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Original paper

Computational model for prediction of particle degradation during dilute-phase pneumatic conveying: modeling of dilute-phase pneumatic conveying

PIERRE CHAPELLE [1,*], NICHOLAS CHRISTAKIS [1], HADI ABOU-CHAKRA [2], IAN BRIDLE [3], M. S. A. BRADLEY [3], MAYUR PATEL [1] and MARK CROSS [1]

[1] *Centre for Numerical Modelling and Process Analysis, University of Greenwich, Old Royal Naval College, Park Row, Greenwich, London SE10 9LS, UK*

[2] *Chemical and Process Engineering, School of Engineering, University of Surrey, Guildford, Surrey GU2 7XH, UK*

[3] *The Wolfson Centre for Bulk Solids Handling Technology, University of Greenwich, Wellington Street, Woolwich, London SE18 6PF, UK*

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Abstract—A complete model of particle impact degradation during dilute-phase pneumatic conveying is developed, which combines a degradation model, based on the experimental determination of breakage matrices, and a physical model of solids and gas flow in the pipeline. The solids flow in a straight pipe element is represented by a model consisting of two zones: a strand-type flow zone immediately downstream of a bend, followed by a fully suspended flow region after dispersion of the strand. The breakage matrices constructed from data on 90° angle single-impact tests are shown to give a good representation of the degradation occurring in a pipe bend of 90° angle. Numerical results are presented for degradation of granulated sugar in a large scale pneumatic conveyor.

Keywords: Dilute-phase pneumatic conveying; numerical simulations; impact degradation; breakage matrices.

1. INTRODUCTION

Particle degradation during dilute-phase pneumatic conveying is often a major concern in industrial practice [1]. Current computational approaches for simulating pneumatic conveying systems concentrate only on the detailed description of the flow of the solids and gas phases, and disregard the damage which occurs to the particle on impact with pipe walls (e.g. [2, 3]). Comparatively, erosive wear of

*To whom correspondence should be addressed. E-mail: p.chapelle@gre.ac.uk

bends by particle impact has been widely studied in the literature and models have been presented for predicting the life time of individual bends in conveying systems (e.g. [4, 5]).

The aim of this paper is the numerical prediction of particle degradation during dilute-phase pneumatic conveying. In a previous work [6], the concept of a breakage matrix has been used to develop a procedure for determining degradation propensity under a range of particle velocities and particle sizes from single-impact tests carried out in a laboratory-scale degradation tester. In the present work, a complete model of dilute-phase pneumatic conveying is developed, which incorporates within a physical flow model of the solids and gas phases through the pipeline, a model for particle degradation based on the analysis presented in [6]. Numerical results are presented for the degradation of granulated sugar in an industrial scale pneumatic conveyor.

2. NUMERICAL MODELS

Since the most extensive damage during dilute-phase pneumatic conveying is caused by collisions of the particles with bend walls [7], only degradation by impact at the bends is considered in the present work.

A number of simplifying assumptions have been made in order to arrive at a working model.

- All particles travel at the same velocity whatever their size and density [8].
- Interparticle collisions are neglected, since only dilute phase conveying systems are considered.
- Degradation of a particle in a bend is represented by a single impact. Indeed, the first impact gives rise to the most severe damage and additional impacts in the same bend after the particle rebounds off the wall are not considered as significant as a first approximation. As will be shown later, the collective effect of a 90° angle bend on a particle size distribution can be represented quite accurately through a 90° angle single-impact event.
- Fatigue phenomena, by which a particle breaks when collisions occur a number of times, are ignored as a first approximation. Hence, each impact of a particle is considered independent of the next.

Two important parameters affecting degradation by impact are the particle conveying velocity and the collision angle (the angle between the particle path and the tangent to the wall at the collision point), which most likely vary along the length of the pipeline [7]. Therefore, it is essential for the evaluation of degradation to first calculate the particle velocity at any point in the system, and hence the velocity of any impact between a particle and the bend wall. A model for particle degradation should additionally consider the effect of the bend structure (type of bend, radius of curvature) upon which the collision angle is critically dependent.

As a consequence, a prerequisite to the modeling of degradation during dilute-phase conveying is the development of a model for particle acceleration within the pipeline, and the capability to correlate the impact angle and the bend structure.

2.1. Flow model

The pipeline is divided up into straight sections and bends.

2.1.1. Straight horizontal pipeline. Observations by tomography of the flow pattern during dilute-phase conveying [9] reveal two distinct modes of flow in a straight pipe element, as represented on Fig. 1. Note that the developed flow modes in the case of a horizontal bend to a horizontal straight pipe are identical to those shown in Fig. 1.

Downstream of a pipe bend, as a result of being slowed down inside the bend, particles are mainly transported in a form of a strand, in which a high concentration layer of particles occupies the lower portion of the pipe. In the upper portion of the pipe, particles are suspended in the transport gas. As particles on the top of the layer are picked up from the surface by the gas stream flowing more quickly, the layer is gradually eroded and decreases in depth along the pipeline. A point is reached where the layer of solids vanishes and all the particles are transported as a homogeneous gas–solid suspension flow (characterized by a uniform dispersion of the particles across the pipe). This mode of flow is called fully suspended flow. It should be mentioned that similar phenomena occur to suspend the particles initially at rest after their injection at the inlet of the conveying pipe [10].

The approach used in the present study for modeling the strand type flow is based on the model of Bradley *et al.* [11]. This model is a one-dimensional model, which describes the flow of two layers (i.e. the dense strand and the suspended flow above the strand) with separate velocity and exchanging momentum between them due to shear forces at their interface. It is based on a balance of forces on elements of the region above the strand and of the strand itself.

