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Demonstration of non-linear model predictive control for optimal flexible operation of a CO_2 capture plant

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ABSTRACT

Due to the penetration of renewable intermittent energy, there is a need for coal and natural gas power plants to operate flexibly with variable load. This has resulted in an increasing interest in flexible and operational issues in the capture plant as well. In the present paper a nonlinear model predictive control (NMPC) system was tested at the Tiller pilot plant in Norway. The most important part of the NMPC software is the dynamic model representing the absorber/desorber plant. A previous first principle (mechanistic) dynamic model of the plant using MEA was modified for a solvent of AMP and piperazine, and then successfully verified by step response tests. The NMPC, which was set up to minimize the deviation from the capture rate setpoint and minimize the specific reboiler duty was then tested in a closed loop with large changes in flue gas flow and CO_2 composition. Even for gas rate variations of more than 300% (110–340 m^3/h) and CO_2 concentration changes of 30%, the dynamic response was satisfactory. A test with frequently occurring constraints on the reboiler duty revealed a need for an extension to include direct control of the lean loading. Test of setpoint changes in total CO_2 recovery showed that the control system managed to rapidly change from one capture rate to another with a time constant of typically 10 min. This might be used in a second layer of optimization, a dynamic real-time optimizer, that minimizes the capture costs during a longer horizon considering varying energy prices.

1. Introduction

1.1. Flexibility of absorption-based capture processes

Power plants built in the 1990′s and early 2000′s were typically designed for base load operation favoring high efficiency, low capital costs and minimum cost of electricity production (IEAGHG, 2012). Due to the liberalisation of the energy market (Newberry, 2015) and the penetration of renewable intermittent energy, natural gas and coal power plants nowadays need to be more flexible in terms of electricity production. Both the volatility of the electricity prices and the need for stabilizing the electricity grid favor power stations that can change the electric power production load quickly and reliably. This has resulted in more flexible power plants. A new lignite coal power plant can today typically change the electric output down to 50% of the capacity with a ramp rate of 2-5%/min (Henderson, 2016) and the ramp rate of an F-class natural gas turbine is about 10%/min. (Abudu et al. 2021).

For post combustion CO_2 capture plants, this might imply operational challenges which have resulted in an increased focus on operational flexibility in the post combustion capture plants as well (Cohen et al., 2011; IEAGHG, 2012; Alie et al., 2016; Bui et al., 2016; IEAGHG, 2016).

1.2. Load-following operation

One important flexibility issue is the ability to follow the changes in the power plant load over time. When the load changes, the amount of flue gas produced changes correspondingly. This means that the capture plant must handle a flowrate reduction down to 50% with a change rate of 3%/min as well. This "load-following" capability is basically a control issue and previous research has focused on discussing the best control structure, including the pairing of control loops Panahi and Skogestad (2011). concluded that the optimal control of a PCC plant for this scenario was to use lean liquid flow rate to control the capture rate, and the steam to reboiler (reboiler duty) to keep a temperature in the upper part of the desorber constant. This result was based on a "self-optimizing

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control" procedure for selecting control variables that by keeping them constant, kept the process close to optimal conditions during various disturbances. Based on relative gain array (RGA) analysis, Luu et al. (2015) proposed that the capture rate could be controlled by the lean solvent flowrate and the SRD to be controlled by the reboiler duty. They also proposed to use a model predictive control (MPC) strategy, where the model was a linearized version of the nonlinear dynamic process model implemented in gPROMS Mejdell et al. (2017). compared two control structures. The first approach was based on a liquid-to-gas ratio control in the absorber and a steam-to-liquid ratio control in the desorber in combination with a slow feedback control on the CO_2 concentration in the gas out of the absorber. The second was like Panahi and Skogestad (2011) to use the lean liquid rate to control capture rate and the reboiler duty to control a desorber temperature. However, the temperature was located in the lower part of the desorber where the temperature is sensitive to deviations from optimal conditions. Both structures showed promising performance during the simulation testing, but the second one was superior, especially with respect to CO_2 recovery control. The fast response observed with this control structure was due to the resulting tight control of lean loading, and the large time constants associated with loading changes were then avoided.

It might be concluded, also from other publications (IEAGHG, 2016; Meechleri et al., 2017; Montañés et al., 2017; Marx-Schubach and Schmitz, 2019), that most authors recommend controlling the CO_2 concentration out of absorber by the lean liquid flowrate and the lean loading by the reboiler duty. Since the lean loading usually is not measured online it might be controlled indirectly by a liquid density sensor, a temperature in the reboiler or in the desorber column Mejdell (2020). discusses the various alternatives for lean loading control.

1.3. Operation under limited availability of steam

Flexibility in power plant output might also be related to the energy requirements in the capture plant. Since the reboiler requires 45-50% of the available low-pressure steam, some or all of this steam might be redirected to the low-pressure steam turbines for shorter or longer periods of the day, for example when the electricity prices are high. The issue with limited steam might be handled temporarily by having additional large lean and rich buffer tanks on the liquid circulation line and postpone the production of lean solvent to periods when there is excess steam available from the power plant (Enaasen Flø et al., 2016; Kvamsdal et al., 2018). However, as recognized by e.g., Van Peteghem and Delarue (2014), such solvent storage might only be profitable if the electricity price fluctuation can justify the increased investment cost. Another similar suggested solution is to utilize the CO_2 capacity of the solvent by temporarily increasing the loadings and the circulation rate Sanchez Fernandez et al. (2016). and Mac Dowell and Shah (2015) found this more profitable compared to buffer tanks. For shorter time periods one might also utilize the latent heat in the desorber by pressure reduction (Ziaii et al., 2011). These latter strategies avoid the need for additional process equipment but will cause the capture plant to deviate from optimal conditions.

If the legal emission limit of CO_2 is specified over a longer period one might also use strategies with varying CO_2 capture rates. One obvious solution is to bypass the absorber partly or totally during periods of limited steam availability (Enaasen Flø et al., 2016; Sanchez Fernandez et al., 2016), another is to reduce the capture rate of the absorber. Optimal control of the capture plant in a regime with limited steam availability was also addressed by Panahi and Skogestad (2012). One suggested solution was to lower the liquid circulation rate to the level of available steam. This will keep the lean loading at an optimal level while the capture rate will be sacrificed until sufficient steam is available again. The advantage of this configuration compared to solvent load accumulation is that the slow and sometimes fluctuating dynamics related to loading changes are avoided, and optimal conditions is quickly reached as soon as enough steam is available.

A similar solution (constant lean loading from the desorber) was proposed by Ziaii et al. (2009). However, they suggested to keep the liquid rate to the absorber constant by a partly liquid bypass of the desorber. Besides additional equipment costs, such mixing of regenerated and unregenerated solvent is not thermodynamically favorable. Both solutions require a change in the control configuration since the lean loading now is controlled by the rich liquid rate and not the reboiler duty.

A redirection of steam for power output increase might be a very efficient manner to allow the power plant to quickly stabilize the grid frequency. In this scenario, described by Haines and Davidson (2014), the reboiler steam is within a few seconds redirected to the low-pressure turbine to stabilize the grid. The redirection is maintained until the power plant has managed to increase the load and sufficient steam is available again.

