Indian Journal of Chemical Technology
Vol. 15, January 2008, pp.85-89

Prediction of minimum slugging velocity, bubbling bed index and range of bubbling fluidization in cylindrical and non-cylindrical gas-solid fluidized beds

R K Singh* & G K Roy

Department of Chemical Engineering, National Institute of Technology, Rourkela 769 008, India

Email: rksingh@nitrkl.ac.in

Received 13 March 2006; revised received 23 October 2007; accepted 30 October 2007

A bubbling fluidization regime exists between the minimum bubbling velocity and the minimum slugging velocity. Experimental investigations have been carried out for the determination of the minimum slugging velocity and the range of bubbling fluidization for non-spherical particles in cylindrical and non-cylindrical beds. In the present paper, correlations have been developed for the prediction of minimum slugging velocity for gas-solid fluidization in cylindrical and non-cylindrical (*viz*. semi-cylindrical, hexagonal and square) beds for non-spherical particles fluidized by air at ambient conditions. The bed materials used have varied diameter (324 to 925 micron) and density (1500 to 4800 kg/m^3). Experimental values of the minimum slugging velocity have been compared with the values calculated by the developed equations and also with the values calculated by Geldart's equation. A fairly good agreement has been obtained between the calculated and the experimental values. Based on the experimental data, it is concluded that, under similar operating conditions minimum slugging velocity and minimum bubbling velocity are maximum in case of a semi-cylindrical conduit and minimum in case of a square one. It is further observed that the range of bubbling fluidization is maximum in case of a semi-cylindrical bed for the identical operating conditions.

Keywords: Minimum slugging velocity, Bubbling bed index, Range of bubbling fluidization, Non-cylindrical beds

With increase in fluid velocity beyond the minimum fluidization condition, gas-fluidized systems exhibit either only a bubbling fluidization or a non-bubbling fluidization. Non-bubbling fluidiza-tion is also known as particulate or homogeneous fluidization and bubbling fluidization is often referred to as aggregative or heterogeneous fluidization. Geldart's classification of powders[1] gives the relationship between the difference of the solid and the gas densities and the particle diameter. Bed behaviour is obtained according to a practical classification, suggested by Abrahamsen and Geldart[2]. When the superficial gas velocity is higher than that necessary to obtain the incipient fluidization, the bed behaviour is classified under any of the following five categories *viz.* particulate, bubbling, slugging, turbulent and fast fluidization. Fluidized beds of fine powders exhibit some particulate behaviour above the minimum fluidization velocity. For these systems, it is possible to obtain homogeneous expansion without bubbles. The upper limit of the velocity for particulate state corresponds to the appearance of the first bubble and is called the minimum bubbling velocity. The gas velocity at which the bubble size equals to the diameter of the bed, the bed material is lifted by the fluid in form of short discrete cylinders and the phenomenon is termed as slug formation. The superficial gas velocity, at which the slug formation starts, is known as the minimum slugging velocity. The span ($U_{\mathrm{ms}} - U_{\mathrm{mb}}$) of effective gas solid fluidization is the bubbling regime. The bubbling fluidization exists between the minimum bubbling velocity and the minimum slugging velocity. When the bubbles formed at the distributor rise, they grow in size. At higher gas flow rate, the size of the bubble is large enough to cover the entire bed diameter and the slug is thus formed. Stewart and Davidson[3] have given a correlation for the minimum slugging velocity as:

$$U_{\mathrm{ms}} = U_{\mathrm{mf}} + 0.07(\mathrm{g}D_{\mathrm{c}})^{0.5} \quad \ldots(1)$$

The bed must sufficiently be deep for coalescing bubbles to attain the size of a slug.

Baeyens and Geldart[4] felt that the above condition is applicable only if $H_{\mathrm{mf}} > 1.3\ D_{\mathrm{c}}^{0.175}$ in SI unit, otherwise the minimum slugging velocity is expressed as:

$$U_{\mathrm{ms}} = U_{\mathrm{mf}} + 0.07(\mathrm{g}\ D_{\mathrm{c}})^{0.5} + 0.16(1.3\ D_{\mathrm{c}}^{0.175} - H_{\mathrm{mf}})^2 \quad \ldots(2)$$

Queiroz *et al.*[5] have verified the possibility of fluidization in the slugging regime by the dynamics of gas bubble growth in the bed. Singh and Roy[6] have presented correlations for the prediction of the minimum bubbling velocity, the fluidization index and the range of particulate fluidization in cylindrical and non-cylindrical beds.