Assuming the gravitational effects and the friction between the strand and the bottom end of the pipe's inner surface being insignificant in the balance of forces acting on a strand element, Newton's second law of motion applied to a strand

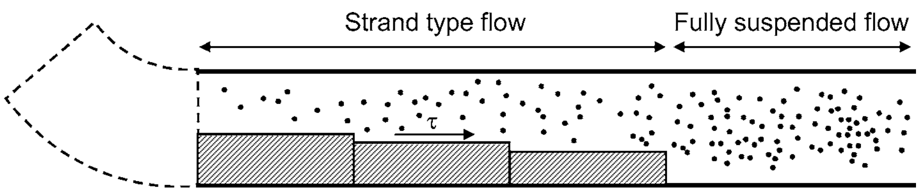


Figure 1. Schematic of the particulate flow downstream of a pipe bend during dilute-phase pneumatic conveying.

element of length δl leads to the following expression:

$$\tau_{\mathrm{str}} S_{\mathrm{str}} = \rho_b \delta l \frac{\pi D^2}{4}(1-\phi) a_{\mathrm{str}}. \tag{1}$$

In this equation, τ_{str} is the shear stress between the strand and the suspended flow above, S_{str} is the strand surface area in contact with the suspended flow, ρ_b is the bulk density of the material in the strand which is assumed to remain constant throughout the strand dispersion process, D is the pipe diameter, $(1-\phi)$ represents the fraction of the pipe cross-sectional area which is covered by the strand and a_{str} is the acceleration of the strand element.

The friction force between the strand and the suspended flow is modeled in analogy to single-phase flow using an equivalent Moody friction factor, as suggested in Bradley *et al.* [11]:

$$\tau_{\mathrm{str}} = \frac{\rho_{\mathrm{air}} f_{\mathrm{str}}}{2}(v_{\mathrm{air}} - v_{\mathrm{str}})^2, \tag{2}$$

where ρ_{air} is the air density, and v_{air} and v_{str} are the air and strand velocities. The value of the equivalent Moody friction factor f_{str} is increased above the one obtained from the Moody diagram [12] in order to account for the additional momentum transfer resulting from the interchange of particles between the surface of the strand and the suspended flow.

The motion of the suspended flow is controlled by a balance between a driving force $\Delta P_{\mathrm{susp}}\phi(\pi D^2)/4$ and the frictional resistances against the pipe wall τ_{susp} and the strand τ_{str}:

$$\tau_{\mathrm{str}} S_{\mathrm{str}} + \tau_{\mathrm{susp}} S_{\mathrm{susp}} = \Delta P_{\mathrm{susp}} \frac{\pi D^2}{4}\phi, \tag{3}$$

where S_{susp} is the contact area between the suspended flow and the pipe wall.

The friction force between the conveying gas and the pipe wall is modeled as for single-phase flow:

$$\tau_{\mathrm{susp}} = \frac{\rho_{\mathrm{air}} f_{\mathrm{susp}}}{2} v_{\mathrm{air}}^2, \tag{4}$$

where the value of the friction factor f_{susp} is assumed, as before, to be higher than the one determined for air alone from the Moody diagram [12], to allow for the contribution to the pressure gradient of the particles present in the suspended flow.

In the fully suspended flow region, the slip velocity between the air and the particles is assumed to be negligible. The usual approach of calculating the total pressure drop along an elemental length consists of splitting the pressure gradient into two parts [13]:

$$\Delta P = \Delta P_{\mathrm{air}} + \Delta P_{\mathrm{solids}}, \tag{5}$$

where ΔP_{air} is the pressure drop that would be expected with air only in the pipe and $\Delta P_{\mathrm{solids}}$ represents the additional pressure drop caused by the particles present in the gas.

ΔP_{air} is calculated as for single-phase flow conditions from Darcy's equation:

$$\Delta P_{\text{air}} = \frac{2f\delta l}{D}\rho_{\text{air}}v_{\text{air}}^2, \tag{6}$$

with the pipe flow friction factor f.

The pressure drop due to the particles can be represented by [14]:

$$\Delta P_{\text{solids}} = K\delta l\rho_{\text{susp}}v_{\text{air}}^2, \tag{7}$$

where ρ_{susp} is the suspension density (defined as the mass flow rate of solids divided by the volume flow rate of air) and K is an empirical coefficient.

The increase in air velocity caused by the expansion of the air is given by:

$$v_{\text{air}} = \frac{4\dot{V}_{\text{air}}}{\pi D^2}, \tag{8}$$

where $\dot{V}_{\text{air}}$ is the volume flow rate of air, which is determined from the ideal gas law to take gas compressibility effect into account:

$$P\dot{V}_{\text{air}} = \dot{m}_{\text{air}}RT_{\text{air}}, \tag{9}$$

with $\dot{m}_{\text{air}}$ the mass flow rate of air, R the gas constant and T_{air} the air temperature.

2.1.2. Straight vertical pipeline. Only vertically-up sections are considered in the present study, since long vertical down sections are not common in industrial practice. For the strand type of flow, the gravitational force must be included in the force balance on a strand element in (1). The approach of Mills [15], who has found that vertically-up sections show pressure gradient approximately twice that of horizontal sections, has been applied to predict the pressure drop under fully suspended flow conditions. Therefore, the pressure drop value obtained from (5) needs to be doubled.

