Hydrodynamics Studies of Gas-Solid Fluidization in Non Cylindrical Conduits for Spherical and Non-Spherical Particles -A review

R.K.Padhi[*], Y.K.Mohanty[*], G.K.Roy[*], B.Sarangi[**]

[*] Chemical Engineering Department, G.I.E.T, Gunupur, Rayagada- 765022, Odisha, India
[**] Department of Chemical Engineering, IGIT, Saranga, Dhenkanal- 759146, Odisha, India

Abstract- Importance of noncylindrical fluidized bed for process operation has been emphasized.. The correlations and the models developed for gas-solid fluidized bed in noncylindrical conduits (square, Hexagonal, semi cylindrical and conical) as compared with cylindrical, with a special focus on the development during the last two decades.Limited investigation have been carried out in noncylindrical bed hence emphasis is given to recent investigation on hydrodynamics study in noncylindricalconduits,while dynamics for cylindrical bes have been exhaustively studied by different authors. Available studies relating to hydrodynamic properties viz. minimum fluidization velocity, bed pressure drop, bed expansion and fluctuation, bubbling and slugging velocity, bubbling and fluidizing index in non-cylindrical conduits have been detailed. The gray areas in non-cylindrical column hydrodynamics for future studies have been identified and potential application of such column have been suggested.

Index Terms- Gas–solid fluidization, non-cylindrical conduits, bed dynamics

I. Introduction

Fluidization is a phenomenon by which fine solids are transformed into a fluid-like state through contact with gas or liquid or by both gas and liquid. It is a fluid-solid contacting technique, which has found extensive industrial applications. Investigations relating to various aspects of fluidization as a novel fluid-solid contacting technique is being carried out since the world war –II and numerous process applications have been made based on these techniques like drying, adsorption and chemical processes such as combustion, carbonization, gasification and solid-catalyzed reaction.

A large number of review paper, reports and other articles show the advancement and improvement in fluidization technology. In order to keep this review to reasonable proportion, it has been confined to gas-solid systems . In the past, a number of excellent reviews have been published on measurement techniques for fluidized beds by Cheremisinoff [1], Grace and Baeyens [2], and Yates and Simons (3). More recently, Louge (4) reviewed experimental techniques for circulating fluidized beds. Werther (5) gave an overview of measurement techniques in fluidized beds, with emphasis on applicability in industrial practice. Chaouki et al. (6) extensively reviewed non-invasive measurement techniques for multiphase flows in general.

Cylindrical gas-solid fluidized beds have been employed extensively in process industries. Apart from the advantages of gas-solid fluidization in cylindrical beds, the efficiency and the quality in large diameter and deep beds suffer seriously due to certain inherent drawbacks such as channeling, bubbling and slugging behavior at gas velocity higher than the minimum fluidization velocity resulting in poor gas-solid contact, homogeneity of the fluid and ultimately quality of fluidization. Hence efforts have been made by the investigators to improve the quality of gas-solid fluidization. To overcome the above-mentioned drawbacks and to improve the fluidization quality techniques such as vibration and rotation of the bed, use of improved distributor and promoter , operation in multistage and Use of non-cylindrical conduits have been recommended by the investigators

Consideration of non-cylindrical conduits, instead of a conventional cylindrical one is considered to be an attractive alternative technique for smooth gas-solid fluidization by Kunni [7], Levey [8]. The introduction of vibration and rotation of the bed , turbulence promoters in a gas-solid fluidized bed enhance the fluidization quality by minimizing bubbling, channeling and slugging but the relative demerits of the above technique is the increase of pressure drop. Vibration and rotation may require additional accessories to the bed with some operational difficulties .Hence to improve the quality of fluidization, the use of noncylindrical conduits has been found to be more effective in controlling fluidization quality as compared to the other methods. Recently the use of non-cylindrical beds has begun to receive much attention for several applications because of a few advantages, like (i) the operation of the fluidizer over a wide range of superficial velocity, (ii) the possibility of fluidizing a wide range of particles of different sizes or densities, and (iii) intensive particle mixing.

Although qualitative explanations relating to fluidization quality have been presented in terms of the bed parameters for cylindrical beds by different authors, their effects in case of a non-cylindrical column remain unexplored. Keeping in view, studies relating to the quantification of fluidization quality in terms of minimum bubbling velocity, fluidization index and range of particulate fluidization in a non-cylindrical beds, viz. the semi-cylindrical, square, hexagonal and conical ones have been taken up and predominantly in this review preference is given for the correlations and models developed for gas-solid fluidized bed in non-cylindrical conduits as compared with cylindrical developed during last ten years.

