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Fluidization of Geldart Type-D Particles in a Swirling Fluidized Bed

[1]Mohd Faizal Mohideen, Suzairin Md Seri and [2]Vijay R. Raghavan

[1]Faculty of Mechanical & Manufacturing ,Universiti Tun Hussein Onn Malaysia
86400 Batu Pahat, Johor, Malaysia

[2]Department of Mechanical Engineering, Universiti Teknologi Petronas,
31750 Tronoh, Perak, Malaysia

[1]mfaizal@uthm.edu.my,suzairin@uthm.edu.my, [2]vijay_raghavan@petronas.com.my

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Abstract—Geldart Type-D particles are often associated with poor fluidization characteristics due to their large sizes and higher densities. This paper reports the hydrodynamics of various Geldart Type-D particles when fluidized in a swirling fluidized bed (SFB). Four different sizes of particles ranging from 3.85 mm to 9.84 mm with respective densities ranging from 840 kg/m3 to 1200 kg/m3 were used as bed material to study the effect of various bed weights (500 gram to 2000 gram) and centre bodies (cone and cylinder) for superficial velocities up to 6 m/s. The performance of the SFB was assessed in terms of pressure drop values, minimum fluidization velocity, Umf and fluidization quality by physical observation on regimes of operation. The swirling fluidized bed showed excellent capability in fluidizing Geldart Type-D particles in contrast to the conventional fluidized beds. The bed pressure drop of increased with superficial velocity after minimum fluidization as a result of increasing centrifugal bed weight. It was also found that the particle size and centre body strongly influence the bed hydrodynamics.

INTRODUCTION

Fluidization is the operation by which solid particles are transformed into fluidlike state through suspension in a gas or liquid. This method of contacting between solid and fluid has some unusual characteristics, and fluidization engineering puts them to good use [1]. The basic mechanism of a fluidized bed can be seen simply as fluid percolation through particle interstices via a distributor, in which particles begin to exhibit fluidlike characteristic upon experiencing sufficient drag force by the fluid. A large number of industrial processes use the fluidization technique in their daily operations, viz., combustion, gasification of solid fuels, drying of particles, particle heating, oxidation, metal surface treatments and catalytic and thermal cracking [2].

The fluidization phenomena of gas-solids systems depend very much on the particle characteristics. Geldart [3] was the first to classify the behavior of solids fluidized by gases into four distinct groups, namely A, B, C and D, characterized by the density difference between the particles and the fluidizing medium, $(\rho_p - \rho_g)$, and by the mean particle size, d_p. Fig. 1 shows Geldart's classification of particles.

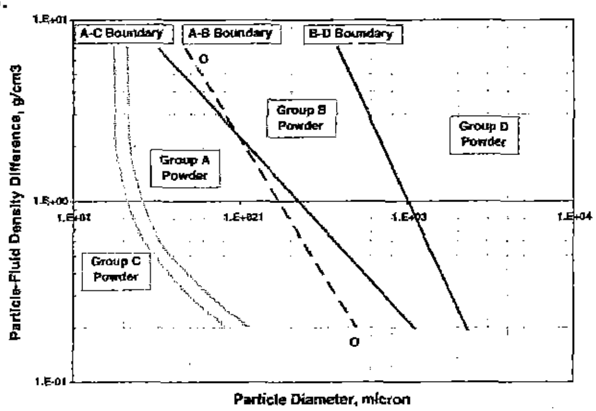

Figure 1: Geldart's classification of particles [*reproduced from* [1]]

Geldart type-D particles are typically large (mean size larger than 0.6 mm) and denser than other categories. They require higher velocities to fluidize the bed than other categories, resulting in the gas flow through the particle voids becoming transitional or turbulent. The bubbles which cause mixing of particles in the bed, now coalesce easily to form larger but fewer bubbles. Hence the Geldart type-D particles are difficult to fluidize, especially for deep beds and do not mix well [1, 2] though spoutable. Despite their use in a large number of applications, especially in food and biomass processing, this type of particles, and its hydrodynamics in particular, have received rather less attention in publications. Cranfield and Geldart [4] studied the fluidization characteristics of large particles (1–2 mm) and discussed advantages of using fluidized beds of large particles for certain applications. Rhodes [5] reviewed a number of research works on coarse particles in discussing his findings on turbulent fluidization. The mechanisms of gas flow and bubble characteristics of fluidized beds of coarse particles were investigated by Glicksman et al. [6].

