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Prediction of minimum bubbling velocity, fluidization index and range of particulate fluidization for gas–solid fluidization in cylindrical and non-cylindrical beds

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Abstract

A uniform fluidization exists between minimum fluidization velocity and minimum bubbling velocity. Experimental investigations have been carried out for determination of minimum bubbling velocity and fluidization index for non-spherical particles in cylindrical and non-cylindrical beds. In the present paper equations have been developed for the prediction of minimum bubbling 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. A fairly good agreement has been obtained between calculated and experimental values. Based on the experimental data it is concluded that under similar operating conditions minimum bubbling velocity and the fluidization index are maximum in case of either semi-cylindrical conduit or hexagonal conduit for most of the operating conditions and minimum in case of square one. It is further observed that the range of uniform (particulate) fluidization is maximum in case of semi-cylindrical bed for identical operating conditions.

Keywords: Gas–solid fluidization; Cylindrical and non-cylindrical beds; Minimum bubbling velocity; Fluidization index; Particulate fluidization

1. Introduction

Particulate fluidization exists between minimum fluidization velocity and minimum bubbling velocity. Generally particulate fluidization occurs with liquid–solid systems, sometimes it also occurs with gas–solid systems when the particles are very fine but over a limited range of velocity. The superficial gas velocity at which bubbles first appear is known as the minimum bubbling velocity. The ratio of minimum bubbling velocity to minimum fluidization velocity, U_{mb} / U_{mf}, is known as the fluidization index, which gives a measure of the degree to which the bed can be expanded uniformly. This ratio tends to be relatively high for Geldart Group-A powders and for gas of high density reported by Davidson and Harrison [1] and Rhodes [2]. Range of gas velocities over which non-bubbling, i.e., particulate fluidization occurs is small. Davies and Richardson [3] have obtained the values for U_{mb} / U_{mf} up to 2.8 using cracker catalyst (d_p = 55 μ, s.g. = 0.95) fluidized in air at atmospheric pressure. Abrahamsen and Geldart [4] correlated the values of minimum bubbling velocity with gas and particle properties as follows:

$$U_{\mathrm{mb}} = 2.07\exp(0.716F)(x_{\mathrm{p}}{*}\rho_{\mathrm{g}}^{0.06}/\mu^{0.347}) \tag{1}$$

where F is the fraction of powder less than 45 μm.

Minimum fluidization velocity for particles less than 100 μm is given by Baeyens equation,

$$U_{\mathrm{mf}} = (\rho_{\mathrm{p}} - \rho_{\mathrm{f}})^{0.934} g^{0.934} x_{\mathrm{p}}^{1.8}/(1100{*}\mu^{0.87}\rho_{\mathrm{g}}^{0.066}) \tag{1A}$$

As Fluidization Index is the ratio of minimum bubbling velocity to minimum fluidization velocity, dividing the Abrahamsen equation by the Baeyens equation, the correlation obtained is

$$U_{\mathrm{mb}}/U_{\mathrm{mf}} = 2300\rho_{\mathrm{g}}^{0.126}\mu_{\mathrm{g}}^{0.523}\exp(0.176F) / d_{\mathrm{p}}^{0.8} g^{0.934}(\rho_{\mathrm{p}} - \rho_{\mathrm{g}})^{0.934} \tag{2}$$

The higher the ratio, the bed can hold more gas between the minimum fluidization and bubbling point. This means that for a correct initial aeration rate between these two values the bed will be less likely to form bubbles for a small increase in velocity and less likely to deaerate due to a reduction in velocity [5].

In a sense a high fluidization index implies that the catalyst has a certain plasticity and can be expanded, contracted and bent around corners. A low fluidization index implies a brittle fluidization state where a small change could cause a break from the uniformly fluidized catalyst to a bubbling regime or a packed bed [5].

In some fluidizer system bubbles occur at velocities, which is very close to minimum fluidization velocity and in others at values much greater than minimum fluidization velocity, sometimes three times. The range of smoothly, quiescent fluidization is extended by using fluid of high density or operating at higher pressure because the fluidization number increases slightly with pressure and viscosity of the fluidizing medium. Many powders of average particle size less than 100 μm will expand uniformly without bubble formation over a limited range of gas velocity greater than minimum fluidization velocity. Fluidization index can vary from a ratio barely distinguishable from unity to as great as 2 or so in special cases. With materials like fine cracking catalyst, the ratio is typically around 1.2.

