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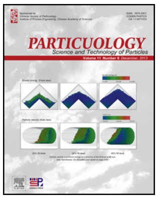

An investigation into flow mode transition and pressure fluctuations for fluidized dense-phase pneumatic conveying of fine powders

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ABSTRACT

This paper presents the results of an ongoing investigation into the fluctuations of pressure signals due to solids–gas flows for dense-phase pneumatic conveying of fine powders. Pressure signals were obtained from pressure transducers installed along different locations of a pipeline for the fluidized dense-phase pneumatic conveying of fly ash (median particle diameter 30 μm; particle density 2300 kg/m^3; loose-poured bulk density 700 kg/m^3) and white powder (median particle diameter 55 μm; particle density 1600 kg/m^3; loose-poured bulk density 620 kg/m^3) from dilute to fluidized dense-phase. Standard deviation and Shannon entropy were employed to investigate the pressure signal fluctuations. It was found that there is an increase in the values of Shannon entropy and standard deviation for both of the products along the flow direction through the straight pipe sections. However, both the Shannon entropy and standard deviation values tend to decrease after the flow through bend(s). This result could be attributed to the deceleration of particles while flowing through the bends, resulting in dampened particle fluctuation and turbulence. Lower values of Shannon entropy in the early parts of the pipeline could be due to the non-suspension nature of flow (dense-phase), i.e., there is a higher probability that the particles are concentrated toward the bottom of pipe, compared with dilute-phase or suspension flow (high velocity), where the particles could be expected to be distributed homogenously throughout the pipe bore (as the flow is in suspension). Changes in straight-pipe pneumatic conveying characteristics along the flow direction also indicate a change in the flow regime along the flow.

1. Introduction

Fluidized dense-phase pneumatic conveying of powders is used in various industries, such as coal-fired thermal power plants, cement production, chemical production, pharmaceutical production, alumina processing, limestone processing and refineries, because of its numerous potential benefits over dilute-phase conveying systems, such as reduced transport gas flow and velocity requirements (from high solids to air mass flow ratio) and thus decreased compressor size and running cost, reduced wear rate of pipes and bends (because of reduced particle velocity), smaller pipeline diameter and support structures, reduced filtration area requirement, and so on. However, despite such merits (and wide range of potential applications), the dense-phase pneumatic conveying of powders has made only limited progress (in terms of installed applications) within industries because the design of such systems is challenging due to the difficulty in determining key design parameters, such as pressure drop in the pipeline and dense- to dilute-phase flow transition criteria. This is because of the highly turbulent and fluidized nature of the two-phase concentrated flow phenomenon inside the pipeline, which involves several mechanisms, such as particle–gas, particle–particle and particle–wall interactions, that are difficult to model based on fundamental principles. Many models have been developed by various previous researchers (Jones & Williams, 2003; Pan, 1992; Pan & Wypych, 1998; Stegmaier, 1978; Williams & Jones, 2006) for calculating the solids friction and the associated pressure drop for horizontal straight pipe sections (which is an important contributor to the total pipeline pressure drop for typical industrial plants), and the majority of the existing models are empirical in nature. These models for the solids friction factor use different dimensionless parameter groupings (generally in the form as given by Eqs. (1) and (2)) and have demonstrated good results when applied to the researchers' own data, but these

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Nomenclature

D	internal diameter of pipe (m)
d_s	particle diameter (m)
V	superficial air/gas velocity (m/s)
d_{50}	median particle diameter (μm)
ρ	density (kg/m^3)
ρ_s	particle density (kg/m^3)
L	pipeline length (m)
S	Shannon entropy
m_f	mass flow rate of air (kg/s)
m^*	solids loading ratio (m_s/m_f)
m_s	mass flow rate of solids (kg/s)
$\tilde{n}_{bl}$	loose-poured bulk density (kg/m^3)
$\tilde{n}_s$	particle density (kg/m^3)
λ_s	solids friction factor through straight pipe
$Fr_m = V_m/(gD)^{0.5}$	Froude number of flow

Superscripts

a, b, c, d	Exponents of the power function (Eqs. (1) and (2))

Subscripts

f	fluid (air)
s	solids

models have not been tested under significant scale-up conditions.

$$\lambda_s = K(m^*)^a(Fr)^b \tag{1}$$

$$\lambda_s = K(m^*)^a(Fr)^b\left(\frac{\rho}{\rho_s}\right)^c\left(\frac{d_s}{D}\right)^d \tag{2}$$

