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FLUID DYNAMIC INVESTIGATION OF DILUTE AND DENSE PHASE FLOW IN A PNEUMATIC TUBE WITH A SPOUTED BED TYPE SOLID FEEDING SYSTEM

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Abstract. *The purpose of this work is to improve the characterization of the different conveying regimes for vertical pneumatic transport with a spouted bed feeder using the pressure gradient diagrams. The range of experimental data investigated were expanded by carrying out measurements of pressure gradients, mean voidages, and solid flow rates for two transport tube diameters (81.4 and 104.8 mm) and different particle diameters (from 0.41 to 1.70 mm). Two different transport regimes have been observed: a dilute and a dense phase flow. The validity of one-dimensional model to predict the pressure gradient on dilute and dense phase flow regimes was analyzed by comparing the predictions with experimental values. The results showed that the one-dimensional model is not sensitive to the changes in flow regimes detected in the pressure diagrams. Finally, a comparison of the specific power consumption at different operational conditions is presented. Although the estimated values of specific power consumption were high, it was not detected a reduction in these values for dense phase conditions.*

Keywords: Pneumatic conveying, Flow regimes, Spouted bed feeder

1. INTRODUCTION

Since the 19th century pneumatic conveying technology has been successfully employed in the industry for the transport of a large variety of particulate materials. Despite its versatility and flexibility for solids transport, the higher power consumption still represents a limiting aspect of pneumatic conveying systems. The efforts to reduce the power consumption led to the dense phase conveying systems, which operate at rather low gas velocities.

On dense phase conveying different flow structures may be reached depending not only on the fluid velocity but also on factors such as the particle and transport tube characteristics. These sub-regimes include the slugging flow, fast fluidized bed, non-slugging flow, plug flow and so on, and can cover a wide range of conditions (Marcus et al., 1990). The characteristics of each regime have been extensively described in literature (for a detailed review see Satija et al., 1985 or Leung, 1989). Since their descriptions are based essentially in qualitative observations, there is no universally accepted definition for the dense phase conveying. Besides, in comparison with the dilute phase, there are relatively few experimental data reported in literature. The lack of a clear definition and the restricted number of experimental data makes the analysis of this type of flow regime much more complex than the one of the dilute phase conveying

A investigation about flow regime transitions in a pneumatic bed with a spouted bed feeder has been carried out by Silva et al. (1997). These authors measured the pressure gradient for different gas velocities and observed a change in the behavior of pressure gradient vs. air velocity curves obtained. They have also reported that, on dilute-phase flow, the pressure gradient decreases as the air velocity is decreased until reaching a minimum value. Beyond this minimum, further reduction of the air velocity led to increase the pressure gradient and the intensity of pressure drop oscillation.

For constant solid flow rates, the relationship between pressure gradient and gas velocity is a criterion for identifying a change of flow regimes (Rizk, 1985). The minimum point of pressure gradient is considered to be the boundary between dense and dilute-phase conveying (Joseph and Klinzing, 1983; Marcus *et al*., 1990). The increase of pressure gradient when the air flow rates are reduced below the minimum point indicates that the pressure drop due to the solids hold-up predominates, instead of the wall friction forces which govern in a dilute flow. So the minimum air flow rate at which solids can be conveyed in the dilute phase regime is set by the air velocity at the minimum pressure drop in the state diagram, which will be referred to as the minimum velocity.

Silva et al. (1997) have demonstrated that these pressure gradient state diagrams can be applied for pneumatic transport with the spouted bed feeder, in spite of solid flow rates are dependent on air flow rates. By keeping the distance z_0 as a constant parameter instead of the solid flow rate, the pressure gradient versus air flow rate curves are similar to those obtained using a classical feeding device. Silva et al. 91997) have also verified that the criterion suggested by Rizk (1985) for identifying the air velocity at minimum pressure gradient conditions predicts well the experimental data obtained in the system with a spouted bed feeder. So they concluded that this criterion can be applied for identifying the minimum air velocity, assuming the existence of the dense phase flow at air velocity smaller than the minimum one.

