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On the influence of aspect ratio and distributor in gas fluidized beds

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Abstract

The influence of distributor on the uniformity of fluidization and the selection of the aspect ratio at a critical fluidization quality which corresponds to the maximum of the experimental to theoretical bed pressure drop ratio have been studied. Two types of multi-orifice distributors and three bed materials having their particle size in the range 70–161 μm have been investigated in an air fluidized bed at ambient conditions. The distributor to bed pressure drop ratio and the operating velocity for achieving uniform fluidization have been analyzed. The aspect ratio is shown to have a significant role in selecting a distributor and fixing the operating velocity particularly when the bed is shallow. A model equation derived from the first principles is proposed to predict the distributor to bed pressure drop ratio in terms of the critical aspect ratio and compared with the data reported in the literature.

Keywords: Aspect ratio; Fluidization; Distributor

1. Introduction

Gas fluidized bed fails to sustain good quality fluidization if the distributor malfunctions. A conventional gas fluidized bed is considered to have three zones, namely: (1) grid zone, (2) bubbling bed zone and (3) bubble erupting zone. Any change in the characteristics of the grid zone of a gas fluidized bed subsequently affects the behavior of the zones above it. A gas distributor affects very much the grid zone. The role and design principle of a variety of distributors can be well understood from a few reviews [1–8]. The reviews generally stress the need for a refined research approach to understand the basics of distributors.

The primary function of a distributor is to support the bed materials and to distribute the fluidizing gas uniformly. Any maldistribution of the fluidizing fluid, defluidization of the bed material on the grid plate, draining of material through the fluid injection sites and attrition of the bed material at the gas issuing sites are undesirable for an efficient distributor. The distributor in fact has an impact [9] in determining the visible bubble flow rate, interfacial area and hence the number of transfer units in a gas fluidized bed. As the grid zone is strongly influenced by the distributor, its design taking into consideration a proper aspect ratio for achieving uniform fluidization is important. There exists a certain criterion for fixing the fluid bed pressure drop for achieving uniform fluidization and this is generally related to the distributor to bed pressure drop ratio, $\Delta P_d/\Delta P_b$ in conjunction with the aspect ratio and operating velocity. To date there is no well defined work on this topic. We attempt in this paper to resolve this issue for the case of multi-orifice plate distributors and further we explore the influence of the aspect ratio, operating velocity and critical distributor to bed pressure drop ratio for achieving uniform fluidization.

1.1. Pressure drop criteria for uniform fluidization

The pressure drop across a distributor is conventionally expressed as its ratio to the bed pressure drop, $\Delta P_d/\Delta P_b$. As a general rule of thumb, this ratio has been chosen [10] at 0.1 for deep beds. This distributor drop ΔP_d is also suggested to be 10–12 in. water column in a shallow bed [10] or generally 100 times the free expansion value [11] for uniform fluidization. The $\Delta P_d/\Delta P_b$ ratio is said [12,13] to fall in the range 0.1–0.4 for uniform operation. The key problem is to select the aspect ratio corresponding to this pressure drop ratio. In a deep fluidized bed pressure drop is high and gas bypasses as large bubbles or slugs which affect in turn heat and mass transfer rates. Shallow fluidized beds have low bed pressure drop. They have low transport disengaging height and high solid expansion ratio. There is insufficient time for the bubbles to grow and form slugs. High rate of heat and mass transfer takes place near the distributor. Shallow beds are used in industries for drying, cooling, waste

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Nomenclature

A	area of cross-section of the column or bed (m^2)
Ar	Archimedes number, $d_p^3\rho_g(\rho_s-\rho_g)g/\mu_g^2$ (–)
C_d	discharge coefficient (–)
d_o	orifice diameter (m)
d_p	diameter of the sphere having the same surface/volume ratio as the particle and also known as surface volume particle diameter, d_{sv} (m)
D	diameter of the column (m)
D_B	bubble diameter (m)
D_{BM}	maximum bubble diameter (m)
D_{BS}	stable bubble diameter (m)
Fr_D	Froude number based on bed diameter, U_c^2/D_g (–)
g	gravitational constant (m/s^2)
h	height above the distributor inside the bed (m)
H	bed height (m)
H_{mf}	bed height at U_{mf} (m)
H/D	aspect ratio (–)
$(H/D)_c$	critical aspect ratio (i.e., H/D at Q_{max}) (–)
n	exponent in Richardson and Zaki, Eq. (1)
ΔP_b	pressure drop across the bed (Pa)
ΔP_d	pressure drop across the distributor (Pa)
ΔP_{ds}	maximum resistance of the stagnant zone during its destruction (Pa)
Q	fluidization quality ($\Delta P_{experimental}/\Delta P_{theoretical}$) (–)
Q_{max}	maximum fluidization quality (–)
R	distributor to bed pressure drop ratio ($\Delta P_d/\Delta P_b$) (–)
R_c	distributor to bed pressure drop ratio R at Q_{max} (–)
Re_{opt}	Reynolds number based on U_{opt} (–)
t	thickness of the distributor plate (m)
U	superficial gas velocity (m/s)
U_c	superficial gas velocity at Q_{max} (m/s)
U_M	superficial gas velocity at which all orifices in a multi-orifice distributor become fully operational (m/s)
U_{mf}	minimum fluidization velocity (m/s)
U_{opt}	optimum fluidizing gas velocity (m/s)
U_t	particle terminal velocity (m/s)
x_i	weight fraction of particle obtained by screen analysis in each size d_i (–)

Greek symbols

ε	bed voidage (–)
ε_{mf}	bed voidage at minimum fluidization velocity (–)
μ	viscosity of the fluidizing gas (Pa s)
ρ_g	density of gas (kg/m^3)
ρ_s	density of solid (kg/m^3)
ϕ	fraction free area of the distributor plate (–)

heat recovery, preoxidation and cooling of iron ore and combustion of powdered coal. Hence Kwauk [14] stressed a need for intensifying research on shallow beds.

1.2. Distributor and its role in estimating operating velocity and aspect ratio

The uniformity of fluidization has relevance to the characteristics of distributors and this has not been explained explicitly hitherto in the literature. Obviously the energy imparted to the fluidization should be dissipated efficiently to bring in a good gas–solid contact. A uniform fluidization in principle should be free from channeling. In such a situation the bed pressure drop should be equal or close to the theoretical value and this in fact can be well related to fluidization stability. As of now it appears that there is no universal definition available for the criterion of fluidization stability. Gupta and Sathiyamoorthy [15] have discussed this aspect in detail. We bring in the following some key issues on stability of fluidization and its relevance to distributor.

