QUANTITATIVE EVALUATION OF INERT SOLIDS MIXING IN A BUBBLING FLUIDIZED BED

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Abstract: This paper presents a method to evaluate the lateral mixing processes of bed material in bubbling fluidized beds. The method combines experiments and mathematical modeling and has the aim to constitute a tool for the investigation of the complex solids mixing phenomenon in fluidized bed units. The experimental method used is based on indirect measurements of a tracer agent. A batch of tracer is fed in one corner of the bed and the amount of tracer agent which traversed the bed is measured over time in a corner, diagonally opposite to the tracer feed point. The mathematical model is based on solving a diffusion-like partial differential equation describing the transient lateral dispersion of particles. From this, values of the lateral dispersion coefficient can be obtained.

The method is applied to evaluate the lateral solids mixing in a fluid-dynamically downscaled 3-dimensional cold model with cross-sectional dimensions of 0.3 m x 0.3 m. The cold model can be operated with a variable bed height up to 0.16 m. Since the cold model is designed according to Glicksman's full set of scaling laws the fluid dynamics is assumed to resemble that of an industrial-scaled bubbling fluidized bed operated at 900°C with cross-sectional dimensions of 1.5 m x 1.5 m and bed heights up to 0.8 m. The results show good qualitative agreement between experimental results and the mathematical modeling and it is concluded that the macroscopic lateral solids mixing behavior in the bed geometry investigated can be described by a diffusion-like partial differential equation. Four superficial gas velocities are investigated and the lateral dispersion coefficients obtained are found to increase steadily over the range of the superficial velocities investigated.

Keywords: Lateral dispersion, bed material

INTRODUCTION

The mixing phenomenon is of great importance in most large-scale chemical processes, such as combustion. Mixing promotes mass and heat transfer and thus governs a significant part of the fuel conversion in the combustion process. In the case of fluidized bed combustion, the lateral mixing of bed material is of special relevance, since it governs the temperature distribution and the fuel mixing which are parameters of crucial importance in the conversion process. For large-scale boilers it is critical to have as homogenous fuel distribution as possible in order to minimize operating costs and emissions. Large-scale units are often built with a large furnace cross-sectional area and operated with low bed heights (*i.e.* a low bed height-to-bed width ratio). Furthermore, mixing in the vertical direction is higher than in the lateral direction (Ito et al., 1999). Thus, lateral mixing is the limiting factor in the overall solids mixing, which implies a risk for uneven fuel distribution in the furnace. This can result in oxygen-depleted zones with incomplete burn out as well as zones with an excess of oxygen. In particular, release of volatile matter can become relatively concentrated in the vicinity of fuel feeding ports, increasing the risk for incomplete volatile combustion in the furnace. This un-combusted volatile species will burn in the cyclone leading to increased, and possibly undesired, cyclone temperatures. In all, this may result in unclear combustion conditions and the need to operate with an un-necessarily high excess air ratio. In addition, presence of strongly reducing zones at furnace walls may enhance the risk for corrosion of furnace walls (*cf.* Niklasson et al., 2003). The mixing of fuel is governed by the mixing of bed material in fluidized beds and therefore lateral mixing of the inert material is of great importance for the operation of large-scale fluidized bed combustors. Thus, there is a need to investigate and develop modeling tools which can describe the solids mixing process under conditions relevant for industrial scale combustors.

This work aims at presenting a method to quantitatively investigate the lateral mixing of bed material in a bubbling fluidized bed through experiments and mathematical modeling. As an example of the possibilities offered by the presented method, the influence of slight variations in one operational parameter (fluidization velocity) on the lateral solids dispersion coefficient is reported. An exhaustive analysis, by means of the presented method, including a wider range of superficial velocities and several other parameters (bed height, pressure drop and arrangement of the distributor plate, forced convective solids flow through the bed) is planned as future work. The experimental work is conducted in a fluid-dynamically downscaled cold flow model of a bubbling fluidized bed combustor (although results of this work should also be relevant for fluidized-bed gasifiers).

