Miniaturisation of the toroidal fluidization concept using 3D printing

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Abstract: We used the stereolithography printing technique to fabricate a toroidal fluidized bed at the smallest scale ever achieved (50 mm diameter with 10 mm annular width). In toroidal fluidization, most of the kinetic energy of the fluidizing gas is used to induce swirling of the particle bed meaning higher gas velocities can be used without entrainment. The end-goal of this research is to use this ‘mesoscale-TORBED’ for screening adsorbents for CO\textsubscript{2} capture, where the intensified heat/mass transfer rates can potentially minimise the troublesome scalability issues encountered in standard packed beds, a consequence of gradient effects. Here, we have performed a comprehensive parametric study to understand the influence of bed loading, gas volumetric flow rate, gas temperature and gas humidity on the swirling bed formations of activated carbon pellets in order to identify appropriate conditions for sorbent screening. We show that desirable ‘uniform packing’ occurs across a broad range of operating conditions and identify the lower bed loading limit as 1200 mg. Other observed bed formations included collapsed, maldistributed and entrained states that caused gas bypassing of the particle bed. At the lowest air flow rate studied (27 L/min), the bed was either in the collapsed state or not swirling at all, whilst swirling was readily observed at the intermediate and high air flow rates (35.5 L/min and 44 L/min respectively). Humidity and air temperature had minimal influence over the flow patterns.

Keywords: Swirling fluidization, micro-fluidized bed, additive manufacturing, temperature, humidity
1 Introduction

Fluidized beds provide a degree of process intensification compared to fixed (packed) beds because the movement of the particles facilitates improved heat and mass transfer. Additionally, conditions within fluidized beds are uniform, meaning there are fewer gradient or dynamic effects such as those found within packed beds. For this reason, Potic et al. (2005) suggested that micro-fluidized bed (MFB) screening results would be preferable. Subsequently, MFBs have been used in a variety of applications for process intensification (Wang et al., 2017) and a commercial micro-fluidized bed reaction analyser has become available (Yu et al., 2010; Guo et al., 2016).

An alternative approach for intensifying gas-solid fluidization is to use a swirling gas stream to suspend the particle bed. Either angled static blades (Shu et al., 2000) or arrays of gas nozzles positioned around the bed (Lavrich et al., 2018) impart tangential momentum to a gas stream before it contacts the particles. The vertical velocity component of the gas provides fluidization, while the horizontal velocity component induces a swirling motion. Particle entrainment is minimized because most of the momentum of the gas stream is dissipated at the base of the bed in the radial and tangential directions. This energy dissipation also produces a high amount of turbulence, which thins the boundary layers around the particles leading to intensified heat and mass transfer rates (Groszek and Laughlin, 2015).

In addition to this improved mixing, other advantages include reduced elutriation, lower bed pressure drops and the capability to process solids with a wide range of particle sizes compared to conventional fluidized beds (Lakshmanan and Dodson, 1998). A further benefit is the gas velocity and gas mass flow rate can be decoupled, enabling the particle bed to be suspended using either a low velocity-high mass flow gas stream, or a high velocity-low mass flow gas stream (Lakshmanan and Dodson, 1998).

Several designations for these swirling gas stream devices have been used in the literature: Swirling Fluidized Beds (Sreenivasan & Raghavan, 2002), Rotating Fluidized Beds (Lavrich et al., 2018), Vortexing Fluidized Beds (Qian et al., 2011) and Toroidal Fluidized Beds (Shu et al., 2000). The latter of these, often abbreviated as TORBED® (Registered trade mark of Mortimer Technology Holdings Ltd.), is the main commercial variant of the technology, supplied by Torftech. There are two sub-variants...
termed the TORBED Compact Bed Reactor (CBR) and TORBED Expanded Bed Reactor (EBR), shown in Figure 1.

![Diagram of TORBED variants](image)

*Figure 1 – Sketches of the two TORBED variants | (a) Compact Bed Reactor (CBR), and (b) Expanded Bed Reactor (EBR)*

The CBR functions predominately as described above. The EBR differs to the CBR by the incorporation of additional cyclonic motion of the particles by angling the outer walls near the blades (see Figure 1b). Entrained particles rise through the centre of the reactor before separating outwards towards the edge of the vessel due to centrifugal forces produced by their swirling motion. These particles then return to the base to be re-entrained. In this regard, the EBR is an alternative arrangement to the Circulating Fluidized Bed (CFB). However, the high turbulence of the EBR inevitably leads to increased attrition (Lakshmanan and Dodson, 1998). Generally, the CBR is suited to solids processing, whilst the EBR is suited to gas-stream processing, though for adsorption-based carbon capture, the CBR might be preferable in order to extend the life of the sorbent.
The CBR facilitates intensified CO$_2$ scrubbing in several ways. Principally, it enables a high throughput of the sorbent and reduces adsorption times through the removal of external heat/mass transfer resistances to the particles. From a practical perspective, the compact bed also enables rapid start-up and real time monitoring due to the uniform processing conditions.

