

Cluster Diameter Determination of Gas-solid Dispersed Particles in a Fluidized Bed Reactor

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Abstract: Clustering is a common hydrodynamic characteristic observed among suspended gas-solid particles in a fast fluidized bed (FFB) regime of a circulating fluidized bed (CFB) system. These suspended particles are unstable in nature and the cluster size is solely dependent on density of the particles. In this paper clustering behavior has been studied with Geldart group B particles like coal and iron ore in a circulating fluidized bed of diameter 0.1016 m and height 5.62 m. Clustering phenomena has been investigated by considering the effects of acceleration and friction into homogeneous gas-solid cluster suspension model. The cluster size when calculated from the homogeneous cluster suspension model seems to match the measured data when the particle concentration in the suspension lies mostly in the dilute phase. It has been found that the acceleration effect has some effect on the cluster diameter; however, the solid friction does not have any significant role. The result of the present study gives a thorough understanding of the FFB. The complex relation between the continuous and disperse phase onto the suspended particles can be modeled using computational fluid dynamics (CFD) module which describes the hydrodynamics i.e., momentum, heat and mass transfer and the reaction module which describes the homogeneous gas phase reactions. These modules give rise to a multi-scale model for a particular clustering process which can be solved using COMSOL Multiphysics and MATLAB software.

Keywords: Fast fluidized bed, circulating fluidized bed, cluster diameter, hydrodynamics, Geldart particles, computational fluid dynamics (CFD).

1. Introduction

Fast-fluidized bed (FFB) has wide range of applications in petrochemical, mineral and metallurgical industries. This particular regime of fluidized bed has gathered importance where

high solid throughputs, transfer rates and thermal uniformity within a reactor are desired. The common hydrodynamic features of a fast fluidized bed in a circulating fluidization system are usually large gas-solid slip velocity, clustering behavior of gas-solid suspension and solid down flow over the wall. Earlier it was believed that in a vertical fluid-particle system, once the superficial fluid velocity (U_0) exceeded the terminal velocity (U_t) of the individual particles, the particles would be carried out of the system. Later, it has been shown that some entrainment do occur at a relatively dense suspension (volumetric solids concentration of the order of 0.1–0.3) with the aid of a cyclone at the top of the system that would capture the solids escaping from the top and return them to the lower part of the system.

The phenomenon of large slip velocity in a FFB has been attributed by Rhodes and Geldart due to the radial nonuniformity in the flowing gas-solid systems which lead to the core-annulus flow model whereas others analyzed the same with the concept of cluster formation. According to this second group of thought the vertical flow of a homogeneous gas-solid suspension are unstable in nature. It has also been indicated that the growth distance of disturbances decreases with increasing suspension density. Therefore, clustering seems to be a characteristic of the suspension in a CFB system.

Assuming the clusters of uniform shape (spherical), size, voidage and velocity Yerushalmi et al. tried to calculate the average drag coefficient of the same. Pritchett et al. tried to fit a model for the clustering suspension using numerical simulation. Yerushalmi et al. stated that the large slip velocities were due to the solid aggregation in the fast bed. According to them, most particles do not remain single but tend to gather in 'clusters', 'strands' or 'packets' which have effective free fall velocities greater than those of individual particles. When the clusters are sufficiently large to be sustained by the rising gas, they will disintegrate and rain back into the

system, thereby resulting in high degree of solid back mixing in the fast bed. Later, Yerushalmi *et al* proposed a simplified homogeneous model for the computation of the cluster diameter with the assumption that the bed was a uniformly dispersed suspension of spherical clusters of diameter, D_c . They computed the cluster diameter, D_c using the experimental data for air-FCC catalyst and air-Dicalite 2000 systems and plotted it against the solid volumetric concentration $(1 - \varepsilon)$.

In the present investigation, an attempt has been made to modify Yerushalmi's model with wide-range data generated in an FFB on drag coefficient and also taking into consideration the effect of acceleration into the force balance equation.

