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REFINED APPROACH TO CALCULATE THE PRESSURE DROP IN SMALL-SCALE FLUIDIZED BED

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Abstract. *This paper presents the methodology for refined calculation pressure drop over a small-scale bed of particles. The article compares the behavior of experimental pressure drop to a 2D CFD model with various particle diameters introducing corrections in the Ergun equation, based on the characteristics of the particles. The purpose was to refine the prediction of maximum pressure drop on fluidized bed columns for small-scale projects of air discharge systems.*

Keywords: *fluidized bed, CFD, pressure drop.*

1. INTRODUCTION

Gas-solid fluidization is the phenomenon whereby a bed of solid particles is submitted to the upward vertical passage of a gaseous fluid whose resulting suspension has a similar behavior to that of a fluid (Yang, 2003). Studies of gas-solid systems have gained strength due to the applications in the areas of drying, mixing, particle coating, powder agglomeration, heating and cooling solid, freezing, roasting coffee, pyrolysis and gasification. During the design phase of a small-scale fluidizer for fluid dynamic studies, the designer must size the gas propulsion system that meets the pressure and speed gradient requirements, capable of generating all fluidization regimes for many particle in bed levels. The most widely used propulsion systems are fans, blowers and compressors. The procedure most commonly found in literature for the prediction of the pressure load in small-scale fluidized beds applies the Ergun equation to the porosity vs. speed curve for each type of particle raised through: open correlations (EMMS Solver GPU group), Ergun BFB / CFB softwares (owner's correlations) or CFD (free such as MFIX, OpenFOAM and owners such as ANSYS, Star CCM +, Comsol Multiphysics). However, the application of the Ergun equation can overestimate the pressure drop of a fluidized bed column with particle diameters above 700 μm. This paper proposes the calibration correlations for each type of particle, with the purpose of refining predictions of pressure drop in fluidized beds to improve the design of small-scale fluidized systems.

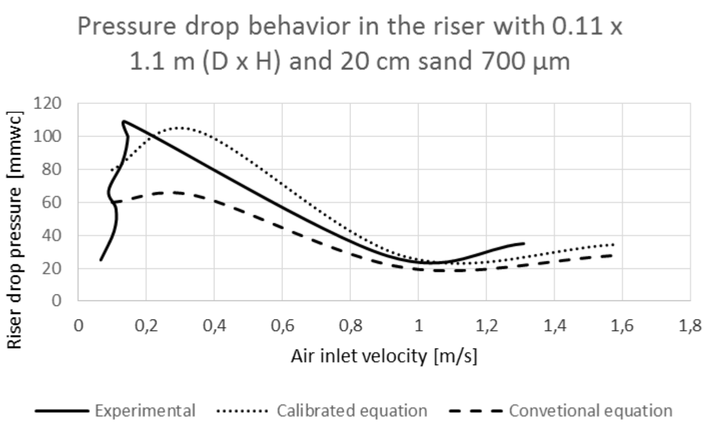

Figure 1: The dashed line shows the underestimation of maximum pressure drop over fluidized bed column using Ergun equation while the dots an estimation adjusted using calibrated Ergun equation.

2. BACKGROUND

The non-circulating fluidized plant is a small-scale process that usually consists of a fan, a flow meter, a windbox/riser set, a U-tube manometer, a cyclone and a particle reservoir for fluid dynamic studies. The non-circulating characteristic implies that the elutriated material does not return to the riser. Figure (2)(a) denotes the equipment that composes the fluidized plant while (b) shows the two-dimensional computational model of the riser.

