COLD FLOW BEHAVIOUR OF A SWIRLING FLUIDIZED BED INCINERATOR

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ABSTRACT
This paper reports the experimental work carried out to determine the cold flow behavior of a swirling fluidized bed incinerator (SFBI) with Quartz sand as bed particles. Three different sizes of sand particles were studied: 425-600µm, 600-850µmand 850-1180µm for four bed heights: 25 mm, 30 mm, 35 mm and 40 mm. The density of the particles was 2450 kg/m³ while the gas superficial velocity (Uₘ) was limited to 3 m/s. It was found that the larger particles exceeded the swirl-able regime and only possessed bubbling regime during fluidization. It was also found that the bed pressure drop was higher for higher bed heights and increased when the bed operates in the swirling regime. The study concludes that bed with the smallest particle size (425-600µm) and highest bed height (40 mm) has the desired characteristics to be used in the SFBI.

Keywords: cold flow behavior, swirling fluidized bed incinerator, pressure drop, minimum fluidization velocity.

INTRODUCTION
Fluidized bed technology is long known for its vast industrial applications such as thermal catalytic cracking of fuels, drying, combustion, gasification, particle heating and many other processes that involve solid-fluid contact. Fluidized beds are simply a bed of particles in which fluid flows through the bed and the drag force by the fluid is large enough to suspend the bed. This intimate contact between the bed and fluid provides huge potential for reaction to take place between the two, in comparison to its packed condition. Many variants of fluidized bed are available in the industry at present where each of them is usually designed for a specific application. Among these variants is one which operates based on swirling principle where swirling motion is created inside the bed to promote lateral mixing of the bed, increase gas residence time and reduce elutriation of finer particles. These types of bed are usually compact by having significant aspect ratio (diameter to height). In general, swirling may be achieved by several ways such as secondary injection into the freeboard using nozzles or using a distributor which provides inclined injection of air to the bed. The latter approach has recently gained interest among researchers particularly in combustion studies. However, before a thermal system is developed for industrial use, its cold flow behaviour must be determined first to estimate the system’s resistance which later translates into operational costs. In the case of a fluidized bed system, it’s cold flow behaviour includes the distributor pressure drop (Δp₂), bed pressure drop (Δp₅), minimum fluidization velocity (Uₘ₅) and regimes of operation.

Kaewklum et al., 2009, have studied the hydrodynamic regimes and characteristics of a conical swirling fluidized bed for Quartz sand of four different sizes and compared the effect of tangential and axial entries into the bed. They have also investigated the effect of various bed heights and proposed a mathematical model to predict minimum fluidization velocity (Uₘ₅) and bed pressure drop (Δp₅). They reported that the bed exhibits four distinct regimes of operation, namely fixed-bed, partially fluidized-bed, fully fluidized-bed with partial swirling and fully swirling fluidized-bed. They concluded that both air entry types influenced the bed’s hydrodynamics significantly and they opted the axial entry bed with annular-blade distributor for their following studies on biomass combustion (Kaewklum and Kuprianov, 2010).

Much earlier, prediction of minimum fluidization velocity and particle trajectory was done by Shu et al., 2000 through a minor modification of existing mathematical models available in the literature. Geldart type A and type C particles were used during experiments in a pilot scale Torbed® reactor. A similar effort was later reported by Sreenivasan and Raghavan, 2002 for large particles (Geldart type-D) in which the authors highlighted their important finding – the increase of bed pressure drop with increasing superficial velocity in swirling mode due to the centrifugal bed weight. An approximate model for predicting the bed’s angular velocity was also proposed by the authors for shallow bed heights by assuming the bed as a lumped system.

