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# Dynamic and control of an absorber - desorber plant at Heilbronn

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**Abstract**

The present work is based on an MEA absorption test campaign performed November 2013–January 2014 at EnBW's amine test pilot located in Heilbronn, Germany. The test campaign included transient step responses in steam, exhaust gas, solvent flow rate and in exhaust gas $CO_2$ concentration. In addition, a test with simultaneous steps in all flow rates (exhaust gas, solvent and steam) was included.

A dynamic model of the Heilbronn plant was then implemented in the dynamic simulator K-Spice® and tested against the step responses at the plant. In spite of some stead state deviations, the model was able to capture the process dynamics very well. For dynamic studies one can therefore assume that the model is representative for the plant. In the work presented here, two different control configurations were tested: 1) Ratio control in combination with a slow feed back control on the $CO_2$ out of the absorber and 2) Control of $CO_2$ out of the absorber by lean liquid flow rate and a temperature sensor up in the desorber packing. Both structures involve only simple PID control loops and are thus easy to implement. The proposed control configurations were tested by simulations with 30% changes in flue gas flow and composition. Both configurations showed good performance during the simulation testing, but the second one was superior as well as excellent with respect to tight $CO_2$ recovery control. This may be an important property in supervisory control schemes.


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*Keywords:* $CO_2$ capture; Pilot plant; Dynamic process modelling; Plantwide control; MEA

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## 1. Introduction

The present work was part of the EU project OCTAVIUS (Optimisation of $CO_2$ Capture Technology Allowing Verification and Implementation at Utility Scale, 2012-2016) which aimed to demonstrate integrated concepts for zero emission power plants covering all components needed for power generation as well as $CO_2$ capture and compression.

A post combustion capture plant has to be flexible and able to cope with various amount of flue gas and $CO_2$ concentrations. This is particularly the case with power plants operating at various loads during the day. At the same time is it important that the $CO_2$ capture process is as optimal as possible in terms of energy demand also under transient periods.

In this paper optimal control of an absorber/desorber amine capture plant located in Heilbronn, Germany is focused. Various step changes were performed at the plant during a campaign running on 30 weight percent mono ethanol amine (MEA). The plant responses were used to compare and validate a dynamic model of the pilot plant. The model has then been used to investigate possible control strategies for optimal control of the plant subjected to large and rapid flue gas changes.

## 2. Heilbronn pilot plant

The amine scrubbing pilot plant is located on site with EnBW's 1000MW coal fired combined heat and power plant in Heilbronn, Germany. The amine test plant extract about 1500 $Sm^3$/hr flue gas from the power plant stack after the wet FGD, where the flue gas typically contain 12-14% $CO_2$. With normal operation approximately 300 kg/hr $CO_2$ is captured. A simplified flow diagram of the pilot plant is given in Fig. 1.