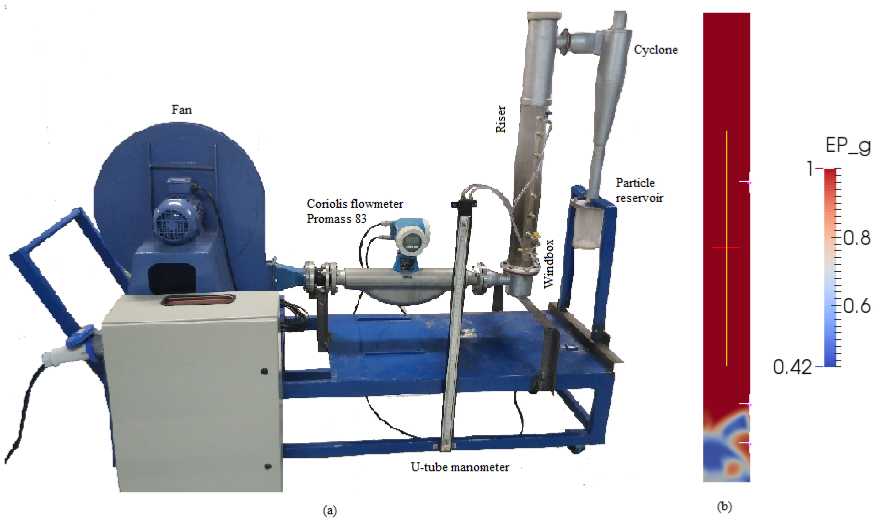

Figure 2: Fluidization plant in laboratorial scale.

To understand the fluid dynamics of a fluidized bed, it is necessary to study the main characteristics of the solid particles forming the bed, among which we highlight the average diameter, d_p, the bulk density, ρ_s, and the sphericity, φ. The porous bed is formed by a set of solid particles leaving voids between particles or conduits, which are occupied by fluid. The porosity ε of the bed (void fraction) is defined as the relationship between the volume of inter-particle voids and the total volume of the bed (solids and voids), so that ε is always less than unity. In 1973, Geldart proposed a classification of particles into four groups depending on their fluidization properties at ambient conditions, as shown in Fig. (3). Based on these properties, each letter (A, B, C, and D) corresponds to a type of Geldart classification.

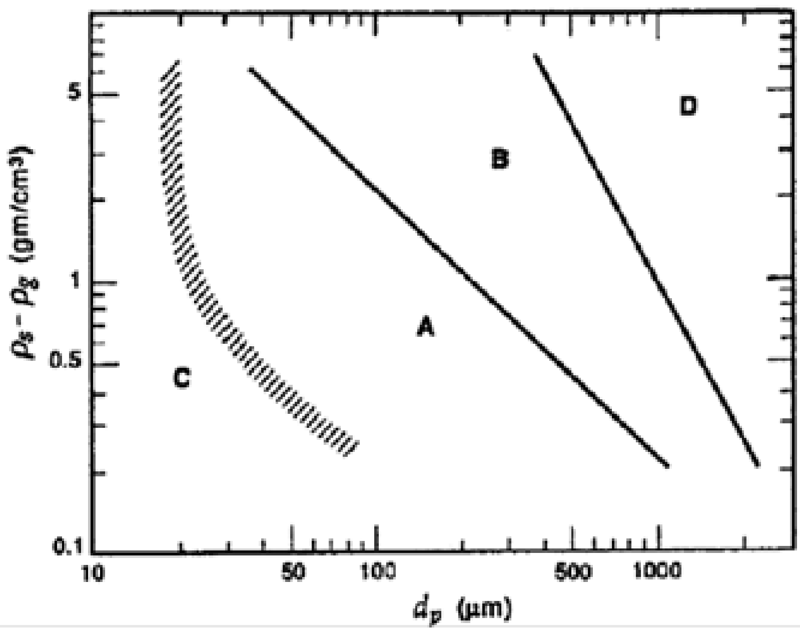

Figure 3: Particle classification according to Geldart (Gupta and Sathiyamoorthy, 1999).

