FLOW OF SLURRIES OF PARTICLES WITH DENSITY CLOSE TO THAT OF WATER

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ABSTRACT

The purpose of this work is to investigate the friction factor for slurries of particles with density close to that of water. The use of particles with density close to that of ice aims to get explanations of new generation of liquid-solid coolant in adiabatic conditions. The solid particles are 3mm-diameter polypropylene sphere, of density 869 kg/m^3. The test section consists of horizontal transparent PVC pipes 22 and 45mm in diameter. Friction losses are measured using differential pressure transducers. Velocity distribution within the water layer is measured using Pitot tube. Flow pattern and particle behaviour are observed using visualisation device.
Different flow pattern can be observed in fully turbulent conditions. The transition between the flow pattern depends on the Froude number. Three predicting models of pressure losses in pipe have been used. The first one consists in correlating experimental data using dimensional analysis. The second one uses rheological model: extended Reynolds number based on the empirical Bingham model is introduced. The Blasius law computed by replacing the classical Reynolds number with the extended Reynolds number is found to be applicable to the two-phase flow in its full range. The third one is a two-layer model that contains the effect of pipe diameter. The model treats the flow as being constituted of two layers: a stationary or a moving bed at the top of the pipe and a heterogeneous flow suspension below. The friction coefficient between the bed layer and the pipe is estimated by performing momentum balance using the experimental results (pressure drop, visualisation and distribution of the local velocity within the heterogeneous flow suspension. The validity of the hypothesis of the two-layer model are discussed.

INTRODUCTION

The hydraulic behaviour of particles with density near that of the carrier fluid are of great importance in the new generation of solid liquid coolant. When the process or the cold requirement are far away from the refrigeration machines as well when the frigorigens have to be used in confined areas (because of their toxicity, their dangerousty or because environmental problems), secondary fluids have to be used to transport the refrigeration energy from the production place to the consumption place. The liquid solid two-phase coolant such as ice slurries of encapsulated phase change materials are very promising because of their enthalpy: systems using latent heat enable greater heat transfer rate than equal mass flow rate system using only heat capacity. For example, the transported energy of ice water slurries with a concentration of 30% by volume is about six times higher than the transported energy of cold water for a classical air cooling system. The distribution of the coolant system can be constructed with smaller diameter pipes. The eat capacity of coolant distribution can be enhance without changing the piping elements by replacing a classical liquid coolant with a two phase coolant (air conditioning of big cities).
The purpose of this work is to study the pressure losses and the two-phase flow structure of concentrated liquid solid two-phase coolant with density particles close to that of the carrier fluid in horizontal pipes and in adiabatic conditions. The liquid phase is water and the solid phase is made with polypropylene particles with density close to unity ρ= 869 kg/m3. The grains are spherical and

3 mm in diameter. The experimental study is based on visualisation, pressure drop and velocity measurements. The results are compared to the reference behaviour of water flow.

The forces affecting the two phase flows play different roles according to the flow pattern, the properties and the concentration of the solid particles (density, size, shape…). The reasons for deviation from pure fluid behaviour are so different that head losses model have to be adapted to each flow configuration (slurries of particles with density close of the carrier fluid, (shook 1985), very high concentration slurries of dense particles (Wilson & al. 1972…)). Many investigations have been performed to predict head losses of solid liquid flows. Different approaches have been developed. Many authors (Turian & Yuan 1977, Zandi & Gavados 1967, Snoek et al. 1996…) have established empirical correlations motivated by dimensional analyses. They can provide very satisfactory interpolation. Extrapolation of the results to flow conditions out of range of the database used has to be taken cautiously. The second approach is to study the reology of the slurries and to establish friction factor – Reynolds number correlation using classical non-Newtonian fluid model. The Bingham model is often used because of its simplicity. The distributions of the plastic viscosity and the yield stress as functions of the solid concentration are determined by applying least square method to the Buckingham-Reiner equation for a laminar flow in a pipe (Ogihara & Miyasawa (1991), Hansen & al. (2000)). By replacing the ordinary Reynolds number by an extended Reynolds number, Ogihara & Miyasawa (1991) found the moody diagram to be applicable to describe the friction losses in laminar and turbulent flows. The third approach is to develop theoretical approaches such as two-layer model or three-layer model (Takahashi et al. 1991, Kelly et al. 1998, Doron and Barnea 1993)

1. CONSTITUTIVE EQUATION

Under isothermal conditions, and no work involved, pressure drop in horizontal piping system can be described by the steady state macroscopic mechanical energy balance given by equation:

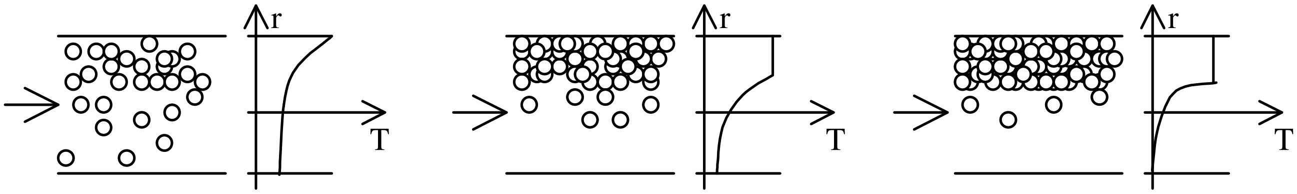
$$\Delta P = P_1 + \alpha_1 \frac{1}{2}\rho\langle V_1\rangle^2 - P_2 - \alpha_2 \frac{1}{2}\rho\langle V_2\rangle^2$$