2. Literature review and use of COMSOL Multiphysics

Various attempts have been made to study the behavior of lean gas–solid suspensions in terms of clustering tendency of the particles. A ‘cluster model’ was developed with the assumption that the cluster was like an impermeable particle with cluster porosity of 0.5. With the help of the model, he explained different hydrodynamic phenomena such as choking, elutriation rate constant and pressure drop gradient for dilute phase transport. Li and Kwauk analyzed the phenomena of energy transport and regime transition based on the method of energy minimization in multiscale modeling of two-phase flow, which consisted of aggregated fluid cavities (gas bubbles) and of aggregated solid particles (clusters). Li *et al.* studied the cluster-size distribution and motion behavior in a FFB by optical fiber probes and a computer data acquisition system. The results revealed that though the local time-averaged cluster lengths remain almost constant radially both in the upper dilute and lower dense region of the cluster length. Noymer and Glicksman observed the motion of clusters of particles at the wall of a CFB by using a thermal-imaging technique for flow visualization. Xu and Kato have tried to establish a simple correlation for calculating the equivalent diameter of clusters in gas–solid flow. They reported that the hydrodynamic equivalent cluster diameter is close to that of the cross-sectional size, but much less than the vertical cluster height experimentally measured. Harris *et*

al. developed correlations to predict the cluster characteristics in large-scale units. Mostoufi and Chaouki analyzed the existence of clusters in dense fluidized bed by the time-position data of a tracer. Nova *et al* studied the particle clustering phenomena using a novel gas–solid optic probe equipped with graded refractive index lens. Wang *et al* acquired the image sequences of the formation, movement and decomposition of the glass beads cluster on a cold CFB.

Thus various models and techniques which have been proposed to study cluster-related and allied phenomena so far can probably be done using the finite element program, Comsol Multiphysics. Moreover, the problem can best be handled numerically, say for a cluster moving in an infinite medium, the dimension of the domain should be large enough to take care of its boundary conditions or so.

3. Clustering nature of gas-solid suspension in fast fluidized bed

It has been observed that FFB consists of uniformly dispersed suspension of spherical particles. The effective voidage $\bar{\varepsilon}$ or the volumetric concentration of clusters $(1 - \bar{\varepsilon})$ has been related to the actual voidage, ε by the following relation.

$$(1 - \varepsilon) = (1 - \bar{\varepsilon})(1 - \varepsilon_{mf}) \quad \dots (1)$$

Neglecting the acceleration and frictional effects, one can equate the pressure drop with gravity and the resulted balance is presented in Eqn (2).

$$\Delta P = (1 - \bar{\varepsilon})\rho_p(1 - \varepsilon_{mf})g \Delta L \quad \dots (2)$$

Writing the pressure drop owing to the drag exerted by the gas upon the clusters,

$$\Delta P = \frac{1}{2} \frac{N}{A_t} \rho_g \left(\frac{U_g}{\bar{\varepsilon}} - U_s \right)^2 C_{ds} f(\bar{\varepsilon}) \frac{\Pi}{4} D_c^2 \quad \dots (3)$$

Noting that, $N \Pi D_c^3 / 6 = (1 - \bar{\varepsilon}) A_t L$,

one can equate the two Eqs (1) and (2) and simplify to get

$$D_c = \frac{3 \rho_g [(U_g / \bar{\varepsilon}) - U_s]^2 C_{ds} \bar{\varepsilon}^{-2n}}{4 \rho_p (1 - \varepsilon_{mf}) g} \quad \dots (4)$$

where N is the number of clusters in a bed section of thickness ΔL and A_t is the cross-sectional area of the bed. U_s is the mean solid

velocity of the clusters. C_{ds} is the familiar drag coefficient for a spherical particle moving at uniform relative velocity. In this case, it is the drag coefficient for a single cluster and functions of Reynolds number based upon cluster diameter and slip velocity, i.e.

$$C_{ds} = C'_{ds} \left[\frac{\rho_g D_c \left(\frac{U_g}{\varepsilon} - U_s \right)}{\mu_g} \right] \quad \dots(5)$$

where μ_g is the viscosity of the gas.

The function $f(\varepsilon)$ has been introduced to take into account the hindered settling effects resulting from the proximity of particles clusters. The effective voidage has been chosen based on Richardson-Zaki equation, viz.