1.4. Non-linear model predictive control (NMPC)

A multivariable controller that uses both inputs and outputs simultaneously might be an alternative to single loop configurations where the optimal pairing depends on the availability of steam. In the present work an advanced control system based on non-linear model predictive control (NMPC) has been used. For simplicity, this involves that a dynamic first principle (mechanistic) process model is utilized to predict future process performance to optimize the control actions. The prediction model is tailor-made for the purpose of online and closed loop control, meaning that calculation speed and robustness is highly focused. The NMPC application used at the Tiller pilot plant is a software tool (CENIT) developed by Cybernetica which is integrated with the basic Siemens PC7 control system through an Open Platform Communications (OPC-UA) server interface. *Cybernetica CENIT* is a widely used tool for NMPC applications, particularly within the polymer, metallurgical industries, and batch processing (Singstad, 2017; Elgsæter et al., 2012).

This control system has earlier been tested successfully at both the Tiller pilot plant in Trondheim, Norway, and at the Technology Centre Mongstad (TCM) for 30% MEA as solvent system. Details of the NMPC algorithm and implementation, the CENIT software and some results from testing of 30 wt.% MEA solvent system at TCM and Tiller can be found in Hauger et al. 2019. Kvamsdal et al. (2018), tested the NMPC in an operating scenario where the electricity price fluctuated during a 24 h period. Besides the NMPC that controlled the process and minimized the SRD for a given capture rate, a dynamic real-time optimizer (DRTO) was implemented as a second layer of optimization that minimized the total cost by dynamically changing the NMPC capture rate setpoint during the period. The test results showed that such operation may enable considerable reduced energy demand in the reboiler and more importantly, such optimal operation is not possible to control manually even for very skilled operators.

In the present work, the NMPC is tested with the CESAR1 solvents system at the Tiller pilot as part of the ALIGN-CCUS project. Like the previous tests with NMPC, the controlled variables *(CV)* were the capture rate and the reboiler duty (electric power to the reboiler). The objective is to minimize SRD (Specific Reboiler Duty) in MJ/kg CO_2 while keeping the capture rate at a given setpoint. The specifications for the two CVs are obtained by manipulating the lean solvent flowrate and the reboiler duty, which are then defined as manipulated variables (*MVs*). Note that in this NMPC setup, the reboiler duty is both an MV (to be manipulated) and a CV (to be minimized).

The NMPC controller has a sampling time of 60 s, meaning that optimized values of MVs are recalculated every minute. The optimal prediction from previous sample is used as a nominal prediction for the nonlinear solver. The NMPC has a prediction horizon of five hours, which means that the optimization criterion is evaluated at certain times during the five following hours. During this prediction horizon, the MVs are allowed to change values four times; immediately, after 15 min, after

45 min and after 90 min. This time discretization, also known as *MV blocking*, is introduced to reduce the size of the optimization problem and thus the calculation time. As the optimization is repeated every 60 sec, the MV blocking has minor effect on the result.

1.5. The Tiller pilot

The pilot plant was built and commissioned in 2010 and is inside a 30 m high building at the SINTEF site at Tiller in Trondheim, Norway (Mejdell et al., 2011). A simplified process flow diagram of the pilot plant is shown in Fig. 1. Flue gas with 11.85% CO_2 (dry) is provided by a propane burner. However, by recirculating some of the captured CO_2 and/or diluting with some air various concentrations of CO_2 in the feed gas can be specified. The absorber packing height is 19.5 m while the packing height in the stripper column is 13.6 m. Mellapak 2X is used as packing materials in both columns and in the two water wash sections in the top of each column. The pilot plant is well instrumented for online measurement of temperature, pressure, flows, densities, and composition. The capacity of the plant is 30–60 kg/h CO_2. The warm parts of the desorber column and reboiler are heat traced such that it operates almost adiabatically. The remaining heat loss is estimated to be 1.5 kW. Around 160 temperature and 110 other tags (pressures, analyzers etc.) are handled by the system.

2. Initial tests for model verification

A two-week test program was carried out in the beginning of the campaign to provide data for model verification. Both steady state data and dynamic step response data were obtained.

2.1. Steady state data

Eight steady state runs were obtained during this part of the campaign. A run is typically performed by adjusting and stabilizing the process parameters during the daytime, then leaving the plant unchanged during the night, and finally take the liquid samples the next morning. All process parameters are averaged over the last hour before liquid sampling.

2.1.1. Optimal L/G

The steady state data were used to find the optimal L/G for 90% capture rate as shown in Fig. 2. In these steady state runs the flue gas was kept constant at 160 m^3/h with a concentration of 13.3% CO_2 dry, typical for coal power plant exhaust.

The minimum SRD was found to be 3.22 MJ/kg CO_2 for an L/G ratio of 1.58 on a mass basis. No heat loss was accounted for. Note that the SRD increases only slightly at higher L/G rates than for the optimal value, while the SRD increases quite rapidly for L/G lower than the minimum. This might be a challenge during the optimization as discussed by Hauger et al. (2019).

The amount of CO_2 captured can be calculated in four different ways:

- CO_2 removed from the flue gas.
- CO_2 absorbed in the absorber liquid phase.
- CO_2 stripped from the desorber liquid phase.

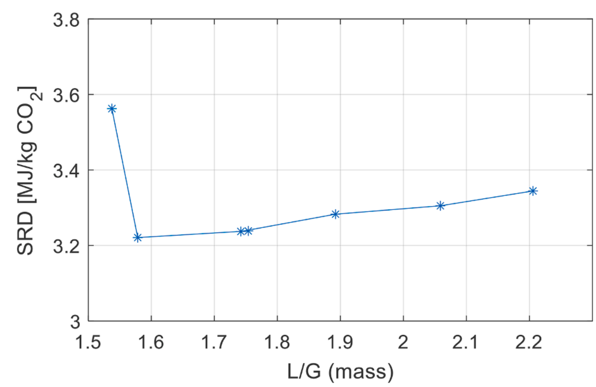

Fig. 2. Specific reboiler duty as a function of liquid gas mass ratio.