However, in systems with annular feeders such as the spouted bed, experimental data on vertical dense phase conveying are scarce. Some data have been reported by Garic et al. (1995) and by Silva et al. (1997), but these cover a narrow range of operational conditions.

In this work we are interested in improving the characterization of the different conveying flow regimes observed in vertical pneumatic transport systems with a spouted bed feeder using the pressure gradient diagrams. The range of experimental data investigated is expanded by carrying out measurements of pressure gradients, mean voidages, and solid flow rates for two transport tube diameters and different particle diameters. The validity of one-dimensional model to predict the pressure gradient on dilute and dense phase flow is then analyzed by comparing the predicted data with experimental values. Finally, a comparison of specific power consumption at different operational conditions is presented.

2. EXPERIMENTAL METHODOLOGY

2.1 Experimental apparatus

The pneumatic conveyor with the spouted bed type feeder is shown in Figure 1. Spherical glass particles with ρ_p=2503 kg/m^3 and mean diameters, d_p varying from 0.41 to 1.70 mm, were transported in galvanized iron pipes, 4.0 m long, and with two different internal diameters, D=81.4 and 104.8 mm. Air to the system was supplied by a 20 HP fan and its volumetric flow rate was measured by means of an orifice plate flow meter. After being pneumatically transported (1 in Fig. 1), the particles returned through a standpipe (3) with diameter equal to 104.8 mm. A reduction nozzle (6) was placed at the air inlet, aiming at reducing both the gas flow rates deviated through the standpipe and the length of the acceleration region on the transport pipe (Silva et al., 1996). The solids flow rates were measured by diverting the flow and collecting the solids in the sample collector (5). The static pressures along the pipe were measured at seven points by pressure taps (2) connected to U-type manometers. While in operation, the standpipe was kept filled with solids and the gas flow rates through it were estimated from the Forchheimer equation, with the pressure drop measured between two points of the pipe. The gas flow rates in the transport pipe are then obtained from the mass balance which considered the difference between the air supplied by the fan and the air deviated through the standpipe, which in most cases can be neglected. The distance between the air inlet and the bottom end of the transport tube (z_0) is a parameter for controlling the solid flow rates and can be changed by a set of flanges of different thickness placed in the transport tube.

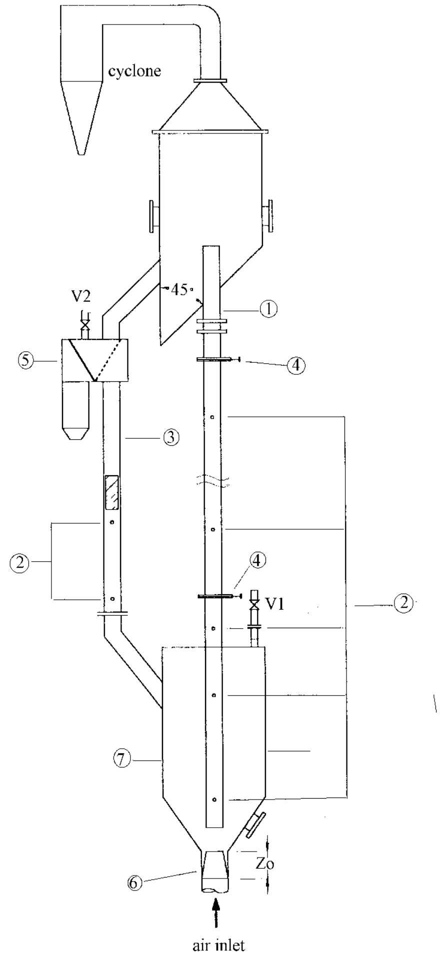

Figure 1 - Experimental apparatus; (1) transport tube, (2) pressure taps; (3) recycle tube, (4) pneumatic trap valves, (5) solids sampler; (6) reduction nozzle; (7) bottom reservoir in the spouted bed feeder.