In order to achieve a channel free fluidization, Siegel [16] developed a simple criterion for a porous type of distributor assuming that the total pressure drop is constant at incipient fluidization. The equation proposed by him for $\Delta P_d/\Delta P_b$ is

$$\frac{\Delta P_d}{\Delta P_b} = \frac{1}{n}\left(\frac{U_{mf}}{U_t}\right)^{1/n}\frac{1}{1-\varepsilon_{mf}} \tag{1}$$

where the exponent n is obtained from the Richardson–Zaki equation [17]. The criterion of Siegel implies that the pressure drop ratio, $\Delta P_d/\Delta P_b$ is constant for a given gas–solid system. This is obvious from Eq. (1) that has constant value of the ratio U_{mf}/U_t for a given gas–solid system. Later, Shi and Fan [18] analyzed this criterion and proposed that the criterion of Siegel should be made valid even for $U > U_{mf}$. Their analysis however lacks in experimental validations. The behavior of gas fluidized bed was studied with respect to the operation of gas issuing sites in the case of multi-orifice distributors [19–21] and tuyeres [22]. The operating velocity when all orifices of a multi-orifice distributor become fully operational was expressed [21] as a function of the distributor to bed pressure drop ratio as given by

$$\frac{U_M}{U_{mf}} = 1 + \left[c\left(\frac{\Delta P_d}{\Delta P_b}\right)\right]^{1/c} \tag{2}$$

where the constant $c = 2$. In proposing this correlation the stability or uniformity of the fluidization was not considered. A later analysis [23] to incorporate the stability criterion (i.e. channel free fluidization) of Siegel [16] in Eq. (2) resulted in a new correlation for the operating velocity ratio

$$\frac{U_M}{U_{mf}} = 2.65 + 1.24\log_{10}\left(\frac{U_t}{U_{mf}}\right) \tag{3}$$

This correlation does not contain the parameter namely distributor to bed pressure drop ratio. However, it was suggested [23] that the constant 2.65 in Eq. (3) could be a function of the bed height. No attempt was made then to analyze how the aspect ratio influenced the uniformity of fluidization. Aspect ratio is conventionally defined as the ratio of the bed height, H to the diameter of the bed, D. The bed height for a fluidized bed is usually taken at the minimum fluidization velocity. If the aspect ratio is more than unity, the bed is usually considered to be a tall or deep bed. On the other hand, a shallow bed is the one that has aspect ratio equal to or less than unity. The exact aspect ratio that is a transition between a deep and shallow fluidized bed is not yet established in fluidization literature. Mori and Wen [24] proposed a correlation to predict the bubble diameter D_B as a function of the aspect ratio, h/D_t which is only a fraction of true aspect ratio. The bubble diameter D_B is assumed to be a certain fraction of fictitious maximum bubble diameter, D_{BM} which does not necessarily correspond to a stable bubble diameter D_{BS} [25]. Hence it is not proper to fix aspect ratio based on the knowledge of bubble diameter.

1.3. Critical velocity for uniform fluidization

Mori and Moriyama [26] attempted to relate the distributor to bed pressure drop ratio with the uniformity of fluidization and hence they linked it to the condition of no drift fluidization corresponding to last nozzle operation in a distributor. They assumed that the cross-sectional area of the fluidized bed section at the condition of no drift in fluidization is same as the total cross-sectional area of the bed and the flow through the stationary bed tends to be the same as minimum fluidization velocity. In other words a nonuniformly fluidized bed is viewed to have two parts namely a fixed bed or stationary section and a fluidized bed section. Experimental evidence to support this hypothesis was given for nozzle type distributor. But no attempt was made to estimate the fraction of fixed (stationary) bed in the fluidized bed. This adds to the problem in assessing the extent of uniformity in fluidization and its relevance to the aspect ratio. An important parameter that is still to be explored is the real minimum fluidization velocity. A real minimum fluidization velocity is proposed [27] as the one, which would give the same distributor pressure drop as of that obtained in an empty column. There has been no attempt to date to estimate this velocity. It appears that this velocity should correspond to the velocity at which all orifices in a distributor become fully operational. Hence all the gas issuing sites at this velocity should offer same resistance to the gas flow. The influence of aspect ratio on this velocity and the corresponding bed quality is yet to be explored. A complete fluidization is said to occur at a critical velocity when resistance caused by stagnant zone above the distributor is overcome. Baskakov et al. [28] proposed a correlation to estimate this critical velocity U_c:

$$\frac{U_c}{U_{mf}} = \left(1 + \frac{\Delta P_{ds}}{\Delta P_d}\right)^{0.5} \tag{4}$$

where U_c is supposed to initiate complete fluidization and ΔP_{ds} the maximum resistance of stagnant zone during its destruction. Since ΔP_d is a function of U_c there is a constraint to estimate U_c directly from Eq. (4).

It can now be realized from the foregoing that uniformity of fluidization is closely related to a critical velocity and the corresponding distributor pressure drop, which again is conventionally expressed as its ratio to bed pressure drop. Further, it can be understood implicitly the fact that the bed height or the aspect ratio has an important role to assess the uniformity of fluidization. Hence we analyze in the following how this aspect ratio can be interpreted from the knowledge of the distributor to bed pressure drop ratio $\Delta P_d/\Delta P_b$.

