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Modeling solids friction factor for fluidized dense-phase pneumatic transport of powders using two layer flow theory

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ABSTRACT

Fluidized dense-phase pneumatic conveying of powders is becoming increasingly popular in various industries, such as power, chemical, cement, refinery, alumina, pharmaceutical, limestone, to list a few, due to the reasons of reduced gas flows and power consumption, improved product quality control, reduced pipeline sizing and wear rate, increased workplace safety etc. An accurate estimation of total pipeline pressure drop is of paramount importance for the reliable design of a pneumatic conveying system. However, because of the highly concentrated and turbulent nature of the gas–solids mixture, fundamentally understanding the flow mechanism and accurately predicting the pressure drop as an important design parameter has only made limited progress so far. This paper results from an ongoing investigation into developing a validated modeling procedure for solids friction factor for the accurate prediction of pressure drop and optimal operating conditions for fluidized dense-phase pneumatic conveying systems. Under the present study, a two-layer based model has been developed by separately considering the solids friction contributions of the non-suspension (dense) bed of powders flowing along the bottom of pipe and the suspension (dilute-phase flow) of particles occurring on top of the non-suspension layer. Volumetric loading ratio and dimensionless velocity have been used to model the non-suspension dune flow layer. A solids impact and friction term and dimensionless velocity have been employed to model the dilute-phase flow due their established reliability. Models have been developed using the straight-pipe conveying data of two types of fly ash, cement and ESP dust (median particle diameter: 7 to 30 μm; particle density: 2300 to 3637 kg/m^3; loose-poured bulk density: 610–1080 kg/m^3). The developed models for solids friction were validated for their scale-up accuracy by using them to predict the pressure drops in five larger and longer pipelines (69 mm I.D. × 168 m long; 105 mm I.D. × 168 m long; 69 mm I.D. × 554 m long, 65 mm I.D. × 254 m long and 80/100 mm I.D. × 407 m long pipes) and by comparing the experimental versus predicted pneumatic conveying characteristics. The two-layer model provided improved accuracy compared to existing models indicating that the model is able to adequately address the dense- to dilute-phase transition criteria.

1. Introduction

Dilute-phase pneumatic transport of powders has been conventionally used for many years in industries, such as power stations, cement, food, chemical, pharmaceutical, petrochemical plants etc., where the gas velocity is kept adequately high to keep the particles suspended in the conveying gas stream [1]. Modeling of dilute-phase mode of conveying is relatively simpler as the principles of suspension flow mechanics can be applied for this type of flow [2]. The high gas velocity requirement (to ensure suspension of particles) calls for larger size of compressors and higher operating power. The high particle velocities cause increased rate of wear of pipelines and bends. The impact of particles with other particles and the pipeline during high velocity flow results in product attrition in case of fragile products (thus poor quality control and product wastage). Moreover, larger gas flow requires bigger size of filtration equipment (e.g. bag filter); hence additional capital and space requirements. To prevail over these limitations of conventional suspension or dilute-phase flow, low velocity dense-phase pneumatic conveying is acquiring popularity within industries in recent years. In this mode of conveying, due to the lower operating gas and particle velocities, the size of the air mover is considerably reduced (so, lower energy consumption). Besides being economical (reduced gas flows and power consumption), low conveying velocity ensures reduced wear rate of pipes and bends and higher product quality control. Moreover it offers the added advantages of safer workplace and smaller sizes of pipes, fittings, support structures and filtration equipment for the same conveying capacities. There can be different types of dense-phase mode of flow

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depending on the air retention capability and permeability of the bulk solids, such as fluidized dense-phase, slug and plug flow [2, 3]. Typically, the fine powders, such as fly ash, cement, pulverized coal etc. that have good air retention capabilities, are good candidates for fluidized dense-phase mode of transport, where the flow through the pipes is in the form of a moving fluidized bed or non-suspension dunes [1]. This mode of conveying can sustain much higher solids to gas mass flow rate ratio (even in the range of 100), resulting in significant benefits of lower gas flow rates, pipe sizes and product and pipeline damage. However, reliable design of such system is still a challenge, which has hindered its widespread industrial installation. Design requirements for fluidized dense-phase pneumatic conveying of powders consist of accurate prediction of the total pipeline pressure drop and minimum transport boundary, i.e. the minimum conveying velocity requirement to ensure flow blockage does not take place. Inaccurate prediction of pressure drop, such as under-prediction would result in reduced throughput, whereas over-estimation of pressure drop would lead to use oversized air movers resulting in increased initial and operating costs [1]. Total pipeline pressure loss includes pressure drops in horizontal straight sections, verticals, bends and acceleration losses. For pipelines having relatively longer horizontal straight pipe run (e.g. fly ash conveying pipelines in coal fired thermal power plants from intermediate surge hopper to remote silo that may have pipe length up to 1 km), accurate prediction of pressure drop for the horizontal straight pipe run is of paramount importance as the major contribution of the total pressure drop comes from the relatively long length of horizontal section. The pressure loss for solids–gas flow through a straight horizontal section of pipe can be expressed using Eq. (1), as given by Barth [4].