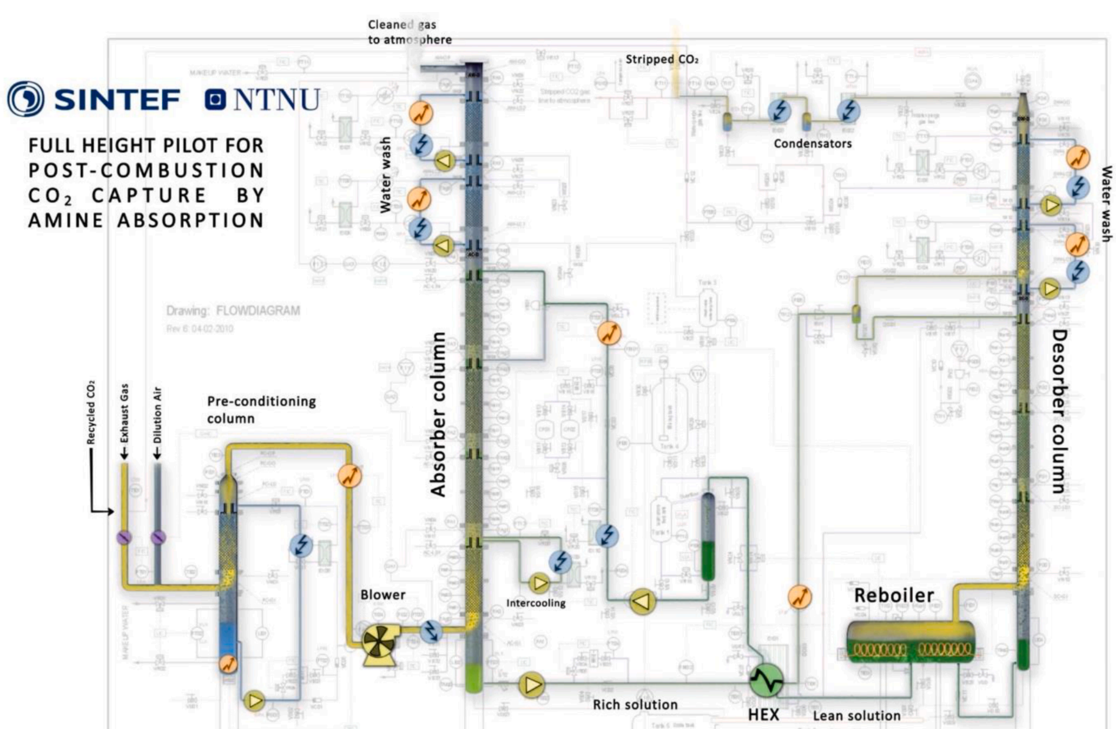

Fig. 1. Simplified process flow diagram of the Tiller plant.

- CO_2 leaving in the top of the desorber as product.

By assuming a perfect mass balance these four values should be equal at steady state. The standard deviation between these four measurements was found to be very small, only 1.8%. The average value was used to calculate the SRD shown in Fig. 2.

2.1.2. Estimator for the CO_2 concentration in the liquid phase

Tait et al. (2016, 2018) addressed the advantage of having an online measurement of CO_2 loading. In this work, we have used the initial eight steady state runs to establish an online estimator for the liquid CO_2 concentration. The input to the model is the liquid density measured by a Coriolis sensor together with a temperature measurement located nearby. The correlation used has the following simple form:

$$\rho = k_0 + k_1 C_{CO_2} + k_2(T - 20)$$

Here$\rho$is the density in g/L,C_{CO_2} is the CO_2 concentration in mol /L and T in °C. Rearranged and solved for $C^*_{CO_2}$in terms of mol/kg we get

$$C^*_{CO_2} = \frac{\rho - k_0 - k_2(T - 20)}{k_1} \frac{1000}{\rho}$$

The model parameters were fitted to laboratory liquid analyses of CO_2 (mol/kg) and to one hour averaged steady state data of the density and temperature. The optimal values of parameters k_0, k_1 and k_2 were found to be 1013.4, 37.94 and -0.7337.

The result fitting the model parameters to the experimental data is shown in Fig. 3.

Fig. 3 shows that the model fits the data very well for both streams. The standard deviation is only 0.015 mol/kg. Since the sensors used in the estimation is on-line measurements, the model is useful during dynamic testing and in the NMPC model since it provides an indirect accurate online estimate of the liquid CO_2 concentration and loading.

2.2. Step response tests

Twelve step response tests were performed during this initial period. These tests included steps in liquid and gas flowrate, CO_2 concentration, reboiler duty, flue gas saturation temperature and stripper pressure. Such tests are important to reveal the dynamics of the plant and for the pairing of control loops see e.g., Mejdell et al. (2017). An example is shown in Fig. 4 where a step down in liquid rate is shown together with the response in the CO_2 concentration in the gas out of absorber. The immediate quick response is that the CO_2 concentration increases, but after a while it reduces again to a new steady state with a concentration lower than before the step change. The reason is that the reboiler duty was unchanged during the test and when less liquid is fed to the desorber the solvent starts to be leaner and this again lowers the gas CO_2 out of absorber. This illustrates that the liquid loading changes are slow while the liquid- and gas flow changes are comparably very fast.

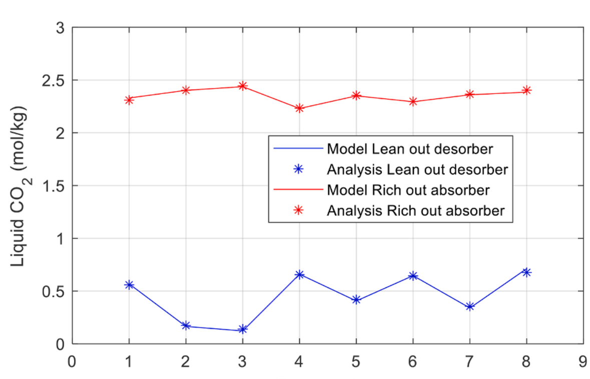

Fig. 3. Measured and estimated values for lean flow out of desorber and rich flow out of absorber.

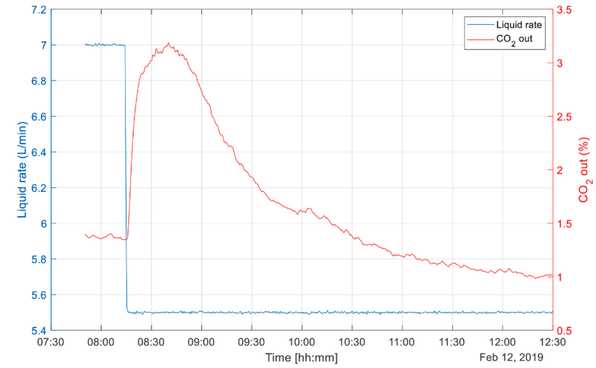

Fig. 4. Step down in liquid rate and the response in gas concentration out of the absorber.

The step changes were primarily used to validate the dynamic process model used in the NMPC. The most important observable process variables to monitor for in the NMPC software, is the production of pure CO_2 in the desorber, the concentration of CO_2 in the flue gas out of the absorber, the lean and rich solvent densities, which are closely linked to the lean and rich loading as shown in the previous section.

As an example, a comparison between the measured and modelled desorber production rate of CO_2 is shown in Fig. 5. The comparison was done during the first three days of the dynamic testing with step variations in the flue gas rate and concentration, liquid flow and reboiler duty.

Though there are some deviations between the data measured in the pilot and the data calculated by the model in the NMPC, the dynamics in the pilot is captured well by the model. During later tests in closed loop such deviations were also taken care of by the NMPC software.

3. Test of the NMPC application

3.1. Overview of the tests

The main objective of the campaign was to challenge the NMPC application with some difficult dynamic scenarios described in the introduction that may apply for power stations with CO_2 capture. An overview of the closed loop (CL) tests is given in Table 1.