2. Prediction of critical distributor to bed pressure drop ratio $\Delta P_d/\Delta P_b$ and the corresponding operating velocity

Mazumdar and Ganguly [29] studied the influence of aspect ratio for water fluidized bed and found that the degree of bed expansion is independent of aspect ratio. Their studies however did not consider the issue on the uniformity of fluidization as influenced by the aspect ratio. Qureshi and Creasy [30] collected a few published data on the distributor to bed pressure drop ratio R $(=\Delta P_d/\Delta P_b)$ for stable or successful operation of prototype and pilot scale gas fluidized bed. They proposed a correlation for estimating the critical pressure drop ratio R_c:

$$R_c = \left(\frac{\Delta P_d}{\Delta P_b}\right)_c = 0.01 + 0.2\left[1 - \exp\left(\frac{0.5D}{H}\right)\right] \tag{5}$$

Geldart and Baeyens [31] later showed that R_c estimated by Eq. (5) did not agree well with the data of Geldart and Kelsey [32] particularly for low aspect ratios and hence proposed a new correlation for $H_{mf}/D < 0.5$:

$$\left(\frac{\Delta P_d}{\Delta P_b}\right)_c \geq \exp\left(-\frac{3.8H_{mf}}{D}\right) \quad \text{for } H_{mf}/D < 0.5 \tag{6}$$

Qureshi and Creasy [30] suggested an approximate empirical correlation to estimate the discharge coefficient C_d of a multi-orifice distributor as

$$C_d = 0.82\left(\frac{d_o}{t}\right)^{-0.13} \tag{7}$$

Hence the equation for pressure drop across a multi-orifice plate type distributor is

$$\Delta P_d = 1.49\left(\frac{d_o}{t}\right)^{1/4} \frac{\rho_g U^2}{2\phi^2} \tag{8}$$

The distributor to bed pressure drop ratio R can be obtained by dividing Eq. (8) by bed pressure drop, ΔP_b and it is equal to R_c:

$$R_c = \frac{1.49}{(\phi^2)(1-\varepsilon)}\left(\frac{d_o}{t}\right)^{1/4} \frac{\rho_g}{\rho_s - \rho_g} Fr_D \frac{1}{H/D} \tag{9}$$

where Froude number, $Fr_D = U_c^2/Dg$ at $U = U_c$ corresponding to maximum quality fluidization. According to Kassim [33] the distributor pressure drop value in the presence of a bed is only 70% of that obtained without the bed. With the porous plate the actual drop is 8% less than the empty bed drop [34]. Hence it is essential to determine experimentally actual ΔP_d in the presence of bed.

Literature to estimate U_c is scanty. We may assume that U_M in Eq. (3) is same as U_c because it has been developed based on the operating characteristics of a multi-orifice type distributor and also conform to the stability criterion of Siegel [16]. Further this velocity U_m has also been tested and proved [35] to be same as U_{opt} at which the wall to bed heat transfer coefficient is maximum. The appropriate correlations for predicting U_{opt} [35] in terms of Re_{opt} are

$$Re_{opt} = 0.008\,Ar^{0.868} \quad \text{for } 1 \le Ar \le 3000 \tag{10}$$

$$Re_{opt} = 0.13\,Ar^{0.52} \quad \text{for } 3000 \le Ar \le 10^7 \tag{11}$$

It is one of our objectives to obtain U_{opt} from Eqs. (10) and (11) and check experimental value of R_c with that predicted by Eq. (9) using U_{opt} in place of U_c.

We may mention here a note of the work by Medlin and Jackson [36] on the fluid mechanical description of fluidized bed to assess the effect of distributor thickness on convective instabilities. Their exclusive theoretical work is based on idealized perturbation analysis of clear fluid and it concerns mainly with porous type of distributor having distributor thickness to bed depth ratio of 0.1. There are several unexplored parameters such as bulk viscosity of particle phase in unperturbed state and gradient of the local mean effective pressure in particle phase with respect to an unperturbed bed voidage. Hence the results of such fluid mechanical theory work of porous plate distributor cannot be extended in our case.

3. Experimental

The experimental setup shown in Fig. 1 consists of a 0.2 m ID fluidized bed column of 1 m height. Two types of multi-orifice distributor having free open area, 0.273% and 0.52% named as types A and B, respectively, were used. The details of these distributors such as the number of orifices, orifice diameter, its spacing and the plate thickness are given in Table 1. The orifices were conical shaped with their top diameter equivalent to orifice spacing and the bottom diameter equal to the respective sizes as given in Table 1. This arrangement ensured the elimination of dead zone of solid particles between the orifices. The length of the orifice is just 0.8 mm only. The distributors were first tested in a very shallow bed and the gas flow distribution was visualized on

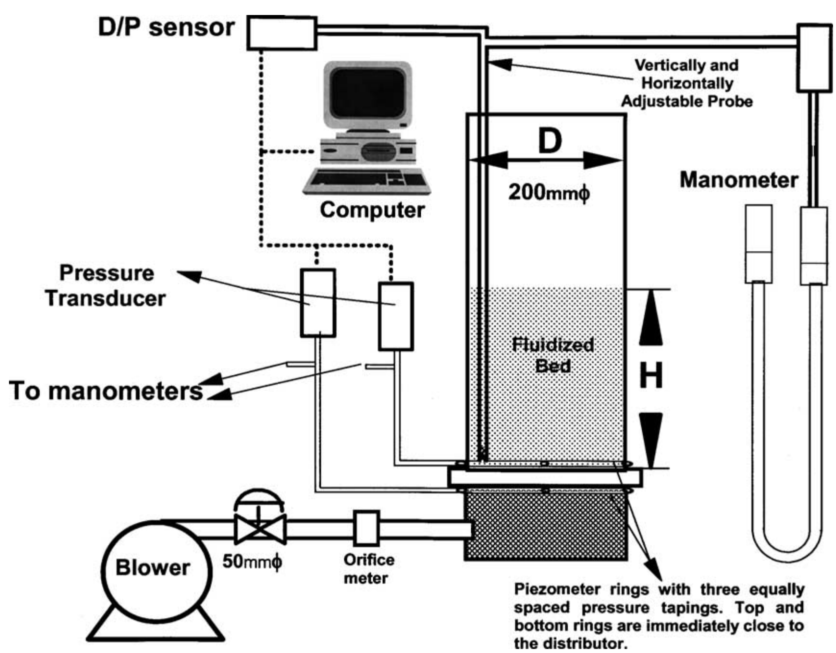


Fig. 1. Schematic representation of experimental setup.

