ISSN(Online) : 2319-8753
ISSN (Print) : 2347-6710

International Journal of Innovative Research in Science, Engineering and Technology

(An ISO 3297: 2007 Certified Organization)

Vol. 5, Issue 7, July 2016

Friction Factor for Fluidized Dense Phase Pneumatic Conveying of Fine Particles

Shijo J.S., Niranjana Behera

School of Mechanical & Building Sciences, VIT University, Vellore, Tamilnadu, India

ABSTRACT:Fluidized dense phase pneumatic conveying through pipelines is dominated by complex interactions such as gas-particle, particle- particle and gas/particle-pipe wall. Pressure drop required for conveying is mostly dependent upon these interactions. Pressure drop can be predicted using a suitable correlation of solid friction factor. But existing correlations or equations of solids friction factor show certain limitations such as large error under scale up conditions. In the present research work an equation for solids friction factor has been developed. These equations have been used for the prediction of pressure drop for higher length or larger diameter pipelines.

KEYWORDS:Pneumatic conveying, Fluidized dense phase, Solids friction factor, Scale-up, Friction factor

I. INTRODUCTION

Fluidized dense phase is defined as the process of pneumatic conveying of fine particles or powders. In this conveying process total pressure drop (ΔP) required for conveying depends upon the frictional forces due to flow of gases and the frictional forces due to flow of solids. The main frictional forces involved during this process are gas-wall friction, gas-particle friction, particle-particle friction and particle-wall friction. The frictional pressure drop due to gases is usually calculated using Blasius correlation (Fox et al. [1]) of friction factor for gases.

Few correlations are available in the literature for calculating the magnitude of solids friction factor. Solids friction factor correlation for dilute phase conveying through vertical or horizontal pipelines was presented by Yang [2]. Klinzing [3] applied equations for particle velocity in order to model an equation of solids friction factor for a dilute mode of flow. For dilute flow, Rautiainen [4] conducted a comparative analysis of existing solids friction factor equations in the literature. Many Solids friction factor equations for dilute mode flow have been presented by the researchers, but only few correlations for fluidized dense phase conveying are available in literature (Stegmaier [5]; Weber [6]; Jones and Williams [7]; Mallick and Wypych [8,9]; Behera et al. [10]).

Solids friction factor depends on important parameters such as solids loading ratio (m^*), gas Froude number (Fr), solids Froude number (Fr_s), ratio of mean particle diameter to pipe diameter (d_p/D) and ratio of particle density to gas density (ρ_s/ρ_a). It may be noted that m^* is the ratio of solids mass flow rate (m_s) to air mass flow rate (m_a). Existing correlations for dense phase flow are presented in *table 1*. These correlations contain the non-dimensional gas Froude number which is calculated at the inlet or at the center distance of the pipeline. Solids loading ratio (which is the ratio of solids mass flow rate to air mass flow rate) does not vary along the length of the pipeline. But the superficial gas velocity and density changes when the flow occurs from inlet to outlet of the pipeline. Due to variation of these parameters the magnitude of solids friction factor varies along the length of the pipeline. Hence the local variation of parameters should be included for developing the correlation of solids friction factor.

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ISSN (Print) : 2347-6710

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Table1: Solids friction factor correlations

Author	Solids friction factor correlation
Stegmaier [5]	$\lambda_s = \frac{2.1 Fr_s^{0.5}}{m^{*0.3} Fr^2} \left(\frac{D}{d_p} \right)^{0.1}$
Weber [6]	$\lambda_s = \frac{2.98}{m^{*0.69} Fr^{1.6}}$
Jones and Williams [7]	$\lambda_s = \frac{83}{m^{*0.9} Fr_i^2}$
Mallick and Wypych [9]	$\lambda_s = 14.5(m^*)^{-0.63} Fr^{-2.01} \left(\frac{\rho_s}{\rho_a} \right)^{-0.207}$

In this work, the solids friction factor equation has been written using important flow parameters with certain number of coefficient or exponents. Optimum values of coefficient and exponents are determined for which the predicted pressure drop is minimum. Scaling to higher pipeline length or higher pipe diameter has been performed in order to predict the maximum conveying distance or maximum conveying pipe diameter.

II. EXPERIMENTAL WORK

Pneumatic conveying tests were performed in a system (of 173m long and 53 mm diameter pipeline (*figure 1*)) using fly ash (FA1) and alumina (AL1). The powder material kept inside the blow tank (of $1m^3$ volume and top discharge) is pressurized by supplying controlled air through a bank of sonic nozzles. Air supply is sub-divided into two ways: part of the air (called as primary air) is supplied through the bottom of blow tank to fluidize the material and another part of the air (secondary air) is supplied at some distance away from the feed point to avoid the blockage of material in the pipeline. Two pressure transmitters were used to measure the pressures of primary air and secondary air. Electrical signals from load cells and pressure transmitters were recorded by a SCXI-1102 data acquisition system (having 32 differential channels). Each pressure transmitter recorded the data with a sampling frequency between 80 and 100 Hz. Solids mass flow rates discharged from the blow tank or collected at the receiving hopper was measured by a series of load cells. Calibration of pressure transmitters and load cells were performed by a Barnett dead weight tester.

