COBEM 99
XV Congresso Brasileiro de Engenharia Mecânica
15th Brazilian Congress of Mechanical Engineering
22 - 26 de Novembro de 1999 / November 22 - 26, 1999 Águas de Lindóia, São Paulo,

PRESSURE DROP BEHAVIOR INSIDE AN ATMOSPHERIC FLUIDIZED BED, FROM MINIMUM TO INCIPIENT FAST FLUIDIZATION, USING DIFFERENT DISTRIBUTOR PLATES

Paiva, João Monney
Dep de Engenharia Mecânica e Gestão Industrial
Esc Sup Tec Inst Politec Viseu
Campus Politécnico 3500 Viseu Portugal
Pinho, Carlos
CEFT- Dep de Engenharia Mecânica e Gestão Industrial
Faculdade de Engenharia da Universidade do Porto
Rua dos Bragas 4099 Porto Portugal
Figueiredo, Rui
Dep de Engenharia Mecânica
Faculdade de Ciências e Tecnologia da Universidade de Coimbra
Largo D. Dinis 3000 Coimbra Portugal

Abstract. *Experiments were performed to study the pressure drop behavior inside an atmospheric vertical fluidized bed with 100 mm internal diameter, divided into 5 sections of 50 mm high each. The fluidized bed under analysis is established by fluidizing two size ranges of silica ballotinis, 355-425 and 600-710 µm, using three distributors types: perspex, metallic mesh and ceramics. The classical force balance equations were modified to account not only for the weight and buoyancy but also for the presence of bubbles and drag on clusters of particles. Different dense phase voidages, as well as different ratios of volumetric flow rate through the bubble phase versus volumetric flow rate through an equivalent dense phase section were considered. Finally, a model is presented covering a range of fluidization regimes for several slices inside the fluidized bed, from minimum fluidization to incipient fast fluidization.*

Keywords: *Fluidization, Fluidized bed, Pressure drop.*

1. INTRODUCTION

In a recent work, Paiva *et al.* (1998), using a two-phase model (Davidson, 1963) for the fluidized bed, recognized a decrease of the *ratio* pressure drop/height of the bed, in opposition to the generally accepted idea of a constant value of this ratio. That behavior explicitly reflects the increase in the average global porosity of the whole bed, with the augmentation of the mass flow rate of the fluidizing medium. In order to have a better understanding of the phenomenon, further studies were developed to lounch some physical insight on the apparent drag reduction recorded.

Using different dense phase voidages throughout the height of the bed, a bubble's voidage deduced from the two-phase theory, with the bubble growth explicitly incorporated, as well as a drag coefficient accounting for the dissipation energy on the incipiently entrained clusters of particles, a model is presented that covers a range of fluidization regimes for the slices inside the bed, from minimum to incipient fast fluidization state.

Finally, the fluid dynamics of the bottom of the bed being of interest for both Stationary and Circulating Fluidized Beds, the pressure drop across the air distributor was measured, using three different types- perspex/acrylic glass perforated plate, metallic mesh and ceramics, to give some insight into fluidization quality of that region of the bed.

2. EXPERIMENTAL

The experimental set up consisted of a fluidized bed made with a perspex tube 100 mm id, 3 mm thick and 500 mm high, as can be observed in Fig. 1.

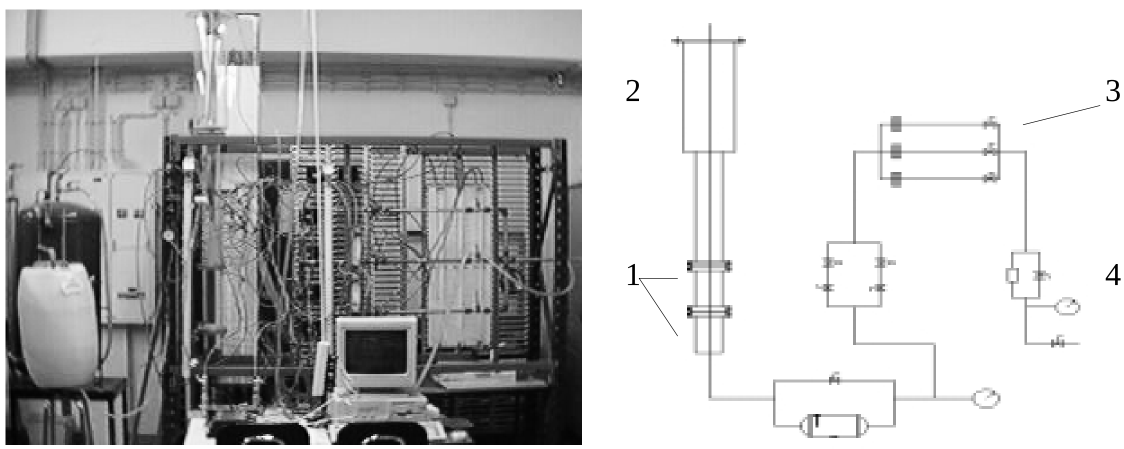


Figure 1- Photography and schematics of the experimental setup: 1- primary beds; 2- transport disengaging zone; 3- thin- plate battery; 4- pressure reduction.