The hydrodynamic characterization and heat transfer studies in the swirling bed was also reported by Mohideen et al., 2012a for extra-large particle sizes (5 mm to 10 mm) for deeper beds. The authors reported the presence of secondary motion of the bed in atoroidal path and highlighted the potential of the bed in biomass processing. Hydrodynamic and drying experiments of oil palm frond (OPF) biomass by Mohideen et al., 2011 confirms this, and the authors highlighted the rapid moisture removal from both leaves and stem (petiole). Based on the hydrodynamics and drying kinetics of OPF, the authors concluded the swirling fluidization is a viable technique for drying of OPF despite its irregular shape and density variation.
Recently, Zakaria et al., 2014 reported the success of drying pliant media using a swirling fluidized bed dryer. In their study, the pliant media dried was an industrial grade sponge which at present is widely used for surface preparation in oil and gas industry. The wet sponge particles were dried at $U_{mf}$ to reduce energy loss due to bed pressure drop. This was because in a typical swirling bed, $\Delta p_b$ increases with increasing $U_s$ due to the centrifugal bed weight. The vigorous mixing of the pliant media allows quick drying without the re-wetting problem that usually occurs in hygroscopic material such as the sponge.

In this study, experimental work were carried out to determine the cold flow behavior of a swirling fluidized bed incinerator (SFBI) with Quartz sand as bed particles. Three different sizes of sand particles were studied: 425-600µm, 600-850µm and 850-1180µm for four bed heights: 25 mm, 30 mm, 35 mm and 40 mm. The findings of this study were imperative to develop a SFBI which is capable in incinerating various types of waste efficiently.

MATERIALS AND METHOD

Swirling Fluidized Bed Incinerator (SFBI)

The SFBI in this study uses an annular-blade type distributor, similar to the previous researchers (Shu et al., 2000), (Sreenivasan and Raghavan, 2002) and (Mohideen et al., 2012a). The plenum chamber provides the air tangentially to the base of the bed while the acrylic column provides good visibility of the bed for characterizing the bed operating regimes. Figure-1 shows the overall SFBI system with auxiliary equipment while Figure-2 depicts the details of the experimental set-up used for the old flow study.

Sample preparation

The Quartz sand in this study was first sieved to three size ranges: 425-600µm, 600-850µm and 850-1180µm. Water displacement method was used to determine the volume of knowns and mass and the density was found to be 2450 kg/m$^3$. This value and the average particle sizes were then used to classify the sand based on Geldart’s diagram for powder classification, which is reproduced from Howard, 1989 as in Figure-3.

Figure-1. The swirling fluidized bed incinerator (SFBI) system with auxiliary equipment.

Figure-2. SFBI detailed design.

Figure-3. Geldart’s classification for powders (Howard, 1989).
Experimental procedure

Sand of a given size was poured into the bed from top. The bed was then fluidized vigorously before the fluidizing air velocity was gradually reduced. The bed height was measured at three different locations using pre-attached scale on the bed column. Additional sand was poured (if necessary) until the bed reached the required height. Air was introduced to the bed gradually while the bed pressure drop was measured (between distributor and top of the bed). The $\Delta p_b$ values were recorded and the bed behaviour was observed closely to establish the regimes of operation. The supply air was limited to 3 m/s since this velocity was found to be appropriate in the present study from preliminary experiments. Upon recording the $U_s$ and $\Delta p_s$ values, experiments were repeated for the other bed heights: 30 mm, 35mm and 40 mm and other sand particle sizes. In the present study, the velocity was measured using a pitot-static probe just before the plenum chamber inlet while the $\Delta p_d$ values were measured using a digital-micro-manometer. Upon completion of experiments with particles, the bed was emptied and the distributor pressure drop, $\Delta p_d$ was recorded with increasing $U_s$.

RESULTS AND DISCUSSIONS

Distributor pressure drop against velocity

As shown in Figure-2, the distributor in this study was made by an array of truncated annular blades with a centre body (cone) in the middle. These blades were inclined 15° to the horizontal plane to provide inclined injection of air into the bed. The total number of blades was 45 which correspond to a fraction of open area (FOA) of 15%. This relatively large FOA yields a low distributor pressure drop, $\Delta p_d$ without compromising the fluidization quality (Shu et al., 2000 and Sreenivasan and Raghavan, 2002). Figure-4 shows the $\Delta p_d$ against $U_s$ in the present study.

$$\Delta p_d = 0.105 U_s^{2}$$

$R^2 = 0.999$

Figure-4. $\Delta p_d$ against $U_s$ for annular blade distributor in the SFBI.