An increased number of fine catalyst particles having diameters less than 40 μm will improve circulation in most units. Higher the fluidization index, the more gas flows interstitially. The amount of gas flowing interstitially is a function of the fines content. The effect of the higher fines content is therefore to give a higher fluidization index.

Correlated values for the fluidization index are useful for monitoring a unit on a given catalyst and can be used to interpret trends which might result in fluidization problem. However correlations are not always meaningful when comparing different catalysts; laboratory measurements are then required. One reason for the differences in calculated and measured values is thought to be due to particle shape and its effects on drag and minimum fluidization velocity. It is better to measure U_{mf} and U_{mb} than to rely on correlations [5].

Although a few qualitative explanations relating to fluidization quality have been presented in terms of the bed parameters for cylindrical beds by previous investigators, their effects in case of non-cylindrical column remains unexplored. With this end in view, studies relating to quantification of fluidization quality in terms of minimum bubbling velocity, fluidization index and range of particulate fluidization for a cylindrical bed and three non-cylindrical beds, viz. the semi-cylindrical, square and hexagonal ones have been taken up.

2. Experimental

The experimental setup is shown in Fig. 1. All the cylindrical and non-cylindrical beds were made of transparent acrylic resin so that the bed behavior 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 properties of the bed materials are given in Table 1 and Table 2, respectively.

1. Experimental set up.

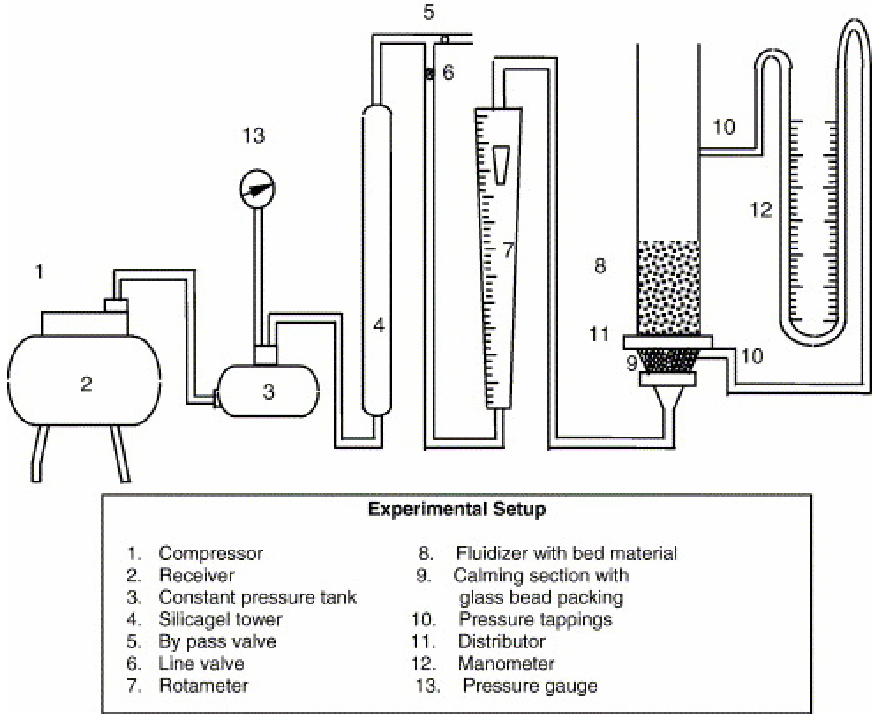

Table 1.