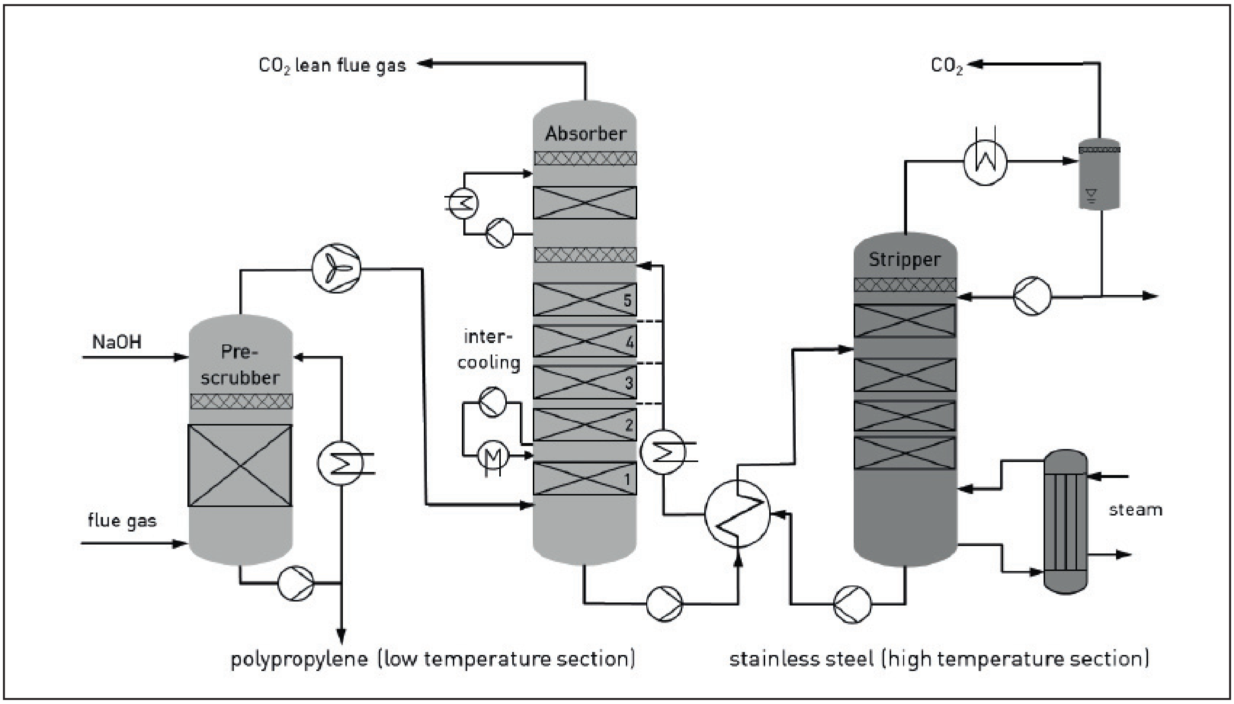


Fig. 1. Simplified flow diagram of the Heilbronn pilot plant

### 2.1. Plant description

The absorber is made of polypropylene and consist of 5 sections with random packing (~24 m total also polypropylene) and with a 0.60m ID, while the desorber is made of stainless steel containing 3 sections of random packing (15m total) and 0.60mID. Each column has in addition a packed section for the water wash. An intercooler is installed between the two lowest sections in the absorber column. Later modifications in the OCTAVIUS project have installed an additional water/acid wash at the top section of the absorber column.

First, the flue gas enters the pre-scrubber upstream the absorber for conditioning of the gas to correct temperature and water saturation. The pre-scrubber acts as a DCC (Direct Contact Cooler) where the outlet gas temperature is indirectly controlled by adjusting the recycle water temperature. Furthermore, an alkaline sodium hydroxide solution is added to the recycle water loop to reduce the $SO_2$ content to a minimum. A pH control is installed to control the absorption rate of $SO_2$ by adjusting the NaOH dosage to the recycle water loop.

After the pre-scrubber the gas is transported by a fan to the absorber column where the lean amine solution and the sour gas flows counter-current over the 5 packed sections (total packing height 24m). $CO_2$ is absorbed from the gas into the lean solution as the solution flow downwards over the packing. After the absorption sections, the gas is further conditioned to correct temperature and moisture in the absorber water-wash section.

Rich solvent is pumped from the absorber bottom/sump and preheated trough a cross-flow heat exchanger, with the hot lean solution from the reboiler, before it is fed to the top of the desorber section. Here, the regeneration of the solvent is aided by rising stripping steam and heat produced from a natural circulating thermosiphon reboiler at the bottom of the desorber. The resulting $CO_2$ rich stream in the overhead stream is purified by removing excessive water in a condenser.

The resulting lean solvent is pumped (lean pump) through the rich/lean exchanger where the hot lean solvent is cooled. A downstream cooler (lean cooler) lower the temperature further before the lean solvent enters the top of the absorber section again, ready for a new absorption cycle.

A more detailed description may be found in [1].

### 2.2. Basic control loops

A simplified P&ID with control loops is shown in Fig. 2 where the pre-scrubber and the absorber intercooling has been omitted. Important to note is that the lean and rich liquid lines has flow control, the absorber sump has level control, and the lean liquid temperature control of the lean trim cooler.

Dynamic modelling of the Heilbronn pilot requires an accurate representation of the PID control loops, hold up, equipment sizing etc.