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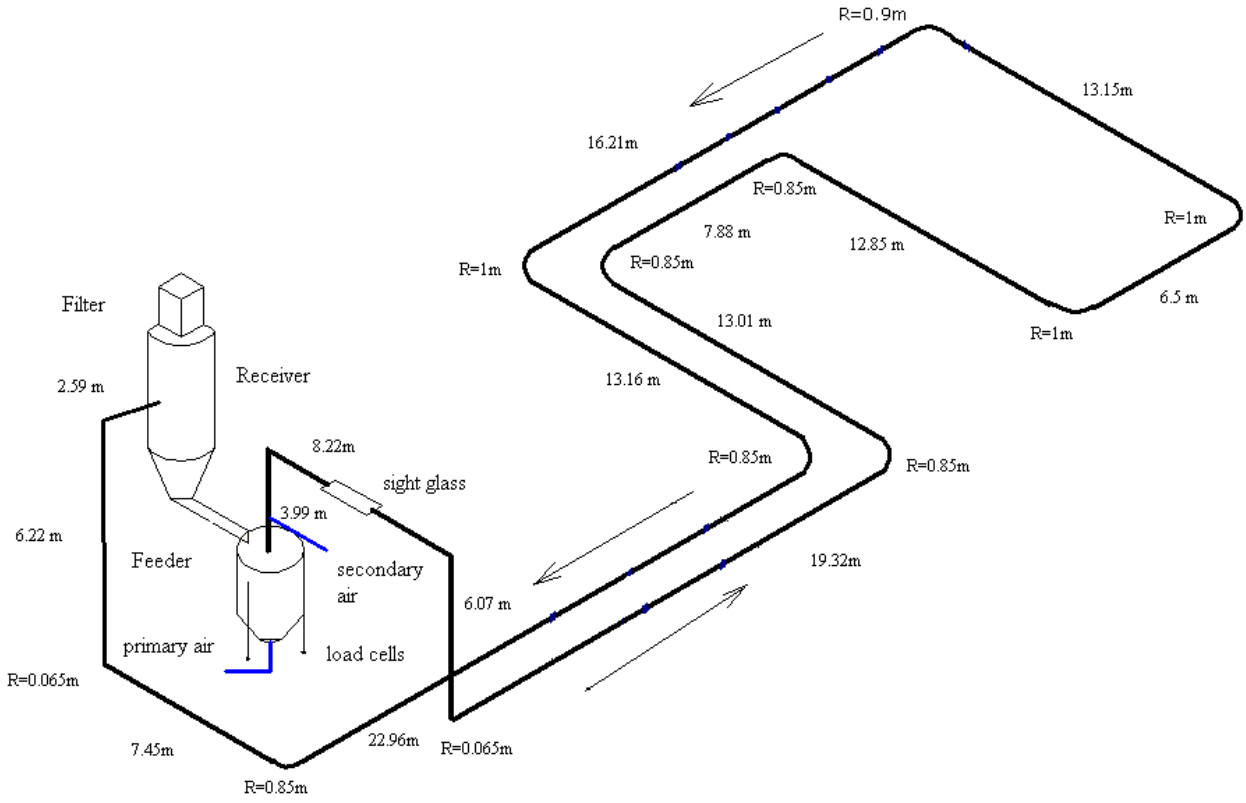

Fig. 1 Schematic diagram of pneumatic conveying set-up with 173 m pipeline configuration

In this work test data (for a pipeline configuration having a 173 m long and 53 mm diameter pipeline) of Williams [11] have been used for analysis. These test data were for three types of conveying materials: fly ash (FA2), alumina (AL2) and cement meal (CM). *Table 2* presents the properties of conveying materials.

Table 2: Conveying material properties

Conveying material	Mean particle diameter (μm)	Particle density (kg/m^3)	Loose-poured bulk density (kg/m^3)	Slip ratio
Fly ash (FA1)	14.91	2096	724	0.939
Alumina (AL1)	37.11	3810	988	0.906
Fly ash (FA2)	19	2530	810	0.931
Alumina (AL2)	79	3300	1050	0.879
Cement meal (CM)	11	3000	930	0.937

III. MATTHEMATICAL MODEL

Although solids friction factor (λ_s) depends on many non-dimensional parameters, but in this model only three important non-dimensional parameters have been used. The equation for this factor has been written by using these three parameters, one coefficient (C) and three exponents (a, b, e).

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$$\lambda_s = \frac{C}{(m^*)^a Fr^b \left(\frac{\rho_s}{\rho_a}\right)^e} \tag{1}$$

Where, ρ_s and ρ_a are the particle density and gas density respectively.

For this model, separate flow continuity equations for gas or solids phase and a combined momentum equation for gas-solids mixture have been written for a straight section of pipeline (having cross-sectional area A and length Δx). Equations (2) to (5) have been written based on following assumptions.

(i) Flow is under steady state
(ii) The Gas phase obeys the ideal gas equation of state.
(iii) No transfer of mass or heat takes place among the phases.
(iv) System temperature is constant.
(v) Flow is dominated in x-direction only.

