, the preferential accumulation of coarse particles near the bottom of the cone with lower specific surface area ameliorates rapid heat release, and the high degree of turbulence due to the higher fluid velocity near the apex of the cone helps heat dissipation. Several researchers have studied the relationship between the pressure drop and gas velocity in conical fluidized beds [Nishi, 1979; Kmiec, 1983; Shi and Fan, 1984; Chen et al., 1997]. Toyohara and Kawamura [1989] have reported fluidization characteristics, such as the pressure drop and particle movement behavior at various apex angles and operating conditions. They have determined the flow regimes from fixed beds to core type partial fluidized beds, core type fluidized beds, and fully fluidized beds have been defined while decreasing gas velocity by visual observation in a two-dimensional column with tapering angles of 3.75, 7.5, 15, 30, and 40°. Peng and Fan [1997] also experimentally determined the maximum pressure drop, the minimum velocities of partial fluidization and full fluidization, and maximum velocities of partial defluidization and full defluidization in two-dimensional liquid-solid tapered beds with tapered angles of 5, 10, 15 and 20°. However, conventional fluidized beds in industrial processes are fabricated in non-visible three-dimensional steel columns. Therefore, in the present study, the flow regimes from fixed beds to fully fluidized beds were determined by both the optical probe method and the difference of bed pressure drop (ΔP) between the increasing and decreasing gas velocity in the beds. The effect of the distributor geometry (opening area fraction and orientation of orifice holes) on the transition velocities among the flow regimes has been determined.

EXPERIMENTAL

Experiments were carried out in a conical gas fluidized bed (0.1 m-i.d.×0.6 m-high) made of a transparent acryl column with an apex angle of 20° as shown schematically in Fig. 1. All experimental runs were performed at room temperature and atmospheric pressure. Air

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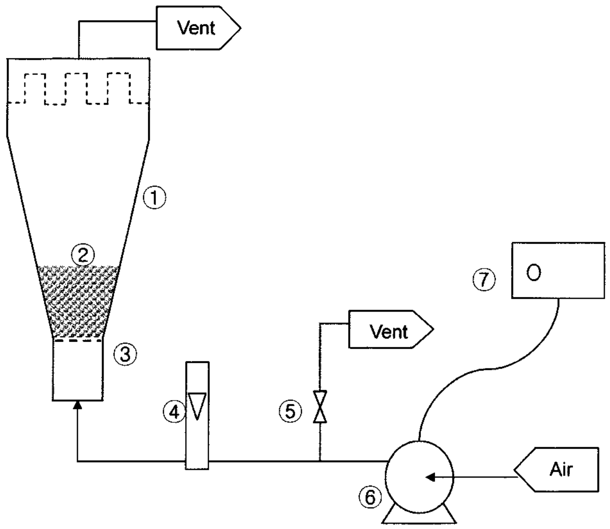


Fig. 1. Schematic diagram of the experimental apparatus.
1. Conical main column
2. Bed materials
3. Distributor
4. Flow meter
5. Bypass valve
6. Blower
7. Blower speed controller

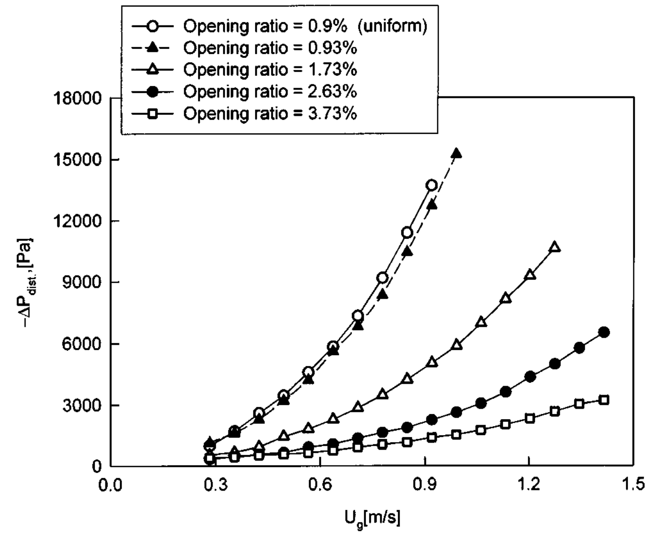


Fig. 2. Effect of gas velocity on pressure drop across the distributors.