II. Bed Dynamics of Gas-Solid Fluidized System in a Conical Bed

Conical fluidized bed is very much useful for the fluidization of wide spectrum of particles, since the cross sectional area is enlarged along the bed height from the bottom to the top. Therefore the velocity of the fluidizing medium is relatively high at the bottom, ensuring fluidization of the large particles and relatively low at the top, preventing entrainment of the small particles. Since the velocity of the fluidizing medium at the bottom is fairly high, this gives rise to low particle concentration, thus resulting in low reaction rate and reduced rate of heat release. Therefore the generation of high temperature zone near the distributor can be prevented. Due to the existence of a gas velocity gradient along the height of a conical bed, it has some favorable characteristics. The conical bed has been widely applied in many industrial processes such as biological treatment of waste water, immobilized bio-film reaction, incineration of waste-materials, coating of nuclear fuel particles, crystallization, roasting of sulfide ores, coal gasification and liquefaction, catalytic polymerization, fluidized contactor for sawdust and mixtures of wood residues and fluidization of cohesive powder.
The study of the hydrodynamic characteristics of fluidization in conical beds is focused on two areas viz. the liquid–solid and the gas–solid systems. Gas –solid systems generally behave in a quite different manner. With an increase in flow rate beyond minimum fluidization, large instability with bubbling and channeling of gas is observed. At higher flow rates agitation becomes more violent and the movement of solids becomes vigorous. In addition, the bed does not expand much beyond its volume at minimum fluidization called an aggregative fluidized bed.. Therefore proper characterization of the bed dynamics for the binary and the multi-component mixtures in gas solid systems is an important prerequisite for their effective utilization, where the combination of particle size ,density and shape factor influence fluidization behavior by Sau. et al [9].

Investigations in the field of dynamic studies relating to various aspects of gas-solid fluidization have been carried out by many investigators. Dynamic studies relating to various aspects of gas-solid fluidization in cylindrical conduits have been carried out by a numbers of investigator but limited study have been carried out in conical conduits. In view of the limited information available for conical gas-solid fluidization system in general the present study has been taken up to investigate a few important work on bed dynamics which are responsible for the quality of fluidization have been discussed thoroughly.

2.1 Minimum fluidization velocity and pressure drop

In view of its potential application in the field of gas solid systems, it is a pre-requisite that the dynamics of the bed be explicitly understood. Two bed characteristics of relevance in this context are the minimum fluidization velocity (V_{mf}) and the pressure drop $(-\Delta P_{mf})$ for a fluidized bed. In the course of experiments it has been observed that, at a particular velocity, the pressure drop reaches a maximum and the particles in the bed are lifted slightly upward by the fluid. This is followed by the particles at the bottom of the bed beginning to fluidize. Once the particles are unlocked there is a sharp decline in the pressure drop. Evidently, fluidization is initiated when the force exerted by the fluidizing medium flowing through the bed is equal to the total effective weight of the particles in the bed. It is assumed that the lateral velocity of the fluid is relatively small and can be neglected. The pressure drop through a packed bed over a differential height of dh is given by Ergun equation [10] as follows:

$$\frac{\Delta P}{H} = \frac{150 U_{mf} \mu_g (1-\varepsilon_{mf})^2}{\phi_s^2 dp_{sm}^2 \varepsilon_{mf}^2} + \frac{1.75 U_{mf}^2 \rho_g (1-\varepsilon_{mf})}{\phi_s dp_{sm} \varepsilon_{mf}^2} \qquad (1)$$

i.e. $dP = (AU + BU^2)dh$

Where, $A = \frac{150 U_{mf} \mu_g (1-\varepsilon_{mf})^2}{\phi_s^2 dp_{sm}^2 \varepsilon_{mf}^2}$ and $B = \frac{1.75 U_{mf}^2 \rho_g (1-\varepsilon_{mf})}{\phi_s dp_{sm} \varepsilon_{mf}^2}$

For conical Bed the overall pressure drop across the bed height, H, is obtained by integrating Ergun equation [1]

$$-\Delta P = \int_{h_0}^{H+h_0} -(dP) = \int_{h_0}^{H+h_0} -(AU + BU^2)dh \qquad (2)$$

For a conical bed with apex angle of α

$$\frac{U}{U_0} = \frac{D_0^2}{D^2}, \frac{D}{D_0} = \frac{h}{h_0} \text{ and } U = U_0 \frac{h_0^2}{h^2} \qquad (3)$$

On integration Ergun equation becomes

$$-\Delta P = AU_0 \frac{H.h_0}{(H+h_0)} + \frac{BU_0^2 h_0}{3}\left[1 - \frac{h_0^3}{(H+h_0)^3}\right] \qquad (4)$$

$$h_0 = \frac{D_0}{2\tan(\alpha/2)}$$

The area of cross-section-of a conical bed increases continuously from the bottom to the top. So the force in a differential bed height of dh is equal to the product of the pressure drop through it, - (dp), and the cross sectional area.