3.2. Test 1: load following change in flue gas flowrate

In this test, the flue gas flowrate was first increased by 3%/min from 160 to 190 m^3/h and then reduced again after 1.5 h to 160 m^3/h with the same change rate. The flue gas concentration was kept constant at 11.85 mol% CO_2 (no recirculation of CO_2). The change rate is typically a

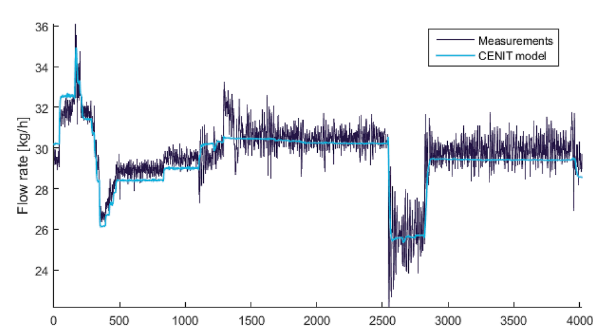

Fig. 5. Model validation showing the production rate of CO_2 from the stripper with ballistic (uncorrected) NMPC model calculations vs. the corresponding plant measurements.

Table 1
Closed loop test plan for the NMPC application.

Test	Test description	Setpoint changes	Frequency/ change rate	Fixed parameters and other notes
1	Load following change in flue gas flowrate	160–190 m^3/h	3% /min	After two hours the flue gas flowrate was reduced to 160 with the same rate of change
2	Load following change in both flue gas flowrate and CO_2 concentration	160–200 m^3/h	3% /min Random changes in concentration	The change in CO_2 concentration introduced by recycling CO_2 from top of the stripper.
3	Load following "worst case". Large variations in flue gas flowrate and CO_2 concentration	110–340 m^3/h	2.5, 0.15, 3.5, 1% /min	Scenario with large variations in flue-gas flow. Various rate of changes. Very frequently changes during one day including flue gas CO_2 concentration.
4	Load following with reboiler duty limitations	160–200 m^3/h	3% /min	Constant CO_2 concentration (12 mol%), change in max value for the available reboiler energy input
5	Grid frequency stabilization with "stripper stop"	160–175 m^3/h	Sudden 75% reduction in reboiler duty	After the reduction, the power plant will increase feed rate to the boiler and thus increase the flue gas and also the available steam to reboiler.
6	Capture rate setpoint changes	85–99%	Every 1–2 h	Constant flue gas of 160 m^3/h, CO_2 concentration in flue gas 12 mol%

max load change rate for coal power stations and represents one challenging scenario. The change in flue gas flowrate together with the controller action on the liquid flow and reboiler duty is shown in Fig. 6 and Fig. 7, respectively.

It is seen that the liquid flowrate and the reboiler duty follow the changes in gas flowrate with a smaller delay. These two manipulated variables are changed by NMPC to keep the capture rate at its setpoint and to minimize the SRD.

Note that the ramp changes in the flue gas flow were done within the Siemens control system. NMPC could read these changes via the OPC server and use this information in the optimization calculations. In a real power plant situation, the flue gas changes are a result of load changes typically initiated by the operators in the power plant and similar

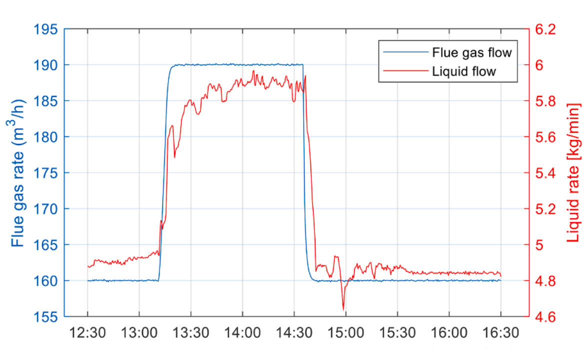

Fig. 6. Ramp step in the flue gas flow and the controller response in liquid flow.

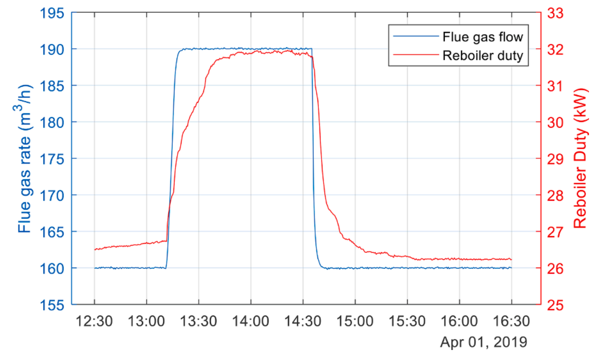

Fig. 7. Ramp step in the flue gas flow and the controller response in reboiler duty.

information might be available from the power plant control system. The result is shown in Fig. 8. Except for some immediate deviations after changes in the flue gas flow, the capture rate is kept almost at the desired setpoint. Also, the SRD is kept almost constant during these changes.

The rich and lean loadings based on the estimator outlined in Section 0 are shown in Fig. 9. The figure shows that the lean and rich loading is almost constant during the test period. This explains the quick responses shown in the test. The large time constants frequently observed in capture plants are generally associated with loading changes.

3.3. Test 2: change in both flue gas flowrate and concentration

In this test, there were variations in both the flue gas flow and CO_2 concentration. As in the previous test, the change in flue gas flowrate is directly monitored by the NMPC as a change in setpoint, while the change in concentration is regarded as a measured disturbance. Thus, the former is seen by NMPC prior to the actual changes and can thus act accordingly and faster to this change, while for the latter any action from NMPC is based on the assumption that the current measured value remains fixed during the prediction horizon. However, both cases give some feed-forward control action.

The change in CO_2 concentration in the inlet gas was initiated at 9:45 by recycling 3.5 m^3/h CO_2 from the top of the stripper to the flue gas at the absorber inlet. This recycling was stopped at 13:40. The change in flue gas flowrate (from 160 to 200 m^3/h at a change rate of 3% /min) was introduced at 10:30 and then at 12:20 the flue gas flowrate was ramped down to 160 m^3/h again with the same change rate. This change in flue gas flowrate also influence the CO_2 concentration as seen in Fig. 10.

As can be seen from Fig. 11, the capture rate was kept at the setpoint value of 90% most of the time except for a shorter time around 10:30 when both the flue gas flowrate and CO_2 concentration were changed simultaneously. The SRD was also kept reasonably constant during this test.

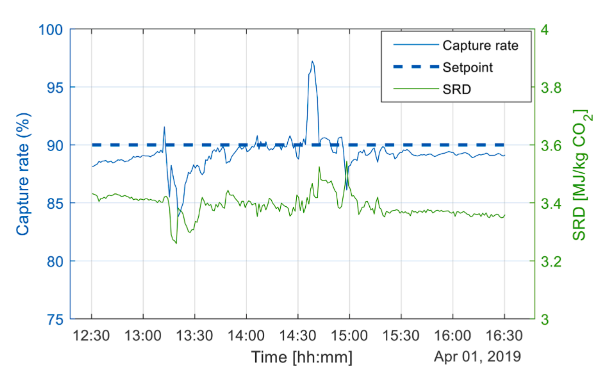

Fig. 8. Response in capture rate and SRD during the test period.

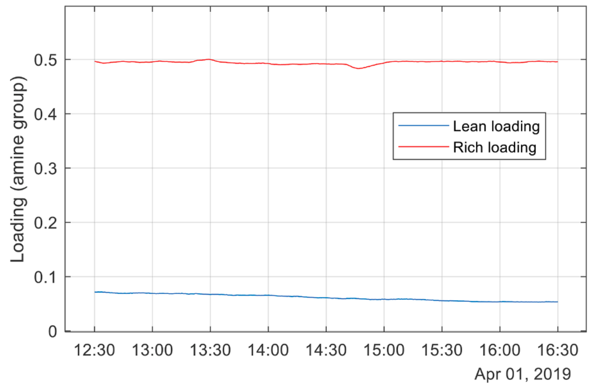

Fig. 9. Response in lean and rich loading (mol CO_2/mol amine group) during test period.

