Dynamic Simulations of the Post-combustion CO_2 Capture System of a Combined Cycle Power Plant

Rubén M. Montañés   Lars O. Nord

Department of Energy and Process Engineering, NTNU – Norwegian University of Science and Technology,
Trondheim, Norway

Abstract

Dynamic process models of the capture unit of a 600 MW combined cycle power plant with post-combustion CO_2 capture were developed in the Modelica language. The process models were utilized to understand the transient response of the capture unit when the plant was initially operated at steady-state conditions at different power plant's loads. Simulations to characterize the open-loop response of main process variables of the process to step-change disturbances in flue gas mass flow rate, solvent circulation mass flow rate and reboiler duty were performed. It was found that the plant was slower when operated at lower loads, i.e., it required longer total stabilization times for the most relevant variables of the process. Simulations revealed that the PCC unit responded significantly faster to an increase in exhaust gas mass flow rate than to a reduction in exhaust gas mas flow rate.

Keywords: transient, carbon capture, gas liquid contactors, operational flexibility, chemical absorption.

1 Introduction

CO_2 capture and storage (CCS) comprises a group of technologies that can significantly reduce the CO_2 emissions from thermal power plants and other industrial sources (IEA, 2016). Post-combustion CO_2 capture based on the chemical absorption-desorption process using amines is a technology that has been technically proven at commercial scale from coal fired power plants in projects such as Boundary Dam in Saskatchewan, Canada, and the Petra Nova project in Texas, USA.

The introduction of large shares of variable renewable energy sources such as wind and solar in power systems is changing the operating patterns of thermal power generation units, including coal power plants and natural gas combined cycle plants (IEA, 2011). Power plants traditionally operated as base load units are operated as load-following units (Montañés, et al., 2016). Therefore, during the last years, interest has grown in the field of operating flexibility of thermal power plants with CO_2 capture technologies (IEA-GHG, 2012).

The low amount of existing commercial-scale post-combustion capture plants (PCC) and the scarcity of published transient performance data of such systems claims for an interest for the development of dynamic process models (Bui, et al., 2014). These models allow studying plant dynamic performance, analyzing various plant transient events as well as developing and implementing optimal control strategies for PCC plants integrated with thermal power plants. Dynamic process simulation provides process insight and contributes to the development of the learning curve for flexible operation of future thermal power plants with CO_2 capture.

The purpose of the study is to provide understanding of the open-loop transient performance of the main process variables of the PCC unit at different load operation points of the power plant. A thermal power plant operated as load-following unit will be operated at part-load conditions during a significant amount of hours during its lifetime (Montañés, et al., 2016). Therefore it is of importance to find out differences in the transient behavior of the process at part-load operating conditions with respect to those of nameplate capacity. In this work, a dynamic process model of the PCC unit of a 600 MW combined cycle power plant with post-combustion CO_2 capture using aqueous monoethanolamine (MEA) as chemical solvent is utilized for providing understanding of the open-loop response of key performance variables to different disturbances applied to the PCC plant. The process insight and understanding developed in this work will be valuable to develop control strategies of the process when integrated with the thermal power plant.

2 Post-combustion CO_2 capture with chemical absorption

2.1 Chemical absorption process

The process of CO_2 capture by chemical absorption is a two-steps regenerative process; one involves the absorption of CO_2 into a solvent, while the other involves the desorption or stripping of CO_2 from the solvent and the regeneration of the solvent. The process conditions change in the absorption and desorption

