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On developing improved modelling and scale-up procedures for pneumatic conveying of fine powders

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ABSTRACT

Pneumatic transport of fine powders in fluidized dense-phase mode is becoming increasingly popular in various industries, such as power, chemical, cement, refinery, alumina, pharmaceutical, limestone, to list a few, because of the reasons of reduced gas flow rate and power consumption, decreased conveying velocities, improved product quality control, reduced pipeline sizing and wear rate, increased workplace safety etc. For the reliable design of a pneumatic conveying system, it is important to accurately predict the total pipeline pressure drop. However, accurate prediction of pressure drop from an improved understanding of the fundamental transport mechanism of fluidized dense-phase flow condition has only made limited progress till now because of the highly concentrated and turbulent nature of the gas-solids mixture. Power plant fly ash (median particle diameter: 30 μm; particle density: 2300 kg/m^3; loose-poured bulk density: 700 kg/m^3) was conveyed through different pipelines (69 mm I.D. × 168 m long; 105 mm I.D. × 168 m long; 69 mm I.D. × 554 m long). 8 different fly ash samples were tested in a fluidizing column for their deaeration characteristics and fluidized bulk densities were determined. Governing equations of flow for the dense-phase pneumatic conveying system of fine powders were solved using Runge-Kutta-Fehlberg (RKF45) method for different fluidized bulk densities of fly ash and air flow rates. The results have shown that the particle and actual gas velocities and the ratio of the two velocities increase in the direction of flow, while a reverse trend was apparent for the solids volumetric concentration. The results were compared against the predictions obtained using existing empirical relations for particle velocity. To develop an improved model for solids friction factor, an existing reliable pure dilute-phase model has been modified for dense-phase flow condition by incorporating sub-models for particle to actual gas velocity and impact and solids friction factor. The developed solids friction factor model was validated by using it to predict the total pipeline pressure drops for larger and longer pipelines and by comparing the experimental and predicted pneumatic conveying characteristics. The results have shown improved reliable predictions and that the model is capable of addressing the gradual transition of flow mechanism from dense- to dilute-phase. The accuracy of prediction is similar (in fact better in certain scale-up cases) when compared to a recently developed two-layer based model (developed by some of the authors). The results demonstrate the importance of incorporating particle and actual gas velocity terms in the model of solids friction factor instead of the prevailing techniques that overly depend on using superficial gas velocities.

1. Introduction

The dense-phase pneumatic transport of fine powders is preferred in several industries, such as coal fired thermal power plants, chemical, pharmaceutical, petrochemical plants, cement, food, etc. over conventional dilute-phase (or suspension flow) due to the reasons of high solids to gas mass ratio, low gas flow requirement (i.e. smaller compressor and savings in operating power), smaller size of pipes and fittings, reduced conveying velocity resulting in lowering of wear rate of pipelines and bends, decreased rate of product attrition, reduced size of solids-gas separation equipment, etc. [1,2]. In this mode of transport, the gas velocities are kept sufficiently low (below the saltation velocities), so that the conveying takes place in a non-suspension mode [1, 2]. Typically, the Geldart Group A type fine powders that have good air retention capabilities, such as fly ash, cement, pulverized coal, etc., are good candidates for dense-phase mode of conveying [3]. Designing a dense-phase pneumatic conveying system includes accurate prediction of the total pipeline pressure loss and the minimum air flow rate

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requirement to ensure flow blockage does not take place. Under-prediction of pressure drop at the design stage would result in reduced material flow rate, whereas over-prediction of pressure drop would result in increased initial and operating costs (due to unnecessary higher conveying velocity). The pressure loss for solids-gas flow through a straight horizontal pipe section can be expressed using Eq. (1), as given by Barth [4].

$$\Delta P = \left((\lambda_f + m^* \lambda_s)\rho_a \, L \, V^2\right)/2 \, D \tag{1}$$

Equation 1 considers the pressure losses due to the gas and solids phases separately. Weber [5] employed this expression (Eq. (1)) for coarse particles in dilute-phase flows. However, various researchers [2,6–10] have subsequently used this representation to estimate the pressure drop for the dense-phase pneumatic transport of fine powders, such as fly ash, pulverized coal, ESP dust, etc. for horizontal straight pipes. In this model, while all other terms can be calculated relatively easily using well established gas only friction factor formula [11], accurate prediction of solids friction factor has been a difficult task due to the limited fundamental understanding of the flow mechanisms of powdered beds. The term solids friction factor is a combined representation of loss of energy due to solids to gas, solids to solids, and solids to pipe wall interactions [1,2]. Whereas, more fundamental modelling methods based on powder mechanics have been developed for certain products and modes of flows, such as low-velocity slug-flow of granular products [12–14] or the dilute-phase flow of bulk solids (for which the suspension flow mechanics can be considered to be applicable with a fair accuracy) [15], the modelling of solids friction factor for the dense-phase conveying of fine powders has been a far more difficult problem to be solved at a similar level of detail. This is due to the highly turbulent and complex nature of dense phase mode of conveying [2], which makes it very difficult to link the particle and bulk properties during actual flow conditions. Because of the limited progress that has been achieved so far towards fundamentally understanding the transport mechanisms and modelling of solids friction factor, empirical power function based models have been popularly used over the years by several investigators [2,6–10], using dimensionless numbers, as given by Eqs. (1) and (2):