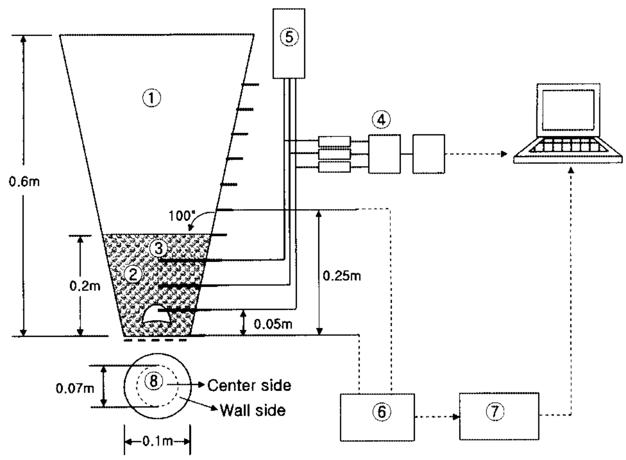


Fig. 3. Schematic diagram of the measuring system.
1. Conical main column
2. Bed materials
3. Optical probe
4. Detector and amplifier
5. Laser source
6. Differential pressure transducer
7. A/D converter
8. Distributor

flow rates were measured by a flow meter (FLT, Korea Flow Cell Co.) in the range of 0-1.4 m/s. Five different types of perforated plate gas distributors were employed. On the perforated plate, a 250-mesh screen was covered on the surface of a distributor to prevent particle weeping from the bed to the plenum chamber.

Details of five different distributors used in this study are shown in Table 1 and Fig. 2. The particles used in this study were 1.0 mm glass beads with a density of 2,500 kg/m^3. To measure the bed pressure drop, 11 pressure taps were mounted on the wall along the centerline of the bed. One pressure probe was connected to the tap immediately above the distributor; the other was located just above the surface of the bed as shown in Fig. 3. The opening of the pressure taps was covered with a screen (250-mesh) to prevent particle leaking from the bed. The pressure transducer was calibrated by a precision manometer. Transducer signals were monitored by a personal computer at a sampling frequency of 1 Hz for 180 s. An optical probe was used to measure bubble passing frequency in the bed.

A schematic diagram of the optical probe measuring system is shown in Fig. 3. The optical probes were installed along the centerline of the conical column. All the experiments were performed with decreasing gas velocity from the fully fluidized state, and then increasing the gas velocity from the fixed bed state. To perform an experimental run, air flow rate through the column decreased or increased incrementally. When a stable state was reached after each increment of gas flow, the pressure drop and the bubble passing signals in the bed were determined. The gas flow rate decreased down to the fixed bed state and then increased to the fully fluidized state.

Table 1. Distributor properties in this study

Distributor no.	Hole dimension/Opening ratio, [%]		Total opening ratio, [%]	Remarks
	Center side	Wall side		
1	φ 2 mm×23 ea		0.92	Uniform distribution
2	φ 3 mm×10 ea/0.8	φ 1 mm×13 ea/0.13	0.93	
3	φ 4 mm×10 ea/1.6	φ 1 mm×13 ea/0.13	1.73	
4	φ 5 mm×10 ea/2.5	φ 1 mm×13 ea/0.13	2.63	
5	φ 6 mm×10 ea/3.6	φ 1 mm×13 ea/0.13	3.73	

