Particuology xxx (2017) xxx–xxx

Contents lists available at ScienceDirect

Particuology

journal homepage: www.elsevier.com/locate/partic

Numerical simulation of fluidized dense-phase pneumatic conveying of powders to develop improved model for solids friction factor

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ARTICLE INFO

Article history:
Received 16 June 2016
Received in revised form
14 November 2016
Accepted 22 November 2016
Available online xxx

Keywords:
Pneumatic conveying
Dense-phase
Numerical analysis
Pressure drop
Scale-up validation

ABSTRACT

Accurate prediction of the solids friction factor through horizontal straight pipes is important for the reliable design of a pneumatic conveying system, but it is a challenging assignment to date because of the highly concentrated, turbulent, and complex nature of the gas–solids mixture. Power-station fly ash was transported through different pipeline configurations. Numerical simulation of the dense-phase pneumatic conveying systems for three different solids and two different air flow rates have shown that particle and actual gas velocities and the ratio of the two velocities increases in the flow direction, whereas the reverse trend was found to occur for the solids volumetric concentration. To develop a solids friction-factor model suitable for dense-phase flow, we modified an existing pure dilute-phase model by incorporating sub-models for particle and actual gas velocities and impact and solids friction factor. The solids friction-factor model was validated by using it for scale-up predictions for total pipeline pressure drops in longer and larger pipes and by comparing experimental and predicted pneumatic conveying characteristics for different solids flow rates. The accuracy of the prediction was compared with a recently developed two-layer-based model. We discussed the effect of incorporating the particle and actual gas velocity terms in the solids friction-factor model instead of superficial air velocity.

Introduction

Fluidized dense-phase pneumatic conveying of fine powders offers several advantages over a conventional dilute phase (or suspension flow), such as a high solids-to-gas mass ratio, a low gas-flow requirement (i.e., a smaller compressor and savings in operating power), smaller pipes and fittings, a reduced conveying velocity that results in a lowering of wear rate of pipelines and bends, a decreased rate of product attrition, and smaller solids–gas separation equipment (Mallick, 2009). As a result, this mode of conveying is being increasingly preferred in several industries, such as coal-fired thermal power plants and cement, food, chemical, pharmaceutical, and petrochemical plants. In fluidized dense-phase transport, gas velocities are kept sufficiently low (below the saltation velocities). As a result, conveying occurs in a non-suspension mode, in the form of a moving fluidized bed or non-suspension dunes (Behera, Agarwal, Jones, & Williams, 2013; Marcus, Leung, Klinzing, & Rizk, 1990). Typically, fine powders with good air-retention properties, such as fly ash, cement, and pulverized coal are good candidates for the fluidized dense-phase mode of conveying (Pan, 1999; Ratnayake, Datta, & Melaaen, 2007). An accurate prediction of the total pipeline pressure drop is an important parameter that requires a reliable estimation at the design stage. The total pipeline pressure drop includes pressure losses in horizontal straight pipes, verticals, bends, and the loss because of initial acceleration. For a typical long-distance pipeline (e.g., the pressure conveying line that runs from the buffer hopper to the remote silo in a coal-fired thermal power plant—typically 500–1000 m long), the major share of total pipeline pressure drop originates from losses in the horizontal straight pipes. An overprediction of pressure drop would increase the initial and operating costs (because of an unnecessary higher conveying velocity and power consumption), whereas, an underprediction of pressure drop would result in a reduced material transport rate. The pressure drop for solids–gas flow through a straight horizontal pipe can be represented by Eq. (1), as given by Barth (1958). This representation considers the pressure drops because of the gas and solids phases separately.

$$\Delta P = \left(\left(\lambda_f + m^* \lambda_s \right) \rho_a L V^2 \right) / 2D. \quad (1)$$

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http://dx.doi.org/10.1016/j.partic.2016.11.006
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Nomenclature

Symbols and abbreviations

A	Cross-sectional area of control volume (m^2)
A_1, A_2	Area of cross-section occupied by dense and dilute phase layers, respectively (m^2)
a, b, c	Exponents of power function
B	Bend loss factor
C	Particle velocity (m/s)
C_{dk}	Drag coefficient for particle in layer k
D	Internal pipe diameter (m)
D_P	Mean particle diameter (μm)
d_p	Particle diameter (μm)
d_{50}	Median particle diameter (μm)
F_{dk}	Body force between particles in layer k(N)
$Fr = V/(gD)^{0.5}$	Froude number of flow
$Fr_i = V_i/(gD)^{0.5}$	Froude number of flow at the pipe inlet
$Fr_s = w_{fo}/\sqrt{gd_{50}}$	Particle Froude number
F_a	Frictional force due to air phase on pipe wall (N)
F_s	Frictional force due to solids phase on pipe wall (N)
f_a	Frictional force per unit volume due to air phase (N/m^3)
f_s	Frictional force per unit volume due to solids phase (N/m^3)
g	Gravitational acceleration (m/s^2)
K	constant of power function
L	Length of horizontal pipe or test section (m)
L_v	Length of vertical pipe or test section (m)
m_f, m_a	Mass flow rate of air (kg/s)
m_s	Mass flow rate of solids (kg/s)
$m^* = m_s/m_a$	Solids loading ratio
N	Number of bends
P	Pressure (Pa)
ΔP	Pressure drop through a straight horizontal pipe or pipe section (Pa)
ΔP_{accel}	Pressure drop due to initial acceleration (Pa)
ΔP_b	Pressure drop due to the bends (Pa)
ΔP_V	Pressure drop due to the verticals (Pa)
r_{sk}	Volume fraction of solids phase in layer k
u_a	Actual gas velocity (m/s)
u_s	Particle or dune velocity (m/s)
V, U_g	Superficial air or gas velocity (m/s)
U_{gk}	Velocity of gas phase in layer k(m/s)
U_{sk}	Velocity of solids phase in layer k(m/s)
$V_a = m_a/\rho_a$	Volumetric flow rate of air (m^3/s)
$V_s = m_s/\rho_{fl}$	Volumetric flow rate of solids (m^3/s)
$VLR = \left\{\left(m_s/\rho_s\right)/\left(m_f/\rho_a\right)\right\}$	Volumetric loading ratio
w_{fo}	Free settling velocity of an isolated particle (m/s)
w_{fo}/V	Dimensionless velocity
x	Horizontal distance in the direction of flow starting at the blow tank (m)
ρ, ρ_a	Density of air (kg/m^3)
ρ_{gk}	Density of gas phase in layer k(kg/m^3)
ρ_s	Particle density (kg/m^3)
ρ_b	Loose-poured bulk density (kg/m^3)
ρ_{fl}	Fluidized bulk density (kg/m^3)
λ_f	Air/gas only friction factor
λ_s	Solids friction factor through straight pipe
$\lambda s*$	Impact and friction factor for solids
$\varepsilon_a = V_a/(V_a + V_s)$	Volume fraction of air
$\varepsilon_s = V_s/(V_a + V_s)$	Volume fraction of solids