Besides being one of the hydrodynamic characteristics, the $\Delta p_d$ also enables the determination of its individual contribution to the overall system resistance. Lower $\Delta p_d$ is always preferred but without any adverse effect on the fluidization quality. The $\Delta p_d$ here may be attributed to the abrupt contraction of the process gas at the inlet of the distributor, flow of the process gas through the opening of the distributor blades and sudden expansion at the distributor outlet. Based on linear regression analysis, the $\Delta p_d$ can be correlated with $U_s$ by the equation: $\Delta p_d = 0.105 U_s^2$ with a correlation coefficient ($R^2$) of 0.999, which indicates a second-order polynomial relation between both parameters.

Regimes of operation

The typical regimes of operation in a bubbling fluidized bed are packed bed, minimum fluidization, bubbling, slugging and finally elutriation. In the SFBI however, the regimes of operation were found to differ due to the swirling motion of the bed. Typical regimes of operation in the SFBI were shown in Figure 5 by taking the 600-850µm sand particles with 30 mm bed height as an example.

Referring to Figure-5, the bed was initially in packed condition whereby the $U_s$ is insufficient to provide enough drag force to suspend the bed. The bed remained in the same condition until random bubbles started to appear on the bed surface with further increase of superficial velocity, $U_s$. Now the bed expands to allow more air to flow through the bed and bubbles were seen on the surface of bed. The $\Delta p_b$ increased with $U_s$ linearly in this region. Further increase of $U_s$ results in more expansion of the bed until it reached the minimum fluidization, $U_{mf}$ where the whole bed was suspended and balanced by the drag force. From $U_{mf}$, a slight increase in $U_s$ resulted in the bed achieving minimum swirling condition denoted as $U_{ms}$ in Figure-5. At $U_{ms}$, the bed was seen to gently swirl and further increase of $U_s$ resulted in faster swirling of the bed. This was true for all bed heights. On the other hand, different regimes were observed for 600-850µm and 850-1180µm bed. For 600-850µm, the bed only swirls for 25 mm bed height and partially swirl for 30 mm bed height. Deeper beds were subjected to bubbling regime only. As for the 850-1180µm bed, the bed only possesses bubbling regime for the whole range of bed height indicating the particle size has exceeded the
swirl-able regime in the present bed configuration. These findings are summarized in Table-1 below:

**Table-1. Operational regime in the SFBI.**

<table>
<thead>
<tr>
<th>Sand particle size (µm)</th>
<th>Bed Height</th>
<th>25mm</th>
<th>30mm</th>
<th>35mm</th>
<th>40mm</th>
</tr>
</thead>
<tbody>
<tr>
<td>425-600</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>/</td>
<td>/</td>
</tr>
<tr>
<td>600-850</td>
<td>/</td>
<td>/</td>
<td>**</td>
<td>**</td>
<td>**</td>
</tr>
<tr>
<td>850-1180</td>
<td>**</td>
<td>**</td>
<td>**</td>
<td>**</td>
<td>**</td>
</tr>
</tbody>
</table>

**Legend**

- / / Excellent Swirling
- // Good Swirling
- / Bubble or partial swirl
- ** Bubbling

**Effect of bed height**

In this study, the three particles studied were from different categories of Geldart’s classification. While the 425-600µm and 850-1180µm particles were from Geldart type B and D, the 600-850µm particles were near the transition region between B and D. The behaviour of these particles was presented in Figures 6 to 8:

**Figure-6.** \( \Delta p_b \) against \( U_s \) for 425-600µm particles.

**Figure-7.** \( \Delta p_b \) against \( U_s \) for 600-850µm particles.

**Figure-8.** \( \Delta p_b \) against \( U_s \) for 850-1180µm particles.

In general, higher bed heights have larger \( \Delta p_b \) due to the larger bed mass acting on the given distributor area. Therefore, larger drag force was required to initiate both fluidization (as well as swirling) in the bed. In the packed region, the gradient of \( \frac{d(\Delta p_b)}{d(\Delta U_s)} \) was found to be constant indicating the linear increase of \( \Delta p_b \) with \( U_s \). However, the 425-600µm particles showed a distinct characteristic in which the \( \Delta p_b \) increased significantly beyond \( U_{mf} \) with a linear trend. The \( \frac{d(\Delta p_b)}{d(\Delta U_s)} \) however, was less than that of the packed region. The significant increase of \( \Delta p_b \) can be explained by the increasing centrifugal bed weight as a result of increased angular momentum at higher \( U_s \). Apart from that, the smaller particles sizes possess larger surface area for a given bed loading, thus translates to higher \( \Delta p_b \) due to augmented inter-particle friction apart from friction with distributor as well as the column wall.