Air was used as the fluidizing medium, at atmospheric conditions, its flow rate being measured with a battery of three thin-plate orifice flowmeters. Differential pressure drop measurements on the orifice plates were sent to a data acquisition system.

The experiments were all carried out with spherical glass ballotinis previously screened and statistically weighed with *t* tests using SPSS, at the 95% confidence level, in the ranges of 355-425 and 600-710 μm. The density of the particles was measured and found to have an average value of 2500 kg/m^3. Fluidizing air superficial velocities were in the 0- 1.4 m/s range, at standard pressure and temperature. The static bed height was of 250 mm above the distributor plate. The distributors used were a perspex plate, perforated with 984, 0.3 mm equally spaced holes, a metallic mesh plate, and a ceramic plate. Their pressure drop measurements can be seen in Fig. 2. To help aiming at an uniform flow distribution, two consecutive primary fixed beds, the first filled with 20 mm and the second with 3 mm diameter silica spheres, were used in the windbox section of the air duct, as can bee seen from Fig. 1 and explained elsewhere (Paiva *et al.*, 1998).

Pressure drop accumulated measurements were made every 50 mm, in order to have slices of 50, 100, 150, 200 and 250 mm high. Three pressure probes were used at each level, set 120º apart, starting just above the distributor plate. Each probe was calibrated against U-tube water pressure gauges. As the probes have a fixed position as bed expands with flow augmentation, the data presented for each bed slice's pressure drop is actually a measure of the global voidage increase of that portion of the bed.

All the experiments were conducted starting with high air flows and after having the bed well fluidized. Then, values for each slice's pressure drop were recorded for consecutively decreasing flowrates, corresponding to U_0/U_{mf} ratios from 5 to 1. The data acquisition system used a sampling frequency of 5 Hz to ensure sufficient accuracy in the statistical analysis, an average of 1000 samples being taken for each spectrum; those readings were then weighed in order to output values corresponding to arithmetically averaged one second intervals. Later on, using a suitable program, these were determined for each position of the flowmeter and the dubious points eliminated according to Chauvenet's criterion (Holman, 1994).

3. RESULTS AND DISCUSSION

As stated on common literature on fluidization, there is a point of minimum fluidization when a force balance leads to the following equation (Kuni and Levenspiel, 1969),

$$\Delta P = (\rho_p - \rho_f)\,(1 - \varepsilon_{mf})\; x_{mf}\; g \qquad (1)$$

in which x_{mf} is the bed's height at minimum fluidization conditions, ε_{mf} the bed voidage at that point and g the acceleration of gravity.

Assuming that the dense phase remains at minimum fluidization conditions, according to the two-phase theory (Davidson and Harrison, 1963), any increase in flow rate beyond that point leads to,

$$\Delta P = (\rho_p - \rho_f)\,(1 - \varepsilon_{mf})\; x_f\; g \qquad (2)$$

with x_f being the bed's height attained in a particular regime, measured from the distributor plate till the free surface of the bed. This equality is usually considered to constitute the fundamental fluidization condition (Couderc, 1985).

One main purpose of many fluidization studies is to understand the behavior of rising bubbles in gas fluidized beds using ordinary distributors. The relative velocity between the dense phase and bubbles should be the rise velocity of single bubbles in beds at minimum fluidizing conditions (Davidson and Harrison, 1963, Davidson *et al.*, 1977), or

$$U_{br} = K' \sqrt{g D_b} \qquad (3)$$

As one can see from the published correlations, namely Rowe (1972), Werther (1976), Yacono (1975), Yasui and Johanson (1958), Park *et al.* (1969), Geldart (1972) and Darton *et al.* (1977), estimating D_b, the equivalent spherical diameter of bubble, is presented, as a general rule, in the form

$$D_b = K\,(U_0 - U_{mf})^{n'} \qquad (4)$$

K being a numerical constant depending on the height of the bed, the particle diameter and/or the number of holes in the distributor plate, and n' a numerical constant. Combining Eqs. (3) and (4), allows expressing U_{br} in a general form:

$$U_{br} = k\,(U_0 - U_{mf})^{n} \qquad (5)$$

again with k and n as numerical constants. Then, the absolute rise velocity of bubbles, U_{ba}, in common bubbling beds (Kunii and Levenspiel, 1969) shall be given by

$$U_{ba} = (U_0 - U_{mf}) + k\,(U_0 - U_{mf})^n \tag{6}$$

On a superficial velocity basis the upward flow of gas in the dense phase is U_{mf} while through the bubble phase is $U_{ba} + \beta.U_{mf}$, (Davidson and Harrison, 1963). Thus, the total gas flow in the bed can be expressed as

$$U_0 = (1 - \delta)\,U_{mf} + \delta\,(U_{ba} + \beta\,U_{mf}) \tag{7}$$

where β is a factor representing the number of times the bubbles are processing the amount of gas passing by an equivalent section of the dense phase (Kunii and Levenspiel, 1969).