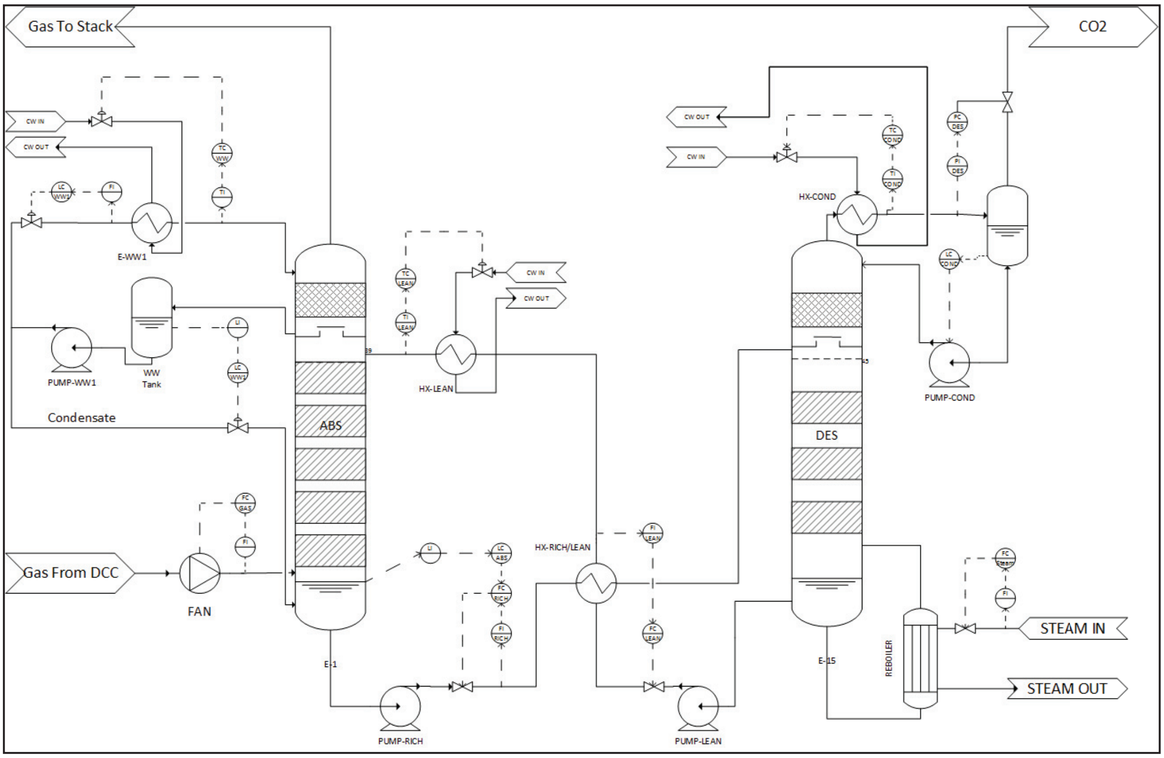


Fig. 2. Simplified process control scheme at the Heilbronn test plant (pre-scrubber and intercooler omitted.)

## 3. Dynamic process modeling with K-spice®

A dynamic model of the pilot plant was developed and implemented in the dynamic simulation program K-Spice® Simexplorer (Kongsberg Oil & Gas Technologies AS). K-spice® is an advanced dynamic process simulator mainly used for modelling and simulation of oil and gas processes including natural gas treatment. It has embedded a powerful library of process units such as mixing tanks, heat exchangers, absorption column sections as well as piping, pumps, valves etc. The library also contains basic instrumentation and control units.

### *3.1. Thermodynamic models*

Process simulation requires thermodynamic methods for calculation of thermodynamic properties and phase equilibrium. In order to enhance calculation speed K-spice® has a built-in thermo- package that takes care of the thermodynamic calculations based on table look-ups. In order to facilitate simulations with 30 wt% MEA additional look-up tables were generated from SINTEF's $CO_2$SIM [2] software.

The mass transfer coefficients and enhancement factor are calculated from the correlations given below.

$$kl_a = \mathrm{L}\left(\alpha, T\right)\left(\mathrm{C}_0 + \mathrm{C}_1\mathrm{G}_\mathrm{g}^{\mathrm{C}_2}\mathrm{G}_l^{\mathrm{C}_3}\right) \quad (1)$$

$$kg_a = \left(\mathrm{C}_0 + \mathrm{C}_1\mathrm{T}\right)\mathrm{G}_\mathrm{g}^{\mathrm{C}_2}\mathrm{G}_l^{\mathrm{C}_3} \quad (2)$$

$$E = \frac{1 + L\left(\alpha, T\right)\mathrm{C}_0}{\mathrm{C}_1 + \mathrm{C}_2\mathrm{G}_\mathrm{g} + \mathrm{C}_3\mathrm{G}_\mathrm{l}} \quad (3)$$

Where L are table values dependent on loading (α) and temperature (T), $C_0$, $C_1$, $C_2$ and $C_3$ are constants and $G_g$ and $G_l$ ([kg/m$^2$s]) are gas and liquid mass fluxes respectively.

### *3.2. Process equipment*

Rotary equipment such as pumps and fans are modelled by pump curves or performance curves. However, since the dynamic response is usually fast (seconds) in such equipment compared to the sampling time in the pilots, there is limited usefulness to model the rotary equipment accurately. Also, completely accurate speed-torque-power simulations are out of scope for the dynamic tests performed within the OCTAVIUS project.

Pipe sizing and elevation are important parameters in the K-Spice® model to predict the fluid holdup/inventory and hydrostatic pressures drop. It is assumed that the pipe volume is homogeneous implying no gradient between inlet and outlet. Valve sizing (valve flow coefficient and valve characteristic) in K-Spice® was done independent from the real plant; however this is not expected to have any impact on the results.

A fixed Heat Transfer Coefficient, (HTC) (dependent on phase) is assumed for both the heated and cooled fluid in the model of the heat-exchanger unit. Fluid volume and net heat transfer area are important input model parameters. The models representing the heated and cooled fluids are discretised into N sections with individual heat transfer, and then solved numerically. Heat-exchanger's data as fluid volume and heat transfer area were provided from the pilot plants, while the HTC for the two sides was determined so they match the measured temperature approach at steady state.

Columns in K-Spice® can either be modelled as trays or packed sections. They are all equilibrium based by default, but a rate-based approach is ensured through the reaction set MEA-$CO_2$. The stage calculation for both trays and packed sections is modelled as a flash operation. A single interface flash calculates the equilibrium between the phases at the interphase. The adjustable parameters for each stage are the gas- and liquid hold-ups and the rates of gas- and liquid- components into the interface. In this study the packed section model was used.