The present study explores the capability of a relatively new technique in fluidization; the swirling fluidization technique in fluidizing the Geldart type-D particles. The swirling fluidized bed (SFB) which is annular in shape with inclined injection of fluidizing gas is used with spherical PVC particles with diameters ranging from 3.85 mm to 9.84 mm and densities ranging from 840 kg/m^3 to 1200 kg/m^3. The bed was investigated for flow regimes, bed pressure drop ΔP_b, minimum fluidization velocity, U_{mf} and minimum swirling velocity U_{ms} experimentally. Various bed configurations were studied - different centre bodies (cone and cylinder) and bed weight from 0.5 kg to 2 kg for superficial velocities, V_s up to 6 m/s.

Related Works

One type of fluidized bed that operates using swirling technique is the swirling fluidized bed (SFB), which is the main focus in this study. The bed is annular type, featuring angular injection of gas and swirling motion of bed material in a circular path as shown in Fig. 2.

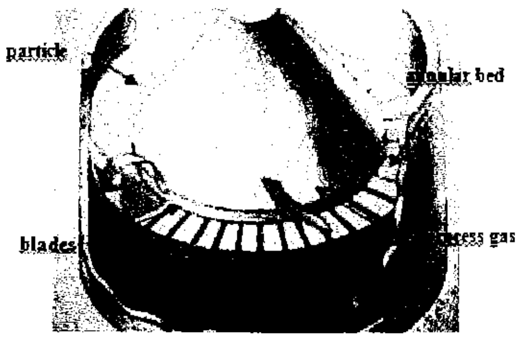

Figure 2: Basic configuration of a Swirling Fluidized Bed (SFB)

The principle of operation is based on the simple fact that a horizontal component of gas velocity in the bed creates horizontal motion of the bed particles. A jet of gas enters the bed at an angle θ_b to the horizontal. Due to angular injection, the gas velocity has two components. The vertical component $U_v = U \sin \theta_b$, causes lifting of the particles. It is this lifting force that is responsible for fluidization. The horizontal component $U_h = U \cos \theta_b$, creates a swirling motion of the particles [7,8,9]. The bed particles are also likely to undergo a secondary motion in a toroid-like path and be well mixed in the radial plane.

This variant of fluidized bed provides an efficient means of contacting between gas and particles. Elutriation of particles which has been a major limiting factor in the operation of the conventional fluidized bed is reduced significantly, since the vertical component of velocity is now only a small fraction of the net gas velocity. The cyclone-like features resulting from the swirling motion of bed particles also contribute to this low elutriation. Hence it is capable in fluidizing a wide variety of shapes of particles including the large ones.

Shu et al [8] studied a similar bed, termed as the toroidal bed, which is taken from the overall shape of the bed in the swirling regime. Relevant hydrodynamic behavior of the bed is measured with various inert materials in a pilot scale 400-mm toroidal fluidized bed reactor. The observed hydrodynamic behavior is found to be predictable at ambient temperature by conventional hydrodynamic models.

Sreenivasan and Raghavan [9] developed an analytical model on the hydrodynamics of a swirling fluidized bed. The model is put forward to predict the angular velocity of the swirling bed at given air flow rate and also the pressure drop of the swirling fluidized bed. In this model, the bed is treated as a lumped system; the whole bed is a single swirling mass of uniform angular velocity. The model was developed based on the conservation of angular momentum principle and the authors validated the model with experimental works.