In view of limited information available on the quantification of bubbling fluidization, in the present work, equations have been developed for the prediction of minimum slugging velocity for non-spherical bed materials in cylindrical and non-cylindrical fluidizers. Also a bubbling bed index and the range of bubbling fluidization have been calculated for a better understanding of the flow dynamics for the said regime in case of gas-solid fluidization.

Experimental Procedure

The experimental set-up has been given elsewhere[7]. All the cylindrical and non-cylindrical fluidizers were made of transparent acrylic resin so that the bed behaviour could be observed clearly. For uniform distribution of fluidizing medium in the bed, a calming section with glass beads was used at the entrance of the column. The dimensions of the beds used and the properties of the bed materials are given in Tables 1 and 2 respectively[8].

A known amount of the bed material was charged to the column from the top. The reproducible static bed was obtained after fluidizing the bed gradually and allowing it to settle slowly.

The compressed dry air was admitted to the column from the constant pressure tank. The bed pressure drop and the bed heights were recorded against the gradual change of flow till the fluidization condition was reached. The minimum fluidization velocity was obtained from the plot of pressure drop versus air mass velocity. The air flow rate was increased slowly after the minimum fluidization condition and the point at which the first bubble appeared was noted as the minimum bubbling velocity. The air flow rate was further increased and the point at which the slug formation started was noted as the minimum slugging velocity.

Development of correlations

The correlations for minimum slugging velocity have been developed with the help of relevant dimensionless groups involving interacting parameters like, the particle diameter, the equivalent diameter of the column, the packed bed height, the density of the particles and of the fluidizing medium.

For dimensional analysis, the minimum slugging velocity, U_{ms}, can be related to the system parameters as follows:

Table 1—Dimension of the beds employed

Type of bed	Cross-sectional area, m^2	Size/diameter, m
Cylindrical	81.07×10^{-4}	0.1016 (diameter)
Semi-cylindrical	88.50×10^{-4}	0.1501 (diameter)
Square	67.24×10^{-4}	0.082 (side)
Hexagonal	64.95×10^{-4}	0.050 (side)

Table 2—Properties of bed materials

Material	Density, ρ_p, kg/m^3	d_p*10^4,m	Minimum fluidization velocity, U_{mf}, m/s	Minimum bubbling velocity, U_{mb}, m/s
A. Cylindrical bed				
Dolomite	2740	9.0	0.71	0.72
Dolomite	2740	7.8	0.64	0.66
Dolomite	2740	6.0	0.49	0.52
Dolomite	2740	4.26	0.28	0.31
Dolomite	2740	3.24	0.15	0.21
Manganese ore	4800	6.0	0.60	0.64
Chromite ore	4050	6.0	0.51	0.53
Coal	1500	6.0	0.31	0.32
B. Semi-cylindrical bed				
Dolomite	2740	9.0	0.69	0.82
Dolomite	2740	7.8	0.54	0.56
Dolomite	2740	6.0	0.43	0.53
Dolomite	2740	4.26	0.37	0.47
Dolomite	2740	3.24	0.15	0.27
Manganese ore	4800	6.0	0.56	0.67
Chromite ore	4050	6.0	0.49	0.60
Coal	1500	6.0	0.23	0.28
C. Square bed				
Dolomite	2740	9.0	0.60	0.61
Dolomite	2740	7.8	0.52	0.53
Dolomite	2740	6.0	0.39	0.40
Dolomite	2740	4.26	0.34	0.37
Dolomite	2740	3.24	0.15	0.22
Manganese ore	4800	6.0	0.54	0.55
Chromite ore	4050	6.0	0.47	0.48
Coal	1500	6.0	0.17	0.18
D. Hexagonal bed				
Dolomite	2740	9.0	0.64	0.76
Dolomite	2740	7.8	0.52	0.62
Dolomite	2740	6.0	0.44	0.54
Dolomite	2740	4.26	0.37	0.44
Dolomite	2740	3.24	0.15	0.26
Manganese ore	4800	6.0	0.59	0.69
Chromite ore	4050	6.0	0.51	0.62
Coal	1500	6.0	0.21	0.25