Weber (1981) used Eq. (1) for coarse particles in dilute-phase flows. Subsequently, various researchers (Jones & Williams, 2003; Pan & Wypych, 1998; Setia & Mallick, 2015) have used the same equation to estimate pressure drop for the fluidized dense-phase pneumatic transport of fine powders. The main challenge in Eq. (1) is to model and/or predict the solids friction factor accurately, which is a combined representation of energy loss because of solids-to-gas, solids-to-solids, and solids-to-pipe wall interactions (Mallick, 2009). This occurs because of the highly turbulent and complex nature of the moving fluidized bed of fine powders (in the form of dunes) at high concentration, which makes it very difficult to link the particle and bulk properties during actual flow conditions. Empirical power-function-based models have been used over the years by several investigators, such as Pan and Wypych (1998) (using dimensionless numbers, as given by Eq. (2)), because of the limited progress towards understanding fundamentally the transport mechanisms under fluidized dense-phase flow conditions and the modeling of a solids friction factor. Some researchers, such as Jones and Williams (2003), have used a condensed version of Eq. (2) considering the value of exponent "c" to be zero. This considers inherently that the typical particle density is much larger than the gas density, and hence the effect of changing gas density on flow mechanism was not taken into account.

$$\lambda_s = K(m^*)^a (Fr)^b \left(\rho_a/\rho_s\right)^c. \tag{2}$$

A previous evaluation (Mallick, 2009) has shown that the above modeling formats of solids friction factor can provide grossly inaccurate predictions under significant scale-up conditions of pipeline length and diameter. Recently, some authors have provided a new two-layer-based model format for solids friction factor (Setia, Mallick, Pan, & Wypych, 2016), as represented by Eq. (3). This model separates the solids friction factor into losses contributed by impact and friction between particles to air/particles/pipe wall and that from keeping particles in suspension.

$$\lambda_s = \tau_1 \left(K(VLR)^a \left(w_{fo}/V\right)^b\right) + \tau_2 \left(\lambda s * C/V + 2\beta_0/\left[\left(C/V\right) Fr^2\right]\right). \tag{3}$$

In the above model, the first term, $K(VLR)^a\left(w_{fo}/V\right)^b$, represents the dense-phase contribution (Setia et al., 2016). The second term, $\lambda_s^* C/V + 2\beta_0/\left[\left(C/V\right) Fr^2\right]$, represents the dilute-phase contribution (Setia et al., 2016; Wypych, Kennedy, & Arnold, 1990). β_0 is given as w_{fo}/V. The dilute-phase portion has been taken from a "pure" dilute-phase model, known as the "Weber A4" model (Wypych et al., 1990). This model (Weber A4) has been reported to provide good predictions for dilute-phase flow for different pipeline configurations (Wypych et al., 1990). The two-layer model format was validated by Setia et al. (2016) for scale-up accuracy for two different fly-ash samples, electrostatic precipitator dust and cement under different pipeline conditions (for pipe internal diameters of 69, 80, and 105 mm and lengths of 168, 254, 407, and 554 m). The two-layer model provides better reliable scale-up predictions compared with other previously known models for solids friction factor. However, the authors believe that little progress has been achieved to date towards understanding the fundamental flow mechanism of fluidized dense-phase solids–gas transport of fine powders. The existing empirical models (Jones & Williams, 2003; Pan & Wypych, 1998; Setia & Mallick, 2015; Setia et al., 2016) for solids friction factor often use the term "superficial gas velocity", instead of the actual gas velocity. A superficial gas velocity can be considered to be an accurate gas velocity representation for dilute-phase flow, where the pipe volume occupied by the particles is minor compared with the rest of the pipe volume available for gas flow. However, for fluidized dense-phase flows under a high solids loading ratio, a considerable portion of the pipe cross-section is occupied by the powders. Therefore, the effect of reduced cross section that is available for the gas-phase flow should not be ignored.

Please cite this article in press as: Kaur, B., et al. Numerical simulation of fluidized dense-phase pneumatic conveying of powders to develop improved model for solids friction factor. *Particuology* (2017), http://dx.doi.org/10.1016/j.partic.2016.11.006