3.1 Gas phase continuity equation

Gas phase continuity equation is

$$\frac{d}{dx}(\varepsilon_a \rho_a u_a) = 0 \tag{2}$$

Where, ε_a and u_a are the void fraction and the gas velocity respectively.

3.2 Solids phase continuity equation

Solids phase continuity equation is

$$\frac{d}{dx}(\varepsilon_s \rho_s u_s) = 0 \tag{3}$$

Where, ε_s and u_s are the solid volume fraction and the particle velocity respectively.

3.3 Conservation of momentum equation for combined gas-solids phase

Forces involved in a fluidized dense phase conveying through a section of pipeline of length Δx and area of cross-section A have been shown in *figure 2*.

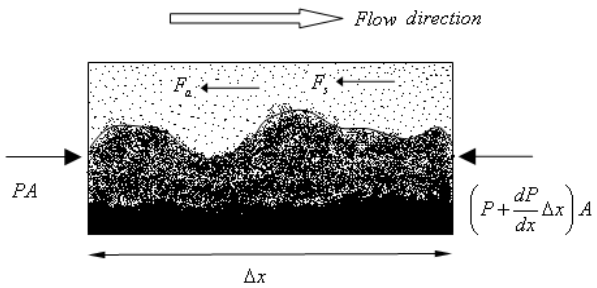

Fig. 2 Forces acting in a dense phase flow through a horizontal pipeline section

Mixture momentum equation is

Rate of change of momentum

= Pressure force – Frictional force due to gas (F_a) – Frictional force due to solids (F_s)

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i.e.

$$\left[(\rho_a \varepsilon_a A u_a)\left(u_a + \frac{du_a}{dx}\Delta x\right) - (\rho_a \varepsilon_a A u_a)u_a\right] + \left[(\rho_s \varepsilon_s A u_s)\left(u_s + \frac{du_s}{dx}\Delta x\right) - (\rho_s \varepsilon_s A u_s)u_s\right] = \left[PA - \left(P + \frac{dP}{dx}\Delta x\right)A\right] - F_a - F_s \quad (4)$$

Where, F is the frictional force.

Dividing both sides of equation (4) by elemental volume ($A\Delta x$)

$$\varepsilon_a \rho_a u_a \frac{du_a}{dx} + \varepsilon_s \rho_s u_s \frac{du_s}{dx} + \frac{dP}{dx} = -f_a - f_s \quad (5)$$

Where, f is the friction force per unit volume.

3.4 Particle velocity equation

Specific particle velocity equation is not available in the literature for fluidized dense phase conveying. In this model, a particle velocity equation applicable for dilute phase (Chambers and Marcus [12]) has been used.

$$\frac{u_s}{u_a} = 1 - 0.008\left(1000 d_p\right)^{0.3} \rho_{bl}^{0.5} \quad (6)$$

3.5 Equation of state

Ideal gas equation can be used to calculate the gas pressure.

$$P = \rho_a RT \quad (7)$$

3.6 Friction force equation

Air only frictional pressure drop can be calculated using the Darcy-Weisbach's equation (Cengel and Cimbala [13]).

$$\Delta P_{fa} = \lambda_a \frac{\rho_a u_a^{\,2}}{2} \frac{\Delta x}{D} \quad (8)$$

Neglecting the roughness of pipe wall, the air only friction factor can be calculated using Blasius equation which is applicable for turbulent flows with Reynolds number less than or equal to 10^5 (Fox et al. [1]).

$$\lambda_a = \frac{0.316}{\mathrm{Re}_a^{0.25}} \quad \text{Where, } \mathrm{Re}_a = \frac{\rho_a u_a D}{\mu} \quad (9)$$

Gas frictional force per unit control volume can be written as

$$f_a = \frac{\Delta P_{fa} \varepsilon_a A}{A \Delta x} \quad (10)$$

Solids only frictional pressure drop can be written like Fanning equation (Barth [14], Wypych and Arnold [15], Jones and Williams [7], Huang et al. [16], Molerus [17]).

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$$\Delta P_{fs} = m * \lambda_s \frac{\rho_a u_a^{\,2}}{2} \frac{\Delta x}{D} \tag{11}$$

Solids frictional force per unit control volume can be written as

$$f_s = \frac{\Delta P_{fs} \varepsilon_s A}{A \Delta x} \tag{12}$$

3.7 Equations for solids volume fraction and void fraction

Initial air and solids volumetric flow rate can be calculated using equations (13) and (14) respectively.

$$\dot{V}_{a0} = \frac{m_a RT}{P} \tag{13}$$

$$\dot{V}_{s0} = \frac{m_s}{\rho_s} \tag{14}$$

If $A\Delta x$ is the elemental control volume of the pipeline, then $\left(V_{a0}/u_{a0}\right)\Delta x$ and $\left(V_{s0}/u_{s0}\right)\Delta x$ are the volume occupied by gas and solids respectively. Hence equations for initial solids volume fraction and void fraction can be expressed as given by Rudinger [18].

$$\varepsilon_{s0} = \frac{\dot{V}_{s0}/u_{s0}}{\dot{V}_{s0}/u_{s0} + \dot{V}_{a0}/u_{a0}} \quad \text{and} \quad \varepsilon_{a0} = 1 - \varepsilon_{s0} \tag{15}$$