$$f(\varepsilon) = \varepsilon^{-2n} \quad \dots(6)$$

where n is the Richardson-Zaki index and is a function of the Reynolds number based upon the terminal settling velocity, $U_{t,c}$ of the cluster. Thus,

$$n = \begin{cases} 4.65; & Re_{t,c} < 0.2 \\ 4.4 Re_{t,c}^{-0.03}; & 0.2 < Re_{t,c} < 1.0 \\ 4.4 Re_{t,c}^{-0.1}; & 1.0 < Re_{t,c} < 500 \\ 2.40; & Re_{t,c} > 500 \end{cases} \quad \dots(7)$$

where $Re_{t,c} = \rho_g DC U_{t,c} / \mu_g$

Equation (4) can be used for computing the cluster diameter, DC , provided the gas velocity, solid rate and fluidized-bed density [or $(1 - \varepsilon)$] are known. The solids, while being transported upward by the gas, form aggregates of clusters and these clusters get dispersed into the gas. The resulting pressure drop occurs due to the interaction between these clusters and gas on one hand and their interaction with the wall on the other hand. On the basis of the measured pressure drop data and calculated voidages, the cluster diameter has been computed using Eqn (4) and plotted in Fig. 1 for coal and iron particles. Figure 2 shows an identical plot for iron ore, FCC catalyst and alumina based on the data reported by Kwauk *et al.* It is generally found that up to a value of $(1 - \varepsilon)$ equal to 0.1, the

cluster diameter predicted by all the workers [present author, Kwauk *et al.*] agree fairly well. Beyond the value of $(1 - \varepsilon)$ given by 0.1, the

computed cluster diameter from Kwauk's data are found to be appreciably higher.

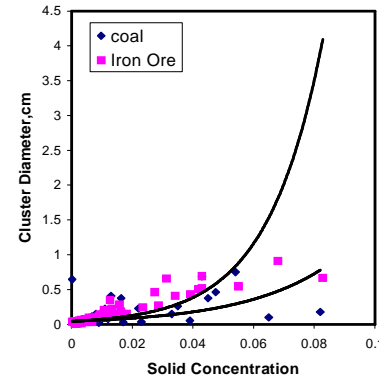


Figure 1: Effect of solid concentration on $(1-\varepsilon)$ cluster diameter (D_c) for different size of particles.

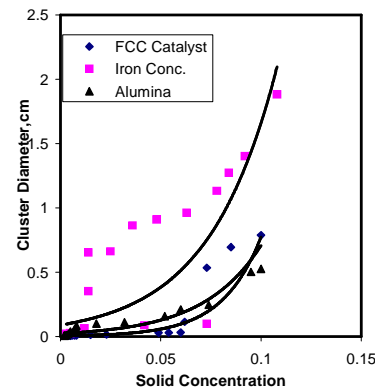


Figure 2: Effect of solid concentration on $(1-\varepsilon)$ cluster diameter (D_c) for different size of particles {Li Kwauk data, mean line drawn}.

4. Modification of Yerushalmi's Theory:

4.1 Effect of acceleration and friction

At the lower end of the riser, the accelerating effect contributes to as much as 9-40% of the total pressure drop, while in the top dilute zone the frictional losses due to solid may be about 5% of the total pressure drop. Accordingly, an attempt has been made to modify Eq.4 for cluster diameter by including the effects of acceleration and friction. The total pressure drop along a length (ΔL) of the riser due to gravity, friction and acceleration is

$$\Delta P = \rho_p g(1-\varepsilon)\Delta L + \left(\frac{4}{D_t}\right)(\tau_F + \tau_p)\Delta L + \rho_p(1-\varepsilon)U_s^2 \quad \dots (8)$$

The wall gas shear stress, τ_F is given by

$$\tau_F = (1/2) f_g \rho_g U_g^2 \quad \dots (9)$$

where f_g is evaluated from Colebrook equation for smooth pipe:

$$\left. \begin{array}{l} f_g = 16 / N_{Re} ; \quad \text{for } N_{Re} < 500 \\ \text{or,} \\ \frac{1}{\sqrt{f_g}} = -4 \log_{10} \left(\frac{1.256}{N_{Re} \sqrt{f_g}} \right) \end{array} \right\} \dots (10)$$

Where N_{Re} is the Reynolds number based on pipe diameter,

$$N_{Re} = D_t U_g \rho_g / \mu_g \quad \dots (11)$$

The wall shear stress, τ_p can be expressed as

$$\tau_p = (1/2) f_s(1-\varepsilon) \rho_p |U_s| |U_s| \quad \dots (12)$$

$$\text{and } f_s = a |U_s|^b \quad \dots (13)$$

Here, Stromberg's relationship is used for which $a=0.003$ and $b=0$.

