Published in Chemical Engineering World, Jan 2006, P 38-42
Archived in Dspace@nitr http://dspace.nitrkl.ac.in/dspace

Hydrodynamic Studies of Gas-Solid Fluidization in Square Bed for Non-spherical Particles.

R. K. Singh
Department of Chemical Engineering
National Institute Of Technology
Rourkela-769008 (Orissa)

Introduction

Fluidization as an established fluid-solid contacting technique has found extensive applications in unit operations like drying, adsorption and in chemical processes Viz. solid-catalized reactions, carbonization, gasification and combustion. A fluidized bed can be achieved by increasing the upward velocity of the fluid through a fixed bed. Fluidized bed technique as compared to fixed bed has specific advantages. Albeit many advantages claimed of fluidization, the efficiency and the quality in large scale and deep gas-solid beds are seriously affected by bubbling, channeling and slugging behaviors, resulting in poor gas-solid contact, lower diffusion and heat transfer rates.

Some remedial measures proposed for the a-fore-said problems and thereby improving the quality of fluidization include the incorporation of baffles in the bed, imparting vibrations to the column, operation in multi-stage unit and the use of conical beds. Investigations in the field of dynamic studies relating to various aspects of gas-solid fluidization have been carried out by many investigators.

First of all a square bed of cross section 250 mm * 250 mm was used by Ostergaard (1) to take a series of cine-photograph of large bubbles emerging from a bed of 20 kg ballotini. Empirical equations were provided for fractional bed volumes occupied by gas bubbles and bubble wakes.

The fractional bed volume occupied by gas bubbles, is related to the bubble velocity (Ub) and the average superficial gas velocity (U_g) by the following expression:

$$\varepsilon_g = U_g * U_b \qquad (1)$$

The fractional bed volume occupied by the bed wakes,

$\varepsilon_w = 0.14\ \varepsilon_g^{0.5}(U_l\text{-}U_{lmf})$ (2)

Nicklin (2) suggested the following relationship between U_b and U_g for bubble flow in water at zero liquid velocity,

$U_b = 21.7\text{-}4.6 \ln U_g$. (3)

(the velocity unit being cm/sec.)

At finite liquid velocities the above equation was modified by Ostergaard (loc. cit.) as

$U_b = 21.7\text{-}4.6 \ln U_g + U_l$ (4)

The fluidization characteristics of large particles (1000-2000 microns) have been studied by Cranfield et al (3) using bed depth up to 30 cm in equipment having a cross-section of 61 cm* 61 cm. Eruption diameters at the 3-D bed surface were determined for a sufficient period of time from cine-film methods to obtain a representative size distribution and mean bubble size calculated from

$D_{\mathbf{b}}=[\sum_1^{Nb} Nd^3 / \sum_1^{Nb} N]^{1/3}$ (5)

They also correlated the bubble rise velocity as

$U_b = 22.25*(d_b)^{1/2}$ (6)

Mixing and segregation was studied in fluidized beds of circular (0.21 m diameter) and square (0.61*0.61 m^2 and 1.22*1.22 m^2) cross-section by Peeler et al (4). Nguyen et al (5) gave some information with respect to the effect of bubble distribution on back mixing in a square bed, 1.22 m * 1.22 m. In such a bed, when a pattern arises in which bubbles are appearing regularly at positions approximately in the four corners then a central downflow is strongest. The pressure differential between centre and wall is at a maximum. Whitehead (1967) found pattern of bubble tracks in a large 1.2 m square bed similar to the pattern observed by Werther: preferred bubble tracks near the walls and corners of a shallow open bed and merging of bubbles towards the bed centre at higher elevation. Newby and Keairns (6) reported a dramatic effect of the tube bundle on bubble size and bubble distribution in a 0.37 m * 0.15 m rectangular fluidized bed. Nguyen et al (5) also studied bubbling patterns at the surface of the bed, 1.22 m* 1.22 m, in presence of horizontal tube bundles. They observed for open beds that the bubbling points are in a pattern of four bubbling zones, with each bubbling zone in one of the corners. When the horizontal tube bundle is present, the bubbles are distributed more or less uniformly over the whole cross-section. The solids down flow can be expected to occur in numerous small streams subject to splitting by the tubes as the stream descend. It was also noted that the bubble eruption diameters are much less in the presence of the tube than without them.

Nguyen et al (7) reported experimental data on gas mixing in a large fluidized bed, 1.22 m * 1.22 m square bed. Nguyen et al (8) extended previously reported data for open beds by examining the effect on backmixing of the insertion in the bed of horizontal or vertical tube assemblies.