In Figure-7 and 8, beyond \( U_{mf} \), the \( \frac{d(\Delta p_b)}{d(\Delta U_s)} \) for 600-850µm and 850-1180µm particles were relatively smaller than that for 425-600µm (Figure 6) due to the absence of swirling motion for all bed configurations (except for the 25 mm bed height for 600-850µm particles). It was also seen that the \( U_{mf} \) values for a given particle size were has a small difference for a given bed height. These \( U_{mf} \) values were summarized in Table-2.

**Table-2. Summary of \( U_{mf} \) values for a given bed height.**

<table>
<thead>
<tr>
<th>Sand particle size (µm)</th>
<th>Bed height (mm)</th>
<th>( U_{mf} ) (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>425-600</td>
<td>25</td>
<td>0.20</td>
</tr>
<tr>
<td></td>
<td>30</td>
<td>0.18</td>
</tr>
<tr>
<td></td>
<td>35</td>
<td>0.18</td>
</tr>
<tr>
<td></td>
<td>40</td>
<td>0.19</td>
</tr>
<tr>
<td>600-850</td>
<td>25</td>
<td>0.27</td>
</tr>
<tr>
<td></td>
<td>30</td>
<td>0.25</td>
</tr>
<tr>
<td></td>
<td>35</td>
<td>0.25</td>
</tr>
<tr>
<td></td>
<td>40</td>
<td>0.30</td>
</tr>
<tr>
<td>850-1180</td>
<td>25</td>
<td>0.53</td>
</tr>
<tr>
<td></td>
<td>30</td>
<td>0.54</td>
</tr>
<tr>
<td></td>
<td>35</td>
<td>0.66</td>
</tr>
<tr>
<td></td>
<td>40</td>
<td>0.69</td>
</tr>
</tbody>
</table>
Effect of particle size

The effect of various particle sizes for a given bed height were presented in Figures-9 to 12. It could be seen that the operating velocity range for 425-600 µm particles was limited to about 1.8 m/s due to elutriation of fines from the bed during experiments.

Typically the $\Delta p_b$ against $U_s$ for smaller particles were slightly larger in the packed region due to the higher surface area of the particles. From Figure-9, 10 and 11 (25, 30 and 35 mm bed height), the $\Delta p_b$ was found to be much higher for larger particles in the initial stages of fluidization.

Apparently, with continuous increase of $U_s$, the $\Delta p_b$ of 425-600 µm particles surpasses the $\Delta p_b$ of the other two particles. This may be explained by the regimes of operation undergone by each particle as addressed in the earlier in part of this paper. The two larger particles were experiencing bubbling regime and hence the $\Delta p_b$ becomes constant with higher $U_s$ beyond $U_{mf}$. In contrast, the 425-600 µm particles swirl vigorously and the centrifugal force of the bed resulted higher frictional force at the bed wall and hence increased the $\Delta p_b$. This effect however was not visible in Figure-12 due to the limited operating velocity of the 425-600 µm particles and significantly higher amount of bed weight for this bed height.

CONCLUSIONS

From the cold flow study carried out in the SFBI, it can be concluded that both particle size and bed height influences the bed behaviour, particularly its operating regimes. The 425-600 µm particles were able to swirl vigorously for all bed heights while the 600-850 µm particles were capable to swirl only for shallow bed height (25 mm). The 850-1180 µm was found to only possess bubbling regimes for all bed heights studied. Due to the swirling of the 425-600 µm particles, its $\Delta p_b$ increased with increasing $U_s$, and the particles have limited operating $U_s$ range up to 1.8 m/s only. From the study, it may concluded that the 425-600 µm particles with 40 mm bed height was advantageous to be used in the SFBI due to the excellent swirling nature of the bed as well as having the largest surface area for heat transfer during incineration.

REFERENCES


