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Experimental evaluation of lateral mixing of bulk solids in a fluid-dynamically down-scaled bubbling fluidized bed

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ABSTRACT

An indirect tracking method for bed material using magnetic separation was applied to a fluid-dynamically down-scaled fluidized bed, to evaluate the influences of different parameters on the lateral dispersion coefficients of the bed material. Solutions to the transient diffusion equation were fitted to the experimental data and showed that the dispersion approach could be used to describe the lateral mixing of solids at the macroscopic level. The values obtained for the dispersion coefficient were scaled-up to be relevant to large-scale boilers operated under high-temperature conditions. The scaled-up lateral solid dispersion coefficients were in the order of 10^{-2} m^2/s, i.e., two orders of magnitude greater than those reported in the literature for smaller sized fluidized and/or fluidized beds operated under ambient-temperature conditions. This paper also considers the mixing phenomena at the mesoscopic level, applying the so-called "mixing cell" concept to elucidate how the mixing of solids is dependent upon the flow characteristics around the main bubble paths.

1. Introduction

In most large-scale chemical processes, such as the combustion and gasification of solid fuels using the fluidized-bed technology, the mixing of inert solids is of major importance [1]. Mixing governs not only how fast the fuel is mixed throughout the unit (mainly through solid–solid interactions), but also how good the contact is between the fuel and the gas phase. In addition, an increase in lateral mixing of the bulk bed material creates a more homogeneous temperature field across the cross-section of the bed owing to the strong thermal inertia of the solids. Thus, the mixing of the solids controls the mass and heat transfer, which in combination with the chemical kinetics governs the conversion of fuel in combustion and gasification processes. Large-scale fluidized bed units can be operated under either bubbling or circulating conditions [2]. Regardless of the mode of operation, large-scale fluidized bed boilers for solid fuel conversion are generally operated with a dense bottom bed [3]. It is important to understand the mixing phenomena in these beds, so as to develop fluidized-bed modeling that is applicable to both combustion and gasification systems. Dense bottom beds in large-scale units have a low aspect ratio of <1 [4], which considered in combination with lateral solids mixing (being at least one order of magnitude lower than that in the vertical direction) [5,6] means that the mixing of solids in the lateral direction is a limiting process that requires closer investigation.

In fluidized-bed combustion, the limitations associated with solids mixing may result in large variations in temperature, and consequently, variations in the combustion rate, across the furnace. This is critical when burning highly volatile fuels, which typically engender strongly reducing conditions at the furnace walls that hold the fuel inlets and regions of oxidation at locations further away from the fuel entrance. These spatial differences in oxidation/reduction may result in high levels of emissions of unwanted species, corrosion, and hot-spots [7]. To prevent these effects, fluidized-bed boilers are operated at higher level of excess air and reduced steam temperatures, as compared with boilers that burn low-grade fuels. Thus, although fuel flexibility is one of the main advantages of the fluidized-bed technology, possibilities remain to improve the efficiency of combustion of low-grade fuels by decreasing the level of excess air and increasing the steam temperature, which can be achieved by improving the lateral mixing of solids. In contrast to direct combustion, the performance of dual fluidized bed systems, such as indirect gasification [8] and chemical looping combustion (CLC) of solid fuels [9], may benefit from limited lateral mixing of solids. Thus, moderate levels of lateral mixing in indirect gasifier beds and in CLC fuel reactor beds increase the fuel residence time, thereby minimizing losses of unconverted char to the secondary reactor [8,10]. However, the rate of mixing of bed materials in such units has to be sufficiently high to maintain the optimal bed temperature across the bed, which enhances the relatively slow and temperature-sensitive reaction rates of gasification. In summary, there is a need to improve current understanding of the phenomena that govern the process of lateral mixing of solids in fluidized beds.

One of the difficulties encountered when interpreting the results of studies reported in the literature that have focused on the lateral mixing of solids is that they often applied gas distributors with a pressure drop that is considerably greater than that typically employed in large-scale

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units used for fuel conversion. As this discrepancy results in patterns of solid flows that may differ from those observed in commercial units, it imposes a limitation on the adoption of such studies as the basis for understanding mixing in large-scale units.

The present work aimed to develop and apply an experimental method for quantifying the lateral dispersion coefficients of in-bed solids in fluidized beds under conditions that are relevant for industrial-sized units. The work applies different operational conditions with respect to fluidization velocity, bed height, and gas distributor pressure drop. The applied method combines mathematical modeling with indirect measurements of a tracer agent in a cold flow model [11], which is fluid-dynamically downscaled according to the set of scaling laws derived by Glicksman et al. [12].

2. Theory

In fluidized beds, the gas bubbles form a number of main pathways [13], around which a corresponding number of horizontally aligned convective solid flow structures have been identified [14]. These solid flow structures are referred to as 'mixing cells'. Three main mechanisms for lateral mixing of solids have been identified [15]: 1) bubble eruptions at the bed surface, with particles being ejected into the splash zone with a certain horizontal velocity component [16,17]; 2) emulsion drift in a downward direction in parallel to the rising bubbles, which also can imply lateral displacement [18]; and 3) the dragging of particles from the surrounding emulsion into the wake region behind the rising bubbles [19]. The first two mixing mechanisms account for most of the lateral mixing. The third mechanism entails strong upward acceleration of the particles; according to Gómez-Barea and Leckner [6], this governs vertical mixing, which is an order of magnitude greater than mixing in the lateral direction.

