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Effect of distributor design on particle distribution in a binary fluidised bed

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Abstract

The influence of distributor design, including orifice size, number and positioning, on the behaviour of a gas-fluidised granular bed has been studied experimentally. The use of Positron Emission Particle Tracking (PEPT) to non-invasively probe the system's three-dimensional dynamics has allowed the extent of (de)fluidisation and the homogeneity of particle distribution to be quantitatively characterised. While increasing the number of orifices demonstrates a clear, positive influence on both fluidisation quality and homogeneity, the data obtained suggest that the orifice size and positioning play a more complex role.

Keywords: Fluidisation; fluidised bed; PEPT; Positron Emission Particle Tracking; Segregation; Mixing

1. Introduction

Fluidised beds are widely used in a variety of industrial processes spanning numerous sectors. Examples include combustion and gasification, powder granulation, polymerisation, spray coating and atomic layer deposition [1–3]. For many of these processes, their effectiveness strongly depends on the ability of the fluidised-bed systems to mix, recirculate and uniformly distribute materials [4–6]. While the mixing and flow behaviours can, to an extent, be inferred from conventional measures such as pressure drop measurements [7], they yield typically only data regarding the broad, bulk behaviours of the bed, and cannot provide reliable information regarding the detailed flow patterns and the spatial distribution of particles within a given system. However, with the advent of powerful three-dimensional imaging techniques such as positron emission particle tracking (PEPT), we are beginning to gain deeper insight into these processes.

It is well known that the dynamics of fluidised-bed reactors can be influenced by the properties of the distributor through which gas is injected into the system. Distributor design directly influences the flow characteristics of, and hence mixing and heat and mass transfer achieved within, a fluidised bed [8] – properties which influence the effectiveness of many applications, including catalysis, combustion, gasification and plastic waste pyrolysis, a process of particular contemporary importance.

Li *et al.* [6] have already provided valuable insight into the effects of varying distributor design. Notably, the authors demonstrated the existence of four distinct flow structures (depicted schematically in Fig. 1) which may be achieved within a fluidised bed, noting that the most efficient mixing was produced by systems possessing quadruple rolls (Pattern D) and the least efficient in the single-roll case (Pattern A).

The work of Li *et al.* [6], however, focused exclusively on sintered metal (SM) distributors and monodisperse beds. Other groups such as Wormsbecker *et al.* [9, 10] or Pavia *et al.* [11] focused on different designs such as perforated, punched or porous plates and have shown different results in drying time and pressure drop. Wormsbecker *et al.* [9, 10] also showed that the hydrodynamics of perforated and punched plates are typically better suited to mixing than a porous distributor due to the lack of momentum induced to wet granular media. Previous studies have also looked at the effect of parameters such as fluidising velocity and feed rate on temperature

distribution for two different multiorifice distributors. [12] The increasing amount of computing power through the last years has facilitated the increased adoption of computational fluid dynamics (CFD) simulations in granular research generally, and in particular in the modelling of fluidised beds [13–16]. Regarding distributor plates, the main focus of the simulations lies on hydrodynamics [17], outlet velocity distribution [18] and bubble forming behaviour [19] for different distributor designs and hole arrangements. In short, while there has been some research into the mixing of binary systems [20], and into the effect of distributor design on the dynamics of *monodisperse* systems, there have, to date, been no studies – either CFD or experimental – explicitly investigating the link between distributor design and the mixing dynamics of binary systems.

In the present study, utilising PEPT imaging, we systematically examine the influence of several design properties on flow dynamics and the spatial distribution of particles within a binary bed, gaining valuable new insight into how distributor design may be used to optimise the behaviour of fluidised granular mixtures.

2. Experimental Details

2.1. Material and Methods

Our experiments utilise a gas-fluidised bed of cylindrical geometry with inner diameter $D = 70$ mm. For all experiments, the system is filled with 1 kg of a binary mixture containing 90 wt% (89 vol%) sand, with intrinsic density 2800 kg/m^3 and size range $0.28 < d < 0.34 \text{ mm}$, and 10 wt% (11 vol%) glass ball-bearings, with intrinsic density 2500 kg/m^3 and size $d = 1 \pm 0.01 \text{ mm}$, giving a static bed height of 140 mm ($H_0 = 2D$). The sand and glass particles belong, respectively, to Geldart groups B and D [21]. Room temperature air ($16 \pm 0.5^\circ\text{C}$) was injected into the plenum, the flow rate being controlled using calibrated rotameters. Pressure taps at the top of the bed and in the plenum were attached to a U-tube manometer allowing the pressure drop across the system to be recorded.