Where P is the static pressure (Pa), ρ the density (kg/m^3), $\langle V\rangle$ the mean velocity in a given section. Subscript 1 and 2 correspond respectively to the upstream and the downstream section; α is a kinetic energy correction factor. The friction losses are expressed in term of friction factor λ defined by:

$$\lambda = \frac{\Delta P}{\frac{1}{2}\rho\langle V\rangle^2}\frac{D}{L}$$

Solid-liquid two-phase flow though straight pipe may be classified in different flow patterns determined by visual observations. In this present study, we use the classification of Doron and Barnea (1993) who define 3 main flow patterns (Figure 1):

1. "Fully suspended flow": the whole solid particles are suspended. This pattern includes the homogeneous flow and the heterogeneous flow, which are defined by a solid phase concentration gradient in the direction perpendicular to the pipe axis.
2. "Flow with moving bed": solid particles accumulate at the top of the pipe. It is characterised by a solid concentration near the maximal "packing". The rest of the pipe is occupied by heterogeneous mixture.
3. "Flow with stationary bed": a stationary bed develops on the top of the pipe. Deposit particles are transported as a separate moving bed on the bottom of the stationary bed. The transition between flow with stationary bed and moving bed is generally associated with a limit deposit velocity corresponding to minimum pressure gradient.



Fully suspended flow Flow with moving bed Flow with stationary bed.

Fig.1 : flow pattern (T is the volume solid concentration)

For high slurry flow rate and solid concentration beneath 30%, all the solid particles are suspended and head losses are similar to the one obtain for liquid flows and the slurry exhibits Newtonian behaviour. If the flow rate is reduced, the solid particles form a moving deposit at the top of the pipe below which flows a heterogeneous mixture with a very low solid concentration. A part and then the whole moving bed become stationary if the flow rate rather decreases. Pressure losses are quite different than that for pure liquid when solid particles tend to agglomerate at the top of the pipe.

Different methods have been developed and tested in the present work to predict the friction losses for the two-phase solid-liquid flow.

1.1 Empirical correlation

Two empirical correlations have been tested. The first one is based on the analysis of the distribution of the friction factor as a function of the Reynolds number:

$$\frac{\lambda - \lambda_l}{\lambda_l.\mathrm{T}} = a.\mathrm{T}^b.\mathrm{Re}^c + \mathrm{d.T^e.D^f}$$

With λ_l friction factor for liquid flow, D the diameter of the pipe, Re Reynolds number define as Re = <V>.D/ν_l (ν_l is the cinematic viscosity of the liquid, <V> the spatial average velocity within a given section of the pipe). The second one is based on the analysis of the distribution of the friction factor as a function of the Froude number:

$$\frac{\lambda - \lambda_l}{\lambda_l.\mathrm{T}} = \alpha.\mathrm{T}^\beta.\lambda_l{}^\chi.\mathrm{Fr}^\delta$$

With Fr Froude number define as Fr = $\langle V\rangle^2$ / [D.g.(1-ρ_s/ρ_l)] with g gravitational acceleration. Subscript s and l correspond respectively to the solid and liquid phase.

1.2 Correlations based on rheological model

The method consists to in assimilating the global motion of two-phase flow with a motion of non-Newtonian fluid whose characteristics are depending of solid concentration (Ogihara and Miyazawa, 1991). Considering the motion law of Bingham fluid, the relationship between the shear stress (τ) and the shear rate ($\delta u/\delta r$) is given by

$$\tau = -\eta \frac{du}{dr} + \tau_c$$

With τ_c yield stress and η plastic viscosity. Reynolds number represents the ratio of inertial forces to viscosity forces. Extended Reynolds number Re_1 based on Bingham fluid may be defined by

$$Re_1 = \frac{-\rho\, u\, du/dr}{\tau / l} = \frac{\rho\, ul / \eta}{1 - \dfrac{\tau_c}{\eta\, du/dr}}$$

Friction losses obtained for heterogeneous suspension flow at high Reynolds number are quite similar to the one obtained with clear water. Therefore the plastic viscosity η is fixed to the dynamic viscosity μ_l of clear water. Considering the Reynolds number Re based on dynamic viscosity μ_l of water, D diameter of pipe and ρ relative density of mixture and using the mean velocity <V> and the diameter of pipe as characteristic dimensions, the relation becomes

$$Re_1 = Re \times \frac{1}{1 + c_1 \tau_c \dfrac{D}{\mu_e \langle V \rangle}} = Re \times \frac{1}{1 + c \dfrac{D}{\mu_e \langle V \rangle}}$$

The dimensionless coefficient c_1 is determined experimentally for each solid concentration in order to have a single relation between the friction factor and the extended Reynolds number Re_1. This correlation should be similar to the one obtained in the case of Newtonian flow in order to be applicable in the all-solid concentration range. Considering the range of Reynolds number studied (Re>10 000) and the pipes used (PVC), the correlation correspond to the Blasius law.