Table 1
Details of multi-orifice distributors

Distributor type	Number of orifices, N	Orifice diameter, d_o (mm)	Free area, ϕ (%)	Orifice spacing in triangular array (mm)	Plate thickness, t (mm)	Plate thickness at the orifice, t_o (mm)
A	121	0.95	0.273	16.6	6.1	0.8
B	325	0.8	0.52	10	5.0	0.8

Table 2
Physical properties of particulate solids

Bed material	Particle size, $\sum(x_i/d_i)^{-1}$, d_p (μm)	Density (kg/m^3)	U_{mf} (m/s)	Voidage at U_{mf}	Sphericity, ϕ_s
Alumina	70	4000	0.0116	0.35	0.99
Rutile	150	4500	0.0477	0.381	0.99
Zircon	161	4550	0.0498	0.378	0.98

the bed surface. The flow distribution was uniform and the bed was fluidized well without any malfunctioning of the orifices in the distributor. The bed materials used were alumina, rutile and zircon and their particle sizes ranged from 70 to 161 μm. The physical properties of the bed materials are given in Table 2. The bed materials selected for this study have industrial importance and fluidization is applied in processing these materials. Air at ambient condition was used as the fluidizing gas. The relative humidity was below 10%. This ensured the absence of inter particle liquid bridge force and capillary forces. The bed materials belong to group B of Geldart classification of solids. Hence the cohesive forces do not influence them very much. The fluidizing gas was metered by rotameters. The fluidizing gas entered into a plenum and was distributed by the multi-orifice distributor. The plenum chamber was 0.25 m long and was packed with 10 mm size ceramic balls in order to distribute the gas evenly prior to its entry through the distributor. The pressure drop across the distributor was measured from the pressure taping of a piezometer ring (a ring type tube) provided just above and below the distributor. Each piezometer ring had three equally spaced pressure tapings. The aspect ratios were increased in small steps up to 2.5. When the column is charged with the materials close to the aspect ratio 2.5 the bed weight was high enough to initiate vibrations at operating velocities even at $3U_{mf}$. As reliable results for obtaining maximum fluidization quality is not possible at tall bed heights we have to restrict the aspect ratio for our studies close to 2.5 ($H = 0.5$ m). Furthermore, the aspect ratio 2.5 corresponds neither to a shallow nor a very deep bed. Even for this limiting aspect ratio the number of incremental steps to raise the bed height were many to determine the Q_{max}, and the corresponding aspect ratio, which we call in this paper as the critical aspect ratio. Pressure drops across the bed as well as the distributors were measured for each aspect ratio. The operating velocity used ranged from $1U_{mf}$ to $5.7U_{mf}$. The measurement of bed pressure drop for shallow bed (when the height is 150 mm and below) will be erroneous due to pressure fluctuations or oscillations caused by the violent agitation of the bed material. Hence care was taken to measure it accurately as per the guidelines suggested by Yang et al. [37]. In order to avoid pressure fluctuations in the measuring manometer limbs, the ends were enlarged and connected with thin bend tubes to achieve an overdamped condition for measuring the equilibrium pressure. A computer was used to record the pressure outputs from calibrated pressure transducers. A vertically as well as horizontally adjustable pressure probe was used to measure the bed pressure at different levels above the distributor. By extrapolation the pressure at the distributor level was determined. The bed pressure drop was then taken as the difference from this value and the surface pressure. The fluidization quality, Q was determined for each aspect ratio at different operating velocities and for all the three bed materials using both types of multi-orifice distributors. Each experiment was carried out in a well expanded and settled fluidized bed. The pressure drop across the distributor was measured in an empty column as well as in a fluidized bed. This was done to account for any pressure reduction that would prevail across the distributor when it supports a bed of fluidized solids above it.

4. Results and discussions

4.1. Distributor pressure drop in empty and filled column

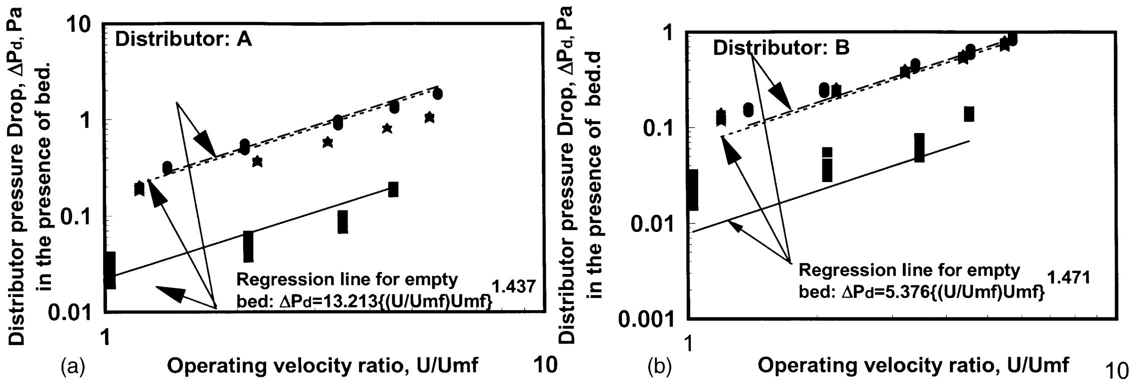
Regression analysis of the experimental data on ΔP_d in empty column for two types of multi-orifice distributors are

$$\Delta P_d = 13.213U^{1.437} \quad \text{for distributor type A (0.273\% free open area)} \tag{12}$$

$$\Delta P_d = 5.376U^{1.471} \quad \text{for distributor type B (0.52\% free open area)} \tag{13}$$

It can be seen from these equations that the behavior of a multi-orifice distributor tends towards that of a porous plate. Hence the conventional pressure drop equation for a single orifice is not advisable for estimating ΔP_d of a multi-orifice distributor plate.

We have already mentioned from literature [33,34] that the presence of bed material above the distributor reduces the pressure drop across the distributor. There has been no quantitative analysis for this conclusion and hence we have attempted here to examine this both for distributors A and B. Experimental ΔP_d values obtained across the distributors A and B in the presence of bed materials such as alumina, rutile and zircon are compared in Fig. 2 with that estimated by Eqs. (12) and (13). Regression equations for estimating ΔP_d are also compared in Fig. 2. It may be noted that the distributor free open area affects the ΔP_d when bed materials are present above the distributor. For example with distributor A (Fig. 2(a)), ΔP_d values estimated by regression equation (12) are greater than that obtained in the presence of bed and the difference is reduced as the bed material becomes coarser and denser. Observation with high free area



Details of legend and symbols

Bed Material	Symbol for data	Legend for Regression line for ΔP in empty bed	Regression equation for ΔP in filled bed Y(A)=aX^b Y(B)=pX^q (Data Points)
Alumina	■	———	a=0.0222,b=1.241,(40) p=0.0193,q=1.132,(72)
Rutile	★	......	a=0.156,b=1.1176,(50) p=0.1019,q=1.150,(95)
Zircon	●	— — —	a=0.2089,b=1.239,(50) p=0.1039,q=1.192,(95)

