Fluid Dynamics Behavior Analysis of Alumina Particles in Fluidized Bed

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Abstract. *Fluidized beds are widely used for particle coating and granulation in the pharmaceutical and chemical industries; among typical examples of fluidized beds some applications could be outlined: the syntheses and catalytic reactions, the catalytic regeneration, the combustion and gasification of coal, coating, drying, etc. Theses applications are due to its excellent mixing properties, high heat and mass transfer rates, and ease of scale-up. This paper presents a experimental and computational study of the hydrodynamics of a gas-solid fluidized bed. Experimental data were obtained by using a fluidized bed~~,~~ containing alumina particles of 110 µm in diameter and 3,500 Kg/m^3 in density. Four simulated data were obtained through the commercial CFD (Computational fluid dynamics), where the porosity profile of fluidized bed was calculated by using an Eulerian multiphase model. Results of numerical and experimental simulation are presented and discussed. The model is validated through comparison to experiment.*

Keywords: *Fluidized Bed, , Alumina.*

1. INTRODUCTION

Processes involving both gas and solid phase are very common in industry. Particle coating and granulation in the pharmaceutical and chemical industries are typical examples of fluidized beds Heat or mass transfer and chemical reaction in such processes depend on the interaction of the two phases within the bed. The solids are mostly contained in a cylindrical or rectangular vessel and gas is introduced through a distributor. Fig. 1 depicts the different contacting regimes as defined by the solid–gas characteristics, the bed geometry and the operating gas flow rate (Smolders and Baeyens, 2001).

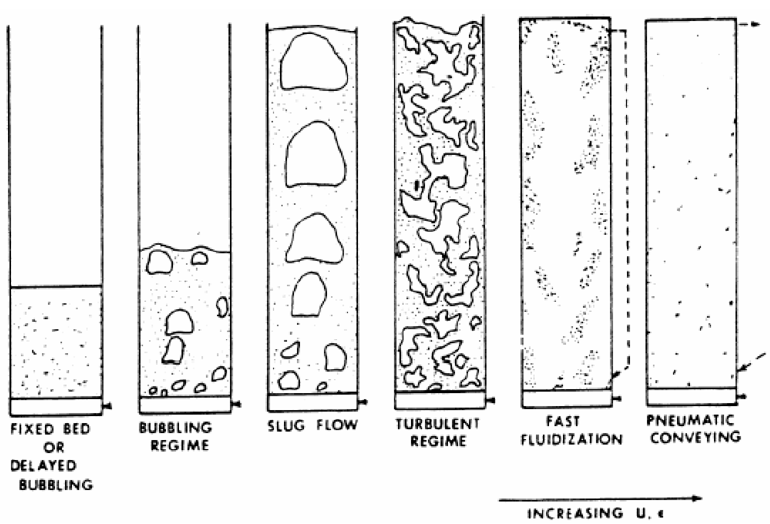

Figure 1. Different contacting regimes.

The packed bed and fluidized bed flow regimes are treated in numerous books and papers Regimes considered in this paper occur at gas velocities approaching the terminal velocities of the particles and refer to mostly Geldart's A and B-class particles only. In this paper these regimes are experimentally investigated, and numerical results are present by using a commercial CFD code..

The numerical simulation technique and CFD studies have become popular in the field of gas-solid two-phase flow. In this research work, the Eulerian multiphase model has been used to simulate the fluid dynamics behavior of the conventional fluidized bed (Duarte et al., 2004).

2. DESCRIPTION OF THE MODEL

The Eulerian multiphase model allows for modeling of multiple separate, yet interacting phases. The phases considered are gas and solid. The Eulerian approach is used for each phase, taking into account all possible intraphase and interphase combinations. The conservation equations can be derived by ensemble averaging the local instantaneous balances for each phase.