The mean voidages (ε) at the transport pipe were measured by means of two pneumatically operated traps (4) which closed off a 2.05 m long section of the pipe. The solids accumulated on the bottom trap were collected by a suction pump and subsequently weighted, providing the average voidage. The accuracy of voidage measurements was estimated using a statistical quality control procedure (Montgomery, 1991) as being ±0.15%. The length of the acceleration region was obtained from experimental data, by plotting the pressure versus axial distance of transport tube and taking the linear portion of these curves. For all the conditions investigated, this length was found to be less than 2.1 m, and the bottom slip valve was placed above such a distance. A glass section was placed at the top end of the transport tube for visual observation of the flow. The air velocity was initially fixed at a maximum value, and the fluid dynamic measurements were carried out. After that, the air velocity was slowly decreased and the experimental procedure repeated, until ceasing the solids transport.

In all experiments, the transport pipe was grounded in order to eliminate electrostatic effects. The investigated range of experimental conditions and measurement accuracy are shown in Table 1.

Table 1. Range of operational conditions

	Range of variation	Accuracy(%)
D (mm)	81.4 and 104.8	-
d_p (mm)	0.41-1.70	-
w_f (kg/s)	0.057-0.155	2.5
w_p (kg/s)	0.207-0.623	6.0
$-dp/dz$ (Pa/m)	148.5-607.44	7.0

2.2 Theoretical Methodology

Fluid dynamics of gas-solid conveying is usually described by the continuity and momentum equations evaluated for each individual phase, according to the classical two-fluid model proposed by Nakamura and Capes (1973). Depending on the particular flow characteristics, some considerations can be made to simplify this model and constitutive relationships can be obtained empirically from experimental data. The simplified model adopted in this work is described as follows. For a comprehensive review and discussion on the fluid dynamic modeling of vertical two phase flow, we can mention the works from Arastoopour and Gidaspow (1979) and Littman et al. (1993).

Dilute phase modeling. The two-fluid one-dimensional model has been solved for the non-accelerating region, where equations are simplified because there are no velocity and voidage gradients. The fluid properties are considered constant and estimated under atmospheric pressure and at the temperature measured for each experiment. The continuity equations for the gas and particle phase are:

$$\varepsilon \rho_f u = G_f \tag{1}$$

$$(1-\varepsilon)\rho_p v = G_p \tag{2}$$

where ε is the mixture mean voidage, ρ_p and ρ_f are the particle and gas densities, u and v are the interstitial fluid and particle velocities.

The individual phase momentum balances, assuming constant fluid properties and the wall friction, are:

$$\varepsilon\left(-\frac{dp}{dz}\right) = \varepsilon \rho_f g + \beta|u-v|(u-v) - F_f \tag{3}$$

$$(1-\varepsilon)\left(-\frac{dp}{dz}\right) = (1-\varepsilon)\rho_p g - \beta|u-v|(u-v) - F_p \tag{4}$$

where $(-dp/dz)$ is the pressure gradient, g is the gravitational acceleration, β is the effective drag coefficient and F_f and F_p are the gas-wall and particle-wall friction forces, given by the following relationships:

$$F_f = 2\, f_f\, \varepsilon\, \rho_f\, u^2 / D \tag{5}$$

$$F_p = 2\, f_p (1-\varepsilon)\, \rho_p\, v^2 / D \tag{6}$$

where D is the transport tube diameter, and f_f and f_p are the gas-wall and particle-wall friction coefficients.