Fig. 2. Comparison of distributor pressure drops in the presence of bed and those in empty column. Data points shown are for all aspect ratios of this study.

distributor B is opposite to what was obtained with distributor A. These results show that it cannot be generalized to state that the distributor pressure drop is reduced if there is a bed of material above it. When the spacing between the orifices is large, there is a good chance for the formation of dead zone between the orifices. A gas jet when formed in an orifice will have the support of the dead zone walls that would give a pressure recovery effect (due to its diverging area). Hence the pressure recovery with distributor A is more easily possible than distributor B because distributor

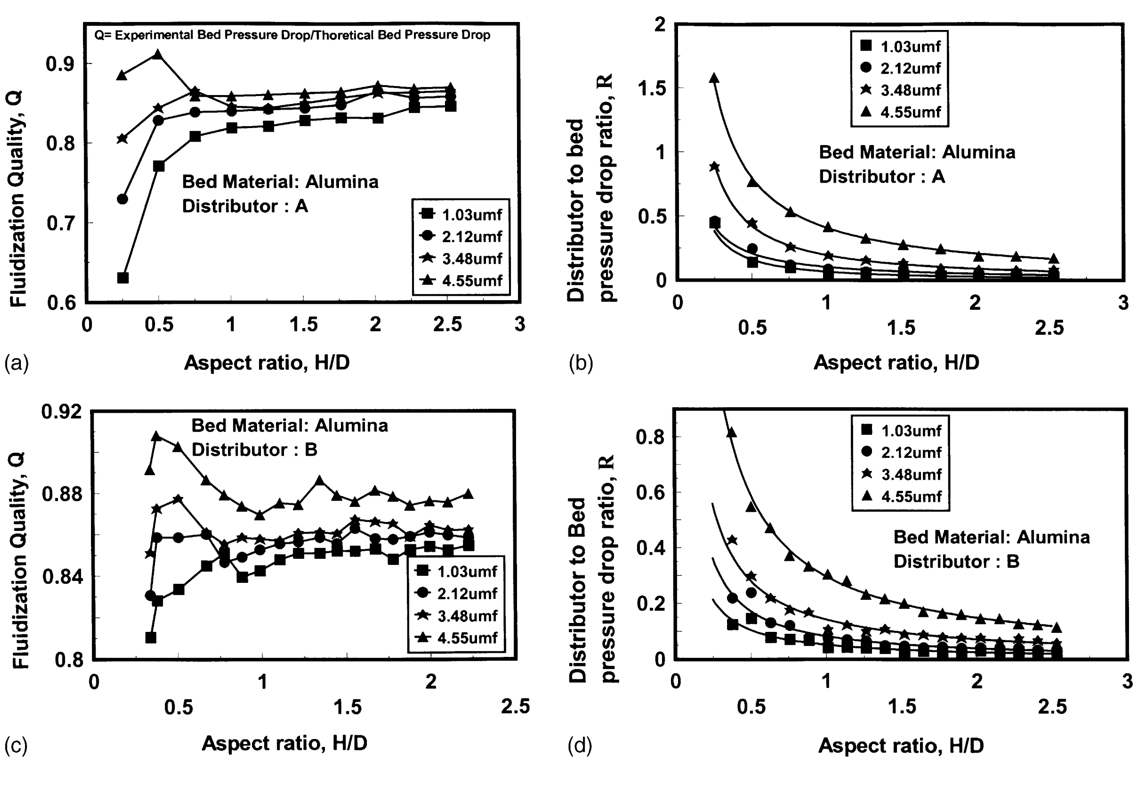


Fig. 3. Variation of fluidization quality, Q and distributor to bed pressure drop ratio, R with aspect ratio for alumina bed and multi-orifice distributors A and B at various velocity ratios, U/U_{mf}.

A has large spacing (16.6 mm) between the orifices as against a close spacing of 10 mm in distributor B. Distributor B has many orifices (321) and this could lead to the formation of more number of jets with relatively low flow of gas through each orifice (as compared with distributor A for a given flow). This situation will be conducive for jet to jet interaction at the distributor site itself leading to pressure loss. Furthermore, the absence of the stable dead zone (due to the close spacing of the orifices in distributor B) cannot normally give a clear conical shape for the jet to expand and to recover the pressure.

Tsukada and Horio [38] studied gas motion and bubble formation at the distributor of fluidized bed and showed that the height of the jet is a function of the particle diameter and not influenced by the bubble gas flow rate. For the same size nozzle the jet length increased with increase in minimum fluidization velocity. This shows that the jet height is longer for coarse and dense material due to their high U_{mf} values and hence there is a less chance for the pressure to recover just at the distributor level in the presence of these materials. As a result ΔP_d is almost same (as shown in Fig. 2(b)) both for empty and charged beds of zircon and rutile, which have relatively high minimum fluidization velocities.

4.2. The effect of distributor, bed material and aspect ratio on fluidization quality, Q

We have already defined the fluidization quality, Q as the ratio of the experimental bed pressure drop to the theoretical value. We prefer to use this term, fluidization quality as it conveys the meaning explicitly. Experimental results obtained on fluidization quality, Q and distributor to bed pressure drop ratio for distributors A and B are shown in Fig. 3(a)–(d) for the bed material alumina and for bed material rutile in Fig. 4(a)–(d). Results with similar trend were observed for the bed materials zircon. As the aspect ratio increases the quality of fluidization increases and then remains nearly constant in most cases. This increase in Q is up to a certain maximum Q_{max} followed by a fall. The Q_{max}, if obtained at high aspect ratio, is close to minimum fluidization condition. If Q_{max} occurs at low aspect ratio, then the corresponding operating velocity is far higher than U_{mf}. The quality of fluidization at Q_{max} that corresponds to a critical aspect ratio had a minimum of 0.85 near U_{mf} for the case of fine powder (i.e. alumina). For bed of rutile and zircon it varied closely from 0.95 to 0.98 but at $U/U_{mf} \gg 1$. The critical aspect ratio decreases with increasing operating velocity and it is also influenced by distributor and the bed material. This can be clearly seen from the plots shown in Figs. 5(a)–(d) and 6(a) and (b).