Experiments were performed using a variety of perforated-plate distributors (see Table 1 and Fig. 2). A total of 15 plates were used, grouped into three distinct sets such that the number, size, and spatial distribution of the gas-outlet holes may be individually, systematically varied. In all cases, the plates were constructed from 1.5 mm thick aluminium and held between flanges,

using rubber gaskets to prevent air loss. The chosen hole diameters and open areas of some plates tested are comparatively large, such that we may explore as wide a range of parameter space as feasible. In particular, the size of the hole diameters used for plates S1-S5 was deliberately chosen to give a pressure drop at the lower end of Geldart's criterion for 'good distribution' [22], so as to amplify the effects of orifice position.

Using standard pressure-drop measurements, the U_{mf} for the glass-sand mixture was found to be 0.09 ± 0.004 m/s (see Fig. 3). Checks were performed to ensure that the measured U_{mf} was consistent for all distributors used. In all experiments, a constant fluidisation velocity $U = 2.4U_{mf} = 0.23$ m/s was utilised in order to ensure reliable bubbling fluidisation. The parameters used were deliberately chosen so as to align with those used in the prior study of Li *et al.* [6]. In order to assure reproducibility, before each run the system was fluidised with a high gas flow rate ($10 U_{mf}$) so as to ensure, as far as possible, an initially well-mixed and fully-fluidised system. It should be noted, however, that preliminary results obtained from systems initiated under different starting conditions (glass on top, sand on top, well-mixed) suggest that, for the range of experimental parameters explored, the steady state dynamics are agnostic to the system's initial conditions.

2.2. Positron Emission Particle Tracking

Data were extracted from our system using positron emission particle tracking (PEPT), a radiation imaging technique capable of tracking the three-dimensional motion of a particle – even within the interior of dense, opaque systems – with sub-millimetre accuracy and millisecond-scale temporal resolution [23, 24]. The ability to image the interiors of opaque systems, combined with PEPT's high temporal and spatial resolution, allow us to study the flow dynamics of our system in a manner not possible using conventional imaging techniques.

In order to perform PEPT, a single particle – physically identical to others of the same species – is taken from the system and 'labelled' with a β^+ -emitting radioisotope, in this case Fluorine-18. Positrons emitted by the ^{18}F atoms rapidly annihilate with electrons in the tracer medium, causing the emission of two high-energy (511 keV) gamma photons possessing opposite trajectories. Once labelled, the particle can then be returned to the system of interest, which is then

placed between the heads of a positron camera (see Fig. 4(a)). If both members of a γ -ray pair emitted due to an annihilation event are detected by the camera's two heads, their linear trajectory can be reconstructed (see Fig. 4(b)). The position of the tracer particle can then be triangulated using multiple reconstructed lines (see Fig. 4(c)). For adequately active tracers, the particle can be successfully located many times per second, allowing its full, three-dimensional trajectory to be recorded.

For monodisperse systems which exhibit ergodic behaviour², the time-averaged behaviour of a single tracer particle provides ensemble information – for example whole-field density distributions, velocity fields, and granular temperature fields [25–27], in addition to many other important physical quantities [28–30]. For binary or polydisperse systems, one may utilise a tracer corresponding to each distinct ‘species’ of particle present within the system. Multiple particles may either be tracked within the system simultaneously [31], or the entire experiment repeated N times for each of the N different tracer species used [28]. According to the principle of ergodicity, the time-averaged results corresponding to each case are expected to be identical, though the latter method (which we adopt here), while significantly more time-consuming, typically allows for higher resolution data. The data acquired for the different species may then be summed to provide whole-field information, or used to provide important information regarding the behaviours of an individual component of the system. For example, the time-averaged spatial distribution of a single species throughout a binary system can provide insight into the quality of mixing within said system. As is clear from the above, PEPT is typically used to study the steady state behaviour of systems using a single particle (though multiple particle tracking methods are increasingly adopted as these techniques become more refined). As such, unlike computational techniques such as CFD, PEPT cannot offer detailed information regarding the transient dynamics of a system, or the collisional forces between particles, for instance. However, unlike CFD, PEPT – being an experimental technique – requires no calibration or validation, and is able to record the dynamics of systems for periods significantly longer than can be realistically simulated using CFD under the constraints of current computer hardware. In the present work, our primary interest lies in the steady-state spatial occupancy distributions and velocity vector fields corresponding to our systems. The manner in which such data may be extracted from PEPT data are described in the following sections.

² The interested reader may find a proof for the ergodicity of a fluidised granular system in our reference [25].

2.2.1. Occupancy Distributions

In order to obtain the spatial distribution of occupancies from PEPT data, one must first subdivide the experimental volume into a series of equally-sized ‘cells’. This may be done in one, two, or three dimensions to produce, respectively, one-dimensional profiles (see Fig. 8), two-dimensional distributions (see Fig. 5) or even 3D distributions – though, for an axially-symmetric system such as that studied here this is not necessary. The total time spent by the tracer in each cell is recorded and divided by the total length of the experimental run performed.

