PARAMETER VARIATION IN A MULTI-STAGE FLUIDIZED BED SYSTEM FOR CONTINUOUS CO_2 CAPTURE BY MEANS OF TEMPERATURE SWING ADSORPTION

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Abstract – Most recently, a new reactor design of a continuous temperature swing adsorption (TSA) CO_2 capture process has been proposed by the authors of this work. The reactor design incorporates interconnected multi-stage fluidized bed columns that act as adsorber and desorber in the TSA process. The feasibility of the proposed CO_2 capture process and the underlying reactor design was already demonstrated in a fully integrated TSA bench scale unit. In the course of this work, a comprehensive parameter variation was performed in the TSA bench scale unit, to study the sensibility of the process performance on key operating parameters. Results show that increasing sorbent circulation rates and decreasing operating temperatures in the adsorber have a beneficial effect on the CO_2 capture rate. The flexibility of the process in respect to different CO_2 concentrations of the treated flue-gas as well as different adsorber feed gas velocities was also investigated. Increasing the CO_2 concentration of the adsorber feed gas increases the driving force for CO_2 adsorption and leads to an increase of the dynamic CO_2 loading of the adsorbent. However, for higher CO_2 feeding rates, the capture performance dropped as a result of insufficient heat transfer surface area and limitations in the achievable sorbent circulation rates. In conclusion, the parameter study revealed a high operating flexibility of the TSA CO_2 capture process and delivered a valuable basis for further improvements of the TSA reactor design.

INTRODUCTION

In the wake of increasing CO_2 levels within the Earth's atmosphere and increasing public awareness of climate change, a great global effort has been attributed to the development of CO_2 abatement measures, especially for electricity generation and industrial processes. Recently, the IEA (2013) proposed a portfolio of technologies that need to be deployed towards 2050 in order to limit the global temperature increase to 2°C as compared to pre-industrial levels. Within this portfolio, carbon capture and sequestration (CCS) plays a significant role, since it is the only suitable CO_2 mitigation measure for large point sources that will continue to utilize fossil fuels in the decades to come. For instance, post-combustion carbon capture technologies (PCCC) allow for removing CO_2 from flue-gas streams that are emitted by coal or gas fired power plants or by processes in the cement or steel industry. The captured CO_2 can than either be utilized as a chemical or sent to a proper storage site to prevent it from being active as a greenhouse gas in the atmosphere.

The currently considered state-of-the-art PCCC technology is amine scrubbing which has been in development since the 1930's. However, the predicted CO_2 capture costs for amine scrubbing processes are comparably high which motivated the development of alternative PCCC technologies. In the past years, it was proposed that temperature swing adsorption (TSA) processes have a great potential to significantly decrease the energy requirement and costs of the CO_2 capture step as compared to amine scrubbing systems (Samanta et al., 2012). This is expected since solid sorbents are utilized instead of an aqueous amine solution and thus, water evaporation in the regeneration step can be avoided completely. Furthermore, more compact reactor designs may be feasible for TSA systems due to faster kinetics and the large specific surface areas of the sorbent materials.

Given the scope of the separation task in PCCC processes (i.e. CO_2 is separated from a diluted gas stream and a highly pure CO_2 product is desired), the CO_2 selectivity of the utilized sorbent plays a central role in the development of TSA processes. For this reason, researchers have established a new class of highly selective chemical sorbents, by immobilizing amines on high-surface-area solid supports (Xu et al., 2003; Gray et al., 2005; Drage et al., 2008; Choi et al., 2009; Sjostrom et al., 2010; Alesi et al., 2012). The amine compounds on the surface of the sorbents chemically react with the CO_2 present in the flue-gas. Hence, these type of sorbent materials feature a high selectivity towards CO_2 as well as a strong temperature dependence of the CO_2 equilibrium loading and are therefore considered as well suited for application in TSA processes.

The chemical reaction with CO_2 also implies that significant amounts of heat need to be exchanged during the adsorption and desorption step. Consequently, it is essential to apply a reactor system and a gas-solid contacting regime that allows for achieving high heat transfer rates between heat exchanger surfaces and the sorbent material. The fluidized bed operating regime is well-known for its superior heat transfer properties and has thus been the suggested optimum contacting regime for the adsorber and desorber of a continuous TSA process (Krutka et al., 2013; Pröll et al., 2016). However, up to the present only little work has been dedicated to the development of suitable TSA reactor designs.