IV. NUMERICAL SIMULATION

Flow equations are solved for the 173 m long straight pipeline by divided it into small sections and using the using the initial conditions of flow parameters at the exit of pipeline section. Flow equations are solved starting from the exit to inlet of the 173 m long pipeline. Unknown parameters P, u_a and ε_a in the equations (2), (3) and (5) can be solved using the fourth order Runge-Kutta method. Objective of solving these equations is to calculate an optimum values of C, a, b and e. Williams [11] have analyzed that the coefficient and exponents vary within a range of values for different conveying materials. Based on these range of values, 44550 numbers of sets of coefficient and exponents have been formed. For the calculation of solids friction force these sets have been used. Optimum data set of C, a, b and e are based on the minimum average error. Equations are solved using the fourth order Runge-Kutta method by writing a Matlab code.

Table 3: Range of values of Coefficient (C) and exponents (a, b and e)

Coefficient or Exponent	Range of values
C	10 to 90
a	0.1 to 1
b	1 to 2
e	0.1 to

V. PREDICTED PRESSURE VS EXPERIMENTAL PRESSURE

For the simulation, test data of five categories of powder materials have been used. Test results of void fraction, superficial gas velocity and absolute pressure are used as initial conditions to carry out the simulation. From the simulation, optimum data set of coefficient and exponents are determined, which are presented in *table 4*. For each

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conveying material, the values of optimum data set were different except the exponent *e* which was same for all the conveying materials.

Table 4: Calculated optimum values of coefficient *C* and exponents (*a*, *b* and *e*) of solids friction factor

Conveying material	Coefficient *C*	Exponent *a*	Exponent *b*	Exponent*e*
Fly ash (FA1)	24	0.1	1.2	0.1
Alumina (AL2)	22	0.1	1	0.1
Fly ash (FA2)	24	0.1	1.5	0.1
Alumina (AL2)	20	0.4	1	0.1
Cement meal (CM)	24	0.2	1	0.1

Predicted and experimental inlet pressure data of fly ash (FA1) are plotted in *figure 3*. Error of data points lie within ±20%, with an average error and standard deviation of error 14.6 % and 7 % respectively.

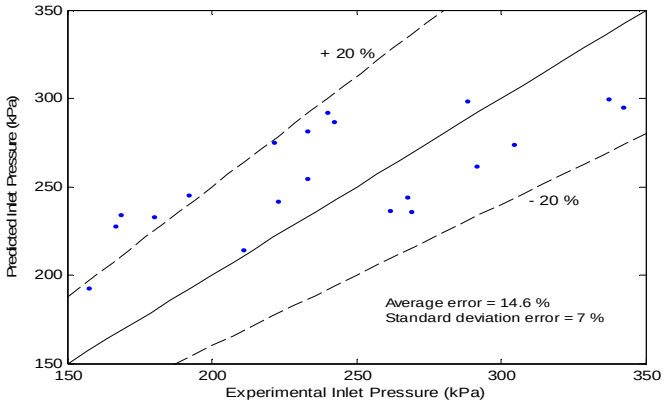

Fig. 3 Comparison of predicted and experimental inlet pressure values for Fly ash (FA1)

Predicted and experimental inlet pressure data of alumina (AL1) are plotted in *figure 4*. Error of data points lie within ±20%, with an average error and standard deviation of error 15.65 % and 9.41 % respectively.

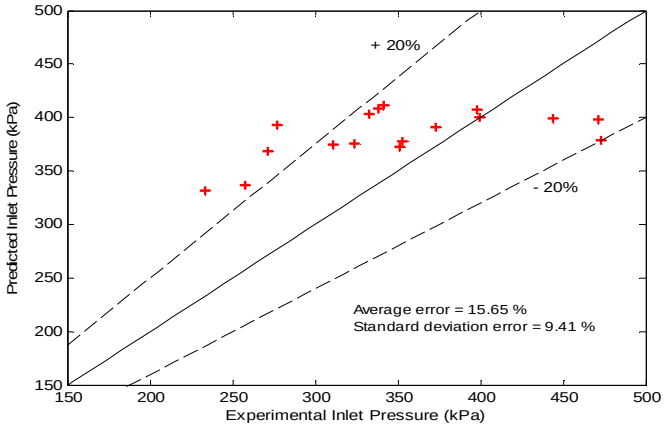

Fig. 4 Comparison of predicted and experimental inlet pressure values for Alumina (AL1)

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Predicted and experimental inlet pressure data of fly ash (FA2) are plotted in *figure 5*. Error of data points lie within ±20%, with an average error and standard deviation of error 15.86 % and 7.39 % respectively.