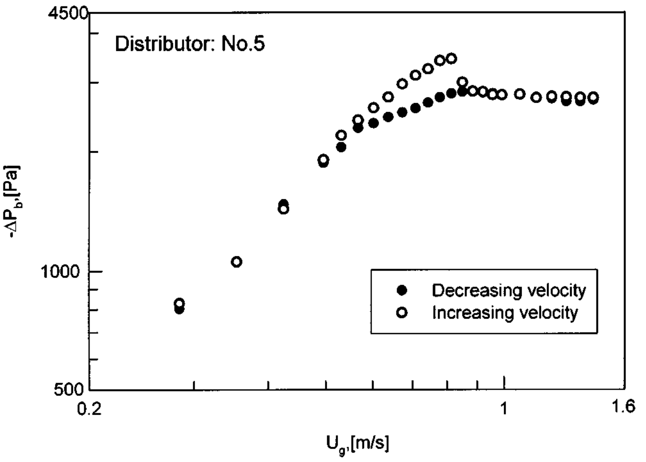


Fig. 4. Variation of the bed pressure drop as a function of gas velocity.

RESULTS AND DISCUSSION

The effect of gas velocity on bed pressure drop with the distributor #5 in the bed by decreasing gas velocity from the fully fluidized state and then increasing gas velocity from the fixed bed state is shown in Fig. 4. Generally, hysteresis in the pressure drop profile with variation of gas velocity can be observed in gas fluidized beds around the incipient fluidization condition [Davidson and Harrison, 1971; Kunii and Levenspiel, 1991; Peng and Fan, 1995]. Specially, the bed pressure drop peaks while gas velocity is increasing, but it does not happen when the gas velocity is decreasing around the minimum fluidizing state unless the density or size of the fluidizing solid particles is extremely low or small [Peng and Fan, 1995]. As can be seen, the bed pressure drop reaches a maximum value, then it decreases abruptly, and afterward it remains constant in the conical bed. This hysteresis of pressure drop is severer than that in conventional three dimensional gas fluidized beds. Peng and Fan [1995] have also reported that hysteresis of the pressure drop was more pronounced in the liquid-solid bed in a tapered column.

Fig. 5 shows the bubble passage with time along the bed height at decreasing gas velocity from the fully fluidized state (Fig. 5a) and that at increasing gas velocity from the fixed bed state (Fig. 5b) with distributor #5 in the conical bed. When the gas velocity decreases, the particles settle down to form a loosely packed fixed bed having voidage at minimum fluidizing condition, ε_{mf} [Kunii and Levenspiel, 1991]. As can be seen in Fig. 5(a), for the entire bed height a bubble was not detected at a gas velocity of 0.57 m/s, while bubble passage could be detected at a height of 0.05 m and U_g=0.64 m/s; however, bubbles were hardly detected at a height of 0.1 m. Therefore, we may claim that the fixed bed state at U_g=0.57 m/s, and the core type partially fluidized bed state at U_g=0.64 m/s. The fully fluidized bed state is reached at U_g=0.99 m/s, since bubbles can be detected along the entire bed height. Toyohara and Kawamura [1993] have studied the solids circulation characteristics in the side regions of tapered two-dimensional fluidized beds. They have measured bubble passages with variation of gas velocity by visual observation. The data obtained in this study is qualitatively in agreement with that of the previous work [Toyohara and Kawamura, 1993].

The bubble passage signals along the centerline of the conical column with the distributors #1 and #2 are shown in Fig. 6. The difference between distributors #1 and #2 is the different arrangement of the orifice holes with similar opening fraction (Table 1). All the experiments were performed by decreasing gas velocity from the fully fluidized state. The bubble passages could be detected by