2.1.3. Bend pipeline. It has been observed that the pressure drop caused by a bend does not occur in the bend itself, but in the straight pipe immediately downstream of the bend [16, 17]. As a consequence, there is no pressure drop in the bend. The deceleration of the particles on their way through a bend occurs mainly as a result of bouncing and sliding contact with the pipe wall under the effect of the centrifugal force. The particle velocity change inside the bend can be described by a simple physical model based on Newton's second law, in which the particle–wall interactions are characterized by a coefficient of friction μ and an overall coefficient of restitution e. This leads to:

$$v = ev_0\exp(-\mu\theta), \tag{10}$$

where v_0 is the particle velocity as it enters into the bend and θ is the angular position of the particle inside the bend. The overall coefficient of restitution e is defined in terms of the initial and rebound velocities v_i and v_r as:

$$e = v_r / v_i. \tag{11}$$

Assuming a constant mass flow rate of solids within the pipeline, it is possible to work out from the particle velocity value the fraction of the pipe cross-sectional area occupied by the strand $(1 - \phi)$ at the outlet section of the bend. The associated increase in the air velocity due to the reduction of the cross-sectional area available for the air can then be obtained from mass conservation.

2.1.4. Parameters of the model. The friction factor between the air and the pipe wall f can be read directly on the Moody diagram for the actual air Reynolds number and surface roughness conditions [12].

The value of the coefficient K characterizing the solids contribution to the pressure gradient in the fully suspended flow can be obtained from the empirical diagram of Hyder, who plotted the values of K versus the particle size and the particle density [14].

Finally, the last two parameters of the model, i.e. the two equivalent Moody friction factors f_{str} and f_{susp}, must be estimated from experimental data, e.g. by fitting the calculated pressure profile downstream of a bend to the measured one. Our determination of the friction factor between the suspended flow and the pipe wall f_{susp} is based on Bradley's analysis, who postulated a linear relationship of f_{susp} against the suspension density [11].

2.1.5. Numerical solution procedure. The solution procedure in a straight section is based on the subdivision of the pipe into elemental lengths. Calculation proceeds forwards along the pipeline, taking into account each element in turn. The boundary condition on pressure is set either at the end of the conveying line (in the case of a positive pressure system) or at the beginning of the pipe (in the case of a vacuum system). Therefore, in the case of a positive pressure system, the calculation procedure works with a guessed value of the pressure at the inlet of the pipe and is iterated to obtain the known value of the pressure at the end of the pipe.

The solution of each element is achieved through an iteration procedure by using a central difference scheme. The procedure for an element in the strand flow region is illustrated in Fig. 2.

The procedure begins with known values of air pressure, air and strand velocities, and strand cross-sectional area at the starting end of the element, and guessed values of these quantities at the opposite end of the element. First, the strand surface in contact with the air S_{str} and the contact area of the air with the inside surface of the pipe S_{susp} are worked out (by geometry considerations) from the cross-sectional area covered by the strand. The friction forces between strand and air τ_{str} and air and pipe τ_{susp} are then determined from (2) and (4), respectively. These values are used to determine the acceleration of the strand element from (1) and the new strand velocity at the opposite end of the element, thus finding the decrease in the

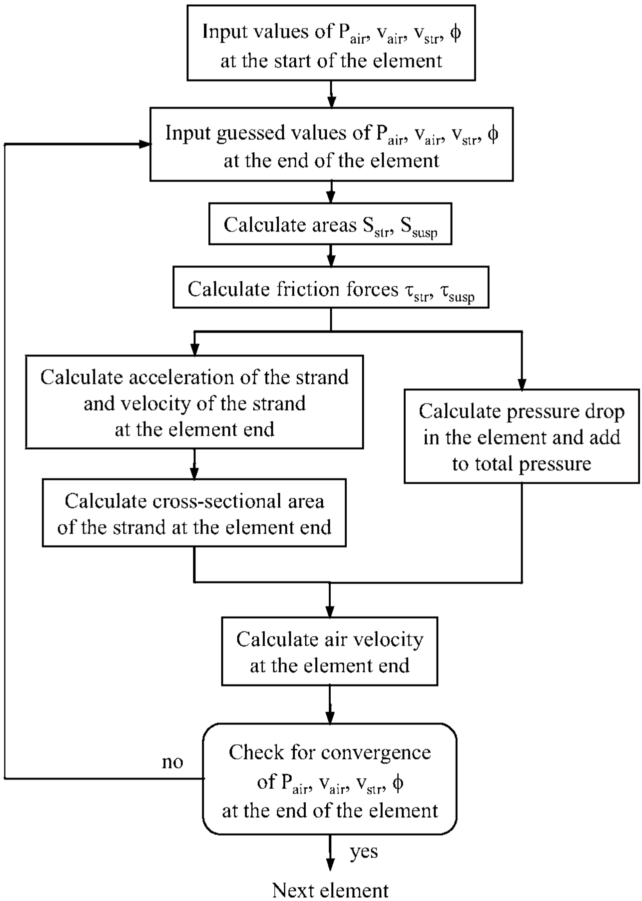


Figure 2. Flow chart for the solution procedure of the governing equations of the strand-type flow.

cross-sectional area of the strand from the conservation of the solids mass flow rate. The pressure of air at the outlet of the corresponding suspended flow element is obtained from the pressure drop in the air element calculated from (3). The value of the air velocity at the outlet of the air element resulting from the expansion of the air and the increase in the cross-sectional area available for the air (due to the reduction of the depth of the strand as its velocity increases) is determined using the conservation of the air mass flow rate in conjunction with the perfect gas law. The overall procedure is repeated until convergence is reached on the values of the air pressure, the air and strand velocities, and the strand cross-sectional area at the opposite end of the element.