A number of different modeling approaches to describe solids mixing exist in literature and an overview of such approaches is provided by Costa and Souza-Santos (1999). The present work uses the random dispersion approach, based on assumptions given by Kunii and Levenspiel (1991). Although the solids mixing in fluidized beds, is in general strongly convective, in large enough geometries with homogeneous nozzle distribution the process can be approached from a macroscopic perspective by a random dispersion process. Figure 1 shows a typical solids flow pattern for a wide bubbling fluidized bed, *i.e.* a bed which has typical geometrical characteristics of a bed of industrial scale. The fluidization gas forms bubbles which rise through the bed. These bubbles induce horizontally-aligned convective flow structures around their rising paths as shown by Pallarès and Johnsson (2006). If bubble coalescence occurs, a pattern with horizontally-aligned flow structures is still found above the coalescence height. Thus, the dispersion model is also able to describe properly cases with bubble coalescence.

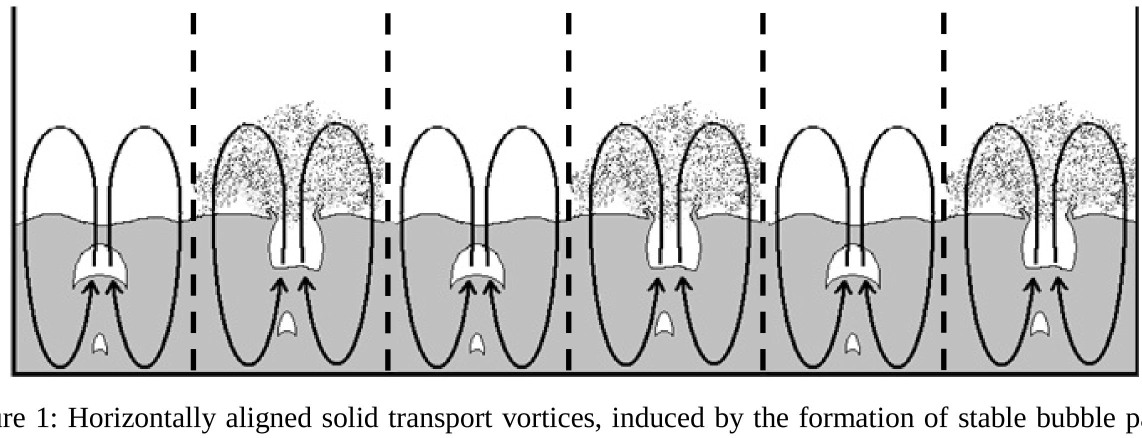

Figure 1: Horizontally aligned solid transport vortices, induced by the formation of stable bubble paths. Particles are ejected at the surface when a bubble erupts. The flow domain can be divided into discrete regions which are indicated by the dashed lines. Based on findings by Pallarès and Johnsson (2006).

The basis of the random dispersion approach is that, in the lateral direction, solid particles are mainly mixed by means of the bubble flow which, given a homogeneous distribution of the nozzles, produces an isotropic solids mixing. The main lateral solids mixing mechanisms are 1) due to bubble eruptions at the bed surface, which ejects particles into the splash zone, as proposed by Kunii and Levenspiel (1991) and 2) due to that solids is dragged into the wake region of ascending bubbles (see Fig. 1). Both these mechanisms can be assumed to be equal in all lateral directions. Thus, the lateral solids mixing in each of the bubble paths is mainly convective but isotropic. Olsson (2011) made an experimental study in a fluidized bed with a large number of bubble paths and showed that it is possible to macroscopically describe the mixing by a random dispersion process, provided enough number of bubble paths are present in the bed. The lateral mixing in such an arrangement of discrete regions, *i.e.* bubble paths (see Fig. 1) where movement is only allowed between neighboring regions, can be seen as a random walk process which, mathematically, can be described by a Markov chain. Diffusion of mass can also be seen as a random walk process and be described similarly. Thus, although being strongly convective, the lateral solids mixing in a fluidized bed can be approached macroscopically by the diffusion equation, which reads:

$$\frac{\partial C}{\partial t} = \frac{\partial}{\partial x}\left(D\frac{\partial C}{\partial x} \right) + \frac{\partial}{\partial y}\left(D\frac{\partial C}{\partial y} \right) \qquad (1)$$

In which the transported scalar C denotes the solids concentration. When applying Eq. (1) to the bed mixing process, D corresponds to the dispersion coefficient which can be empirically-determined, but then includes contribution from both convection and diffusion.