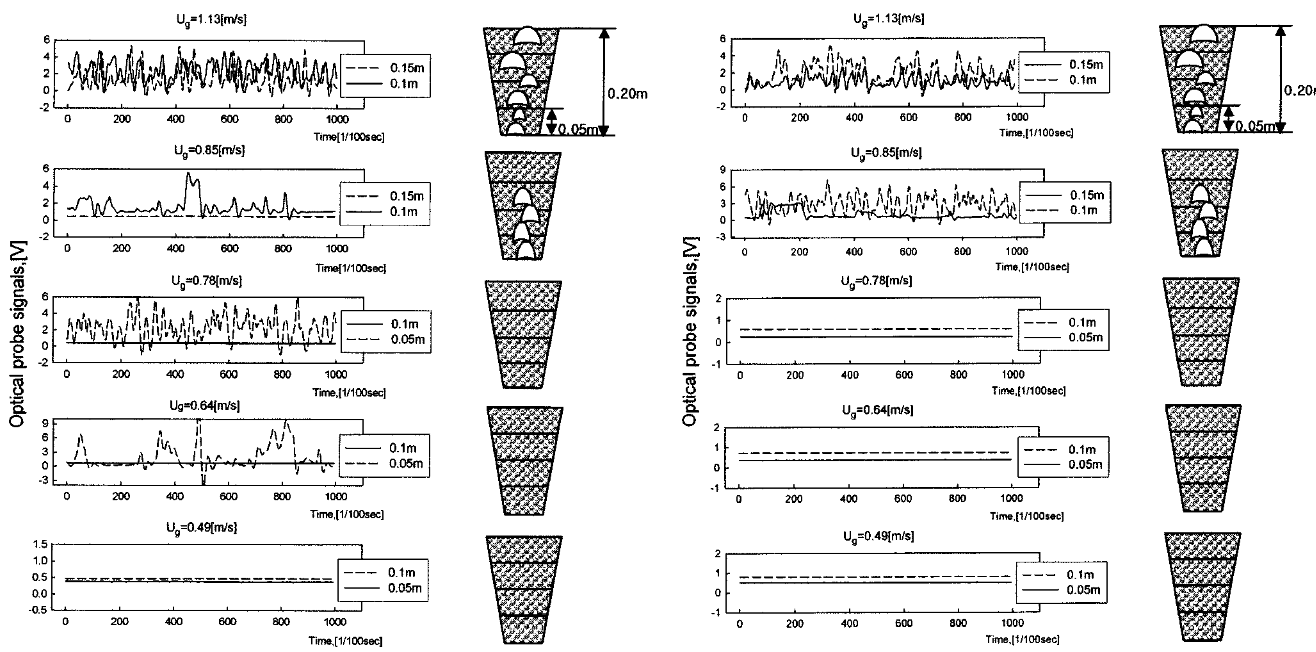
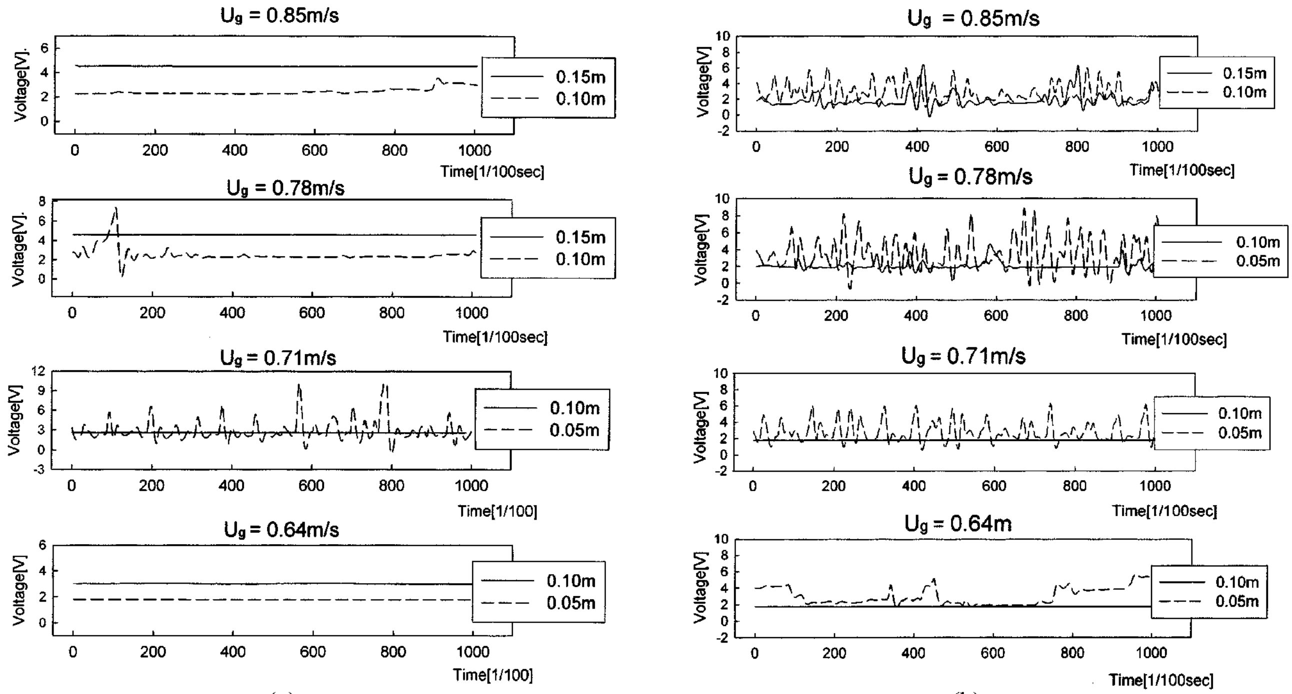