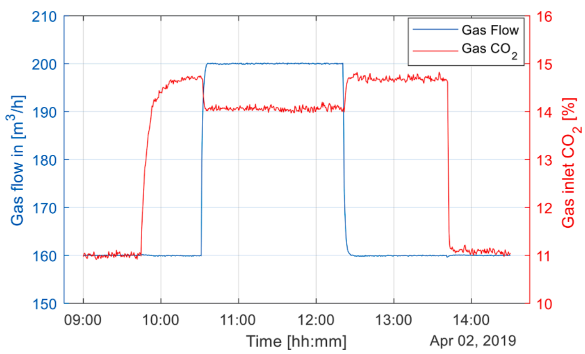

Fig. 10. Steps in both flue gas flow and composition.

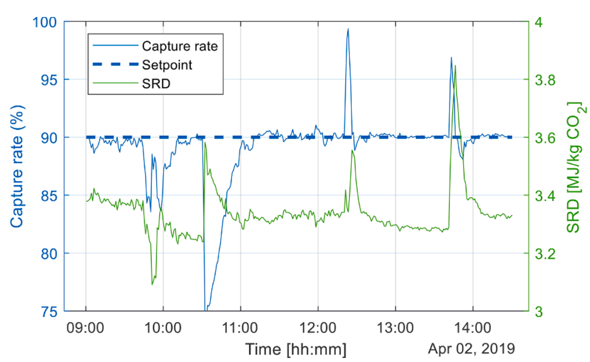

Fig. 11. Response in capture rate and SRD during the test 2 period.

3.4. Test 3: load following worst case scenario

This controller test is denoted "worst-case scenario" because it was designed to stress-test the NMPC application and the plant by rapidly changing the plant conditions towards operating limitations. Fig. 12 shows the variation in flue gas flow and CO_2 composition.

The flue gas flowrate was first increased from 200 to 230 m^3/h at a change rate of 5 m^3/h/min (2.5%/min). Then it was decreased to 110 m^3/h at a change rate of 3 m^3/h/min (0.15%/min). Then the flue gas flowrate was increased with 6 m^3/h/min (3.5%/min) in about 10 min and then at a change rate of 3 m^3/h/min (1.0%/min) up to 340 m^3/h. The maximum flue gas flowrate was then 3 times higher than the minimum flue gas flowrate, which serves to illustrate the flexibility of the plant itself and the control system. As in test 1 and 2, the change in flue gas flowrate is directly observed by NMPC as a change in setpoint, while the change in concentration is regarded as measured disturbances.

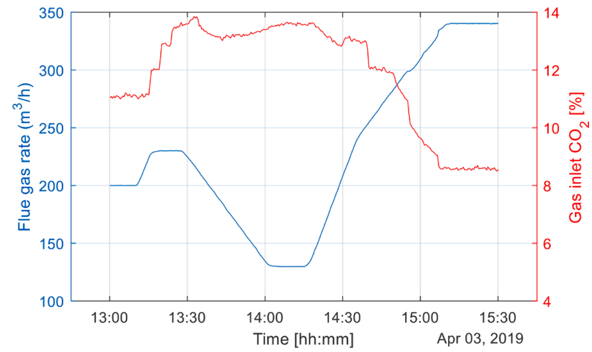

Fig. 12. Ramp changes in flue gas flow and CO_2 composition disturbances.

The two manipulated variables, the liquid flowrate and the reboiler duty, are shown in Fig. 13. These variables respond according to the flue gas flowrate since the amount of CO_2 fed to the absorber is mainly affected by flowrate, and to a less extent by concentration. Note that Fig. 14 shows that the large variations in this test hardly affect the lean and rich loading.

In Fig. 15, it is seen that it was difficult to keep the actual capture at the targeted setpoint of 90%. However, it can obtain almost 100% for the lowest flue gas flowrate, while it is much lower for the maximum flue-gas flowrate. Nevertheless, the average capture rate is around 87% for the whole period, which is not bad considering the very large changes in flue-gas flowrate and CO_2 concentrations. This shows that an NMPC-based control system like the NMPC system is very useful for capture plants that need to operate in a highly flexible manner, but it also shows the need for strategies in which the CO_2 capture rate is allowed to change dependent on the operation of the upstream plant.

3.5. Test 4: change in flue gas flowrate with reboiler limitations

In this test, the flue gas flowrate to the absorber inlet was changed from 160 to 200 m^3/h with a change rate of 3% /min (see Fig. 16) while the CO_2 inlet concentration was kept constant.

However, an electric power limit of 30 kW was periodically set as shown in Fig. 17. This would resemble a shortage of steam from a power-plant due to increased electric output demand. This limitation was relaxed between 10:37 and 11:21 and after 12:49.

It can be seen from Fig. 18, that it is not possible to keep a setpoint of 90% capture rate during the periods with constrained reboiler duty. It is a key feature of the NMPC implementation that the setpoint for capture rate is *deliberately* disobeyed when the desorber reboiler duty is strictly constrained. It would be possible to maintain the capture rate temporarily, but that would lead to inability to maintain the appropriate lean loading over time, eventually leading to a severe decrease in capture rate. Instead, the NMPC quickly backs down to a *sustainable* level of

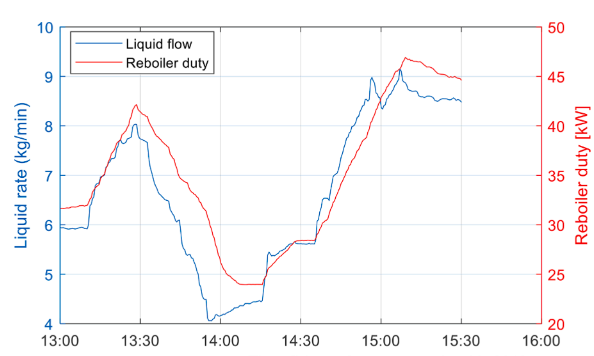

Fig. 13. The controller responses in liquid flowrate and reboiler duty to the changes in Test 3.

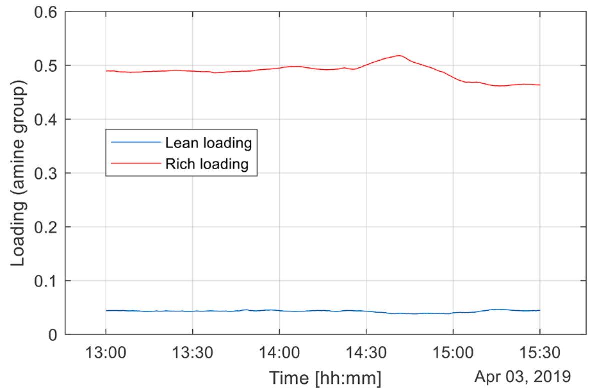

Fig. 14. Responses in lean and rich loading (mol CO_2/mol amine group) during Test 3 period.