Although the TORBED version of the technology has readily been commercially demonstrated, the literature regarding swirling fluidization is still sparse. Different flow regimes have been described for ‘spiral’-type static distributors, created by stacking flat metal blades with a slide overlap at angles usually around 15°. For example, Kaewklum and Kuprianov (2010) describe four main regimes when considering the pressure drop behaviour: (1) fixed-bed (defined by the Ergun equation), (2) partial fluidization, (3) full fluidization with partial swirling (constant pressure drop with increasing gas velocity) and (4) fully swirling fluidization. The latter regime corresponds to an increase in the pressure drop as the gas velocity is increased, which is a result of the fluidizing gas also supporting the centrifugal weight of the bed (Sreenivasan & Raghavan, 2002). Sreenivasan and Raghavan (2002) also describe four different regimes based on observations of the particle bed itself: (1) bubbling (similar to conventional fluidization), (2) wave motion (dune formation at the bed surface), (3) two-layer swirling (swirl velocity gradient across the height of the bed) and (4) stable swirling. Other studies have also proposed models for the minimum swirling fluidization velocity (Sheng et al., 2012; Kaewklum et al., 2009; Shu et al., 2000), pressure drop (Sreenivasan & Raghavan, 2002; Sheng et al., 2012) and transport velocity (Sheng et al., 2012; Shu et al., 2000). However, in all of these studies the blade diameters were between 250–300 mm with an accompanying blade width of ~50 mm. Both of these dimensions exceed the typical boundaries that constitute micro-fluidization behaviour (Wang et al., 2018; do Nascimento et al., 2016; Guo et al., 2009); the concept of ‘mesoscale’ or microscale swirling fluidization therefore remains experimentally untested.

Reducing the swirling bed geometry to scales more attributable with mini- and micro- fluidized beds (order of cm to mm (Wang et al., 2017)) would support sorbent screening experiments with substantially less material. This is especially desirable for screening operations where only small amounts of the solid
sorbent are available, or where the sorbents are expensive. The further improved heat and mass transfer rates compared to the ‘conventional’ fluidized bed might also further improve the scalability of the results by ensuring the rate-limiting step for adsorption is internal diffusion within the sorbent, which will be the same for larger scale operations. One reason for the lack of small-scale swirling fluidization concepts might be the previous manufacturing constraints rather than the lack of applications.

In the present study, we have collaborated with Torftech, the commercial supplier of a swirling fluidized bed variant, the ‘TORBED’, to miniaturise the technology using 3D printing for the purpose of screening sorbents for CO₂ capture applications (McDonough et al., 2018). Here we report the bed formations of activated carbon pellets against a wide range of operating conditions to develop a comprehensive understanding of the fluid dynamics to guide the selection of appropriate operating conditions of the small-scale device (Torftech, 2019).

2 Methodology

2.1 TORBED Design and Additive Manufacturing

Figure 2 compares the CAD model of the mini-TORBED with the final experiment prototype used for imaging the bed formations. All mini-TORBED components were 3D-printed in-house using the SLA technique with a Form2 printer using the proprietary High Temperature Resin (V2). A similar method was used to manufacture micro-fluidized beds (McDonough et al., 2019). CAD models of the freeboard, plenum, central cone and ‘T50’ blade assembly were provided by Torftech, and modified where necessary using Google Sketchup.
The T50 blade ring consisted of 45 equally spaced ~0.9x10 mm slots, and had an o.d. of 50 mm. Due to the angle of the slots, the slot depth (distance between opposing walls) was only ~0.4 mm. Through several test prints, it was found that only one third of the slots were printable when orientating the ring at an angle of 45° using the 100 μm layer height setting, corresponding to these slots being vertically orientated. This required splitting the T50 ring into three segments to avoid channel blockages to ensure uniform gas distribution in the final prototype (Figure 2b). As discussed in the introduction, modification of the blade design enables a decoupling of the air velocity from the volumetric flow rate.

Figure 2 – (a) Exploded CAD model of the small-scale TORBED, (b) 3D-printed T50 blade segments, (c) fully assembled 3D-printed high temperature polymer prototype
In practice however, it was not possible to print narrower slots that would have enabled smaller gas flow rates. And larger slots, although easier to fabricate, would have required a much higher volumetric gas flow rate for fluidization rendering the concept limited in terms of experimental applications (e.g. sorbent screening for CO$_2$ capture).

The plenum region beneath the bed consisted of an inverted cone and an off-centre gas inlet in order to induce a swirling gas stream prior to the static T50 blades. The freeboard was a straight-sided 50 mm diameter, 50 mm length cylinder with a Perspex viewing window bonded to the top. A 12 mm diameter chimney allowed the gas to exit through this window. A small foil cowl then diverted the exhaust away from the high-speed camera lens.