On integration for a conical bed, we get the overall effective weight of the particle in the bed is

$$G = \frac{K[(H+h_0)^3 - h_0^3]}{3} = C_1 \qquad (5)$$

Where, $K = \frac{g(1-\varepsilon)(\rho_s - \rho_f)\pi D_0^2}{4 h_0^2}$

The overall force exerted by fluidizing medium on the particle is

$$F = A_n(-\Delta P) = A_n(AU + BU^2)H \qquad (6)$$

According to the proposed model as given by Agarwal et al [11] , the particles at the bottom start fluidizing when F=G. Therefore the minimum fluidization velocity, Umf can be found out by equating F =G, that give

$$A_1 U_{mf}^2 + B_1 U_{mf} = C_1$$

$$U_{mf} = \frac{-B_1 \pm \sqrt{B_1^2 + 4A_1C_1}}{2A_1} \qquad (7)$$

$Where A_1 = A_n BH$, $B_1 = A_n AH$, $C_1 = g(1-\varepsilon)(\rho_s - \rho_f)A_n H$

For conical fluidized Bed

$A_1 = \frac{\pi BHD_0^2}{4\left(D_0 + 2H\tan\frac{\alpha}{2}\right)}$, $B_1 = \frac{\pi AHD_0^2}{4}$ and $C_1 = \frac{K[(H+h_0)^3 - h_0^3]}{3}$

ISSN 2250-3153

Putting $U_{0} = U_{mf}$ and after integration of Ergun equation the pressure drop is given as

$$-(\Delta P_{mf}) = AHh_0 U_{mf} + \frac{Bh_0\left[(H+h_0)^2 - h_0^2\right]U_{mf}}{2(H+H_0)^2} \qquad (8)$$

The porosity of the bed was calculated from the measurement of height and area of the bed, mass and density of the particles. The sphericityhas been calculated from the correlation of Narsimhan [12] for binary and ternary mixtures ,

$$\frac{1-\varepsilon}{\varphi_s} = 0.231 \log dp_{sm} + 1.417 \qquad (9)$$

On the basis of Ergun's equation and Baskakov and Gelperm's modification for cone geometry,a packed bed pressure drop equation for conical beds has been developed by Biswal.et.al.[13], for gas solid systems as under,

$$-\Delta P_{mf} = \cos\frac{\alpha}{2}\left[37.17[\tan(\alpha)]^{-0.47}\frac{\mu\,\varepsilon(1-\varepsilon)^2 R_0(R-R_0)}{dp_{sm}{}^2 \varepsilon R} + \frac{0.75\rho f(1-\varepsilon)R_0\left[1-\left({R_0}/{R}\right)^3\right]U_0{}^2}{3dp_{sm}\varepsilon^3}\right] \qquad (10)$$

In order to develop correlations for ternary systems, it is necessary to define the particle diameter and the density of the ternary systems. In this study, they are defined byGoossens et al. [14] as,

$$Dp_{sm} = \frac{1}{\Sigma \frac{X_i}{dp_i}}$$

For heterogeneous ternary mixture $\frac{1}{dp_{sm}\rho_{sm}} = \frac{w_1}{dp_1\rho_{s1}} \frac{w_1}{dp_1\rho_{s2}} \frac{w_1}{dp_1\rho_{s3}}$

For homogeneous ternary mixture $\frac{1}{dp_{sm}} = \frac{w_1}{dp_1} + \frac{w_1}{dp_2} + \frac{w_1}{dp_3}$

Sau et.al. [15] have developed a correlation for maximum pressure drop in gas–solid tapered fluidized beds as,

$$Fr = 0.2741(Ar)^{0.2197}(\sin\alpha)^{0.6092}\left(\frac{\varepsilon_0}{\varphi_s}\right)^{-0.6105} \qquad (11)$$

$$\Delta P_{max} = 7.457\left(\frac{D_1}{D_0}\right)^{0.038}\left(\frac{dp}{D_0}\right)^{0.222}\left(\frac{H_s}{D_0}\right)^{0.642}\left(\frac{\rho_s}{\rho_f}\right)^{0.723} \qquad (12)$$