$$\lambda_s = K(m^*)^a (Fr)^b \tag{2}$$

$$\lambda_s = K(m^*)^a (Fr)^b (\rho_a/\rho_s)^c \tag{3}$$

Previous investigation [2] has demonstrated that the above formats of modelling solids friction factor can unexpectedly result in grossly inaccurate predictions under significant scale-up conditions of pipeline length and diameter. Very recently, some of the authors have provided a new two-layer based model format for solids friction factor [16], which includes separate contributory terms for the non-suspension and suspension flows. The format is provided in the following:

$$\lambda_s = \tau_1 \left(K(VLR)^a (w_{fo}/V)^b\right) + \tau_2 \left(\lambda_s^* C/V + 2\beta_0/\left[(C/V)Fr^2\right]\right) \tag{4}$$

In the above model, the first term, $K(VLR)^a(w_{fo}/V)^b$, seems to better represent the dense-phase contribution [10,11,16], whereas the second term, $\lambda_s^* C/V + 2\beta_0/[(C/V)Fr^2]$, represents the dilute-phase contribution [15,16]. β_0 is given as w_{fo}/V. This portion has been taken from a pure dilute-phase model, known as "Weber A4" model [15]. This model (Weber A4) has been reported [2,15] to provide good predictions for dilute-phase flow for different pipeline configurations. This model separates the solids friction factor into losses contributed due to the impact and friction between particle to air/particle/pipe-wall and that due to keeping the particles in suspension. τ_1 and τ_2 represent coupling factors, whose values depend on the Froude number (based on superficial air velocity). The two-layer model format was validated for two different fly ash samples, ESP dust and cement under significant scale-up conditions of pipeline diameter and length (viz. pipe internal diameters of 69, 80, 105 mm and lengths of 168, 254, 407 and 554 m). The results indicated that the two-layer model could provide better reliable scale-up predictions than other previously known models. The predicted pneumatic conveying characteristics provided relatively better 'U'-shaped curves that possibly represent the dense- to dilute-phase flow transition [2]. In spite of such recent developments, the authors are of the opinion that very little progress has been achieved so far towards fundamentally understanding the transport mechanism of dense-phase solids-gas flow of fine powders and the solids friction factor models rely heavily on the empirical approaches involving the use of dimensionless numbers. The existing models [2,6–10,16] for solids friction factor use "superficial gas velocity", instead of actual gas velocity. Actual gas velocity can be considered to be adequately represented by superficial gas velocity for dilute-phase flow, where the pipe volume occupied by the particles is minor compared to the rest of the pipe volume available for gas flows. However, for dense-phase type flows, due to higher concentration of solids, the lack of cross section available for the flow of gas-phase should not be ignored. Also, most of the existing models (e.g. Eqs. (2) and (3) do not address particle velocity and the slip velocity (the difference between gas and particle velocities) and rely on the superficial velocity of the gas. The authors consider that it would be prudent to include a particle velocity term in modelling solids friction factor. The aim of this paper is to solve governing equations for gas-solids flows (under dense-phase condition) for important parameters such as particle and actual gas velocities and make use of these parameters in developing reliable scale-up validated model for solids friction factor. Some studies using numerical methods have been carried out in recent times [17,18]. One-dimensional model including particle size distribution was developed [19] to simulate the pneumatic conveying of fine powders through straight pipelines. A set of one-dimensional flow equations were solved in order to predict the flow parameters at the inlet of the pipeline by using the experimental data as initial conditions at the exit of the pipeline. Using the developed model, pressure drop values through the pipeline were predicted and compared with experimental data. Predicted values were found to be in good agreement with the experimental data. One of the key aspect that requires attention is particle velocity [20]. Effects of material properties such as friction and restitution coefficients of particles on flow behaviors in terms of particle flow pattern, gas pressure drop, solid concentration, particle velocity and transition of flow regime were studied [21]. However, such studies have considered particle velocity relation applicable only for dilute-phase (and not applicable for the dense-phase flow), which would tend to provide significant over-estimation of particle velocity when applied to dense-phase cases. The present paper attempts to provide solution for particle velocity for the dense-phase flow condition. It must be also noted that experimental measurement of particle velocity during actual flow condition is a difficult task and in the knowledge of the authors, very little research has been conducted to develop a reliable measurement method. Sight glass observations only provide limited information due to the sticking of the powders on the inner surface of glass. Most of the existing models (majorly empirical) for particle velocity are applicable for coarse particles being transported in dilute-phase [21].

2. Experimental setup

Power station fly ash was conveyed from dense- to dilute-phase through pipelines of different lengths and diameters (i.e. 69 mm I.D. × 168 m long, 105 mm I.D. × 168 m long, 69 mm I.D. × 554 m long) at the Bulk Materials Handling Laboratory of University of Wollongong, Australia. The fly ash was conveyed with different air and solids mass flow rates. Physical properties of the fly ash and pipeline conditions are given in Table 1. Particle size distributions were measured using