The volume fraction of fluid phase α_f is calculated from the continuity equation:

$$\frac{\partial}{\partial t}\left(\alpha_f\right) + \nabla.(\alpha_f \vec{v}_f) = 0 \tag{1}$$

where $\vec{v}_f$ is the velocity of fluid phase f. The volume V_f of the phase f, is defined by:

$$V_f = \int_V \alpha_f \, dV \tag{2}$$

The solution of Equation (1) for the fluid phase, along with the condition that the volume fractions add up to one, allows for calculating of the solids-phase volume fraction α_s . This treatment is common for fluid-fluid and granular flows.

The conservation of the fluid-solids momentum equation is given by:

$$\frac{\partial}{\partial t}\left(\alpha_s \rho_s \vec{v}_s\right) + \nabla.(\alpha_s \rho_s \vec{v}_s \vec{v}_s) = -\alpha_s \nabla p + \\ -\nabla p_s + \nabla.\overline{\overline{\tau}}_s + \alpha_s \rho_s \vec{g} + K_{fs}(\vec{v}_f - \vec{v}_s) \tag{3}$$

Where, the momentum exchange coefficient (K_{sf}) between solid phase *s* and fluid phase *f* can be written in the following general form:

$$K_{sf} = \frac{\alpha_s \rho_s f_d}{\tau_s} \tag{4}$$

where f_d is the drag and τ_s, the "particulate relaxation time" given by

$$\tau_s = \frac{\rho_s d_s^2}{18\mu_f} \tag{5}$$

where d_s is the diameter of particles in the solid phase (*s*) and μ_f is the fluid viscosity.

The definition of f_d includes a drag function (C_D) that is based on the relative Reynolds number (R_{es}), which is given by Equation (8).

$$K_{sf} = \frac{\rho d_s \left(\vec{v}_s - \vec{v}_f\right)}{\mu_f} \tag{6}$$

In this research work the Gidaspow et al. (1992) model was used, which is a combination of the Wen and Yu (1966) and the Ergun (1952) equations.

The solids-fluid exchange coefficient K_{sf} is of the following form
for $\alpha_f > 0.8$:

$$K_{sf} = \frac{3}{4} C_D \frac{\alpha_s \alpha_f \rho_f \left| \vec{v}_s - \vec{v}_f \right|}{d_s} \alpha_f^{-2.65} \tag{7}$$

where:

$$C_D = \frac{24}{\alpha_f \, \mathrm{Re}_s} \left[1 + 0.15 \left(\alpha_f \, \mathrm{Re}_s \right)^{0.687} \right] \tag{8}$$

for $\alpha_f \leq 0.8$:

$$K_{sf} = 150 \frac{\alpha_s (1 - \alpha_f) \mu_f}{\alpha_f d_s^2} + 1.75 \frac{\rho_f \alpha_s \left| \vec{v}_s - \vec{v}_f \right|}{d_s} \tag{9}$$

The solids pressure in Equation (3) is obtained according to Equation (10)

$$p_s = \alpha_s \rho_s \Theta_s + 2 \rho_s \left(1 + e_{ss} \right) \alpha_s^2 g_{0,ss} \Theta_s \tag{10}$$

where: the radial distribution function (Ogawa et al., 1980), Θ_s : the granular temperature (Ding and Gidaspow, 1990), e_{ss} : the coefficient of restitution (in this work was used $e_{ss} = 0.9$). The radial distribution function, g_0, is a correction factor that modifies the probability of collisions between grains when the granular phase becomes dense. This function may also be interpreted as the dimensionless distance between spheres:

$$g_{0,SS} = \frac{D_s + d_s}{D_s} \tag{11}$$

Where Ds is the distance between the particles. This distance is not constant in the several positions of the bed, because the voidage is also variable.

2. METHODOLOGY

2.1. Materials

The materials used in the experiments were alumina particles with properties shown in the table 1.

Table 1: Property of the materials

Property	Observed values
ρ_s (Kg/m^3)	3.500,0
d_p (μm)	110
ϕ	0.7

Figure 1 shows an image of the material used in this work.