The dimensionless correlations for the critical fluidization velocity of mixture of irregular particles and the maximum pressure drop are given by Sau et al [16],

$$Re_c = 301.416(Ar)^{0.1272}\left(\frac{\rho_s-\rho_f}{\rho_f}\right)^{0.2507}\left(\frac{dp_{sm}}{D_0}\right)^{1.582}(\sin\alpha)^{0.1917} \qquad (13)$$

$$-\frac{(\Delta P_{max})}{\rho_{sm} g H_s} = 0.0204(Ar)^{0.02}\left(\frac{H_s}{D_0}\right)^{-0.2273}(\sin\alpha)^{0.2947} \qquad (14)$$

$$Fr = \left(\frac{U_{mf}}{\sqrt{g\,dp_{sm}}}\right) \qquad (15)$$

The original correlation for minimum fluidization velocity of Chiba et al. [17] as reported by Clarke et al.[18] for completely mixed bed of homogeneous binary mixture of particles is given by,

$$U_{vf} = U_{pk} - (U_{pk} - U_f)X_f \qquad (16)$$

The critical fluidization velocity as given by Sau et al [19] for binary mixture of regular particles in tapered fluidized bed as ,

$$Re_c = 16.364(Ar)^{0.1989}\left(\frac{\rho_s-\rho_f}{\rho_f}\right)^{0.2507}\left(\frac{dp_{sm}}{D_0}\right)^{1.1854}(\sin\alpha)^{0.0585} \qquad (17)$$

$$-\frac{(\Delta P_{max})}{\rho_{sm} g H_s} = 0.081(Ar)^{0.1492}\left(\frac{H_s}{D_0}\right)^{-0.2337}(\sin\alpha)^{0.2826} \qquad (18)$$

Based on the above correlations and equation, the computational models were successfully applied for predicting the hydrodynamic characteristics of fluidization viz. the minimum fluidization velocity as well as the dependence of the pressure drop across the bed for a conical gas-solid bed. The developed correlations for cylindrical conduits could not be used in case of fluidization in conical bed with a varying cross sectional area. In most of the investigations, it is observed that the experimental values of U_{mf} and ΔP_{mf} are generally more than that in case binary mixture of irregular particles. Moreover it has also been observed that the bed pressure drop and the minimum fluidization velocity are less in case of irregular particles than that for regular particles because of less bed voidage and reduced channeling and slugging in a conical bed.

2.2.Bed fluctuation ratio(r)

The fluctuation may be defined as the ratio of the highest and the lowest level of the top of the bed for any fluidizing gas mass velocity. Bed fluctuation and fluidization quality being inter-related, efforts have been made to correlate fluctuation ratio in terms of static and dynamic parameters of the system. Experimentally it is given by,

$$r = \frac{\textit{highest level the top of bed occupies}(h_2)}{\textit{lowest level the top of bed occupies}(h_1)}$$

A correlation for fluctuation ratio in conical vessels for regular particles has been developed by Biswal et al[20] using dimensional analysis approach based on four dimensionless groups neglecting the effect of density of gas and solid particles. The correlations reported by Biswal et al [20] for the fluctuation ratio of regular and irregular particles (21) are given by equations (19) and (20).

$$r = 3.168\left[\left(\frac{d_p}{D_0}\right)^{0.14}\left(\frac{d_c}{h_s}\right)^{-0.18}\left(\frac{h_s}{D_0}\right)^{-0.24}\left(\frac{G_f G_{mf}}{G_{mf}}\right)^{0.17}\right] \qquad (19)$$

$$r = 9.48\left[\left(\frac{d_p}{D_0}\right)^{0.27}\left(\frac{d_c}{h_s}\right)^{-0.82}\left(\frac{h_s}{D_0}\right)^{-0.15}\left(\frac{G_f G_{mf}}{G_{mf}}\right)^{0.32}\right] \qquad (20)$$

Singh and Roy et al [22] proposed the correlation for bed fluctuation ratio in case of binary homogeneous and heterogeneous binary mixtures of regular and irregular particles in conical beds are,

For homogeneous binary mixture of spherical particles

$$r = 9.8 \times 10^{-2}\left(\frac{G_f}{G_{mf}}\right)^{1.088}\left(\frac{D_0}{dp_{sm}}\right)^{-1.97}\left(\frac{h_s}{D_0}\right)^{-0.25}(\tan\alpha)^{-0.25} \qquad (21)$$

For heterogeneous binary mixture of spherical particles

$$r = 0.44\left(\frac{G_f}{G_{mf}}\right)^{0.58}\left(\frac{\rho_{sm}}{\rho_f}\right)^{-0.10}\left(\frac{D_p}{D_0}\right)^{-0.08}(\tan\alpha)^{-0.17} \qquad (22)$$

For both homogeneous and heterogeneous binaries mixture of non-spherical particles

$$r = 3.42\left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{-0.02}\left(\frac{D_0}{d_{p_{sm}}}\right)^{-0.180}\left(\frac{h_s}{D_0}\right)^{0.033}\left(\frac{\rho_{sm}}{\rho_f}\right)^{-0.023} \qquad (23)$$

Current review of literature deals on the development of mathematical model for fluctuation ratio for conical bed.By critically reviewing , It has been observed that very little work has been reported for binary mixtures in conical beds and practically no work has been carried out for ternary mixtures in conical bed. Thus there has been ample of opportunity for the study of the hydrodynamic characteristics of binary and ternary mixtures in a conical fluidized bed.