1.3 Two layer model

A two-layer model has been developed by Takahashi et al. (1991) to estimate pressure losses for concentrated slurries of particles with density near that of the carrier fluid. The flow is described as follow: a solid particle bed develops on the top of the pipe, below which flows a heterogeneous mixture with very low solid concentration. The model is based on the following assumptions: the shear stress between the particle layer and the pipe wall is assumed to be K times the shear stress between the heterogeneous layer and the pipe wall. The velocity of the particle bed V_{pb} is assumed to be negligible than that of the heterogeneous layer V_{hl}. The mean velocity of the heterogeneous layer $\langle V_{hl} \rangle$ is estimated by:

$$\langle V_{hl} \rangle = \frac{\pi}{(\pi - \theta + \sin\theta\cos\theta)^2} \frac{Q(1-T)}{A}$$

Q is the volume flow rate, A the cross section of the pipe. Angle θ shows the thickness of the moving layer (figure 2). θ is estimated using empirical equation ($\theta = \alpha\,(0.059\,Fr^{-1} - 0.09)$ for $Fr \le 38$; $\theta = 0$ for $Fr \ge 38$)

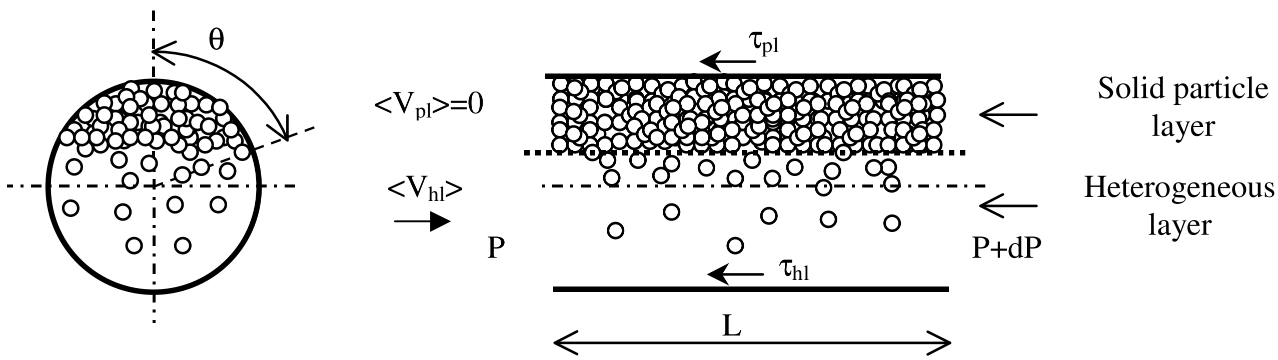


Fig. 2: Schematic diagram of the two-layer model

Subscripts pl and hl correspond respectively to the solid particle layer and the heterogeneous layer. The friction coefficient λ is given by (λ_l is the friction coefficient for pure liquid flows) :

$$\lambda = \frac{\left(1 + KR\left(\cos\theta - \dfrac{\sin\theta}{\theta}\right)\right)\pi\theta}{(\pi - \theta + \sin\theta\cos\theta)^2} \lambda_l + \frac{\pi(\pi - \theta)}{(\pi - \theta + \sin\theta\cos\theta)^2} \lambda_l$$

2. EXPERIMENTAL SETUP

Fig.3 shows a schematic diagram of the experimental apparatus used in this study. A 120 litres storage tank (2) allows setting the particles concentration by introducing or by getting back particles from the feed tank (1) of 300 litres. A pump driven by a variable speed motor and connected to the feed tank enables accurate control of the delivered solid concentration. A vortex pump (3) connected to a speed variator set the flow inside the test section. Two sets of valves (4) and (5) allow accurate flow regulation. Two test sections (6) (7) allow pressure losses measurements in horizontal straight pipes. A « Rosemound Elite 200 » mass flowmeter (8) measures the mass flow rate Qm, the density of the mixture ρ flowing throughout the test sections and the temperature with the following accuracy : $\Delta Qm = \pm 1 kg/h \quad \Delta\rho = \pm 0.5 \mathrm{kg/m^3}$

The solid concentration T is given by the relation: $T = \dfrac{\rho_e - \rho}{\rho_e - \rho_b}$. The maximal uncertainty is less than 2% Pressure drop are measured using 4 differential DRUCK pressure transducers. The relative uncertainty is estimated at about 0.1% of the measurement range (0-2bars, 0-200mbars, 0-20mbars, 0-2mbars). Data acquisition is obtained by a Keithley DAS 1700HR card with a resolution of 32 bits i.e. ΔV = ± 5mV on the measure range of 0-10V. This card is installed on a CyrixInstead computer. The sampling frequency is 500 Hertz. Continuous measurements are performed for 10s. A HI resolution SONY camera (9) with 16x enlargement is linked to a Pentium II Intel MMX computer. It is implemented with an acquisition card AV Master. Video acquisitions are made with 'Fast capture" software. The visualisation of the flow is carried out in real time; 100 pictures by seconds can be registered. A Pitot tube (10) mounted on the 45mm tube allow local velocity measurements. Estimate of the volume flow-rate computed with the Pitot and mass flow-meter measurements agree in their range of uncertainty.

