DESIGN AND SIMULATION OF A CIRCULATING FLUIDIZED BED TO CLEAN THE PRODUCTS OF BIOMASS GASIFICATION

Moisés Uchôa Neto, uchoa.moises@gmail.com
Yuri de Araújo Carvalho, yuricarvalho@hotmail.com
Departamento de Engenharia Mecânica, Faculdade de Tecnologia, Universidade de Brasília, Campus Darcy Ribeiro, Asa Norte, Brasília - DF, 70910-900.

Taygoara Felamingo de Oliveira, taygoara@unb.br
Manuel Barcelos, manuelbarcelos@gmail.com
Faculdade do Gama, Universidade de Brasília, Área Especial 2, Lote 14, Setor Central, Gama - DF, 72405-610.

Abstract. *The main goal of this work is to design a workbench circulating fluidized bed to study the cracking of tar in gases from the processes of biomass gasification. For this, a design methodology based on analytical results and empirical correlations for fluidized beds was employed. In parallel, a numerical code of open source technology (MFIX) for the solution of the transport equations of the multiphase flow in the column of a fluidized bed was used to give support to the choice of the design elements. The whole project of the workbench fluidized bed was completely developed, whose operation parameters such as bed geometry, gas velocity, circulating ratio and void fraction characterize a fast fluidization process. A preliminary mesh convergence study was executed with the numerical tool, that was validated comparing with analytical results. Among the most important results, the code computed the predicted value for the minimum fluidization velocity of the bed.*

Keywords: *Multiphase flow, Fluidized Bed, Fast Fluidization, Design Methodology and Numerical Simulation.*

1. INTRODUCTION

The relative flow between solid particles and a fluid (gas or liquid), through a vertical tube, make the particles to move up, promoting a multiphase flow, known as fluidized bed. This type of flow has many applications in the industry field, mainly in chemical and processes industries (Basu, 2006).

The regime of fast fluidization is characterized by high flow velocities, elevated friction between gas and solid particles and a considerable process of circulation of solids. Such characteristics are very appropriated to promote the clean of gases from processes of gasification of biomass, due to the cracking of tars. In this way, the energy of the resulting gas at the end of the process of gasification is higher than at the entrance of the fluidized bed, besides that, there is a lower concentration of hydrocarbons that can condensate on the equipment walls, which use the gas as fuel.

In order to study this regime of fluidization and its application on cleaning gases, the objective of this work includes the design of a workbench fluidized bed, including the reactor, the gas-solid separator, and a solid recycle system. Numerical analysis of the flow conditions in the specification of the project also will be realized.

2. DESIGN METHODOLOGY OF A CIRCULATING FLUIDIZED BED

2.1 Selection of solid particles

The first step in a fluidized bed design needs the physics characterization of the gas and solid particles of the bed. In the light of fluidization experience, Geldart (1973) classified solids broadly under four groups, A, B, C, and D. This classification is a function of the diameter and density of the particles. This classification is very important, since particles of different groups have different behaviors at similar conditions.

According to Basu (2006), circulating beds commonly use particles of group A, which are particles in the range from 30 to 100 μm. The particles of group A fluidize softly at low velocities and have a controlled bubbling at higher velocities. Particles of group B are used in circulating fluidized beds as well.

Taking in account all these information, the workbench circulating fluidized bed design will consider sphere particles of glass with 50 μm (group A) e 200 μm (group B), that will be fluidized by the atmospheric air.

2.2 Fluidized Bed Characteristic Velocities

2.2.1 Selection of solid particles

Considering a vertical tube, in which there are an amount of particles, it is possible to imagine an experiment where air is injected by the bottom of the tube at increasing velocities. While the velocities are sufficient low, the particles remain stationary, and the bed has a behavior of a porous mean, such that the pressure loss across the bed is given by Darcy's law.

As the upwards flow increases, it reaches a critic point, at which the behavior of the bed changes and it suffers a small expansion. This velocity is known as minimum fluidizing velocity. At this point, the drag force becomes equal to the apparent weight of the bed. That can be written as

$$\Delta p\, A = A\, L_{mf}(1-\varepsilon_{mf})(\rho_p-\rho_g)g, \tag{1}$$

in which Δp is the pressure loss across the bed, A is the cross section area of the tube, L_{mf} the height of the bed at the minimum fluidization, ε_{mf} is the void fraction at minimum fluidization, rho_p is the density of solid particles, rho_g the density of gas and g the gravity acceleration.