The model of acceleration of the strand type flow has been previously validated through comparisons against pressure measurements obtained in an industrial-scale pneumatic conveying system [11].

2.2. Degradation model

As discussed earlier, particle degradation is considered to occur only by impact at the bends. By analogy to the traditional analysis of grinding processes by population balance models (e.g. [18]), the variation of all particles sizes in an impact event can be represented in a concise way by means of a breakage matrix based on the following matrix equation [6]:

$$\mathbf{B} \cdot \mathbf{i} = \mathbf{o}, \tag{12}$$

where **i** and **o** are column vectors, whose entries are the particle mass fraction in each size class, respectively before and after impact, and **B** is the breakage matrix, whose elements b_{ij} define the mass fraction of particles of size class j which after impact ends up in size class i. Chapelle *et al.* [6] developed an interpolation method for calculating the output particle size distribution over a range of impact velocities and particle sizes, based on the measurement of the breakage matrix for a limited number of velocities in a laboratory scale degradation tester.

In this paper, the overall change in the particle size distribution between the inlet and the outlet of a bend is described by an equation equivalent to (12). Assuming that degradation in a bend can be represented by a single impact, the variables **i** and **o** in (12) are now defined as the particle size distribution at the inlet and outlet of the bend, respectively. Numerical evaluation of the outlet particle size distribution **o** is performed by means of the method presented in [6], for the impact velocity involved in the bend, which is considered to be the particle velocity at the inlet section of the bend calculated from the flow model. Experimental breakage matrices must be constructed from results of impact experiments carried out at an impact angle representative of the angle of collision in the bend, which is primarily dependent on the angle of the bend involved. However, it is not possible to evaluate directly the angle at which a particle impacts on the wall, without performing a Lagrangian description of the particulate flow within the bend. Therefore, a simplified model is proposed, in which a given value of the bend angle is assumed to correspond to a given value of the particle impact angle. As it will be demonstrated later, this approach enables a satisfactory characterisation of degradation processes in 90° angle bends.

3. VALIDATION OF THE DEGRADATION MODEL AGAINST PNEUMATIC CONVEYING TESTS IN THE LABORATORY

3.1. Test facility

A pilot-sized pneumatic conveyor test facility was developed to predict degradation in pneumatic conveying systems (see Fig. 3). This test facility allows material to be conveyed around one bend, with certain critical variables regulated. The entire sample of the conveyed material is then captured and analyzed to assess the amount of degradation caused under the designated test conditions. Care has been taken

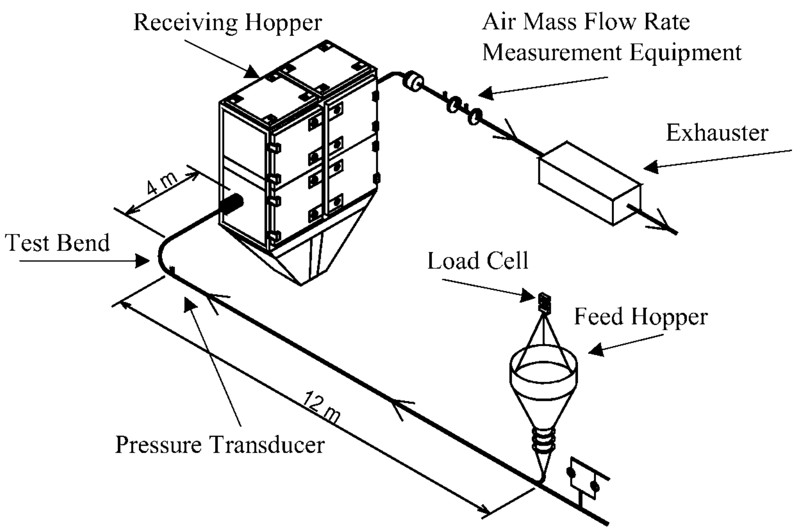


Figure 3. Schematic of the pilot-sized pneumatic conveyor test rig.

with the design of this conveying section to ensure that the acceleration length provided is adequate for the particles to reach a steady-state velocity [8]. The pneumatic conveyor test rig is described in detail in [19].

3.2. Conveying tests

The aim of this test rig is to assess as accurately as possible the degradation in a single bend. The rig works on a negative pressure principle that minimizes degradation due to feeders. Particles were accelerated along a 38-mm internal diameter pipeline over a length of 12 m. Material was conveyed around one bend into a receiving hopper. Once the entire sample of conveyed material had been collected from the hopper, a particle size analysis was applied to assess the amount of degradation caused under the designated test conditions. An advantage of using this test facility is that it is possible to collect all the conveyed material, as the filter and the internal walls of the receiving hopper are easily accessible and cleaned.

A series of tests using granulated sugar (of solids density of 1600 kg/m^3) at varying velocities (9, 16 and 22 m/s) for a short radius ($R/D = 3.6$) 90° angle bend geometry was undertaken. A constant suspension density equal to 15 kg/m^3 was used in all experiments. The retrieved sample was reduced to a smaller representative quantity using a dividing riffle box. For the means of measuring degradation, the particle size analysis was performed according to British Standard [20] procedure for conducting a sieve test. The experimental uncertainty has been calculated to be 5%.