The reactive $CO_2$ component is treated separately in an enhancement model based on either a tray or packed model. In the latter case the full packed section is discretized in packed stages and it is recommended to keep the height in each individual packed stage less than HETP (Height Equivalent to a Theoretical Plate).

### *3.3. Heilbronn model*

A flow sheet of the K-Spice® model for the Heilbronn pilot is given in Fig. 3. Since the intercooling is bypassed during the dynamic test campaign at Heilbronn it was excluded from the process model in K-Spice®. For model simplicity, the pre-scrubber is omitted as well.

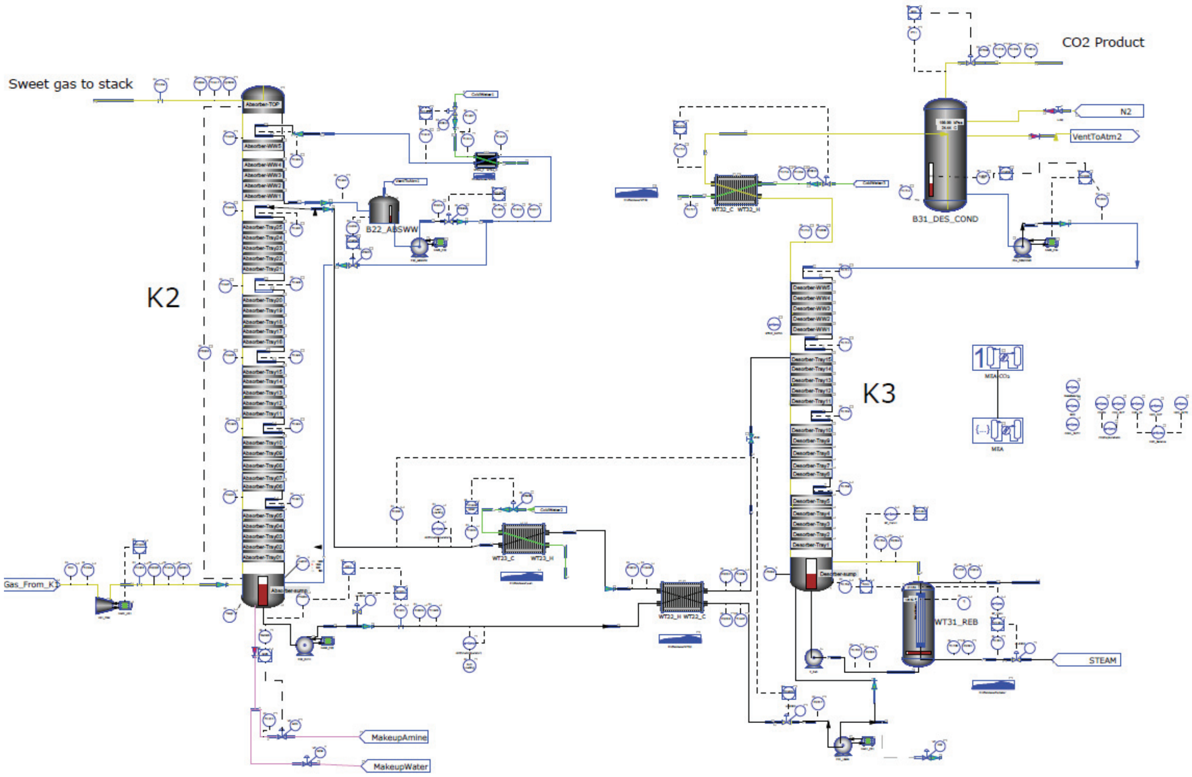


Fig. 3. K-SPICE model for the Heilbronn pilot

The process control system at Heilbronn logs about 120 parameters from instruments measuring flow, temperature, pressures etc., and most of these items were included into the K-spice® model for direct comparison. The sampling rate was every 120 second in the logging software of the pilot. Additionally, manual sampling of the solvent was done during the dynamic tests as this is not an online measurement. The amine concentration and $CO_2$ solvent loading of the samples were analysed.

## 4. Test runs comparison with simulations

A test campaign using 30 wt% MEA as solvent system was performed at the pilot plant at Heilbronn, Germany [1] from November 2013 to March 2014. The campaign included dynamic tests with transient step responses in steam, exhaust gas, solvent flow rate and finally a step in exhaust gas $CO_2$ content. In addition to single step responses, a test with a simultaneous step in all flow rates (exhaust gas, solvent and steam) was also included into the test campaign.

All these tests were also simulated with the K-Spice® model for comparison and model validation. To illustrate the performance of the dynamic model the results from some of these tests are shown in the following sections.