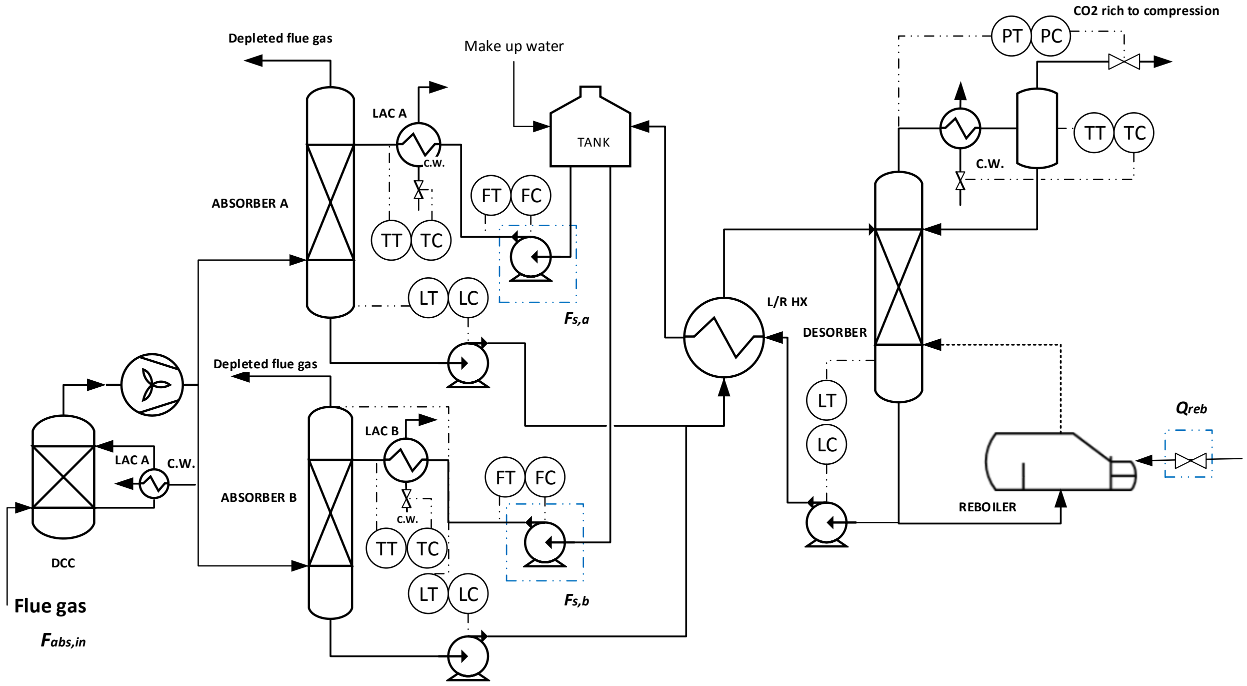

Figure 1. Process configuration of the post combustion CO_2 capture unit (PCC) of the natural gas combined cycle power plant studied in this work. Includes temperature (T), level (L), flow (F) and pressure (P); transmiters (T) and controllers (C).

process, being main changes temperature and pressure, and also solvent concentrations and pH. For absorption, low temperature and high partial pressure of CO_2 is desired, while for desorption, high temperature and low partial pressure of CO_2 is desired.

When the process is utilized for flue gas treatment from a power plant, the exhaust gases are normally cooled down by means of a direct contact cooler (DCC), that reduces the flue gas temperature and the water content. A fan allows overcome the gas pressure drop in the absorber, which is operated slightly above atmospheric conditions, and at around 40 ºC, refer to Figure 1. In the absorber column, the exhaust gas flowing upwards meets the chemical solvent flowing downwards. Packing material allows having a thin film of liquid with high surface contact area for heat and mass transfer between the gas and liquid phases, and the exothermal chemical absorption process. Depleted flue gas leaves the absorber at the top through a stack, normally after flowing through a water wash section that allows keeping the water mass balance of the process and reduces chemical solvent emissions due to solvent droplets or solvent vapor carry over. The rich solvent, i.e., solvent with a lot of bounded CO_2, accumulates in the absorber sump and is then pumped towards the top of the stripper. An intermediate heat exchanger allows for heat integration between the absorber and stripper columns. The rich solvent is heated up by the lean solvent coming from the stripper bottom towards around 110 ºC and then enters the stripper at the top of the column. This heat integration allows reducing reboiler and cooling duties. A mixing tank allows for accumulation of the solvent at different operating conditions of the plant.

The desorption process normally occurs at around 100 to 130 ºC. Steam supplied from the power plant provides the reboiler duty required to regenerate the solvent (endothermal desorption process), and to generate the stripping vapors flowing upwards in the stripper column, consisting mainly of H_2O and CO_2. The regenerated lean solvent is sent to the absorber inlet via the heat integration exchanger and a lean amine cooler that controls the temperature of the solvent at the inlet of the absorber to around 40 ºC. At the top of the stripper there is a condenser and a cooler where the solvent and steam condenses. The condensate is conducted back to the column via a reflux. The product CO_2 rich flow the top of the stripper is conducted to the compression section where it will be conditioned for transport and storage purposes.

2.2 Process configuration

The PCC unit was designed to treat flue gas from a 611 MW combined cycle power plant. The gas turbine (GT) of the power plant was the heavy duty Mitsubishi JAC 701, and the steam cycle consisted of a three-pressure reheat (3PRH) configuration. The design of the PCC unit included the process integration with the power plant through the flue gas line from the HRSG outlet and a steam extraction from the steam turbine's IP/LP crossover. The steam extraction was utilized to feed the reboiler duty required to produce the stripping vapors needed for chemical desorption in the stripper column. The design point chosen for the post-combustion unit was 100% GT load under ISO conditions, which, for the gas turbine, corresponded to flue gas with a mass flow rate of 887.1 kg/s with 4.33 vol % CO_2 (wet). The chemical solvent utilized was 30%wt aqueous MEA and the target capture rate was 90%. Further details on design aspects of PCC units for combined cycle power plants can be found in (Dutta, et al., 2017).

The resulting process configuration of the PCC unit consisted of a two absorbers and one stripper layout, as shown in Figure 1. Each absorber column had dimensions of 16.3 m in diameter and 23.2 m height, while the desorber had a 9.7 m diameter with 10 m height. The process equipment included absorber columns, desorber column and reboiler, overhead condenser, internal lean/rich heat exchanger, mixing tank for water and MEA makeups, direct contact coolers and circulation pumps. A fan was included in the process to overcome the pressure drop imposed by the absorber column.