The construction of such a distribution allows us to gain insight into the spatial distribution of particles within our system as, for an ergodic system such as those explored here, the time-averaged occupancy distribution of a tracer particle is directly proportional to the density distribution of all particles of the same species [25].

2.2.2. Velocity Vector Fields

The pseudo-instantaneous velocity, $\vec{v}_i$, of a tracer particle at its i^{th} location, $\vec{r}_i$, can be determined as:

$$\vec{v}_i = \frac{\vec{r}_{i+j} - \vec{r}_{i-j}}{t_{i+j} - t_{i-j}} \quad (1)$$

where the ideal value of j to use is dependent on the data rate – for a lower data rate, a smaller j must be used in order to ensure that an accurate pseudo-instantaneous velocity is achieved. In the present work, the values of j implemented varied from $j = 1$ at a data rate of 30 Hz to $j = 4$ at a data rate of 100 Hz. It should be noted, however, that in the present work our interest does not lie in the accurate determination of particles’ individual velocity magnitudes, rather only their time-averaged directions, which are significantly less sensitive to the precise value of j used. Sensitivity analyses show that the forms of the velocity vector fields shown in Figs. 5 and 7 are insensitive to small changes in the precise value of j chosen.

In order to determine a spatial distribution of velocities, the experimental volume must – as above – be subdivided into a series of cells. For each cell within the system, the relevant mean velocity may be determined by taking the average of all values of $\vec{v}_i$ corresponding to locations, $\vec{r}_i$, falling within said cell. For a two-dimensional distribution, these velocities may be used to

create velocity vector fields such as those presented in Fig. 7. Note that all velocity vector fields presented in this paper have been normalised to give an equal, unit length, such that their orientation may be easily interpreted for both low and high velocity magnitudes.

2.2.3. Further Information

Although not necessary to the comprehension of this paper, the interested reader may find further details regarding the PEPT technique, its underlying algorithms, and its application to monodisperse and binary granular systems in references [23–30].

3. Results

3.1. Dead zones

One of the simplest indications that a distributor is not adequately fluidising a bed (and hence not achieving a homogeneous distribution of particles) is the appearance of ‘dead zones’ or ‘stagnant solids’ [32, 33] near the base of the system. PEPT is an effective method of detecting such dead zones, allowing us to directly and accurately determine their three-dimensional extent in a manner not possible with conventional, optical imaging techniques.

As a particle cannot – by definition – either leave or enter a dead zone, such zones appear as blank regions in the three-dimensional map of our system produced using PEPT (see, as an example, Fig. 5). As such, the total ‘dead volume’, V_d , for a system can be calculated by summing the volumes of all unoccupied cells in the vicinity of the distributor, providing a useful scalar metric by which to measure the effectiveness of a given distributor in fluidising our beds. An unoccupied cell is defined simply as any voxel falling within the experimental volume which is never visited by the tracer particle.

In Fig. 6, we show the variation in dead volume as a function of the three main parameters explored here: distributor hole number, hole size, and hole spacing (and hence position). Beginning with the simplest set of observations, we see in the leftmost panel of Fig. 6 a monotonic, decreasing trend between V_d and N_h . In other words, as one may well expect, increasing the number of orifices through which air is injected decreases the extent of dead zones, with plate N1 (Fig. 5(a)) showing the most significant dead zones and N5 (Fig. 5(c)) the most complete fluidisation.

Despite the clarity of the trend observed, the origin of the improved fluidisation observed with increasing N_h cannot be reliably ascertained from this data set alone, as in increasing the number of orifices, we are simultaneously increasing the distributor's open area *and* the spatial distribution of orifices. As such, in order to better understand our system, we perform additional tests in order to individually probe these two factors.

In order to explore the effect of orifice size, a series of distributors was constructed using a fixed number of holes placed at a radius equivalent to the radial centre of the holes used in distributor N5. As is clear from Fig. 6, the relation between V_d and d_h is more complex than that relating V_d and N_h . For $d_h \gtrsim 2$ mm, the data suggest an increase in the quality of fluidisation (i.e. a decrease in V_d) with increasing orifice size. The lowest value of V_d , however, is in fact observed for the *smallest* orifice size tested. The origin of this unexpected deviation from the trend observed can be explained by the fact that, for this orifice size, the system follows Pattern *D* (see Fig. 1) which, as noted by Li *et al.* [6] is expected to induce superior mixing as compared to the Pattern *B* flow profiles exhibited by distributors S2-S5. The flow patterns for distributors S1 and S2 are shown in Fig. 7.