### *4.1. Step down in reboiler duty at liquid flow rate 4 m³/h*

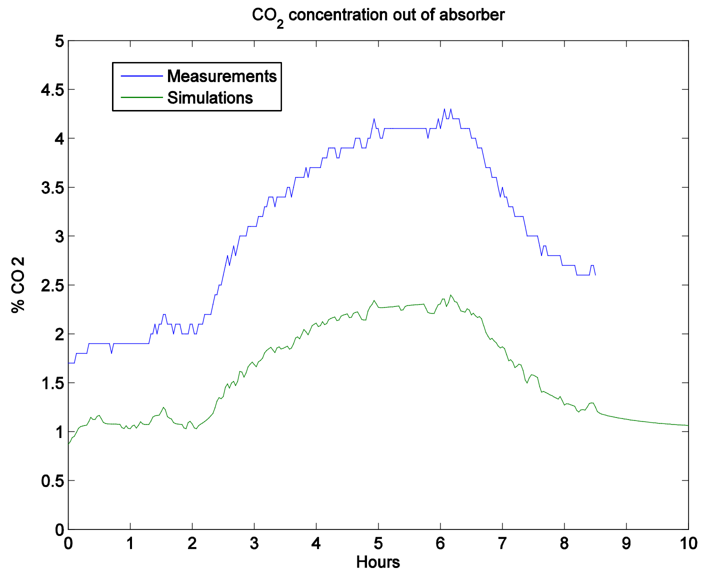


Fig. 4. Response in the $CO_2$ gas concentration out of the absorber by 20% step down in reboiler duty after 2.5hrs. Step back at 6.5h

In Fig. 4 the response from a 20% step down in reboiler duty is shown. The responses are very slow and it takes more than 6 hours to achieve a new steady state. Although there are some deviations in the $CO_2$ gas concentrations the response is very similar dynamically.

### *4.2. Step down in reboiler duty at liquid flow rate 6 m³/h*

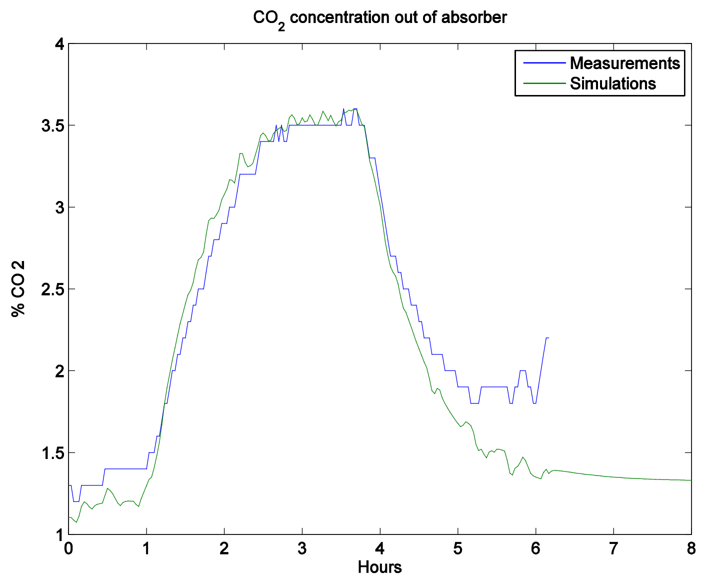


Fig. 5. Response in the $CO_2$ gas concentration out of the absorber by 20% step down in reboiler duty after ~1 hrs. Step back at ~4hrs

In Fig. 5 a similar step with 20% reduction in reboiler duty is shown with a higher solvent circulation rate. Again the responses are very similar dynamically, while the steady state deviations are also smaller. Note also that the responses are faster due to the higher liquid circulation rate. These two step changes show that the holdups in the plant are modeled reasonably well.

### 4.3. Step down in solvent flow rate

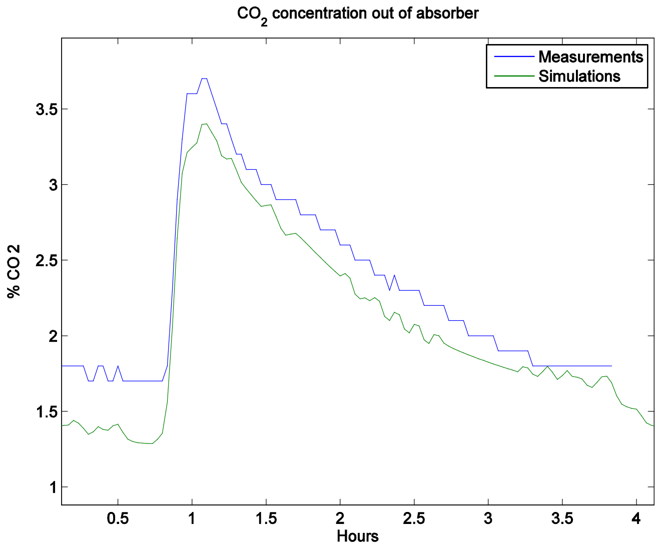


Fig. 6. Response in the $CO_2$ gas concentration out of the absorber by 20% step down in liquid flowrate.

In Fig. 6, the response of a step down in liquid circulation rate is shown. The two responses, simulated and measured, are very similar. Note that the response in terms of impact on the $CO_2$ out is very fast. However, the effect decreases after some time because the lean loading starts to decrease when less liquid comes into the desorber and the reboiler duty is constant. However, the effect decreases after some time because the lean loading starts to decrease when less liquid flows into the desorber and the reboiler duty is constant. However, the ability to control the $CO_2$ out of the absorber in short term will be used for control purposes as shown in Section 5.

