EXPERIMENTAL STUDY OF GAS DISTRIBUTORS FOR FLUIDIZED BEDS CONTAINING SAND-BAGASSE MIXTURES

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Abstract. *This paper presents an experimental study of the performance of gas distributors in a bubbling fluidized bed using mixtures of sugar cane bagasse and sand particles as solid material. Tuyere type distributors with orifice diameters of 3, 4 and 5 mm were tested inside fluidized beds with biomass-to-inert mass ratio from 0 to 0.4%. Measurements of bed pressure drop as a function of superficial gas velocity were made for each tested gas distributor and biomass-to-inert mass ratio. Therefore this work presents the behaviour of bagasse/sand mixtures in a fluidized bed using tuyere type gas distributors in order to study the influence of these parameters on fluidization quality. Tests were conducted in a fluidized bed made of glass, acrylic and carbon steel sections allowing visual observation of the gas-solid flow. Results showed significant erosion at the top of each tuyere after two hours operation and a uniform bed was obtained at operational conditions near the predicted minimum slugging velocity for all orifice diameters studied.*

Keywords*: fluidized bed, gas distributors, sugar cane bagasse, biomass, bio-energy*

1. INTRODUCTION

The increasing demand for energy claims for improvements on energy generation processes. Among these processes, fluidized bed technologies applied to combustion, gasification or pyrolysis processes have been applied in thermal electric industries having the advantage of being able to operate with a wide range of solid fuels as coal, oil shale, urban wastes and different types of biomass. Researchers have been directed to create an equilibrium between economical and environmental aspects in order to obtain the best operational condition concerning each process. Fluidized bed technologies are characterized by the high conversion efficiencies and low emissions of atmospheric pollutants. Also, regarding countries where there is high quantity of sugar and alcohol mills, the sugar cane bagasse can become a good solution for energy co-generation using the fluidized bed technology (Cortez *et al*., 2008; Johnsson, 2007). However, the lack of knowledge about the process, high costs and power consumption involved are problems to be solved before industrial applications (Joyce *et al*., 2006; Basu, 1984).

Currently, bagasse boilers uses technological systems that provide low combustion efficiencies such as pin-hole and travelling grate, so, the application of fluidized bed technology on these systems will certainly improve combustion efficiency.

The air distribution grade is an important component of fluidized beds not only for supporting the bed material but also to promote a uniform gas distribution through the solid material. A proper design of the grid plate is, therefore, very important from an operational as well as an economic point of view.

Wormsbecker *et al.* (2007) made experiments in order to verify the influence of bed loading and superficial gas velocities in a fluidized bed dryer performance using three types of gas distributors. Their results showed that the distributor design affected the drying time and temperature profile.

Sobrino *et al.* (2009) studied the distributor plate effects on the hydrodynamics characteristics of turbulent fluidized beds through pressure and radial voidage profiles analysis. They made experiments using bubble-cap and perforated plate distributors. Their results also showed the influence of the gas distributor design on the fluidized bed structure.

Zhu *et al.* (2010) verified the influence of gas distributor design on the gas-solids flow structure in a CFB riser. The results showed that the distributor design had significant effects on the solids distribution by measuring the solids concentration at different lateral and axial positions using a multi-fibre optical probe. They concluded that the flow structure in the entrance region is critical to the overall particles mixture in the fluidized bed and more research is needed about this subject.

The design of the gas distributor is still a challenge for equipment design for pyrolysis, gasification and combustion processes whose fluidized beds are usually composed by an inert material (sand or ash) and a fuel (coal, biomass, oil shale or waste residues). Problems like erosion, non-uniform distribution, bed collapse and segregation are usually present, mainly because of low particle density and irregular shape of biomass particles (Karri and Werther, 2003; Zhu, 2010). Currently, sugar cane bagasse is the most interesting biomass for energy production in Brazil, representing almost 74% of total energy generated from biomass and about 3% of the total energy capacity of the country.