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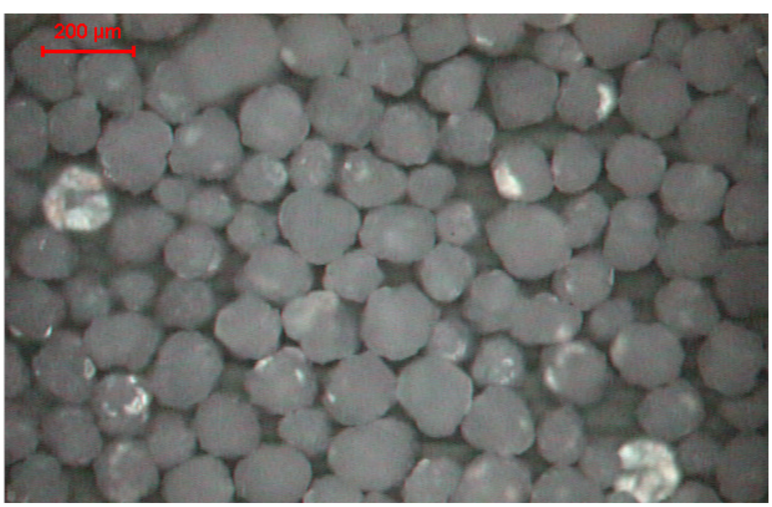

Figure 1. Image of the Alumina used in this work

2.1. Experimental Apparatus

Figure 2 shows a scheme of the experimental apparatus used in this research work and Table 2 shows the characteristics of this bed.

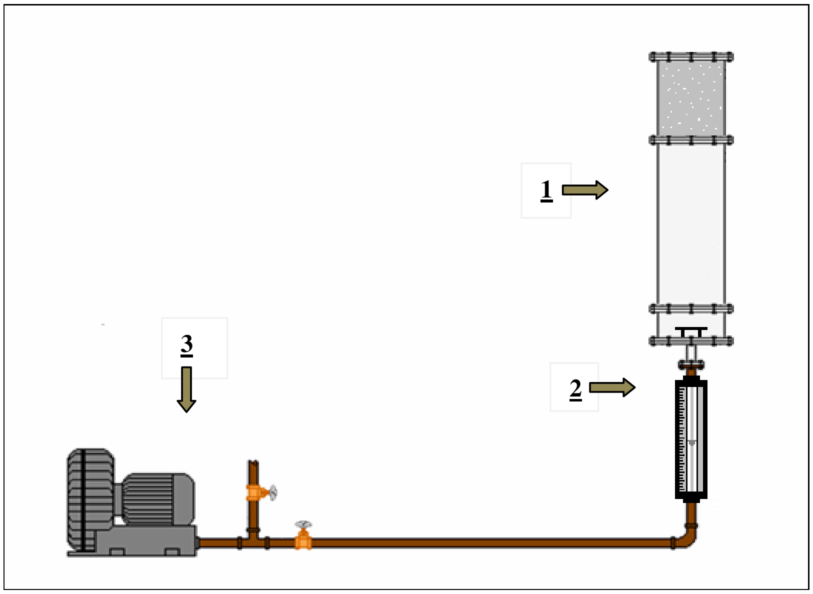

Figure: 2. Experimental Apparatus

The fluidized bed (1) was built in Plexiglas. Air was supplied by a 7.5 hp blower (3). The air flow rates were measured by a rotameter (2). The pressure drop was measured at the wall using a "U" tube manometer connected to a pressure tap located on the plenum and superior part of the column.

Table 2: Characteristics of the fluidized bed

Height of the plenum (mm)	138
Height of the cylindrical part (mm)	700
Height of Deflector Plate (mm)	62
Diameter of Deflector Plate (mm)	100
Diameter of gas inlet (Di) (mm)	21
Diameter of the column cylindrical part (Dc) (mm)	140

3. RESULTS AND DISCUSSIONS

3.1. Experimental results

Figure 3 shows the experimental fluidized bed characteristic curves for Alumina particles with 110 µm in diameter, for three different heights of the particles bed.