EXPERIMENTAL SETUP

The unit used for the experiments is showed in Fig. 2. It is a cold model with cross-sectional dimensions of 0.3 m x 0.3 m which is fluidized at ambient temperature (and pressure) with the air taken from the in-house pressurized-air system with the flow controlled by a rotameter.

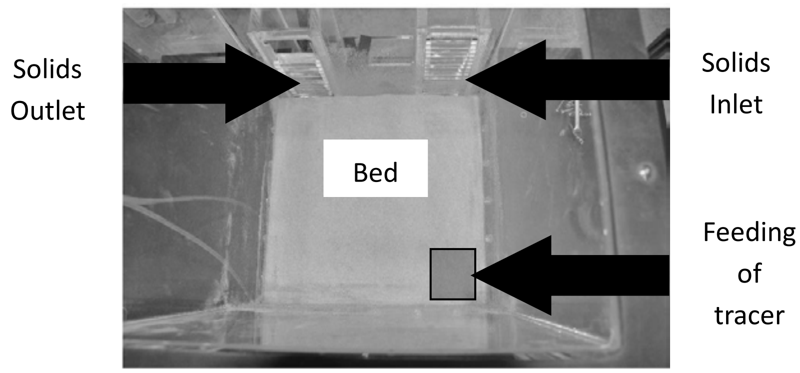

Figure 2: A picture of the cold unit used in the experiments (view from top of unit). The tracer agent is inserted in the bottom-right corner and the concentration was measured over time at the outlet located in the top-left corner.

Glicksman's full set of scaling laws (Glicksman, 1994) are applied to define the geometry and the solids used and, thus, the unit is assumed to fluid-dynamically resemble an industrial-scaled bubbling fluidized bed operated at 900°C with cross-sectional dimensions of 1.5 m x 1.5 m. The unit can be operated with different bed heights by changing the number of blocks located by the outlet (see Fig. 2). Bed heights up to 0.16 m can be applied in this manner, which would correspond to a bed height of 0.8 m in the large-scale unit, according to the scaling relationships used. However, during the experiments reported in this work the initial bed height was fixed to 0.06 m, which corresponds to a bed height of 0.3 m in the large-scale unit.

The full set of scaling laws proposed by Glicksman (1994) was used to select bed material. It should be pointed out that the scaled bed does not correspond to a specific industrial unit, but have dimensions chosen to correspond to a typical fluidized-bed combustor or gasifier of some 10MW of size. Table 1 lists the main parameters as obtained from application of the scaling laws. As ambient air was selected as fluidizing agent, a length scaling factor of 5 was obtained from the scaling laws. Bronze powder was used as the bulk bed material and iron powder as tracer agent due to that it is magnetic and, thus, can be separated from the bronze powder in the samples collected. Also, the iron used has similar size distribution as the bronze powder. As can be seen from Table 1 it is not possible to completely fulfill the scaling laws used since densities of the two materials (bronze and iron) are too low (16% and 26%, respectively).

Table 1: Comparison between the parameters for the large-scale unit, those given by the scaling laws and those applied in the experimental work.