3 Dynamic process model development and validation

The Modelica library Gas Liquid Contactors (GLC) (Modelon AB, 2016), from Modelon AB, was utilized as a basis to develop the dynamic process model of the PCC unit. The library contains dynamic process models of the main equipment for systems' level modeling of the absorber-desorber process with monoethanolamine (MEA) as chemical solvent. That equipment includes absorber and desorber columns, sumps, reboiler, condensers, water wash sections, pumps, valves, mixing tank, and property media packages.

The chemical absorption-desorption process within packed segments was modelled considering the two-film theory approach for heat and mass transfer. Chemical equilibrium for reactions was assumed, and mass transfer was modeled considering rate-based models with enhancement factor (Kvamsdal, et al., 2009). Detailed description of the dynamic process models included in the GLC library has been presented previously in literature (Prölß, et al., 2011).

The dynamic process models included in the GLC library have been previously validated with large-scale experimental data by (Montañés, et al., 2017). The validation consisted of modeling the whole absorber-desorber system of the demonstration scale chemical absorption plant at CO_2 Technology Centre Mongstad (TCM DA), in Norway. The amine plant at TCM DA was configured to treat exhaust gases coming directly from the exhaust of a natural gas fueled combined heat and power (CHP) plant placed at Mongstad's refinery. The exhaust gas from two GE 9001E gas turbines contains about 3.5 %vol CO_2, and around 3% of the total exhaust gas mass flow rate is conducted to the amine plant for CO_2 absorption. The PCC plant at TCM can treat up to 60 000 Sm^3/hr of exhaust gas and can capture around 80 ton CO_2/day at nameplate capacity when configured to treat CHP gas. The experimental data utilized for validation includes steady-state data for a wide range of operating conditions and multiple transient events. The plant was operated with 30 wt % aqueous MEA. The conclusion of the work in (Montañés, et al., 2017) is that the process models can capture, with sufficient accuracy, the steady-state and transient phenomena of the process at the demonstration plant scale. In addition, it gives confidence towards using the models for simulation and analysis of the transient performance of the scaled-up process to commercial scale of 4770 ton/day CO_2 captured.

Rules for consistent inventory control (Aske & Skogestad, 2009) were applied to design the regulatory control layer of the PCC unit in Figure 1. It included level controllers for absorbers and stripper sumps, overhead condenser pressure control, lean solvent temperature at absorbers inlet, and exhaust gas temperature at absorber inlet. The controllers were tuned by means of the SIMC tuning rules.

The supervisory control layer for this process has three degrees of freedom, consisting of the two solvent mass flow rates at absorber inlet $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$, and the reboiler duty $\dot{Q}_{reb}$.

4 Process simulations description

Generally, a combined cycle power plant is brought to part-load operating conditions by reducing the GT load and consequently the steam turbine's power output will be reduced. A GT load reduction results in reduced GT exhaust gas mass flow rate sent to the HRSG of the combined cycle and to the absorbers of the PCC unit. The open-loop transient performance of the plant is studied for three steady-state operating conditions of the power plant, corresponding to 100%, 80% and 60% GT load.

4.1 Steady-state operating conditions at 100%, 80% and 60% GT load

In order to obtain the steady-state operating conditions of the PCC unit at the three operating points, simulations were run with different flue gas mass flow rates as input boundary conditions to the dynamic process model, corresponding to different GT loads, refer to Table 1. The exhaust gas temperature and composition of the absorber was considered constant as boundary condition (input). Note that the exhaust temperature at the inlet of the absorber is normally controlled by the DCC, and that it was observed that exhaust gas composition did not change considerably for the purpose of this study, considering the part load range analyzed of 100% to 60% GT load, and for the specific GT utilized in this work. In addition, a decentralized control structure for the supervisory control layer was included. Several studies, including the one based on self-optimizing control theory by (Panahi, 2011), suggest that keeping the capture ratio *Cap* and a temperature in the stripper column constant can lead to efficient operation of the process for varying loads of the absorption-desorption process. Therefore, the available degrees of freedom for operation where utilized to control these process variables. Solvent mass flow rates $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$ were utilized to control the respective CO_2 capture rates Cap_a and Cap_b at the top of the absorbers to the design value of 0.9, while reboiler duty was used as manipulated variable to control reboiler temperature T_{reb} to the value 119 °C. CO_2 capture rates are calculated for each absorber column at the top, by using Equation (1), where $\dot{F}_{abs,in}$ is the exhaust flue gas at the inlet of the absorber column, $X_{abs,in}$ is the CO_2 mass fraction in the exhaust gas at the absorber inlet, $\dot{F}_{abs,out}$ is the depleted flue gas mass flow rate at the absorber stack and $X_{abs,out}$ is the CO_2 mass fraction in the flue gas at the absorber stack. The resulting operating conditions of the PCC at different GT loads are shown in Table 1 and Table 2.