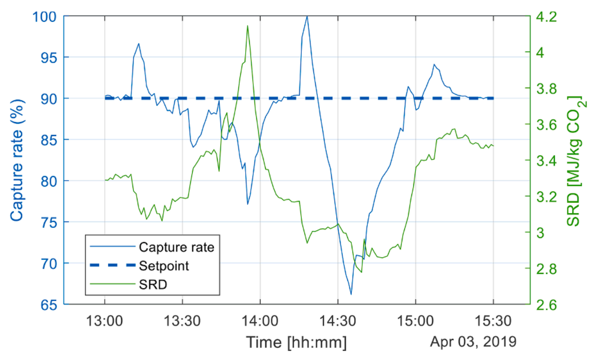

Fig. 15. Response in capture rate and SRD during the Test 3 period.

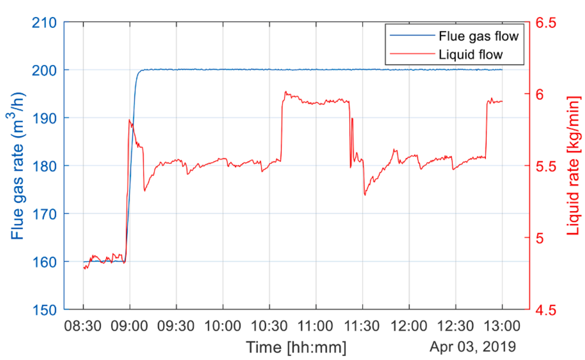

Fig. 16. Ramp step in the flue gas flow and the controller response in liquid flow during limitations in reboiler duty.

capture, given the reboiler duty constraint.

3.6. Test 5: grid frequency stabilization with "stripper stop"

An electrical grid that is served with substantial solar and wind energy will face situations where it is desirable that the coal or natural gas power plants can maintain the grid frequency very quickly. An instant power boost is required over a very short period, typically in the range of 10 s, and it will not be possible for the power plant to ramping up its load that fast. However, as discussed by Haines and Davidson (2014), this change in power demand can be countered by temporarily redirecting the reboiler steam to the low-pressure steam turbine until the power plant has managed to ramp up the power production. Such grid reliability services might give the power plant additional income (Cohen et al., 2011). In Test 5 at Tiller, the heat was not stopped completely, but

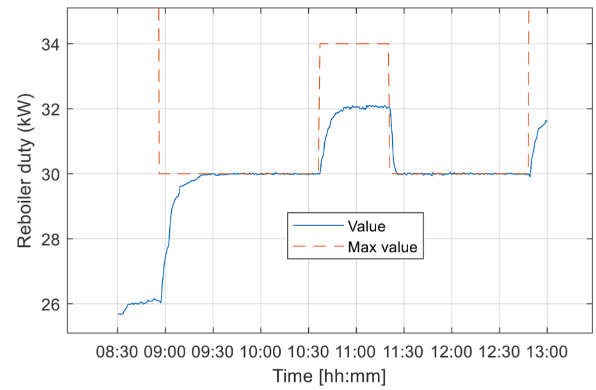

Fig. 17. Actual reboiler duty and maximum energy input from a resembled power-plant.

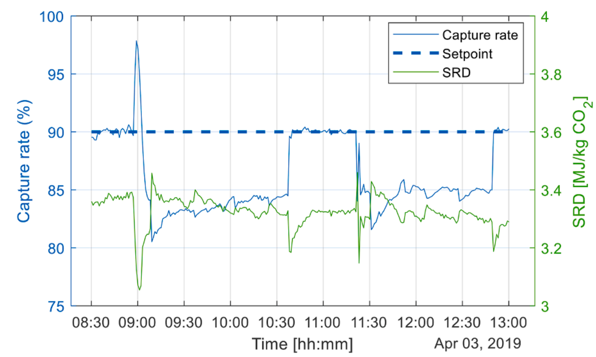

Fig. 18. Response in capture rate and SRD during the Test 4 period.

reduced to 25% within 10 s.

On the first day of testing, the stripper stop test was conducted three consecutive times. In Fig. 19 the reboiler duty and flue gas flow is shown. At 10:00 the reboiler duty was quickly decreased during 10 s to 7 kW. This low heat input was kept for 30 s before the flue gas flowrate increased by 3%/min from 160 to 176 m^3/h. This was due to the assumed increase in the power plant load. The energy to the reboiler was increased accordingly up to 28 kW to mimic that more "low-pressure steam" energy from the power plant is produced and available for the reboiler. At 11:17 the flue gas flowrate went back to the original value and new tests were conducted at 13:04 and 13:52.

From Fig. 19 it can be observed that the reboiler duty increased to a final level that was higher than in the beginning due to the increased flue gas flow it had to treat. When the flue gas was reduced back to 160 m^3/h to prepare for the next test around 11:15, however, the reboiler duty went down again, but not to the same level as before the test began. Judging from the lean loading, as shown in Fig. 21, the cause of the

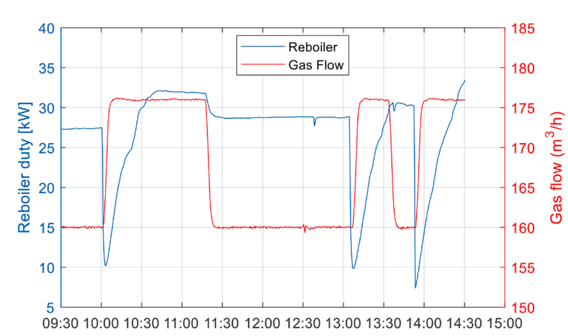

Fig. 19. The reboiler duty and flue gas flowrate during Test 5.

slightly increased reboiler duty was the desire to lower the lean loading over time.

The capture rate is kept well, and the SRD shown in Fig. 20 is kept at moderate values during the test period.

Fig. 21 shows the liquid circulation flowrate together with the lean loading level. It is shown that, like the case with the reboiler duty, the liquid solvent flowrate and the lean loading were not the same after reducing the flue gas flowrate back to 160 m^3/h. Instead, it remained at an elevated level, and it seemed to increase after each consecutive reboiler stop. After the last test it increased steadily and, at the end, the NMPC control was deactivated by the operator.

We can conclude that the control by NMPC led to undesired plant conditions during these tests, in the sense that it was unable to fully recover from one reboiler stop before the next one occurred. Possible reasons are discussed in Section 4, but the main reason was believed to be too short breaks between each consecutive test, leading to a loading build-up, which agrees with the lean loading as shown in Fig. 21.

To investigate this hypothesis, the pause between each consecutive reboiler stop was increased the following day of testing. In addition, the rate of change for the flue gas flowrate was decreased from 4.8 to 3.2 m^3/h/min. The results are shown in Figs. 22–24.

The gas flow and reboiler duty shown in Fig. 22 and the capture rate and SRD in Fig. 23 are quite similar to the corresponding responses in Figs. 19 and 20 the first day. On the other hand, looking at the loading estimates as shown in Fig. 24 there are some changes.