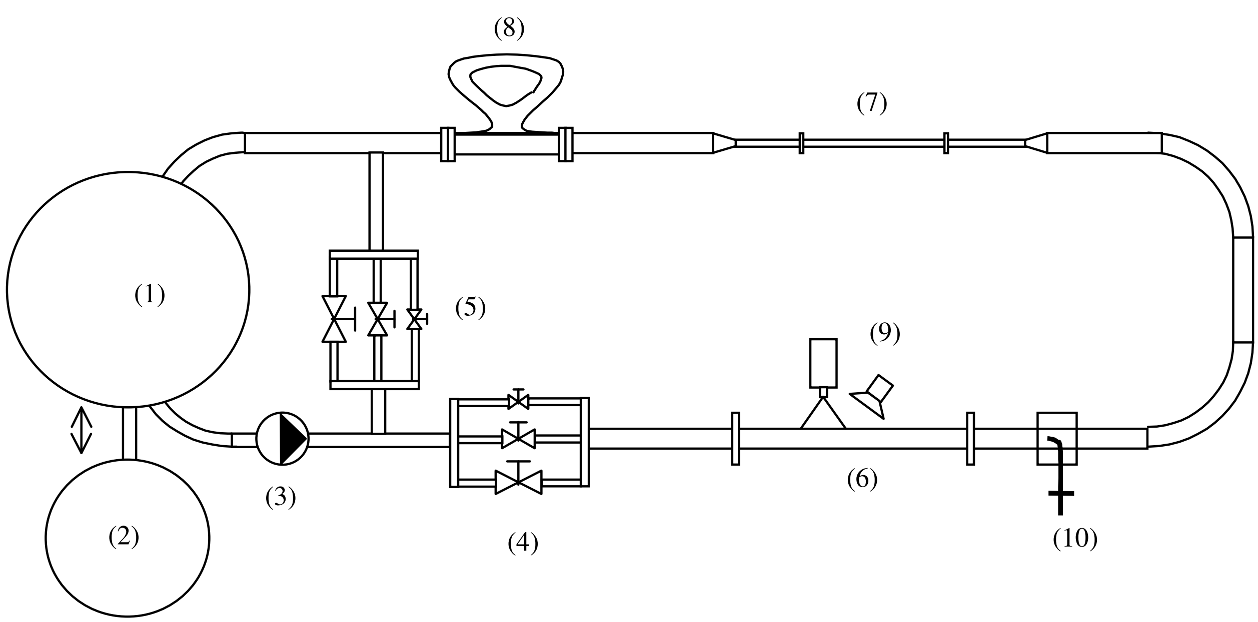


Fig. 3: schematic diagram of the experimental set-up

The solid particles used in this study are polypropylene sphere 3mm in diameter and of relative density 0.869. Their density was determined by suspending them in methanol-water solution to produce neutrally buoyant mixtures. Maximum packing concentration of 0.68 was determined in a cylinder of diameter 4cm in air.

3 RESULTS AND COMENTS

Different flow pattern may be observed as a function of the Froude number (Figure 4)

1. Fr > 15: heterogeneous suspension flow.
2. 0.2 < Fr < 15: moving bed.
3. Fr < 0.2: stationary bed.

heterogeneous suspension flow	moving bed	stationary bed

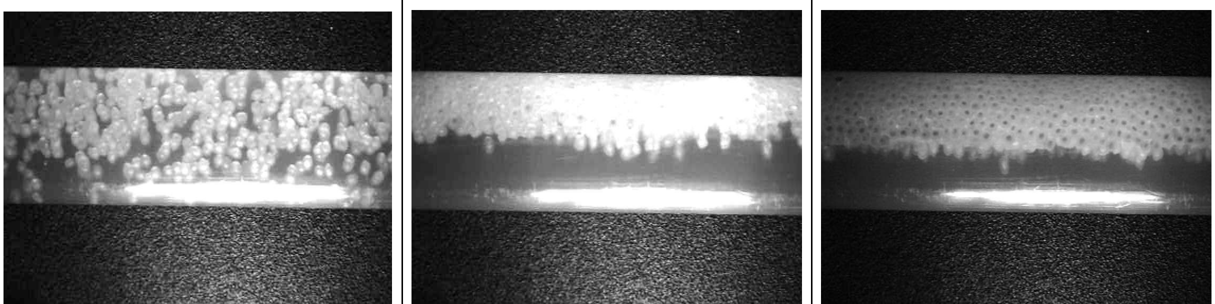


Fig. 4: Flow pattern

The distribution of the friction factor λ as a function of the Reynolds number and the distribution of the pressure loss coefficient ϕ as a function the Froude number are shown in figure 5 and 6. The pressure loss coefficient ϕ is defined as $\phi=(\lambda-\lambda_w)/(\lambda_w{*}T)$

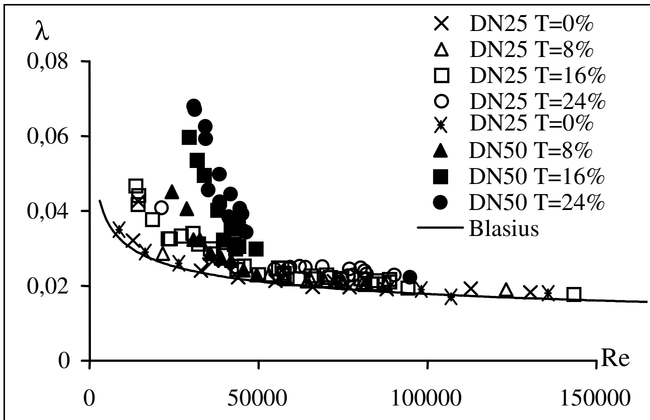


Fig. 5 : Distribution of the friction factor λ as a function of the Reynolds number

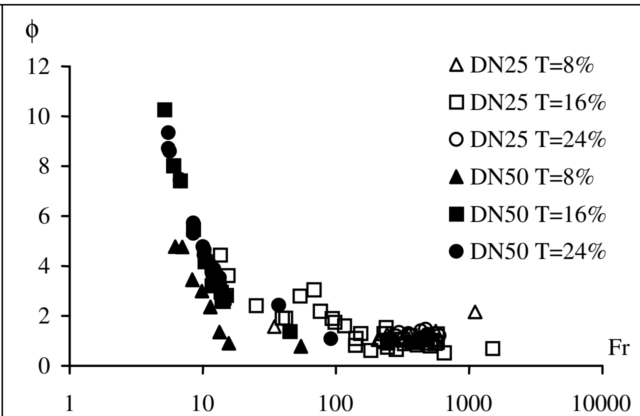


Fig. 6 : Distribution of the pressure loss coefficient ϕ as a function of the Froude number

3.1 Empirical correlations

The constants of the empirical correlations are determined by fitting the experimental data. The comparison between the experimental results and the calculated results are shown in figure 7 and 8.