The gas-wall friction coefficient, f_f, was determined using the equation proposed by Colebrook (Bird et al., 1960), with the mean relative roughness obtained experimentally and equal to $2.3\text{x}10^{-3}$ for the 81.4 mm tube and to $1.33\text{x}10^{-3}$ for the 104.8 mm tube. The equation proposed by Yang (1978) with modified parameters was used to estimate the particle-wall friction coefficient, f_p:

$$f_p = 0.0004 \frac{(1-\varepsilon)}{\varepsilon^3} \left(\frac{(1-\varepsilon)}{u-v} u_t \right)^{-1.467} \tag{7}$$

where u_t is the particle terminal velocity. The modified parameters were adjusted by Silva (1997) from experimental data obtained for the transport of glass spheres in the pneumatic bed with spouted bed feeder. Equation (7) is valid for particle diameters varying from 0.24 to 2.85 mm and tube diameters from 81.4 to 148.00 mm.

In the non-accelerating region of dilute pneumatic transport the drag coefficient, C_d, can be related to the effective drag coefficient, β, using the following equation (Littman et al., 1993):

$$\beta = \frac{3(1-\varepsilon)\, \rho_f\, C_d}{4\, d_p} \tag{8}$$

Several empirical equations are available in the literature for predicting C_d in gas-solids suspensions (Lee, 1987; Grbavcic, 1991). In this work we employed the equation from Turton and Levenspiel (1989) for the standard drag curve. At high mixture voidage, the use of standard drag curve is well-accepted and has been adopted by authors such as Arastopoour and Gidaspow (1979).

Once the relations for predicting the solid-wall friction factor and the drag coefficient have been provided, the model is reduced to a set of four nonlinear algebraic equations, which must be solved using an iterative method to yield the variables u,v, ε and $-(dp/dz)$. In this work, the solution was obtained using the BROYDN subroutine described in Raman (1985). The diameter of the transport tube, physical properties of the solid and gas phases, gas and solid flow rates, as well as initial values of the variables u,v, ε and $-(dp/dz)$ were inserted in the computational program.

Dense phase modeling. Preliminary investigations on the pneumatic tube with spouted bed feeder indicated dense phase flow patterns with characteristics of non-slugging flow. Although the definitions of this type of flow regime change from one author to another, Marcus et al. (1990) describe this flow as being characterized by extreme turbulence, particles travelling in clusters and strands which break up continuously and extensive backmixing of solids. Little information is available in literature for predicting the pressure drop in the non-slugging dense phase mode. Yerushalmi and Cankurt (1979) have suggested that the friction loss could be neglected on the momentum balances. Then the total pressure drop for the gas-solid mixture can be estimated considering only the solids hold-up:

$$-\frac{dp}{dz} = \left[\rho_p(1-\varepsilon) + \rho_f\,\varepsilon\right]g \tag{9}$$

A correlation for estimating the voidage is suggested by Yerushalmi and Cankurt (1979). They observe, however, that it needs further validation and do not recommend it for general use.

3. RESULTS

Typical curves of pressure gradient versus air flow rates are shown in Fig. 2.

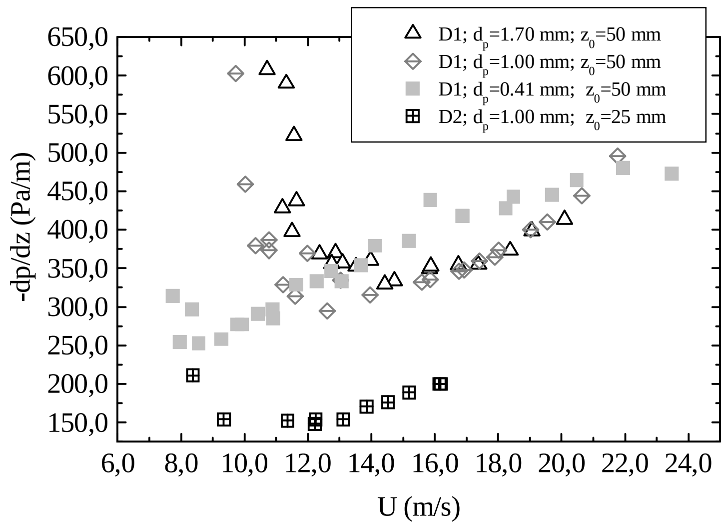

Figure 2 - Pressure gradients versus superficial air velocities for the pneumatic transport tubes with spouted bed feeder; D1=81.4 mm and D2=104.8 mm.