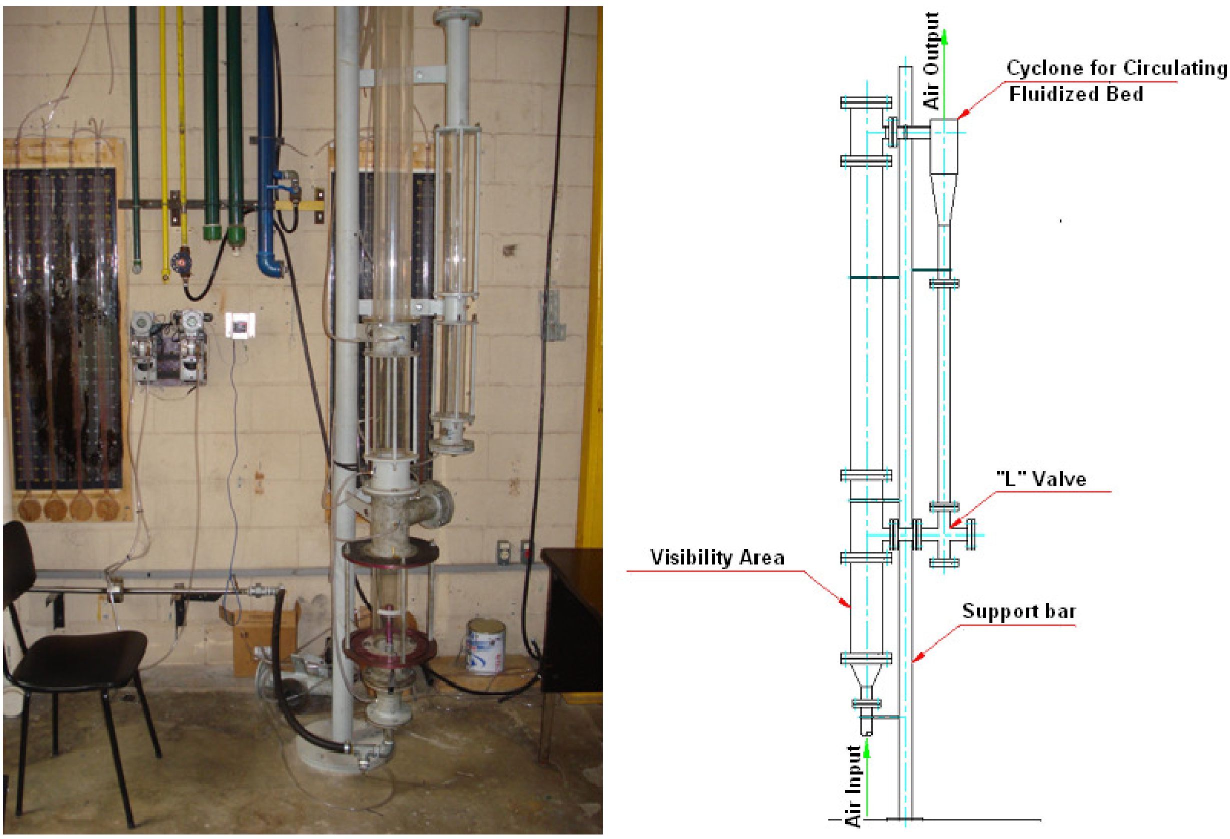
Concerning fluidized beds applied to mixtures of solid particles with different diameters and density, a good mixture between the particles must be observed in order to obtain a good reactor performance (Zhang *et al*., 2009). However, few works can be found in literature involving fluidization behavior of mixtures of bagasse and inert material. Rasul and Rudolph (2000) studied the fluidization of bagasse-pumice and bagasse-FCC mixtures. These authors proposed a mixture and segregation model on bagasse fluidization that allows obtaining a good operational condition for industrial processes using this material.

This work aims to contribute in understanding the fluidization behavior of bagasse-sand mixtures, through experimental studies in a fluidized bed system using three different tuyere (or nozzle) type distributors, with orifices diameters of 3, 4 and 5 mm, in order to verify the influence of this parameter on fluidization quality.