2.3 Bed expansion ratio (R)

Bed expansion ratio (*R*) is defined as the ratio of the average height of a fluidized bed to initial static bed height at a particular flow rate of the fluidizing medium above the minimum fluidizing velocity. It is an important parameter for fixing the height of a gas-solid fluidized bed when recommended for process applications

Expansion of gas-solid fluidized beds may in general result from the volume occupied by bubbles and from increase in voidage of the dense phase. It is given by the expression,

$$R = \frac{h_1 + h_2}{h_s} = \frac{h_{avg}}{h_s} \qquad (24)$$

The bed expansion ratio of a gas-solid fluidized bed depends on excess gas velocity (Gf −Gmf), particle size (dp) and initial static bed height (hs). Bed expansion is substantially greater in a two-dimensional bed than in a three-dimensional one. The use of a square fluidized bed has been advocated for some specific applications in view of its certain advantages as mentioned by Singh et.al. [22].The bed expansion reported by different investigators have different meanings because of the varied methods of measurement adopted. A number of investigations have been made with respect to the prediction of bed expansion in case of cylindrical conduits but limited information is available on the improvement of fludization quality interms of bedexpansion ratio for conical conduit. With respect to the bed expansion, the effects of column geometry have to deal extensively.

III. Bed Dynamics of Gas-Solid Fluidized System in Non-Cylindrical Conduits Other Than the Conical

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 as discussed earlier in this paper. Therefore some dynamic studies have, therefore, been made in hexagonal, square and semi-cylindrical beds for their potential application in gas-solid fluidization.

3.1 Minimum fluidization velocity and pressure drop

The overall pressure drop across a bed height, H is obtained by integrating the Ergun's equation

$$(-\Delta P) = \int_0^H [(AU + BU^2)]\,dh \qquad (25)$$

For the bed whose cross sectional area does not vary with the height equation (25) on integration gives,

$$-\Delta P = (AU + BU^2)H \qquad (26)$$

According to the proposed model as discussed in the above by Agarwal et al [11] the particles start fluidizing when F=G..Therefore, the minimum fluidization velocity, Umf can be found out by equating F =G

$$A_1 U_{mf}^2 + B_1 U_{mf} = C_1$$

$$U_{mf} = \frac{-B_1 \pm \sqrt{B_1^2 + 4A_1C_1}}{2A_1} \qquad (27)$$

The pressure drop for minimum fluidization is calculated as,

$$-\Delta P_{mf} = (AU_{mf} + BU_{mf}^2)H \qquad (28)$$

Values for minimum fluidization velocity and pressure drop are calculated using the above equations. 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 For hexagonal, semi-cylindrical, square and cylindrical bed, cross-sectional area remains constant with the height of bed.

3.2 Correlations for bed fluctuation ratio

For any bed the general form for the bed fluctuation ratio is given by Singh et al [22][23], The correlations has been developed with the help of relevant dimensionless groups involving interacting parameters like bed height, equivalent diameter of the column, particle diameter, density of the particle, density of the fluidizing medium and fluid mass velocity

$$r = k\left[\left(\frac{d_p}{D_c}\right)^a \left\{\left(\frac{d_c}{h_s}\right)\left(\frac{\rho_f}{\rho_s}\right)\right\}^b \left(\frac{G_f - G_{mf}}{G_{mf}}\right)^c\right]^n \qquad (29)$$

Where, *k* is the coefficient and *a*, *b*, *c* and *n* are the exponents. The effects of the individual groups on fluctuation ratio, *r* have been separately evaluated for the different conduits and values of *a*, *b*, and *c* have been obtained from the slope of the plots .