Data showed in Fig. 2 have been obtained by reducing the air flow rate from the maximum value provided by the blower to the minimum one required to maintain the transport of particles. For both tube diameters, the pressure gradient measured at constant z_0 initially decreases with increasing the air velocity, until reaching a region in which these curves change their behavior and the pressure gradient starts increasing with the increase of air velocity. It was observed during the experiments that for the highest air flow rate no oscillations were detected in the water manometers connected to the tube. However, as the air rate was reduced, the oscillations started to increase, and reached values of about ± 1.0 cm at the smallest air velocity. These observations clearly indicate an alteration on flow characteristics, and the air velocity at the minimum point is assumed as being a transition between the two different flow regimes. The value of the air velocity at this minimum point on each one curve has been calculated by adjusting linear equations to the points located on the left side and to those located on the right side of the minimum pressure gradient. The intersection between these two lines defines the air velocity at this transition. So data situated on the left side of this intersection are in the dense phase flow while data located on the right side are in the dilute flow. A reduction of pressure gradient is expected when the tube diameter is increased because the gas and solid flow rates decrease.

Analyzing the curves obtained for the 81.4 mm diameter tube we see that the minimum air velocity is reduced when the particle diameter is reduced (from 13.0 m/s for the 1.70 mm particles to 12.1 m/s for the 1.00 mm particles and to 9.0 m/s for the 0.41 mm particles). It is also observed that the effect of particle diameter on the pressure gradient depends on the flow regime. In the dense phase region, the pressure gradient increases with particle diameter. This is expected because higher air velocities are required to transport larger particles. But the dependency of pressure gradient on the particle diameter changes when we analyze the dilute phase region. For particles of 1.00 and 1.70 mm diameters, the pressure gradient shows no dependency on the particle diameter while the pressure gradient values measured for the 0.41 mm particle are higher than the ones measured for larger particles. We observe that particles of 1.00 and 1.70 mm belong to group D on the diagram proposed by Geldart (1973), while the 0.41 mm particles belong to group B. The characteristics of solids to be conveyed affect the way these particles behave when are fluidized, and in this case the solid characteristics may affect also the solids feeding at the tube entrance. Note that the feeder used here is a particular type of fluidized bed and the dependency between solid and air flow rates may be affected by the particle diameter. We believe that the differences in particle characteristics may be a possible reason for the change in curves behavior, but a better understanding about these mechanisms still requires further investigation..

Experimental values of mean voidage are shown in Fig. 3 as a function of solid to gas mass ratio (w_p/w_f). In this figure, the flow regime has been identified following the criterion previously described.

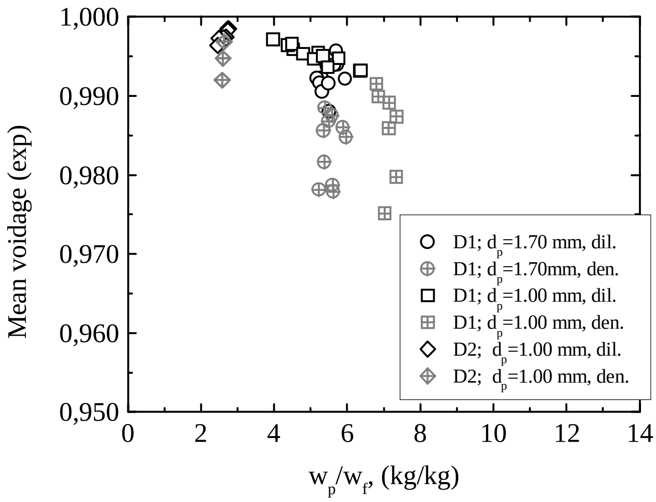

Figure 3 – Experimental mean voidage as a function of the solid to gas mass flow ratio for pneumatic transport tubes with a spouted bed feeder; D1=81.4 mm and D2=104.8 mm; dil.=dilute phase flow; den.=dense phase flow.