Recently, Kaewklum & Kuprianov [10] investigated the hydrodynamic regimes and characteristics of a conical fluidized bed operating with annular-blade distributor. Using quartz sand with four different particle sizes, they compared the pressure drop and minimum fluidization velocity with four-nozzle tangential entry system which also generates swirling motion in the bed. From the cold model tests, they concluded that the method of air injection substantially affected the hydrodynamics and fluidization regimes. However, the swirling effect can only be achieved at superficial velocities higher than 3 U_{mf} and limited to low bed heights.

Though comprising such merits, the SFB also comes with certain drawbacks. Towards attending to these deficiencies through the proposal of a novel distributor, it is imperative to first understand fully the existing bed characteristics and other bed configurations which have not been addressed in the literature. Hence, the following section reports the findings from experimental studies in a SFB.

Methodology

A well-planned methodology is important to meet the desired project objectives. In this section, details of the experimental apparatus including distributor design, blower selection, flow measurement and experimental procedure for pressure drop measurement are discussed.

Annular Distributor. The distributor assembly consists of lower and upper flanges together with inner plates respectively, holding the blades firmly to form the annular distributor, somewhat similar to that used by [7,8,9,10]. The centre body is important to avoid the possible creation of 'dead zone' at the centre of the bed during operation with bed materials. Air enters the plenum chamber via tangential entry and expands before entering the annular blade distributor.

Particle Properties and Experimental Configuration. Particles used in the experiments are large spherical PVC beads, which fall in Geldart type D particles as proposed by [3]. Four different sizes of particles are used, with their respective density and diameters are shown in Table 1 below:

TABLE 1: PARTICLE PROPERTIES AND EXPERIMENTAL CONFIGURATION

Particle	Size (mm)	Density (kg/m^3)	Fraction of Open Area (FOA, %)	Centre Body	Bed Weight (kg)	Superficial Velocity (m/s)
1	3.85	3954	12.9	Cone	0.5 – 2	0 – 5
2	5.75	950	12.9, 17.2	Cone, Cylinder	0.5 – 2	0 – 6
3	7.76	918	12.9, 17.2	Cone	0.5 – 2	0 – 6
4	9.84	840	12.9, 17.2	Cone, Cylinder	0.5 – 2	0 – 6

Results and discussion

This section presents the results of the experimental investigations on the SFB. The findings from hydrodynamic characteristics for various bed configurations are presented. The following aspects are addressed:

i. Regimes of operation and hydrodynamics of SFB
ii. Investigation of various configurations in SFB:

- Variation of bed loading (0.5 to 2 kg)
- Variation of particle sizes (3.85 mm to 9.84 mm)
- Variation of centre body (cone & cylinder)

Hydrodynamics. Typical regimes of operation in a conventional fluidized bed include packed bed, minimum fluidization, bubbling, slugging and finally elutriation. While operating a SFB, one can distinguish different regimes of operation as shown in Fig. 3. Though the packed regime (Regime I) still exists, progressive increase of fluidizing gas flow rate upon minimum fluidization led to a condition suitably designated as the minimum swirling condition (Regime II) where the bed almost swirls. A few particles started swirling gently at this point. Further increase in fluidizing gas flow rate results in the fully swirling motion of the bed (Regime III). At this condition, the bed is subjected to both fluidization and swirling where vigorous mixing occurs and interaction between gas and particles are intense. A similar regime was reported by [7], but designated as '*Torbed regime*'. This regime was the largest regime where the particles tend to swirl faster with the increase of fluidizing gas (thus increasing pressure drop) until finally reaching elutriation (Regime IV) for shallow beds (bed weight less than 1000 gram). Beyond the U_{mf}, the ΔP_b slightly reduced to U_{ms} where the particles rearrange themselves to allow better gas flow through the bed. Beyond the U_{ms}, in contrary to conventional fluidized beds, the ΔP_b was found to increase with V_s.