Dimension of bed 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 [6]

Material	**Density, kg/m^3**	**d_p * 10^4,m**	**Static bed porosity, ε_0**	**Minimum fluidization velocity, U_{mf}, m/s**
A. Cylindrical bed				
Dolomite	2740	9.0	0.550	0.71

Material	**Density, kg/m^3**	**d_p * 10^4,m**	**Static bed porosity, ε_0**	**Minimum fluidization velocity, U_{mf}, m/s**
Dolomite	2740	7.8	0.538	0.64
Dolomite	2740	6.0	0.526	0.49
Dolomite	2740	4.26	0.520	0.28
Dolomite	2740	3.24	0.515	0.15
Manganese ore	4800	6.0	0.568	0.60
Chromite ore	4050	6.0	0.522	0.51
Coal	1500	6.0	0.543	0.31
B. Semi-cylindrical bed				
Dolomite	2740	9.0	0.468	0.69
Dolomite	2740	7.8	0.470	0.54
Dolomite	2740	6.0	0.488	0.43
Dolomite	2740	4.26	0.516	0.37
Dolomite	2740	3.24	0.537	0.15
Manganese ore	4800	6.0	0.527	0.56
Chromite ore	4050	6.0	0.546	0.49
Coal	1500	6.0	0.563	0.23
C. Square bed				
Dolomite	2740	9.0	0.490	0.60
Dolomite	2740	7.8	0.492	0.52
Dolomite	2740	6.0	0.496	0.39
Dolomite	2740	4.26	0.512	0.34
Dolomite	2740	3.24	0.520	0.15
Manganese ore	4800	6.0	0.531	0.54
Chromite ore	4050	6.0	0.476	0.47
Coal	1500	6.0	0.516	0.17

Material	Density, kg/m^3	d_p * 10^4,m	Static bed porosity, ε_0	Minimum fluidization velocity, U_{mf}, m/s
D. Hexagonal bed				
Dolomite	2740	9.0	0.502	0.64
Dolomite	2740	7.8	0.513	0.52
Dolomite	2740	6.0	0.520	0.44
Dolomite	2740	4.26	0.535	0.37
Dolomite	2740	3.24	0.547	0.15
Manganese ore	4800	6.0	0.501	0.59
Chromite ore	4050	6.0	0.498	0.51
Coal	1500	6.0	0.526	0.21

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 obtained. The minimum fluidization velocity was obtained from the plot of pressure drop versus gas velocity. The gas velocity was increased slowly after the minimum fluidization condition and it was noted when the first bubble appeared as minimum bubbling velocity.

3. Development of correlations

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

For dimensional analysis the minimum bubbling velocity can be related to the system parameters as follows:

$$U_{mb}=f(d_p/D_c, D_c/h_s, \rho_p/\rho_f) \quad (3)$$

Eq. (3) can be rewritten as

$$U_{mb}=k\cdot((d_p/D_c)^a(D_c/h_s)^b(\rho_p/\rho_f)^c)^n \quad (4)$$

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

The effect of individual groups on minimum bubbling velocity have been separately evaluated for different conduits by plotting minimum bubbling velocity versus individual group and 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_{mb} against log $((d_p / D_c)^a(D_c / h_s)^b(\rho_p / \rho_f)^c)$ for different conduits.

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;

$$U_{mb}=0.5231(d_p/D_c)^{1.13}(D_c/h_s)^{-0.0384}(\rho_p/\rho_f)^{0.74} \quad (5)$$

Semi-cylindrical bed;

$$U_{mb}=0.168(d_p/D_c)^{0.994}(D_c/h_s)^{-0.1849}(\rho_p/\rho_f)^{0.80} \quad (6)$$

Hexagonal bed;

$$U_{mb}=0.15(d_p/D_c)^{0.5733}(D_c/h_s)^{-0.0887}(\rho_p/\rho_f)^{0.5384} \quad (7)$$

Square bed;

$$U_{mb}=0.168(d_p/D_c)^{0.27}(D_c/h_s)^{-0.0132}(\rho_p/\rho_f)^{0.2825} \quad (8)$$

With the help of Eqs. Figs. (5), (6), (7) and (8), the minimum bubbling velocities have been calculated for different experimental data points and have been compared with their experimental values.