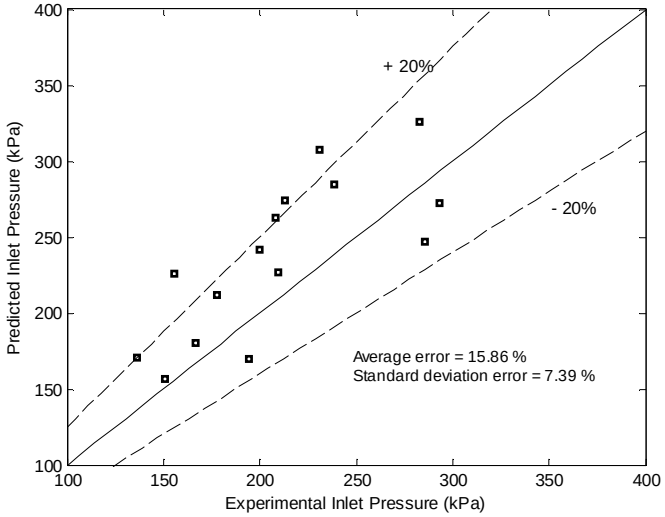

Fig. 5 Comparison of predicted and experimental inlet pressure values for fly ash (FA2)

Predicted and experimental inlet pressure data of alumina (AL2) are plotted in *figure 6*. Error of data points lie within ±20%, with an average error and standard deviation of error 12.55 % and 12.61 % respectively.

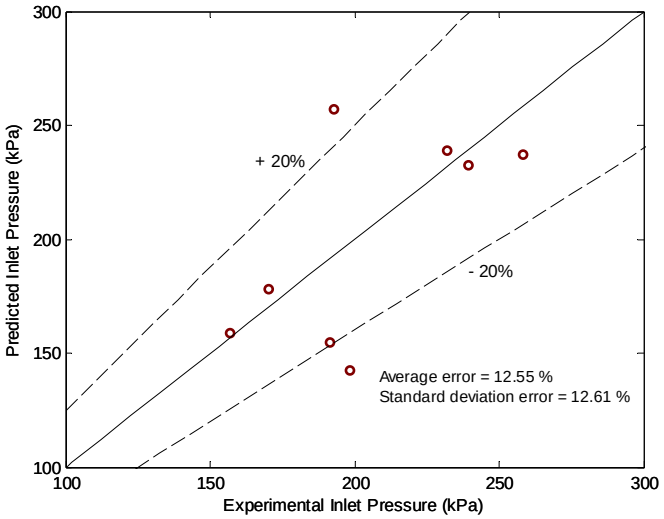

Fig. 6 Comparison of predicted and experimental inlet pressure values for alumina (AL2)

Predicted and experimental inlet pressure data of cement meal (CM) are plotted in *figure 7*. Error of data points lie within ±20%, with an average error and standard deviation of error 5.44 % and 5.06 % respectively.

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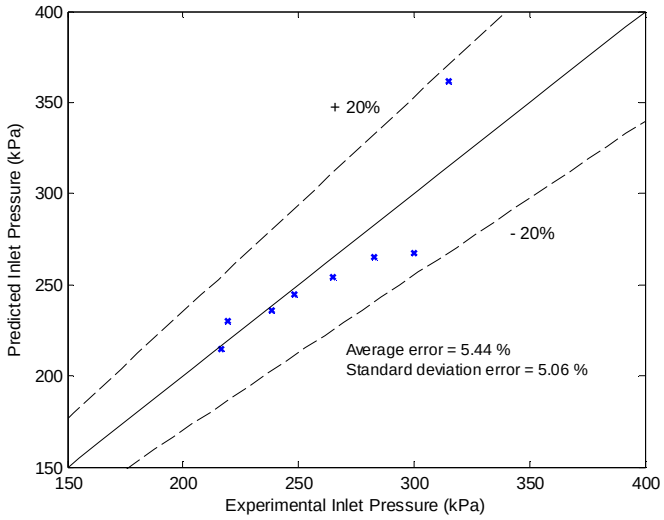

Fig. 7 Comparison of predicted and experimental inlet pressure values for cement meal (CM)

VI. FLOW PARAMETER VARIATION ALONG THE LENGTH OF PIPELINE

In this section, variations of different parameters along the length of 173 m long pipeline have been analyzed.

a. Absolute pressure, gas velocity and void fraction

Variations of flow parameters absolute pressure, void faction and gas velocity are plotted in *figure 8.* Absolute pressure decreases, but void fraction and gas velocity increase along the length of the pipeline. Gradual expansion of gas from inlet to exit causes increase in the gas velocity and the void fraction. As flow occurs from upstream to downstream, the material gets aerated and at some location the flow may also be converted from dense to dilute.