Considering a bed of solid particles within a circular cross section column deposited on a perforated plate, one can observe different fluidization regimes as the fluid velocity increases. When the fluid motion is upward with low airflow, it passes through the spaces between particles, which remain stationary; characterizing the fixed bed regime. Increasing the fluid velocity, the particles distance themselves independently by small vibrations, being called as "expanded bed" in this state. With a larger increase of the velocity, a condition in which all the particles are suspended by upward flow of gas or liquid is reached. Thus, the pressure drop at a given bed volume is equal to the weight of the existing particles per unit area. The bed is called "fluidized bed" at minimum fluidization. Increasing the gas velocity, the stirring becomes intense and the movement of the particles becomes more vigorous. The bed does not expand much more than the minimum fluidization volume, and this regime is called aggregative, bubbling or heterogeneous fluidization. Other fluid-particle contact arrangements may occur when the gas bubbles collide with each other as they ascend through the bed, and reach almost the entire diameter of the column. It can happen in beds of great length and small diameter. This phenomenon is known as slugging regime. When the superficial gas velocity is increased considerably in a gas-solid system, one may surpass the value of the terminal velocity of the drag producing particles of solid material which causes the disappearance of the upper surface of the bed. This state is called a turbulent fluidized bed, according to Fig. (4).

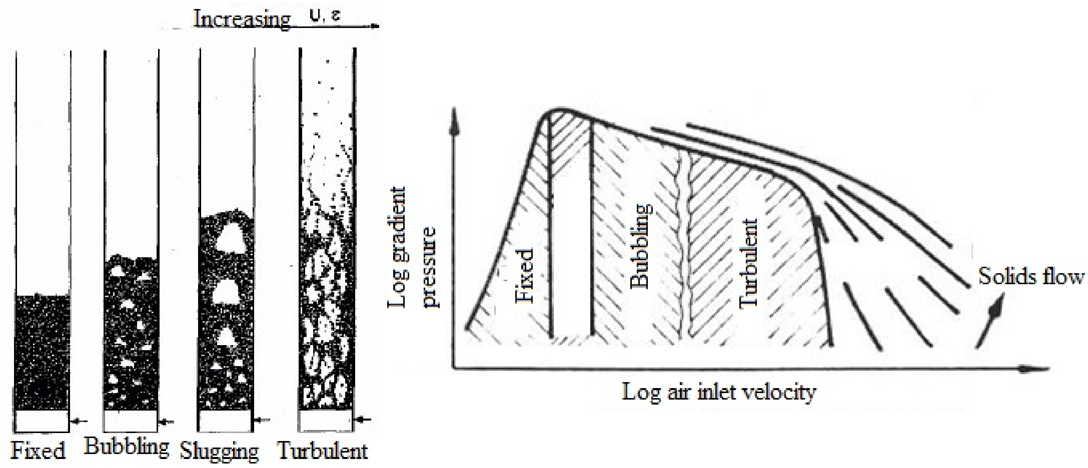

Figure 4: Scheme of possible fluidization regimes in non-circulating risers. (Yang, 2003)

The upward movement of the gas through a bed of particles results in a characteristic pressure drop (see Fig. (5)) of fluidized beds due to the resistance that the particles provide to the passage of gas. Designers of fluidized bed plants use the maximum pressure of this characteristic curve for the dimensioning of the gas propulsion system.