Parameter	Large-scale unit	According to scaling laws	Experimental
Temperature [°C]	900	20	20
Kinematic gas viscosity [m^2/s]	$1.57 \cdot 10^{-4}$	$1.50 \cdot 10^{-5}$	$1.50 \cdot 10^{-5}$
Gas density [kg/m^3]	0.29	1.21	1.21
Bed Geometry [m]	L	L/5	L/5
Particle diameter (average) [µm]	300	60	75
Solids density [kg/m^3]	2600	10600	Iron – 8900 Bronze - 7800
Superficial velocity [m/s]	u	sqrt(u^2/5)	sqrt(u^2/5)

The following procedure is used to obtain the lateral dispersion coefficient: Batches of the iron powder are fed to the bubbling bed in the designated solids feeding corner indicated in Fig. 2. Samples of the bed material mixture are then collected over time at the solids outlet, located in the diagonally opposite corner (see Fig. 2). Samples are collected during a period of 60 seconds, with 2-seconds interval between samples, and their respective iron concentrations are determined by magnetic separation. The dispersion coefficient is found by fitting solutions to Eq. (1) to the measured tracer concentration. The fluidization velocity in the tests ranged from 0.13 m/s to 0.37 m/s in the cold model (corresponding to 0.29 m/s to 0.83 m/s in the large-scale unit), while all other parameters were kept constant.

RESULTS AND DISCUSSION

Figure 3 exemplifies results from one of the conducted experiments (fluidization velocity of 0.37 m/s). The initial iron concentration is approximately zero (small amounts of iron powder from previous experiments can remain at the beginning of each new experiment). The outlet iron concentration is increasing rapidly while the tracer solids is mixed into the bed and dispersed horizontally through the bed. Once the tracer material is well mixed the measured tracer concentration levels out. As can be seen, solutions of Eq. (1) can be used to fit the experimental data with good accuracy. For the example shown, the fit gives a dispersion coefficient D=0.0017 m^2/s. Similar agreement to what is illustrated in Fig. 3 was obtained in all four cases investigated. As mentioned above, samples are collected over a period of 60 seconds. However, the magnetic separation of the tracer agent requires up to 20 hours. The reason for this is the particle size: The iron particles intertwine themselves with the bronze particles making it difficult to separate them after the measurement runs. This is partly due to that the magnetic force between the magnet and the small sized particles is rather weak. Since it is the transient which is important to capture in order to model the dispersion coefficient, a non-zero initial tracer concentration is assumed acceptable, although significant effort is put into keeping the amount of remaining iron as low as possible.

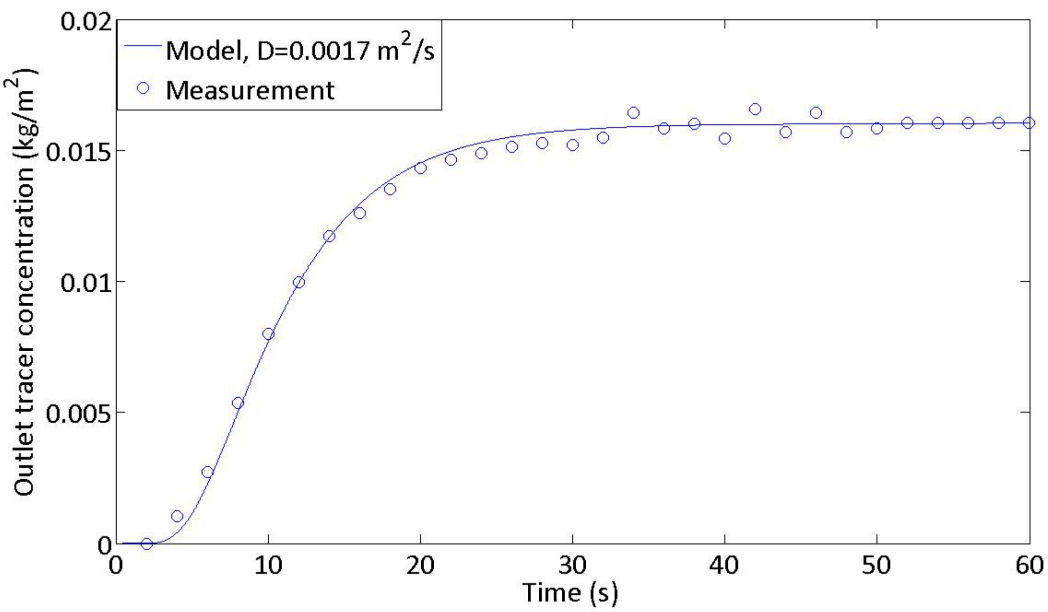

Figure 3: Results from a run at U_0 =0.37 m/s. The dispersion coefficient was determined to 0.0017 m^2/s.