Singh[23] studied the effect of various system parameters on fluctuation ratio in case of unpromoted non-colummar beds, viz square, hexagonal and semi cylindrical ones and proposed the following correlations

For Cylindrical Bed

$$r = 1.95\left[\left(\frac{d_p}{D_c}\right)^{0.04}\left\{\left(\frac{d_c}{h_s}\right)\times\left(\frac{\rho_f}{\rho_s}\right)\right\}^{0.04}\left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.05}\right] \qquad (30)$$

For semicylindrical bed

$$r = 2.323\left[\left(\frac{d_p}{D_c}\right)^{0.08}\left\{\left(\frac{d_c}{h_s}\right)\times\left(\frac{\rho_f}{\rho_s}\right)\right\}^{0.04}\left(\frac{G_f - G_{mf}}{G_{mf}}\right)^{0.07}\right] \qquad (31)$$

For hexagonal bed

$$r = 2.3\left[\left(\frac{d_p}{D_c}\right)^{0.08}\left\{\left(\frac{d_c}{h_s}\right)\left(\frac{\rho_f}{\rho_s}\right)\right\}^{0.05}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.08}\right] \quad (32)$$

For square bed

$$r = 2.55\left[\left(\frac{d_p}{D_c}\right)^{0.09}\left\{\left(\frac{d_c}{h_s}\right)\times\left(\frac{\rho_f}{\rho_s}\right)\right\}^{0.04}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.05}\right] \quad (33)$$

3.3 Correlations for bed expansion ratio

Singh et al. [22,23] studied the effect of various system parameters on bed expansion ratio in case of cylindrical and non-cylindrical beds and proposed the following correlations.

$$R = K\left(\frac{d_p}{d_c}\right)^{a}\left(\frac{d_c}{h_s}\right)^{b}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{c} \quad (34)$$

The values of k, a, b, c for various types of bed for bed expansion ratio is given by Singh et al [23]

For Cylindrical Bed

$$R = 2.55\left(\frac{d_p}{d_c}\right)^{0.11}\left(\frac{d_c}{h_s}\right)^{0.21}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.18} \quad (35)$$

For semi-cylindrical

$$R = 5.46\left(\frac{d_p}{d_c}\right)^{0.28}\left(\frac{d_c}{h_s}\right)^{0.02}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.21} \quad (36)$$

For hexagonal bed

$$R = 2.422\left(\frac{d_p}{d_c}\right)^{0.12}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.25} \quad (37)$$

For square bed

$$R = 6.09\left(\frac{d_p}{d_c}\right)^{0.24}\left(\frac{G_f-G_{mf}}{G_{mf}}\right)^{0.27} \quad (38)$$

3.4 Prediction of bubbling velocity and fluidization index in non-cylindrical bed for gas–solid fluidization.

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, Umb / Umf, is known as the fluidization index, which gives a measure of the degree to which the bed can be expanded uniformly .

Wilhelm et al. [24] proposed the use of Froude number (Fr_{mf}) as a criterion for bubbling or aggregative fluidization. A value of $Fr_{mf} > 1.0$ induces a bubbling behavior in the bed, where

$$Fr_{mf} = \frac{U_{mf}^2}{d_p g} \quad (39)$$

A correlation for minimum bubbling velocity was suggested by Geldart [25] as

$$U_{mb} = K_{mb} d_s \quad (40)$$

Where $d_s = \frac{1}{\Sigma_i\left(\frac{x_i}{d_{si}}\right)}$

and K_{mb} is Constant whose value is 100 in C. G. S. unit.

Abrahamsen and Geldart [26] correlated the values of minimum bubbling velocity with gas and particle properties as follows:

$$U_{mb} = 2.07\left[\exp(0.716F)\frac{x_p \rho_g^{0.06}}{\mu_f^{0.347}}\right] \quad (41)$$

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

$$U_{mf} = \frac{(\rho_p-\rho_f)^{0.934} g^{0.934} x_p^{1.8}}{(1100\mu^{0.87}\rho_g^{0.066})} \quad (42)$$

As Fluidization Index is the ratio of minimum bubbling velocity to minimum fluidization velocity, dividing the Abrahamsen equation by the Baeyensequation[27], the correlation obtained is

$$\frac{U_{mb}}{U_{mf}} = \frac{2300\rho_g^{0.126}\mu_g^{0.523}\exp(0.716F)}{d_p^{0.8} g^{0.934}(\rho_p-\rho_g)^{0.934}} \quad (43)$$

This means that higher the ratio, the bed can hold more gas between the minimum fluidization and bubbling point. The correlations for minimum bubbling velocity developed by Singh[28]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. The correlations obtained for different conduits are as follows

Cylindrical bed;

$$U_{mb} = 0.5231\left(\frac{d_p}{D_c}\right)^{1.12}\left(\frac{D_c}{h_s}\right)^{-0.0254}\left(\frac{\rho_p}{\rho_f}\right)^{0.74} \quad (44)$$

Semi-cylindrical bed;

$$U_{mb} = 0.168\left(\frac{d_p}{D_c}\right)^{0.994}\left(\frac{D_c}{h_s}\right)^{-0.1849}\left(\frac{\rho_p}{\rho_f}\right)^{0.80} \quad (45)$$

Hexagonal bed;

$$U_{mb} = 0.15\left(\frac{d_p}{D_c}\right)^{0.5733}\left(\frac{D_c}{h_s}\right)^{-0.0887}\left(\frac{\rho_p}{\rho_f}\right)^{0.5284} \quad (46)$$

Square bed; $U_{mb} = 0.168\left(\frac{d_p}{D_c}\right)^{0.27}\left(\frac{D_c}{h_s}\right)^{-0.0132}\left(\frac{\rho_p}{\rho_f}\right)^{0.2825}$ (47)

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 by Singh [28] under identical operating conditions. 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.