As observed in dense phase flow conditions most of the mean voidage values varied from 0.975 to 0.990 while in dilute phase conditions these values were always greater than 0.990. From these results we conclude that the mean voidage cannot be used as a criterion for identifying the flow regimes since the values of solids hold-up in dense phase flow are not too different from those considered as typical of the dilute flow (Leung, 1989; Mok et al., 1989). Changes in flow characteristics detected in the pressure diagrams do not cause sharp variation in the voidage and the transition between the flow regimes seems to occur in a diffuse way. The voidage for the 0.41 mm particles could not be measured because the pneumatic traps did not operate well for such small particles.

The two phases one dimensional model described in section 2.2 was solved to predict the pressure gradient. Since this model is valid for high voidage conditions, it has been applied initially for all experimental data, including those related to dilute and dense flow regimes. The aim was to verify if this model is sensitive to the changes in flow characteristics detected in the pressure diagrams. Deviations between the predicted and measured values of pressure gradient are showed in Fig. 4.

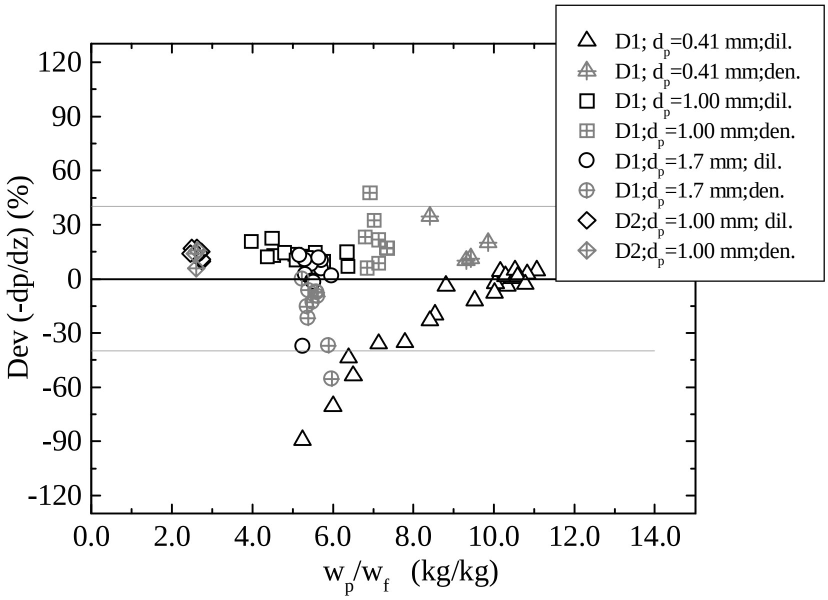

Figure 4 – Deviations between predicted and experimental pressure gradients versus the particle to gas mass ratio, D1=81.4 mm and D2=104.8 mm.

It can be observed in Fig. 4 that most of the deviations are in a range of ±40%, with a randomly distribution. Deviations values do not decrease in the dense flow conditions, as would be expected. The limitations of one-dimensional model for predicting the pressure gradient and mean voidage in the dilute flow have already been discussed in previous works (Ferreira et al., 1996). Probably a better adjustment between predicted and experimental values of *(-dp/dz)* would be obtained if the experimental data obtained were used to estimate the solid-wall friction coefficient and the drag force. Note that these parameters are normally used to adjust the model to experimental data. The empirical equations used to determine these two parameter have been obtained by Silva (1997) in the same equipment, but for a narrower range of experimental conditions. The adjustment might improve if a larger number of experimental data were considered. However, this is not our objective at the moment. The point to be focused is that the one dimensional model is not sensitive to the changes of flow regimes observed in the pneumatic tubes with spouted bed feeder. If suitable empirical equations were provided for the solid-wall friction and the drag force, the model should predict better the pressure gradient for all the range of air velocity used in the pressure diagrams.