While the effect on rich loading is rather small there is a significant effect on the lean loading. In this test the NMPC control system managed to run the plant back to the optimal lean loading condition, but it takes almost two hours. This contrasts with the first day of testing where the stripper stops occurred more frequently with the results that the lean loading steadily increased. In that case the controller increased the liquid flowrate to maintain the capture rate which in turn increased the lean loading and the plant went unstable. This emphasises the importance of having additional control with the lean loading. The reason is that the SRD is only a weak function of lean loading at higher liquid flowrates as shown in Fig. 27. Some direction for improvements and further testing of events like stripper stop is discussed in the Section 4.

3.7. Test 6: capture rate setpoint changes

For base-load operation of the upstream power plant, the capture rate will normally be kept constant, but for more flexible operation, it may be important and more optimal to change the capture rate depending on availability of steam and/or variations in electricity prices as discussed previously. Thus, tests with consecutive changes in the setpoint from 85 to 99% for the capture rate were conducted. A test of changing the capture rate setpoint every hour between 90 and 99% is shown in Fig. 25 together with the resulting SRD. After a setpoint change the capture rate changes smoothly and reaches the new setpoint within about 15 min. It is also noteworthy that 99% capture rate is attainable

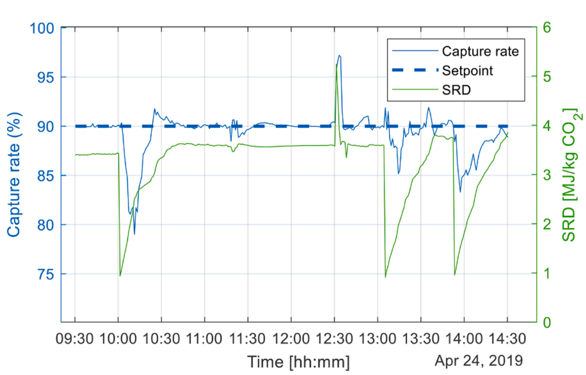

Fig. 20. Capture rate and SRD during the stripper stop test.

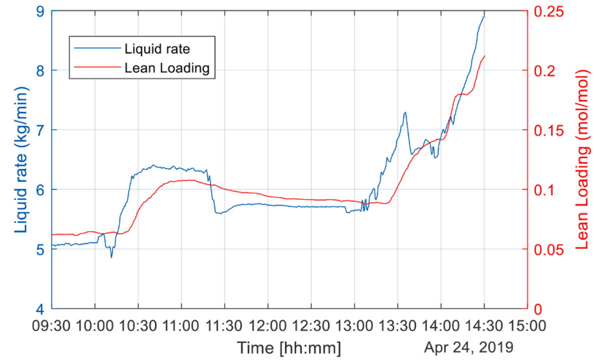

Fig. 21. Liquid rate and lean loading during the stripper stop test.

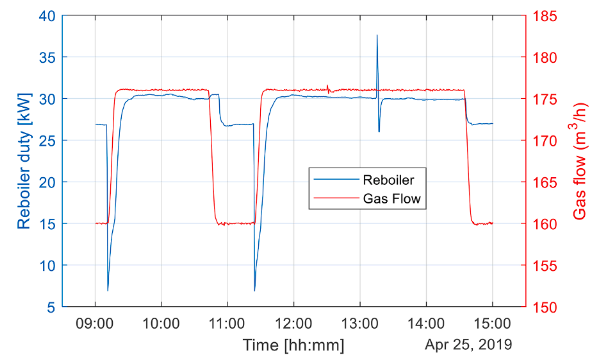

Fig. 22. The reboiler duty and gas flow.

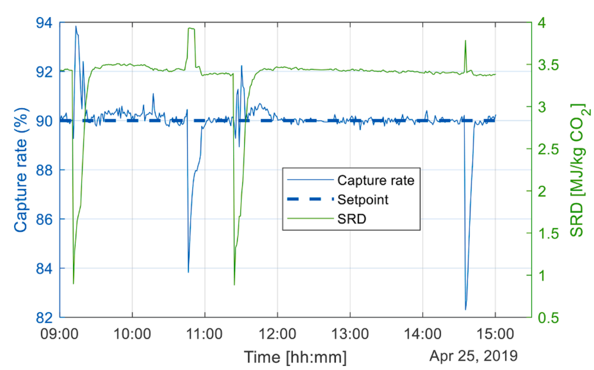

Fig. 23. Capture rate and SRD during the stripper stop test.

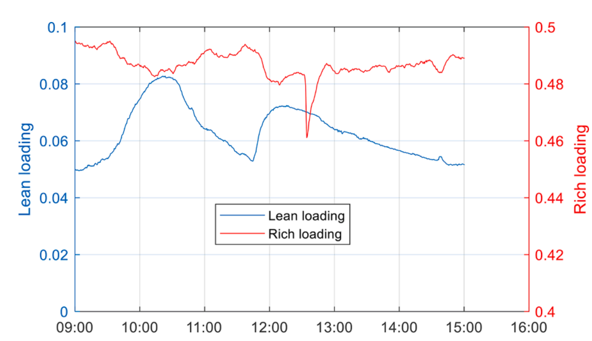

Fig. 24. Lean and rich loading during the stripper test.

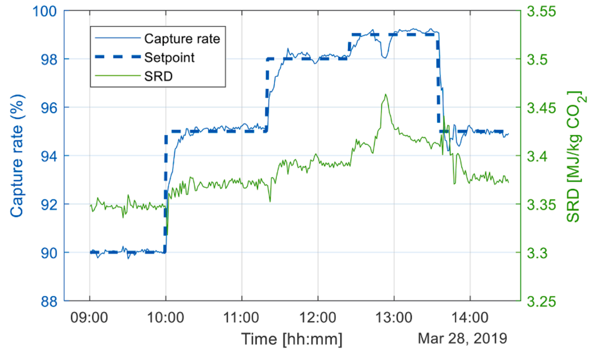

Fig. 25. Capture rate setpoint changes and resulting control performance. Capture ratio calculated based on absorber inlet and outlet CO_2 concentration in the flue gas.

without any operational challenges within the same timeframe.

Both the liquid solvent flowrate and the energy input to the stripper reboiler follows the same pattern as the capture rate setpoints as shown in Fig. 26. This is according to the expectations. The reboiler duty will increase since the increased capture rate will imply higher amounts of produced CO_2. The additional energy per kg CO_2 required due to the higher capture rate will imply higher SRD. However, this increase is rather small, as can be seen from Fig. 25. Typical values are 3.35, 3.37, 3.39 and 3.42 MJ/kg CO_2 for 90, 95, 98 and 99% capture rate.

The response in real capture rate shows that it can easily change the capture rate in the pilot plant. The NMPC effectively adjusts the liquid flowrate so that SRD is minimized as can be seen from Fig. 26.

4. Discussion

4.1. General evaluation of the control system

The NMPC based control system manipulated the liquid circulation rate and the reboiler duty setpoints in the Siemens PC7 system. The low-level controllers (typically PI) in PC7 were tuned sufficiently fast to track the setpoints well. These two manipulated variables were used to control the CO_2 capture rate and minimize the SRD in the plant.

The test results showed that the NMPC control system basically behaved according to the expectations. With the NMPC, it was possible to shift from one capture rate to another, as prescribed by the plant operator. The transitions were very smooth and the objective of minimizing the reboiler duty was obtained throughout the tests.