3.5 Prediction of slugging velocity and bubbling bed index in non-cylindrical bed for gas–solid fluidization

The gas-solid fluidization is characterized by the formation of bubble. The size of the bubble increases and sometimes even its diameter may become equal to that of the column. When the bubble diameter approaches the column diameter, it is termed as slugging. The superficial gas velocity at which slug formation starts is known as minimum slugging velocity. An aggregative fluidized bed in a column of small diameter operated at sufficiently high gas velocity will show continuous slug flow. Slugging affects adversely the fluidization quality. Slugging increases the problem of entrainment and lowers the performance potential of the bed. Slugging is especially serious in long narrow fluidized beds.

Identical to fluidization index,it is proposed to define 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.

Stewart et al. [29] have given a correlation for minimum slugging velocity as

$$U_{ms} = U_{mf} + 0.07\sqrt{gD_c} \qquad (48)$$

The bed must be sufficiently deep for coalescing bubbles to attain the size of a slug. Baeyens et al. [30] concluded that the above condition is applicable only if $h_{mf} > 1.3D_c^{\,0.175}$ in SI units, otherwise the minimum slugging condition is given by

$$U_{ms} = U_{mf} + 0.07\sqrt{gD_c} + 0.16\left(1.3D_c^{\,0.175} - h_{mf}\right)^2 \qquad (49)$$

The correlation for minimum slugging velocity have been developed by Singh et al.[31] with the help relevant dimensionless group involving interacting parameter like particle diameter, Equivalent diameter of the column, the packed bed height, the density of the particle and of the fluidizing medium. As by the author the correlations obtained for different conduits are as follows:

Cylindrical bed,

$$U_{ms} = \left[0.136\left(\frac{d_p}{D_c}\right)^{0.6324}\left(\frac{D_c}{h_s}\right)^{0.044}\left(\frac{\rho_p}{\rho_f}\right)^{0.6559}\right] \qquad (50)$$

For semi-cylindrical bed,

$$U_{ms} = \left[0.269\left(\frac{d_p}{D_c}\right)^{1.0245}\left(\frac{D_c}{h_s}\right)^{0.2012}\left(\frac{\rho_p}{\rho_f}\right)^{0.8234}\right] \qquad (51)$$

For hexagonal bed,

$$U_{ms} = \left[0.290\left(\frac{d_p}{D_c}\right)^{0.6478}\left(\frac{D_c}{h_s}\right)^{0.215}\left(\frac{\rho_p}{\rho_f}\right)^{0.5292}\right] \qquad (52)$$

For square bed,

$$U_{ms} = \left[0.863\left(\frac{d_p}{D_c}\right)^{1.25}\left(\frac{D_c}{h_s}\right)^{0.057}\left(\frac{\rho_p}{\rho_f}\right)^{0.78}\right] \qquad (53)$$

Minimum slugging velocity and minimum bubbling velocity are maximum in case of semi cylindrical bed and minimum in case of other noncylindrical conduits like square by Singh [31] and Hexagonal by Padhi et al.[32] under identical operating conditions based on the experimental data. It is further observed that the bubbling bed index is maximum in case of a square bed but the range of bubbling fluidization is maximum in case of a semi cylindrical bed for identical operating conditions. Since the range of bubbling fluidization is maximum again in case of semi cylindrical bed therefore a better substitute for conventional one when bubbling fluidization is desired to meet the process requirements.

IV. CONCLUSION

By reviewing the investigation of different authors it is concluded that the developed correlations for cylindrical conduits could not be used in case of fluidization in noncylindrical conduits.. Moreover it is also observed that that bed pressure drop and minimum fluidization velocity in case of noncylindrical conduits is less in case of irregular particles than that in regular particles because of less bed voidage and decreased channeling and slugging. However there is some deviation seen when compared with the equations with the fluidization of single particle and found that some of the parameters are less for noncylindrical conduits than that for conventional cylindrical conduits.