The expression of the volume fraction of bubbles in a single slice of the bed, $(V_{bubbles}/V_{bed})_i$, is then:

$$\delta = \frac{U_0 - U_{mfi}}{k\left(U_0 - U_{mfi}\right)^n + U_0 + \left(\beta - 2\right)U_{mfi}} \tag{8}$$

where U_{mfi} is the minimum fluidization velocity for the slice i, n takes a value of 0.8 and k is a function of x, as deduced from Darton *et al.*(1977) work, in the form

$$k = 0.384\,g^{0.3}\left(x + 2\sqrt{\frac{\pi D^2}{N_o}}\right)^{0.4} \tag{9}$$

In the above equation D is the bed diameter and N_o is the plate's number of orifices. Introducing the pressure drop at minimum fluidization conditions, again for a certain slice i,

$$\Sigma\Delta P_{mfi} = (\rho_p - \rho_f)\,(1 - \varepsilon_{mf})\;x_i\;g \tag{10}$$

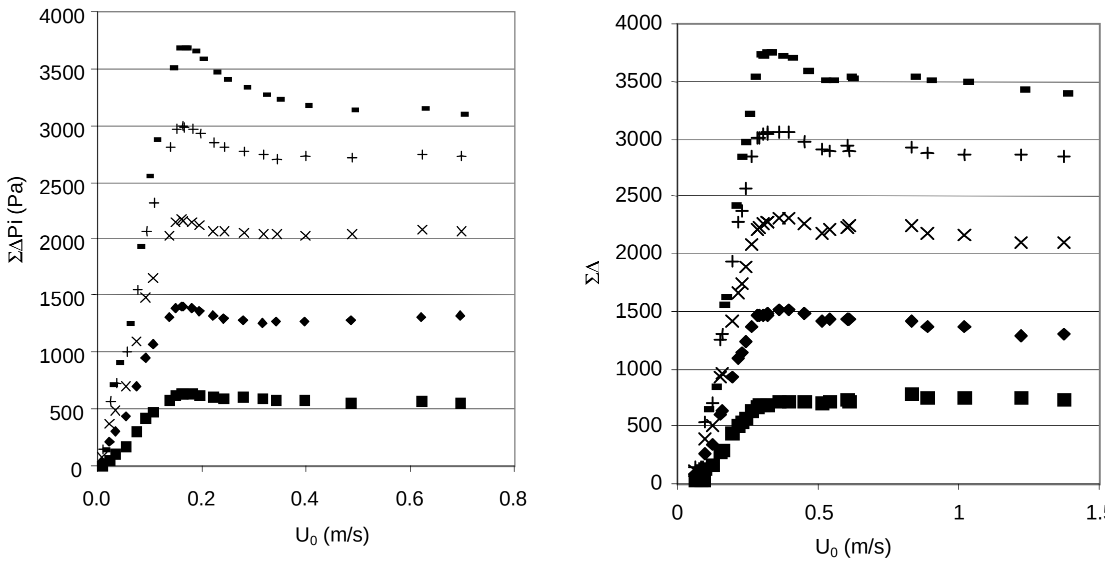
where x_i is the distance from the distributor plate to the slice under analysis. Equation (2) can be adapted so that

$$\Sigma\Delta P_i = \Sigma\Delta P_{mfi}\left(1 - \delta\right) \tag{11}$$

It is therefore no longer suitable to account for the pressure drop due to the weight of particles in a certain section of the bed solely on a volume basis corrected by the dense phase voidage, when fluidization states pass from incipient to bubbling regimes. The effect of bubbles on the overall voidage should be accounted for.

Typical quantitative data, representing accumulated slice's pressure drop, $\Sigma\Delta P_i$, *versus* superficial velocity values, U_0, are presented, regardingg two size ranges: 355-425 μm and 600-710 μm, Figs. 2 (a) and (b).

As can be noticed, pressure drop through slices of constant height reveals a decreasing value as mass flow rate goes beyond the minimum fluidization point. Two factors influence cumulatively on this behavior: one is the referred leaning of particles in the slices and the other is the incipient entrainment of solids, as velocities increasingly approach 2 ... $3.U_{mf}$.