Table 1. Values of PCC unit input variables at different power plant's load operating conditions. Note that both absorber columns were operated in parallel, so $\dot{F}_{s,a}$ was equal to $\dot{F}_{s,b}$.

GT load [%]	$\dot{F}_{abs,in}$ *[kg/s]*	$\dot{F}_{s,a}$ *[kg/s]*	$\dot{Q}_{reb}$ *[MW]*
100	887.1	613.3	205.9
80	765.1	535.2	176.2
60	653.5	464.1	149.6

$$Cap_a = \frac{\dot{F}_{abs,in} \cdot X_{abs,in} - \dot{F}_{abs,out} \cdot X_{abs,out}}{\dot{F}_{abs,in} \cdot X_{abs,in}} \qquad (1)$$

Table 2. Values of most relevant process variables of the PCC unit at different operating conditions of the power plant. Note that both absorber columns were operated in parallel, so Cap_a was equal to Cap_b (in the table shown as *Cap*). It also resulted in same value of solvent loading at absorbers inlets ($L_{i,abs}$).

GT load [%]	*Li,abs*	*Li,str*	*Cap*	*Prod [kg/s]*
100	0.280	0.501	0.9	55.2
80	0.280	0.497	0.9	47.6
60	0.279	0.493	0.9	40.7

4.2 Open-loop step response simulations

The simulations consisted of step-changes of ±10% of main PCC inputs, or disturbances, when the plant was at steady-state operating conditions at the three GT operating points. Step-changes were applied to each process input at a time, keeping the remaining process inputs constant. The output in main process variables was recorded and dead times and 10% settling times were calculated.

- Dead time Θ describes how long it takes before a process variable begins to respond to a change in the process input. With begins to respond it is meant that the trajectory of the process variable moves out of the band defined by the initial steady-state value of the process variable y_0, and a ±1% change in the process variable Δy, *i.e.:* $-0.01\,\Delta y + y_0 < y_0 < 0.01\,\Delta y + y_0$, for the first time.
- The 10% settling time t_s is the time it takes from the instant in which the process variable begins to respond to the input change, until it remains within an error band described by the final steady-state value of the process variable y_∞, and 10% of the change in the process variable Δy, i.e.: $-0.1\,\Delta y + y_\infty < y_\infty < 0.1\,\Delta y + y_\infty$.
- The resulting total stabilization time t_{sta} is the sum of the dead time and the settling time. In addition, the relative change *RC* in the process variable is calculated as in Equation (2), where y_0 is the initial steady-state value of the process variable.

$$RC\ (\%) = 100 \cdot \frac{y_\infty - y_0}{y_0} \qquad (2)$$

The main inputs/disturbances applied to the process in this analysis were:

- Flue gas mass flow rate $\dot{F}_{abs,in}$. Note that the flow was split and the absorber columns were operated in parallel. This means that each absorber column treated an exhaust gas mass flow rate of $\dot{F}_{abs,in}/2$.
- Solvent mass flow rates at absorbers inlets $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$.
- Reboiler duty $\dot{Q}_{reb}$.

The responses of the main process variables of interest in this analysis were:

- Solvent lean CO_2 loading at absorbers inlet $L_{i,abs}$.
- Solvent rich CO_2 loading at stripper inlet $L_{i,str}$.
- CO_2 capture rate at absorbers stacks Cap_a and Cap_b.
- CO_2 product mass flow rate *Prod*, at the outlet of the overhead condenser of the desorber. This is the CO_2 rich product flow of the PCC unit sent to conditioning, compression and transport.
- Temperature at stripper column bottom T_{reb}.

The difference in solvent loading at inlet and outlet of the absorber determines the capability of the solvent to carry CO_2. This in turn depends on the absorber size, operating conditions, regeneration of the solvent and CO_2 partial pressure. Solvent CO_2 loading *L* is defined as the ratio between moles of CO_2 and moles of solvent (mol/mol) in Equation (3).