Substituting the values of τ_F and τ_p from Eqs. (9) and (12) respectively and using Eqs.(2) and (8), one gets after rearrangement:

$$D_c = \frac{3}{4} \frac{\rho_g(1-\bar{\varepsilon})(U_g/\bar{\varepsilon}-U_s)^2 C_{ds} \bar{\varepsilon}^{-2n}}{\rho_p(1-\bar{\varepsilon})(1-\varepsilon_{mf})g + \left(\frac{4}{D_t}\right)(\tau_F + \tau_p) + [\rho_p(1-\bar{\varepsilon})(1-\varepsilon_{mf})U_s^2/L]} \quad \dots (14)$$

4.2 Comparison of calculated cluster diameter based on homogeneous suspension model with the measured data of Horio

Horio et al. reported cluster diameter with 1.5 times cluster length pierced by the probe from the analysis of the raw signals of reflected light intensity from the optical fiber probe. The cluster size was found to be small (5-40 mm when the bed solid concentration ranges from 0.05 to 0.2). The cluster size was almost constant in the core irrespective of the height, while it was large (more than 10mm) and grew in size as it moved downwards in the annulus.

As already stated, the cluster diameter was recalculated by incorporating the effects of solid friction with wall and acceleration of the particles at the bottom of the riser. It is observed that while the accelerating force has some effect on the cluster diameter, the solid friction factor does not have any significant role.

Binary and multicomponent solid mixtures flowing in a vertical pipe, the solid wall friction effect is smaller in comparison with that for the flow of single-sized particles. These may be due to collisions between the particles of different sizes that accelerate the heavier particle but decelerate the lighter one. On the basis of the assumption that the cluster-phase particle fraction ($\varepsilon_{p,cl}$) is of the order of 0.2, the cluster diameters have been recalculated for the present data and those reported in the literature. Figure 3 gives such a plot, which shows that the recalculated cluster diameter is 1 order higher than the earlier case (when the cluster-phase voidage was assumed to be ε_{mf}). The cluster diameter calculated from homogeneous cluster suspension model appears to tally with the measured data when the particle concentration in the suspension lies between 0.01 and 0.10, i.e. the gas-solid suspension remains in the dilute phase.

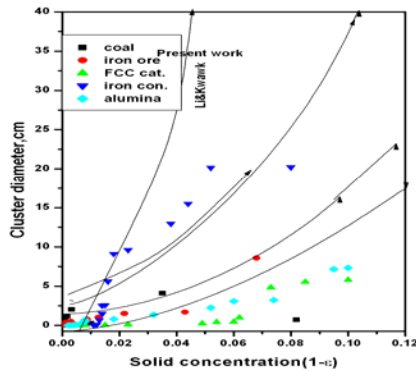


Figure 3: Recalculated cluster diameter with $\varepsilon_{p,cl} = 0.2$.

5. Conclusions

In the present investigation, the cluster behavior of fast fluidization have been studied thoroughly by using Geldart's group A as well as group B type of particles in a typical 'N' type CFB system. The homogeneous model proposed by Yerushalmi *et al* to predict the cluster diameter has been modified by including the effects of acceleration and frictional forces. It is observed that the acceleration has some effect on the cluster diameter. However, it is interesting to note that the solid friction does not have any significant effect on cluster diameter. The cluster diameter calculated from homogeneous cluster suspension model appears to tally with the measured data when the particle concentration in the suspension lies below 0.02, i.e. the gas-solid suspension remains in the dilute phase.

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7. Nomenclature

- ΔP pressure drop, kg/m²
 g acceleration due to gravity, m/s².
 L height of the column, m.
 N number of clusters per unit volume
 A_c cross-sectional area of the riser column, m².
 U_0 superficial fluid velocity, m/s.
 U_g gas velocity, m/s.
 U_s solid velocity, m/s.
 U_t terminal settling velocity of individual particles, m/s.
 G_s solid circulation rate, kg/m²s.
 C_{ds} drag coefficient.
 D_c cluster diameter, m.
 D_t tube diameter, m.
 n Richardson-Zaki index.
 $Re_{t,c}$ cluster Reynolds number.

Re_t Reynolds number at terminal settling.
 $U_{t,cl}$ terminal settling velocity of clusters,
m/s.
 f_g gas friction factor
 f_s solid friction factor
 N_{Re} Reynolds number based on pipe
diameter.

Greek Letters

ε voidage.
 $\bar{\varepsilon}$ average voidage.
 ε_{mf} minimum fluidization velocity.
 $\varepsilon_{P,cl}$ cluster-phase particle fraction.
 ρ_p solid density, kg/m³.
 ρ_g gas density, kg/m³.
 μ_g gas viscosity, kg/m²s.
 τ_g shear stress due to gas, kg/m².
 τ_s shear stress due to solid, kg/m².