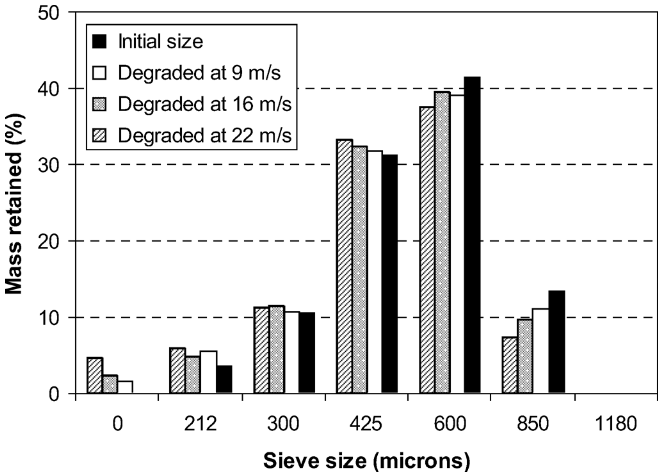


Figure 4. Particle size distribution of sugar samples at different particle velocities using a pilot-sized pneumatic conveyor test facility.

3.3. Test results

Figure 4 shows the degradation behaviour of the sugar for a range of particle velocities. As expected, the amount of degradation increases as the velocity increases. However, it seems likely that below 22 m/s the amount of degradation is relatively small, which might be due to the fact that the strength of the particle is high enough to resist the shock of subjected impacts; only peripheral fractures occur. This is typified by the corners of the sugar having a tendency to break from the particle.

3.4. Comparison between the model and the test results

For the above-mentioned inlet particle size distribution, the outlet particle size distribution has been calculated at each investigated velocity by means of the presented degradation model using the breakage matrices determined in a previous work [6]. These results are compared against the measurements from the pilot-sized pneumatic conveyor rig (see Fig. 5).

Good agreement is found between the particle size distribution measured and that calculated for each impact velocity. These results demonstrate that the degradation model based on 90° angle single impact data does give a suitable representation of the degradation that takes place in a 90° angle bend.

The calculated air pressure at the bend inlet (0.997×10^5 Pa) is in reasonable agreement with the measured value (0.918×10^5 Pa), with a relative difference of about 8%.

Experimental tests performed in the pilot-sized pneumatic conveyor rig for 90° single bends need not be repeated with bends of different curvatures, since it has been experimentally observed [19] that the bend curvature does not have a

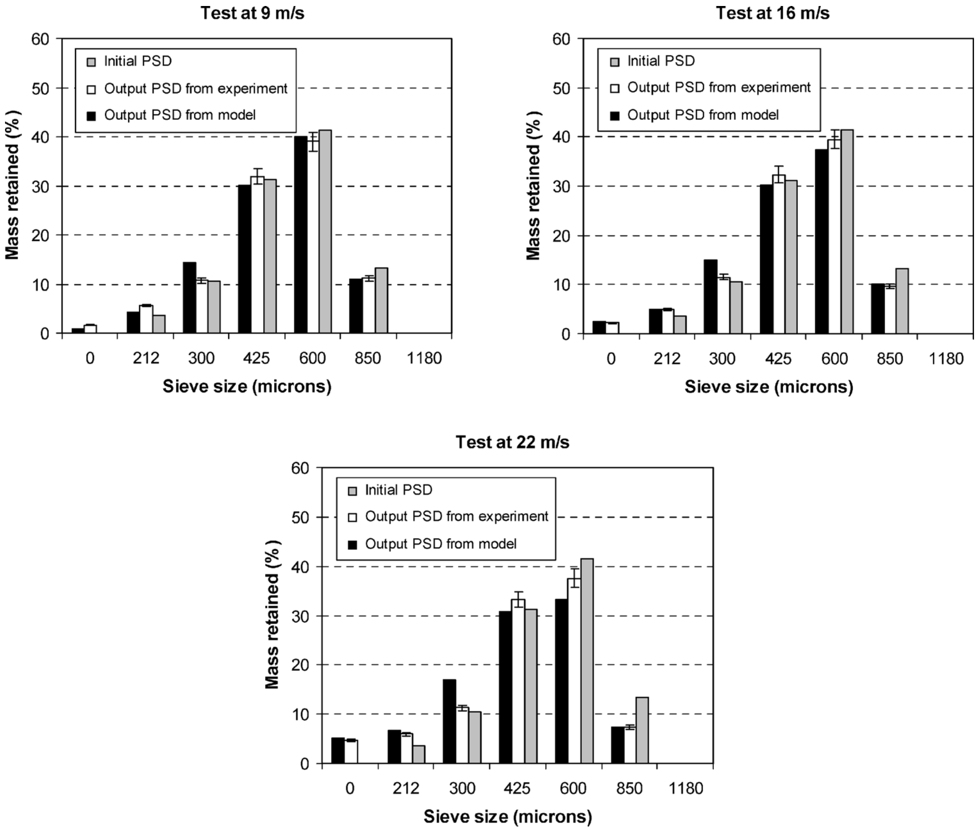


Figure 5. Comparison between the experimental results and the prediction of the model.

significant effect in the extent of particle degradation for granulated sugar being pneumatically conveyed through a series of 90° bends.

4. APPLICATION TO A LARGE-SCALE PNEUMATIC CONVEYING PROCESS

As an illustrative application, the model presented above has been used to study degradation in an industrial-scale dilute-phase pneumatic conveying system. The layout of the pipe system is shown in Fig. 6. The pipeline with internal diameter of 0.053 m includes various elements, i.e. horizontal pipes, 90° elbow bends (of curvature $R/D = 3.6$) and a vertically upward pipe.