(a) (b)

Figure 2- Accumulated pressure drop per slice *versus* superficial velocity, ballotinis 355-425 μm (a) and 600-710 μm (b).
(slices: ■: 1, ♦: 1+2, x: 1+2+3, +: 1+2+3+4, -: 1+2+3+4+5)

From literature, the transition from bubbling to turbulent fluidization is gradual and occurs over a wide range of gas velocities, depending either on its properties or on equipment scales and it has been the subject of a number of investigations (Lanneau (1960), Kehoe and Davidson, 1971, Massimilla, 1973, Yerushalmi *et al.*, 1976, 1978, Cankurt and Yerushalmi, 1978, Yerushalmi and Cankurt, 1978, Turner, 1978, Avidan and Yerushalmi, 1982).

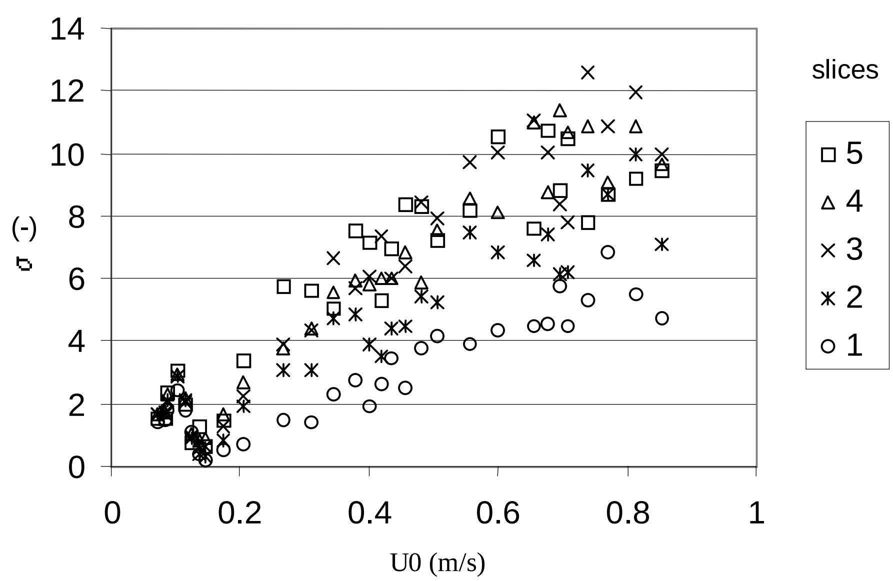


Figure 3- Standard mean deviation of pressure drop measurements, σ, *versus* superficial air velocity U_0, 355-425 μm.

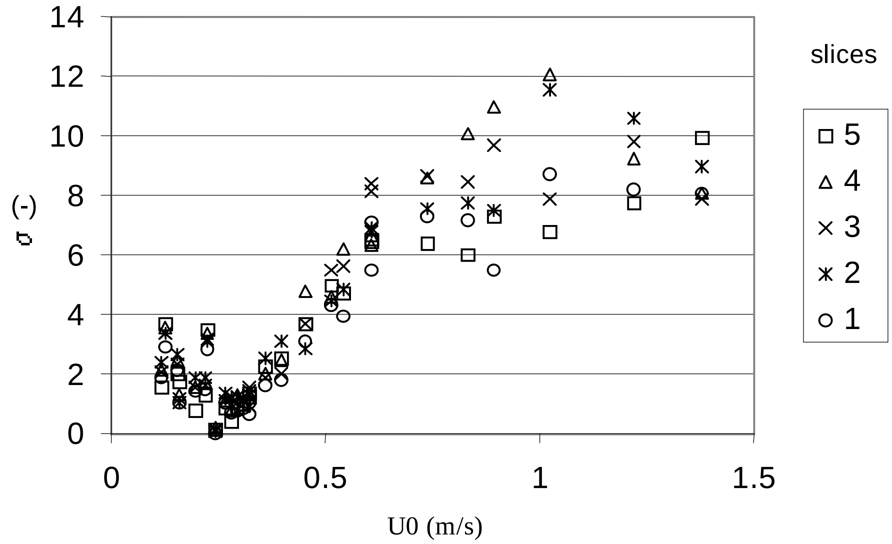


Figure 4- Standard mean deviation of pressure drop measurements, σ, *versus* superficial air velocity U_0, 600-710 μm.

Yerushalmi and Avidan (1985) characterize the transition to turbulent fluidization by velocities U_c– the velocity at which amplitude of pressure fluctuations peak, and U_k– the velocity at which amplitude of pressure fluctuations level off, marking the onset and end of the transition, over a significant range of values. Particularly, Canada *et al.* (1976), and Yerushalmi *et al.* (1976, 1978), who studied glass spheres with densities from 2420 to 2480 kg/m^3, pressures ranging from 1 to 10 atm and mean diameters of 157, 650 and 2600 μm, report a large span of U_c, U_k and U_k/U_t values, U_t being the particle's terminal velocity.

