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Prediction Of Minimum Fluidization Velocity And Bed Pressure Drop In Non-circular Gas-solid Fluidized Bed

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Abstract

The onset of fluidization velocity and corresponding pressure drop are. based on the balance between the forces which include the gravitational and fluid frictional forces exerted on the particles and the total effective weight of the particles. The equations for minimum fluidization velocity and corresponding pressure drop were derived for non-circular sections of square, hexagonal and semi-cylindrical shape and compared with the experimental ones. Under identical operating conditions, onset of fluidization velocity is minimum for a square bed while the semi - cylindrical one exhibits minimum bed pressure drop at the onset of fluidizaion.

Introduction

Cylindrical or columnar fluidized beds have been employed extensively in process industries. In spite of many advantages claimed by fluidization in columnar beds, the efficiency and quality in large scale gas-solid fluidized beds are seriously affected by bubbling, slugging and channeling behaviour at gas velocities higher than the minimum fluidization velocity. These result in poor gas-solid contact and lower mass and heat transfer rates. Various techniques viz. introduction of baffles, operation in a multi - stage unit, imparting vibrations and rotations have been advocated for improved performance of a gas-solid fluidized bed. Application of non-cylindrical conduits, instead of a conventional cylindrical one, is also considered to be an attractive alternative technique for smooth gas-solid fluidization (2, 3). In this paper, some dynamic studies have, therefore, been made in hexagonal, square and semi-cylindrical beds for their potential application in gas-solid fluidization.

The objectives of the present work include (i) Development of a mechanistic model for the condition of incipient fluidization in non-columnar beds based on Ergun's equation (1), (ii) Derivation of equations for predicting the minimum fluidization velocity, U_{mf} and minimum bed pressure drop, **$(-\Delta P_{mf})$**, for gas-solid fluidization (4,5) and (iii) Comparison of the developed equations with experimental data.

Theoretical Analysis

With the assumption that friction is negligible between particles and bed walls,fluidization is initiated in a fluid-solid bed when the force exerted by the fluidizing medium flowing through the bed is equal to the total effective weight of the material in the bed. For simplicity, we further assume that the lateral velocity of the fluid is sufficiently small to be neglected and that the vertical velocity of the fluid is uniformly distributed on the cross-sectional area (4).

The pressure drop through a packed bed with differential height dh is equal to **(1)**

$$dP = (Au + Bu^2)dh \qquad (1)$$

where,

$$A = \frac{150(1-\varepsilon_o)^2 \mu}{\varepsilon_o^3 \; (\phi_s \; d_p)^2}, \quad B = \frac{1.75(1-\varepsilon_o)\rho_f}{\varepsilon_o^3 \; \phi_s \; d_p}$$

The overall pressure drop across the entire bed height, H, is obtained by integrating equation (1), i. e.,

$$(-\Delta P) = \int_o^H Au+Bu^2)dh \qquad (2)$$

for bed whose cross-sectional area does not vary with height, equation (2) can be integrated to give

$$-\Delta P = (Au+Bu^2)H \qquad (3)$$

The overall force exerted by the fluidizing fluid on the particles is

$$F = A_n (- \Delta P) = A_n (Au + Bu^2) H \qquad (4)$$

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The overall effective weight of the particles in the bed is,

$$G=\int_0^H g(1-\varepsilon_o)(\rho_s-\rho_f)A_n dh=g(1-\varepsilon_o)(\rho_s-\rho_f)A_n H \quad (5)$$

The particles begin to fluidize when F = G. Substituting for u as U_{mf}, the minimum fluidization velocity and equating (4) and (5), the relation obtained is,

$$U_{mf} = -B' + (B'^2 + 4 A' C')^{\frac{1}{2}} / (2A') \quad (6)$$

where,

$A' = A_n BH$, $B' = A_n AH$, $C' = g(1-\varepsilon)(\rho_s - \rho_f) A_n H$

The pressure drop (ΔP_{mf}) for minimum fluidization is calculated as,

$$-\Delta P_{mf} = (Au_{mf} + Bu_{mf}^2) H \quad (7)$$

Values for minimum fluidization velocity and pressure drop are calculated using equations 6 and 7. For hexagonal, semi-cylindrical, square and cylindrical bed, cross-sectional area remains constant with the height of bed. The porosity of the bed was calculated from the measurement of height and area of bed, mass and density of the particles. The sphericity was calculated using packed bed data of cylindrical bed (7) or by the equation,

$$(1 - \varepsilon) / \emptyset_S = 0.231 \log d_p + 1.417 \quad (8)$$

Where, d_p is the particle diameter in feet (6).