Zhou et al (9) obtained the voidage profile in a circulating fluidized bed of square cross-section. Branislav and Mladen (10) investigated the effect of fluidization velocity on heat transfer fluidized bed and inclined heat transfer surfaces. Their investigation was conducted on laboratory scaled apparatus, of the square section 160*160 cm. to define fluidization velocity, temperature measurement on front, lateral and back side of the heater, relating to the fluidization air flowing direction for different angle of inclination.

Cohesive forces between particles can have a significant effect on the bulk flow in fluidized beds. These forces can come from a number of sources including liquid bridging, van der Waals forces and electrostatic forces. Mild cohesive forces can lead to changes in the minimum fluidization velocity, the minimum bubbling velocity and the bed expansion. High levels of cohesion can lead to total de-fluidization and the formation of slugs and channels.

Though bubble behavior and mixing and segregation have been studied by different investigators, the complete hydrodynamic studies have not been done. In the present investigation the complete hydrodynamic studies like, the minimum fluidization velocity, the pressure drop across the bed, the expansion ratio, the fluctuation ratio, the minimum bubbling velocity, the minimum slugging velocity and the range of bubbling fluidization have been studied and these behaviors have been compared with those in cylindrical beds.

Experimental:

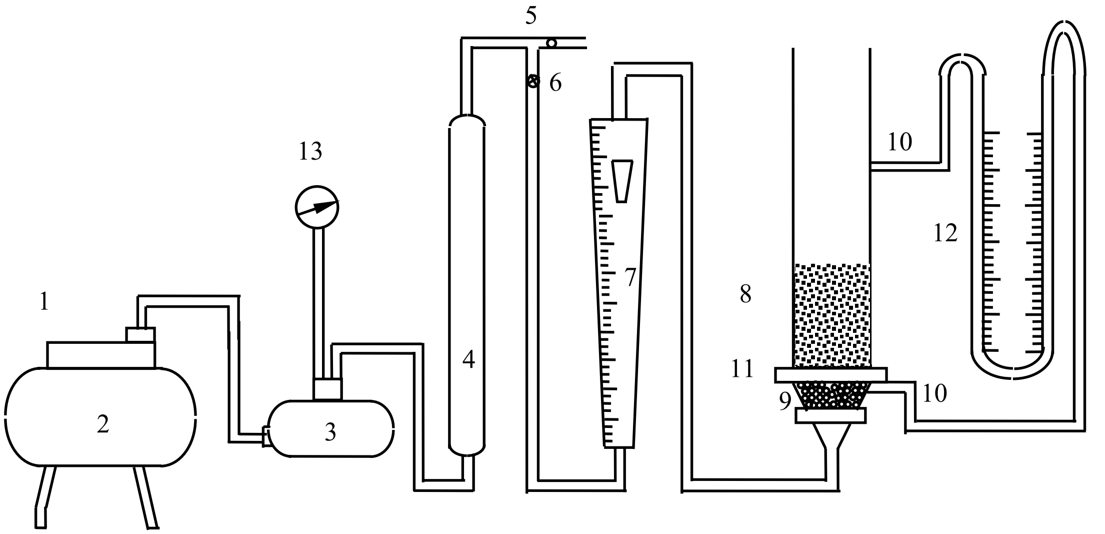
The experimental setup is shown in Fig. 1. A square bed of cross-section 8.2 cm * 8.2 cm, made of transparent acrylic resin was used so that the bed behaviors could be observed clearly. For uniform distribution of fluidizing medium in the bed, a calming section with glass beads was used at the entrance of the column. Air was used as fluidizing medium after drying it in silica-gel tower. The properties of the bed materials are given in Tables 1.

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 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.

In the fluidized state, as the top layer of the bed was fluctuating, both levels (maximum and minimum) and the bed pressure drop were noted against flow rate.