The conveyed materials is granulated sugar of solids density 1600 kg/m^3 and with an initial size distribution identical to that used for the laboratory tests described in Section 3. The experimental breakage matrices determined in [6] for this material have been used in the degradation model. The simulations were performed for operational conditions usually employed in industrial practice, i.e. a conveying velocity of 18.5 m/s and a suspension density of 15 kg/m^3. In the inlet cross-section of the pipeline, particles were assumed to be homogenously dispersed, and the gas

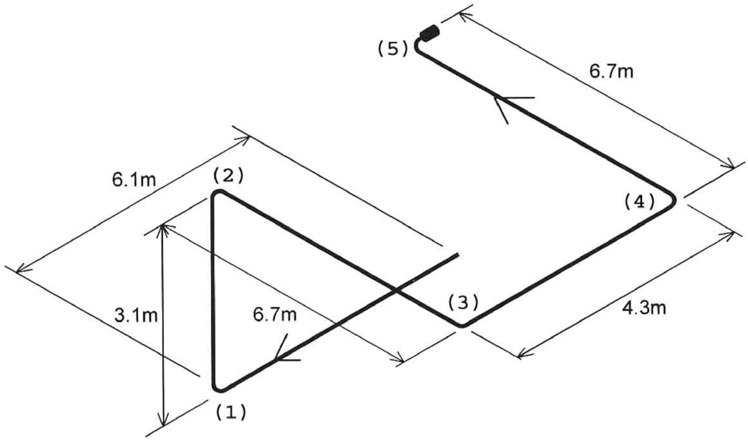


Figure 6. Layout of the industrial-scale pneumatic conveying line.

Table 1.
Values of the parameters in the flow model used for the calculations

Parameter	Value
Inlet air velocity (v_{air})	18.5 m/s
Mass flow rate of particles ($\dot{m}_s = \rho_{susp}\dot{V}_{air}$)	0.615 kg/s
Friction factor strand/suspended flow (f_{str})	3
Friction factor suspended flow/pipe wall (f_{susp})	0.0115
Friction factor air/pipe wall (f)	0.0044
Coefficient characterising the pressure drop caused by the particles in the gas (K)	1.6×10^{-6}
Bend wall friction coefficient (μ)	0.25
Coefficient of restitution (e)	0.8

and particle velocities were assumed to be identical. Table 1 summarizes the values of the parameters in the flow model used for the calculations. The dependence of the friction factor between the suspended flow and the pipe wall on the suspension density was assumed to follow the relation obtained by Bradley *et al.* [11]:

$$0.004 + 0.005\rho_{susp}. \tag{13}$$

An elemental length of 0.2 m was used for solving the governing equations of the strand-type flow. It was verified that an additional reduction in the value of the elemental length had no significant influence on the results of the simulation.

The predicted profiles of the air pressure, and the air and particle velocities along the conveying pipeline are presented in Fig. 7. The air expands along the pipe, with a total pressure drop over the whole pipeline of the order of 1.5×10^4 Pa. It can be seen that the pressure distribution downstream of each bend (the bends being located at a conveying distance from the inlet of 6.1, 9.2, 15.9 and 26.9 m, respectively) is characterized by a high-pressure gradient in the section immediately

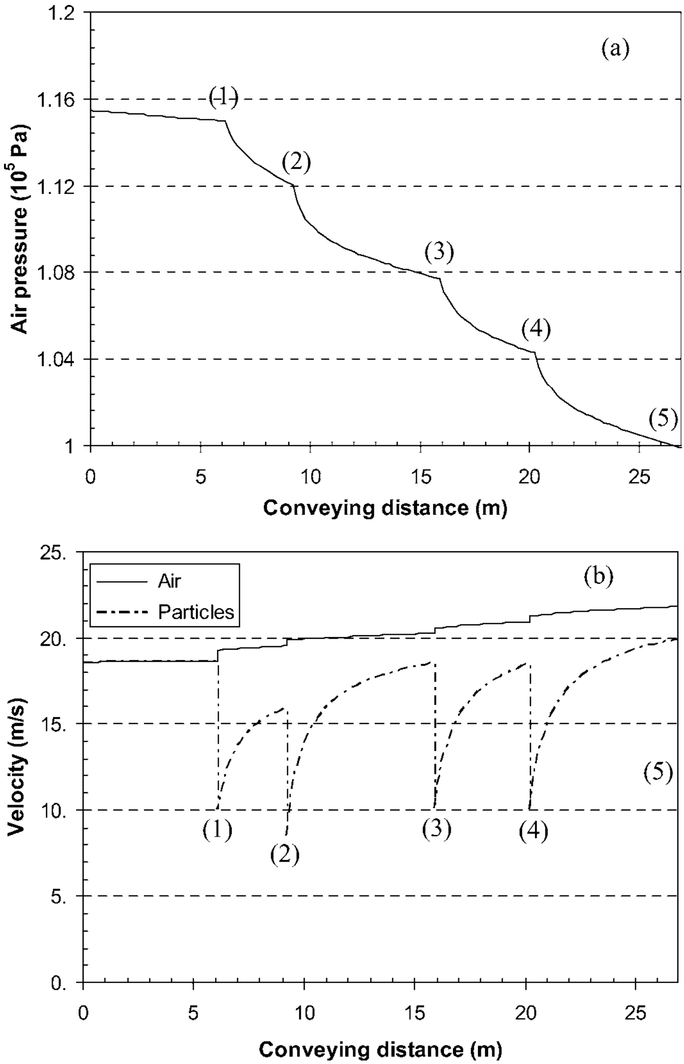


Figure 7. Prediction of the numerical simulation for: (a) air pressure, and (b) air and particle velocities profiles along the conveying line. The indexes (i) $(i = 1, 5)$ refer to the location of the bends along the conveying line.