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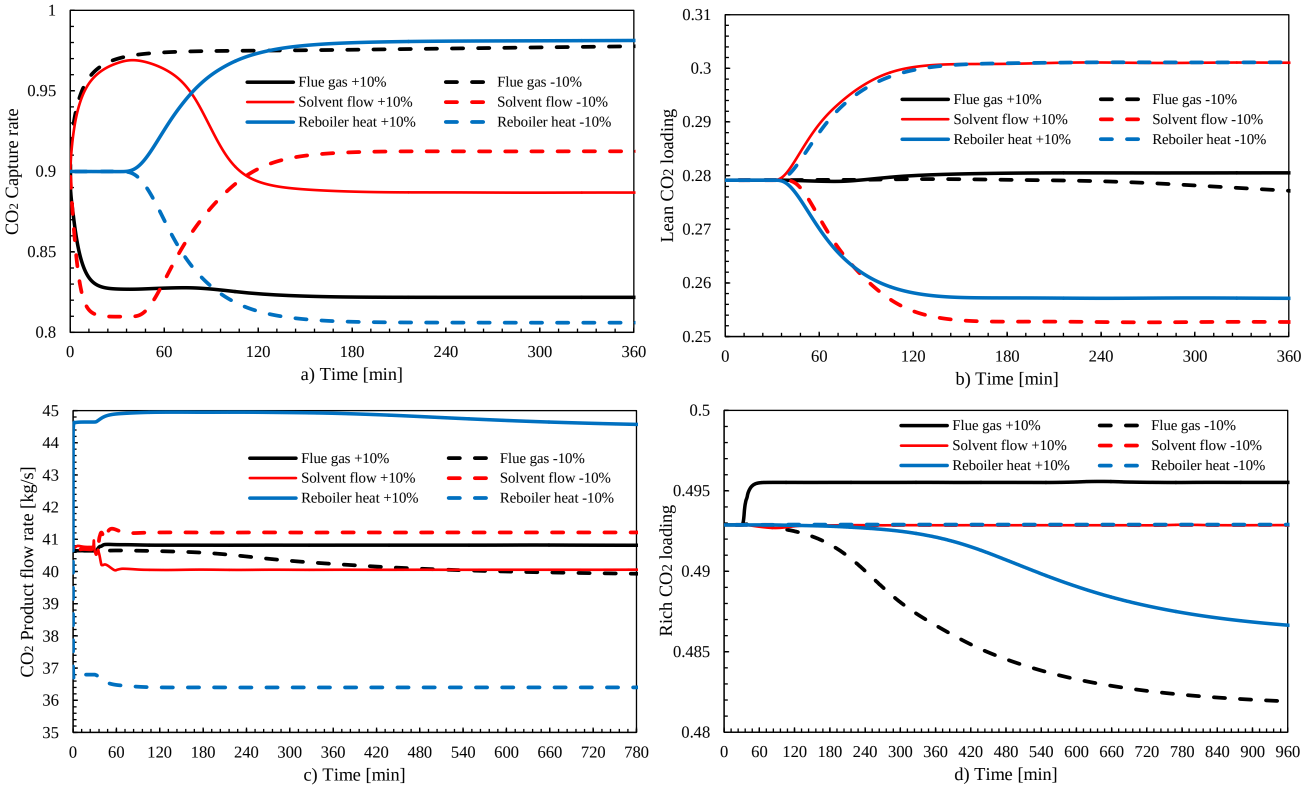

Figure 2. Transient responses of the relevant process variables to different step-changes in process inputs. These simulations correspond to the initial steady-state operation of the PCC unit for 60% GT load. Step-changes were applied at t = 0 min.

$$L = \frac{mols\ of\ CO_2}{mols\ of\ solvent} \qquad (3)$$

5 Results and discussion

5.1 Response to step changes in flue gas mass flow rate $\dot{F}_{abs,in}$

The resulting response times of the PCC unit's main process variables to step-changes in flue gas mass flow rate are shown in Table 3 and Table 4. Figure 2 shows the transient response of the main process variables for the different step changes studied in this work. In addition, Figure 3 shows trends of total stabilization times t_{sta} for the main variables of the process when operating the plant at different loads.

It can be observed that CO_2 capture rate *Cap* stabilized relatively fast, within 1 h, after a disturbance in flue gas mass flow rate. The CO_2 capture rate decreased for increased flue gas mass flow rate (+10%). A faster response in *Cap* was observed when the flue gas flow rate was increased (+10%) than when it was decreased (-10%), showing the non-linear performance of the PCC system. This behavior was consistent at the different operating points of the PCC plant. The dead time of this response was negligible, since the flue gas mass flow rate was included in the calculation and naturally changes when a step change is applied.

The CO_2 product flow rate *Prod* required larger stabilization times than *Cap*. This shows the differences in performance of the absorbers and desorber columns during transient conditions when a disturbance is applied to the PCC unit. The dead times observed in the CO_2 product mass flow rate can be explained by the residence time imposed by the solvent hold-ups in the cold side of the internal heat exchanger's piping and rich solvent piping. These residence times resulted in dead times in convectively transported variables of the liquid solvent from absorber outlet to stripper inlet, including rich solvent loading at the stripper inlet $L_{i,str}$. Note that the dead times of $L_{i,str}$ and *Prod* responses are similar in Table 3 and Table 4. Stabilization of the *Prod* was significantly faster when increasing flue gas mass flow rate (around 1 h) than when flue gas mass flow rate was decreased (9 to 11 h). It can also be observed that the *Prod* response was slower at lower power plant loads, refer to Figure 3.