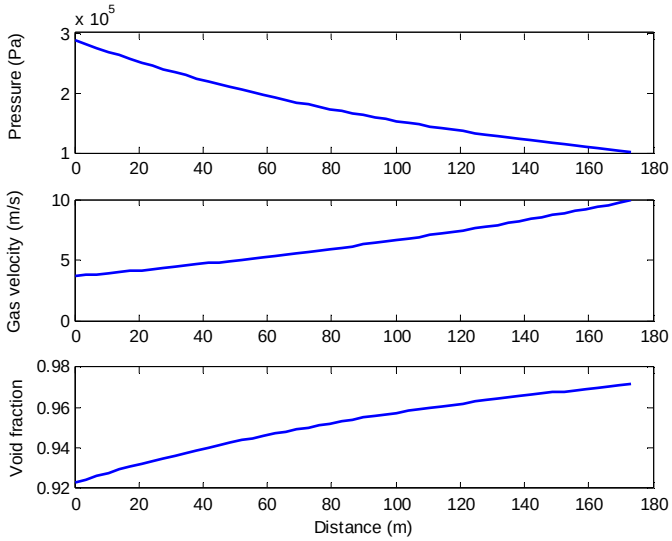

Fig. 8 Variation of flow parameters along the length of pipeline

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b. Pressure drop ratio

Total pressure drop is the sum of air only pressure drop (ΔP_{fa}) and solids pressure drop (ΔP_{fs}). Variation of solids to air pressure drop ratio along the length of the pipeline is plotted in *figure 9*. The ratio is more for a dense flow as compared to a dilute flow. For a dilute mode of flow, the air only pressure drop is almost equal to the solids pressure drop. Hence for a dilute flow, the ratio does not change along the length of the pipeline.

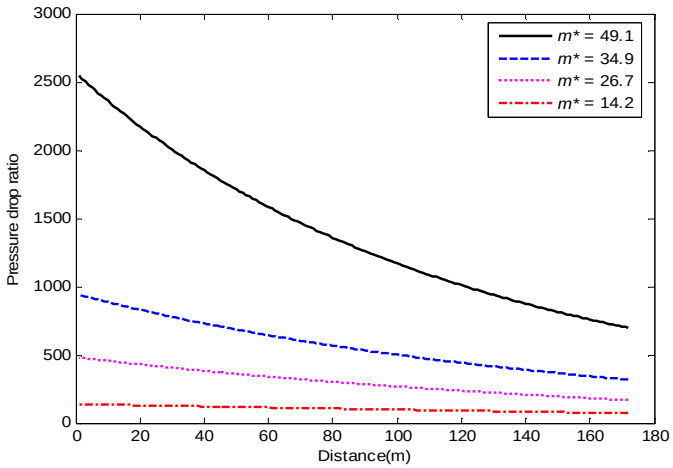

Fig. 9 Variation of pressure drop for fly ash (FA1)

$$\text{Pressure drop ratio} = \frac{\Delta P_{fs}}{\Delta P_{fa}} \tag{16}$$

c. Friction factor ratio

Variation of solids friction factor (λ_s) to air friction factor (λ_a) ratio along the length of the pipeline is plotted in *figure 10*. For a dense flow the ratio is higher as compared to a dilute flow. For a dilute flow, the difference between the values of solids and air friction factor are negligible.

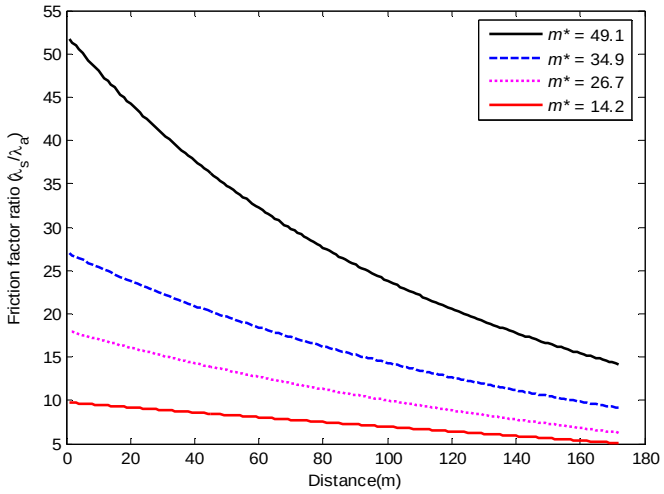

Fig. 10 Variation of friction factor ratio for fly ash (FA1)

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$$\text{Friction factor ratio} = \frac{\lambda_s}{\lambda_a} \quad (17)$$

d. Specific power consumption

Power consumed per unit mass of conveying material is the specific power consumption. Variation of local specific power consumption along the length of the pipeline is presented in *figure 11*. For the case of dilute flow (m^* = 14.2), local specific power consumption increases since the flow becomes more and more dilute when flow moves towards downstream of the pipeline. Fora a dense flow (m^* = 49.1) the specific power consumption does not vary much significantly along the length of the pipeline. For a dense phase flow (m^* = 26.7), the specific power consumption does not vary up to certain distance, but beyond a location it increases which may be due to change in flow mode from dense to dilute.

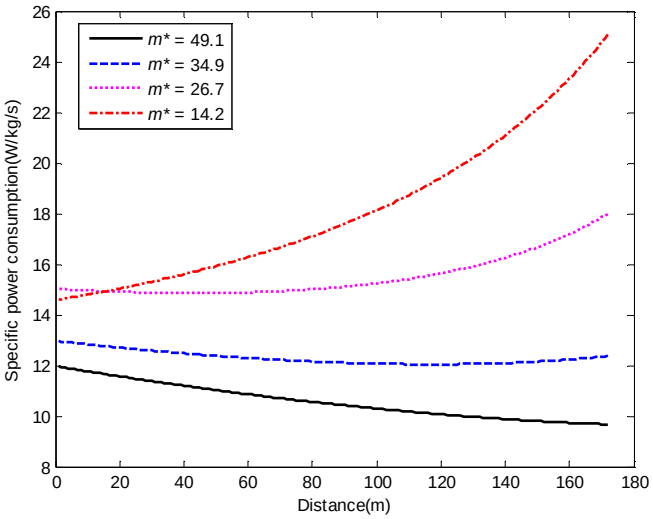

Fig. 11 Variation of specific power consumption for fly ash (FA1)

VI. LENGTH AND DIAMETER SCALE-UP ANALYSIS

The pipeline configuration used for a pilot plant setup is of smaller length or diameter as compared to the plant pipeline configuration. The experimental data for the pilot plant can be scaled-up to achieve the experimental data of a plant pipeline setup. In this way it is possible to calculate the maximum possible conveying distance or maximum possible conveying diameter for flow condition of constant solids or air mass flow rate.