A piston-type O-ring seal was the most effective solution for preventing gas leakage around the bladeset whilst also enabling regular disassembly/assembly of the TORBED for the experiments. Based on the 50 mm diameter of the T50 ring, a standard size 228 silicone O-ring was sufficient (AS 568 A 228). A groove for the O-ring was incorporated into the interior wall of the upper freeboard section as shown in Figure 2a, where the dimensions of the groove were taken from standard design tables. A small amount of high-vacuum silicone grease placed on the O-ring lubricated the joint for easier assembly. In addition, two custom stainless steel collars (held in place with three sets of M4 studding) prevented the TORBED from popping open under pressure. Figure 2c shows the fully assembled prototype used in the experiments; the red O-ring that provided the seal is visible here.

2.2 Activated Carbon Pellets

This study used activated carbon (AC) pellets to investigate the influence of different operating conditions on the swirling fluidization characteristics. Carbonaceous sorbents offer several advantages. They are readily accessible and affordable (Bui et al., 2018), and they can be regenerated at ambient conditions (Peng et al., 2019). Further, carbonaceous adsorbents are thermally and chemically stable, and they can be doped to enhance the adsorption capacity (Olajire et al., 2010).
Figure 3 summarises the particle size distribution of the AC pellets expressed in terms of the mean particle diameter ($D_m$), defined as the diameter of a circle with the same projected area. This was computed in ImageJ using the “Analyze Particles” function using an image of ~600 AC pellets evenly dispersed across a plain white backdrop (Figures S1–S3). The mean pellet size was $D_m = 1.40 \pm 0.37$ mm, and the circularity as viewed in these 2D images was $0.38 \pm 0.06$. The measured tapped bulk density ($\rho_b$) was $0.487 \pm 0.006$ g/cm$^3$, which placed the particles on the boundary between Geldart D (spouting) and Geldart B (sand-like).

![Image of Probability Distribution Function (PDF) of the mean activated carbon pellet diameter]

*orange line shows smoothed experimental data*

2.3 High Speed Camera Imaging and PIV Analysis

2.3.1 High Speed Camera Imaging and Experiment Conditions

Figure 4 summarises the arrangement of the experiment equipment used for observing the effects of the gas volumetric flow rate, temperature and humidity on the AC bed formations. Compressed air was first regulated to 3.0 bar using a pneumatic miniature regulator (model: P3A-EA12BEBNP) before the volumetric flow rate was set using a mass flow controller (Bronkhorst, model: EL-FLOW F-202BB). A pair of 3-way valves then allowed the air to either be diverted to a blank stainless steel tube or through
a Nafion shell-in-tube membrane for humidification (Perma Pure, model: MH-110-12S-4). Deionised water saturated the shell side of this membrane, and the fluid head (controlled using the height of a liquid reservoir) enabled partial control over the amount of humidification of the air stream. An in-line digital sensor (Sensirion EK-H4 Evaluation Kit) positioned after the second 3-way valve recorded the absolute and relative humidity. The temperature of the humidified/un-humidified air was adjusted using a 200 W in-line gas heater (Omega, model: AHP-3742) with the control thermocouple positioned directly at the heater outlet. Another thermocouple at the TORBED inlet measured the final reported gas temperatures. The TORBED connected to the outlet of this tubing arrangement using a ¼ in. NPT fitting integrated into the inlet of the plenum (see Figure 2).

![Diagram of Experiment Flowsheet](image)

**Figure 4 – Experiment flowsheet**

A summary of the experiment conditions is included in Table 1, whilst the supplementary materials include a full summary of the specific conditions and corresponding results. Preliminary experiments identified the lowest volumetric air flow rate as around 27 L/min for the onset of swirling fluidization at the lower loadings. It was not possible to achieve uniformly packed behaviour below bed loadings of 800 mg due to excessive gas bypassing of the pellets. The upper air flow rate used (44 L/min) was mainly based on practicality; with high gas flow rates it would be difficult to observe meaningful adsorption kinetics because of the reduced contact/residence time between the gas and solid material.
The absolute humidity of the raw air supply ranged from 0.2–0.4 g/m³, which could be boosted up to 6 g/m³ by diverting the air through a shell-and-tube membrane saturated with deionised water on the shell side. The maximum humidity was limited by the short residence times imposed on the humidifier. The epoxy used to bond the Perspex viewing window to the freeboard limited the maximum temperature to around 120 °C.