Recent investigations into the scale-up accuracy of these existing and/or popular models on the basis of comparison between predicted pneumatic conveying characteristics and experimental data have proven that the models are generally inaccurate under scale-up conditions of pipeline length and/or diameter. These models are based on the assumption that particle velocity is equal to gas velocity. Because of the large difference between particle and gas velocity, especially for dense-phase pneumatic conveying, it would be inappropriate to use these models for pressure drop calculations. Moreover, the pressure drop in the pipeline fluctuates continuously as depicted by the pressure versus time trace obtained by Dhodapakar and Klinzing (1993), whereas current modeling techniques for the prediction of pressure loss inside pipes are based on steady-state analysis, which could be another major reason for the inaccuracy of existing models. These inaccuracies indicate that the existing non-dimensional approach of empirical modeling is based on parameter groupings that are too simple to accurately describe the dense-to-dilute flow mechanism. The highly turbulent solids–gas flow phenomenon requires a more fundamental approach, e.g., a study on the transient nature of the fluctuating solids–gas flows, especially in fluidized dense-phase. Compared with the numerous works that have been carried out, which are based on empirical modeling approach using steady-state pressure data, only a few efforts have been made toward the characterization of the transient pressure signals (because of the complex turbulent solids–gas flows) and extracting valuable information from such signals for modeling solids friction and pressure drop. This could include characterization of pressure transients for suspension (dilute) and non-suspension flows (dense-phase), i.e., an attempt to identify certain signal attributes related to a specific mode of conveyance, such that the flow mechanisms could be predicted from the knowledge of pressure signal fluctuation. To obtain more detailed information from the pressure peaks, such as frequency, amplitude and variation of these signals along the pipeline length and varying operating conditions (e.g., mass flow rates, gas velocity, solid loading ratio, etc.), an appropriate method to analyze pressure signals needs to be chosen. In the past, researchers have proposed certain methods to analyze pressure signals for pneumatic conveying of bulk solids, including statistical analysis based on square designs, cross-correlation coefficient and skewness factor (Jama, Klinzing, & Rizk, 2000), wavelet transforms (Li, 2002), electrical capacitance tomography (Fuchs, Zangl, & Wypych, 2007; Jaworski & Dyakowski, 2002), Hurst's rescaled range analysis to identify flow mode (Jama, Klinzing, & Rizk, 1999), phase-space diagram (Pahk & Klinzing, 2008a) and power spectral density (Pahk & Klinzing, 2008b). These existing works on the analysis of pressure fluctuation for pneumatic conveying were either based on coarse particle dilute-phase transport or for slug-type dense-phase flows (for particles such as the plastic pellets) for which the flow mechanisms are relatively well defined. Very limited work has been reported in the area of analysis of pressure signal fluctuation for fine powders (capable of being aerated fluidized dense-phase flow, typically Geldart group A powders). Recently, Williams, Jones, and Yadav (2007) conducted pneumatic conveying trials with cement meal ($d_{50} = 11$ μm, solid loading ratio in the range of 51–70) and determined the average velocity from the time delay difference of peaks and troughs of the pressure signals corresponding to two different locations of pipeline. The average velocity from pressure fluctuations was found to pertain to the specified testing conditions and the product conveyed. Williams et al. (2007) concluded that additional analysis was required for varied conveying conditions, thus signifying the need for further research.

Shannon entropy is a relatively new tool in the field of pressure signal analysis for solid–gas flow to measure the information content and complexity of time series. It is utilized to predict the degree of uncertainty involved in predicting the output of a probabilistic event. From a given signal or time series, a discrete data set of $X(t)$ is obtained as $X(t) = \{X(t_1), X(t_2), \ldots, X(t_n)\}$, and corresponding values of X are denoted as $X_1, X_2, \ldots, X_n$. The probability P of any value of X is $P(X_i) = X_i/n$. Thus, the entire set of probabilities, i.e., $P(X_1), P(X_2), \ldots, P(X_n)$, are calculated from the original data set. The Shannon entropy of a pressure signal in the pneumatic conveying is calculated as follows:

$$S(X) = \sum_{i=1}^{n} P(X_i)\log_b P(X_i), \tag{3}$$

where n is the length of the signal. When $b = 2$, e, and 10, the unit of S is a bit, nat, and hart, respectively. $P(X_i)$ is the probability of every component in the signal.

$$\sum_{i=1}^{n} P(X_i) = 1.$$

In the past, Shannon entropy has been used to correlate the flow characteristics of two-phase systems, such as when Zhong et al. (2009) performed Shannon entropy increment analysis on pressure fluctuations of a biomass fluidized bed to study flow patterns and dynamic behavior. The Shannon entropy values of steady fluidization were found to be small, whereas it was found to have increased for turbulent fluidization. Recently, Duan and Cong (2013) used the Shannon entropy method to compare the dynamic behavior of certain Geldart group B and Group D particles in fluidized condition. The Shannon entropy of Geldart group B particles increased steadily to a maximum (during uniform expansion of bed), whereas the Shannon entropy of the Geldart group D particles exhibited only a slight increase during initial fluidization (due to limited expansion

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