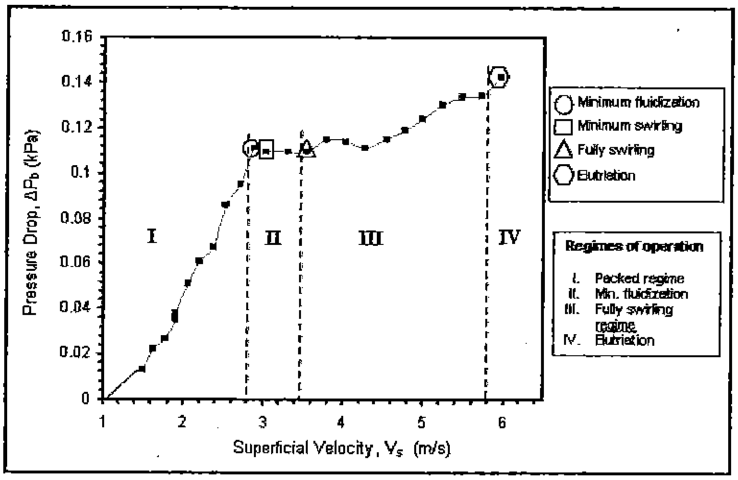

Figure 3: Regimes of operation in the SFB for shallow bed (1000 gram bed loading) with cylinder as centre body

For deeper beds, 1500 gram bed loading for instance, a two-layer bed is observed as reported by [8]. In the two-layer bed which occurs at a static bed height greater than 45 mm, a continuously swirling bottom layer and a vigorously bubbling top layer are visible upon minimum swirling velocity, U_{ms}. This is because the horizontal component of the velocity is attenuated and finally vanishes at the interface between the two layers as a result of continuous momentum transfer inside the bed. This regime is shown in Fig. 4.

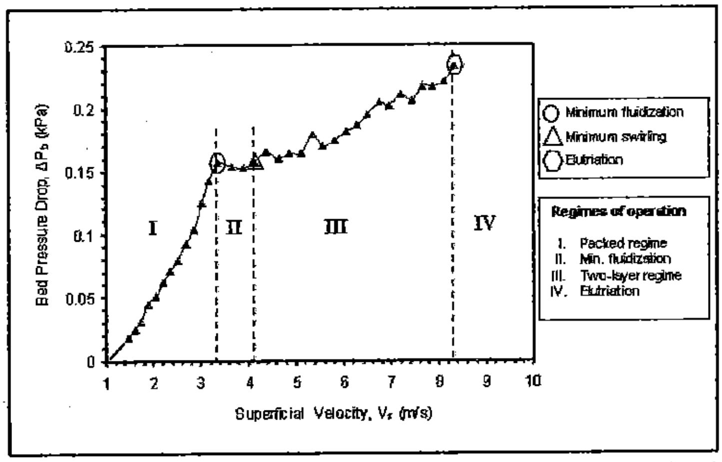

Figure 4: Regimes of operation in the SFB for deep bed (1500 gram bed loading) with cone as centre body

The two-layer regime of operation is unlikely to be favorable for actual industrial use due to the gradient in momentum transfer between the bottom swirling layer and the upper bubbling layer. This may cause significant variation in heat and mass transfer in the bed, affecting the quality of the end product. Hence, many researchers [7-10] working on similar beds opt for shallow bed operations. It was also found that similar regimes of operation were obtained for all particles investigated in the current work.

Effect of Various Bed Configurations. Various configurations of the swirling fluidized bed as in Table 1are studied through batch experiments and presented in the following section.

Effect of Variable of Bed Loading. Bed loadings were increased from 500g to 2000g in steps of 500 grams to investigate the effect of variable bed loading, which also corresponds to its respective bed height. Fig. 5 shows ΔP_b against V_s with cone as centre body. As mentioned earlier, ΔP_b increased with the increase of V_s upon U_{ms}. This distinct feature differentiates the SFB from conventional beds. The reason for this feature is the increase of centrifugal bed weight, which results in higher wall friction as proposed by [6], apart from increasing friction between particles. For deep beds, i.e. higher bed loadings, a two layer bed appeared, a swirling bottom layer and bubbling top layer as discussed earlier. Higher bed loading naturally impose higher pressure drops. Throughout the swirling regime, intense particle mixing and a very high degree of solid-gas contact is observed.