The left side of Eq.(1) represents the drag on the bed, and the right side represents the weight of the particles. Developing this relation together with the Ergun's relation, we obtain the following expression.

$$Re_{mf} = \frac{(U_{mf} d_p \rho_g)}{\mu} = [C_1^2 + C_2 Ar]^{0.5} - C_1 \tag{2}$$

in which Re_{mf} is the Reynolds number of particles at the minimum fluidization condition, U_{mf} is the minimum fluidization velocity, d_p is the particle diameter, μ the viscosity of gas, C_1 e C_2 are constants recommended by Grace and Ar is the Arquimeds number, defined by Grace (1982).

$$Ar = \frac{\rho_g(\rho_p-\rho_g)g d_p^3}{\mu^2} \tag{3}$$

The expansion of the bed means that L_{mf} and ε_{mf} are different at minimum fluidization from they were before. The minimum fluidization velocity marks the transition from fixed to fluidized bed.

2.2.2 Transport Velocity

The transport velocity is the velocity under which the fast fluidization cannot occur, independent of the solid circulation rate. From this point, the velocity of the gas through the bed is higher than the terminal velocity of the single particles. Thus, a fluidized bed which operates at fast fluidization regime needs a system of separation of particles from the suspension and another system that provides a continuum flux of these particles back to the basis of the bed, such way the bed does not get empty.

A empiric relation for the transport velocity, based on experiments on a 92mm diameter column, is given by Perales et al. (1991),

$$U_{tr} = \frac{1.45 \mu Ar^{0.484}}{\rho_g d_p}, \quad 20 < Ar < 50000 \tag{4}$$

Considering the equations (4) and (2), the transport velocities obtained for the particles used in this work are showed at Tab(1). Additionally, in Fig.(**??**), the full circles mark the point which the velocity of gas reaches the transport velocity.

Table 1. Characteristics Velocities of the selected particles

Diameter	U_{mf}	U_{tr}
50μm	0.0024	1.394
200μm	0.0388	2.610

2.2.3 Choking Velocity

The choking velocity marks the beginning of fast fluidization, immediately after the turbulent fluidization regime, what makes this parameter of extreme importance to the design of the workbench circulating fluidized bed. The choking velocity is directly connected to the solid circulation rate. The higher is the solid circulation rate, more stable is the bed. The Yang's correlation (Yang, 1983), shown in Eq.(5) developed to particles of group A, at ambient temperature and sing columns of pneumatic transport of small diameter (D<0.3m), can be used for a first approximation of this parameter.

$$\frac{U_{ch}}{\varepsilon_c} = U_t + \sqrt{\frac{(2gD(\varepsilon_c^{-4.7}-1)\rho_p^{2.2})}{(6.81 * 10^5 \rho_g^{2.2})}} \tag{5}$$

in which U_{ch} is the choking velocity, ε_c the void fraction at this velocity and U_t the terminal velocity of the particles. The Eq.(5) must be solved simultaneously with a definition of the solid circulation rate at a velocity equals to the choking velocity, given by Eq.(6),

$$G_s = (U_{ch} - U_t)(1 - \varepsilon_c)\rho_p \tag{6}$$

In which G_s is the solid circulation rate, given in kg/sm^2. For each value of G_s, the Eq.(5) and (6) are solved simultaneously to determine U_{ch} and ε_c. The result of this process can be observed at Fig (1).