**Table 1 – Summary of experiment conditions and accuracy**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Conditions</th>
<th>Accuracy</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bed Material</td>
<td>Activated Carbon (AC)</td>
<td>N/A</td>
</tr>
<tr>
<td>AC Pellet Size (mm)</td>
<td>1.40</td>
<td>± 0.37 mm</td>
</tr>
<tr>
<td>Bed Loading (mg)</td>
<td>800 – 3500</td>
<td>± 1 mg</td>
</tr>
<tr>
<td>Air Flow Rate (L/min)</td>
<td>27 – 44</td>
<td>± 1%</td>
</tr>
<tr>
<td>Air Temperature (°C)</td>
<td>20 – 120</td>
<td>± 0.1°C @ 20°C (Ambient)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>± 6.5°C @ 65°C</td>
</tr>
<tr>
<td></td>
<td></td>
<td>± 5.7°C @ 120°C</td>
</tr>
<tr>
<td>Absolute Humidity (g/m³)</td>
<td>0.3 – 6</td>
<td></td>
</tr>
<tr>
<td>Relative Humidity, RH (%)</td>
<td>0.5 – 35.0</td>
<td>± 3% of RH</td>
</tr>
</tbody>
</table>

2.3.2 PIV Analysis

A high-speed CCD camera (Photron FastCam) positioned directly above the TORBED (Figure 4) recorded the different bed formations. Images were captured at 1000 fps for ~5.5 s (RAM limit of the camera) through Photron FASTCAM Viewer (PFV) software, and then downloaded to an external HD for later transfer to a larger storage server. The shallowness of the bed (a design advantage of the CBR variant) meant that Particle Image Velocimetry (PIV) analysis could be directly applied to the activated carbon pellets without the requirement for tracer particles or a laser light sheet. A 150 W halogen lamp was used to compensate for the low light sensitivity using the fast shutter speed.

The raw images captured during the experiments were first imported into ImageJ and cropped, then the central portion of the images were deleted so that only the annular bed was present in the images (Figure 5a). This reduced the file size significantly. These modified images were subsequently analysed in PIVLab v2.0 in Matlab (Thielicke and Stamhuis, 2014). A mask was applied so that only the annular bed was included in the analysis (Figure 5b) and the images were filtered using the “Enable CLAHE” (window size: 20 px), “Enable highpass” (kernel size: 15 px) and “Enable intensity capping” options
A 2-pass FFT deformation was used for all experiment conditions, with an initial integration area of 64×64 px and step size of 32 px, and a second pass integration area of 32×32 px with step size of 16 px. Figure 5d shows an example of the resulting velocity vectors produced. Any erroneous velocity vectors were detected through cross-validation and replaced via interpolation (Figure 5e). Calibration of the velocity vectors was applied by selecting the outer diameter of the bed as a reference distance (Figure 5e; 50 mm) and specifying the time difference based on the image capture rate (1 ms). Figure 5f shows an example velocity magnitude field obtained after calibration. These velocity fields could be exported as text files for further processing (e.g. data averaging).

Figure 5 – Processing steps during PIV analysis | (a) cropped raw image with mask applied to central cone, (b) mask applied in PIVLab to exclude regions from analysis, (c) image filtering in PIVLab, (d) velocity vector field obtained from PIV analysis, (e) erroneous velocity vectors removed and replaced with interpolated values (orange); red line shows reference distance for calibration, and (f) velocity magnitude field computed after calibration with reference distance and high speed camera framerate.
The ~5500 captured images were divided into ~2750 consecutive image pairs, producing ~2750 velocity vector fields for each of the experiment conditions studied. The images were recorded at a resolution of 1280×1084 px, which corresponded to 11 px/mm (or \( \delta = 90 \mu \text{m/px} \)). The uncertainty with the analysis is usually taken as 0.1 px (Afolabi, 2012). Therefore, for a time step size of \( dt = 1 \text{ ms} \), the systematic uncertainty in both the x- and y-velocity components was \( e_v = 0.1 \delta / dt = 0.009 \text{ m/s} \) (around 3% of the maximum reported velocity). Section 3.3 provides further analysis based on these velocity fields.

3 Results and Discussions

3.1 Bed Formations

The miniature TORBED fabricated for this project operates in the compact mode. Here, the activated carbon pellets formed a shallow bed near the blade distributor, meaning bubbling and slugging structures were not observable because there was insufficient room for their development. This is in contrast to the observations of deeper particle beds, where faster swirling layers occur at the base of the bed with slower swirling layers at the top (Sreenivasan and Raghavan, 2002).

The meso-TORBED produced several distinguishable swirling bed states depending on the pellet loading, and gas flow rate/temperature/humidity:

- ‘Un-swirled’. Partial fluidization of the pellets occurred, but the gas velocity was insufficient to induce any additional noticeable bulk swirling motion. This was a consequence of either over-loading or under-airing of the bed.
- \textit{Collapsed}. The collapsed state consisted of two sub-states where part of the annular bed remained unfluidized. This collapsed region either remained fixed in one location on the bladeset (static collapsed), or moved around the distributor in the opposite direction to the swirling particle flow (dune collapsed).
- \textit{Maldistributed}. Here, the entire particle bed was swirling, but a slower aggregated region formed that moved around the bed slower than the average bulk motion. Here, the particles...
captured insufficient energy from the gas stream, resulting in gaps in the pellets that provided a path of least resistance to the gas flow, creating a negative feedback loop that promoted further maldistribution.