Over the last 60 years, theoretical and experimental investigations have been conducted in the field of solid mixing in fluidized beds, for a review, see Breault [20]. The underlying physical phenomena of the mixing process are obviously governed by the fluid dynamics of the gas and solid phases, for which a set of partial differential equations can be derived to describe the conservation of mass and momentum for each phase [21], if these phases are assumed as continuums. As there is no general analytic solution for such governing equations, numerical approximations and high computational power are required to obtain a numerical solution [22]. To study the phenomena of solid mixing at lower computational cost, different approximations of the governing equations are employed in literature, which Costa [23] divided into empirical, dispersive (this work), and counter-current.

The present work applies the dispersive approach to evaluate tracer measurements. This approach is based on the assumption made by Rowe and Partridge [19] that the lateral mixing of solids in a fluidized bed, despite the convection driven by the rising gas bubbles, can be represented as a random isotropic dispersion process. Olsson et al. [24] showed through experiments in a large-scale unit that as long as the cross-section allows a sufficient number of mixing cells, the lateral mixing of fuel particles at the macroscopic scale can be described in terms of a dispersive pattern. The availability of a sufficient number of mixing cells is a reasonable assumption for fluidized beds with low aspect ratios. Using the dispersive approach, the lateral mixing of solids at a macroscopic level, i.e., on a spatial scale greater than the length of the mixing cells, can be characterized by a single dispersion coefficient, D:

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

Thus, the dispersive approach combines the contributions from convection and diffusion with the mixing into the lateral dispersion coefficient, D, which can be determined experimentally [15] or through simulations [25]. However, this type of macroscopic description of lateral mixing of solids does not resolve the process of mixing within the above-mentioned mixing cells.

Olsson et al. [24] defined a cell mixing factor, α, which quantifies the fraction of solids inside a mixing cell that remains within the same mixing cell after a bubble cycle. However, the present work instead applies a 'bubble mixing factor', β, which is defined as the fraction of solids that leaves the original mixing cell after a bubble cycle; this is more representative for expressing the mixing, since, as opposed to α, high values of β yield high mixing rates:

$$\beta = 1 - \alpha = \frac{2 \cdot D}{f \cdot L^2} \tag{2}$$

In Eq. (2), f is the characteristic bed frequency, which can be obtained from pressure fluctuation measurements, and L is a characteristic length of the mixing cell, which in the present work is evaluated through visual observations of bubble burst locations at the dense bed surface. Note that Olsson et al. [24] applied the concept of cell mixing factor derived to fuel mixing, while the present work focuses on the mixing of bed material by means of the bubble mixing factor. The bubble mixing factor accounts for the bubble flow which induces solid mixing, since the characteristic bubble frequency of the bed is included in the expression of the bubble mixing factor. Note that the bubble mixing factor is used to describe the lateral solid mixing on a macroscopic level, which is the scope of this work, and therefore the solid mixing within each mixing cell is not resolved.

Several experimental techniques for studying solid mixing in fluidized beds have been developed [26] and their main advantages and disadvantages are summarized below. Although the in-bed flow of a single tracer particle can be assessed using techniques based on X-rays or γ-rays, these measurement techniques are expensive and have high safety requirements [26], and they do not confer reasonable resolution when metal powder is used as the bed material (which is typically the case in fluid-dynamically down-scaled tests using combustion or gasification units). For flows that are confined to 2-dimensional geometries, Particle Image Velocimetry (PIV) provides detailed information on the particle flow and has been employed to investigate different solid mixing mechanisms, e.g., wake mixing [19], splashing [16,17], and emulsion drift [18]. Solid mixing can also be studied by indirect methods, which provide limited information and no data on velocity fields but are simpler, inexpensive, and sufficient when the goal is to characterize the solid mixing on a macroscopic level using a dispersion coefficient. Examples of indirect methods described in literature are: the use of sublimating CO_2 solids as a tracer [27,28]; and the sampling of bed material at several positions inside the bed *via* tubes [29]. The present work applies a method based on the sampling of an inert tracer using magnetic separation.

Fluid-dynamically down-scaled fluidized bed units make it possible to carry out studies at the ambient temperature with simpler and more flexible operation and with direct visual access to the solid flow, while maintaining similar fluid-dynamics as in a fluidized bed operating at high temperature. Glicksman [30] and Horio et al. [31] have proposed different sets of rules for fluid-dynamically scaling fluidized beds. The scaling laws proposed by Glicksman [30] can be simplified for beds that are operated with either small particles at low velocity (viscous regime) or large particles at high velocity (inertial regime). Furthermore, Glicksman et al. [12] showed that the two sets of scaling laws proposed by Horio et al. [31] and Glicksman [30] are equivalent for bubbling fluidized beds operating in a viscous regime (i.e. for particle Reynolds numbers <4). The scaling laws proposed by Glicksman et al. [12] have been experimentally validated in several studies [32–34]. In the scaling rules proposed by Glicksman et al. [12], the characteristics of the gas distribution system are not considered. However, to achieve dynamic similarity between two fluidized beds, the same fluidization regime must be maintained in the two systems. This is achieved by having the same ratio of gas distributor pressure drop to bed pressure drop in