The calculated results show good agreement with the experimental data. The average of the absolute relative deviation between the correlation and the experimental results $(\lambda_{exp}-\lambda_{cor})/\lambda_{exp}$ is lower than 4% for the correlation based on Froude number and lower than 8% for the correlation based on Reynolds number.

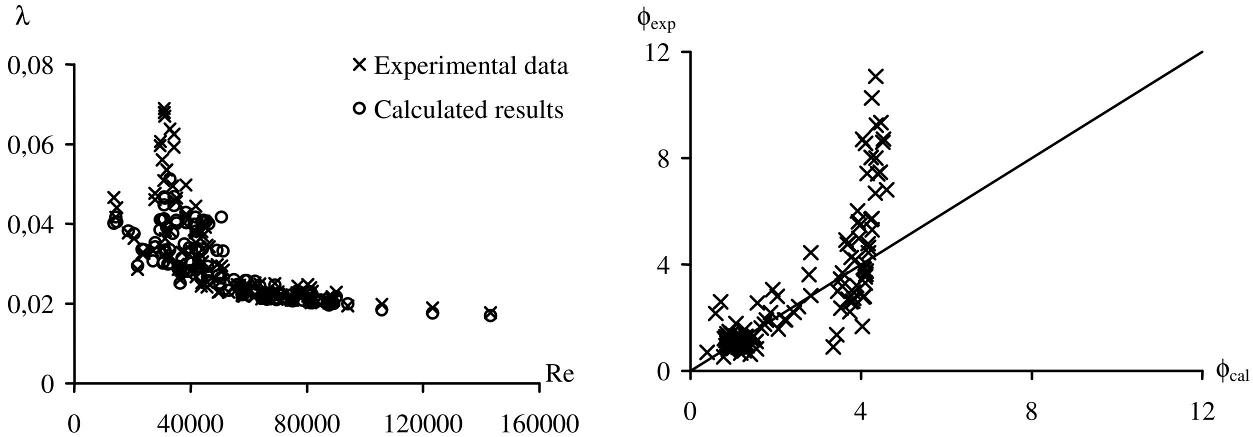


Fig. 7 : Prediction of friction coefficient λ based on Reynolds number correlation

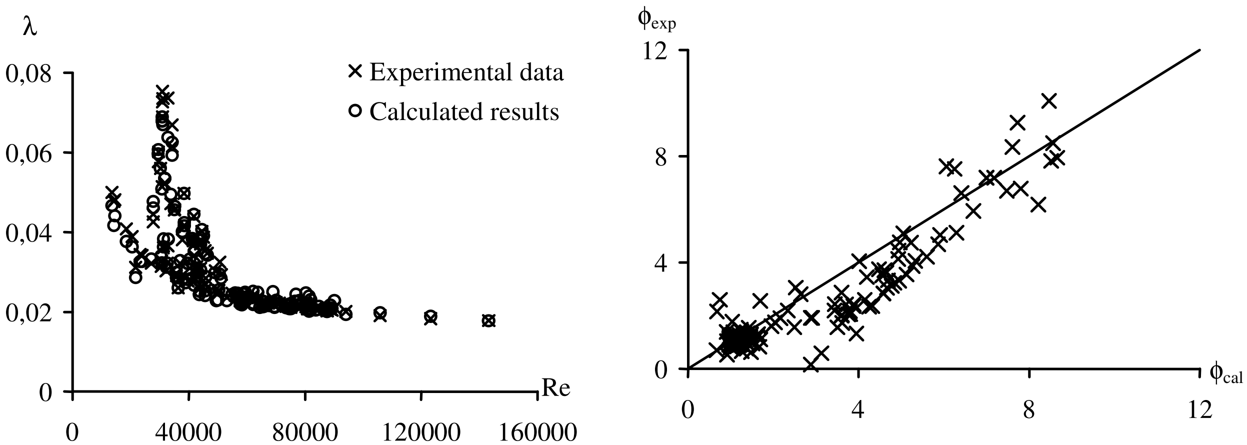


Fig. 8 : Prediction of friction coefficient based on Froude number correlation

The final form of the correlation based on Froude number is the following:

$$\phi = 576.\mathrm{T}^{-0,27}.\lambda_e^{1,5}.\mathrm{Fr}^{-0,31}$$

The results obtained by Turian and Yuan are quite different. Such results are based on a large number of data points collected from the literature and also taken from their slurry flow experiment. Contrary to our experiments, the solid density of the entire body of their data points is greater than the water density.

The final form of the correlation based on Reynolds number is the following:

$$\phi = 48,49.\mathrm{T}^{0,235}.\mathrm{Re}^{-0,105} - 2,63.\mathrm{T}^{0,263}.\mathrm{D}^{-0,44}$$

Such empirical form based on power law can provide very satisfactory correlation of experimental data. Extrapolation of the results to flow conditions out of range of the database used have to be taken cautiously. Otherwise the pressure drop correlation is sensible to the flow regime and need to develop an associated quantitative regime delimitation scheme.

3.2 Correlations based on rheological model

The value of the c coefficient included in the extended Reynolds number is determined by applying least square method in order to adjust for a given solid concentration the Blasius law. The relation between the c coefficient and the solid concentration is shown in figure 9 and 10 for internal diameter 0,45 and 0,22.