An example for recent research dedicated to fluidized bed reactor systems used for a TSA process is the dual fludized bed system investigated by Zhao et al. (2013). Here, the adsorber and the desorber were both designed as bubbling fluidized beds. The adsorber and the desorber were fluidized by a N_2-CO_2-mixture and N_2 respectively. Furthermore, solids transport between the reactors was facilitated by a riser, a cyclone and two loop-seals. In a process feasibility test of the dual fluidized bed reactor system, a capture efficiency of around 84.4% could be achieved at a rather low dynamic sorbent working capacity of about 3,5%$_w$. The rather low dynamic sorbent loading is the result of thermodynamic limitations associated with the excellent mixing behavior in the fluidized bed. Consequently, the sorbent CO_2 loading is at best in equilibrium with the outlet CO_2 partial pressure. Hence, in such systems a high CO_2 capture efficiency comes at the cost of relatively low CO_2 loadings of the sorbent material exiting the adsorber. As shown by Pröll et al. (2016) this thermodynamic limitation can be overcome by implementing counter-current contact of gas and solids. In an adsorber consisting of multiple fluidized bed stages, the lean sorbent material enters at the top and is gradually loaded with CO_2 as it moves downward, allowing for a higher CO_2 loading of the sorbent at the adsorber solids outlet compared to a single stage adsorber. In a similar way, counter-current flow of gas and solids can promote CO_2 desorption in a desorber when a stripping agent (e.g. N_2 or steam) is used as a fluidization gas. Krutka et al. (2013), Nelson et al. (2014) and Veneman et al. (2016) have investigated TSA reactor systems where multi-stage fluidized beds are being implemented. In work carried out by Krutka et al. (2013), a multi-stage fluidized bed reactor with three stages was implemented in the adsorber. The desorber was designed as a single bubbling fluidized bed since CO_2 served as a fluidizing agent. In work published by Nelson et al. (2014), the adsorber as well as the desorber each are designed as columns consisting of three fluidized bed stages. Veneman et al (2016) investigated a TSA process configuration comprising a gas-solids trickle flow adsorber and a multi-stage fluidized bed desorber where N_2 was used as a stripping agent. Pröll et al. (2016) introduced a TSA reactor system comprising two multi-stage fluidized bed columns and two solids transport lines which connect the bottom of each column with the top of the adjacent column. Each of the columns consists of five bubbling fluidized bed stages and solids transport from stage to stage is facilitated by internal downcomers. Furthermore, the solids transport lines interconnecting the columns consist of an L-Valve located at the solids outlet of each column, a solids transport riser and a gravitational gas-solids separator, which directs solids into the top stage of the adjacent column.

A bench scale unit based on the underlying continuous TSA process was designed, constructed and put into operation. An initial test of the unit delivered satisfactory results as well as a proof of concept (Schöny et al., 2016b). N_2 was used as a stripping agent and a capture efficiency of over 90% could be achieved from a flue-gas containing 5,1vol% of CO_2. It could be shown that the CO_2 capture performance was limited by insufficient heat transfer surface in the fluidized bed stages. Furthermore, the absence of CO_2 in the adsorber off-gas within the initial minutes after the introduction of CO_2 into the adsorber feed-gas showed that mass transfer as well as adsorption kinetics were sufficiently fast under the prevailing operating conditions. Nevertheless, steam which is the designated stripping agent had not yet been tested and the sensitivity of the CO_2 capture performance towards changes of relevant operating conditions was largely unknown. The aim of this work is to demonstrate the flexibility of the continuous TSA process by conducting parameter variations in the existing TSA bench scale unit and to establish dependencies between operating conditions and the CO_2 capture performance of the unit.

EXPERIMENTAL SETUP AND METHODS

The TSA bench scale unit

Fig. 1 shows a basic scheme of the TSA bench scale unit (BSU). Both, the adsorber and the desorber consist of five consecutive fluidized bed stages that in turn feature a perforated plate type gas distributor, an exchangeable weir plate, a downcomer that directs sorbent material into the next lower stage and a heat exchanger. The heat exchangers are immersed in the bubbling fluidized beds formed above the gas distributors in each stage and are used to extract and provide heat from the adsorber and to desorber, respectively. A water-glycol mixture is used as coolant for the heat-exchanger coils in the adsorber stages.