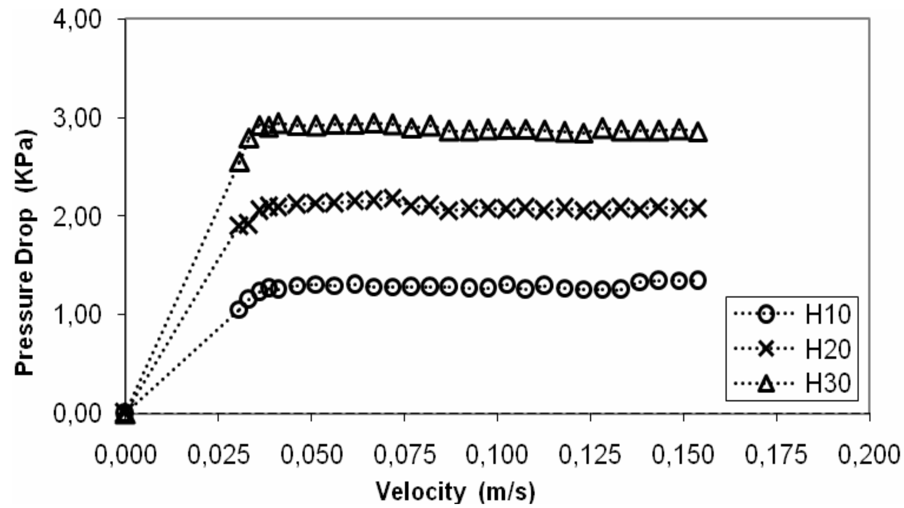

Figure 3. Characteristic curves for Alumina particles with 110 μm in diameter and different height.

From the experimental data of the pressure drop versus velocity, it was possible to identify the minimum fluidization condition as well as, the formation and the disappearance of the internal cavity. The minimum fluidization velocity (U_{mf}) was about 0.038 m/s for three heights. For heights 10, 20 and 30 centimeter the pressure drop at the minimum fluidization condition was 1.26, 2.09 and 2.9 KPa, respectively.

The expansion of bed, alternate of the 2.5 at 3.6 cm for three heights studied are presented in Figure 4. The expansion in a minimum fluidization was 1.2, 2.1 and 3.0 cm for 10, 20 and 30 cm in height respectively.

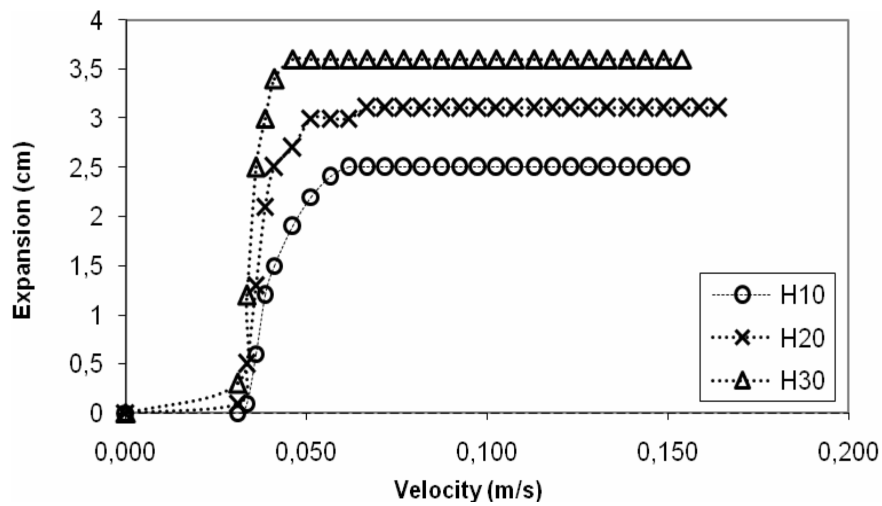

Figure 4. Expansion curves for alumina particles with 110 μm with different heights.