From Figs. 3 and 4, representing the variation of standard mean deviation of pressure drop measurements, σ, one seems to recognize a peak either in sizes 355-425 as in 600-710 μm, around 0.75 and 1 m/s superficial gas velocity, respectively.

Recently, Bi and Grace (1995) proposed a unified flow regime diagram in gas-solid fluidized beds with little or no overflow of solids, supported on experimental findings from Grace (1986), Bi *et al.* (1993, 1995) and Bi and Grace (1994, 1995), based in $Ar^{1/3}$ and $Re/Ar^{1/3}$ (where $Re = \rho_f U_0 d_p/\mu_f$ and $Ar = \rho_f(\rho_p-\rho_f) g d_p^3/\mu_f$ are the Reynolds number and Archimedes number, respectively, with ρ_f and ρ_p being the gas and particle density and μ_f the gas viscosity). In such diagram, the present studied cases lie close to the border line of definite bubbling and turbulent zones, $Ar^{1/3}$ values varying in the range of 0.44 to 0.75, enforcing the proposed relation between fixed points measured pressure drop decrease and incipient entrainment.

The overall bed pressure drop must thus be modeled through a balance of forces, by means of the cumulative effect of weight minus buoyancy- $\Delta P_{W\text{-}B}$, plus the incipient entrainment- ΔP_{IT}:

$$\Delta P_T = \Delta P_{W\text{-}B} + \Delta P_{IT} \qquad (12)$$

The first term shall be affected by the increase of fractional void of bubbles in the bed's slices, as stated in Eq. (11); as to the second term, a coefficient C_{Dc} must account for the drag effect on clusters of particles, as referred by Klinzing (1981). Then:

$$\Delta P_T = (\rho_p - \rho_f)(1-\varepsilon_{mf})g(1\text{-}\delta)x + C_{D_c}\frac{1}{2}\rho_F (U_f - U_p)^2 N_p \qquad (13)$$

U_f and U_p being the interstitial and particle velocities, respectively, $U_f – U_p$ standing for the slipping velocity of the incipiently entrained particles. N_p is the number of particles defined by the ratio $(V_{particles}/V_{1particle})_i$, in a single slice of the bed.

Combining Eqs. (1), (8) and (13), and putting it in differential form, one can yield:

$$\frac{dP}{dx} = \left[(\rho_p - \rho_f) g + \frac{3}{4} \lambda \left(\frac{U_0}{\varepsilon_{mf}} - U_p \right)^2 \rho_f \frac{D^2}{d_p^{\,3}} \right] (1 - \varepsilon_{mf})(1 - \delta) \tag{14}$$

which can be numerically integrated within each slice's limits, taking λ as a factor that represents the drag on the particle clusters, and assuming that the slipping velocity is, roughly, the terminal settling velocity of a single particle; the consideration for the neighboring particles effects is done through a corrected drag coefficient $C_D = C_{Ds} \varepsilon_{mf}^{\ -c}$, as suggested by Klinzing (1981).

The result of applying Eq. (14) to the experiments that conducted to the data expressed in Fig. 2 is shown by the solid lines in Figs. 5 and 6.