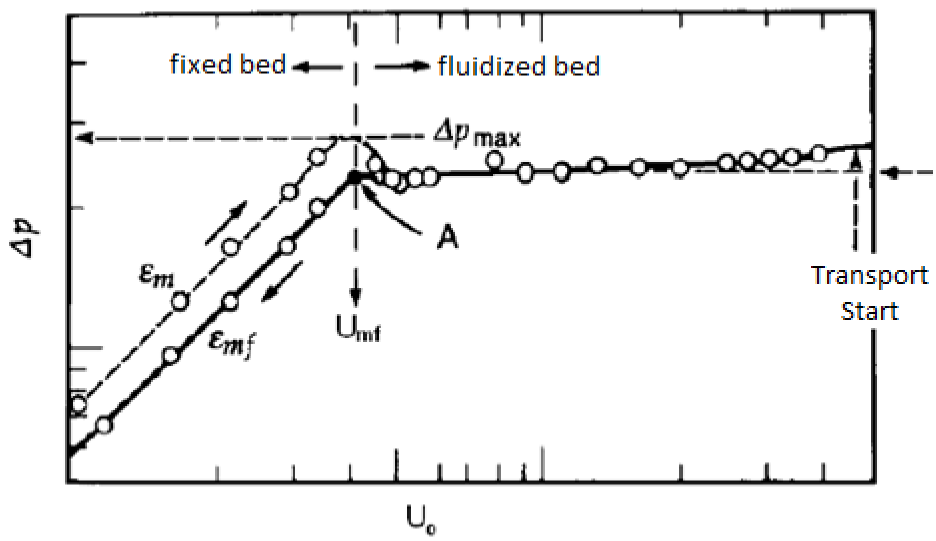

Figure 5: Pressure drop characteristics of fluidized bed particles as a function of superficial velocity input. (Yang, 2003)

In 1952, the chemical engineer Sabri Ergun proposed his famous correlation, Eq. (1), to estimate the pressure gradient between two pressure outlets considering the particle (sphericity, diameter) and fluid (viscosity and density) properties, and the average porosity in the same place.

$$\frac{\Delta p}{L} = \left(\frac{150(1-\varepsilon)^2\mu}{\varepsilon^3\varphi^2 d_p^2} U + \frac{1.75(1-\varepsilon)\rho_f}{\varepsilon^3\varphi d_p} U^2 \right) \tag{1}$$

The Ergun equation is a direct superposition of two asymptotic limits: lower limit for the laminar flow modeling in Darcy's scheme described by the Blake-Kozeny equation and an upper limit for modeling the turbulent flow, presumably, described by the Burke-Plummer equation (Cai *et al.*, 1998) (Ozahi *et al.*, 2008) and (Nemec and Levec, 2005). It is known from literature that the failure of Darcy's law is attributed to the turbulent regime (Macdonald *et al.*, 1979). Many scientists conducted numerous experiments on the flow through packed beds to check the validity of the Ergun equation. Based on empirical findings with some smooth particles, they proposed coefficient values of 180 and 1.8 (Woudberg, 2012) for the Ergun equation rather than the values of 150 and 1.75.

2.1 Ergun Correlation uncalibrated

For $A = 150$ and $B = 1.75$, the Eq. (2), generally gives good agreement with experimental data for 0.3 âĽď Îţ âĽď 0.5. However, experimental results show that the real porosity range goes beyond the stated range, during slug flow regimes and on.

$$\frac{\Delta P}{L} = \left(A\frac{(1-\varepsilon)^2\mu}{\varepsilon^3\varphi^2 d_p^2}U + B\frac{(1-\varepsilon)\rho_f}{\varepsilon^3\varphi d_p}U^2 \right) \tag{2}$$

Studies conducted by Du Plessis (Du Plessis, 1993), revealed a proposal for the correction coefficients A and B according to ζ, which is packed bed porosity of the formula by Ergun. This correlation relates A, Eq. (3) to B, Eq. (4), with only the porosity of each particle, resulting in determined values for each particle (Leva, 1947).

$$A = 41\frac{(1-\zeta)^2}{(1-\zeta)^{\frac{2}{3}}(1-(1-\zeta)^{\frac{1}{3}})(1-(1-\zeta)^{\frac{2}{3}})} \tag{3}$$

$$B = \frac{\zeta^2}{(1-(1-\zeta)^{\frac{2}{3}})^2} \tag{4}$$

2.2 Computational Fluid Dynamics

A real gas-solid system is complex, three-dimensional, transient, multiscale spatio-temporal and form dissipative heterogeneous structure (clustering) through nonlinear interactions in non-equilibrium conditions (Li *et al.*, 2013). In general, the multiphase flow may be modeled by computational fluid dynamics (CFD) via Euler (TFM) or the Lagrangian (discrete particle model, DPM) approaches, a direct type (DNS) or non-direct (large eddy simulations, LES, or Reynolds-averaged Navier-Stokes, RANS), homogeneous (by Wen-Yu correlation application without compromises between scales) or heterogeneous (EMMS model), hot (with chemical reactions or thermal exchanges) or cold, stationary or transient and 1-3D (dimensions) (Marin, 2011).