- **Uniformly packed.** This is the desired bed state where the blade distributor remained fully covered so that no by-passing of the gas occurred, and the particles all moved around the bed with a uniform velocity with no aggregated or maldistributed regions.

- **Entrained.** This is an over-aired condition where a higher gas velocity resulted in a higher centrifugal force, forcing the particles toward the outer wall. Although the particles maintained an apparent uniform velocity in the tangential direction, part of the distributor remained uncovered by the particles. Thus, similar to the maldistributed state, this would be problematic for CO$_2$ adsorption because CO$_2$ rich gas could bypass the adsorbent.

The supplementary data online includes representative animated videos of each of the six bed states, whilst Table 2 presents a comprehensive flow map summarising the conditions that resulted in each bed formation.

Table 2 shows that the meso-TORBED was either under-loaded (maldistributed state) or ‘over-aired’ (entrained state) for particle bed weights of 800 mg and 1000 mg. Under-loading reduces the absorption of the kinetic energy of the fluidizing gas (further discussed in Section 3.3), whilst ‘over-airing’ increases the centrifugal force resulting in the particles swirling mainly at the periphery of the blades. Both lead to gas bypassing around the particles. Thus, the lower design point corresponds to a loading of 1200 mg, which produced the desirable uniform packing behaviour across most of the operating conditions. Although, this loading could include a mixture of inert and active materials for screening purposes. This concept has been similarly exploited in micro-fluidization reaction analyzers to reduce the materials requirements to as low as 10–50 mg (Yu et al., 2010; Guo et al., 2016). However, the materials requirement in the TORBED will also depend on the process and sensor sensitivities because of the higher gas flow rates required compared to conventional fluidized beds.
At the lowest gas flow rate considered (27 L/min), bed loadings exceeding 1400 mg would be unsuitable for sorbent screening due to the reduced mixing because of either collapsed or non-swirling behaviour. However, for intermediate and high gas flow rates (35.5 L/min and 44 L/min respectively), the ‘T50’ blade configuration appears to be suitable for a wide range of bed loadings. Based on the qualitative flow map in Table 2, it also appears that the gas temperature and humidity have negligible influence over the bed formations. Section 3.3 provides more nuanced discussions of the swirling fluidization quality.

Table 2 – Bed formations observed qualitatively at each operating condition | (N) un-swirled, (C_d) collapsed (dune), (C_s) collapsed (static), (M) maldistributed, (U) uniformly packed, (E) entrained

<table>
<thead>
<tr>
<th>Operating Conditions</th>
<th>Bed Loading (mg)</th>
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<tbody>
<tr>
<td>L/min</td>
<td>°C</td>
</tr>
<tr>
<td>27</td>
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<td>27</td>
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<td>27</td>
<td>120</td>
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<td>27</td>
<td>120</td>
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<td>35.5</td>
<td>20</td>
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<td>35.5</td>
<td>70</td>
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No obvious degradation of the activated carbon pellets was observed during the experiments (i.e. no chipping or splitting). However, the 3D printed polymer did darken slightly over time, suggesting a minor amount of surface disintegration of the particles that could not be measured quantitatively. Here surface disintegration refers to the loss of small fragments along the natural grain boundaries (Ray et al., 1987). Degradation in fluidized beds is driven by two mechanisms: surface wear caused by shear between bubble and particle phases, and impact damage due to either particle-particle interactions or particle-wall collisions (Scala et al., 2007). In the compact-mode TORBED, there is insufficient height for the development of bubbling structures. Additionally, in the uniformly packed regime, the particles
are swirled uniformly meaning there is little slip between individual particles. *I.e.* the shear in the radial direction is low (see Figure 6m-p). Consequently, we infer that it is only the particle-wall collisions that would be responsible for any particle degradation effects.