Both, the coolant forward flow temperature as well as the flow rate to the individual adsorber stage heat exchangers can be adjusted. This allows for good control of the adsorber stage temperatures during continuous CO_2 capture tests. In the desorber stage heat exchangers thermal oil is used as a heating medium. Unlike in the adsorber, the heating circuit does not exhibit a manifold to adjust the thermal oil flow rate to the individual desorber stages. Instead, the thermal oil flows remain constant during operation and the heat input to the desorber can only be controlled by adjustment of the thermal oil forward flow temperature. As can be seen from Fig. 1, any mixture of CO_2, N_2, air and steam can be used as simulated flue-gas mixture to fluidize the adsorber, whilst N_2, steam or any mixture thereof can be used as stripping and fluidizing agent in the desorber. Sorbent transport between the adsorber and the desorber is facilitated by two individual transport lines. In both transport lines, L-valves are used to control the circulation rate of sorbent material. During normal operation, the L-valve below the adsorber column is operated with constant aeration to achieve the desired sorbent circulation rate (SCR). The aeration of the L-valve located below the desorber column is controlled to maintain a constant distribution of sorbent material between both columns, which is indicated by the measured pressure drops over the individual columns. Both L-valves carry the sorbent material into a transport riser that lifts the sorbent material to the top of the columns. At the top of each riser, the sorbent material enters a gas-solids separator and is separated from the transport gas of the riser and further directed into the top stages of the columns. The transport gas exits the gas-solid separator at the top and gets recycled to the bottom of the transport risers. For the sake of simplicity, the transport line transporting lean solids from the desorber to the adsorber will henceforth be called “lean transport line” whereas the respective riser in this line will be called “lean riser”. Conversely, the other transport line and riser will be called “rich transport line” and “rich riser”, respectively. During the experiment, the SCR can be calculated from the pressure drop over the rich riser. To prevent gas slip from the columns into the transport lines, small amounts of purge gas (“7” in Fig. 1) can be introduced into the moving bed sections between both columns and the transport lines. More detailed information on the BSU process setup and design parameters (including the geometry of the columns and the heat exchanger coils) as well as on the start-up procedure, can be found elsewhere (Schöny et al., 2016a; Schöny et al., 2016b). The measurement equipment applied to the BSU should, however, be briefly presented in the following.

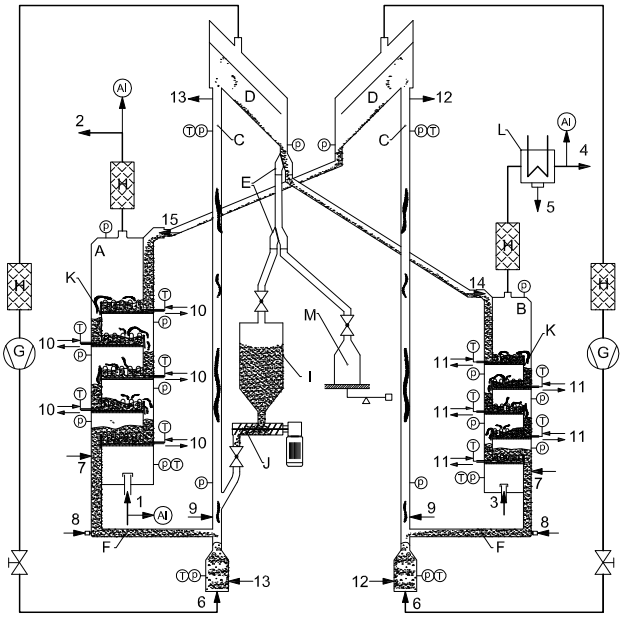

Fig. 1: Configuration of the bench scale unit for continuous temperature swing adsorption CO_2 capture from simulated flue gases.