This article worked with a two-fluid model (TFM) for both gas and solid phases are described as fluids in continuous interpenetrating, allowing larger system simulations (for a small-scale plant) and variables of interest such as ε, the porosity, and Δp, the pressure drop over the bed of particles is reasonably predictable and which consumes less computational resource. The equations of motion are described by the Navier-Stokes equation and the transfer of interfacial moment is modeled by Eq. (6), (7) and (9).

$$\frac{\partial}{\partial t}(\varepsilon_f\rho_f) + (\nabla\varepsilon_f\rho_f\bar{u}) = 0 \tag{5}$$

$$\frac{\partial}{\partial t}(\varepsilon_f\rho_f\bar{u}) + (\nabla\varepsilon_f\rho_f\bar{u}\bar{u}) = -\varepsilon_f\nabla p - (\nabla\varepsilon_f\bar{\bar{\tau}}_f) - \bar{S}_p + \varepsilon_f\rho_f\bar{g} \tag{6}$$

where $\bar{u}$ is the gas velocity, p is pressure, $\bar{g}$ is gravity, ε_f is porosity of fluid, ρ_f is the density of fluid and $\bar{\bar{\tau}}_f$ represents the gas phase stress tensor. The sink term $\bar{S}_p$, represents the drag force exerted on the particles.

$$\bar{S}_p = \beta(\bar{u} - \bar{v}) \tag{7}$$

where β is the momentum transfer coefficient, $\bar{u}$ is the gas velocity, and $\bar{\bar{\tau}}_f$ represents the gas phase stress tensor. The sink term $\bar{S}_p$, represents the drag force exerted on the particles.

$$\frac{\partial}{\partial t}(\varepsilon_s\rho_s) + (\nabla\varepsilon_s\rho_s\bar{v}) = 0 \tag{8}$$

$$\frac{\partial}{\partial t}(\varepsilon_s\rho_s\bar{v}) + (\nabla\varepsilon_s\rho_s\bar{v}\bar{v}) = -\varepsilon_s\nabla p - (\nabla\varepsilon_s\bar{\bar{\tau}}_s) - \bar{S}_p + \varepsilon_s\rho_s\bar{g} \tag{9}$$

where $\bar{v}$ is the solid velocity, ε_s is the solids hold up, ρ_s is the density of solid and $\bar{\bar{\tau}}_s$ represents the solid phase stress tensor. This model was implemented in Fortran code using MFIX, two-dimensional non-circulating riser, transient, structured orthogonal grid and boundary condition of free-slip wall particles and no-slip wall for gas.

3. METHODOLOGY

Based on information collected in the literature review of this article, the methodology was divided into three stages: the experimental procedure, the CFD modeling, and model validation. The experimental procedure consists of describing the assumptions during measurements, handling equipment and operating conditions. In CFD modeling the settings and criteria used to create a TFM-2D model with equal fidelity to the experiment are approached. Finally, the model validation process consists of the comparison between the simulated and experimental situations.

3.1 Experimental Procedure

Experimental tests were carried out using different particle diameters, sphericity and bulk density. Table (1) summarizes the measurements taken of their individual and collective characteristics.

Table 1: Characteristics of particles and porosity measurements. specimens.