### 3.2 Velocity Vector Fields

Figure 6 shows the velocity vector fields and velocity magnitude contours of the five swirling bed states; each row corresponds to four stages of one revolution of the particles around the annulus. Additionally, each image has been time-averaged from at least five different sets of velocity data from the same point during each revolution of the particles to reduce the influence of random noise. For reference, the figure caption includes the specific experiment conditions that produced each image set.
Figure 6 – Example velocity vector fields observed in the 3D printed TORBED | the colorbars refer to
the velocity magnitude in m/s | (a-d) Collapsed Static: bed loading = 1800 mg, flow rate = 27 L/min,
humidity = 0.89 g/m$^3$, temperature = 22.6°C | (e-h) Collapsed Dune: bed loading = 2000 mg, flow
rate = 27 L/min, humidity = 0.27 g/m$^3$, temperature = 23.4°C | (i-l) Maldistributed: bed loading =
800 mg, flow rate = 27 L/min, humidity = 3.2 g/m$^3$, temperature = 122°C | (m-p) Uniformly Packed:
bed loading = 2000 mg, flow rate = 35.5 L/min, humidity = 0.18 g/m$^3$, temperature = 65.2°C | (q-t)
Entrained: bed loading = 800 mg, flow rate = 44 L/min, humidity = 0.3 g/m$^3$, temperature = 113°C
Figures 6a-d show velocity vector fields from the static collapsed state where an ‘unswirled’ region developed at a fixed point on the blade distributor, resulting in a fixed near zero-velocity zone. This is a result of over-loading or ‘under-airing’ of the bed, where insufficient kinetic energy was absorbed from the gas stream. The other collapsed state that produced a moving ‘de-swirled’ dune is shown in Figures 6e-h. Here the collapsed region migrated in the opposite direction to the swirling particle direction (noted by the orange arrows). Here, the swirling zone deposited particles at the leading edge of the collapsed zone, whilst the particles at the opposite side of the collapsed zone became re-entrained. Thus, the collapsed zone would decay at one side and grow at the other, resulting in the formation of the moving dune. This bed state only occurred in the room temperature experiments at the lowest gas flow rate for bed loadings between 1600–2200 mg.

The maldistributed state occurred at the lowest bed loadings and shared some similarities with the collapsed states, particularly the collapsed state with moving dune. Although the entire bed was swirling, there was incomplete coverage of the blades by the particles. Two distinct regions formed: a slower-moving agglomerated region and a faster-moving sparsely populated region. In the velocity vector fields in Figures 6i-l, the higher velocity region actually corresponds to the denser agglomerated zone, because the sparser region did not have a sufficient particle density to enable robust PIV analysis, resulting in an apparent low/zero-velocity. However, as a result these figures do indicate the relative size of the agglomerated region compared to the full annulus, and show that this agglomerated region moves in the same direction as the velocity vectors.

The velocity vector fields for the desired uniformly packed state (Figures 6m-p) are clearly distinct from the other swirling states. The velocity vectors indicate a highly regular circular path of the particles with near uniform velocity magnitude in the tangential direction and at each averaged time point during a single revolution of the particles. The velocity also exhibits negligible radial variation, indicating that there was minimal slippage/shear between the particles along with full blade coverage by the particles.
Finally, Figures 6q-t shows the velocity vectors for the entrained state. ‘Over-airing’ of the bed resulted in an enhanced centrifugal force, reducing the coverage of the blades at the inner periphery. The apparent strong radial velocity profile is a result of reduced particle density at the inner periphery, making it difficult to calculate the velocities using the PIV method. Thus, similar to the maldistributed state (Figures 6i-l), the velocity magnitude also provides an indication about the particle distribution in the entrained state. Here, the particle velocities are tangentially uniform and exhibit minimal variation over time (indicated by the similar velocity fields at each point from the revolution of the particles).

3.3 Swirling Quality and Particle Velocity

Further analysis of the velocity vector fields provided additional quantitative understanding of the swirling fluidization characteristics as a function of the operating conditions. The two approaches considered were swirling quality (velocity uniformity) and average particle velocity.

Regarding the swirling intensity, Chyang et al. (1997) proposed the vortex number \( V_{or} \) for swirling fluidized beds shown in equation 1. Here, \( v_\theta \) is the tangential velocity, \( v_z \) is the axial velocity, \( r \) is the radius of the fluidized bed and \( A_0 \) is the cross-sectional area of the freeboard region. This equation is similar to the swirl number typically used for characterising swirl flow in combustors (Eldrainy et al., 2009), and compares the ratio of the mean tangential to mean axial velocities of the gas stream per unit flow area.

\[
V_{or} = \frac{\int \int (v_\theta / v_z) r \, dr \, d\theta}{A_0}
\]

In the CBR configuration of the TORBED, the particles exhibit minimal vertical motion, meaning the majority of their motion is in the radial and tangential directions. Further, vortexing fluidized beds can independently adjust the tangential and axial gas velocities because the swirling motion is controlled via a secondary gas flow (usually using angled gas nozzles placed around the periphery of the particle bed). In the meso-TORBED however, the static blade geometry resulted in a fixed \( v_\theta / v_z \) ratio. Therefore, applying the vortex number metric in equation 1 provided no additional intuition beyond the
use of an average particle velocity. Instead, a simpler uniformity index (equation 2) was sufficient for characterising the swirling homogeneity (quality). Here, $\sigma$ is the standard deviation of the velocity magnitude of the particles and $\bar{v}$ is the mean particle velocity. Based on this definition, smaller values of $I_u$ indicate worse uniformity, whilst perfect uniformity corresponds to $I_u = 1$ (where $\sigma = 0$). The standard deviation was calculated using equation 3, which used the absolute values of the velocities to account for the positive and negative velocity components at various points around the annulus.