It may be inferred that a deep bed of alumina (fine solids) with low free area distributor maintains good quality fluidization. It also implies that a low free area distributor is a right choice to process a large amount of fines without loosing fluidization quality. Conversely, shallow beds tend to have good quality fluidization with distributor having a higher percent free area even at operating velocity close to U_{mf}. Bed materials like rutile and zircon particles are coarser than alumina. These two beds, as compared to alumina bed, do not have a steep fall in the aspect ratio with increase in U/U_{mf}. This indicates that a good fluidization is possible with these bed materials over wide range of velocity. The

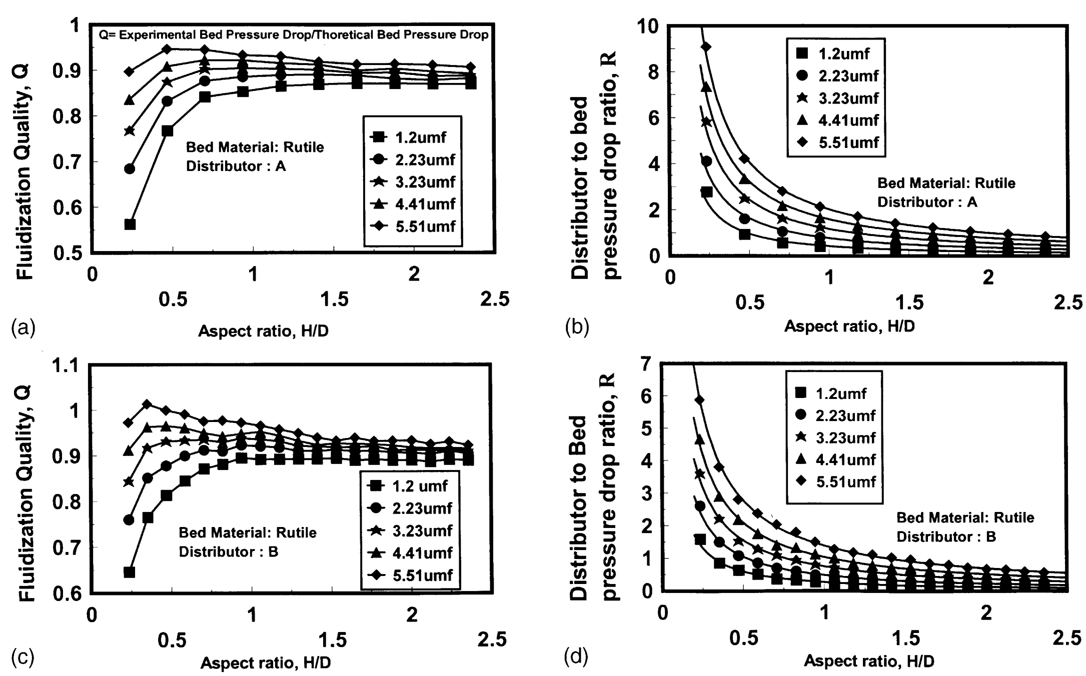


Fig. 4. Variation of fluidization quality, Q and distributor to bed pressure drop ratio, R with aspect ratio for rutile bed and multi-orifice distributors A and B at various velocity ratios, U/U_{mf}.

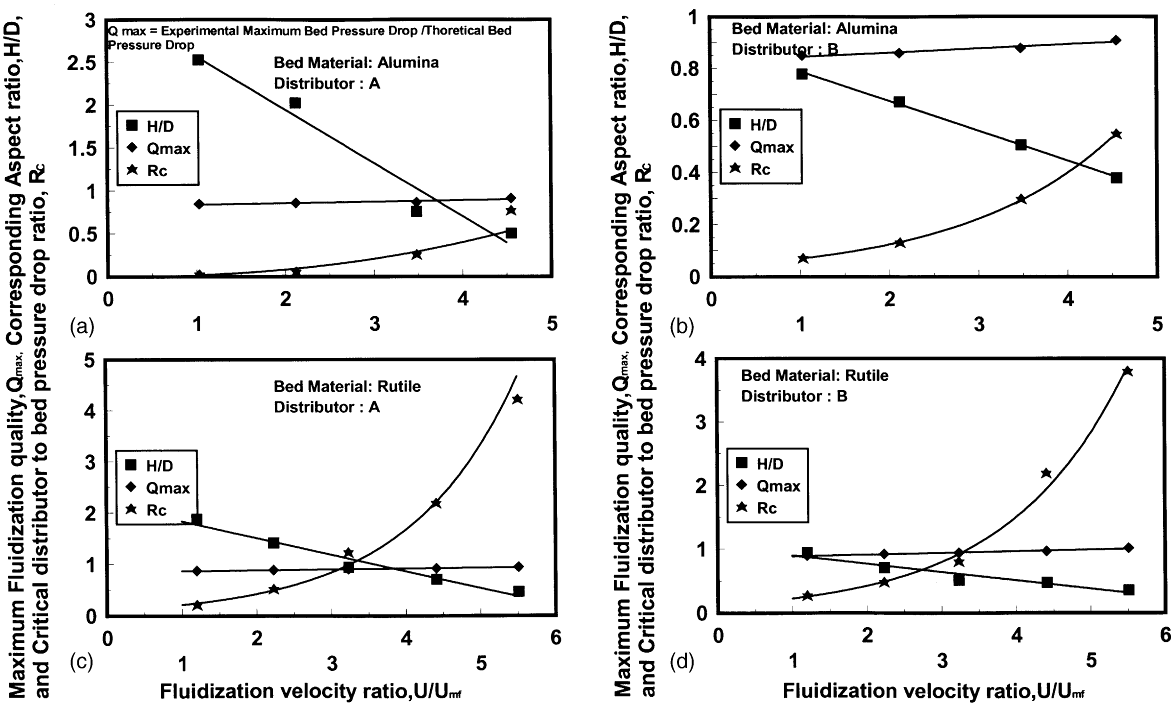


Fig. 5. Maximum fluidization quality, Q_{max}, corresponding aspect ratio, H/D and critical distributor to bed pressure drop ratio, R_c for alumina and rutile with distributors A and B.

observations on the critical aspect ratio show that it decreases linearly with the operating velocity ratio. Hence a high operating velocity is required for a shallow bed to maintain good quality fluidization. This is an important guideline for selecting the operating velocity and the appropriate distributor for shallow beds.