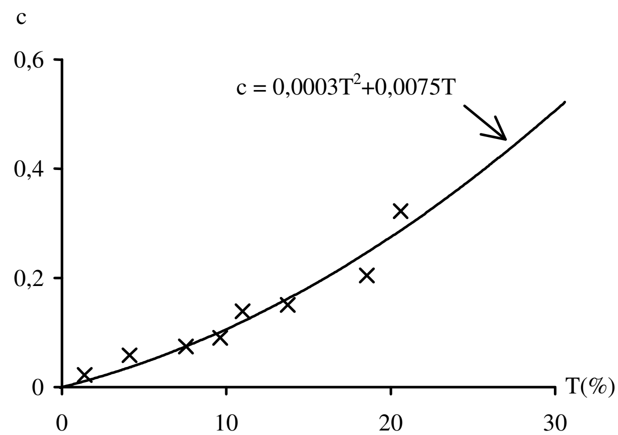


Fig. 9 : Distribution of coefficient c as a function of the solid concentration (diameter 22mm)

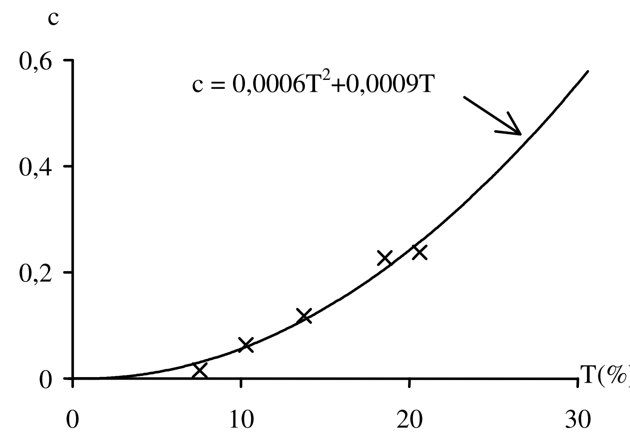


Fig.10 : Distribution of coefficient c as a function of the solid concentration (diameter 45mm)

The distribution of the friction factor as a function of the extended Reynolds number is shown in Figure 11. The calculated results show good agreement with the experimental data. The average of the absolute relative deviation between the correlation and the experimental results is lower than 6%.

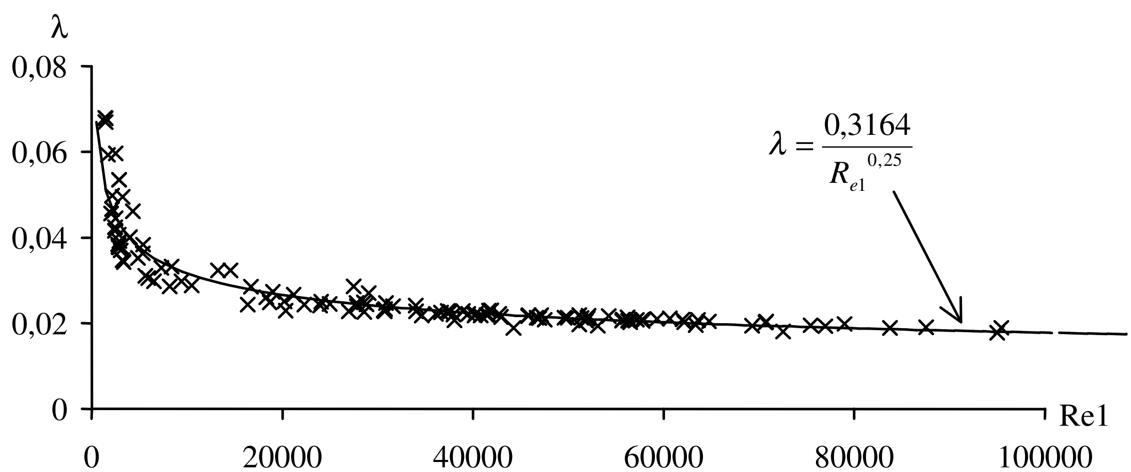


Fig. 11: Distribution of the friction factor as a function of the extended Reynolds number Re_1

3.3 Two layer model

The distribution of the friction factor estimated using the two-layer model as a function of the Reynolds number is shown in Figure 12. The average of the absolute relative deviation between the correlation and the experimental results is lower than 10%. The minimum deviation between the experimental results and the calculations is obtained for shear stress between the particle layer and the pipe wall equal to 50 times the shear stress between the heterogeneous layer and the pipe wall. (Takahashi found a value of K equal to 20)

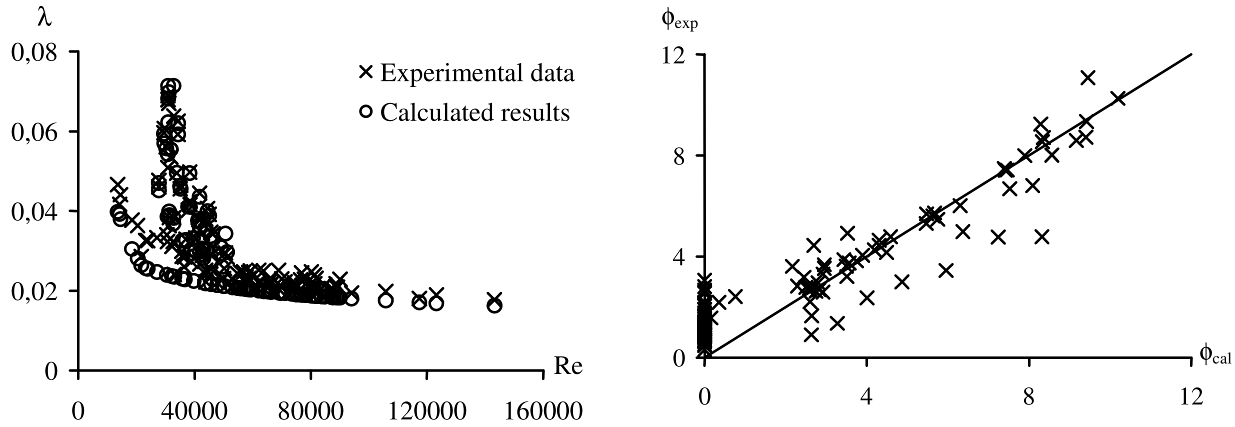


Fig. 12 : Prediction of friction coefficient based two layer model

3.4 Local measurements

Typical distribution of the local velocity of water within the pipe is shown in Figure 13.

