Powder Technology xxx (2016) xxx–xxx

Contents lists available at ScienceDirect

Powder Technology

journal homepage: www.elsevier.com/locate/powtec

Heat transfer challenge and design evaluation for a multi-stage temperature swing adsorption process

G. Hofer [a,*], J. Fuchs [b], G. Schöny [b], T. Pröll [a]

[a] University of Natural Resources and Life Sciences, Department of Material Sciences and Process Engineering, Peter-Jordan-Str. 82, 1190 Vienna, Austria
[b] TU Wien, Institute of Chemical Engineering, Getreidemarkt 9/166, 1060 Vienna, Austria

ARTICLE INFO

Article history:
Received 25 April 2016
Received in revised form 28 October 2016
Accepted 19 December 2016
Available online xxxx

Keywords:
Temperature swing adsorption
TSA
Carbon capture
Heat transfer
Fluidized bed
Heat exchanger design

ABSTRACT

Functionalized solid amine-based temperature swing adsorption (TSA) processes have recently been proposed as a potential way to reduce the energy-penalty of post-combustion carbon capture processes. Thereby, multi-stage fluidized bed contactors with immersed heat exchanger surfaces and counter-current flow of solids and gas phase may solve the heat transfer challenge while maintaining the thermodynamic process requirements. Hence, the present work develops design requirements for TSA stages based on achievable heat transfer rates in bubbling fluidized beds. The considered particles are Geldart Type B. It is shown that the pressure drop of multi-stage fluidized bed TSA units for flue gas CO_2 capture is practically determined by the heat exchange requirement. Scalability, maintainability and durability of different heat exchanger geometries are considered. The net movement and mixing of particles within the bubbling bed stage must be maintained in spite of the immersed heat exchangers concerning possible dead zones in the area of the tube bundles. Comprehensive models are used to predict heat transfer coefficients for tubes immersed in fluidization. A heat transfer measurement test device for optimization of the heat exchanger geometry has been put into operation and heat exchange measurement results are compared to calculated heat transfer coefficients. It is shown that experimentally obtained heat transfer rates for single tubes are in good agreement with modeled values. A model proposed for Geldart A particles is used to estimate heat transfer rates for two particular tube bundles with a tube diameter of 25 mm and horizontal tube spacing of 2.2 and 2.8. It is shown that the calculated results represent heat transfer rates qualitatively and quantitatively for tube bundle heat exchangers immersed in Geldart Type B particle fluidized beds. Although this article has been motivated by heat exchange in TSA, it may be of interest for other applications concerned with heat transfer between bubbling fluidized beds and immersed heat exchanger surfaces.

1. Introduction

A double loop, multi-staged fluidized bed system operated with functionalized solid amine sorbents, by means of temperature swing adsorption (TSA), has been introduced for continuous post-combustion CO_2 separation tasks [1,2]. In this process configuration, staged bubbling fluidized bed columns are used for the adsorber and desorber, respectively. For continuous operation it is necessary to extract the heat of adsorption from the adsorber and to supply about the same amount of heat into the desorber. In addition, the sensible heat caused by the temperature swing has to be transferred. Thermodynamically, the contactors could be favorably designed as fixed or moving beds. However, heat transfer rates are poor for fixed and moving bed regime and heat transfer is crucial in TSA [3]. Hence, fluidized bed contactors with counter-current flow of solids and gas may solve the heat transfer challenge while maintaining the thermodynamic process requirements.

On the one hand, it is important to achieve high heat transfer coefficients in order to limit the pressure drop across the adsorber column, because of the major influence on blower power consumption. On the other hand, high heat transfer rates are equally important with concern to the desorber. The restriction on stripping steam demand for desorption has limiting influence on fluidization and, therefore, on fluidized bed volume to accommodate heat exchanger geometries. Furthermore, with increasing pressure drop difference between adsorber and desorber it gets harder to seal them against each other, and the possibility of purging supply lines increases. Since adsorption kinetics are known to be fast and mass transfer is efficient in fluidized beds, we formulate the hypothesis that the required heat exchanger surface will determine the dimensions of the fluidized bed stages. This would mean that the practically achieved heat transfer rates, the resulting compactness of the heat exchanger bundles and the operational expenditures would determine the overall costs of such a continuous TSA device.

* Corresponding author.
E-mail address: gerhard.hofer@boku.ac.at (G. Hofer).

http://dx.doi.org/10.1016/j.powtec.2016.12.062
0032-5910/

Fig. 1 shows the principle of the double loop fluidized bed system including the relevant heat exchangers in the adsorber and desorber reactor columns featuring five stages each for efficient CO_2 separation with resulting capture efficiencies up to 90% or more. The top-down moving sorbent particles are fluidized by introducing raw exhaust gas at the bottom of the adsorber column. While contacting the sorbent in counter-current flow, CO_2 is progressively removed from the flue gas. After the separation process, when the rich sorbent reaches the bottom of the adsorber, these particles are lifted through a riser system to the desorber, that is operated at higher temperatures. For regeneration, stripping steam is used to fluidize the top-down streaming sorbent in the desorber. On top of the desorber, a gas mixture containing steam and CO_2 is obtained. In order to obtain pure CO_2 the steam is condensed downstream of the desorber. To close the particle circulation loop, the lean sorbent is lifted from the bottom of the desorber column to the adsorber for further CO_2 separation.