4.3. *The effect of aspect ratio on distributor to bed pressure drop ratio, R*

The distributor to bed pressure drop ratio, R as shown in Fig. 4(b) and (d), decreases with increase in aspect ratio and it increases with the operating velocity. With a low free open area distributor (i.e., A), the ratio R is always higher than that obtained by using high free open area distributor (i.e., B). These results are generally expected and can also be interpreted by Eq. (9). The R-values steeply increase with the decrease in aspect ratio particularly when H/D is below 1 and its value is even higher than unity. This is contrary to the distributor to bed pressure drop ratio of 0.1 recommended [10] in literature. The R-value actually becomes low at high aspect ratios. Thus it can be seen from our results that a rule of thumb value of 0.1 for R is not valid for aspect ratios less than 1. This is due to a different

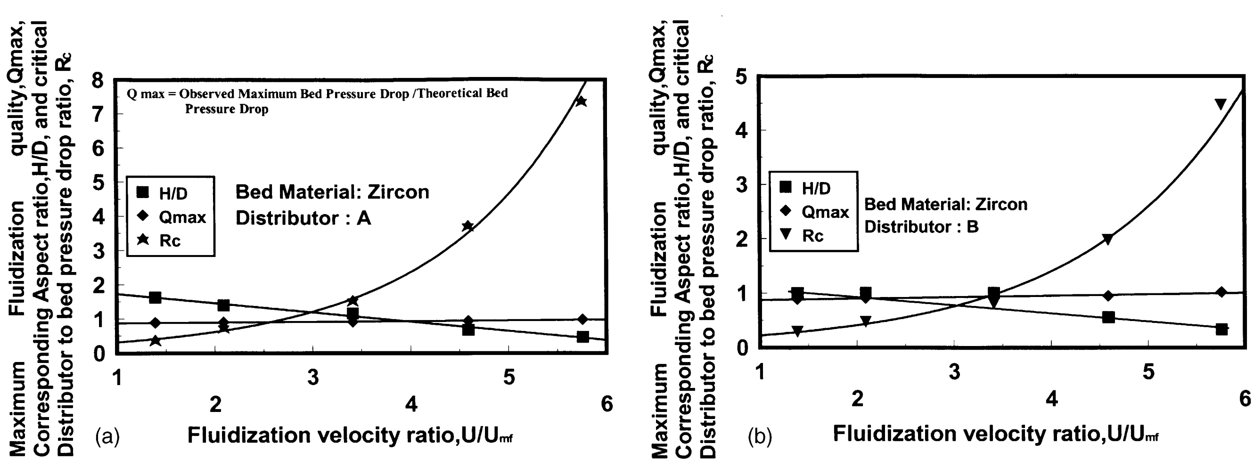


Fig. 6. Maximum fluidization quality, Q_{max}, corresponding aspect ratio, H/D and critical distributor to bed pressure drop ratio, R_c for zircon with distributors A and B.

hydrodynamics of very shallow beds as against the conventional bed.

4.4. Maximum fluidization quality, Q_{max} and corresponding R_c

The influence of the distributor on the stability of fluidization has not yet been well understood. The plots shown in Figs. 5 and 6 depict the maximum fluidization quality, Q_{max} that is found to exist for a given distributor at a given operating velocity and aspect ratio. We have chosen R corresponding to Q_{max} and named it as the critical ratio, R_c. It can be seen that R_c increases with increase in the operating velocity. Q_{max} decreases with operating velocity and there is a steep increase of R_c-values at high operating velocity ratios. An interesting point to note here is the magnitude of the R_c-values, which has a certain range in lower level, and this range falls from 0.022 (see Fig. 5(a)) to 0.037 (see Fig. 6(a)) in our system when the operating velocity at Q_{max} is close to minimum fluidization velocity. This is the range of value often cited in the literature and recommended generally when the aspect ratio is greater than unity. If the aspect ratio is less than unity the operating velocity ratio is high and it falls in our case in the range 3–5. The two factors namely high operating velocity and low aspect ratio, elevate the R_c-values which ranged from 0.55 (see Fig. 5(b)) to 7.36 (see Fig. 6(a)). This range of values has not usually been cited in the literature, the reason being the lack of analysis of the R_c-values for shallow beds considering the relevant factors, which influence the quality of fluidization. Now it can be learnt that if the R_c-value is to be lowered for low aspect ratio bed while the operating velocity is high, it is essential to have a distributor of high free open area.

4.5. Prediction of R_c in terms of aspect ratio or operating velocity U_c

Fig. 7 shows the plot of the R_c-values at Q_{max} and also the corresponding operating velocity ratio U/U_{mf} as a function

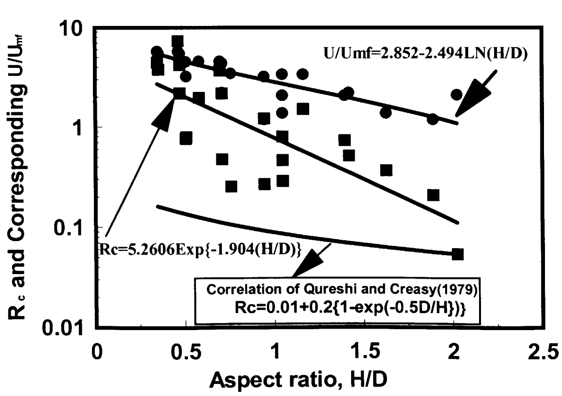


Fig. 7. Correlations for distributor to bed pressure drop ratio R_c (legend: dark square) and corresponding U/U_{mf} (legend: dark circle).

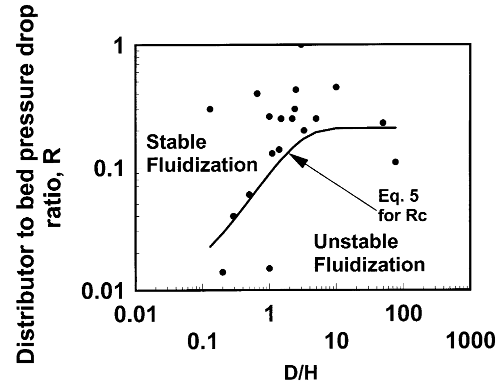


Fig. 8. A replot of distributor to bed pressure drop ratio R from the literature data [30].