Instrumentation

The gas flow rates of N_2, air, and CO_2 used to fluidize the adsorber, are controlled by three mass flow controllers (MFCs) of the type “Bürkert 8746” with the nominal ranges of 10-500Nl/min, 10-500Nl/min and 1.6-80Nl/min respectively. Furthermore, “Bürkert 8741”-type MFCs are used to control the N_2-aeration of both L-valves, which have a nominal N2, flow rate of 0.1-5Nl/min. All of the utilized MFCs have an accuracy of 0.3% of the full-scale. The steam flow rate to the adsorber and the desorber is controlled by a

configuration consisting of a needle valve and a variable area flowmeter. “Krohne H250 M40”-type flow meters are used for adsorber feed gas humidification and control of the stripping steam flow rate. These variable area flowmeters have a measuring range of 0.5-5m³/h and 1.7-17m³/h, respectively and an accuracy of 1.6%. Furthermore, the flow rate of the purge gas introduced below the desorber is controlled by needle valves and a Krohne DK800-type variable area flowmeters in the range of 0.032-0.32Nm³/h which has an accuracy of 2.5%. “Kalinsky DS2”-type pressure transmitters in the range of 0mbar to 10mbar up to 0mbar to 100mbar are used for pressure measurement at several locations. Differential pressure measurements over each stage as well as over a one-meter segment of each riser are measured using pressure transmitters in the range of 0mbar to 10mbar. The linear error of these pressure transmitters is up to 1% of the measuring range. “JUMO CANtrans T” temperature sensors in the range of -50°C to 150°C are used to measure the temperature of all fluidized beds in the system. These sensors are PT-1000-type resistance temperature detectors (RTD) and are in the tolerance class B according to DIN EN 60751. Coolant and thermal oil temperatures are measured using PT-100-type RTDs which are also in the tolerance class B of the DIN EN 60751 standard. The analyzers used to measure the inlet and outlet CO_2 concentration of the adsorber are “SIEMENS ULTRAMAT 23” NDIR-type analyzers. The measuring ranges of these analyzers are 0-25vol%CO_2 and 0-5vol%CO_2 and the linearity error of these analyzers is within 1% of the measuring range. In an effort to further reduce the error of the CO_2 analyzers, three different calibration gases in the range of 0-5.1vol%CO_2 are used before and after an experiment. Additionally, a paramagnetic O_2 analyzer, with a measuring range of 0-25vol%O_2, with a reproducibility limit below 1%, is used to monitor the inlet O_2 concentration of the adsorber. The solids circulation rate was typically measured after an experiment by directing solids into the set-up for SCR measurement (“M” in Fig. 1) for at least thirty seconds. Subsequently the weight of the accumulated solids was measured using a “KERN CKE 2000-2”-type scale. The accumulated mass of solids was then divided by the accumulation time to calculate the SCR. The average SCR over at least three repetitions of this procedure was then taken as the solids circulation rate for the corresponding operating point. In this procedure, care was taken to have pressure equalization between the container and the atmosphere, to avoid any build-up of pressure in the transport line that could impact the SCR.

Process performance evaluation

In PCCC the amount of captured CO_2 relative to the amount of CO_2 in the feed-gas is one of the most important performance parameters and is referred to as the capture efficiency (η_{capt}). The capture efficiency can be simply calculated when the inlet and outlet flow rates of CO_2 are known. However, since the current experimental setup does not allow for direct measuring the flow rate of CO_2 at the outlet of the adsorber, the capture efficiency must be calculated using the measured inlet and outlet CO_2 concentrations of the adsorber together with the flow rates of the gases that are assumed to not being adsorbed by the sorbent material (i.e. air and excess N_2). For the sake of simplicity, the purge gas flow introduced in the moving bed section of the desorber is assumed to completely end up in the adsorber off-gas. Furthermore, it is assumed that gas leakages from the adsorber into the rich transport line can be neglected. This assumption was confirmed by dedicated leakage tests performed prior to the presented experimental campaign. The assumptions given above make it possible to calculate the capture efficiency according to Eq. (1).

$$\eta_{capt} = 1 - \frac{y_{CO2,ads,out} \cdot (F_{N2,ads,in} + F_{Air,ads,in} + F_{N2,des,purge}) \cdot (1 - y_{CO2,ads,in})}{y_{CO2,ads,in} \cdot (F_{N2,ads,in} + F_{Air,ads,in}) \cdot (1 - y_{CO2,ads,out})} \tag{1}$$

In addition to the capture efficiency, the total CO_2 capture rate $\dot{m}_{capt}$ (given in kilograms of CO_2 captured per 24 hours) is used as a second process performance indicator. The capture rate was calculated according to Eq. (2).