For flue gas flow rate increase (+10%), the relative change in solvent loadings was small. This is because the solvent capacity was close to the limit under these operating conditions. In general, it was found that lean solvent loading at the inlet of the absorber $L_{i,abs}$ required larger stabilization times t_{sta} than rich loading at stripper inlet $L_{i,str}$. This can be explained by the buffering effect introduced by the mixing tank placed in the recycle loop (from stripper sump to absorber liquid inlet). In addition, larger dead times to this specific disturbance were found for $L_{i,abs}$ than for $L_{i,str}$, due to the additional

Table 3. Response to +10% step increase in flue gas mass flow rate $\dot{F}_{abs,in}$ at various GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	74	53	127	0.45
	$L_{i,str}$	26	11	36	0.45
	Cap	0	12	12	-8.79
	Prod	26	5	31	0.29
	T_{reb}	0	127	127	-0.05
80	$L_{i,abs}$	36	105	141	0.47
	$L_{i,str}$	27	12	39	0.50
	Cap	0	15	15	-8.74
	Prod	28	45	72	0.35
	T_{reb}	0	54	54	-0.05
60	$L_{i,abs}$	68	88	156	0.48
	$L_{i,str}$	34	13	46	0.54
	Cap	0	17	17	-8.70
	Prod	34	28	62	0.42
	T_{reb}	0	60	60	-0.06

Table 4. Response to -10% step decrease in flue gas mass flow rate $\dot{F}_{abs,in}$ at various GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		-10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	47	578	626	-2.50
	$L_{i,str}$	26	538	564	-2.62
	Cap	0	55	55	8.55
	Prod	27	556	583	-0.98
	T_{reb}	3	572	575	0.30
80	$L_{i,abs}$	50	639	689	-2.31
	$L_{i,str}$	29	592	621	-2.53
	Cap	0	58	58	8.84
	Prod	30	603	633	-1.90
	T_{reb}	0	625	625	0.27
60	$L_{i,abs}$	129	619	748	-1.99
	$L_{i,str}$	131	529	661	-2.26
	Cap	0	39	39	8.88
	Prod	163	503	666	-1.85
	T_{reb}	5	667	672	-2.83

residence time introduced by liquid hold-ups in desorber packed segments and sump, lean amine piping and hot side piping of the integral heat exchanger, mixing tank and lean amine cooler. Again, the plant response in solvent CO_2 loadings was faster when flue gas mass flow rate was increased for all power plant loads studied, refer to Figure 3. It must be mentioned that the relative change in process variables to step-changes is more significant the step-down than step-up of the flue gas flow rate. This can be explained by the fact that the solvent rich loading at the steady-state operating conditions is close to the solvent limit CO_2 loading capacity, which is limited by stoichiometry.

5.2 Response to step-changes in solvent mass flow rate $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$

The resulting response times of the PCC unit's main process variables to step-changes in solvent circulation mass flow rates are shown in Table 5 and Table 6.

Table 5. Response of the main process variables to 10% step increase in solvent circulation mass flow rate $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$ at the inlet of the absorbers, for different GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	27	50	77	8.04
	$L_{i,str}$	25	118	143	-0.02
	Cap	0	131	131	-1.46
	Prod	21	40	60	-1.48
	T_{reb}	0	25	25	-1.04
80	$L_{i,abs}$	31	113	144	8.19
	$L_{i,str}$	37	112	149	-0.01
	Cap	0	137	137	-1.77
	Prod	25	21	46	-1.78
	T_{reb}	0	35	35	-1.05
60	$L_{i,abs}$	35	67	102	7.85
	$L_{i,str}$	29	813	842	0.00
	Cap	0	161	161	-1.47
	Prod	31	22	52	-1.46
	T_{reb}	0	39	39	-0.99

Solvent circulation mass flow rate step changes resulted in inverse responses in CO_2 capture rates, refer to Figure 2. This can be explained by the coupled operation of the absorbers and desorber columns via the recycle loop. When increasing the solvent circulation flow rate (10%), the *Cap* increases during the first part of the transient. However, since the reboiler duty is kept constant, the lean loading at the inlet of the absorber $L_{i,abs}$ will increase (more solvent being circulated for the same regeneration energy introduced in the process $\dot{Q}_{reb}$), resulting in a reduction of *Cap*, with a delay imposed by solvent hold-ups (residence time) through piping and mixing components in the recycle loop.

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Observe the large dead time in $L_{i,abs}$ in Figure 2. An analog explanation could be used for the inverse response observed when solvent circulation mass flow rate was reduced. Larger stabilization times were required when the plant was operated at lower loads, see Figure 3.

For these disturbances, CO_2 product mass flow rate *Prod* stabilizes relatively faster (around 1 h) than CO_2 capture rate *Cap* (2–3 h). Similar stabilization times t_{sta} were noted when increasing (10%) and when decreasing (-10%) the solvent circulation mass flow rates $\dot{F}_s$.

The relative change in stripper inlet rich solvent loading $L_{i,str}$ was very small, so it can be considered constant when changing the solvent circulation rate by 10%. It shows that the solvent's capacity was working at the limit. However, lean loading $L_{i,abs}$ relative change was large. A large dead time was observed in $L_{i,abs}$ (27-47 minutes), due to the large amount of solvent inventory within the plant (residence time), and in the recycle loop. In addition, a settling time of 1 to 2 hours was observed, this is likely due to the buffering effect introduced by the absorber tank and other mixing components, such as, desorber and absorber sumps.