adjacent to the bend, followed by a lower-pressure gradient far downstream of the bend. This pressure drop profile around a bend is similar to the one observed in the experiments of [17]. Experiments for a wide range of materials and operating conditions show that the length over which the development of the pressure drop induced by a bend occurs (i.e. the distance which is necessary for the pressure drop to approach a steady gradient) is of the order of 8 m (e.g. [8, 14]), which is in reasonable agreement with the value of 6 m predicted by the present model.

Expansion of the air causes an increase in the air velocity, and thus in the particle velocity as a result of the momentum transfer between the gas and solids phases. In a pipe bend, particles are considerably slowed down due to centrifugal forces and particle–wall interactions. The reduction of particle velocity around a bend is on average about 45%. In the downstream straight section of a bend, the particles, which are conveyed in the form of a strand, are progressively re-accelerated towards the air velocity. It may be noted that the drop of the particle velocity in the bend, and the subsequent creation of a strand, is accompanied with a marginal increase of the air velocity at the bend exit due to the reduction of the pipe cross-sectional area occupied by the air. For this example, none of the straight pipe sections is long enough to disperse completely the strand before reaching the next bend. Such dispersion of the strand can be observed in Fig. 8, which shows the profiles of the air and particle velocities along the pipeline for the same simulation conditions, except the length of the last straight pipe of the conveying system which has been set to 20 m in the present simulation (instead of 6.7 m previously). It can be seen that a transition from the strand-type flow to a fully suspended flow occurs in the last straight pipe at a conveying distance of about 14 m from the outlet section of the upstream bend.

Figure 9 shows the change in the harmonic mean size of the particle distribution after each bend. The harmonic mean size X_h was defined by the equation:

$$\frac{1}{X_\mathrm{h}} = \sum \frac{w_i}{d_i}, \tag{14}$$

where w_i is the fraction of material retained between sieves of mean size d_i.

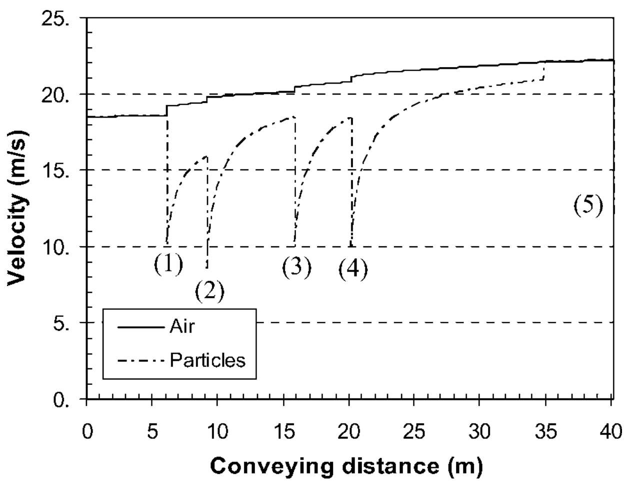


Figure 8. Calculated profiles of the air and particle velocities along the conveying line for a conveying system with a last straight pipe section of 20 m length. The indexes (i) ($i = 1, 5$) refer to the location of the bends along the conveying line.

The materials experiences significant degradation as the harmonic mean decreases at almost half of its initial value after conveying through five bends. The rate of degradation is approximately constant with respect to the number of bends, since particles in the present case have similar impact velocity in each bend (see Fig. 7).

The full particle size distribution after each bend is presented on Fig. 10. As expected, degradation in the bend considerably alters the particle size distribution. It can be seen that as the number of passes through a bend increases, the fractions of the largest particles (>425 μm) decrease, whereas the fractions of the smallest particles (<425 μm) increase. At the end of the conveying line, almost 18% of

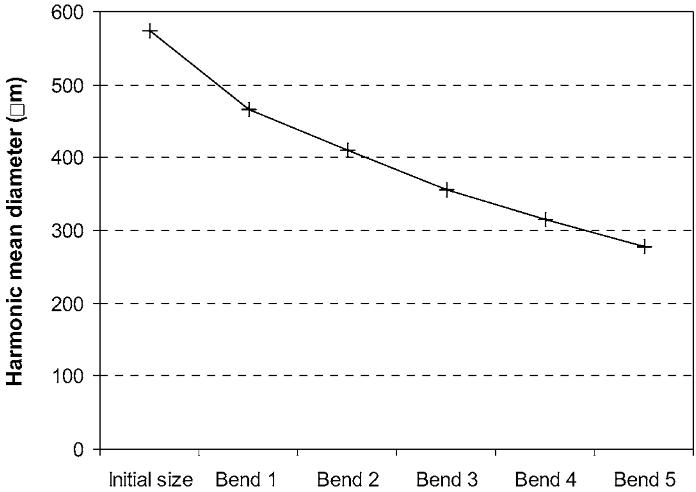


Figure 9. Harmonic mean sizes of the particle distribution after each pipe bend.

