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Research article

Investigating wall-to-bed heat transfer in view of a continuous temperature swing adsorption process

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ABSTRACT

Heat transfer between bubbling fluidized beds and immersed heat exchanger surfaces is studied in view of continuously operated temperature swing adsorption processes for post-combustion CO_2 capture. A novel heat transfer measurement test device was used to measure wall-to-bed heat transfer coefficients. The present work focuses on the comparison of experimentally obtained and calculated heat transfer coefficients. Heat transfer at horizontal single tubes and tube bundles immersed in fluidized particle beds of glass beads with 140 µm and 200 µm in Sauter mean diameter is investigated. It is shown that the experimental results for single tubes are in mediocre agreement to established mathematical models, such as the ones proposed by Natusch et al. (1975) and Molerus et al. (1995), and that heat transfer is significantly influenced by the tube diameter. The model by Petrie et al. (1968) was considered to take the effect of the tube diameter into account, which lead to promising results. Furthermore, measured heat transfer coefficients at tube bundles of different geometries are compared to predictions using the models by Natusch et al. (1975) and Lechner et al. (2013). Some of the tube bundle reduction factors predicted by the model by Lechner et al. (2013) are larger than one, which stands in contrast to the finding that the highest heat transfer coefficients occur at single tubes. However, both models lead to adequate results when calculating heat transfer coefficients for different tube bundle geometries.

1. Introduction

Previously conducted studies have shown that heat transfer has a dominant or even limiting effect on the CO_2 capture performance of continuous temperature swing adsorption (TSA) processes [1,2]. For reasons concerning the overall process economy it may be necessary to achieve shallow bubbling beds with minimized pressure drop across the reactor stages. The pressure drop across a fluidized bed is known to be practically constant in the range of U_{mf} (minimum fluidization gas velocity) to U_t (terminal gas velocity). However, the fluidization gas velocity has a major influence on the bed expansion and, thus, on the available space for the placement of in-bed heat exchangers. Concurrently, the fluidization rate affects the achievable heat transfer coefficient between the immersed heat exchangers and the fluidized particles. The particles considered for the application in the TSA process are of Geldart Type B.

In general, it is recognized that there are three mechanisms of heat transfer between a fluidized bed and immersed heat exchanger surfaces – namely (1) particle convection, (2) gas convection and (3) radiation. Due to the relatively low temperatures occurring in the TSA process radiation may be neglected [3]. In most dense gas-solid fluidized beds solids mixing is the primary cause for the particle convective heat transfer [4]. Thus, heat transfer coefficients are low at low superficial gas velocities, where particles are in the state of a fixed bed. With an increase of the gas velocity and the formation of bubbles the increase in particle movement results in a sharp rise of the heat transfer coefficient until a characteristic maximum is reached, as shown in Fig. 1. A further increase in gas velocity yields to a decrease in heat transfer, which may be pronounced to a greater or lesser extent. The reason for the described behavior is based on the alternation between the particle and gas convective heat transfer [4]. The particle convective heat transfer reaches a maximum at the optimal gas velocity U_{opt}, whereas the gas convective heat transfer increasingly takes on greater significance beyond this point.

For relatively deep bubbling fluidized beds with a height/diameter-ratio larger than one and Geldart Type B particles it is widely accepted that the solids flow occurs in upward movement in the bed center. This behavior is explained by the growth and coalescence of bubbles while they are rising. Concurrently, a downward flow is observed at the walls enclosing the bubbling bed. We may conclude that a certain lateral

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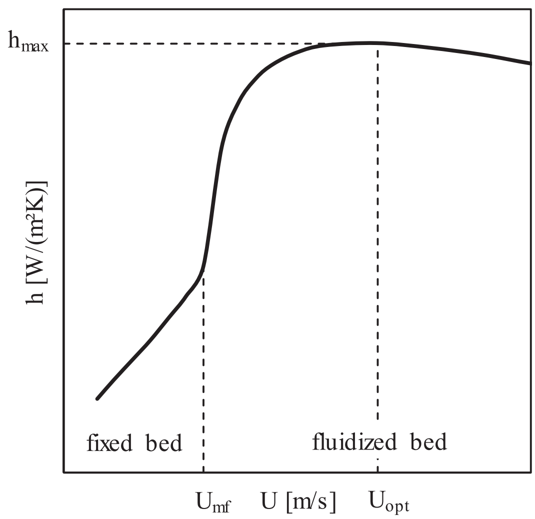

Fig. 1. Typical behavior of wall-to-bed heat transfer as a function of gas velocity in bubbling fluidized beds of Geldart Type B particles.

distribution of axial particle velocities exists. With an increase in gas velocity the lateral particle velocity becomes more uniform [5]. The described pattern of the solids flow is also observed while heat exchanger surfaces, i.e. tube bundles, are immersed in emulsified particle beds [5,6]. Yao et al. [7] correlated all the described phenomena to the packet renewal model, which was published first by Mickley and Fairbanks in 1955 [8]. Furthermore, we conclude that heat transfer may be not uniform across the fluidized beds cross-section. However, this phenomenon is not investigated in this study.