4. Results and discussions

The values of minimum bubbling velocity calculated with the help of Eqs. Figs. (5), (6), (7) and (8) have been compared with their respective experimental values in Fig. 2, Fig. 3, Fig. 4 and Fig. 5. Fairly good agreement has been found to exist between calculated and experimental values. Minimum bubbling velocity, fluidization index and range of particulate fluidization have been compared in Table 3, Table 4 and Table 5. For identical operating conditions minimum bubbling velocity and fluidization index are maximum in case of either semi-cylindrical bed or hexagonal bed for most of the operating conditions and least in case of square bed. The range of particulate fluidization is maximum again in case of semi-cylindrical bed and less in case of other beds used. Hence where particulate fluidization is the requirement of the operation, semi-cylindrical column is a better substitute.

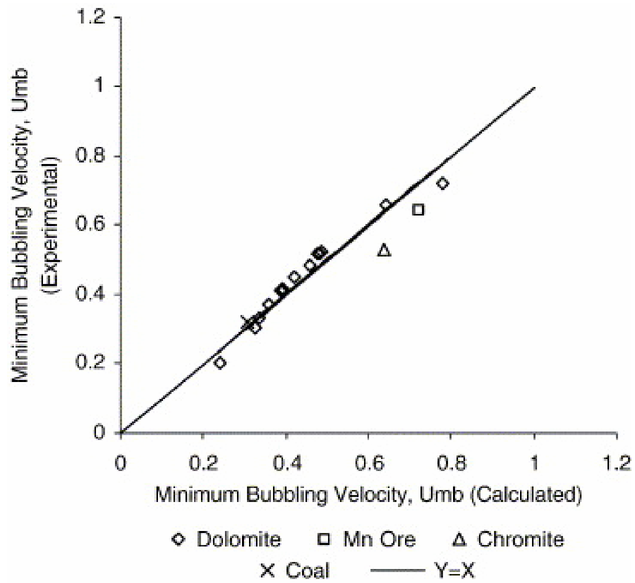

Fig. 2. Comparison of experimental and calculated values of minimum bubbling velocity: cylindrical bed.

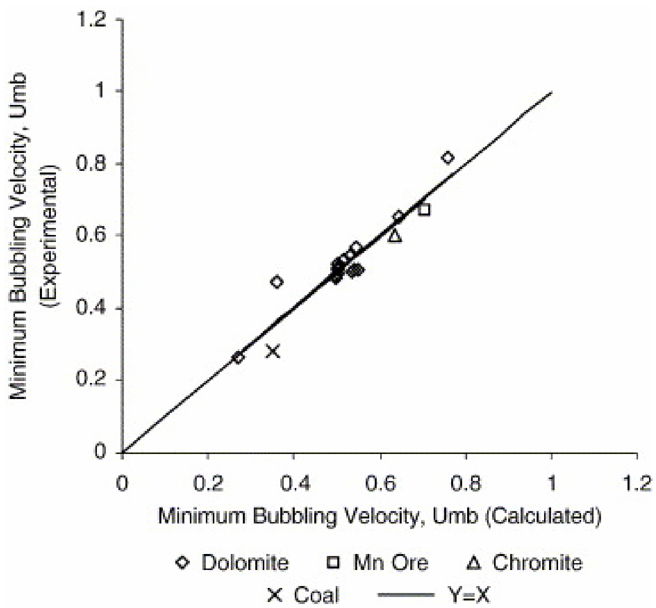

Fig. 3. Comparison of experimental and calculated values of minimum bubbling velocity: semi-cylindrical bed.

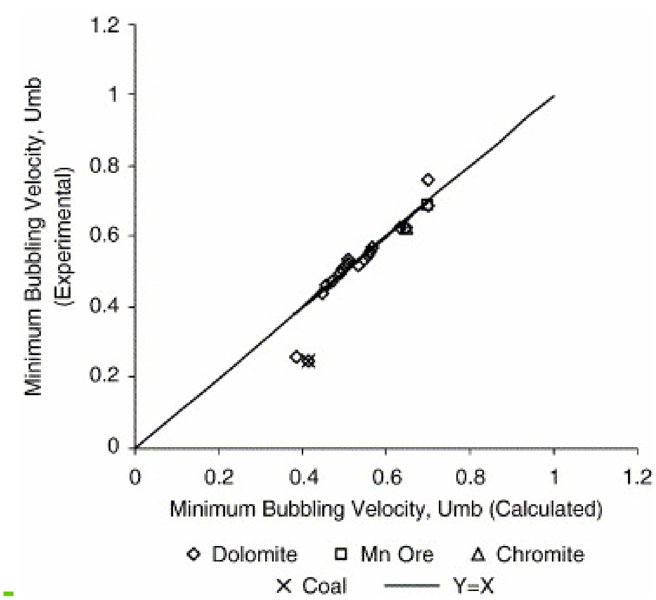

Fig. 4. Comparison of experimental and calculated values of minimum bubbling velocity: hexagonal bed.