2. MATERIAL AND METHODS

2.1. Experimental system

A laboratory-scale fluidized bed was strategically constructed to operate in different fluidization regimes, since fixed bed to fast fluidized bed. It is composed meanly by plenum, gas distributor (3 tuyere per plate with 8 orifices each, 0.0254 m tube diameter), riser (0.100 m internal diameter, 0.10m height), cyclone and solid circulation valve (L type). A centrifugal fan supplies the air to the system and an orifice meter provides the measurement of the air flow rate into the system. Air temperature is measured by a thermocouple (J-type) located between the orifice meter and the plenum. Figure 1(a) shows the photography of the experimental system and Fig. 1(b) shows a schematic view of the main loop. Three different tuyere type gas distributors were constructed using orifices diameters of 3, 4 and 5 mm. The system was made of glass, acrylic and carbon steel sections allowing visual observation of the gas-solid flow.



(a) (b)

Figure 1. Experimental System

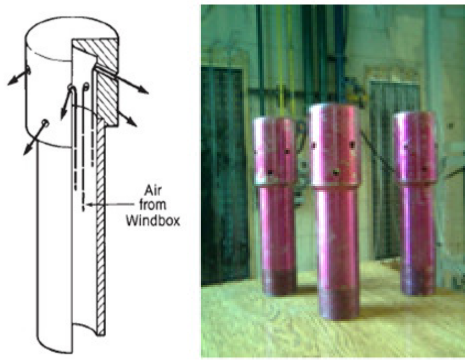


Figure 2. Tuyeres or nozzles type gas distributor

The tuyeres distributor, Fig. 2, were dimensioned to prevent particles agglomeration (Basu, 1984), and were based on design suggestions by Karri and Werther (2003) and Whitehead (1971).

2.2. Solid particles

Tests were made using silica sand and sugar cane bagasse as solid material. Mean particle diameter and particles density for each material were determined using sieve analysis according to ABNT NBR 6508 and picnometry, respectively. The sugar cane bagasse, before drying, presented 13% fiber and 51% moisture content (wet basis). Tests were conducted after drying the sugar bagasse to 0.9 % moisture content (wet basis). Table 1 shows the properties of the solid materials used in the experiments. Both particles were classified as group B concerning Geldart (1973) classification.

Table 1. Mean Sauter diameter (dp) and particles density (ρ_p) of sand and sugar bagasse

Material	dp (10^{-6} m)	ρ_p (kg/m^3)
Sugar bagasse	1501	165
Silica sand	313	2525

The average particle diameter (d_m) and average particle density (ρ_m) of the binary-solid fluidized bed were calculated according to Gibilaro *et al.* (1986) by Eq. (1) and (2), respectively.

2.3. Methodology

Fluidization tests were carried out to each nozzle configuration, at four mixture mass ratio between sugar cane bagasse and sand (M_A/M_B) in the range from 0 to 0.4%. The mean diameter (d_m) and the particle density (ρ_m) for the bagasse/sand mixture were obtained from Rasul and Rudolph (2000) correlations indicated by Eqs. (1) and (2).

The minimum fluidization velocity for each mixture mass ratio was obtained experimentally from bed pressure drop measurements as a function of superficial gas velocity (U_o).

$$d_m = \frac{d_A.d_B.(\alpha_A + \alpha_B)}{\alpha_A.d_B + \alpha_B.d_A} \tag{1}$$

$$\rho_m = \varepsilon.\rho_g + \alpha_A.\rho_A + \alpha_B.\rho_B \tag{2}$$

Where d_A, d_B and α_A, α_B are the mean diameter and volume fraction of solid species A (sugar cane bagasse) and B (sand), respectively, ε is the bed voidage calculated by Eq. (3) and ρ_g and μ_g are the gas density and gas dynamic viscosity, respectively.

$$\varepsilon = 1 - \alpha_A - \alpha_B \tag{3}$$

Three tests were done for each operational condition in order to reduce experimental uncertainties.