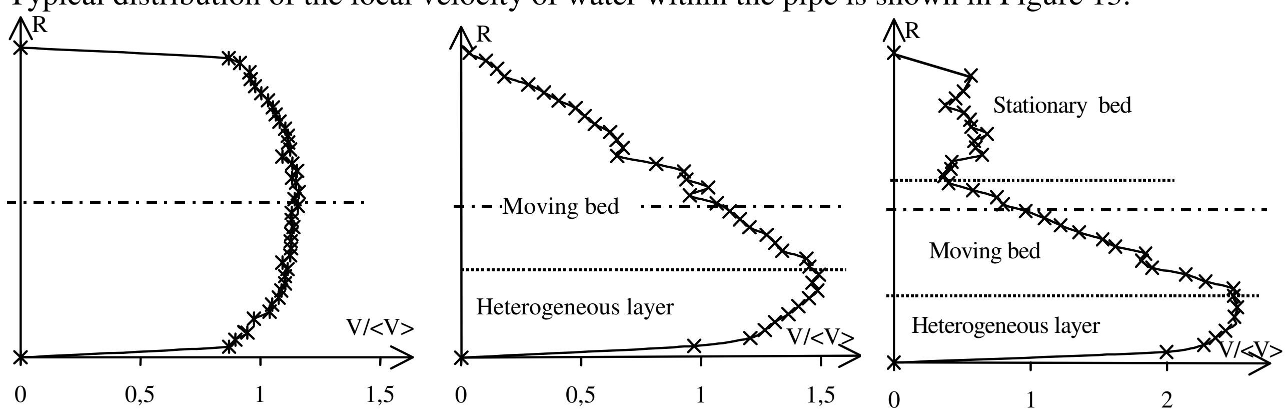


Fig. 13: Velocity distribution within straight pipe under developed flow conditions

Mass and momentum balances may be determined using the experimental results in order to check the relevance of the two layer model. Mass and momentum balance in the x direction are performed under developed flow conditions with the following assumptions: the solid concentration in the heterogeneous layer is negligible. The solid concentration in the moving bed is assumed to be T_{pl}=0.52 for cubic packing. Let us consider the tube flow system in figure 14.

The continuity equation for the liquid phase is: $\langle V_{hl} \rangle .A_{hl} + \langle Vl_{pl} \rangle .(1-0{,}52).A_{pl} = \langle V \rangle .(1-T).A$

The continuity equation for the liquid phase is: $\langle Vs_{pl} \rangle .0{,}52.A_{pl} = \langle V \rangle .T.A$

The integral balance of momentum is: $\frac{dP}{dx}.A + \tau_{pl} L_{pl} + \tau_{hl} L_{hl} = 0$

With L_{pl}=2.θ. R, L_{hl}=2.(π-θ). R. The shear stress between the turbulent heterogeneous layer and the pipe wall is:

$$\tau_{hl} = \frac{1}{2} \rho_l \langle V_{hl} \rangle^2 f_{hl}$$

With f_{hl} the friction coefficient $f_{hl} = 0{,}046.Re_{hl}^{0.02}$, Re_{hl} the Reynolds number $Re_{hl} = \frac{D_{hl}.\langle V_{hl}\rangle}{\nu_l}$ based on the hydraulic diameter $D_{hl} = \left(4.A_{hl}/(L_{hl}+L_{hl-pl})\right)$ $\quad L_{hl-pl} = 2R.\sin(\theta)$

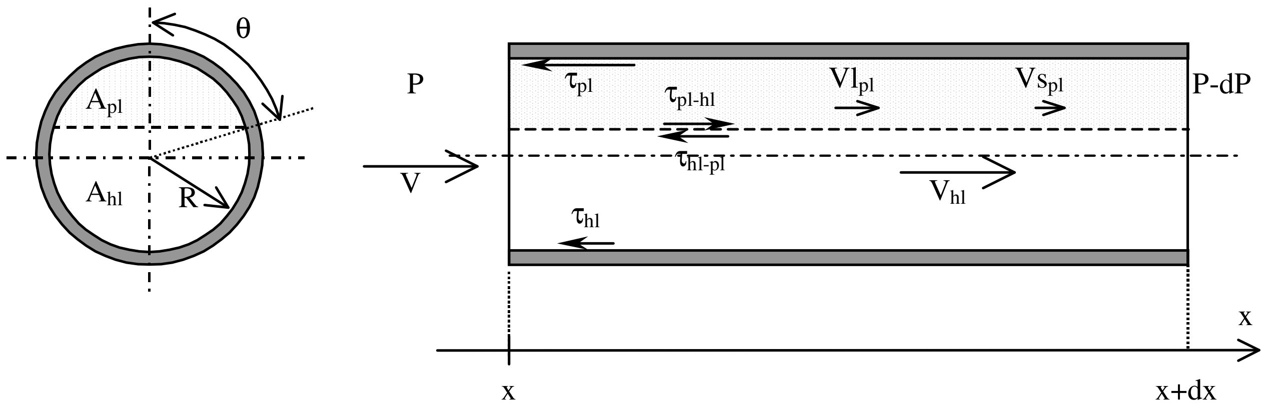


Fig. 14: Schematic diagram of force balance

Vl_{pl} and Vs_{pl} are respectively the liquid and the solid velocity within the solid particle layer. The global mass and momentum balances show that the hypothesis of the two-layer model of Takahashi are erroneous: The liquid flow-rate within the heterogeneous layer is lower than 70% for relative height of the particle bed layer greater than 50%. The liquid flow rate in the particle bed is not negligible. Significant slip velocity appears between the solid particles and the liquid within the solid particle layer (Figure 16). τ_{pl} is about 3.2 times τ_{hl} for flows with stationary bed, and about 5.8 time τ_{hl} for flows with moving bed. (Figure 15).