Particles	Diameter $[\mu m]$	ρ_s $[kg/m^3]$	Sphericity	U_t $[m/s]$	Porosity
Sand II	700	2650	0.85	4.084	0.443 ± 0.010
Glass	1000	2285	1	6.54	0.41
Drill Cuttings	1320	2400	0.55	3.59	0.440 ± 0.014
Sand III	2200	2600	0.88 ± 0.22	8.34	0.435

Each particle was deposited in the riser and subjected to air inlet speed steps. After 60 seconds, the flow reading (flow meter Coriolis PROMASS 83) and pressure drop over the bed (between the positions 95 and $185mm$ relative to the base of the riser) and the freeboard (between the positions 185 and $705mm$ from the riser base) was checked at each velocity.

3.2 CFD modelling

The simulation used herein, being of the type TFM 2D, is characterized by transient structured orthogonal grid, the riser dimension $0.11 \times 1.1m$ ($5mm$ /cell in the x and y directions and 4840 cells), Wen-Yu drag model. A test consists of multiple simulations with different gas inlet velocities, starting from $0.059m/s$ (below the minimum fluidization speed) to the terminal velocity U_t of each particle (See Fig. (6)). Therefore, in each test performed there was a velocity profile capable of exciting the bed in all of its fluidization regimes.

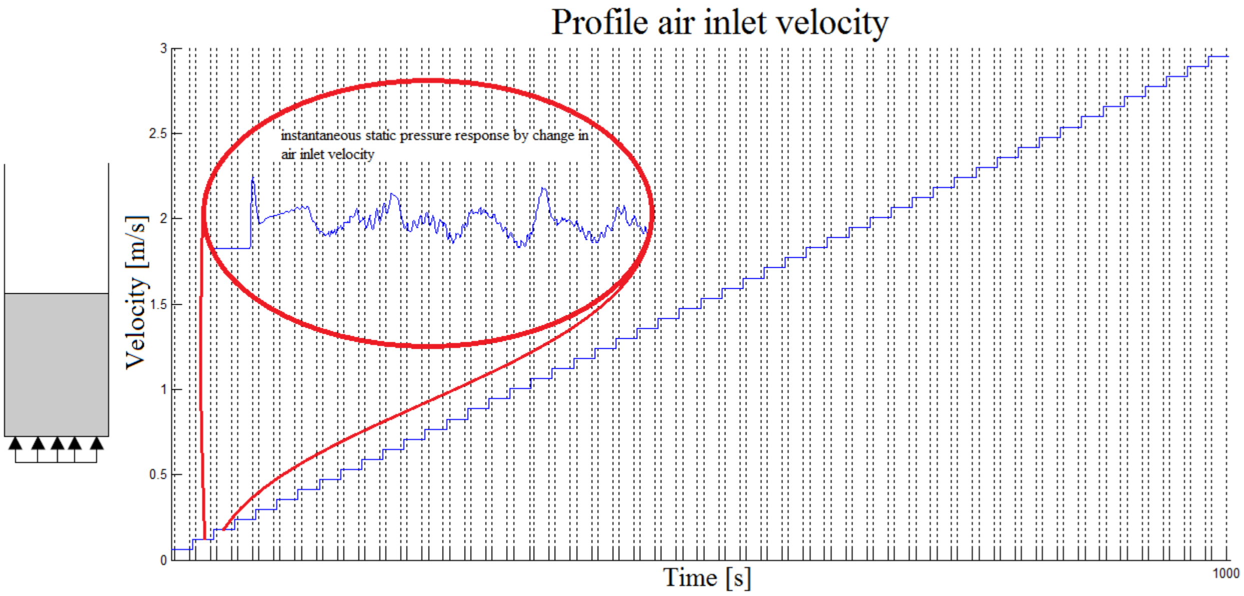

Figure 6: Inlet velocity profile with numerical simulation, similar to the experimental profile.

For post-processing, the 2D riser was divided into two control surfaces: the bed and the freeboard sections. The instant porosity and the lower and upper differential pressure were extracted from each section. The same was done in the freeboard section. The pressure drop information was separated to validate the model with the experimental plant. The validation of the model presented in this article seeks to compare the riser pressure drop (bed + freeboard) with the experimental data for the purpose of confirming the validity of the proposed model.