$$U_{ms} = f\,[d_p/D_c,\ D_c/h_s,\ \rho_p/\rho_f\,] \qquad \ldots(3)$$

Eq. (3) can be rewritten as

$$U_{ms} = k[(d_p/D_c)^a(D_c/h_s)^b(\rho_p/\rho_f)^c]^n \qquad \ldots(4)$$

where k is the coefficient and a, b, c and n are the exponents.

The effect of the individual group on the minimum slugging velocity has been separately evaluated for different conduits by plotting of U_{ms} against the individual group on logarithmic plots and the values of exponents a, b and c have been obtained from the slope of these plots. The values of k and n have been obtained by plotting log U_{ms} against log $[(d_p/D_c)^a(D_c/h_s)^b(\rho_p/\rho_f)^c]$ for different conduits presented in Table 3.

On putting the values of a, b, c, k and n in Eq. (4), the correlations obtained for different conduits are as follows:

Cylindrical bed;

$$U_{ms} = 0.136(d_p/d_c)^{0.6324}(D_c/h_s)^{0.044}(\rho_p/\rho_f)^{0.6559} \qquad \ldots(5)$$

Semi-cylindrical bed;

$$U_{ms} = 0.269(d_p/D_c)^{1.0245}(D_c/h_s)^{0.2013}(\rho_p/\rho_f)^{0.8234} \qquad \ldots(6)$$

Hexagonal bed;

$$U_{ms} = 0.290(d_p/D_c)^{0.6478}(D_c/h_s)^{0.215}(\rho_p/\rho_f)^{0.5292} \qquad \ldots(7)$$

Square bed;

$$U_{ms} = 0.863(d_p/D_c)^{1.25}(D_c/h_s)^{0.057}(\rho_p/\rho_f)^{0.78} \qquad \ldots(8)$$

With the help of Eqs (5) to (8), the minimum slugging velocities have been calculated for different data points and compared with their experimental values.

Bubbling bed index

Identical to fluidization index which gives the idea of the range of particulate fluidization in the form of a ratio of the minimum bubbling velocity to minimum fluidization velocity, it is proposed to define a term “Bubbling Bed Index” which is the ratio of the minimum slugging velocity and the minimum bubbling velocity and can predict the range of bubbling fluidization for gas-solid system.

Results and Discussion

The values of minimum slugging velocities calculated with the help of Eqs (5) to (8) have been compared with their respective experimental values and also with the values calculated by Geldart's Equation (Eq. 2) in Figs 1 to 4. Fairly good agreement has been found to exist between the calculated and experimental values. The values of minimum slugging velocity calculated by Geldart's equation are close to the experimental values for smaller and lighter particles, but the values calculated by the proposed

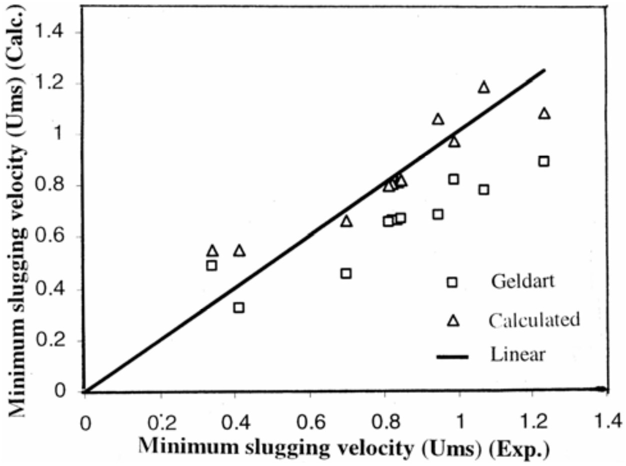

Fig. 1—Comparison of experimental and calculated values of minimum slugging velocity – cylindrical bed