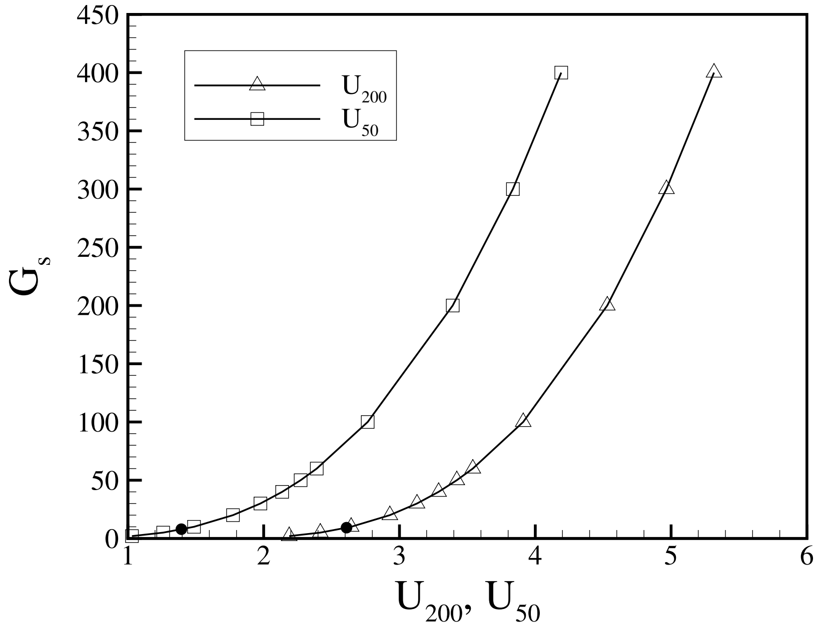

Figure 1. Choking Velocities for different particles. The full circles mark the transport velocity values. The units of G_s and U are respectively kg/m^2s and m/s

2.3 Workbench circulating fluidized bed design

2.3.1 Reactor

The reactor is the region where the bed remains, where the contact between gas and particles occurs (Ideally dolomite or another catalyst). The more important factor to reach determinate fluidization regime is the gas velocity. To control this parameter it is possible to modify the flow and the cross section area of the tube. This way, it is possible to choose a cross section that suits the project, only observing the possibility of installation of a fan that provides the necessary velocities. It is important to note that tubes of smaller diameters have more tendencies of present radial variation of flow properties, due to the enhancement of wall effects in these cases. The column height is important when chemical reactions are to be considered, since it will affect the contact time of gas and solids. The reactor will be made of acrylics to permit the visualization of the internal part. The chosen diameter for the tube was 12 cm, to operate with velocities of magnitude close to 4 m/s.

2.3.2 Gas-Solid Separator

In fast beds matters, the more used type of separator is the cyclone. The process of solids separation is done by centrifugal forces, the solids migrate towards the wall when a circular movement is imposed to the suspension. The selected type of separator for this bed was the vertical axis tangential entry cyclone. The superior part of the cyclone has a cylindrical shape, while the inferior part has a conic shape. The gas-solid suspension enters tangentially in the cyclone by a rectangular duct. The cylindrical shape of the cyclone and the tangential entry of the gas cause the gas-solid suspension to flow in two concentric vortices around the cyclone. The outer vortex travels downwards and the inner one travels upwards. Solids, being denser than the gas, leave the outer vortex to migrate towards the wall under the action of centrifugal force with a radial velocity V_r. The viscous effects are higher near the wall, making the suspension velocity to decrease, which is no longer able to carry the solids, which fall down and are collected by the inferior part of the cyclone, from where they are taken back to the reactor basis.

The solid movement in the radial direction is given by the equilibrium of the centrifugal force with the viscous drag, given by Stokes Law. Developing the equation, we get the radial velocity

$$V_r = \frac{d_p^2 \rho_p V_e^2}{18 \mu r}, \tag{7}$$

in which V_e is the gas velocity at the entry of the cyclone and r the cyclone radius.

In an ideal situation, solids reaching the wall are fully captured. The efficiency of the cyclone, thus, will be directly proportional to the solid migration velocity V_r. The faster the particle moves towards the wall, greater is its probability of being captured before it is entrained. But, the real operation is quite different from this ideal situation. It is very difficult to define a cut-off diameter where all the particles with smaller diameters will be entrained, and the bigger will be collected. Designers have defined a practical cut-off size known by d_{50}. Particles larger than this size have 50% chance of being collected. Pell and Dunson (1997) define d_{50} considering the number of spirals that a single particle travels in the cyclone, as shown in Eq.(8),

$$d_{50} = \sqrt{\frac{18 \mu L}{2 \pi N_c V_c (\rho_p - \rho_g)}} \tag{8}$$

where L is the width of the rectangular duct at entry and N_c is the effective number of spirals done by the gas inside the cyclone given by Zenz (1997).