Figure – 1: Experimental Setup

1. Compressor
2. Receiver
3. Constant pressure tank
4. Silicagel tower
5. By pass valve
6. Line valve
7. Rotameter
8. Fluidizer with bed material
9. Calming section with glass bead packing
10. Pressure tappings
11. Distributor
12. Manometer
13. Pressure gauge



Results and coclusions:

Minimum fluidization velocity and pressure drop across the bed at incipient fluidization have been obtained from the pressure drop versus fluid mass velocity plot. Minimum fluidization mass velocity (G_{mf}) depends on the porosity of the bed corresponding to the minimum fluidization conditions. Minimum fluidization velocity and bed pressure drop at incipient condition have been compared in Table 2**.** It has been found that under identical operating conditions, minimum fluidization mass velocity was less in case of a square bed. Since operating mass velocity of industrial fluidizers depends on the minimum fluidization mass velocity, a lower value of G_{mf} for a square bed will have an edge over the conventional circular bed, thus permitting the use of a lower operating velocity for identical process situation. It is also observed that pressure drop at minimum fluidization condition is also less in case of square bed, thereby comparatively reduced energy input for the fluidizing medium.

Bed expansion ratio have been compared for cylindrical and square beds with identical excess velocity ratio and wall effect for an aspect ratio value of nearly unity in Table 3. Under similar operating conditions, fluidized bed height will be more in case of square bed.

Bed fluctuation ratios have been compared for cylindrical and square beds in Table 4. It is observed that under similar operating conditions, fluctuation ratio is higher in case of square bed.

Minimum bubbling velocity, fluidization index and range of particulate fluidization have been compared in Table 5.For identical operating conditions minimum bubbling velocity, fluidization index and the range of particulate fluidization are less in case of square bed.

Minimum slugging velocity, bubbling bed index and the range of bubbling fluidization have been compared in Tables 6. Based on the experimental data it is concluded that under similar operating conditions minimum slugging velocity is less in case of square bed. It is further observed that the bubbling bed index and range of bubbling fluidization are more in square bed than cylindrical bed.

In view of the fact that the range of bubbling fluidization is more in case of the square, this will, therefore, be a better substitute for the conventional one when bubbling fluidization is desired to meet the process requirement.

Table 1 Properties of bed materials[11]

A. Cylindrical bed

Material	Density, kg/m^3	d_p*10^4,m	Static bed Porosity, ε_0
Dolomite	2740	9.0	0.550
Dolomite	2740	7.8	0.538
Dolomite	2740	6.0	0.526
Dolomite	2740	4.26	0.520
Dolomite	2740	3.24	0.515
Manganese ore	4800	6.0	0.568
Chromite ore	4050	6.0	0.522
Coal	1500	6.0	0.543

B. Square bed

Material	Density, kg/m^3	d_p*10^4,m	Static bed Porosity, ε_0
Dolomite	2740	9.0	0.490
Dolomite	2740	7.8	0.492
Dolomite	2740	6.0	0.496
Dolomite	2740	4.26	0.512
Dolomite	2740	3.24	0.520
Manganese ore	4800	6.0	0.531
Chromite ore	4050	6.0	0.476

Coal	1500	6.0	0.516

Table 2: Comparison of minimum fluidization mass velocity and pressure drop at incipient fluidization

d_p*10^4,m	Density, kg/m^3	G_{mf},Cyl. $Kg/hr.m^2$	G_{mf},sqr. $Kg/hr.m^2$	ΔP_{mf} Cyl	ΔP_{mf} Sqr
9.0	2740	3300	2800	840	760
7.8	2740	3000	2400	800	730
6.0	2740	2300	1800	780	740
4.26	2740	1300	1600	760	700
3.24	2740	700	680	700	660
6.0	4800	2800	2500	1350	1300
6.0	4050	2350	2200	1200	1180
6.0	1500	1450	800	500	460

Table 3: Comparison of the bed expansion ratio

d_p/D_c	D_c/h_s	$(G_f-G_{mf})/G_{mf}$	Expansion ratio Cylindrical bed	Expansion ratio Square bed
0.0059	1.069	0.50	1.3065	1.4734
0.0059	1.069	0.75	1.4055	1.6439
0.0059	1.069	1.00	1.4802	1.7769
0.0059	1.069	1.25	1.5408	1.8870
0.0059	1.069	1.50	1.5923	1.9823

Table 4: Comparison of the Bed Fluctuation Ratio

Sl. No.	$(D_p/Dc)*103$	$(D_C/h_S)*(\rho_f/\rho_S)*10^4$	$(G_r–G_{mf})/G_{mf}$	Cylindrical bed	Square bed
1	5.9	5.04	0.25	1.07	1.12
2	5.9	5.04	0.50	1.11	1.19
3	5.9	5.04	0.75	1.13	1.23
4	5.9	5.04	1.00	1.15	1.26
5	5.9	5.04	1.25	1.16	1.28
6	5.9	5.04	1.50	1.17	1.3