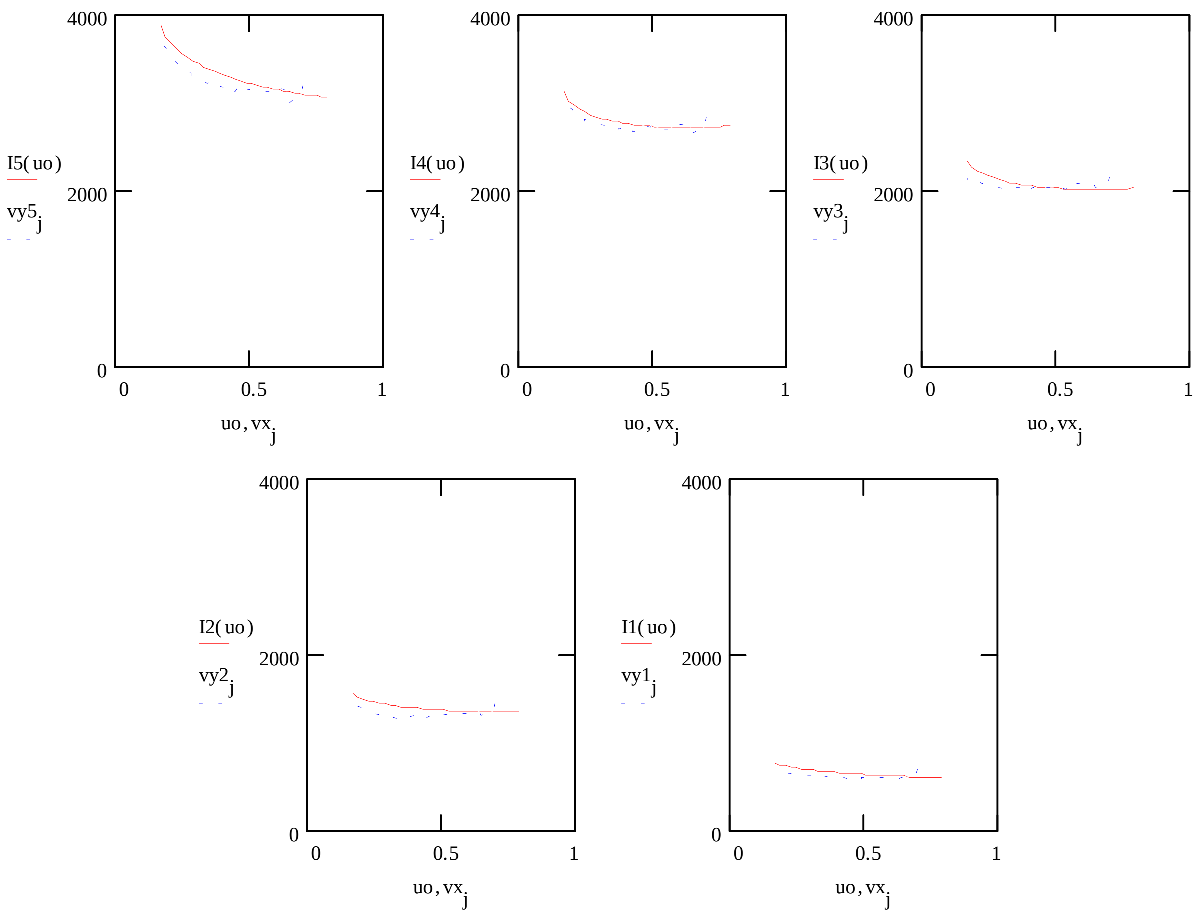


Figure 5- Curve fitting (solid lines) according to Eq. (14), ballotinis 355-425 µm. ($I1..5(u_0)$- integrated values of Eq. (14), *i. e.*, integrated values of velocity against bed height, for slices 1, 1+2,...,1+2+3+4+5; $u_0 \equiv U_0$; $vy1..5_j$- measured pressure drop *versus* vx_j- measured superficial velocity– dashed points)

In general there is a good correlation between experimental data and values obtained from Eq. (14), exception made for the first slice where the presence of jets, flowing out the distributor orifices, hides the global behavior of the pressure drop.