3.2. The voidage profile

The profiles of solids volume fraction obtained by simulation can be observed in Figure 5 for alumina particle with 110 μm in diameter, for height of the 10 cm particle bed. Through these Figures behavior observation of the fluidized bed regime is possible

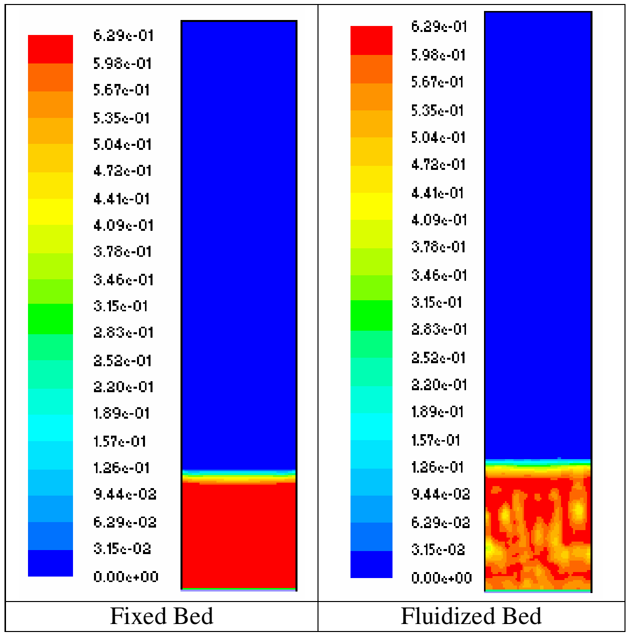

Figure 5. Distribution of solids volume fraction for Alumina particles (d_p =110 µm and H10).

3. FINAL REMARKS

In this paper the experimental pressure drop, fluidization velocities and expansion of bed for three different heights, as well as a CFD simulation by an Eulerian multiphase model were used to investigate the fluidized bed flow regime of the Alumina particles. From the experimental data of the pressure drop it was possible to identify the minimum fluidization condition. Through the profiles of solids volume fraction obtained by simulation using the Eulerian multiphase model, it was possible to identify the fluidized bed flow regime and the characteristic of this moving bed. The Eulerian multiphase model can also be used to simulate the fluidized bed flow regime in other configurations, with different geometries, operating conditions and particle types.

4. ACKNOWLEDGEMENTS

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5. REFERENCES

Duarte, C. R., Santana, R. C., Neto, J. L. V., Borges, J. E., Murata, V.V. e Barrozo, M.A.S.(2004) Estudo experimental e de simulação sobre a movimentação das partículas em um leito de jorro, CD-ROM Proceedings, XXXI – Congresso Brasileiro de Sistemas Particulados, Uberlândia, Brazil, October 24-27, 2004.

Ding, J. and Gidaspow D. (1990). A Bubbling Fluidization Model Using Kinetic Theory of Granular Flow. AIChE Journal, Vol. 36, pp. 523-538.

Ergun, S. (1952). Fluid Flow Through Packed Columns. Chem. Eng. Prog., Vol. 48(2), pp. 89-94.

Gidaspow, D., Bezburuah, R., and Ding, J. (1992). Hydrodynamics of Circulating Fluidized Beds: Kinetic Theory Approach. Proc. of the 7th Engineering Foundation Conference on Fluidization , pp. 75-82.

Mathur, K.B. & Epstein, N.(1974) Spouted Bed, New York, Academic Press, 304 p.

Ogawa, S., Umemura, A., and Oshima, N., J. (1980). On the Equations of Fully Fluidized Granular Materials. Appl. Math. Phys., Vol. 31, pp. 483

Smolders K. and Baeyens J.(2001). Gas fluidized operating at high velocities: a critical review of occurring regimes. Powder Technology. Issue 119. pp.269-291.

Wen, C. Y. and Yu, Y. H. (1966). Mechanics of Fluidization. Chem. Eng. Prog. Symp. Series, 62, 100-111.