Fig. 5. Optical probe signals and the schematic diagram of bubble passage.
(a) Decreasing gas velocity condition, (b) Increasing gas velocity condition



(a) (b)

Fig. 6. Comparison of the optical probe signals between the distributors # 1 and # 2.
(a) Distributor #1, (b) Distributor #2

the optical probe along the bed height by increasing the gas velocity exceeding the core type partial fluidized condition. At the corresponding superficial gas velocity (i.e., U_g=0.85 m/s), the bubble passages of distributor #2 could be detected at a bed height of 0.15 m, but bubbles could not be detected with the distributor #1 at the same bed height. This may indicate that gas flow through the distributor #2 shifts toward the center of the distributor due to the presence of large holes at the center of the distributor.

Variation of the axial bubble frequency with gas velocity is shown in Fig. 7. As can be seen, the bubble frequency at H_{probe}=0.05 m is more frequent than that at H_{probe}=0.15 m due to bubble coalescence at the partial and full fluidized states. At higher gas velocities, bubble size increases with the excess gas velocity, U_g-U_{mf} [Kunii and Levenspiel, 1991] along the bed height. Therefore, bubble frequency decreases with increasing the bed height at the given gas velocity. As can be seen, the bubble frequency is zero at the condition that U_g is lower than 0.54 m/s and an optical probe height of 0.05 m. The flow regimes between the fixed beds and the core type fluidized beds can be divided by the bubble frequency at the max- imum velocity of full defluidization (U_{mfd}). When U_g is higher than 1.13 m/s, the bubble frequency remains a constant value ($N_B \cong 4.0$/s) with increasing gas velocity. Therefore, the flow regime between the core type fluidized beds and the full fluidized beds can be divided by the bubble frequency data at the minimum velocity of full fluidization (U_{mff}) as shown in the figure.

The effect of gas velocity on the difference of bed pressure drop

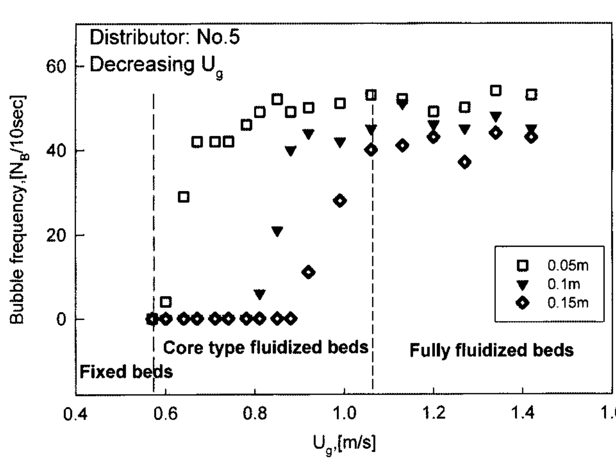


Fig. 7. Variation of the axial bubble frequency with gas velocity using the distributor #5.