It is apparent from Fig. 6 that the trends observed from our varying-size data are not as robust as those observed for varying orifice number or varying position, as our data for this parameter span a relatively small range of V_d values, and possess comparatively large error bars. These data suggest, perhaps surprisingly, that of the parameters explored in this study, the hole diameter is of the least importance in terms of achieving fluidisation. The lack of statistically significant influence of the hole diameter becomes more striking still when we consider the variations in pressure drop, ΔP_d , across the different distributors relative to the pressure drop across the bed, ΔP_b . The distributors were designed so as to give pressure drop values varying from $\frac{\Delta P_d}{\Delta P_b} < 0.1$ (for which one would expect significant maldistribution [22]) to $\frac{\Delta P_d}{\Delta P_b} \gg 1$ (for which one would expect excellent distribution [22]). This result clearly illustrates the fact that a knowledge of open area alone is not sufficient to predict the efficacy of a distributor.

Finally, in order to explore the influence of orifice distribution, we utilised a series of distributor plates (S1-S5) with fixed orifice number and diameter – and hence also fixed open area – but varying orifice positioning (see Fig. 2).

The observed trend relating the dead volume, V_d , to the hole spacing, R_h , again suggests a monotonic decrease, with increased mixing observed as the holes are placed further from one another and the centre of the system. This trend can be fairly easily understood by considering the flow fields within the system. For distributors S1-S4, the net flow within the system followed pattern *B* with particles flowing upward in the centre of the system and down near the outer wall. This pattern will clearly create a tendency for particles to settle near the walls of the system, where the horizontal flux redirecting particles toward the centre of the bed (and hence the distributor orifices) is weakest. This effect will be strengthened by the fact that the absence of orifices near the wall means that such regions are susceptible to becoming defluidised. The more centrally the orifices are placed, the less likely a particle coming to rest near the system's outer wall is to be entrained by a bubble emanating from said orifices, or indeed to be fluidised at all, leading to larger dead zones for smaller R_h .

Conversely, the further from the system's centre the orifices are placed, the more likely descending particles are to come into contact with rising bubbles, and hence remain fluidised. In the case of distributor S5, for which the lowest V_d (i.e. the best fluidisation) is observed, we observe that the system again follows Pattern *D*.

Perhaps the most interesting outcome of this section as a whole is that, through a careful choice of orifice diameter and orifice distribution, it is possible to achieve strong fluidisation even with a very small number of individual orifices – the systems explored here are far removed from the case of the sintered metal plates, but nonetheless show good mixing potential.

3.2. Particle distribution

In the previous section, we have shown various ways in which the presence of dead zones within a fluidised bed may be reduced by varying the size, number and/or distribution of orifices belonging to the system's distributor. However, an absence of dead zones does not necessarily equate to a well-mixed system – it is well known that even a fully-fluidised bed may produce significant segregation between large/light and small/dense particles in a binary system [1, 34–36]. In this section, we attempt to assess quantitatively the degree to which each of our distributors provides a homogeneous distribution of larger particles throughout our system. In order to do so, we first use our PEPT data to produce – for each data set – a vertical occupancy distribution of glass particles,

as described in section 2.2.1. These one-dimensional profiles show the relative fraction of time spent by our tracer at any given height within our system. For ergodic systems such as those explored here, this is directly proportional to the mean concentration of glass particles [25], i.e. the data shown can be thought of as a density distribution of the larger particle species.

Fig. 8 shows the vertical occupancy profiles for all distributors explored. If the glass particles were homogeneously distributed within the system, then all cells within the bulk of the bed should possess identical occupancy or ‘residence fraction’ values, τ_i , equal to the mean value, τ_m , across all such cells. Conversely, for a poorly-distributed system, the measured occupancies will deviate strongly from their expected mean values. As such, we may define a ‘Distribution Index’, I_D , by which we may quantify the quality of large-particle distribution throughout a given system. For a system divided into N_c cells (see section 2.2.1), the distribution index may be calculated as

$$I_D = \frac{1}{N_c} \sqrt{\sum_i^{N_c} (\tau_i - \tau_m)^2} \quad (2)$$

For a perfectly-distributed system, $I_D = 0$, with larger values of the index corresponding to increasingly less-homogeneous distributions. The variation of the distribution index with the various parameters explored may be seen in Fig. 9.

It is important to note the distinction between the distribution index, I_D , discussed here and the dead volume index, V_d , discussed in the previous section. The distribution index provides a measure of the homogeneity of mixing between different particle species, and is agnostic to the dynamic state of the system. The dead volume, meanwhile, is agnostic to the degree of mixing between different species, depending only on the volume of the bed which are defluidised, irrespective of the specific compositions of the fluidised and defluidised regions.