Experimental

All the non-columnar vertical beds including the cylindrical ones were made of transparent acrylic resin so that the solids' behaviour could be observed clearly. For proper distribution of air, a calming section with glass beads was used at the entrance of column. Above the glass beads, a copper screen of 48 mesh was used.

The dimensions of the beds employed are listed in Table 1. Properties of the beds and the fluidized particles are listed in Tables 2 and 3 respectively. Ambient air (approximately 30°C) dehydrated through silica gel tower was used as the fluidizing medium. The flow rate was measured by a rotameter and the pressure drop by a manometer. The grid pressure drop has been found to be negligible.

Table 1 — Dimension of bed employed

Type of bed	Cross-sectional area, m^2	size/dia m
Cylindrical	81.07×10^{-4}	0.01016 (dia.)
Semi-cylindrical	88.5×10^{-4}	0.1501 (dia.)
Square	67.24×10^{-4}	0.082 (side)
Hexagonal	64.95×10^{-4}	0.050 (side)

Table 2 — Properties of beds

Material	Type of bed	voidage (ε_o)
Dolomite	cylindrical	0.515 to 0.550
Chromite ore	cylindrical	0.522
Coal	cylindrical	0.543
Sago	cylindrical	0.437
Manganese ore	cylindrical	0.568
Dolomite	Square	0.490 to 0.520
Chromite ore	Square	0.476
Coal	Square	0.516
Ramdana	Square	0.446
Manganese ore	Square	0.531
Dolomite	Semi-cylindrical	0.468 to 0.537
Chromite ore	Semi-cylindrical	0.546
Coal	Semi-cylindrical	0.563
Sago	Semi-cylindrical	0.467
Manganese ore	Semi-cylindrical	0.527
Urea	Semi-cylindrical	0.439 to 0.484
Dolamite	Hexagonal	0.502 to 0.547
Chromite ore	Hexagonal	0.498
Urea	Hexagonal	0.412 to 0.432
Coal	Hexagonal	0.526
Sago	Hexagonal	0.416
Manganese Ore	Hexagonal	0.501

Table 3 . Properties of Fluidized Particles

Material	$d_p \times 10^3$, m	ρ_s, kg/m^3	$\emptyset_S$
Dolomite	0.600 to 0.925	2740	0.555 to 0.660
Chromite ore	0.600	4050	0.654
Coal	0.600	1500	0.655
Manganese ore	0.600	4948	0.527
Ramdana	0.60 to 1.85	930	1.0
Sago	0.60 to 1.85	1420	1.0
Urea	0.60 to 1.85	1300	1.0

A known weight of the particles was poured into the bed and then the particles were fluidized fully by air at a certain velocity. The loading of particles for each experiment ranged from 0.5 to 1.5 kg. When a stable state of fluidization was established, the velocity was gradually reduced to zero. After the velocity reached zero, the velocity was gradually increased and the values of velocity and pressure drop were again recorded. Using the data while increasing the velocity, the pressure drop-velocity curve was constructed. The minimum fluid mass velocity (G_{mf}) and the minimum pressure drop, (ΔP_{mf}) for fluidization were determined from the curve.

Results And Discussion

The experimental results have shown that for the increasing fluid velocity cycle, the pressure drop - velocity relation for non-columnar beds yield curves (Figures 1 and 2) for which the pressure reaches a

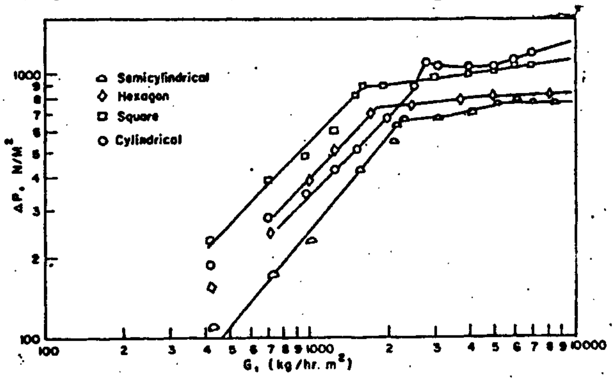

Figure 1 Pressure drop-mass velocity relations for circular and non-circular beds with same m/An for non-sphesical particles.

maximum value and then decreases tor a while after which it again increases slowly. In the decreasing fluid