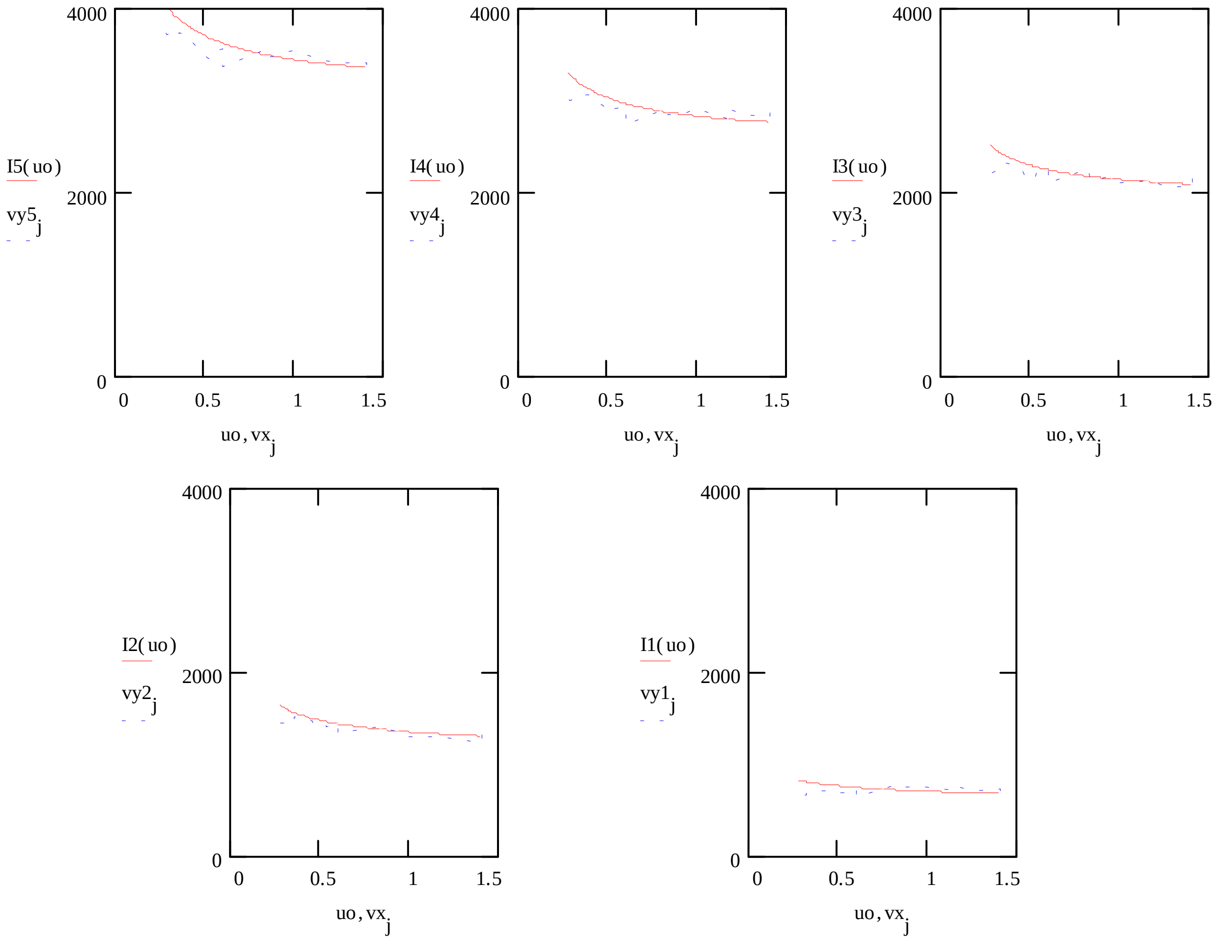


Figure 6- Curve fitting (solid lines) according to Eq. (14), ballotinis 600-710 µm. ($I1..5(u_0)$- integrated values of Eq. (14), *i. e.*, integrated values of velocity against bed height, for slices 1, 1+2,...,1+2+3+4+5; $u_0 \equiv U_0$; $vy1..5_j$- measured pressure drop *versus* vx_j- measured superficial velocity– dashed points)

Literature (Lewis, 1949, Shanon, 1961, Wen, 1966 e Wilhelm, 1948) refers to ε_{mf} experimental data varying from 0.36 to 0.46 for spherical particles. Dense phase voidages varying with bed's height used in Eq. (14) are given in Table 1. The *c* exponent, affecting the drag coefficient by means of the dense phase voidage ε_{mf}, takes a value of 0.3.

Table 1. Parameters used in Eq. (14)

		1st slice	*1st+2nd*	*1st+..+3rd*	*1st+..+4th*	*1st+..+5th*
	xi	0.05	0.10	0.15	0.20	0.25
355-425	ε_{mf}	0.41	0.41	0.42	0.42	0.42
(µm)	β	3	4	3	3	3
	$\lambda.10^5$	0.7	0.7	0.8	0.8	1.1
600–710	ε_{mf}	0.40	0.42	0.42	0.42	0.42
(µm)	β	4	3	3	3	3
	$\lambda.10^5$	0.9	1	1.1	1.1	1

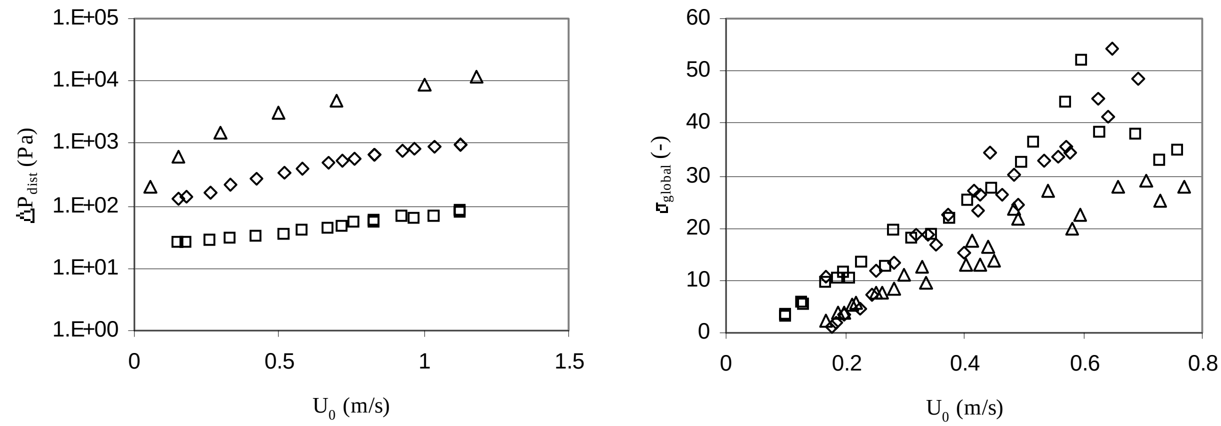


Figure 7- Pressure drop (a) and standard mean deviation (b) values for perspex, perforated with 984, 0.3 mm equally spaced holes (◊), metallic mesh (∇), and ceramic plate( ).