In order to calculate a meaningful value of I_D , it is important that we utilise only cells in the bulk of the bed, i.e. below the disengagement zone, where the assumption of a uniform solids concentration is approximately valid. To do so, we implement a cut-off height determined as the point above which the distribution first dips below a designated minimum occupancy fraction, τ_r^{min} , and consider only cells falling below this height. As the ideal value of τ_r^{min} is not clearly defined, each data point in the left-hand panels of Fig. 9 represents the average of 8 individual I_D

values, each calculated with a different τ_r^{min} in the range $\tau_r^{min} \in [0.0025, 0.02]$. Though, in all cases, the absolute value of I_D shows a clear dependence on the chosen threshold, for I_D vs. N_h and I_D vs. d_H , the qualitative trends between the distribution index and the relevant control variable remain broadly consistent across all τ_r^{min} (see Fig. 9, right-hand panels), suggesting these trends to be robust and physically meaningful.

The same degree of reproducibility is not, however, apparent for the case of I_D vs. R_h . Moreover, the mean size of the error bars associated with the data for I_D vs. R_h shown in Fig. 9 is comparable to the range of the data plotted. This suggests that – for the systems explored here – while the radial positioning of the distributor orifices has a clear and significant impact on the *fluidisation* of the bed (i.e. the presence or absence of dead zones), it does not exert a strong influence over the quality of mixing observed. It is possible that a stronger variation in I_D with R_h may be masked by the transition from Pattern B to Pattern D between $R_h / R = 0.71$ and $R_h / R = 0.86$. However, even if this were the case, such an alteration would only affect the value of the plot's final data, and hence point would not be enough to produce a clear trend in the data observed.

By contrast, the number of orifices, N_h , demonstrates a strong and clear influence on the distribution of large particles through the system, with said distribution improving monotonically with N_h (see Fig. 9). The improvement of the glass-particle distribution with increasing N_h can also be seen from the top panel of Fig. 8, where we see a clear transition from a strongly peaked distribution for $N_h = 10$ to a highly-uniform distribution for $N_h = 50$. Interestingly, for the small orifice numbers in particular, the larger, less-dense particle species shows a notable tendency to segregate toward the *top* of the system. This suggests that for these systems the segregative process is driven by differences in density, as opposed to differences in size [34, 36] – despite the particles' size ratio ($\tilde{d} = \frac{d_1}{d_2} = 3.3$) being significantly larger than the density ratio ($\tilde{\rho} = \frac{\rho_2}{\rho_1} = 1.12$). Though theoretical models for both gas-fluidised [37] and vibrofluidised [38] beds suggest that the direction of segregation should be opposite for this combination of size and density ratio, making this a somewhat surprising result, it is not without precedent, as prior experimental studies have also shown systems with comparable $\tilde{d}$ and $\tilde{\rho}$ values to exhibit similar behaviour [39, 40].

Finally, for the case of varying orifice diameter, we observe two distinct peaks in the observed mixing quality (i.e. two local minima in the distribution index). The first, unsurprisingly, occurs for the smallest hole diameter $d_h = 1$ mm (distributor R1), where – as discussed in section 3.1 – the Pattern D flow field (see Fig. 7) is expected to cause increased mixing. A more unexpected result is the second minimum in I_D for distributor R4 ($d_h = 8$ mm), contrasting somewhat with our results in section 3.1. Some insight into the origin of this minimum may be gained by considering the forms of the vertical distribution profiles presented in Fig. 8, in particular the observed downward shift in the peak in τ_r with increasing orifice diameter. The profiles suggest a tendency for the larger, less-dense component of the system to segregate *above* the smaller, denser component for smaller orifice diameters but *below* it for larger diameters (though, in all cases, the glass particles seem to avoid the region immediately adjacent to the distributor. For $d_h = 8$ mm, we seemingly observe a ‘balance’ wherein neither upward nor downward segregation is preferred. At present, however, we can offer no concrete explanation of this observation, leaving it an open question for future research.

It is finally valuable to explicitly note that the clear distinctions observed between the trends in dead volume detailed in section 3.1 and those shown for the distribution index in the current section clearly demonstrate that conditions favourable to *fluidisation* are not, necessarily, conducive to *mixing*, and vice-versa. This is an important fact to consider for any project involving distributor design for binary particulate systems.

4. Conclusions

In this paper, positron emission particle tracking has been utilised to provide detailed information regarding the flow patterns, fluidisation quality and particle distribution within a gas-fluidised granular bed.

The number of orifices possessed by a distributor has shown a strong, positive correlation with both fluidisation quality and the homogeneity of particles within the bed; the relationships between other quantities, however, were found to be more nuanced.