$$\dot{m}_{capt} = 24 \cdot F_{CO2,ads,in} \cdot \eta_{capt} \cdot \rho_{n,CO2} \tag{2}$$

In continuous TSA processes it is important to achieve the desired CO_2 separation task at minimum sorbent circulation, since the sensible heating demand of the circulating solids contributes to the overall process energy demand. Since a minimum sorbent circulation rate in turn corresponds to a maximal exploitation of the sorbent’s working capacity, the “dynamic sorbent loading” (dSL) is used as third process performance parameter in this work. The dSl can be calculated by relating the CO_2 capture rate to the SCR as given in Eq. (3).

$$dSL = 100\% \cdot \frac{\dot{m}_{capt}}{24h/day \cdot \dot{m}_{sorb}} \tag{3}$$

The solids inventory of the BSU consists of approximately 7kg of an amine functionalized mesoporous sorbent. Over the course of the experimental campaign, different operating parameters were varied to assess their impact on the process performance parameters defined above. Base case operating conditions were defined prior to the experiments and the corresponding process performance parameters served as a benchmark for all other operating conditions during the parameter variations. In the base case operating conditions, the CO_2 concentration in the adsorber feed gas was set to 3vol%, similar to the CO_2 concentration found in the flue-gas of a gas turbine. Furthermore, the adsorber feed gas velocity was adjusted to around 0.3m/s and the stripping steam rate to around 8.5kg/h. The purge gas flow rate to the desorber moving bed section was 0.032Nm³/h of N_2 and the thermal oil forward flow temperature for the desorber heating was set to 110°C. The cooling of the adsorber column was controlled to reach an average column temperature of approximately 50°C-55°C.

Starting from these operating conditions the CO_2 concentration in the adsorber feed gas was varied to mimic flue-gases from different sources. The lowest point, with a CO_2 concentration of 3vol% would mimic the off-gas from a gas turbine, whereas 12vol%CO_2 would represent the off-gas from coal combustion. Subsequently, the adsorber feed gas velocity was varied to demonstrate the flexibility of the TSA process with regard to changing loads of a CO_2 emitting process. Here, the adsorber feed gas velocity was varied between 0.32m/s and 0.68m/s at a constant CO_2 concentration of 3vol%CO_2. In the last two variations the sensitivity of the capture performance was investigated by varying the average adsorber stage temperature and the sorbent circulation rate. It is worth mentioning, that apart from the latter parameter variation, the SCR was adjusted under each operating condition to achieve the target capture efficiency of around 90%. Furthermore, for some of the presented operating points, it was not possible to reach the target capture efficiency since the SCR has an upper limit in the current BSU setup.

RESULTS AND DISCUSSION

Table 1 gives a summary of the results obtained from the base case operating point and the performed parameter variations. It is worth to mention that the given values are time-averaged over a time span of at least twenty minutes after reaching steady-state conditions

Table 1: Summary of results gained from the variation of operating conditions

	Adsorber feed gas velocity	Adsorber feed gas CO_2 concentration	Average adsorber temperature	Sorbent circulation rate	Average desorber temperature	CO_2 capture efficiency	CO_2 capture rate	Dynamic sorbent loading
	[m/s]	*[vol%]*	*[°C]*	*[kg/h]*	*[°C]*	*[%]*	*[kg$_{CO2}$/day]*	*[wt%]*
Base-case	0.3	2.9	53.2	14,2	105.2	89.5	21,2	6,2
Variation of adsorber feed gas CO_2 concentration	0.3	7.9	52.1	36,7	100.6	91.2	58,4	6,6
	0.3	12.0	51.8	36,7	100.6	66.9	64,2	7,3
Variation of adsorber feed gas velocity	0.5	3.4	52.3	24,1	105.2	91.7	34,4	5,9
	0.7	3.3	51.4	38,8	103.0	90.8	47,4	5,1
Variation of average adsorber temperature	0.3	3.3	42.4	12,1	103.5	90.1	21,4	7,3
	0.3	3.0	58.3	16,3	107.3	90.9	21,6	5,5
Variation of the solids circulation rate	0.3	3.0	52.7	14,7	105.9	92.2	21,9	6,2
	0.3	3.1	52.3	11,6	106.3	73.4	17,4	6,3

In the following, the results gained from each parameter variation will be presented and discussed in more detail. The presented graphs show the column stage temperatures, the solids circulation rate as well as the process performance indicators as a function of the varied process parameter.