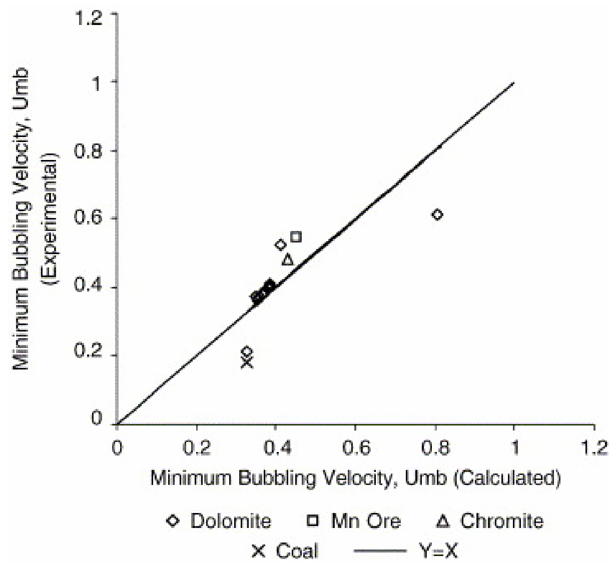

Fig. 5. Comparison of experimental and calculated values of minimum bubbling velocity: square bed.

Table 3.

Comparison of minimum bubbling velocity (m s^{-1}) in different beds

d_p * 10^4, m	**Density, kg/m^3**	**Cylindrical**	**Semi-cylindrical**	**Square**	**Hexagonal**
9.0	2740	0.72	0.82	0.61	0.76
7.8	2740	0.66	0.56	0.53	0.62
6.0	2740	0.52	0.53	0.40	0.54
4.26	2740	0.31	0.47	0.37	0.44
3.24	2740	0.21	0.27	0.22	0.26
6.0	4800	0.64	0.67	0.55	0.69
6.0	4050	0.53	0.60	0.48	0.62
6.0	1500	0.32	0.28	0.18	0.25

Table 4.

Comparison of fluidization index in different beds

d_p * 10^4, m	Density, kg/m^3	Cylindrical	Semi-cylindrical	Square	Hexagonal
9.0	2740	1.014	1.188	1.017	1.188
7.8	2740	1.031	1.037	1.019	1.192
6.0	2740	1.061	1.233	1.026	1.227
4.26	2740	1.107	1.270	1.088	1.189
3.24	2740	1.400	1.800	1.467	1.733
6.0	4800	1.067	1.196	1.0185	1.169
6.0	4050	1.039	1.224	1.021	1.216
6.0	1500	1.033	1.217	1.059	1.190

Table 5.

Comparison of range of particulate fluidization ($m\ s^{-1}$) in different beds

d_p * 10^4, m	Density, kg/m^3	Cylindrical	Semi-cylindrical	Square	Hexagonal
9.0	2740	0.01	0.13	0.01	0.12
7.8	2740	0.02	0.12	0.01	0.10
6.0	2740	0.03	0.10	0.01	0.10
4.26	2740	0.03	0.10	0.03	0.07
3.24	2740	0.06	0.12	0.07	0.11
6.0	4800	0.04	0.11	0.01	0.10
6.0	4050	0.02	0.11	0.01	0.11
6.0	1500	0.01	0.05	0.01	0.04

Nomenclature

d_p

Particle diameter [m]

D_c

Equivalent 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 density [kg m^{-3}]

ρ_p

Particle density [kg m^{-3}]

U_{mf}

Minimum fluidization velocity [m s^{-1}]

U_{mb}

Minimum bubbling velocity [m s^{-1}]

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