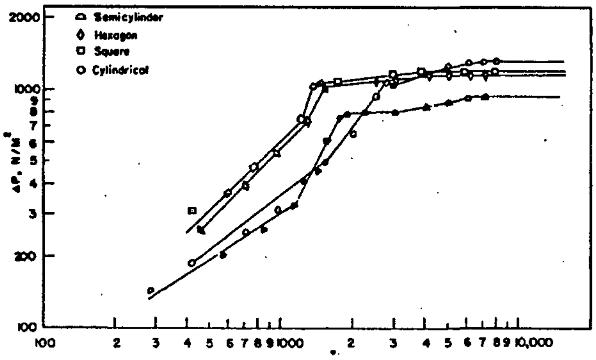

Figure 2Pressure drop-mass velocity relations for circular and non-circular beds with same m/An for spherical particles.

velocity cycle, however, the pressure drop decreases continuously and the peak is not recovered. This is similar to the pressure drop-velocity diagram of the columnar fluidized bed. The difference between the diagrams of columnar bed and that of non-columnar bed is that the peak region decreased sizably in case of non-columnar beds.

Figure 3 compares the minimum fluidization mass velocity calculated with the help of equation 6 with the experimental ones. To obtain the calculated G_{mf} values, static bed data were used in equation 6. In Fig. 4, values for the minimum fluidization pressure drop (calculated by Equation 7) have been compared with the experimental data. It is evident from these figures that the agreement between the calculated and experimental values is satisfactory for engineering calculations, indicating that the Ergun's method can be generalized to non-columnar beds as long as the cross-sectional area does not vary with height. Higher experimental values for minimum fluidization mass velocity and pressure drop in a number of cases may

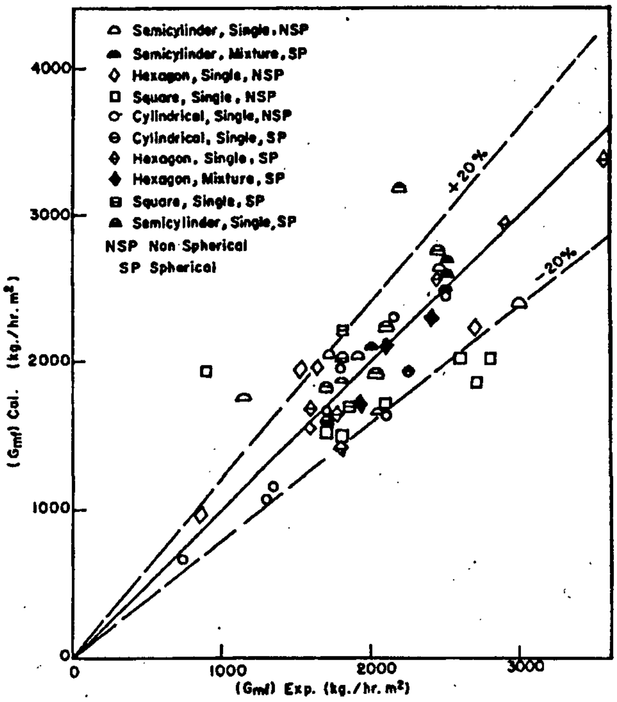

Figure 3 Comparison of predicted minimum fluidization velocities with experimental data.

be attributed mostly to porosity change at minimum fluidization. The columns were large enough to eliminate wall effects since the minimum ratio of column diameter to particle diameter was fifty five.

The values of minimum fluidization velocity and pressure drop for different conduits at equal amount of

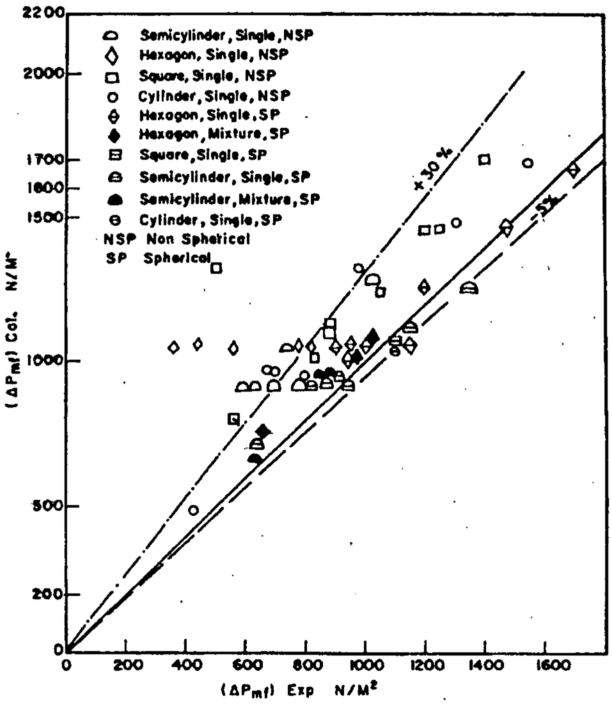

Figure 4 Comparison of predicted pressure drop at minimum fluidization velocities with experimental data.