Variation of the sorbent circulation rate

As shown the temperatures in the columns stay similar despite changes in the solids circulation rate. In the adsorber this could be achieved by adapting the cooling duties in the corresponding fluidized bed stages. The graph showing the process performance as a function of the solids circulation rate, displays an improved capture efficiency and capture rate at increasing SCRs. However, a slight decrease in the dynamic sorbent loading was observed. High capture efficiencies correspond to low CO_2 partial pressures in the off-gas of the adsorber compared to the CO_2 partial pressure in the treated flue-gas. These low CO_2 partial pressures in turn correspond to a relatively low average loading of the sorbent material present in the uppermost stage of the adsorber. Since an increase of the sorbent circulation rates allows to achieve lower CO_2 loadings in the upper-most adsorber stage, the capture efficiency increases.

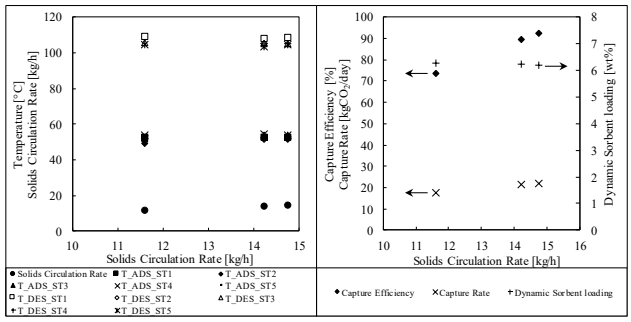

Fig. 2: Results from variations of the sorbent circulation rate

Variation of the average adsorber stage temperature

An increased temperature in the fluidized bed stages of the adsorber, reduces the maximum achievable CO_2 loading of the sorbent material in each stage and thus the rich sorbent loading at the adsorber solids outlet decreases. This in turn reduces the dynamic sorbent loading and increases the required SCR to achieve the separation task. However, the slight increase of the desorber temperature and the reduction in the axial temperature dispersion across the desorber column (see Fig. 3), indicate a lower sensible heat demand in the desorber despite a higher SCR required to reach a capture efficiency of 90%.

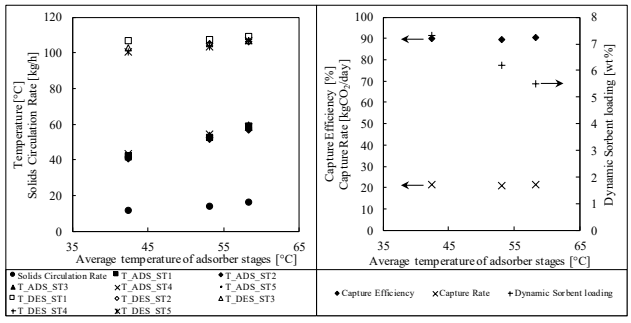

Fig. 3: Results from variations of the average adsorber stage temperature

Variation of the adsorber feed gas CO_2 concentration

It can be seen that the SCR of 36.7kg/h could not be surpassed when the flue-gas CO_2 concentration was above 8vol%. A reduction of the desorber stage temperatures at flue-gas CO_2 concentrations above 3vol% indicate an increased heat demand in the desorber and an insufficient heat transfer surface area in the desorber stages. Furthermore, an increase of the stage temperatures as well as of the axial temperature