Gas-solid fluidization as contacting techniques in noncylindrical conduits has tremendous potential for industrial use. In spite of a lot of research activities carried out for the understanding of fluidization technology in the past few decades, several aspects relating to the effect of distributor, irregular and regular shape particles as bed material, liquid viscosity and surface tension, scaling up of a developed system for industrial application are not fully investigated. Further the various alternative proposition for smooth bed hydrodynamic with improvement in fluidization quality for cylindrical vice versa noncylindrical conduits have not been extensively explored. While some investigation to understand the various bed hydrodynamics of cylindrical and tapered bed have been conducted , the studies relating to alternative column configuration viz. square,hexagonal and semicylindrical extremely limited. In addition, CFD analysis of such bed is absolutely necessary to have an explicit understanding of bed dynamics in view of their potential process application in the year ahead.

Gas-solid system continues to raise the fundamental question, difficult practical dilemmas and opportunities for innovation. Much work has been carried out to understand the hydrodynamics characteristics in cylindrical conduits but limited work has been done in noncylindrical conduits and hence it need more attention However, it is premature to draw more conclusions with respect to dynamic behavior of non-circular beds at this stage and investigations are on for auxiliary enhancement.

WHERE

A =Erguns constant

A_n =Cross-sectional area of non-column bed m^2

Ar =Archimedes number $= \frac{g\,dp_{sm}^{3}\rho_{sm}^{2}}{\mu_f^{2}}$

B =Erguns constant

dp_{sm} =Average particle diameter of ternery mixture

dh =Differential bed height , m

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

D_o= Bottom diameter of tapered bed , m

D_c =Equivalant column diameter , m

d_p =Particle diameter , m

D_t =Diameter of column , m

Fr =Frouds number

F=Force exerted by the fluidizing medium on particle in the bed , N

G=Effective weight of the material in the medium , N

G_{mf} =Superficial fluid mass velocity at minimum fluidiation , Kg/hrm^2

G_f =Flow rate of fluid at fluidization condition , $\frac{m^3}{hr}$

G_{mf} =Flow rate of fluid at minimum fluidization condition , $\frac{m^3}{hr}$

g =Gravitational acceleration , 9.81 m/s^2

g_c =Gravitational constant

H=height of the bed , m

h_{max} =Maximum height of fluidized bed , L
h_{min} =minimum height fluidization bed , L
h_s =initial static bed height , L
H_f =Height of expanded bed , m
H_s =Height of initial static bed , m
K = Constant
m = Correlation factor , dimensionless
r = Bed Fluctation ratio (dimensionless)
R = Bed expantion ratio (dimensionless)
Re_c =Critical Reynolds number
U = Superficial fluid velocity ,$\frac{m}{sec}$
U_0 =Incipient fluidizing velocity ,$\frac{m}{sec}$
U_{vf} =Superficial velocity of gas where both components in a binary mixture are fluidized, m/sec
U_{pc} = Minimum fluidization velocity of packed components in single component fluidized beds, m/sec
U_f =Minimum fluidization velocity of fluid in single component fluidized beds ,$\frac{m}{sec}$
U_{mf} =Superfecial minimum fluidization velocity ,$\frac{m}{s}$
U_{mb} = Minimum bubbling velocity ,ms^{-1}
U_{ms} = Minimum slugging velocity , , ms^{-1}
V_{mf} = Minimum fluidization velocity
V_0 =fixed bed the superficial velocity
V_t = Particle terminal velocity
X_f: Volume fraction of fines in the mixture

Greek word
ΔP_{mf} = Pressure drop at minimum fluidization velocity ,$\frac{N}{m^2}$
ΔP = Pressure drop through particle bed , , $\frac{N}{m^2}$
$\emptyset_s$ = Sphericity of particle
ρ_p = Density of particle ,$\frac{kg}{m^3}$
ρ_f = Density of air ,$\frac{kg}{m^3}$
ε = Porosity of the fluidizing bed
ϵ_0 = Voidage at packed condition
μ = Viscosity of gas phase
ρ_f = Density of fluid phase ,$\frac{kg}{m^3}$
ρ_s = Density of a particle ,$\frac{kg}{m^3}$
$\emptyset_s$ =Sphericity of a particle

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AUTHORS

First Author: Mr.RadhaKrushnaPadhi, MTech, Chemical Engineering Department, G.I.E.T, Gunupur, Rayagada, Odisha, 765022 .Mail ID: rkpadhi_2006@rediffmail.com

Second Author: Dr. G.K.Roy, PhD,Chemical Engineering Department, G.I.E.T, Gunupur, Rayagada, Odisha, 765022. Mail ID: gkroy1234@ gmail.com

Third Author: Dr.Yashobanta Mohanty, PhD, Chemical Engineering Department, G.I.E.T, Gunupur, Rayagada, Odisha, 765022. Mail ID:ykmohanty@rediffmail.com