Varying the radial position of a fixed number of orifices at a fixed open area and orifice diameter shows that the placement of orifices nearer the outer walls of the system reduces the volume of dead zones within the bed, but does not clearly influence the homogeneity of the bed as

a whole.

Perhaps most surprisingly, varying the the orifice diameter at fixed orifice position and number shows no statistically significant impact on the dead-zone volume, despite a more than twenty-fold variation in open area. Conversely, the orifice diameter shows a distinct but decidedly non-monotonic influence on the spatial distribution of particles. At present, the physical origin of this behaviour has not been identified; it is hoped that this observation – and indeed others within this work– will provide a catalyst for further research in this area.

Declaration of interests

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

References

- [1] B. Formisani, R. Girimonte, Experimental analysis of the fluidization process of binary mixtures of solids, *KONA Powder and Particle Journal* 21 (2003) 66–75 (2003).
- [2] E. Teunou, D. Poncelet, Batch and continuous fluid bed coating–review and state of the art, *Journal of food engineering* 53 (4) (2002) 325–340 (2002).
- [3] D. M. King, J. A. Spencer II, X. Liang, L. F. Hakim, A. W. Weimer, Atomic layer deposition on particles using a fluidized bed reactor with in situ mass spectrometry, *Surface and Coatings Technology* 201 (22-23) (2007) 9163–9171 (2007).
- [4] K. Shibata, M. Shinizu, S.-i. Inaba, R. Takahashi, J.-i. Yagi, Pressure loss and hold-up powders for gas-powder two phase flow in packed beds, *ISIJ international* 31 (5) (1991) 434–439 (1991).
- [5] H. Yurong, L. Huilin, S. Qiaoqun, Y. Lidan, Z. Yunhua, D. Gidaspow, J. Bouillard, Hydrodynamics of gas–solid flow around immersed tubes in bubbling fluidized beds, *Powder technology* 145 (2) (2004) 88–105 (2004).
- [6] Y. Li, H. Fan, X. Fan, Identify of flow patterns in bubbling fluidization, *Chemical Engineering Science* 117 (2014) 455–464 (2014).
- [7] S. Chiba, T. Chiba, A. Nienow, H. Kobayashi, The minimum fluidisation velocity, bed

- expansion and pressure-drop profile of binary particle mixtures, *Powder Technology* 22 (2) (1979) 255–269 (1979).
- [8] T. H. Pedersen, F. Conti, Improving the circular economy via hydrothermal processing of high-density waste plastics, *Waste management* 68 (2017) 24–31 (2017).
- [9] M. Wormsbecker, T. S. Pugsley, H. Tanfara, The influence of distributor design on fluidized bed dryer hydrodynamics (2007).
- [10] M. Wormsbecker, T. Pugsley, J. R. v. Ommen, J. Nijenhuis, R. Mudde, Effect of distributor design on the bottom zone hydrodynamics in a fluidized bed dryer using 1-d x-ray densitometry imaging, *Industrial & engineering chemistry research* 48 (15) (2009) 7004–7015 (2009).
- [11] J. Paiva, C. Pinho, R. Figueiredo, The influence of the distributor plate on the bottom zone of a fluidized bed approaching the transition from bubbling to turbulent fluidization, *Chemical Engineering Research and Design* 82 (1) (2004) 25–33 (2004).
- [12] K. S. Rao, G. V. Reddy, Effect of distributor design on temperature profiles in fluidized bed during the combustion of rice husk, *Combustion science and technology* 179 (8) (2007) 1589–1603 (2007).
- [13] H. Kumar, S. Mohapatra, R. Singh, Three-dimensional cfd modelling of a fluidised bed combustor fuelled by biomass and coal, *Materials Research Innovations* 19 (sup8) (2015) S8–118 (2015).
- [14] H. Kuo, H. Zhang, R. Hsu, Cfd modelling of a countercurrent staged fluidised bed, *Chemical Engineering Journal* 137 (3) (2008) 664–676 (2008).
- [15] L. Li, B. Li, Z. Liu, Modeling of spout-fluidized beds and investigation of drag closures using openfoam, *Powder Technology* 305 (2017) 364–376 (2017).
- [16] S. Karimipour, T. Pugsley, Application of the particle in cell approach for the simulation of bubbling fluidized beds of geldart a particles, *Powder technology* 220 (2012) 63–69 (2012).
- [17] I. Afrooz, C. Sinnathambi, S. Karuppanan, D. Ching, Cfd simulation of bubbling fluidized bed: Effect of distributor plate orifice pattern configuration on hydrodynamics of gas-solid mixing, *ARPN J. Eng. Appl. Sci* 11 (2006) 11954–11959 (2006).
- [18] F. Depypere, J. G. Pieters, K. Dewettinck, Cfd analysis of air distribution in fluidised bed equipment, *Powder Technology* 145 (3) (2004) 176–189 (2004).