Transition velocities between fluidization regimes were calculated from literature correlations using mean diameter and the particle density of the bagasse/sand mixture in order to compare with experimental results.

The minimum fluidization velocity (U_{mf}) was determined using Wen and Yu (1966) recommended equation as showed by Eq. (4).

$$\mathrm{U_{mf}} = \frac{\mu_g}{\rho_g . d_m} . \sqrt{33.7^2 + 0.0408.Ar - 33.7} \tag{4}$$

Where Ar is the Arquimedes number given by Eq. (5).

$$\mathrm{Ar} = \frac{\rho_g . d_m{}^3}{\mu_g{}^2} \left(\rho_m - \rho_g\right) . g \tag{5}$$

Where g is the acceleration of gravity.

The minimum slugging velocity (U_{ms}) was determined from the Eq. (6) from Stewart and Davidson (1967):

$$\mathrm{U_{ms}} = \mathrm{U_{mf}} + 0.07.\sqrt{g.d_m} \tag{6}$$

The maximum velocity of slugging (U_c) and velocity to transition to turbulent regime (U_k) was determined by Eq. (7) and (8) given by Horio (1990):

$$\mathrm{U_c} = \frac{\mu_g . 0.936.Ar^{0.472}}{\rho_g . d_m} \tag{7}$$

$$\mathrm{U_k} = \frac{\mu_g . 1.46.Ar^{0.472}}{\rho_g . d_m} \tag{8}$$

3. RESULTS AND DISCUSSION

Figure 3 shows some photos of the sand-bagasse mixture from the fixed bed regime to the bubling fluidized bed regime. At low gas flow rate, the particles of sugar bagasse were located at the top of the fixed bed as it was still observed at minimum fluidized bed condition. The complete mixture between sugar cane bagasse and inert material was only observed at bubbling fluidized bed regime for superficial velocities around twice U_{mf} for all tested orifice diameter.

(a) fixed bed (b)minimum fluidization condition (c) bubling fluidization regime

Figure 3. Fluidization regimes at bed containing the bagasse/sand mixture

Table 2 shows the transition velocities calculated from Eqs. (4) to (8) using Rasul and Rudolph (2000) correlations for mean particle characterization inside the bed and the experimental values obtained for minimum fluidization velocity ($U_{mf,exp}$) and for the mixture velocity ($U_{mix,exp}$) which represents the gas superficial velocity necessary to the complete mixture between solids particles into the bed as shown in Fig. 1 (c).

Table 2. Transition velocities

$M_A .10^3$ (kg)	M_A/M_B (%)	α_A (%)	$U_{mf,exp}$ (m/s)	$U_{mix,exp}$ (m/s)	U_{mf} (m/s)	U_{ms} (m/s)	U_c (m/s)	U_k (m/s)
0	0	0	0.05	-	0.0765	0.0804	1.91	2.98
4.69	0.17	4.0	0.05	0.0896	0.0776	0.0815	1.90	2.96
9.81	0.25	8.4	0.05	0.105	0.0781	0.0820	1.89	2.95
15.9	0.39	13.6	0.05	0.107	0.0785	0.0824	1.89	2.94

Figure 4 shows the pressure drop for each tested nozzle as a function of superficial gas velocity. It can be observed that there is no difference between the results obtained for 4 and 5 mm orifice diameter. As expected, the distributor presenting the smallest orifice diameter produced the biggest pressure drop.