### 4.4. Step down in gas flow rate

In Fig. 7, a 20% step down in the gas flow rate is shown. The simulated effect is somewhat larger than the measured ones although the differences are quite small.

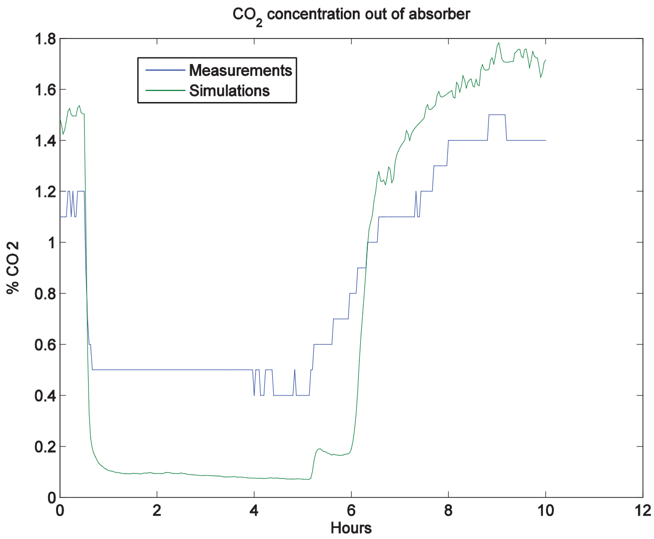


Fig. 7. Response in the $CO_2$ gas concentration out of the absorber by 20% step down in gas flow rate after ~1hr. Step back at ~5.5 hr.

### *4.5. Simultaneous step in flow rates*

In this test a simultaneous and proportional reduction in flue gas flow rate, liquid circulation rate and reboiler duty (steam flow rate) was performed. The measured and simulated responses are shown in Fig. 8 and show almost no effect on the $CO_2$ concentration in the gas out of the absorber. Thus ratio control can potentially be a good control strategy in an industrial $CO_2$ absorption plant (see Section 5.1).

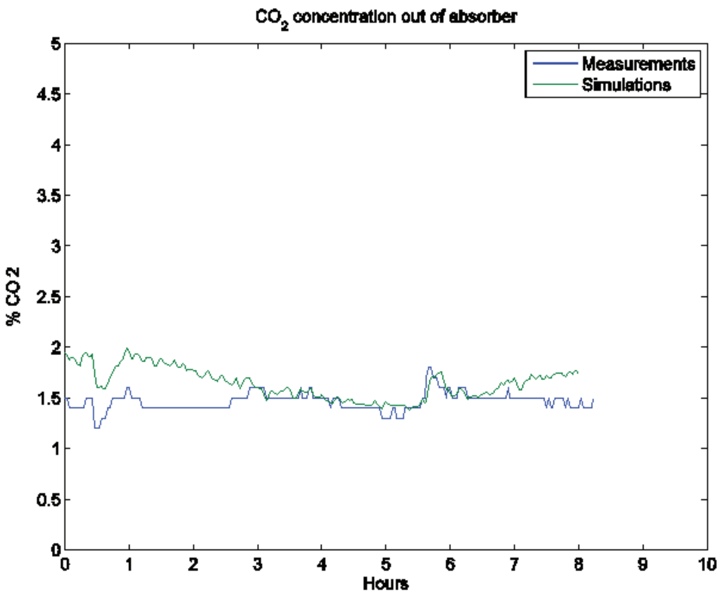


Fig. 8. Response in the $CO_2$ gas concentration out of the absorber for a simultaneous step of 20% in flue gas feed rate, steam rate and liquid circulation rate.

The results from the step change testing shows that the response in liquid $CO_2$ concentration is very slow due to the large solvent inventory in the pilot plant. Lowering solvent flow rate gave further delayed responses. $CO_2$ production rate and internal desorber column temperatures gave a fast response to a step in a reboiler steam flow rate. Simultaneous ratio responses in steam and solvent flow rate gave very little overall disturbance with a step-response in exhaust gas flowrate into the plant.

A comparison between the responses from the plant and K-spice® model shows some steady state deviations, but the dynamics are very similar. This was typical also for the other tests performed. The simulation results that will be presented in the next section should therefore be quite representative for the Heilbronn plant if the same control structure had been implemented there.

## 5. Control study

### 5.1. Control configurations

Two different control structures, referred hereafter as "Con1" and "Con2", were tested with the K-Spice® model. "Con1" was based on the results in Fig. 8 with a ratio control such that the liquid flow rate and steam flow rate were changed in proportion to the amount of $CO_2$ in the gas to the absorber. An extra feedback control loop was implemented to assure 90 % recovery in a long term.