For the design of the cyclone, Basu (2006) suggests the utilization of a table with dimensional relations depending on the main purpose of the cyclone. In this work we have chosen to adopt the relations for high efficiency cyclone given by Swift (1969). The detail of the separator may be seen in Fig.(2b).

2.3.3 Solid Recycle System

As soon as the solids are separated of the gas, they need to be launched to the bed again. If the solids flow rate can be controlled, that will give a great freedom in experiments. To make that, a mechanism was chosen to dose the quantity of particles that enters the bed. This mechanism works with a transport screw connected to a stepper motor, that can be viewed at Fig.(2c), with details to the sections of entry and exit of particles. When the screw rotates, the particles are transported continuously with a known rate. In Fig.(2a) it is shown the drawing of the complete workbench with all its components.

3. PHYSICAL MODELING

The proposed study in this work uses the MFIX model (Multiphase Flow with Interphase eXchange) to evaluate the physical processes which occur inside the fluidized bed reactor. The MFIX models the physical problem using a set of main equations related to mass, momentum, energy and the chemical species balance to the gas phase and multiple solid phases, complemented by secondary equations for granular tension based on kinetics theory, on flow with friction theory and on chemical reactions (Syamlal et al. (1993)). In other words, the MFIX model is based on hydrodynamics theory of fluidization with capacity of describe chemical reactions and heat exchange in dense or diluted solid-fluid flows.

In MFIX, the different steps are described mathematically as a continuum interpenetrating using modeling by averages to represent the problem variables. The variables are the averages upon a region, which is large when compared to the particle size, but is quite smaller than the domain of the fluid. Thus, a field variable is introduced the phases volume fractions (ε), allowing the trace of the average volume fraction occupied by several phases. The sum of all volume fractions must be equal to one, such that

$$\varepsilon_g + \sum_{m=1}^{M} \varepsilon_{sm} = 1 \tag{9}$$

The subscripts $(\,)_g$ and $(\,)_s$ represent the gas and solid phases, respectively, and M is the number of solid phases and the index $(\,)_m$ identifies a specific solid phase. The effective densities are $\rho_{g'} = \varepsilon_g \rho_g$ to the gas phase and $\rho_{s'm} = \varepsilon_{sm} \rho_{sm}$ to the solid phase.

The conservation of mass is expressed by the equations 9 and 10, respectively relatives to the continuity for a gas phase and for multiple solid phases.

$$\frac{\partial}{\partial t} (\varepsilon_g \rho_g) + \nabla . (\varepsilon_g \rho_g \vec{v}_g) = \sum_{n=1}^{N_g} R_{gn} \tag{10}$$