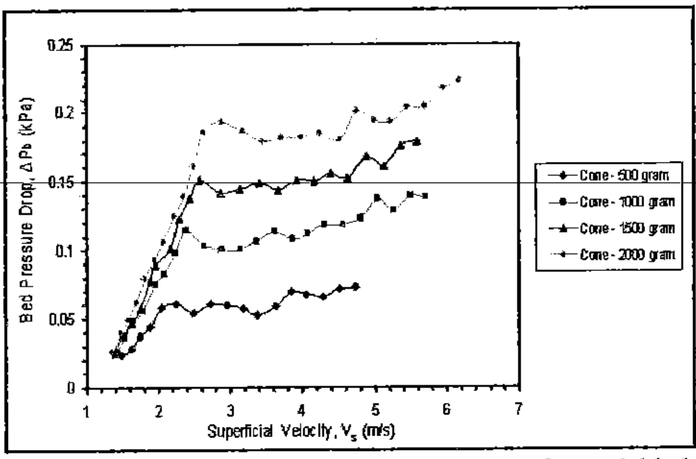

Figure 5: Bed pressure drop against superficial velocity for variable bed loading

Effect of Particle Size. Batch experiments were conducted with two different particles, 5.75 mm and 9.84 mm for two different bed weights as in Fig. 6. The effect of particle geometry was the aim of this experiment

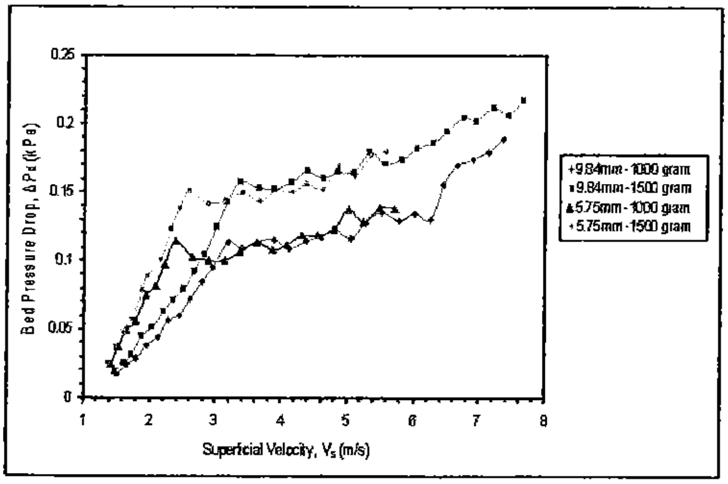

Figure 6: Bed pressure drop against superficial velocity for different bed weight

It can be seen that in the packed region, larger particles have lower pressure drop for both bed weights. This is due to the fact that smaller particles actually have a larger surface area. Larger particles, on the other hand, are capable of withstanding higher superficial velocity and hence longer swirling. Similar trends are found for 1500 gram bed weight for both particle sizes.

Effect of Different Centre Bodies – Cone and Cylinder. In a SFB with annular blade distributor, presence of a centre body is desirable to avoid 'dead zone' in the centre of the bed. The 'dead zone' may simply occur since the angular momentum of fluidizing gas at the centre of the bed is insufficient to either fluidize or impart swirling motion to the particles. Thus, the centre body is imperative in a SFB and as a consequence of the presence of centre body, the bed is now annular in shape. Two centre bodies investigated in the present study are cone and cylinder and some of the results are as in Fig. 7 and Fig. 8.

Fig. 7 shows bed pressure drop, ΔP_b for cone and cylinder, as a function of the superficial velocity, V_s. Two bed loadings, 1 and 1.5 kg are investigated for 5.75 mm particles. For the configuration shown, the ΔP_b is apparently high for higher bed loading since the bed is now deeper and the fluidizing air has to flow through longer distance from the distributor at the bottom.