$$I_u = 10^{-\left(\frac{\sigma}{\bar{v}}\right)}$$  \hspace{1cm} (2)

$$\sigma = \sqrt{\frac{\sum(|v| - |\bar{v}|)^2}{N}}$$  \hspace{1cm} (3)

The uniformity index was calculated across the full annular cross-section of the TORBED for each of the ~2750 sets of velocity fields per condition, then time-averaged to produce a single indicator of swirling quality. Table 3 summarises these time-averaged uniformity indices for each combination of operating conditions. The blacked-out areas corresponded to the ‘un-swirled’ state. Additionally, the operating condition values shown in Table 3 correspond mainly to ‘low’, ‘medium’ and ‘high’ states (which varied slightly between each bed loading). The exact conditions are included in the supplementary materials document.
Table 3 – Time-averaged uniformity index obtained at every operating condition | the blacked-out areas correspond to the un-swirled state where the uniformity index had no useful meaning

<table>
<thead>
<tr>
<th>Operating Conditions</th>
<th>Bed Loading (mg)</th>
</tr>
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<tbody>
<tr>
<td></td>
<td>L/min</td>
</tr>
<tr>
<td>27</td>
<td>20</td>
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<tr>
<td>27</td>
<td>3</td>
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<td>44</td>
<td>120</td>
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</table>

Table 3 illustrates that the uniformity index is a good predictor of the swirling quality because the largest values ($I_u > 0.3$) align with the qualitatively identified uniformly packed regime shown in Table 2. However, the advantage of the $I_u$ metric is more subtle nuances within the uniformly packed regime emerge. Clearly, the maximum uniformity at a given bed loading (above 1000 mg) was a function of the gas volumetric flow rate. For example, the best volumetric flow rate for the 1400 mg, 2000 mg, and 3000 mg loadings were 27 L/min, 35.5 L/min and 44 L/min respectively. Thus, for each bed loading there is an optimal volumetric flow rate. Too low, and the bed becomes under-aired leading to the collapsed and un-swirled states, and too high, the bed becomes over-aired leading to the entrained state. Equation 4 predicts the optimal volumetric flow rate ($Q$) based on the bed loading ($m_s$), within the experimental range explored in this study. The humidity and temperature had negligible effect on the uniformity index.

$$Q = 0.6762 m_s^{0.5163}$$

Figure 7 shows how the average velocity magnitudes of the AC pellets changed as a function of the bed loading at each gas temperature. For the majority of the experiments, the gas humidity had minimal effects.
influence over the behaviour of the bed formations, so both sets of humidity were plotted together. Again, the temperature had minimal effect on the velocity, suggesting that the gas density/viscosity should also negligibly influence the swirling fluidization characteristics.

Figure 7 – Average particle velocities as a function of bed loading at different air volumetric flow rates: (a) 27 L/min, (b) 35.5 L/min, and (c) 44 L/min.
At the lowest air volumetric flow rate (Figure 7a), the average particle velocity initially rose with increased bed loading (up to 1400 mg), before decreasing. This likely occurred because not all of the kinetic energy of the gas was utilised to swirl the bed at the lowest loadings. Here the bed was under-loaded (or over-aired) resulting in an unstable condition whereby maldistribution of the particles produced a path of least resistance for the gas, that then further promoted maldistribution. With increased bed loading, there was reduced capacity for gas bypassing to occur. Thus, the initial rising particle velocities as the bed loading increased caused the bed to absorb a greater proportion of the kinetic energy of the gas. Further increases in the loading then reduced the average velocity because the bed became overloaded, leading to the collapsed and un-swirled states.

Interestingly, for the intermediate air flow rate (Figure 7b) the average particle velocities remained approximately constant as the loading increased, whilst the average velocities increased slightly at the highest air flow rate (Figure 7c). Based on the velocity, there was not a definitive turning point where overloading became apparent, though the velocity at the 3500 mg loading was slightly lower. These results reflect the qualitative observations of Table 2; i.e. these operating conditions all produced the desirable uniformly packed state.

Therefore, for sorbent screening operations involving Geldart D pellets, it is the recommendation of this study to use bed loadings between 3000–3500 mg to maximise the total adsorption capacity. The corresponding ‘optimal’ gas flow rate is perhaps more debatable. Although the highest volumetric flow rate produced a slightly higher velocity with increased uniformity at these higher loadings, the contact time would also be lower.