4. RESULTS

First, the validation process occurred by comparing the behavior of non-spherical particles (real) with a TFM-2D simulation of spherical particles (MFIX code does not allow sphericity specification). Then the equation has been made of the calibration coefficients for each particle. Finally, they were summarized comparative between measurements, conventional Ergun equation and calibrated.

4.1 Model validation

The comparison between the pressure drop in the simulated and experimental riser shown in Fig. (7) indicates good predictability of the TFM-2D CFD model when submitted to the same inputs, indicating a tendency to predict the behavior of the porosity of the bed and the freeboard accurately.

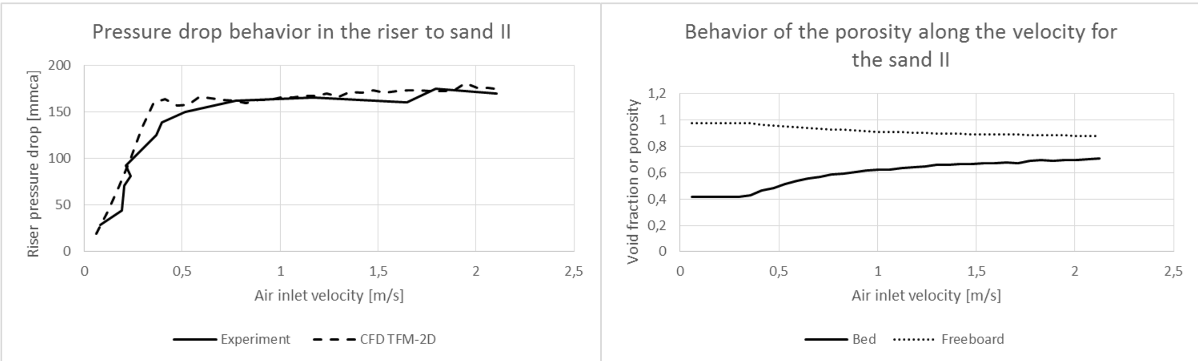

Figure 7: Comparison between pressure drops in experimental and simulated riser, and the behavior of the porosity simulated in the bed and freeboard.

4.2 Calibration

According to the measured data of the particle, it performed the calibration of Ergun's equation for each type of particle. Table (2) summarizes the calibration coefficients, using the method suggested by Du Plessis (1993).

Table 2: Calibration coefficients of the Ergun equation for each particle.

Particles	Coefficients	
	A	B
Sand II	207.36	1.88
Glass	204.85	1.99
Drill Cuttings	207.58	1.88
Sand III	206.93	1.89

The porosity results were validated and applied to the conventional and the calibrated Ergun equation, as shown in Tab. (3).

Table 3: Comparative in mmwc, the maximum pressure drops between the experimental values and calibrated Ergun equation.

Particles	$\Delta P_{experimental}$	ΔP_{CFD}	$\Delta P_{Ergun-Calibrated}$	$\Delta P_{Ergun-Conventional}$
Sand II	127	163.92	132.92	102.44
Glass	153	143.6	139	74.125
Drill Cuttings	144.5	144.52	146.06	96.93
Sand III	165	163.02	166.4	15.87

In all cases, the calibrated correlations presented results closest to experimental values for the maximum pressure drop.

5. CONCLUSIONS

For diameters above 700 μm the conventional Ergun equation underestimated the maximum pressure drop in the riser, proving the need for correction or calibration. A CFD modeling validation has provided a good porosity profile along different air velocities for each particle. In this profile porosity Du Plessis's methodology was applied to perform a calibration for each particle and to estimate the higher pressure drop over the riser. This procedure may comprise the designer's methodology for dimensioning the air propulsion in small-scale fluidizing systems.

6. ACKNOWLEDGEMENTS

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