solid particles of monosize and mixed size per unit cross-sectional area of bed are presented in Table 4 for comparison.

Conclusion

The equations for the prediction of minimum fluidization velocity and bed pressure drop, developed for the circular beds on the basis of Ergun's packed bed equation, can also be used to calculate the above mentioned quantities for non-circular beds of constant cross-section. The experimental values of minimum fluidization velocity for circular section when compared with those of Wen and Yu correlation, give an error of around thirty five percent.

For identical operating parameters, fluidization onset velocity in a square bed at a fluid mass velocity is less than that of circular bed while bed pressure drop for the same condition measures to be lower for semi-cylindrical bed than that of a circular bed (based on $\Delta P/H$). However, it is premature to draw more conclusions with respect to dynamic behaviour of non-circular beds at this stage and investigations are on. Other factors like structural capability, fluidization quality and finally, the acceptability by the users, who are often guided by conventionalism, are to be considered for the selection of the type of bed. The deviations are ± 20 percent for Figure 3 and + 30 to -5 percent in case of Figure 4. The error on one side only in Figure 4 is due to the difference in porosity calculation.

Notation

A = Ergun's constant

A_n = cross-sectional area of non-columnar bed, m^2

B = Ergun's constant

dh = differential bed height, m

d_p = particle diameter or mean particle diameter, m

dP = pressure drop through bed height dh, N/m^2

F = force exerted by the fluidizing medium on the particles in the bed, N

G = effective weight of the material in the bed, N

g = gravitational accelaration, m/s^2

G_{mf} = superficial fluid mass velocity at minimum fluidization ($kg / hr. m^2$)

H = height of the bed, m

m = amount of particles, kg.

ΔP = pressure drop through particle bed, N/m^2

ΔP_{mf} = pressure drop at minimum fluidization velocity, N/m^2

u = superficial fluid velocity, m/s

u_{mf} = superficial minimum fluidization velocity, m/s

Greek Letters

ε_o = voidage at packed condition

μ = viscosity of gas phase, kg (m. s)

ρ_f = density of fluid phase, kg/m^3

ρ_s = density of particle, kg/m^3

ϕ_s = sphericity of particles.

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Table 4 . Onset of Fluidization Conditions for Various Conduits

m / A_n	Type of system	Hexagonal			Square			Semicylinder			Cylinder		
		G_{mf}	ΔP_{mf}	ΔP_{mf}/ H	G_{mf}	ΔP_{mf}	ΔP_{mf}/ H	G_{mf}	ΔP_{mf}	ΔP_{mf}/ H	G_{mf}	ΔP_{mf}	ΔP_{mf}/ H
10.4	Single	2800	830	9880	1600	800	10389	1900	680	8831	1900	800	10666
10.8 (SP)	Single	2000	1150	8846	1700	1100	8029	1900	800	6504	2300	1100	8148
13.0	Single	2400	1100	10476	1850	1000	10752	2000	1040	10097	2150	980	10316
10.8	Single	2050	800	9523	1650	960	11428	1800	840	9032	2300	1100	12643
14.8	Single	2100	1200	10256	1850	1200	10619	2000	1220	10166	2250	1250	10965
8.3	Single	1900	540	8709	1700	560	9330	1900	580	8529	—	—	—
16.9	Single	2300	1400	11200	—	—	—	2000	1450	10902	2250	1550	11538
10.5	Mixture	1900	780	9750	1350	780	9397	1700	700	9210	2200	780	10400
13.0	Mixture	1800	960	9600	1650	960	9600	1700	950	9047	1900	950	11176
15.0	Mixture	1800	1200	10434	1300	1140	9913	1700	1130	9741	2250	1240	12040
17.4	Mixture	1900	1400	10769	1400	1350	10000	1600	1340	10551	2250	1480	12333

SP - Spherical Particles, others are non-spherical particles.

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