As mentioned previously, heat exchange is expected to be the dominant limiting factor when carrying out TSA. Therefore, the present work focuses on the application of established heat exchange calculation methods for immersed surfaces in bubbling fluidized beds; in particular single tubes [6–9] and tube bundles [10–14]. Previously defined TSA process design parameters [1,2,4] are considered as input data. Furthermore, the calculated results are used to develop an awareness for practical heat exchanger design possibilities. In spite of already accomplished investigations regarding heat exchange in bubbling fluidized beds, a heat transfer measurement test device (HTMT) has been deployed to conduct heat exchange measurements at immersed single tubes and, preferably, tube bundles in various settings. It is shown, that the theoretically achievable heat transfer coefficients calculated with the model proposed by Lechner et al. [19], valid for Geldart Type A particles only, is also able to predict coefficients for Geldart Type B bulk material.

2. Heat exchange in TSA

2.1. Prediction of necessary heat exchanger surface area

Based on Fourier's law the heat flow Q to be transported in the adsorber and desorber can be written as

$$Q = h \cdot A_{hex} \cdot \Delta T_m \tag{1}$$

where h is the overall heat transfer coefficient applied to the outside diameter d_o of the heat exchanging tube

$$h = \left[\frac{d_o}{d_i} \cdot \frac{1}{h_i} + \frac{d_o}{2 \cdot \lambda} \cdot \ \ln\left(\frac{d_o}{d_i}\right) + \frac{1}{h_{fb}}\right]^{-1} \tag{2}$$

For thin tube walls with high heat conductivity λ and turbulent flow of liquid in the tubes we may assume

$$\frac{d_o}{2 \cdot \lambda} \cdot \ \ln\left(\frac{d_o}{d_i}\right) \ll \frac{d_o}{d_i} \cdot \frac{1}{h_i} \approx \frac{1}{h_i} < \frac{1}{h_{fb}} \tag{3}$$

and thus

$$h \approx \left[\frac{1}{h_i} + \frac{1}{h_{fb}}\right]^{-1} \tag{4}$$

Assuming that the reactor design has a rectangular cross sectional area with the lengths a and b and, furthermore, 100% of this area is used to accommodate heat exchanger tubes, the total heat exchanging surface area A_{hex} can be estimated by

$$A_{hex} = \frac{a \cdot b \cdot \pi}{d_o \cdot s_h \cdot s_v} \cdot H_{fb} \tag{5}$$

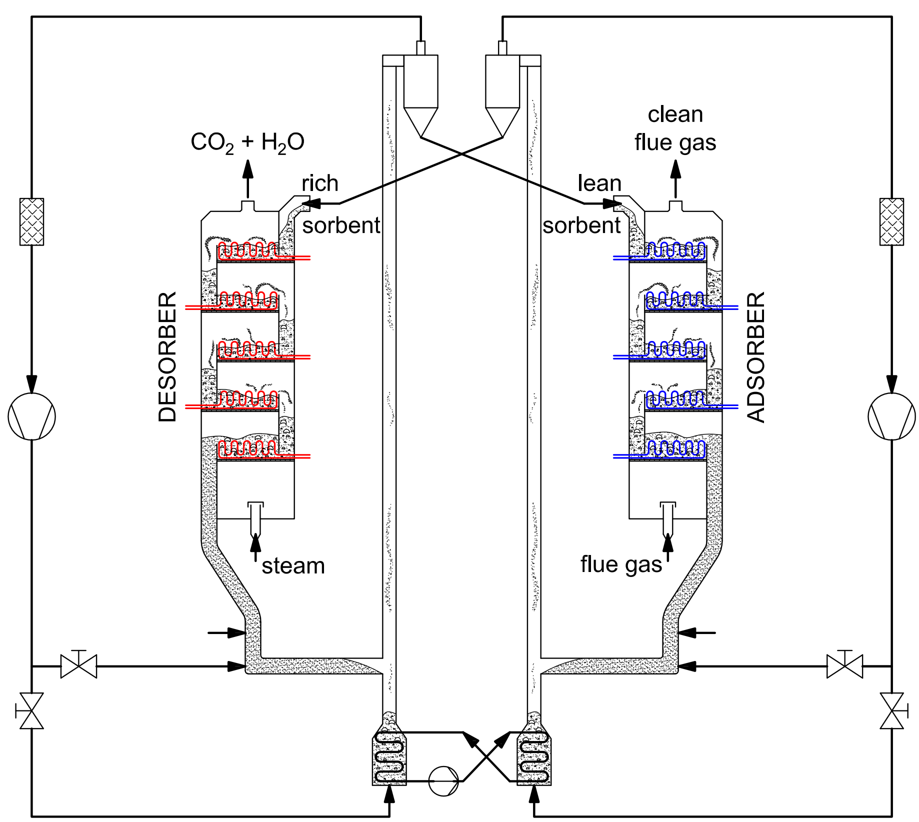

Fig. 1. Principle of the continuous TSA CO_2 separation process with relevant heat exchange requirements featuring five stages in the adsorber and desorber (blue = cooling requirement, red = heating requirement). (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)

Please cite this article as: G. Hofer, et al., Heat transfer challenge and design evaluation for a multi-stage temperature swing adsorption process, Powder Technol. (2016), http://dx.doi.org/10.1016/j.powtec.2016.12.062