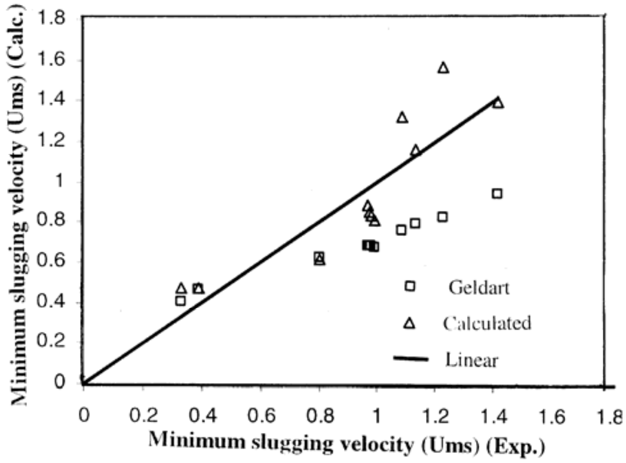

Fig. 2—Comparison of experimental and calculated values of minimum slugging velocity – semi-cylindrical bed

Table 3—Values of constants and exponents for different beds

Bed	Constant, *k*	Exponent, *a*	Exponent, *b*	Exponent, *c*	Exponent, *n*
Cylindrical	0.136	0.9394	0.0654	0.9747	0.9143
Semi-cylindrical	0.269	1.2117	0.2381	0.9739	0.8455
Hexagonal	0.290	0.8464	0.2811	0.6919	0.7649
Square	0.863	1.4314	0.0652	0.8935	0.8732

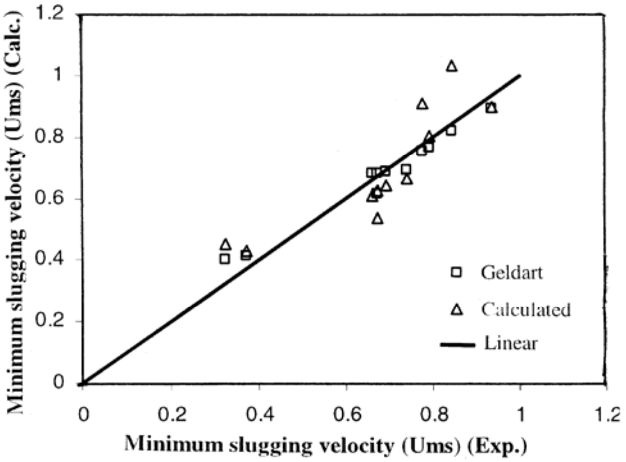

Fig. 3—Comparison of experimental and calculated values of minimum slugging velocity – hexagonal bed

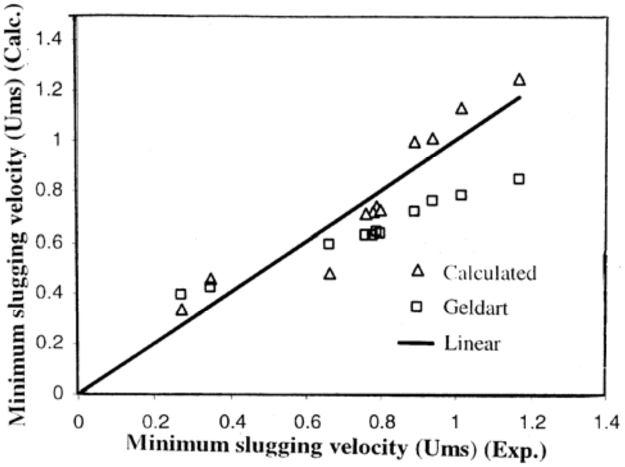

Fig. 4—Comparison of experimental and calculated values of minimum slugging velocity – square bed

