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Hydrodynamic modeling of downward gas–solids flow. Part I: Counter-current flow

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ABSTRACT

The one-dimensional model of accelerating turbulent downward counter-current gas–solids flow of coarse particles was formulated and experimentally verified by measuring the pressure distribution along the transport tube. The continuity and momentum equations were used in the model formulation and variational model was used for the prediction of the fluid–particle interphase drag coefficient.

Experiments were performed by transporting spherical glass particles 1.94 mm in diameter in a 16 mm i.d. acrylic tube at constant solids mass flux of 392.8 kg/m^{2}s. Tube Reynolds number ranged from 170 to 5300 and the slip Reynolds number from 650 to 1060. Under these conditions loading ratio (G_p/G_f) varied between 66 and 2089. Visual observations show that particles flow downward in apparently homogenous dispersion. Experimental data for the static fluid pressure distribution along the transport tube agree quite well with the model predictions. The mean voidage and the particle velocity decrease, while the slip velocity increases with the increase in gas superficial velocity.

The values of the pressure gradient, porosity, particle velocity and slip velocity along the tube were calculated according to the formulated model. In these calculations, particle–wall friction coefficient was determined indirectly by adjusting the f_p value to agree with the experimental data. The effect of the value of f_p on the model calculations was significant.

Calculations show that the acceleration length for the same particles (1.94 mm) in downward counter-current gas–solids flow is about two times higher than the acceleration length in upward co-current gas–solids flow. In the system investigated, “choking” occurs at slip velocity which is about 73% of the single particle terminal velocity.

1. Introduction

Gas–solids contact in vertical tube in dilute regime (solids volume concentration less than about 5%) can be achieved in three different modes: co-current up flow, co-current down flow and counter-current down flow. Co-current up-flow (risers or pneumatic transport) usually operate in dilute regime with both gas superficial velocity and slip velocity higher than the single particle terminal velocity. If riser tube diameter is relatively large solids back mixing and non-uniformity in radial distribution of particle velocity and voidage often appear since a core–annulus structure is formed [1]. Compared to co-current risers, gas–solids co-current downers usually have the advantages of much less solid back mixing and shorter gas–solids contact time, but suffer the disadvantage of very low solids holdup [2]. As in this configuration both gas and solids travel in the direction of gravity, the flow accelerates quickly. The solids holdup that can be achieved in gas–solids co-current downers is usually below 1%. They are therefore used for the reactions with short residence time and low solids/gas ratio. The application of the downer reactors for the systems requiring higher solids/gas ratio can be achieved by introducing the gas upwards–solids downwards counter-current flow configuration. This configuration is expected to combine the advantages of the co-current risers and co-current downers: higher solids holdup of the risers and low solids back mixing of the downers. Gas–solids counter-current downers have a principal advantage compared to co-current risers and downers since they have higher solids hold-up, low solids back mixing and residence time of solids is more uniform and can be varied in a wider range [2,3]. Co-current risers and downers were investigated over the years [4–7]. Despite the high potential of gas–solids counter-current downers, they attracted much less attention, especially for coarse particles.

Luo et al. [8] investigated axial pressure gradient profiles and actual solids holdups in 25 mm i.d. counter-current downward flow with fine particles. They observed recirculation at the wall especially at high gas velocities. Their experimental results indicated that the pressure gradient could be used for the estimation of solids holdup in the fully developed flow region.

There are several papers regarding the heat transfer applications of the counter-current downers. The concept of falling particle heat exchanger was proposed by Decher [9]. The concept was further

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developed by Thayer and Sekins [10] who used the melting particles falling down the tube to carry out experimental and numerical studies. Gat [11] proposed the two-phase equations of motion for the flow of spherical particles through the fluid by treating the particles as a pseudo-fluid. The numerical analysis of a direct-contact heat exchange using particle–suspended gas as a heat transfer medium was performed by Park et al. [12]. Experimental studies on counter-current gas–solids heat exchanger system for steel and alumina particles have been carried out by Sagoo [13]. Rajan et al. [14] performed the simulation of the counter-current gas–solids heat exchanger using the two-fluid model. They investigated the effect of solids loading ratio and particle size on the heat transfer rate, temperature profile and thermal effectiveness of the gas. According to their simulation, the heat transfer rate increased with increasing solid loading ratio and decreasing particle size.

There are also several papers in Chinese language which investigate the counter-current downers in the field of coal topping process [ref. from [2]]. Schmid et al. [15] investigated the dual circulating fluidized bed for biomass gasification. The fuel gasification reactor investigated was a circulating fluidized bed with the characteristic of counter-current flow conditions for the gas phase and bed material particles.

Peng et al. [2] conducted a CFD study on the gas–solids flow in counter-current downer reactor. Solid phase used were FCC catalyst particles 217 μm in diameter. In order to improve the simulation results, the authors introduced a new interphase drag correlation which takes into account particle clustering effect which is important for small particle systems. The proposed drag correlation is then integrated into Eulerian model to study the hydrodynamics of the gas–solids flow in a counter-current downer reactor. It was shown that the model used can predict the main features of the counter-current downer investigated. The measured axial pressure distribution and radial solid concentration profiles were reproduced reasonably well. Jia et al. [3] conducted similar numerical study of counter-current gas–solids flow in FCC disengager and stripper with the main aim to investigate conditions for the prevention of coke formation in the FCC reactor. These authors also used the effective cluster diameter of the particles in the calculations. Their general conclusion was that simulations agree well with the measurements conducted in an oil refinery.