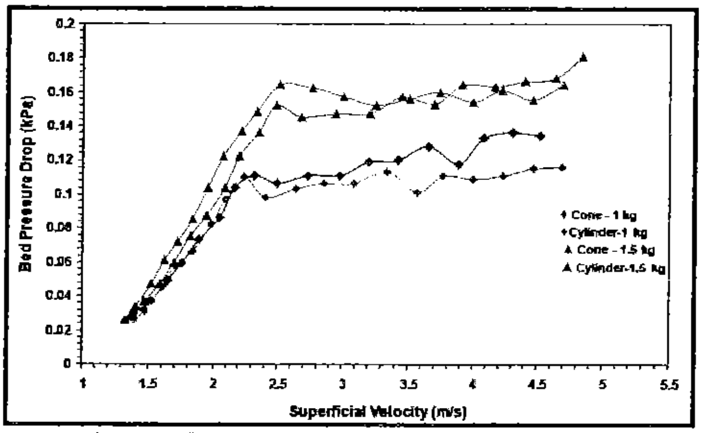

Figure 7: Bed pressure drop against superficial velocity for different centre bodies (5.75 mm particle)

In the packed region, ΔP_b is higher for bed with cylinder as centre body compared to cone. This is due to the increasing bed cross-sectional area for cone in the vertical direction, and hence providing less resistance for flow. Thus one can expect lower percolating velocity of air and shorter percolation route in the bed, compared to that of cylinder. Similar result is obtained with larger particles, 9.84 mm as depicted in Fig. 8.

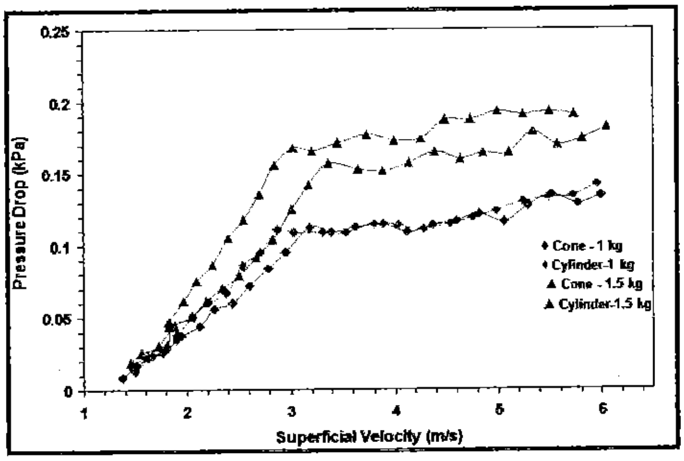

Figure 8: Bed pressure drop against superficial velocity for different centre bodies (9.84 mm particle)

On the other hand, as the bed reaches U_{ms} and starts swirling, the ΔP_b for bed with cone becomes higher (though there is a delay for 1.5 kg bed weight). This can be explained by the fact that the increasing cross-sectional area for cone that addressed earlier, also result in longer retention time of angular momentum. As a result, higher rate of momentum transfer occurs, compared to bed with cylinder as centre body. This is also supported by visual observation on the bed. Therefore, higher ΔP_b in swirling regime for cone should not be viewed as a demerit in this case. For larger particles, 9.84 mm as in Fig. 8, the ΔP_b in both packed regime and swirling regime for cylinder is substantially high since the bed is now deepest among all configurations and the effect of centre body is more clearly visible.

From the findings, it is apparent that the cone has advantage over cylinder as centre body and can be generally accepted as a better centre body and preferable for use in the swirling fluidized bed.

Conclusion

In conclusion, the SFB has been investigated through batch operations. The findings indicate that the sequence of flow regimes in swirling fluidized bed are packed bed, minimum fluidization, swirling regime, two-layer regime and finally elutriation or transport regime. Deep beds are prone to form partially fluidized regime and two-layer beds. Various configurations were also investigated and the study concludes as below:

i. The hydrodynamics of swirling fluidized bed are different from other conventional fluidized bed, in which the pressure drop increases with the mass flow rate of fluidizing gas
ii. Larger particles have lower pressure drop and capable of withstanding higher superficial velocity and hence, larger swirling regime
iii. Both particle size and centre body strongly influences the SFB behavior
iv. Cone is observed to have better advantage as centre body in a SFB compared to a cylinder.

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