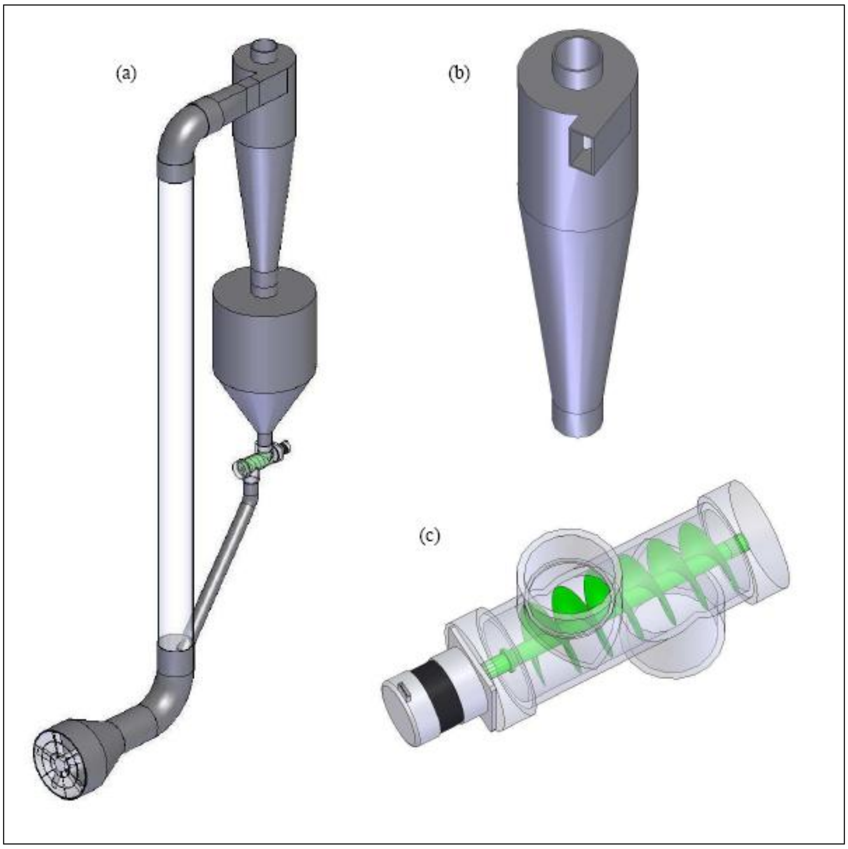

Figure 2. Circulating Fluidized Bed

$$\frac{\partial}{\partial t}\left(\varepsilon_{sm}\rho_{sm}\right)+\nabla.\left(\varepsilon_{sm}\rho_{sm}\overrightarrow{v}_{sm}\right)=\sum_{n=1}^{N_{sm}}R_{smn} \tag{11}$$

In equations 9 and 10 the symbol $\overrightarrow{v}$ represents the velocity of the phases. The term $R_{()}$ in the right side represents the mass transfer through the interface due to different processes such chemical reactions and evaporation. The subscript N takes in account the total number of processes, and the subscript $()_n$ indicates a specific process. The fluid phase equations are complemented by state equations, with the fluid being modeled by a gas obeying to ideal gas law or a incompressible fluid with constant density.

In conservation of momentum, the balance of momentum follows the same separation used in continuity equations, so that to the gas and solids phases we have

$$\frac{\partial}{\partial t}\left(\rho_g\varepsilon_g\overrightarrow{v}_g\right)+\nabla.\left(\rho_g\varepsilon_g\overrightarrow{v}_g\overrightarrow{v}_g\right)=\nabla.\overline{\overline{S}}_g+\rho_g\varepsilon_g\overrightarrow{g}-\sum_{m=1}^{M}\overrightarrow{I}_{gm}+\overrightarrow{f}_g \tag{12}$$

$$\frac{\partial}{\partial t}\left(\rho_{sm}\varepsilon_{sm}\overrightarrow{v}_{sm}\right)+\nabla.\left(\rho_{sm}\varepsilon_{sm}\overrightarrow{v}_{sm}\overrightarrow{v}_{sm}\right)=\nabla.\overline{\overline{S}}_{sm}+\rho_{sm}\varepsilon_{sm}\overrightarrow{g}-\sum_{l=1,\,l\neq m}^{M}\overrightarrow{I}_{ml} \tag{13}$$

In which $\overline{\overline{S}}_{()}$ represents the phases tensor, $\overrightarrow{f}_g$ is fluid friction due to internal porous surface, $\overrightarrow{I}_{gm}$ are force interactions between gas and solid phases, $\overrightarrow{I}_{ml}$ are the force interactions between the solids phases and $\overrightarrow{g}$ is the gravitational force vector.

The MFIX model only takes in consideration the buoyancy ($-\varepsilon_{sm}\nabla P_g$, where P_g is the pressure in gas phase), the drag force, (($\overrightarrow{F}_{gm}\left(\overrightarrow{v}_{sm}-\overrightarrow{v}_g\right)$, where $\overrightarrow{F}_{gm}$ is a parameter that links correlations of terminal velocity to drag correlations) and momentum variables due to mass transfer ($R_{gm}\left[\xi_{gm}\overrightarrow{v}_{sm}+\left(1-\xi_{gm}\right)\overrightarrow{v}_g\right]$, with $\xi_{gm}=1$ to $R_{gm}<0$ and $\xi_{gm}=0$ to $R_{gm}>0$), as interactions between gas and solids because these are the most significant. The force interactions between solid and solid are assumed just as drag functions between the solids phases, what is quantified in MFIX model based on kinetics theory.