There are no experimental data or theoretical studies for the coarse particles flow in the counter-current downer reactors in the literature, despite the potential for their practical usage. The application of the counter-current downers with coarse particles include the heat transfer applications, draft tube spouted beds with shallow annular section, first stage dryers of fertilizer granules in order to avoid the particles agglomeration in the second step fluidized bed dryers. The study of the counter-current downer rectors is also important from phenomenological point of view. For example, there is no data in the literature about the flow regimes in the counter-current downers, or about the "chocking" phenomena which occurs in the system at certain flow conditions. The data of the particle–wall friction coefficient f_p in downer reactors is also missing. The flow of the fine particles in the counter-current downers is often accompanied by particles clustering due to cohesive and electrostatic effects. In these cases the concept of effective cluster diameter was proposed [2], but the general correlation for the cluster diameter calculation is still not available.

The aim of this study is the formulation of the one-dimensional mathematical model of dilute accelerating turbulent downward gas–solids flow of coarse particles and its experimental verification by measuring the pressure distribution along the transport tube. Essentially, the model of upward mixture flow proposed earlier [16] was modified for the application in downward counter-current gas–solids flow.

2. Experimental

The experiments were carried out in an air–particle system, using the pneumatic transport tube schematically shown in Fig. 1. In the

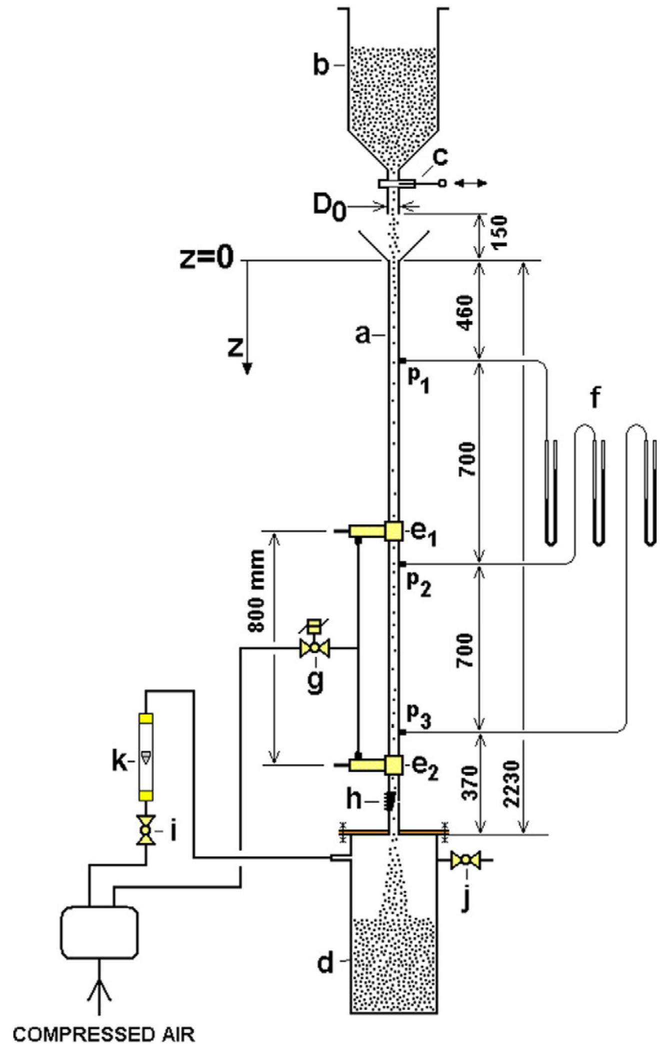

Fig. 1. Schematic diagram of the experimental system. a — transport tube, 16 mm i.d.; b — feeding reservoir (hopper); c — knife valve; d — receiving reservoir; e_1, e_2 — traps; f — manometers; g — electromagnetic valve for closing the traps; h — connection point; i — valve; j — on–off valve; k — rotameter; p_1…p_3 — pressure taps.

counter-current experimental setup in this paper, the particles move downwards by free falling from the feeding reservoir (b) while the upwards air flow is induced using the compressor. The transport tube (a) was of internal diameter 16 mm and 2230 mm of length. The pressure profile along the tube was measured at three points (p_1, p_2 and p_3) using manometers. The concentration of the particles in the transport tube was measured using two closing traps placed at a distance of 800 mm from each other (e_1 and e_2). The traps were operated using an electromagnetic valve (g) in such a way that they closed simultaneously. The traps were made of perforated plates in order to collect the particles, and to allow the air flow through them. The collected particles were weighed and the solids concentration was calculated using these data. The other elements of the pneumatic transport tube are shown in Fig. 1.

Spherical glass particles 1.94 mm in diameter were used in the experiments. The density of the particles was $\rho_p = 2507$ kg/m^3. The terminal velocity of the particles according to the correlation of Kunii and Levenspiel [17] was $U_t = 11.25$ m/s. Minimum fluidization parameters (U_{mF} and ε_{mF}) required to calculate variational constants for the prediction of the drag coefficients were determined experimentally using standard fluidization column 60 mm in diameter. The values of the minimum fluidization velocity and the porosity at minimal fluidization were $U_{mF} = 0.985$ m/s and $\varepsilon_{mF} = 0.422$ [16]. The physical characteristics

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