"Con2" was based on the result in Fig. 6 and utilizes the lean liquid rate for the control of the $CO_2$ recovery ($CO_2$ out of absorber) and the steam flow rate (reboiler duty) for maintaining optimal conditions in the desorber. This control structure was also investigated by [3, 4]. They found that this control structure could be close to an optimal MPC, but much simpler to implement. Close to optimal conditions could be obtained by keeping an internal temperature in the desorber column constant.

### 5.2. Optimal start values

The dynamic simulations are easiest to interpret if they are started when the process is at steady state. In the present study we also wanted to start at a steady state condition that was optimal such that we could evaluate if the control system could maintain such an optimal condition.

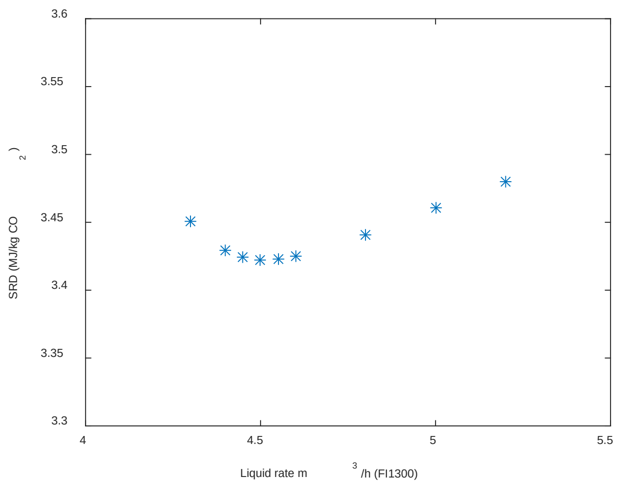


Fig. 9. Optimization of lean liquid flow rate for the implemented model, 90% recovery.

In Fig. 9 the result of the steady state simulations in K-spice is shown. The figure shows a minimum specific reboiler duty (SRD) of 3.420 MJ/kg $CO_2$ at a lean liquid rate of 4.5 $m^3/h$. Compared to a similar study at the plant the values are somewhat different, however the deviations was not large. The reason is as previously stated that there are some steady state deviations between the K-spice® model and the plant.

Fig. 10 show desorber temperature profiles for liquid flow rates 4.3, 4.5 and 5.0 $m^3/h$. The red line with 5.0 $m^3/h$ has a clear temperature pinch in the top of the desorber which is typical for a heat-limited case [5]. The green line has its steepest temperature profile in the bottom and is thus close to the stripping-limited case. The optimal case in blue is somewhere in between.

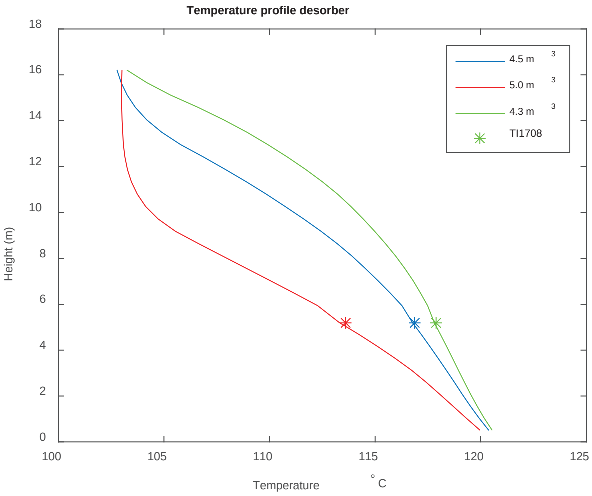


Fig. 10. Desorber temperature profiles for liquid flow rates 4.3, 4.5 and 5.0 $m^3/h$. Stars: Temperature sensor "TI1708".

The values of temperature sensor TI-1708 located between packed section one and two in the desorber are shown as stars in Fig. 10. This temperature is sensitive for the variations in liquid flow rates and is controlled by varying the reboiler duty in the "Con2" control configuration.

### *5.3. Step changes during the simulations*

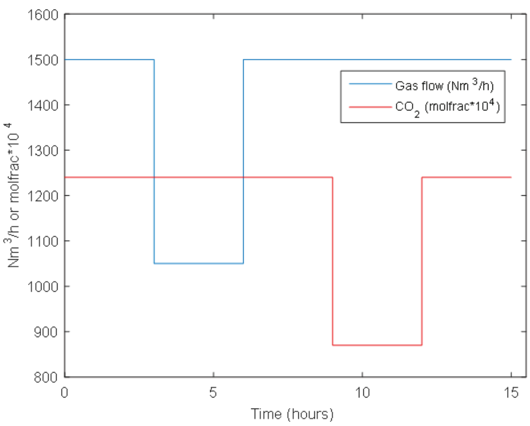

Fig. 11. Feed gas flow rate changes at 3h and 6h. Absorber $CO_2$ gas composition changes at 9h and 12h.

The two control strategies were tested by implementing them in K-Spice® and introduce 30% step changes in exhaust gas flow rate and exhaust $CO_2$ composition as shown in Fig. 11.