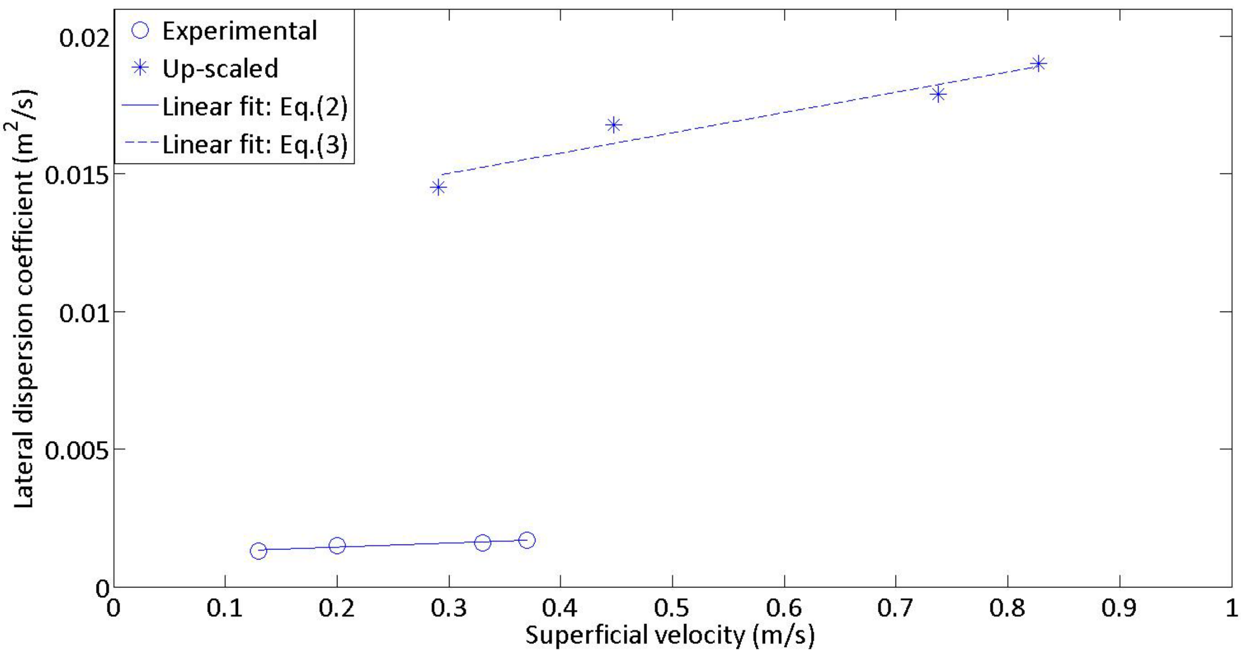

Figure 4: Lateral dispersion coefficients for different velocities from experiments, as measured and recalculated back to large-scale conditions.

Figure 4 summarizes the dispersion coefficients found for the four different fluidization velocities investigated, expressed both as experimental values and as rescaled back to the large-scale conditions. Over the range of velocities investigated a linear function is fitted to the data points as expressed by Eqs (2) and (3). The linear fits are obviously only intended for the range where the measurements are made. The purpose of the linear fit is to estimate the dispersion coefficient for velocities other than the measured, but within the interval where measurements were made (*i.e.* for interpolation purposes). Linear regressions of the data in Fig. 4 yield:

$$D_{cold} = 0.001474 \cdot u_0 + 0.001145 \quad u_0 \in [0.13, 0.37] \tag{2}$$

$$D_{l\arg e\ scale} = 0.007372 \cdot u_0 + 0.01281 \quad u_0 \in [0.29, 0.83] \tag{3}$$