- [19] A. Al-Akaishi, A. Valera-Medina, C. Chong, R. Marsh, Cfd analysis of the fluidised bed hydrodynamic behaviour inside an isothermal gasifier with different perforated plate distributors, *Energy Procedia* 142 (2017) 835–840 (2017).
- [20] S. Cooper, C. J. Coronella, Cfd simulations of particle mixing in a binary fluidized bed, *Powder Technology* 151 (1-3) (2005) 27–36 (2005).
- [21] D. Geldart, Types of gas fluidization, *Powder technology* 7 (5) (1973) 285–292 (1973).
- [22] D. Geldart, *Gas fluidization technology*(1986).
- [23] D. Parker, R. Forster, P. Fowles, P. Takhar, Positron emission particle tracking using the new birmingham positron camera, *Nuclear Instruments and Methods in Physics Research Section A: Accelerators, Spectrometers, Detectors and Associated Equipment* 477 (1-3) (2002) 540–545 (2002).
- [24] D. Parker, Positron emission particle tracking and its application to granular media, *Review of Scientific Instruments* 88 (5) (2017) 051303 (2017).
- [25] R. Wildman, J. Huntley, J.-P. Hansen, D. Parker, D. Allen, Single-particle motion in three-dimensional vibrofluidized granular beds, *Physical Review E* 62 (3) (2000) 3826 (2000).
- [26] R. Wildman, J. Huntley, D. Parker, Granular temperature profiles in three-dimensional vibrofluidized granular beds, *Physical Review E* 63 (6) (2001) 061311 (2001).
- [27] R. Wildman, J. Huntley, D. Parker, Convection in highly fluidized three-dimensional granular beds, *Physical review letters* 86 (15) (2001) 3304 (2001).
- [28] C. Windows-Yule, T. Weinhart, D. Parker, A. Thornton, Effects of packing density on the segregative behaviour of granular systems, *Physical review letters* 112 (9) (2014) 098001 (2014).
- [29] C. Windows-Yule, A. Rosato, A. Thornton, D. Parker, Resonance effects on the dynamics of dense granular beds: achieving optimal energy transfer in vibrated granular systems, *New journal of physics* 17 (2) (2015) 023015 (2015).
- [30] C. Windows-Yule, B. Scheper, A. van der Horn, N. Hainsworth, J. Saunders, D. Parker, A. Thornton, Understanding and exploiting competing segregation mechanisms in horizontally rotated granular media, *New journal of physics* 18 (2) (2016) 023013 (2016).
- [31] A. Nicusan, C. Windows-Yule, Positron emission particle tracking using machine learning, Accepted for publication in *Review of Scientific Instruments* (2020).

- [32] A. NIENOW, N. NAIMER, T. CHIBA, STUDIES OF SEGREGATION/MIXING IN FLUIDISED BEDS OF DIFFERENT SIZE PARTICLES, *Chemical Engineering Communications* 62 (1-6) (1987) 53–66 (dec 1987).
doi:10.1080/00986448708912050. URL <https://doi.org/10.1080/00986448708912050>
- [33] Y. Zhang, B. Jin, W. Zhong, B. Ren, R. Xiao, DEM simulation of particle mixing in at-bottom spout-fluid bed, *Chemical Engineering Research and Design* 88 (5-6) (2010) 757–771 (may 2010). doi:10.1016/j.cherd.2009.11.011. URL <https://doi.org/10.1016/j.cherd.2009.11.011>
- [34] B. Formisani, R. Girimonte, T. Longo, The fluidization process of binary mixtures of solids: development of the approach based on the fluidization velocity interval, *Powder Technology* 185 (2) (2008) 97–108 (2008).
- [35] B. Formisani, R. Girimonte, V. Vivacqua, Fluidization of mixtures of two solids differing in density or size, *AIChE Journal* 57 (9) (2011) 2325–2333 (2011).
- [36] B. Formisani, R. Girimonte, V. Vivacqua, Fluidization of mixtures of two solids: a unified model of the transition to the fluidized state, *AIChE Journal* 59 (3) (2013) 729–735 (2013).
- [37] F. P. Di Maio, A. Di Renzo, V. Vivacqua, A particle segregation model for gas-fluidization of binary mixtures, *Powder technology* 226 (2012) 180–188 (2012).
- [38] D. C. Hong, P. V. Quinn, S. Luding, Reverse brazil nut problem: competition between percolation and condensation, *Physical Review Letters* 86 (15) (2001) 3423 (2001).
- [39] A. P. Breu, H.-M. Ensminger, C. A. Kruelle, I. Rehberg, Reversing the brazil-nut effect: competition between percolation and condensation, *Physical review letters* 90 (1) (2003) 014302 (2003).
- [40] A. Rao, J. S. Curtis, B. C. Hancock, C. Wassgren, Classifying the fluidization and segregation behavior of binary mixtures using particle size and density ratios, *AIChE Journal* 57 (6) (2011) 1446–1458 (2011).