The distribution of the shear stress between the particle layer and the pipe wall as a function of the mean velocity of the two-phase flow is shown in Figure 17. The experimental data agree well with the empirical equation given by $\tau_b = 4{,}48 \langle V\rangle^{1{,}46}$

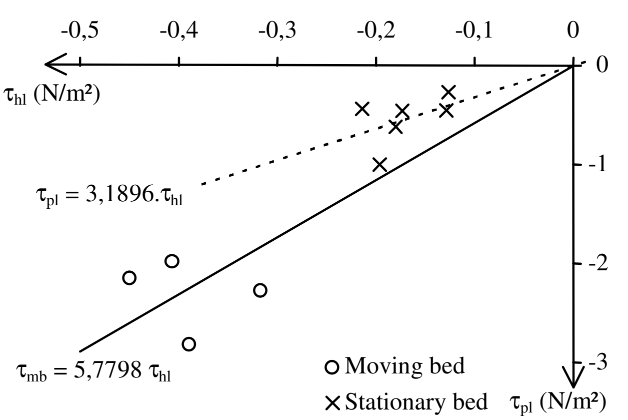


Fig.15 : shear stress between the particle layer and the pipe wall as a function of the shear stress between the heterogeneous layer and the pipe wall

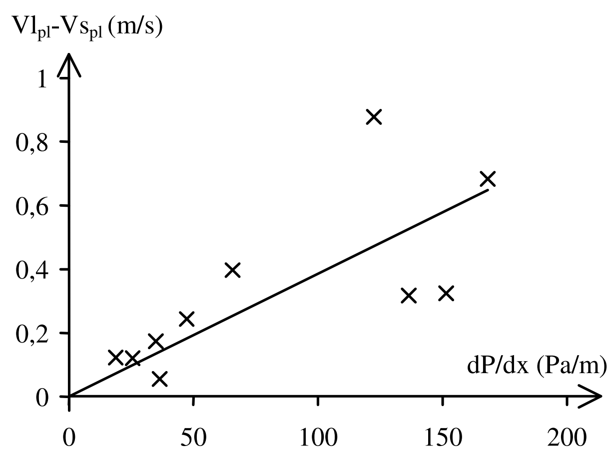


Fig. 16 : Distribution of the mean slip velocity within the solid particle layer as a function of the pressure drop

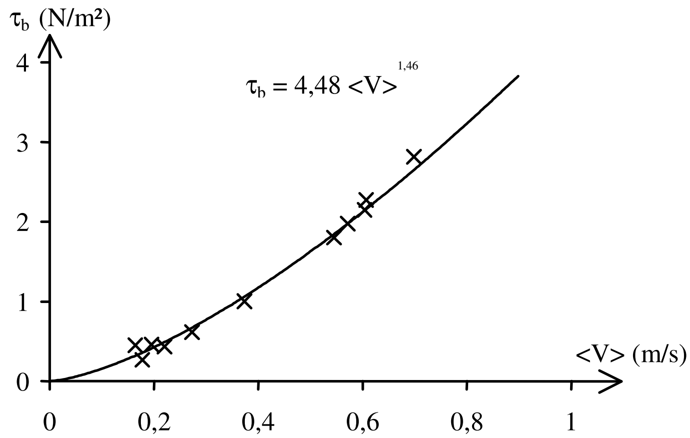


Fig. 17: Distribution of the shear stress between the particle layer and the pipe wall as a function of the mean velocity

CONCLUSION

Flows of concentrated slurries of polypropylene particles through straight pipe have been studied. The diameter of the particles was about 3mm and the density close to 869. Different flow patterns can be observed in fully turbulent conditions. The transition between the flow pattern depends on the Froude number. Heterogeneous suspension flow are present in the piping elements for Froude number greater than 15. Moving bed takes place in the straight pipes for Froude number between 15 and 0.2 and stationary beds appear for lower Froude number.

The friction losses depend on the flow pattern: for heterogeneous flow suspension flows and solid concentration below 30%, they are quite similar to the one obtained with liquid flows. The Blasius law is then appropriate to compute pressure losses and the slurry exhibits Newtonian behaviour. Conversely for flows with moving beds, the friction losses are greater and depend on the solid concentration, the flow rate, the pipe diameter and the Froude number.

Different methods have been used to predict pressure losses though straight pipe. Two correlations based on Froude number and Reynolds number have been tested and give satisfactory results. A Bingham fluid model has been used to compute extended Reynolds number. The Blasius law computed by replacing the classical Reynolds number with the extended Reynolds number is found to be applicable to the two-phase flow in its full range.

Integral balance of mass and momentum based on local measurements show the existence of a significant liquid flow-rate within the particle layer. The shear stress between the particle layer and the pipe wall mainly depends on the mean velocity of the two-phase flow. Such observations should be introduced in the two-layer model tested.

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