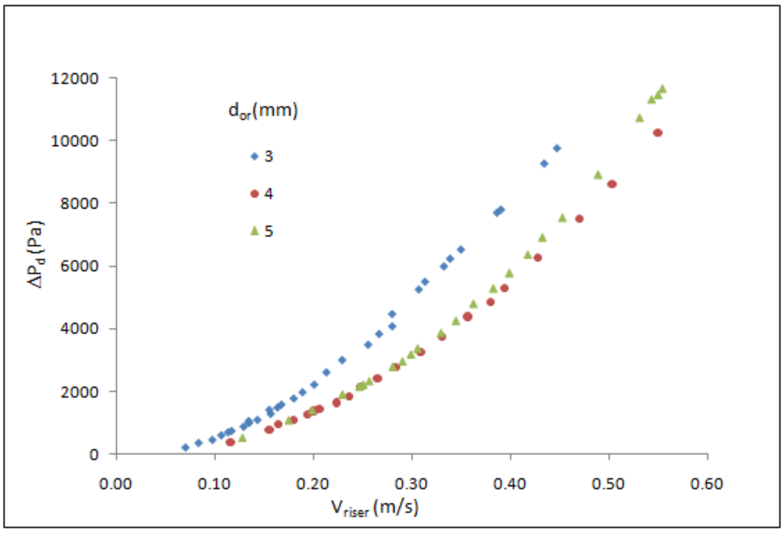


Figure 4. Gas distributor pressure drop (ΔP_d) for tested orifices diameters (no solid material inside the riser).

The nozzles were submitted to a test of friction and reflux of material at room temperature. After two hours of operation at 0.303 m/s gas superficial velocity in the riser, it was observed that all tested nozzles showed intense reflux of particles causing accumulation of material in the wind box. Another important factor observed in this test was the major wear at top of the nozzle, as shown in Fig. 5. As a result we verified the importance of using materials more resistant to abrasion in this region.

(a) Before the operation

(b) After the operation

Figure 5. Friction at distributor after two hours operation.

Figures 6 to 8 show the pressure drop between the wind box and the top of the riser which includes the distributor and the bed pressure drop ($\Delta P_{d+riser}$) as a function of superficial gas velocity for the three orifices diameters studied. Results were obtained for different mass of bagasse inside the bed as indicated in Tab. 2.

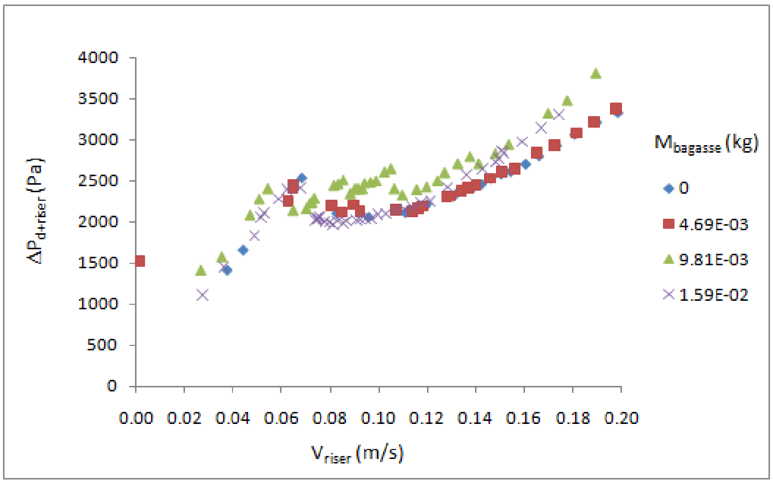


Figure 6. Bed pressure drop plus distributor pressure drop ($\Delta P_{d+riser}$) for 3 mm orifice diameter at the nozzle

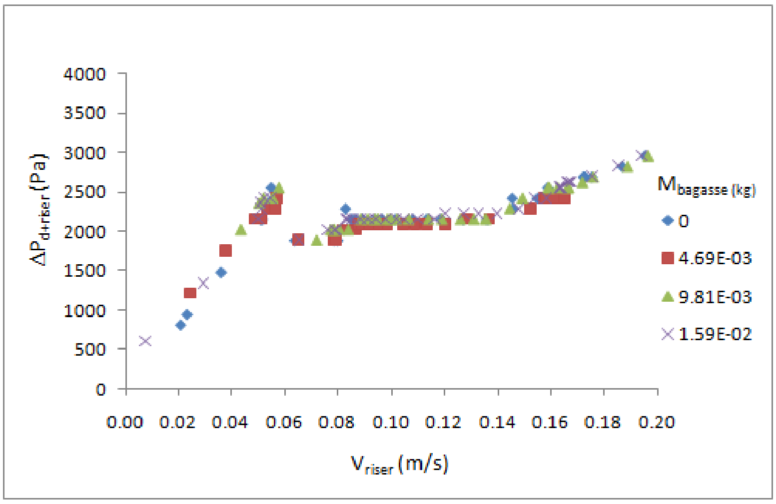


Figure 7. Bed pressure drop plus distributor pressure drop ($\Delta P_{d+riser}$) for 4 mm orifice diameter at the nozzle