Table 6. Response of the main process variables to -10% step decrease in solvent circulation mass flow rate $\dot{F}_{s,a}$ and $\dot{F}_{s,b}$ at the inlet of the absorbers, for different GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		-10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	35	53	88	-10.26
	$L_{i,str}$	29	150	180	0.00
	Cap	0	118	118	2.05
	Prod	26	29	55	2.03
	T_{reb}	0	8	8	1.09
80	$L_{i,abs}$	35	74	108	-10.60
	$L_{i,str}$				0.00
	Cap	0	125	125	2.11
	Prod	30	33	64	2.85
	T_{reb}	0	37	37	1.13
60	$L_{i,abs}$	43	71	115	-9.47
	$L_{i,str}$	37	788	825	0.00
	Cap	0	166	166	1.38
	Prod	35	28	63	1.40
	T_{reb}	0	9	9	0.99

5.3 Response to step-changes in reboiler duty $\dot{Q}_{reb}$

Simulations in which reboiler duty $\dot{Q}_{reb}$ was changed with step-changes by ± 10% were performed. Flue gas conditions and solvent circulation flow rates were kept constant at each operating point of the plant. The resulting response times of the PCC unit's main process variables are shown in Table 7 and Table 8.

Table 7. Response of the main process variables to 10% step increase in reboiler duty $\dot{Q}_{reb}$, for different GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	28	384	412	-10.00
	$L_{i,str}$	172	526	697	-1.80
	Cap	31	69	100	9.54
	Prod	0	322	322	9.70
	T_{prod}	0	335	335	1.09
80	$L_{i,abs}$	33	419	451	-9.42
	$L_{i,str}$	247	531	778	-1.52
	Cap	35	67	102	9.06
	Prod	0	332	332	9.67
	T_{reb}	0	353	353	1.02
60	$L_{i,abs}$	37	457	494	-8.91
	$L_{i,str}$	335	539	874	-1.34
	Cap	40	87	126	9.24
	Prod	0	606	606	9.44
	T_{reb}	0	368	368	0.96

Table 8. Response of the main process variables to -10% step decrease in reboiler duty $\dot{Q}_{reb}$, for different GT loads. Dead times Θ, settling times t_s and total stabilization times t_{sta} are shown.

		-10 %			
GT load [%]		***Θ [min]***	***t_s [min]***	***t_{sta} [min]***	***RC [%]***
100	$L_{i,abs}$	28	56	85	8.448
	$L_{i,str}$	44	694	739	0.008
	Cap	29	52	81	-10.81
	Prod	0	24	24	-10.84
	T_{reb}	0	11	11	-1.10
80	$L_{i,abs}$	29	66	96	8.1519
	$L_{i,str}$	92	685	777	0.0058
	Cap	34	65	99	-10.62
	Prod	0	27	27	-10.63
	T_{reb}	0	13	13	-1.05
60	$L_{i,abs}$	37	72	109	7.88
	$L_{i,str}$	93	403	496	-0.006
	Cap	39	75	113	-10.44
	Prod	0	33	33	-10.44
	T_{reb}	0	12	12	-1.01