### 5.4. Test results

The response in $CO_2$ recovery for the control configuration "Con1" is shown in Fig. 12. It shows that the $CO_2$ recovery has some temporary deviation from 90% which are larger than expected from the test with simultaneous step in flow rates (see section 4.5). It also shows inverse responses that make it difficult to tune the outer loop faster. On the other hand, the average value during the simulations was 90.05%, which is very close to the objective.

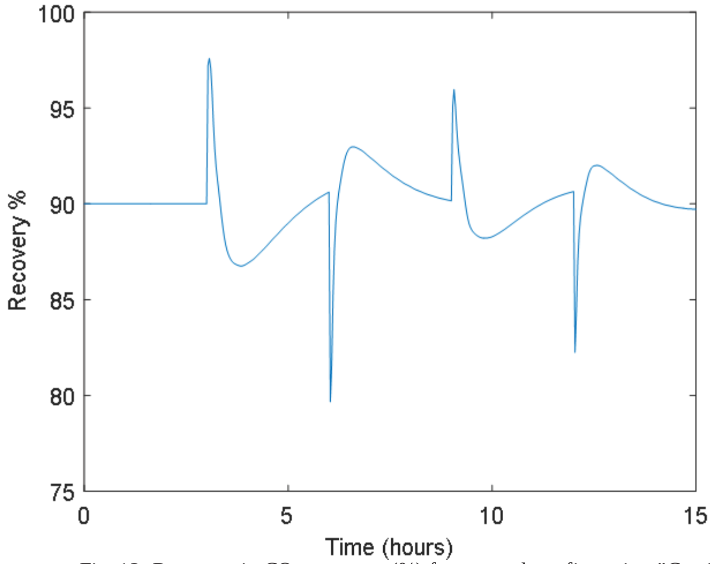


Fig. 12. Response in $CO_2$ recovery (%) for control configuration "Con1".

Fig. 13 shows the response in $CO_2$ recovery for the control configuration "Con2". We may notice that the absorber $CO_2$ recovery is excellent controlled with only small spikes of about 10 min for these large disturbances. The average value is 90.00%.

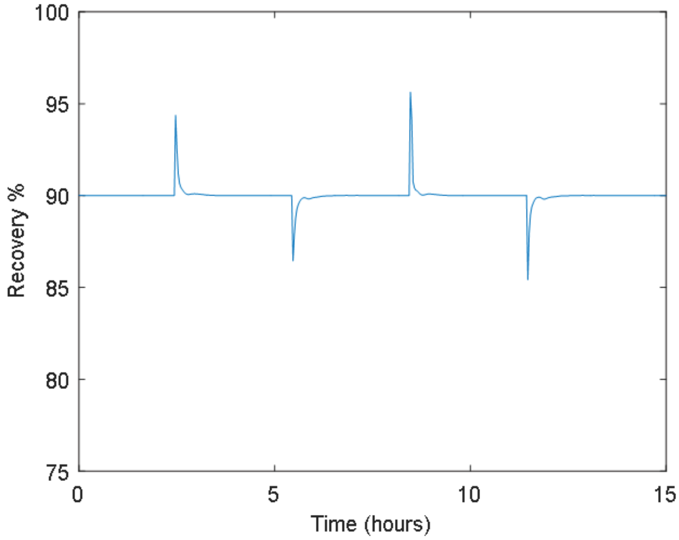


Fig. 13. Response in $CO_2$ recovery (%) for control configuration "Con2".

## 6. Discussion and Conclusion

In Table 1 the statistics during the simulations are shown.

Table 1. Statistics for the two control configurations

| | Recovery (%) | | SRD (MJ/kg $CO_2$) | |
|---|---|---|---|---|
| Control configuration | Mean | Std | Mean | Std |
| Con1 | 90.05 | 1.755 | 3.426 | 0.052 |
| Con2 | 90.00 | 0.521 | 3.420 | 0.077 |

The average $CO_2$ recovery was extremely close to 90 % for both configurations and the average SRD was very close to the optimal 3.420 MJ/kg $CO_2$. Even if "Con2" has slightly better statistics this will not be significant in a real plant.

However, the control structure "Con2" gives a much tighter control of the absorber $CO_2$ recovery, which may be important as a part of supervisory control. In supervisory control one may want to optimize the cost during the day and have a lower recovery when the electric prices are high and higher recovery when they are low. The fast and accurate control of the recovery for "Con2" is very well suited for such purposes. On the other hand, "Con1" gave quite significant inverse responses for the $CO_2$ recovery and is not suited for in those cases.

The temperature location in the desorber for the second loop in "Con2" could be further tested. In [3, 4] they found that a temperature in the upper part of the desorber was the most optimal. However, this temperature might much easier be non-sensitive to the steam rate if the desorber comes into the heat-limited regime. The dead time will also be larger in this loop due to the longer distance to the steam rate valve.

An alternative is also to use a weighted sum of several temperatures in the desorber similar to the work of [6], but this must be further tested before it can be recommended.

## Acknowledgements

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