There are several investigations of dispersion coefficients in literature with values differing with several orders of magnitude between the different works, Breault (2006). These notable differences can be explained by the use of different values in some operational parameters. For example, Shi and Fan (1984) conducted experiments in a 2-dimensional model with fluidization velocities in a similar range as this work, finding dispersion coefficients two orders of magnitude lower than the present values. The reason for this difference can be found in that Shi and Fan used bed heights in the range 0.02 - 0.05 m, which is far from the bed height used in this work (0.3 m, on up scaled basis) and carried out their investigations in a 2-dimensional bed. Berruti et al. (1986) made investigations in a cylindrical fluidized bed operated at similar conditions (bed height and gas velocity) as those in this work, obtaining dispersion coefficients two orders of magnitudes lower than the present ones. Possible explanations for this are the difference in cross-sectional bed size (Ø=0.267 m against the present 0.3 m x 0.3 m), and the arrangement of the gas distributor. It should be noted that estimations of the solids dispersion coefficient gain in reliability as bed size increases (both because of the better statistics obtained and the improved suitability of the dispersion approach in large geometries). In the work by Schlichtaerle and Werther (2001), the range of superficial velocities originally investigated by Bellgardt and Werther (1986) are extended and a quadratic correlation between the dispersion coefficient and the superficial velocity is found. In the original work by, Bellgardt and Werther (1986), investigating almost the same range of superficial velocities as the present work, they found a linear relation when using a bed height of 0.41 m and a quadratic relation when using a bed height of 0.82 m. Both cases used particles with a diameter of 0.45 mm. This should be compared with the present work which uses an up-scaled bed height of 0.3 m and particles with average diameter of 0.38 mm. In the case with the lower bed height the dispersion coefficients found by Bellgardt and Werther varies linearly from $0.7 \cdot 10^{-3}$ to $1.25 \cdot 10^{-3}$ m^2/s, *i.e.* an order of magnitude lower than the present work. This discrepancy can possible be explained by the fact that neither Bellgardt and Werther nor the authors referenced above applied fluid dynamic scaling laws to their experimental facility, i.e. their experiments were carried out at cold conditions while the present ones upscale to hot conditions.

Since both the bed material and tracer agent used have somewhat too low densities (bronze 16% and iron 26%) compared to the density required to fulfill the scaling laws, the scaling cannot be seen as exact. It is reasonable to assume that lighter particles are easier to lift and thus back-mixing by particle ejection after bubble eruption is enhanced. This should imply an overestimation of the lateral solids dispersion coefficient. On the other hand, the experiments are carried out with coarser particles than the size determined by the scaling laws, which should have the opposite effect *i.e.* underestimating the dispersion coefficient. It is uncertain which of these two effects is dominating or if they cancel out each other. The terminal velocity for a single particle with an ideal size and density is 1.02 m/s (60 µm and 10600 kg/m^3 according to the scaling laws). For a single particle with ideal density but a diameter of 75 µm the corresponding velocity is 1.44 m/s. The terminal velocities for the bronze powder and iron powder used in the experiments are 1.24 m/s and 1.11 m/s respectively, which is in-between the other two values as expected.

CONCLUSIONS

A model for determining the lateral solids dispersion in a downscaled fluidized bed combustor has been developed and applied to investigate the influence of fluidization velocity on the dispersion of inert material in a fluidized bed. Over the range of superficial velocities investigated (corresponding to 0.29 m/s to 0.83 m/s in the large size unit) a linear relationship between the lateral dispersion coefficient (increasing from 0.015 m^2/s to 0.019 m^2/s in large-scale) and the superficial velocity is found. Due to the large difference in the literature values of dispersion coefficients it is not straight forward to compare different works. Although other authors have found similar trends under comparable conditions, further investigation is required in order to quantify the lateral solids dispersion coefficient, especially with respect to expanding the range of superficial velocities and investigating the influence of bed heights and nozzle characteristics and distribution.

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