Though quite different in pressure drop values, as can be seen in Fig. 7 (a), the distributors didn't affect significantly bed overall behavior. Still, there is an influence in registered standard mean deviation of pressure drop measurements, Fig. 7 (b). No explanation for this discrepancy having been found yet, it will be the object of future works.

4. CONCLUSIONS

This model represents a step further in the physical insight regarding the apparent pressure loss reduction in bubbling fluidized beds beyond minimum fluidization point, as was previously presented by Paiva *et al.* (1998). Recognizing the need for a relation that would account for the combined influence of the particle diameter, the height of the bed considered for analysis and the regimes developed in the bed, there was a comprehensive switch from a crude initial approach towards a better understanding of involved physical phenomena, with a focus on entrainment of clusters of particles. For superficial velocities values of 0.3...0.5 m/s, typical of terminal velocities of particles in the range 300...800 μm, a drag coefficient λ was defined, accounting for both the incipient entrainment and the effects of neighboring particles. Different ε_{mf} dense phase voidages, as well as different β ratios, were considered for each slice of the bed.

REFERENCES

Avidan, A. A. and Yerushalmi, J., 1982, Powder Technol., 32, 223.

Bi, H. T. and Grace, J. R., 1994, Transition from bubbling to turbulent fluidization, AIChE Annual Meeting, San Francisco, 13-18 November.

Bi, H. T. and Grace, J. R., 1995, Effects of measurements methods on velocities used to demarcate the transition to turbulent fluidization, Chem. Engng. J. 57, 261-271.

Bi, H. T. and Grace, J. R., 1995, Flow regime diagrams for gas-solid fluidization and upward transport, Int. J. Multiphase Flow, Vol. 21, No 6, pp. 1229-1236.

Bi, H. T., Grace, J. R. and Zhu, J. X., 1993, On types of choking in pneumatic systems, Int. J. Multiphase Flow 19, 1077-1092.

Bi, H. T., Grace, J. R. and Zhu, J. X., 1995, Regime transitions affecting gas-solids suspensions and fluidized beds, Chem. Engng. Res. Des. 73, 154-161.

Cankurt, N. T. and Yerushalmi, J., 1978, Gas backmixing in high velocity fluidized beds, in Fluidization, eds Davidson, J. F. and Keairns, D. L., p. 387, Cambridge University Press, London and New York.

Darton, R. C., LaNauze, R. D., Davidson, J. F. and Harrison, D., 1977, Trans IchemE, **55**, 274.

Davidson, J. F. and Harrison, D., 1963, Fluidised Particles, Cambridge University Press.

Grace, J. R., 1986, Contacting modes and behavior classification of gas-solid and other two-phase suspensions, Can. J. Chem. Engng. 64, 353-363.

Holman, J. P., 1994, Experimental Methods for Engineers, 6th edition, McGraw-Hill, Inc.

Kehoe, P. W. K. and Davidson, J. F., 1971, Continuously slugging fluidized beds, Inst. Chem. Eng. Symp. Ser., 33, 97.

Klinzing, G. E., 1981, Gas-Solid Transport, MacGraw-Hill Book Company.

Lanneau, K. P., 1960, Gas-solid contacting in fluidized beds, Trans. Inst. Chem. Eng., 38, 125.

Lewis, W. K., E. R. Gilliland and W. C. Bauer, 1949, Ind. Eng. Chem., 41, 1104.

Massimilla, L., 1973, Behavior of catalytic beds of fine particles at high gas velocities, AIChE Symp. Ser. 69, Nº 128, 11.

Paiva, J. M., Pinho, C. T. and Figueiredo, R. A., 1998, Evolution of pressure drop inside different layers of an atmospheric fluidized bed, Proceedings of the 7th Brazilian Congress of Engineering and Thermal Sciences, November 3-6, Rio de Janeiro, vol 2, pp. 827-831.

Wen, C. Y. and Y. H. Yu, 1966, Chem. Eng. Progr. Symposium Ser. No. 62, 62.

Wilhelm, R. H. and M. Kwauk, 1948, Chem. Eng. Progr., 44, 201.

Yang, W. C., 1978, AIChE Journal, vol. 24, n. 3, pp. 548-552.

Yerushalmi, J. and Avidan, A., 1985, High Velocity Fluidization, in Fluidization, eds Davidson, J. F., Clift, R. and Harrison, D., 2nd edition, Academic Press.

Yerushalmi, J. and Cankurt, N. T., 1978, High velocity fluid beds, Chemtec 8, 564.

Yerushalmi, J., Cankurt, N. T., Geldart, D. and Liss, B., 1978, Flow regimes in gas-solid contact systems, AIChE Symp. Ser. 74, Nº 176, 1.

Yerushalmi, J., Turner, D. H. and Squires, A. M., 1976, The fast fluidized bed, Ind. Eng. Chem. Process Des. Dev., 15, 47.