Table 5**:** Comparison of minimum bubbling velocity ($m.s^{-1}$), fluidization index (F.I.) and range of particulate fluidization ($m.s^{-1}$) .

d_p*10^4, m	Density, kg/m^3	U_{mb} Cylindrical	U_{mb} Square	F.I. Cylindrical	F.I. Square	Range Cylindrical	Range Square
9.0	2740	0.72	0.61	1.014	1.017	0.01	0.01
7.8	2740	0.66	0.53	1.031	1.019	0.02	0.01
6.0	2740	0.52	0.40	1.061	1.026	0.03	0.01
4.26	2740	0.31	0.37	1.107	1.088	0.03	0.03
3.24	2740	0.21	0.22	1.400	1.467	0.06	0.07
6.0	4800	0.64	0.55	1.067	1.0185	0.04	0.01
6.0	4050	0.53	0.48	1.039	1.021	0.02	0.01
6.0	1500	0.32	0.18	1.033	1.059	0.01	0.01

Table 6 Comparison of Minimum Slugging Velocity ($m.s^{-1}$), Bubbling Bed Index and Range of Bubbling fluidization.

D_p*10^4, m	Density, $kg.m^{-3}$	Cyl.	Square	Cyl.	Square	Cylindrical	Square
3.24	2740	0.412	0.2737	2.019	1.274	0.2079	0.0589
4.26	2740	0.7023	0.6647	2.294	1.788	0.3962	0.2930
6.0	2740	0.8469	0.8015	1.643	2.017	0.3313	0.4041
7.8	2740	0.9915	0.9384	1.513	1.782	0.3362	0.4121
9.25	2740	1.2393	1.1730	1.722	1.915	0.5196	0.5607
6.0	4800	1.0741	1.0166	1.678	1.820	0.4391	0.4580
6.0	4050	0.9502	0.8929	1.656	1.807	0.3766	0.3988
6.0	1500	0.3388	0.3512	1.051	1.923	0.1066	0.1686

NOMENCLATURE :

d_p: particle diameter	[m]
D_c: column diameter	[m]
G_f: fluid mass velocity	[$kg.hr^{-1}.m^{-2}$]
G_{mf}: fluid mass velocity at minimum fluidization	[$kg.hr^{-1}.m^{-2}$]
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. Ostergaard, K., “On bed porosity in gas-liquid fluidization”, Chem. Engg. Sc., 20,165-167 (1965)
2. Nicklin, D.J., Chem. Engg. Sc., 17, 693 (1962)
3. Cranfield, R.R. and Geldart, D., “Large particle fluidization”, Chem. Engg. Sc., 29, 935-947 (1974)
4. Peeler, J.P.K. and Huang, J.R., “Segregation of wide size range particle mixtures in fluidized bed”, Chem. Engg. Sc., 44, 5, 1113-1119 (1989)
5. Nguyen, H.V., Potter, O.E. and Whitehead, A.B., “Bubble distribution and eruption diameter in a fluidized bed with a horizontal tube bundle”, Chem. Engg. Sc., 34, 1163-1164 (1979)
6. Newby, R.A. and Keairns, D.L., “Fluidized bed heat transfer between parallel horizontal tube bundles”, “Fluidization” Edited by Davidson, J.F. and Keairns, D.L., Cambridge Univ. Press, 320 (1978)
7. Nguyen, H.V., Whitehead, A.B. and Potter, O.E., “Gas back mixing, solid movement and Bubble activities in large scale fluidized bed”, A.I.Ch.E. Jl. 23, 913-922 (1977)
8. Nguyen, H.V., Potter O.E., Dent, D.C. and Whitehead, A.B., “Gas back mixing in large fluidized bed containing tube assemblies”, A.I.Ch.E. Jl., 27,3, 508-514 (1981)
9. Zhou, J., Grace, J.R., Qin, S., Brereton, C.M.H., Lim, C.J. and Zhu, J., “Voidage profile in a circulating fluidized bed of square cross-section”, Chem. Engg. Sc., 49, 3217-3226 (1994)
10. Branislav, S. and Mladen, S. “The influence of fluidization velocity on heat transfer between fluidized bed and inclined heat transfer surfaces”, The Scientific Journal FACTA Universitatis Series, Mechanical Engineering, Vol. 1, No. 6,703-710 (1999)
11. Singh, R.K, “Studies on certain aspects of gas-solid fluidization in non-cylindrical beds.” Ph. D. Dissertation, Sambalpur University, 1995.