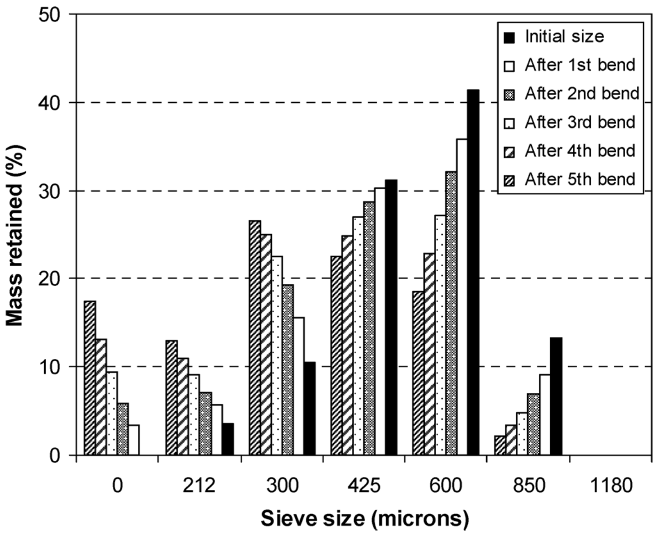
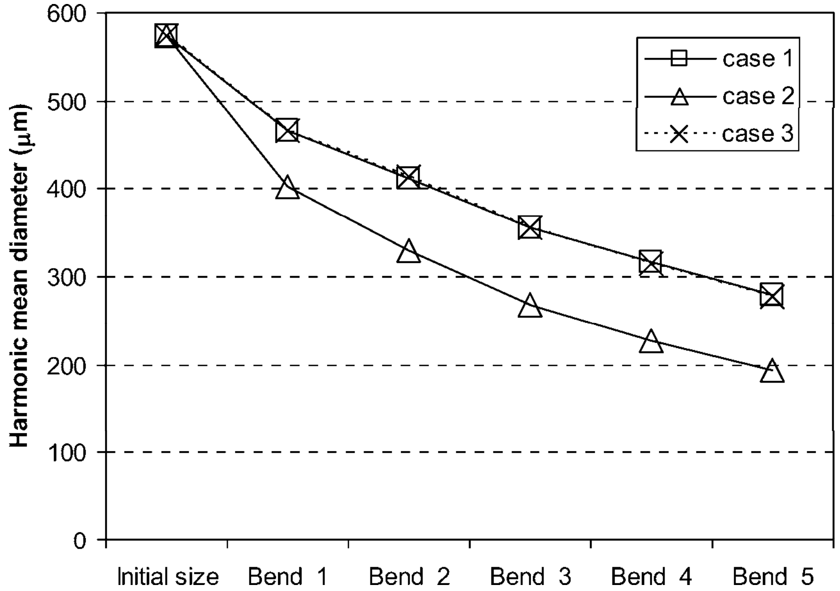


Figure 10. Particle size distributions after each pipe bend.



	Air velocity at inlet (m/s)	Mass flow rate of particles (kg/s)
case 1	18.5	0.615
case 2	24.1	0.615
case 3	18.5	0.800

Figure 11. Harmonic mean sizes of the particle size distribution after each pipe bend under different inlet air velocities and mass flow rates of particles.

the material turns out to be dust (defined as particles smaller than 200 μm) and the fraction of particles larger than 850 μm decreases to 80% of its initial value. The high degradation observed in our simulation can be explained by the relatively high conveying velocity used and the large bend angle considered.

Numerical predictions for the particle harmonic mean diameter after each bend under different inlet gas velocities and particles mass flow rates are shown in Fig. 11. It can be seen that the inlet gas velocity has a very significant effect on particle degradation. Indeed, a 30% increase of the inlet air velocity under the same particle mass flow rate (cases 1 and 2) causes a drop of approximately 30% of the particle mean diameter at the outlet of the conveying system. Figure 11 shows also that the particle mass flow rate has a negligible influence on particle degradation over the range of values investigated, as the particle mean diameter after each bend remains essentially unchanged between cases 1 and 3. The primary influence of the air velocity on degradation is in agreement with the experimental observations commonly reported in the literature (e.g. [21, 22]). The fact that no significant effect of the particle mass flow rate on particle degradation is predicted in the simulation may be attributed to two reasons. First, according to our analysis, the particle mass flow rate plays a role only in the fully suspended flow mode [see (7)], which controls the flow in the present simulation in a limited portion of the conveying system, i.e. the first straight pipe (see Fig. 7). Second, the effect of collisions of particles with other particles is not accounted for in the model in its present form. Interparticle collisions at high particle mass flow rate may cause a decrease in the particle velocity and hence reduce the level of degradation undergone by the particles.

5. CONCLUSIONS

Calculation of the propensity of particle degradation due to impact, based on the experimental determination of breakage matrices, has been combined with a simple physical flow model of the gas and solids phases to predict degradation during dilute-phase pneumatic conveying. By comparing the results of the degradation model to those obtained in a pilot-sized pneumatic conveyor test rig, it has been demonstrated that degradation occurring in a 90° angle pipe bend can be adequately described using 90° angle single-impact experimental data, hence supporting the main assumption of the developed degradation model. Numerical simulations for a large scale pneumatic conveying system were presented and discussed.

6. ONGOING WORK

The performance of the model developed in this work is promising and further experimental work is planned to obtain data sufficient to validate the model for an industrial-scale pneumatic conveyor. Moreover, work is currently underway in order to enable the model to account for different bend angles, multiple impacts in a bend and the effect of the number of impacts on the particle degradation behavior. This should lead to a powerful engineering tool, which can be employed for process control and the prediction of the optimum operating conditions.

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