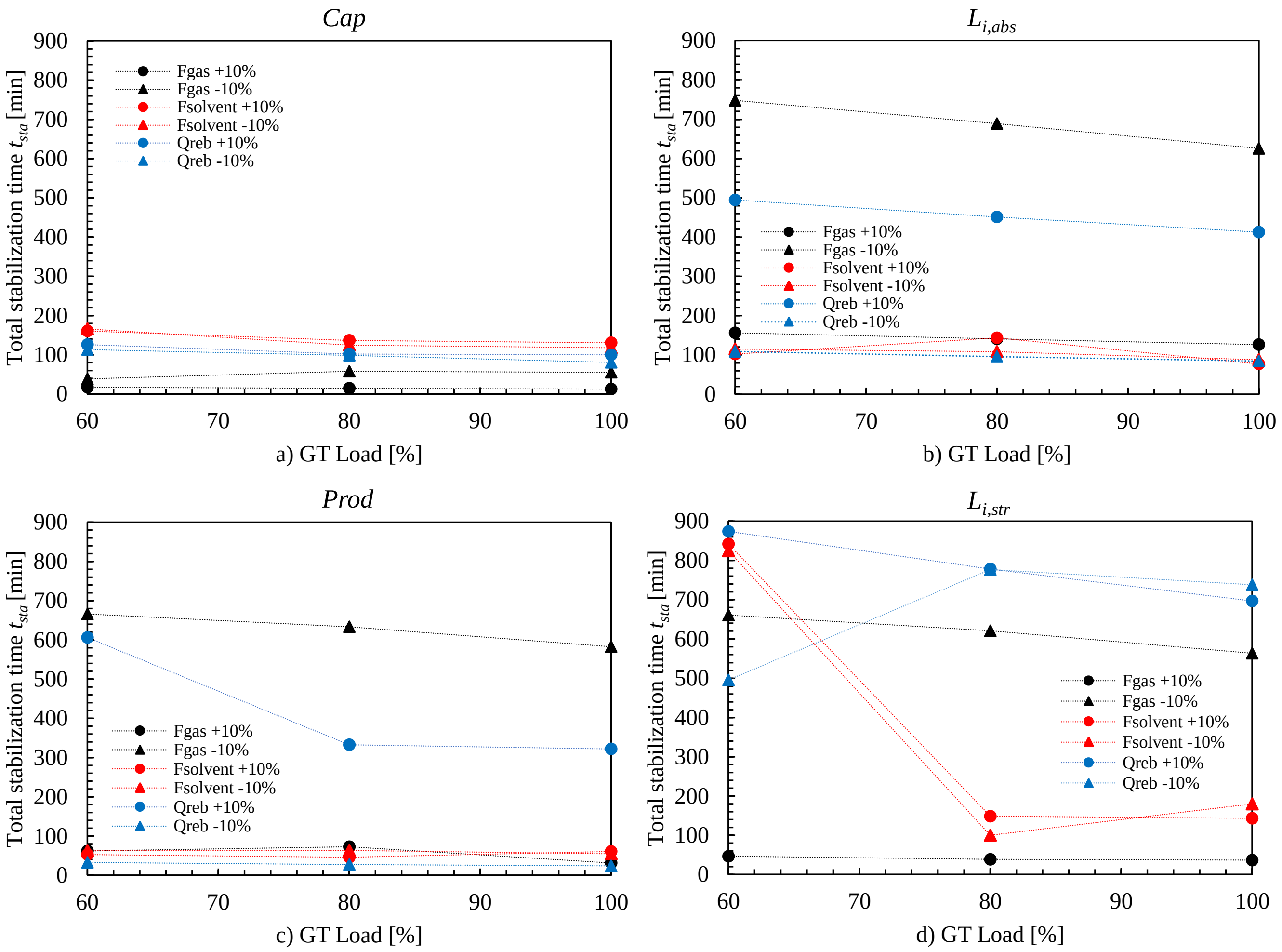

Figure 3. Trends in total stabilization times of main process variables of the PCC unit, when disturbed by the different plant input step changes, at different GT loads. a) CO_2 capture rate *Cap*; b) Solvent CO_2 loadings at absorbers inlets $L_{i,abs}$; c) Product CO_2 mass flow rate *Prod*; and d) solvent CO_2 loading at stripper inlet $L_{i,str}$.

Increasing the reboiler duty will result in increased CO_2 capture rate *Cap* due to the lower resulting lean loading at the inlet of the absorber $L_{i,abs}$. Reducing reboiler duty will result in reduced *Cap* due to the increase in $L_{i,abs}$. A relatively large dead time in the *Cap* response of 28–37 min was found. This dead time was larger when the plant was operated at lower power plant loads. This is because at lower power plant loads solvent circulation rates are smaller (refer to Table 1), resulting in larger residence time though piping and mixing tank in the recycle loop.

The relative change in CO_2 product mass flow rate *Prod* was also large, but with practically no dead time. This is because the reboiler duty introduced in the reboiler is physically closer to the overhead of the stripper. However, the recycle loop and coupled operation of the absorber and desorber makes the total stabilization time t_{sta} of the *Prod* longer than for *Cap*. Observe the slow response in $L_{i,str}$ in Figure 3. In general, longer total stabilization times were found for both *Cap* and *Prod* when the plant was operated at lower loads, refer to Figure 3.

The relative change was also significant for lean loading at absorber inlet $L_{i,abs}$ with a large dead time, as previously mentioned. The dead times were even larger for rich loading at the inlet of the stripper $L_{i,str}$, and longer total stabilization times than for $L_{i,abs}$ were observed.

6 Conclusions

The open–loop transient performance of the main process variables of the plant were studied when the plant was operated at different power plant's load conditions, and for different disturbances to the PCC unit. In general, it is found that the plant was slower when the plant was operated at lower loads, i.e., it required longer total stabilization times for the main variables of the process. In general, CO_2 capture rate stabilized relatively faster (1–3 h) than other process variables (1–11 h).

In addition, it was found that the PCC unit responded significantly faster to the increase in flue gas mass flow rate than to reductions in flue gas mas flow rate. This could have significant implications on efficient operation of the PCC unit when ramping down the power plant's load, due to long stabilization times require of the process and the resulting inefficient operation during transient conditions, if a suitable control structure cannot be implemented.

Process variables respond differently to different disturbances. For the same process disturbance and process variable, the response was different when increasing or decreasing the input. This shows the non-linear behavior of the process. The recycle loop in the process from desorber outlet to absorber inlet connects the operation of the absorbers units and the stripper, and the resulting dynamic interaction between the absorption and desorption unit resulted in long stabilization time of main process variables, up to 11 h.

Current and future work includes the integration of the PCC unit with a dynamic process model of the power plant. That will allow the study of dynamic interactions between the power plant and the PCC unit under transient events of the power plant, and to analyze optimal control structures and operation of the integrated process for efficient flexible operation.

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