Few researchers (Mallick and Wypych [8]) have suggested that there is a minimum value of superficial gas velocity or gas Froude number below which the conveying is possible. This value varies with conveying material. The minimum Froude number ($Fr_{i,min}$) for cement and fly ash ranges from 4 to 5 and 3 to 5 respectively. For the simulation, this minimum value (which is occurring at the inlet of the pipeline) is taken to be 4.5 and 4 for cement meal (CM) and fly ash (FA1) respectively. On the other hand, The solids volume fraction should not exceed the value of packing limit $\varepsilon_{s,max}$ = 0.63. But in the present simulation, the allowed solids volume fraction (for which pipeline blockage does not occur) at the inlet of pipeline is equal to or less than 0.55.

Above two limiting conditions were applied in the simulation, for the prediction of the maximum conveying distance (using a 53 mm pipe diameter) in section 7.1 and the maximum conveying diameter (using a 173 m long pipeline) in section 7.2.

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a. Calculating the maximum conveying distance by length Scale-up

Predicted pressure drop for different conveying distances or pipeline lengths under different flow conditions for cement meal (CM) have been plotted in *figure 12*. In this figure, four curves have been shown which are for different air mass flow rates, but for the same solids mass flow rates. It can be observed that the required pressure drop increases with increase in pipeline length.

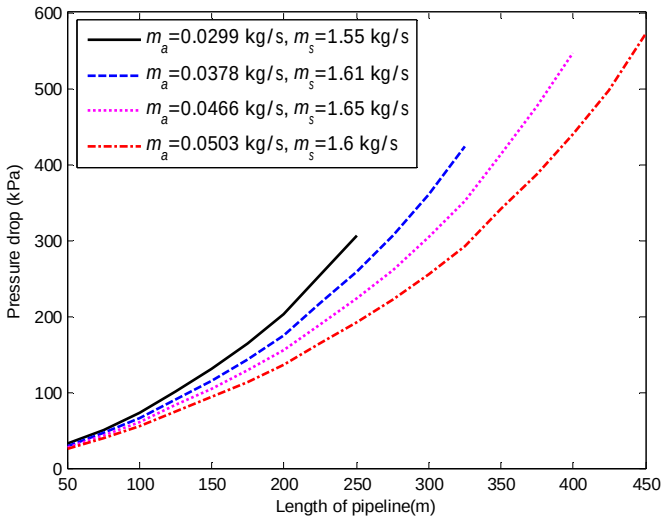

Fig. 12 Pressure drop vs pipeline length for cement meal under different flow conditions

For further analysis, the predicted pressure drop for different conveying distances or pipeline lengths under different flow conditions for fly ash (FA1) have been plotted in *figure 13*. In this figure, four curves have been shown which are for different air mass flow rates, but for the same solids mass flow rates. It can be observed that the required pressure drop increases with increase in pipeline length.

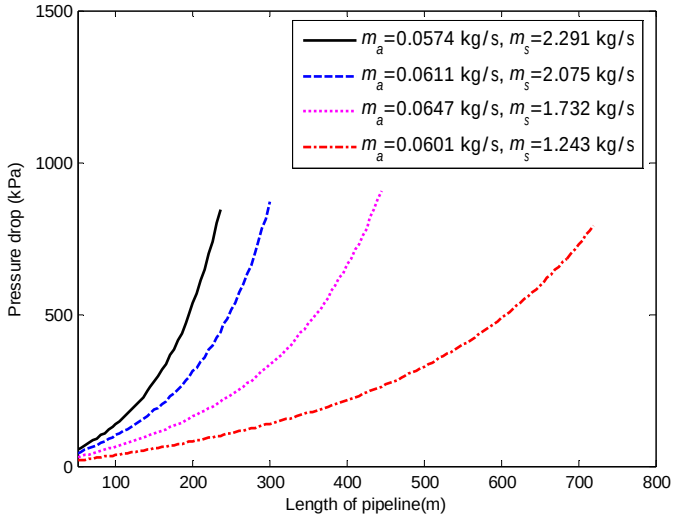

Figure 13: Pressure drop vs pipeline length for fly ash (FA1) under different flow conditions

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b. Calculating the maximum conveying pipe diameter by diameter scale-up

Predicted pressure drop for different conveying pipeline diameters under different flow conditions for cement meal (CM) have been plotted in *figure 14*. In this figure, four curves have been shown which are for different solids mass flow rates, but for the same air mass flow rates. It can be observed that the required pressure drop decreases with increase in pipeline diameter.