Figure 1: The four flow patterns described by Li *et al.* [6].

Figure 2: Schematic images (to scale) showing the various distributor designs detailed in Table 1.

Figure 3: Plot showing the pressure drop curve from which the system's U_{mf} was extracted. The

values shown represent an average taken across 10 repeat measurements, with the error bars representing the standard deviation of the values acquired.

Figure 4: Schematic diagram illustrating the manner in which PEPT locates a single tracer particle. If an active tracer particle (highlighted red for clarity, though in reality identical to all others of its species) emits a pair of back-to-back γ photons (panel (a)) that are both simultaneously recorded by the detectors of the positron camera, their trajectory can be reconstructed algorithmically (panel (b)). By finding the point of intersection of several such lines (panel (c)), the position of the tracer's centre may be determined.

Figure 5: Data showing azimuthally-averaged fractional occupancy or 'residence time' (τ_r) distributions (colours) and velocity vector fields (arrows) for (a) distributor N1, (b) distributor N3 and (c) distributor N5. White regions near the base of the system represent apparent 'dead' regions not visited by the tracer. For clarity, the velocity vectors shown are normalised to unit length.

Figure 6: Variation in the total dead volume, V_d , within our systems as a function of the number of orifices in the tested distributor plate (N_o), the diameter of said orifices (centre) and the radial position of their centres on the plate, normalised by the internal radius, R , of the system.

Figure 7: Velocity vector fields for distributors S1 and S2, illustrating the two- and four-roll patterns B and D in the r - z coordinate system. For clarity, the velocity vectors shown are normalised to unit length.

Figure 8: One-dimensional axial occupancy profiles for the cases of varying orifice number (top), orifice diameter (centre) and radial position (bottom).

Figure 9: Variation in the value of the distribution index, I_D (see main text) as a function of the number of orifices in the tested distributor plate (top), the diameter of said orifices (centre) and the radial position of their centres on the plate, normalised by the internal radius, R , of the system. The different coloured symbols in the right-hand column of images correspond to values

calculated with differing τ_r^{min} in the range $\tau_r^{min} \in [0.0025, 0.02]$. Data sets calculated with τ_r^{min} values closer to the lower (higher) end of this range correspond to colours further to the red (blue) end of the spectrum used. Data in the left-hand images correspond to the averages of the data shown in the equivalent right-hand plots.

Table 1: Details of the distributor designs used in experiment, alongside the shorthand names by which each individual design is referred to in the main text, the number of holes in the plate (N_h), the diameter of the holes (d_h) and their spacing (R_h), the open area of the distributor, and the pressure drop across it at U_{mf} (ΔP). Pressure drop values marked by asterisks correspond to readings too high to be measured using our equipment.

Name	N_h	d_h (mm)	R_h (mm)	Open area (%)	ΔP (Pa)
Diameter 1 (R1)	5	2	26	0.4	*
Diameter 2 (R2)	5	4	26	1.4	2400
Diameter 3 (R3)	5	6	26	3.2	550
Diameter 4 (R4)	5	8	26	5.7	240
Diameter 5 (R5)	5	10	26	8.9	160
Number 1 (N1)	10	2	6.1	0.7	*
Number 2 (N2)	20	2	5.0-9.1	1.4	2600
Number 3 (N3)	30	2	5.0-12.3	2.1	1400
Number 4 (N4)	40	2	5.0-15.4	2.8	1100
Number 5 (N5)	50	2	5.0-18.4	3.6	590
Spacing 1 (S1)	5	8	11.7	5.7	240
Spacing 2 (S2)	5	8	17.5	5.7	240
Spacing 3 (S3)	5	8	23.3	5.7	240
Spacing 4 (S4)	5	8	29.2	5.7	240
Spacing 5 (S5)	5	8	35	5.7	240

Graphical abstract

Highlights

- Influence of distributor orifice size, number and position on fluidised bed dynamics investigated.
- Positron emission particle tracking used to quantify extent of defluidisation and bed homogeneity
- Increasing orifice number found to increase both fluidisation quality and bed homogeneity.
- Changing orifice size at fixed orifice number found to have little impact on fluidisaion.

Figure 1

Figure 2

Figure 3

Figure 4

Figure 5

Figure 6

Figure 7

Figure 8

Figure 9