The conservation equations for the gas are complemented with common equations of the fluid state, and the tensor of the gas is modeled with hydrostatic ($-P_g\overline{\overline{I}}$) and viscous ($\overline{\overline{\tau}}_g$) components. However, the tensor of multiples solids phases is modeled as the solids phases were plastic granular flows (slow shears, continuous contact, momentum transfer by friction) and viscous flows (fast shears, transient contacts, momentum transfer by collision) having similar components to fluid flows, such as the tensor of the fluid ($\overline{\overline{S}}_{sm} = -P_{sm}\overline{\overline{I}} + \overline{\overline{\tau}}_{sm}$). The tensor model depends on packaging condition of the bed ε_g^* ($\varepsilon_g \leq \varepsilon_g^*$ For plastic granular flows, $\overline{\overline{S}}^p_{sm}$, $\varepsilon_g > \varepsilon_g^*$ for viscous flows, $\overline{\overline{S}}^v_{sm}$).

As the flows modeled with MFIX for the proposed study for this work are of isothermal nature, and the main interest is the flow solution with interaction between gas and solid particles, but without occurrence of chemical reactions. The equations related to the energy conservation will not be introduced because they do not make part of the utilized modeling.

All conjunct of equations showed here can be discretized over non uniforms meshes, using systems of Cartesian Coordinates or Cylindrical coordinates and solved for method of finites volumes. The core of numerical code is based in SIMPLE (Semi-Implicit Method for Pressure Linked Equations) (Syamlal, 1988). Small modifications are present in SIMPLE for improve the stability and speed of calculus. First, the MFIX has implemented in the code one equation to correction of fraction of solid volume, which improves the convergence when the solids are with low density of packing and stabilize the calculus in regions with high density of packing too. Besides that, the MFIX has adapted to the numerical code an automatic adjust of time step for ensure high speed of execution. The accuracy of numerical discretization is improved using numerical schemes of second order to discretize in space the terms of convection, which help to control numerical wrongs of diffusion, and as direct result, there is the prediction of pattern in flows that represents in a form more realistic the physical of studied problem. The MFIX code is written in a FORTRAN language and has a friendly structure to the user, as entry files of easy manipulation, optimized files to minimum occupation of storage space, tools for acceleration of recovery of dates and a complete report of errors.

4. PRELIMINARY RESULTS OF NUMERICAL SIMULATION

4.1 Simulation of Hagen-Poiseuille flow

To validate the code, it was simulated an axissimetric air flow inside a tube with diameter 14 cm and length 200 cm. The boundary conditions were: Pressure defined in entry and exit sections and non-slipping wall. The result found agrees with the analytic solution, how we can see in the Fig(3)

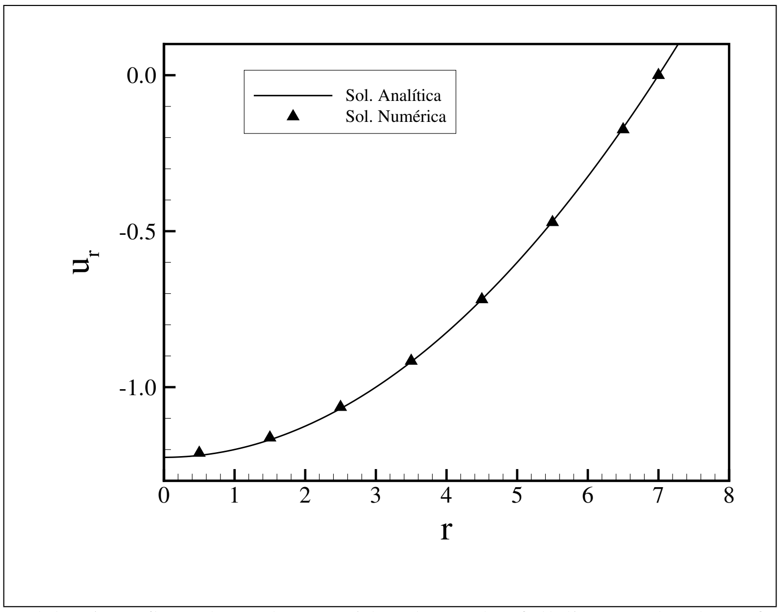

Figure 3. Fully developed laminar flow through a straith tube. The full line marks the referential analytical result: $u = -\dfrac{\partial p}{\partial}\dfrac{R^2}{4\mu}(1 - \dfrac{r^2}{R^2})$. The units of U_r and r are respectively m/s and cm.