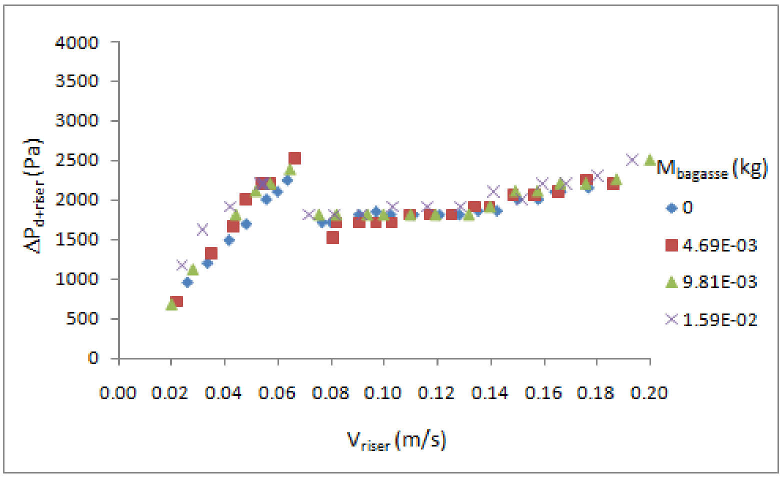


Figure 8. Bed pressure drop plus distributor pressure drop ($\Delta P_{d+riser}$) for 5 mm orifice diameter at the nozzle

The minimum fluidization velocity ($U_{mf,exp}$) for each mixture mass ratio was obtained experimentally from Figs. 6 to 8 as recommended by Kunii and Levenspiel (1991). It can be observed that the bagasse load and the orifice diameter do not affect the minimum fluidization velocity and values around 0.05 m/s were obtained. However, visual observation showed that the effective particles mixture only occurred at higher gas velocities ($U_{mix,exp}$), depending on bagasse load.

Figures 6 to 8 also show a pressure drop increase at gas velocities around 0.12 m/s due to the increment of the distributor pressure drop contribution on the measured value, as expected.

Results from Eq. (4) showed that minimum fluidization velocities (U_{mf}) were around 0.08 m/s, for all mixtures compositions and orifices diameter of the distributor, which are bigger than the experimental values obtained. These results show that Eqs. (1) to (4) must be revised. Also, the beginning of bed homogeneity occurred at velocities near the predicted minimum slugging velocity (U_{ms}) showed in Tab. 2.

4. CONCLUSIONS

The minimum fluidization velocity was not affected by the bagasse load or orifice diameter of the tuyere gas distributor for the tested bagasse-sand mixtures. Application of Eqs. (1) to (4) to minimum fluidization velocity prediction is not recommended for bagasse-sand mixtures.

The studied tuyere configurations were able to fluidize sand-bagasse mixtures at velocities around 0.05 m/s and the bed uniformity was reached for velocities in the range of 0.09 to 0.10 m/s, depending on the biomass load.

The studied fluidized beds showed a desired bagasse-sand mixture and bed uniformity at superficial gas velocities around the minimum slugging velocity. This information is very important for fluidized bed boiler designers as it can be used to estimate the operation range of air mass flow rate in industrial equipment.

All tested nozzles distributors, after two hours of operation, presented erosion due to attrition with sand. This fact could be solved using special materials in the nozzle manufacturing process.

The presence of sugar cane bagasse in a bed containing sand particles increases the mean particle diameter of the bed material and decreases the bed density.

5. ACKNOWLEDGEMENTS

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