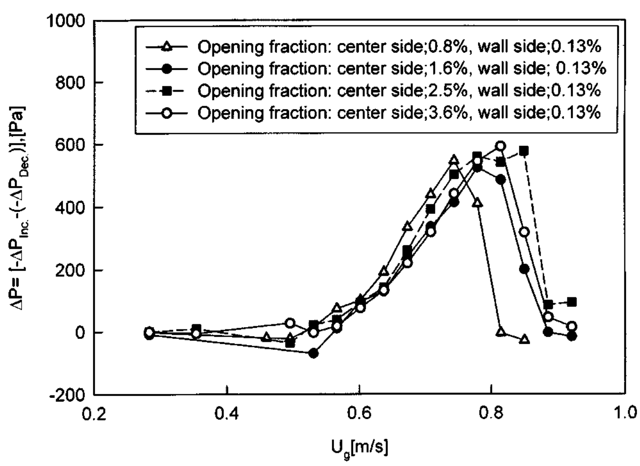


Fig. 8. Variation of the difference of bed pressure drop with increasing or decreasing gas velocity.

(ΔP) between the increasing and decreasing gas velocity as a function of opening fraction of the perforated distributor is shown in Fig. 8. As can be seen, ΔP increases as gas velocity (U_g) is increased from zero to a maximum value and then decreases as the gas velocity is further increased. The full defluidization occurs until the ΔP starts to increase, the minimum partial fluidization velocity (U_{mpf}) is defined at the maximum pressure drop point, and the minimum velocity of the full fluidization condition is defined as that the ΔP decreases to the zero point. It is also found that the ΔP becomes higher with the larger hole diameter at the inner side of the distributor because more gas can go through near the center portion of the distributor at a given gas flow rate. Peng and Fan [1997] applied the Ergun equation to the conical fluidized beds and derived an expression for the maximum bed pressure drop. From their expression for the conical beds, the gas velocity required to reaching the maximum ΔP increases with increasing the opening fraction of the distributor because ΔP also increases with gas velocity.

A relationship between ΔP and U_g in the conical bed of 1.0 mm glass beads having a static bed height of 0.2 m with distributor #5 is shown in Fig. 9. As gas velocity decreases from the fully fluidized state to the fixed bed, four flow regimes can be observed, namely fixed bed (I), core-type partial fluidized bed (II), core-type fluidized bed (III), and fully fluidized bed (IV). With increasing gas velocity, three flow regimes can be observed from the fixed bed (I'), core-type fluidized bed (II') to fully fluidized bed (III') state. Peng and Fan [1997] reported that if the liquid velocity was increased beyond the fully fluidized bed regime in the liquid-solid tapered fluidized beds, the transition flow regime from the fully fluidization to the turbulent fluidized bed regime could be observed. However, those flow regimes were not observed because we limited the gas velocity below 1.4 m/s.

The effect of the opening fraction of the distributors on the minimum velocities of the partial fluidization and the full fluidization, and the maximum velocities of the partial defluidization and the full defluidization is shown in Fig. 10. As the opening fraction is increased, the velocity difference between the maximum velocity of partial defluidization (U_{mfd}) and the minimum velocity of partial fluidization (U_{mff}) increases because the gas tends to flow towards the center of the column with larger opening fraction of the distributor.