Table 4—Comparison of minimum slugging velocity (m.s^{-1}) in different beds

D_p*10^4, m	Density, kg.m^{-3}	Cylin-drical	Semi-cylindrical	Square	Hex-agonal
3.24	2740	0.412	0.3300	0.2737	0.325
4.26	2740	0.7023	0.8033	0.6647	0.670
6.0	2740	0.8469	0.9690	0.8015	0.740
7.8	2740	0.9915	1.1343	0.9384	0.790
9.25	2740	1.2393	1.4179	1.1730	0.937
6.0	4800	1.0741	1.229	1.0166	0.842
6.0	4050	0.9502	1.087	0.8929	0.775
6.0	1500	0.3388	0.3880	0.3512	0.370

Table 5—Comparison of bubbling bed index in different beds

D_p*10^4, m	Density, Kg.m^{-3}	Cylin-drical	Semi-cylindrical	Square	Hex-agonal
3.24	2740	2.019	1.239	1.274	1.261
4.26	2740	2.294	1.700	1.788	1.521
6.0	2740	1.643	1.841	2.017	1.377
7.8	2740	1.513	1.731	1.782	1.259
9.25	2740	1.722	1.737	1.915	1.228
6.0	4800	1.678	1.788	1.820	1.225
6.0	4050	1.656	1.807	1.807	1.378
6.0	1500	1.051	1.389	1.923	1.722

Table 6—Comparison of range of bubbling fluidization (m.s^{-1}) in different beds

D_p*10^4, m	Density, kg.m^{-3}	Cylin-drical	Semi-cylindrical	Square	Hex-agonal
3.24	2740	0.2079	0.0636	0.0589	0.0670
4.26	2740	0.3962	0.3307	0.2930	0.2296
6.0	2740	0.3313	0.4427	0.4041	0.2029
7.8	2740	0.3362	0.4791	0.4121	0.1627
9.25	2740	0.5196	0.6015	0.5607	0.1734
6.0	4800	0.4391	0.5415	0.4580	0.1545
6.0	4050	0.3766	0.4855	0.3988	0.1519
6.0	1500	0.1066	0.1087	0.1686	0.1552

equations agree well for all the range of particles used in the experimentation. Minimum slugging velocity, bubbling bed index and the range of bubbling fluidization in different conduits have been compared in Tables 4 to 6 respectively. Based on the experimental data it is concluded that, under similar operating conditions, the minimum slugging and minimum bubbling velocities are maximum in case of the semi-cylindrical conduit and minimum in case of the square one. It is further observed that, the bubbling bed index is maximum in case of square bed but the range of bubbling fluidization is maximum in case of the semi-cylindrical bed for identical operating conditions. In view of the fact that the range of bubbling fluidization is maximum in case of semi-cylindrical bed, this will, therefore, be a better substitute for the conventional one when bubbling fluidization is desired to meet the process requirement.

Nomenclature

d_p:	Particle diameter	[m]
D_c:	Column diameter	[m]
G_f:	Fluid mass velocity	[kg.h^{-1}.m^{-2}]
G_{mf}:	Fluid mass velocity at minimum fluidization	[kg.h^{-1}.m^{-2}]
h_s:	Static bed height	[m]
ρ_f:	Fluid (medium) density	[kg.m^{-3}]
ρ_p:	Particle density	[kg.m^{-3}]
U_{mf}:	Minimum fluidization velocity	[m s^{-1}]
U_{mb}:	Minimum bubbling velocity	[ms^{-1}]
U_{ms}:	Minimum slugging velocity	[ms^{-1}]

References

1 Geldart D, *Powder Technol*, 7 (1973) 285.

2 Abrahamsen A R & Geldart D, *Powder Technol*, 26 (1980) 35.

3 Stewart P S B & Davidson J F, *Powder Technol*, 1 (1967) 61.

4 Baeyens J & Geldart D, *Chem Eng Sci*, 29 (1974) 255.

5 Queiroz C A R, Cavalho R J & Moura F J, *Brazilian J Chem Eng*, 22(1) (2005) 117.

6 Singh R K & Roy G K, *Powder Technol*, 159 (2005) 168.

7 Singh R K & Roy G K, *Indian J Chem Technol*, 13 (2006) 139.

8 Singh R K, *Studies on certain aspects of gas-solid fluidization in non-cylindrical beds*, Ph. D. Thesis, Sambalpur University, 1997.