The NMPC control system performed also well in terms of disturbance rejection. The main disturbances studied were the flue gas flowrate and the CO_2 composition. These have an immediate effect on the absorber conditions, especially the capture rate, which is one of the most important variables of the plant. Changing the flowrate of the lean amine solvent into the absorber was an efficient way to counteract these disturbances on the vapor side of the column, while the effect on the liquid composition was much slower. Consequently, the disturbance rejection was both effective and favorable in terms of energy usage, but not necessarily without minor, temporary setpoint deviations.

The test program investigated the possibility to predict changes in the flue gas inlet flowrate, and to forward this information to the NMPC. This enables improved feed-forward control, and it is realistic to assume that a post-combustion capture plant connected to a power plant would have access to such information, i.e., changes in the power plant load. The results show that this will help to make the disturbance rejection even better. Predictability for changes in the flue gas CO_2 concentration has not been explored in this study but is expected to hold the potential for even better control through transitions.

The robustness of the NMPC control system was tested by inflicting large variations in the disturbances, pushing the plant towards its limits. The control system performed as expected both during periods with high load and periods with low load. The capture rate setpoint and the energy efficiency were obeyed to a satisfying degree throughout these tests.

The results show that the energy penalty for a higher capture rate is quite low for the CESAR1 solvent system compared to MEA, which was tested in a previous project (Hauger et al., 2019), meaning that increasing above 90% capture rate does not significantly increase the SRD. This is important both with respect to a trend towards higher (than 90%) capture rates in post combustion plants, but also for flexible operation of the capture plant. The latter is important when using a dynamic real-time optimizer (DRTO), for which it typically must be possible to alternate between very high and moderately low capture rates.

4.2. Lean loading control

The capture rate is readily controlled by changing the liquid solvent flowrate. This is a fast and responsive method, but it requires that the lean loading is kept reasonably constant by the reboiler duty. The lean loading itself is not a degree of freedom to the controller, nor was it a controlled variable with constraints or a reference, but it is an important property of the plant, which is also predicted by the dynamic process model. The lean loading was an *indirect* part of the plant optimization, in the sense that the NMPC control system implicitly determines the optimal lean loading to achieve its goals. This is illustrated in Fig. 27 which is a corresponding curve to Fig. 2 but with lean loading as x-axis instead of liquid flowrate.

The optimal L/G corresponds to a lean loading of 0.033, but larger lean loading up to 0.08 will have only slightly influence on the SRD.

In Test 4 with constrained energy input to the reboiler the NMPC was forced away from what would be the optimal conditions in the unconstrained case (Fig. 18). When the reboiler duty was constrained, the NMPC chose to reduce the flowrate of amine solvent as well (Fig. 16), leading to a decrease in capture rate and inability to obey the capture

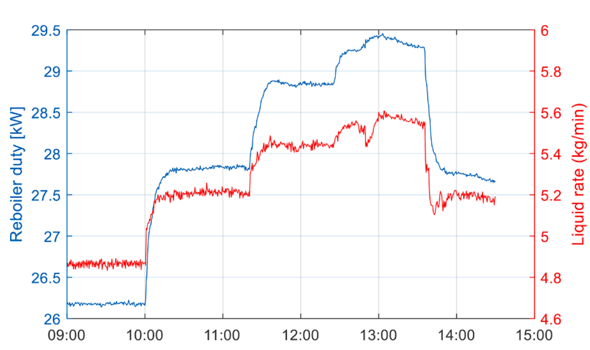

Fig. 26. Reboiler duty and liquid rate during the setpoint changes.

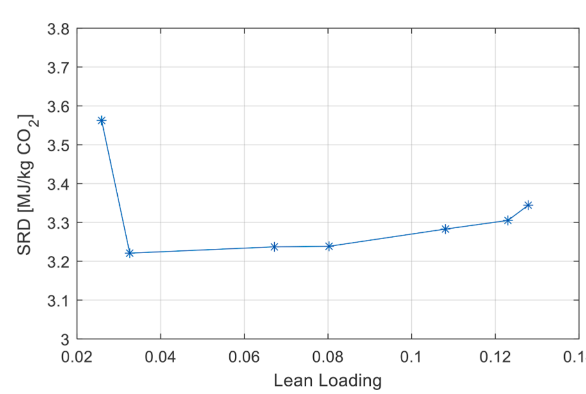

Fig. 27. Lean loading corresponding to the optimization curve (Fig. 2).

rate setpoint. That could seem counter-intuitive, but it is the optimal solution for the predicted future horizon (5 h) given the circumstances, because it is the best way to maintain the lean loading over time. The opposite strategy, which would be to force the high capture rate by slowly and steadily increasing the solvent flowrate in the lack of sufficient reboiler duty, is not sustainable and will lead to severe issues after a short time. Note that using the liquid flowrate to keep the lean loading constant is the same strategy that has been suggested in the literature when operating under steam limitations (see Section 1.3).

When the control system became unstable during the first day of "stripper stop" tests it was because it did not manage to keep the lean loading at the desired level. On the second day, the lean loading went back to its initial value before the next stripper stop test was introduced and NMPC control system managed the scenario much better. However, the time before reaching the desired lean loading after this unpredictive disturbance was more than 2 h and this is not acceptable. An explanation for this long time is that NMPC optimizes SRD, and the improvement in SRD is quite small when decreasing the liquid rate and the lean loading as shown in Fig. 27.

To improve the "stripper stop" scenario the liquid flowrate should be decreased in proportion to the available steam like it did in Test 4 and thus keep the optimal lean loading. This could be done by putting a penalty on "high" values of lean loading in the optimization objective such that deviations from the optimal value are punished harder. The effect would be reduced liquid flow rates, reduced capture rates, and thus reduced values of the lean loading. The lean loading control could be implemented by controlling the lean liquid concentration estimator or even faster by controlling the temperature profile in the desorber (Mejdell et al., 2017). We will explore such extensions of the NMPC optimization criterion in later research projects.

5. Conclusions

In the ALIGN-CCUS project, the NMPC based application was tested for the CESAR1 solvent system at the Tiller pilot. Based on the closed loop tests, the following can be concluded:

- A smooth transfer of capture rate from 85 up to 99% capture was obtained, which facilitates the possibility of using a dynamic real time optimizer that can minimize the capture cost over a longer horizon.
- Changes in flue gas flowrates due to changes in the electric output from the power plant were well controlled. NMPC could use the planned change in the flue gas flowrate in its predictions for future control steps.
- The pilot plant could handle flow changes from 130 to 340 m^3/h without any operational problems. This shows that a PCC plant can be very flexible also for part load power plants.
- The NMPC control system managed also pure disturbances like CO_2 concentrations changes well.
- The "stripper stop" concept is a realistic way to stabilize the grid frequency. However, NMPC needs to expand the control criterion to also include control of lean loading to avoid unstable conditions. The exact implementation is left for further investigation and future tests.
- The dynamic process model of the plant was found to be very well suited for control purposes.

CRediT authorship contribution statement

T. Mejdell: Conceptualization, Methodology, Investigation, Writing – original draft, Writing – review & editing. **H.M. Kvamsdal:** Project administration, Conceptualization, Writing – original draft. **S.O. Hauger:** Methodology, Software. **F. Gjertsen:** Methodology, Software. **F.A. Tobiesen:** Methodology, Software. **M. Hillestad:** Conceptualization.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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