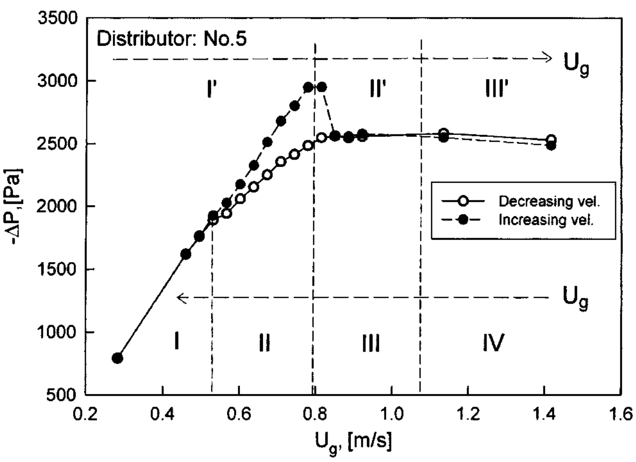


Fig. 9. Relationship between the bed pressure drop and gas velocity with the distributor #5.
Open symbol: increasing gas velocity condition; Closed symbol: decreasing gas velocity condition

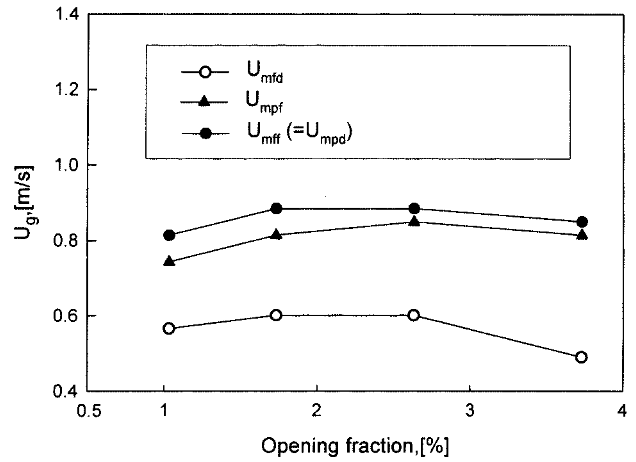


Fig. 10. Variation of the various velocities (U_{mfd}, U_{mpf}, U_{mff} and U_{mpd}) with gas velocity in the conical beds.

CONCLUSION

The effect of air distributor geometry (the opening fraction of 0.009-0.03, different hole size and number) on the fluidization characteristics of 1 mm glass beads has been determined in a conical gas fluidized bed. The flow regimes of the conical fluidized beds were determined for gas velocity increasing and decreasing conditions. The pressure drop increases with increasing gas velocity (U_g) from zero to a maximum value and then decreases as the gas velocity further increases. From the relationship between the differential bed pressure drop and gas velocity, the minimum velocities of the partial fluidization and the full fluidization, and maximum velocities of the partial defluidization and the full defluidization have been determined. In the conical beds, gas velocity that corresponds to the maximum bed pressure drop increases with increasing the opening fraction of the distributors.

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NOMENCLATURE

d_p : particle diameter [mm]
H_{probe} : height of the optical probe [m]
H_s : static bed height [m]
ΔP : difference of bed pressure drop when increasing or decreasing gas velocity [Pa]
ΔP_b : bed pressure drop [Pa]
$\Delta P_{Dec.}$: bed pressure drop when decreasing gas velocity [Pa]
$\Delta P_{dist.}$: pressure drop of the distributor [Pa]
$\Delta P_{Inc.}$: bed pressure drop when increasing gas velocity [Pa]
U_g : gas velocity through the bottom of the bed [m/s]

U_{mb} : maximum bubbling velocity [m/s]
U_{mf} : maximum fluidization velocity [m/s]
U_{mfd} : maximum velocity of full defluidization [m/s]
U_{mff} : minimum velocity of full fluidization [m/s]
U_{mpd} : maximum velocity of partial defluidization [m/s]
U_{mpf} : minimum velocity of partial fluidization [m/s]
ε_{mf} : voidage at the minimum fluidization velocity [-]

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