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

This above representation considers the pressure drop due to the gas and solids separately. In this model, while all other parameters can be calculated relatively easily based on well established gas only friction factor formula [5], accurate modeling of solids friction factor is a challenging task due to the limited fundamental understanding of the flow mechanisms of powdered bed. The solids friction factor term is a combined representation of energy loss due to solids to solids, solids to gas and solids to pipe wall interactions [1]. Weber [6] employed this model for coarse particles in dilute-phase type flows. However, various other researchers [7–13] have subsequently employed this expression to predict the pressure loss for the dense-phase pneumatic transport of fine powders, such as fly ash, pulverized coal, ESP dust etc. for horizontal straight pipes. However, due to the highly turbulent and complex nature of the moving fluidized bed of particles under high solids to gas mass ratio (in the form of dunes), it is very difficult to link the particle and bulk properties and the above interactions to the actual operating conditions and modeling the design parameters. Hence, only limited progress has been achieved so far towards fundamentally understanding the flow mechanisms and modeling of solids friction factor. Because of such difficulties, empirical power function type modeling has been popularly employed over the years by several investigators [7–14]. One of the most popular forms of solids friction factor model is provided in Eq. (2):

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

This format has been applied by various researchers [7–10] and can provide good accuracy when applied to researchers' own data. Also, some researchers, such as Pan and Wypych [9] included the particle to air density ratio term in the above format to account for the change in gas density along the pipeline. The model format is given as follows:

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

Previous investigations by the authors [10–13] have shown that the above formats of modeling provide gross inaccuracy under significant scale-up conditions of pipeline length and diameter. Very recently, the authors have provided a new model format by using volumetric loading ratio [12, 13] and dimensionless velocity [13] as the flow defining parameters. Both parameters were considered to be better representatives of flow conditions and the following format was provided to model for solids friction factor.

$$\lambda_s = K(VLR)^a(w_{fo}/V)^b \tag{4}$$

Extensive experimental results were used [12, 13] for the proper scale up validation and there seemed to be a considerable improvement in the overall accuracy of predictions over certain zones of the pneumatic conveying characteristics (medium air flow range). However, the predicted PCC (pneumatic conveying characteristics) did not provide adequate 'U' shaped characteristics, i.e. it could not follow the gradual change in flow mechanism from fluidized dense- to dilute-phase pneumatic conveying (i.e. non suspension to suspension flow mechanism). As a result, the pressure drop prediction lines were not adequately turning upward in very low and high velocity zones, i.e. instead of 'U'-shaped characteristics (with change in slope from low to high air flows), the predicted PCC had more pronounced regions of flat characteristics. Hence, further studies are required to accurately model solids friction factor to address the changes in flow mechanism for the pneumatic conveying of fine powders and to provide pressure drop prediction characteristics that closely follow experimental plots both in values and trends.

2. Experimental data

Various fine powders were conveyed for different solids and air mass flow rates from fluidized dense- to dilute-phase. ESP dust and Australian power station fly ash were conveyed at the Bulk Materials Handling Laboratory of the University of Wollongong, Australia. Different samples of cements and fly ash were conveyed at the pneumatic conveying test facilities of Fujian Longking Co., Ltd., China. The physical properties of the products and pipeline lengths and diameters are provided in Table 1.

Australian power station fly ash and ESP dust were conveyed through three test rigs of the University of Wollongong, Australia (pipelines 1, 2 and 3, refer to Table 1). Typical schematic (for one pipeline) of the test setup is shown in Fig. 1. The 69 mm I.D. × 168 m long pipeline included one 7 m vertical, five 90° bends having 1 m radius of curvature and a 150 mm N.B. tee-bend connecting the end of the pipeline to the receiver bin. For fly ash, static pressure measurement tapping points, P8, P9, P10, P11 and P12, were employed along the length of all the pipes. The P8 tapping location was used to measure total pipeline pressure drop. P11–P12 tapping points were used to obtain differential pressure loss, from where models for solids friction have been generated in this paper. Static pressure measurements points for ESP dust were installed at P8, P9 and P10 locations of the 69 mm I.D. × 554 m pipe from where total pipeline pressure drop and straight pipe pressure data were measured. A 6 m^3 receiving bin with insertable pulse-jet dust filter was provided on top of the blow tank. All other necessary instrumentation for data recording and analysis were provided using a portable PC-compatible data acquisition system. Detailed description of the test set-up and conveying program are provided in [10].

Schematics of the test pipelines used to convey cement and fly ash in Fujian Longking Co., Ltd., China (pipelines 4 and 5) are shown in Figs. 2 and 3. A rotary screw compressor (with air drier and receiver) was used having the maximum delivery pressure of 750 kPa and 660 m^3/h of capacity (Free Air Delivery). A bottom discharge type blow-tank (having 0.75 m^3 empty volume) was used to feed the product into the pipeline. A receiving bin of 2 m^3 capacity was installed on top of the blow tank and fitted with bag filters having a reverse pulse jet type cleaning