In a qualitative manner, the immersed tubes have an influence on bubble growth and, concurrently, on solids mixing. Rüdisüli et al. [6] conducted experiments to examine the lateral bubble distribution in fluidized beds with immersed vertical tubes. In a quantitative manner, they reported that the number of bubbles decreases with increasing bed height if there is no bundle of tubes present. This can be attributed to the typical coalescence of bubbles. It is mentioned that the number of bubbles remains almost constant over bed height if a tube bundle is inserted to the bubbling bed. Rüdisüli et al. [6] concluded that '*bubble coalescence in beds with vertical tubes is either inhibited or compensated by more frequent bubble splitting*'.

In the past, a great number of experimental and computational studies on wall-to-bed heat transfer were conducted in fixed beds, fluidized beds in bubbling and turbulent regime as well as in circulating fluidized beds. Lately, Merzsch et al. [9] and Lechner et al. [10] have contributed an extensive literature and empirical research concerning the influence of tube diameter, tube bundle arrangement and particle moisture on bubbling bed heat transfer for Geldart Type A and C bulk materials.

It is summarized that there are numerous models relevant for single tubes, such as those introduced by Zabrodsky [11], Noack [12], Mathur et al. [13], Kunii et al. [14] or Stefanova et al. [5] – to name just a few. Most of the available models are designed to estimate maximum heat transfer coefficients for a certain optimum fluidization gas velocity. Effectively, just a couple of models are able to describe the overall fluidized bed heat transfer coefficient as a function of superficial gas velocity, which is of special importance when it comes to dimensioning heat exchangers for TSA. As to that, the models of Natusch et al. [15], Molerus et al. [16] and Martin [17] may be of special interest.

However, the mentioned models disregard the influence of the tube diameter on heat transfer. Natusch et al. [15] reported that – according to Vreedenberg [18], Gel'perin and Einstein [19] and Zabrodsky [11] – the influence is negligible for tube diameters larger than 15 mm. Nevertheless, Petrie et al. [20], Molerus et al. [21] as well as other authors have shown that the tube diameter has significant influence on heat transfer and suitable models were developed.

Regarding heat transfer at tube bundles immersed in fluidized beds, many models were introduced by different authors. Some of the proposed tube bundle models lead, in analogy to single tube models, to maximum heat transfer coefficients correlated to optimum gas velocities. However, other models are designed to calculate so-called tube bundle reduction factors [10,15], which are applicable multiplicatively to available single tube models. The expression 'tube bundle reduction factor' implies, that heat transfer is reduced by the implementation of in-bed tube bundle heat exchangers. Hence, heat transfer coefficients decrease at some point, i.e. if the tube spacing is low enough. This leads to the conclusion that particle movement is hindered by the placement of in bed heat exchangers.

This work will investigate and discuss the difference between

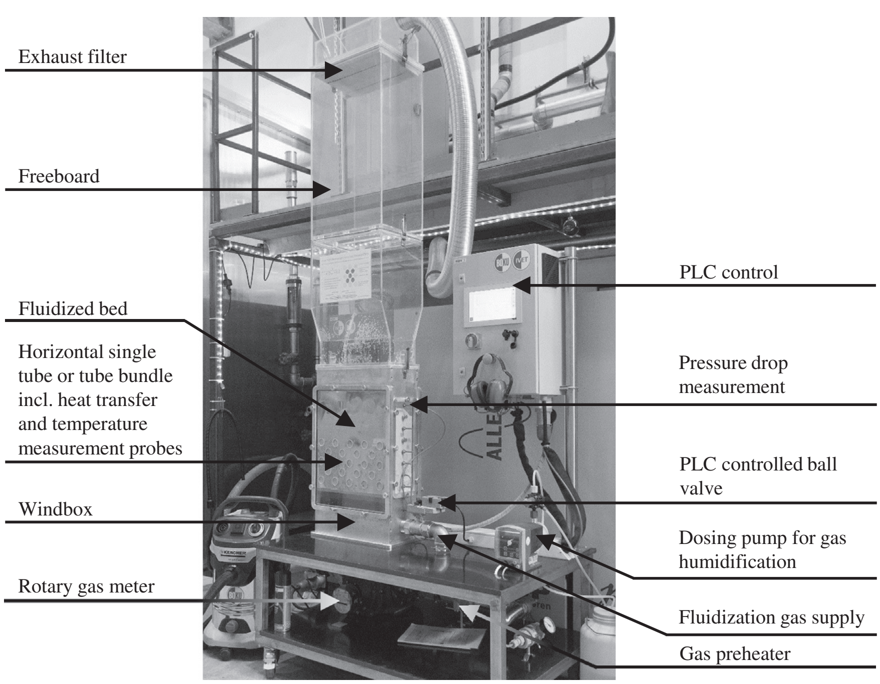

Fig. 2. Heat transfer measurement test device (HTMT).