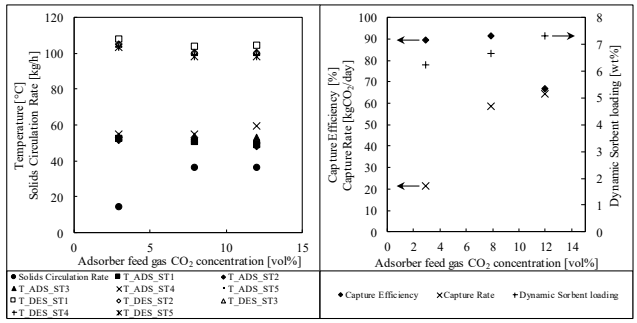

Fig. 4: Results from variations of the adsorber feed gas CO_2 concentration

dispersion in the adsorber show that the released heat could not be sufficiently extracted despite an increase of the coolant flow rates to the respective coils. Both of these effects can be explained by an increased latent and sensible heating and cooling demand as a result of a significant increase of the capture rate and SCR. It can be assumed that a further increase of the SCR at a flue-gas CO_2 concentration of 12vol% may have led to a further increase of the axial temperature dispersion in the adsorber and to a lower average temperature across the desorber. It can be seen that the dynamic sorbent loading steadily increases in the course of this variation. As the driving force for adsorption is the partial pressure of CO_2, an increase of the rich sorbent loading is expected. Since the dynamic sorbent loading increases at an increased CO_2 partial pressure in the adsorber feed gas, the rich loading undergoes a stronger increase than the lean sorbent loading. This indicates that the desorption step does not limit process performance at the operating conditions tested herein.

Variation of the adsorber feed gas velocity

To achieve the target capture efficiency of 90%, the solids circulation rate needed to be increased with increasing feed gas velocity. The increasing capture rates together with the increasing SCR had a similar effect on the desorber stage temperatures as during the variation of the CO_2 feed concentration. If a constant lean sorbent loading at the outlet of the desorber is assumed, the decrease of the dynamic sorbent loading indicates a reduced rich loading of the sorbent at increasing adsorber feed gas velocities. As the temperature in the adsorber could be kept in a narrow range, there are two possible explanations for a reduction in the dynamic sorbent loading. Either the contact time between solids and gas was too short to allow for sufficient CO_2 adsorption or a slip of gas through the bubble phase of the fluidized bed stages occurred at high adsorber feed gas velocities.

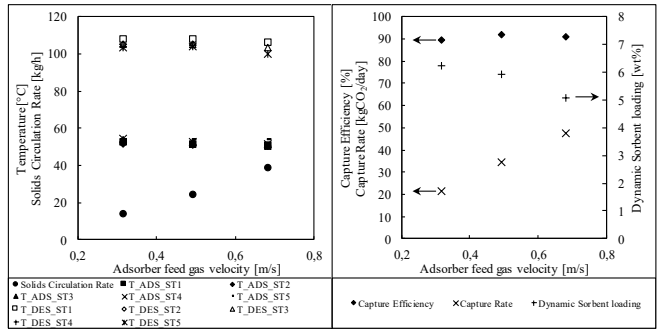

Fig. 5: Results from variations of the adsorber feed gas velocity

CONCLUSIONS

From the variation of operating conditions presented in this work, the following conclusions are drawn:

- Higher solids circulation rates improve the CO_2 capture performance at the cost of a lower dynamic sorbent loading.
- The CO_2 capture performance is particularly sensitive towards changes in the operating temperature of the adsorber, as this also impacts the maximum achievable CO_2 loading of the sorbent.
- A significant increase in the solids circulation rate is required to achieve a capture efficiency of 90%, when increasing the CO_2 concentration of the adsorber feed gas. Furthermore, an increase of the dynamic sorbent loading in the course of the CO_2 concentration variation of the adsorber feed gas indicates that the desorption step does not limit process performance at the operating conditions tested herein.
- An increase of the adsorber feed gas velocity at a constant CO_2 concentration entails a reduction of the dynamic sorbent loading, at a capture efficiency around 90%. This behavior indicates either insufficient contact time between solids and gas for CO_2 adsorption, or a slip of gas through the bubble phase of the fluidized beds in the adsorber.
- A maximum of approximately 64kg_{CO2}/day could be captured from an adsorber feed gas consisting of 12vol%CO_2 at a gas velocity of 0.32m/s in the adsorber. At this operating point, the capture efficiency could not be increased beyond 67% as the solids circulation rate in the current BSU setup has an upper limit.

In future work, long-term tests will be performed to investigate the stability of the utilized sorbent material as well as process emissions. Furthermore, an experimental assessment of the process energy demand will be carried out.

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NOTATION

η_{capt} capture efficiency, %
y_{CO2} volumetric CO_2 concentration, vol%
F flow rate at normal conditions, $Nm^{-3} \cdot h^{-1}$
$\dot{m}_{capt}$ CO_2 capture rate, $kg \cdot day^{-1}$
$\rho_{n,CO2}$ Density of CO_2 at normal conditions, $kg \cdot Nm^{-3}$
dSL dynamic sorbent loading, wt%
$\dot{m}_{sorb}$ adsorbent circulation rate, $kg \cdot h^{-1}$

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