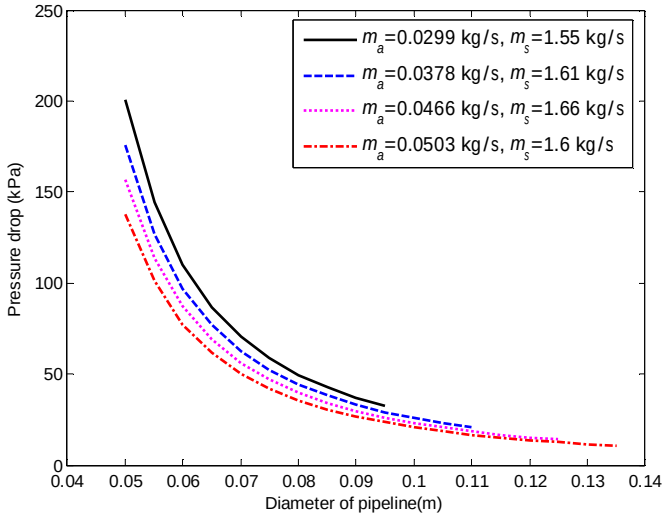

Fig. 14 Pressure drop vs pipeline diameter for cement meal (CM) under different flow conditions

For further analysis, the predicted pressure drop for different conveying pipeline diameters under different flow conditions for fly ash (FA1) have been plotted in *figure 15*. In this figure, four curves have been shown which are for different solids mass flow rates, but for the same air mass flow rates. It can be observed that the required pressure drop decreases with increase in pipeline diameter.

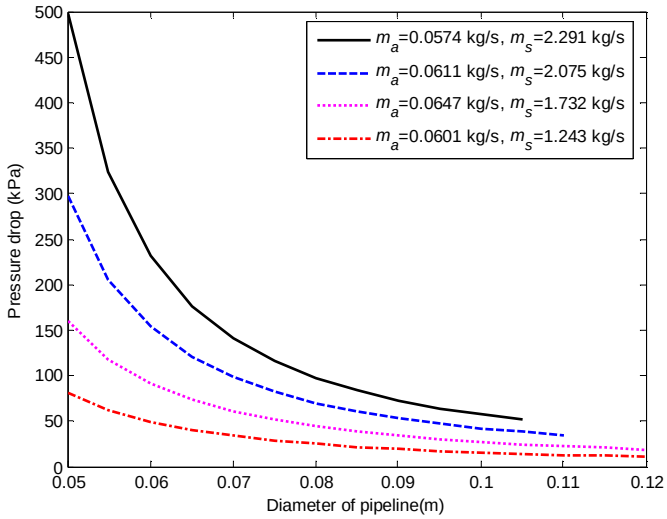

Fig. 15 Pressure drop vs pipeline diameter for fly ash (FA1) under different flow conditions

ISSN(Online) : 2319-8753
ISSN (Print) : 2347-6710

International Journal of Innovative Research in Science, Engineering and Technology

(An ISO 3297: 2007 Certified Organization)

Vol. 5, Issue 7, July 2016

VII. CONCLUSION

A model of solids friction factor has been developed which consists of a number of flow equations. Solids friction factor has been written by using certain numbers of non-dimensional numbers such as gas Froude number, solids loading ratio and particle-gas density ratio. The flow equations are solved to calculate the optimum values of coefficient and exponents used in the expression for solids friction factor. Comparing the coefficient and exponents for five conveying materials, it has been found that they have a common exponent of density ratio. Conveying limiting conditions have been imposed to predict the maximum conveying distance or pipeline length and maximum pipeline diameter. It has been observed that the required pressure drop increases with increase in pipeline length and decreases with increase in pipeline diameter.

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ISSN(Online) : 2319-8753
ISSN (Print) : 2347-6710

International Journal of Innovative Research in Science, Engineering and Technology

(An ISO 3297: 2007 Certified Organization)

Vol. 5, Issue 7, July 2016

BIOGRAPHY

Shijo J.S. graduated in 2010 from Kerala University, Thiruvananthapuram in India with a first class Degree in Mechanical Engineering. He Acquired Master's Degree in 2012 as a topper in Machine Design at SCT College of Engineering & Technology, Thiruvananthapuram under Kerala University. He has worked as Assistant Professor at UKF College of Engineering & Technology, Kollam, Kerala. Since 2014 he has been doing PhD studies at VIT University, Vellore. He also attended several conferences in PAN India level related to his PhD studies.

Dr. Niranjana Behera graduated in 1998 from VSSUT University, Burla in India with a first class degree in Mechanical Engineering. He first worked as a lecturer in MITS, Rayagada, Orissa, India. He undertook M.Tech programme at IIT, Kharagpur with specialization Mechanical system Design in 2004. After M.Tech he worked in JITM, Paralakhemundi, Orissa, India and ITER, Bhubaneswar, Orissa, India. He obtained PhD degree from IIT Delhi. In 2009 he visited University of Newcastle, Australia for six months to carryout experimental work on pneumatic conveying. Presently he is working as Associate Professor at VIT University, Vellore (India).