3.4 Minimum Swirling Velocity

Shu et al. (2000) proposed equation 5 for the prediction of the minimum fluidization velocity for the TORBED, which modifies the minimum fluidization velocity for a conventional fluidized bed, $u_{mf}$, with the blade angle, $\theta_i$ (defined from the horizontal). Here, $u_{mf}$ uses the form originally presented by
Wen and Yu (1966) (equation 6), where $Ar$ is the Archimedes number (equation 7) and $C_1 = 33.7/C_2 = 0.0408$ (Shu et al., 2000).

$$u_{mf,r} = \frac{u_{mf}}{\sin \theta_i}$$  \hspace{1cm} (5)

$$Re_{mf} = \frac{\rho g u_{mf} d_p}{\mu_g} = \sqrt{\left[C_1^2 + C_2 Ar\right] - C_1}$$  \hspace{1cm} (6)

$$Ar = \frac{\rho_g d_p^2 (\rho_b - \rho_g) g}{\mu_g^2}$$  \hspace{1cm} (7)

The cross-sectional flow area of the bed at the base of the cone was 1,257 mm$^2$, corresponding to superficial air velocities of 0.36, 0.47, and 0.58 m/s for volumetric flow rates of 27, 35.5, and 44 L/min respectively. Equations 5–7 predict $u_{mf,r} = 0.43$ m/s for the AC pellets based on a bulk density of 487 g/cm$^3$ and the blade geometry shown in Figure 2. The temperature had minimal effect on $u_{mf,r}$. Thus, equations 5–7 imply that at a gas flow rate of 27 L/min, the bed was slightly below the minimum fluidization point, which agrees with the observations of Table 2 that showed uniform packing for gas flows required $Q \geq 35.5$ L/min. Consequently, equations 5–7 appear to be valid for the meso-TORBED.

4 Conclusions

The toroidal fluidized bed (TORBED) uses static angled blades to induce an additional tangential velocity component to a fluidized bed in order to realise intensified heat and mass transfer rates. In this study, we successfully miniaturised the technology to a 50 mm diameter scale through the application of the stereolithography (SLA) 3D-printing technique. Here we stress that it would not be possible to fabricate the blade distributor geometry using conventional manufacturing methods in a practical or cost-effective timeframe at this scale.

To understand the swirling fluidization characteristics, we performed a comprehensive parametric study by varying the bed loading, gas volumetric flow rate, gas temperature and gas humidity across a total of 252 combinations. Six different bed states were subsequently categorised: un-swirled, collapsed static, collapsed dune, maldistributed, uniformly packed and entrained. The desirable uniformly packed
state occurred at bed loadings of $m_s \geq 1200$ mg and gas flow rates of $Q \geq 35.5$ L/min. Humidity and air temperature had minimal influence over the flow patterns. For sorbent screening experiments, we would recommend the higher bed loadings to maximise the potential adsorption capacity. Although, this loading could include a combination of inerts and active sorbent to further minimise the material requirements.

Although the higher gas volumetric flow rate produced desirable uniform packing, the reduced gas residence time might make the observation of meaningful kinetic data more difficult. Continued development of the SLA printing technology will eventually enable the fabrication of smaller slot sizes, which will eventually allow smaller gas volumetric flow rates to produce the desired uniform packed state.

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Potential Bias
Torftech are the only commercial vendor of the TORBED technology.

Nomenclature

$A_0$    Cross-sectional area of the freeboard region (m$^2$)
$Ar$    Archimedes number ($= \rho_g d_p^3 (\rho_b - \rho_g) g / \mu_g^2$)
$d_p$    Particle diameter (m)
$g$    Gravitational acceleration (9.81 m/s$^2$)
$I_u$    Uniformity index ($= 10^{-(\sigma/\mu)}$)
\( m_s \)  Bed loading (g)  
\( N \)  Number of data points  
\( Q \)  Gas volumetric flow rate (L/min)  
\( r \)  Radial position (m)  
\( Re_{mf} \)  Minimum fluidization Reynolds number (= \( \rho g u_{mf} d_p / \mu g \))  
\( u_{mf} \)  Minimum fluidization velocity in a conventional fluidized bed (m/s)  
\( u_{mf,T} \)  Minimum fluidization velocity in the TORBED (m/s)  
\( v \)  Velocity magnitude (m/s)  
\( v_z \)  Axial velocity (m/s)  
\( v_\theta \)  Tangential velocity (m/s)  
\( V_{or} \)  Vortexing number (= \( \int \int (v_\theta / v_z) r \ dr \ d\theta / A_0 \))

**Greek Letters**

\( \theta \)  Tangential position (rad)  
\( \theta_i \)  Blade angle (rad)  
\( \mu_g \)  Gas viscosity (Pa.s)  
\( \rho_b \)  Bulk density of the particles (kg/m\(^3\))  
\( \rho_g \)  Gas density (kg/m\(^3\))  
\( \sigma \)  Standard deviation of velocity (m/s)

**References**


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