of the critical aspect ratio. The regression line for predicting the ratio R_c corresponding to Q_{max} is

$$R_c = 5.2606 \exp\left(-\frac{1.904H}{D}\right) \tag{14}$$

The above regression equation is the best possible fit and has a poor correlation coefficient $r = 0.5734$. The literature for choosing the aspect ratio for stable operation is scanty. Qureshi and Creasy [30] proposed Eq. (5) only as a guideline for selecting R_c and there is no well-defined stability criterion in proposing this equation. Actually, Eq. (5) is only an approximate boundary line that separates the stable and unstable fluidization region and this can be seen in Fig. 8 which is plotted using the data of Qureshi and Creasy [30]. Only three data points fell in the unstable region. In obtaining some of these data these authors had to assume the missing values, as some relevant data were not available directly from the literature. It is not clear if the data on distributor pressure drop is for the empty column or fluidized bed. The important data of concern is the distributor pressure drop and the corresponding operating velocity. It is apparent that Eqs. (5) and (14) are not a function of the parameters that take into consideration of the distributor design and the operating gas velocity. The regression equation for estimating the velocity ratio, U/U_{mf} as a function of the critical aspect ratio $(H/D)_c$ is

$$\frac{U}{U_{mf}} = 2.852 - 2.494 \ln\left(\frac{H}{D}\right)_c \tag{15}$$

This regression equation has the best correlation coefficient 0.7097 among various curve fittings of the data. In principle U predicted by Eq. (15) is U_c and this can be used in Eq. (9). But it may not be appropriate to choose U from Eq. (15) as this has poor correlation coefficient.

4.6. Prediction of R_c by appropriate correlation

A comparison of the predictions of R_c by Eqs. (5), (6), (9) and (14) is shown in Fig. 9. It is obvious that no equation seems to give a reasonably good prediction. The predictions

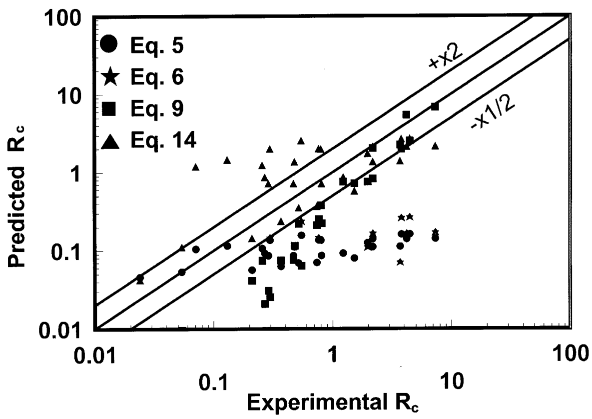


Fig. 9. Predicted R_c vs. experimental R_c of fluidized bed.

made by Eqs. (5) and (6) are far away from the actual. We have already pointed out the limitations and problems associated with Eq. (5). However, it was the first to give a relation of this kind for selecting an aspect ratio for stable fluidization. Eq. (6) was suggested for shallow beds but without any experimental verification. Prediction of R_c by Eq. (9) deviates much from the experimental values and this is due to the use of Eq. (8) in deriving Eq. (9). It may be recalled from our earlier discussions that Eq. (8) is not the appropriate choice for predicting pressure drop across a multi-orifice plate distributor. Hence the regression equations (12) and (13) in conjunction with Eq. (9) can be employed to estimate R_c to have meaningful results. But there are limitations in using these regression equations as they have been proposed from the data of distributor pressure drops obtained in an empty bed. In order to use Eq. (9) to predict the ratio R_c for a given aspect ratio, further data on distributor design parameters and the operating velocity corresponding to uniform fluidization are needed. It is seen in Eq. (9) that predictions made by Eq. (14) have some scattered points that correspond to tall and shallow beds.

Fig. 10 shows the results of the predictions by Eq. (9) in which operating velocity, U_{opt} as predicted by Eqs. (10) and (11) have been used. It can be seen that the predicted

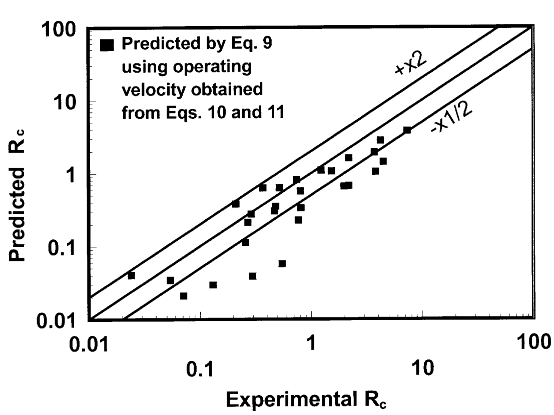


Fig. 10. Predicted R_c vs. experimental R_c of fluidized bed.

R_c-values are very close to the experimental values. There are four data points of R_c that fall less than unity and they are underestimated by Eq. (9). These data points were obtained for bed material alumina and with using distributor B. Actually for fine powder a low free open area distributor initiates a smooth and good quality fluidization. Hence distributor B that has relatively high free area is not a favorable choice for fine powder like alumina. This could be the reason for unrealistic prediction of R_c by Eq. (9) for the case of alumina particles.

Our studies are concerned with the aspect ratios keeping the bed diameter constant. We certainly anticipate some effects if bed diameter is a variable parameter. Generally large diameter industrial scale fluidized beds employ shallow beds. Hence it is very important to estimate the fluidization quality of very shallow beds precisely using a well-instrumented system. Such studies may have more academic values. But its significance in practice is yet to be assessed. Hence we recommend future studies to investigate bed stability and to evaluate the relevant R_c-value and optimum velocity for shallow beds by increasing the bed diameter. Obviously, this will be a challenge for future research. A better approach to resolve this issue may be a numerical simulation technique.

5. Conclusions

The distributor type, aspect ratio and operating velocity influence the quality of fluidization. There exists a critical aspect ratio where the fluidization quality is maximum and this critical aspect ratio is influenced by operating velocity as well as distributor type. The critical aspect ratio is found to fall linearly with the increasing operating velocity. The distributor to bed pressure drop ratio has been expressed as a function of critical aspect ratio and the results are compared with the literature data. A guideline to choose the appropriate value of this ratio to achieve a good quality fluidization has been presented. The pressure drop across a multi-orifice distributor provided with a number of orifices tends to approach the behavior of a porous plate and it is affected when bed materials are present above it. Pressure drop across the distributor during fluidization is not changed much from its empty bed value for coarse and dense materials at operating velocities much above the minimum fluidization velocity. Distributor influences a shallow bed remarkably. Experiments using large diameter columns and numerical simulation technique are suggested for future investigations. We also recommend studies at high temperatures as this has relevance for the pebble bed high temperature gas reactor.

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