4.2 Finding the minimum fluidization Velocity

To estimate the U_{mf}, a simulation of a bed with glass particles of diameter 50um in a tube of diameter 12 cm and length 200 cm was realized. The simulation was realized in 2 dimensions with axissimetric profile and with initial conditions:

Gas velocity and particle velocity equal to zero. For the boundary conditions we declared the pressure on top and the velocity of gas constant in the bottom of the bed as well the non-slipping wall condition. The simulation begun with velocity zero and increased at time steps. In this part it was taken carefully with the time steps, because was necessary that the flow been in steady state for each velocity. In the end it was written the difference of pressure in the tube for each velocity simulated. Initially it was realized a study of mesh convergence for more accurate results. All the simulations were executed with 3 meshes as can be seen in the Fig(4). From mesh 1 to mesh 2 was obtained a more accurate result, but in the mesh 3, the code did not converged from a determinate point and the improvements in relation of mesh 2 were not so significant. With this result the mesh 2 was utilized in the simulations. In the Fig(5) were made 2 distinct linear adjustments: For velocities below 0.0028 m/s, was obtained a constant line, feature of fluidized beds. The crossing of lines can be used to estimate the minimum fluidization velocity. In this case, was found the value of 0,0024 m/s, which is agree with the estimate shown in the Tab(1).

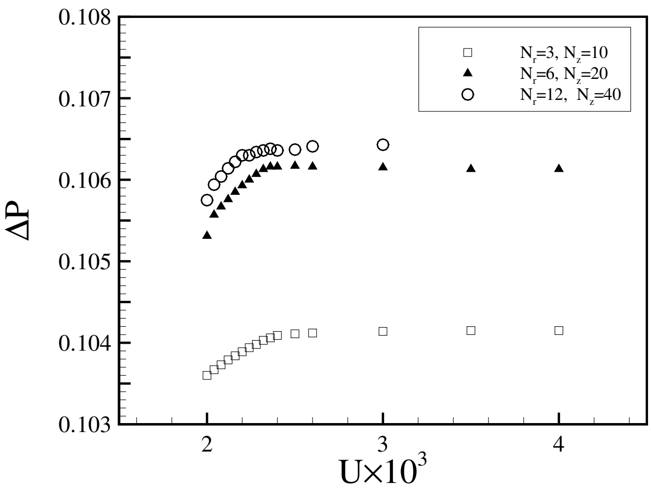

Figure 4. Mesh convergence study.

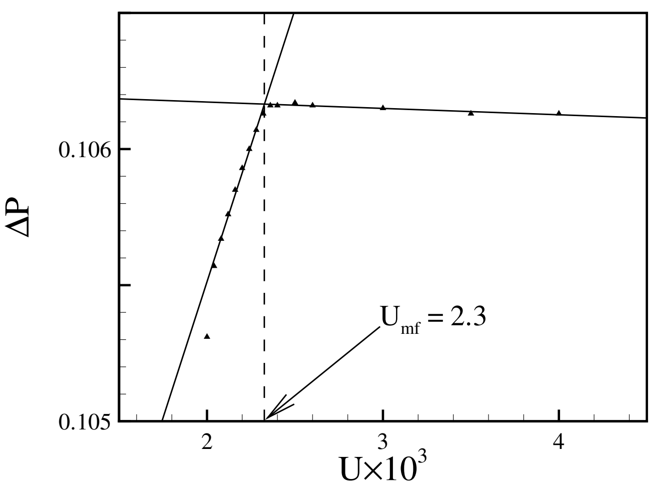

Figure 5. Minimum fluidizing velocity. The units of ΔP and U in figures 4 and 5 are respectively $Pa * 10^5$ and m/s

5. CONCLUSION

A project of a circulating fluidized bed was designed. For this purpose, was used a methodology that combine analytical techniques with aid of empirical relations and numerical techniques for simulation of flows. Using the analytical techniques the main parameters of project were specified, so a complete drawing of the workbench could be made with all constructive details. The numerical simulations were realized using the free code MFIX. This tool was validated with success, by comparison with experimental results. The velocity of minimum fluidization was determinate using this code, and the result founded agreed with the previsions realized in the first step.

Future works will search improvements of numerical analisys by simulations of fast fluidization. In this case, boundary conditions with entry and exit of particle must be defined to control the circulation rate. When the parameters of the bed in a quick fluidization state have been detailed, the workbench will be construed for the realization of experimental tests, correspondents to fluidization state.

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