Equation (9) could not predict well the pressure gradient in the dense phase flow. A possible reason is that the correlation used to estimate the voidage values (the one proposed by Yerushalmi and Cankurt, 1979) is not suitable for particles used here. Even when Eq. (9) has been solved using experimental voidage values, the pressure gradient deviations were up to ±80%. So we conclude that the simplifying assumptions used to derive Eq. (9) are not valid for the conditions used here.

A comparison among the specific power consumption obtained at different particle to gas mass ratios can be seen in Fig. 5. The specific energy consumption for each experimental condition was estimated from the equation suggested by Mills (1990).

It can be observed that the solid to mass gas ratios vary from values close to 2 (for the 1.00 mm particles in the 104.8 mm transport tube) to 11 (for the 0.41 mm particles in the 81.4 mm transport tube). The values of specific power consumption obtained are high, but compatible with data reported in the literature for similar ranges of solid to gas mass ratios (Taylor, 1998).

The results show that, in the range of solid to gas mass ratios investigated, there is no reduction in the specific power consumption for dense flow conditions in the pneumatic tubes with spouted bed feeder solid feeding system.

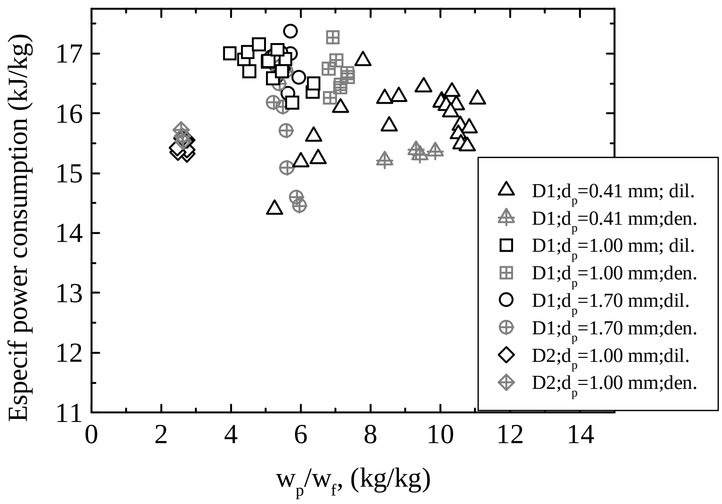

Figure 5 – Specific power consumption versus particle to gas mass ratio, D1=81.4 mm and D2=104.8 mm.

4. CONCLUSIONS

Experimental data obtained in this work allow the following conclusions:

- the pneumatic tube with spouted bed feeder may be operated in two different flow regimes, the change between flow regimes is indicated by the change in the behavior of pressure gradient versus air velocity curves. At high air velocities the pressure gradient decreases when the air velocity is reduced but at a certain air velocity the pressure gradient start to increase with decreasing air velocity. The transition point is given by the air velocity at the minimum pressure drop, which is dependent on the particle and transport tube diameters;

- the curves of pressure gradient versus air flow rate are qualitatively similar for all the particle diameters investigated, but the dependency of these curves on the particle diameter is different for each flow regime. In dense phase region the pressure gradient for a given air velocity increases with increasing particle diameters, but in dilute region the greatest pressure gradient was obtained for the smallest particle. Data obtained until now do not allow a clear understanding of this behavior and further investigation about the influence of particle characteristics on the pressure gradient and on feeder performance is still needed;

- for the conditions investigated the comparison between the measured pressure gradient and the estimated values using the two-fluid one dimensional model in the dilute flow shows that the model is not sensitive to changes in flow regimes detected in the pressure diagrams;

- the modified one-dimensional momentum equation suggested by Yerushalmi and Cankurt (1979) does not predict well the pressure gradient in the dense phase flow, even if the experimental voidage values are used in the model;

- the pneumatic bed with spouted bed feeder provides high specific power consumption, and no reduction in this values were detected for dense phase flow conditions.

Akwnoledgements

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