Chemical Engineering and Processing 52 (2012) 112–124

Contents lists available at SciVerse ScienceDirect

Chemical Engineering and Processing: Process Intensification

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

Economically efficient operation of CO_2 capturing process. Part II. Design of control layer

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ARTICLE INFO

Article history:
Received 24 February 2011
Received in revised form 18 July 2011
Accepted 8 November 2011
Available online 17 November 2011

Keywords:
Plantwide control
Stability
Decentralized controllers
MPC
Large disturbances

ABSTRACT

In part I of this study, control structures were proposed for different operational regions of a post-combustion CO_2 capturing process using the top-down steady-state economic part of the plantwide procedure. In the current study, the bottom-up part of the complete procedure is considered. For this purpose, dynamic simulation is used to validate the proposed control structures. Different control configurations using decentralized controllers and model predictive control (MPC) are considered. At the end, a simple control configuration is proposed which keeps the process close to the optimum in all operational regions without the need for switching the control loops.

1. Introduction

We study optimal operation of a post-combustion CO_2 capturing process, where the objective is to minimize the sum of the energy cost and penalty cost for releasing CO_2 to the atmosphere. In part I of this work [1], a top-down analysis of the complete plantwide control procedure (Table 1) was performed to identify different operational regions of active constraints as a function of the throughput (flow rate of flue gas) and to select self-optimizing controlled variables (CVs) in each region. In region I, the flue gas flow rate is given at its nominal value and there are two unconstrained degrees of freedom (DOF), which may be considered as the CO_2 recovery in the absorber and the CO_2 mole fraction at the bottom of the stripper. The best associated CVs were found to be the CO_2 recovery in the absorber and the temperature on tray no.16 in the stripper [1] (see Fig. 1). However, these CVs are not necessarily the best in all operating regions and for larger flowrates of the flue gas (region II), where the reboiler duty reaches its maximum, the temperature of tray 13 in the stripper was found to be the best unconstrained CV [1]. For even larger flue gas flowrates, one reaches the minimum allowable CO_2 recovery of 80% and we have reached the bottleneck where a further throughput increase is infeasible (region III) [1].

In this work, we focus on the bottom-up part of the procedure in Table 1, where we want to identify a control structure (as simple as possible) that implements in practice the steady-state control objectives from part I over the entire feasible throughput range. A main issue is to handle transition from region I to region II where an unconstrained degree of freedom is lost due to a constraint becoming active. In general, some logical switching or reconfiguration of control loops would be necessary to manage the transition. Here, using prior knowledge of the constraint that becomes active, we synthesize a simple control structure that does not require any loop reconfiguration and thus provides near-optimal operation over the entire feasible throughput range. The details on how we arrived at the chosen pairings are the main issue for the present paper.

Towards the synthesis of such a simple structure, four different control configurations (including the one in Fig. 1) using decentralized PI controllers, briefly summarized below, are considered.

- Alternative 1: The two unconstrained self-optimizing CVs for region I are controlled using the most obvious pairings (Fig. 1).
- Alternative 2: The two self-optimizing CVs for region I are controlled using the reverse pairings compared to Alternative 1.
- Alternative 3: The best self-optimizing CV for region II is controlled.
- Alternative 4: Recommended modification of Alternative 2 which provides near optimal operation over the entire throughput range (regions I and II).

Also, the possibility of using multivariable control is considered.

To validate the proposed control structures, dynamic process simulation is performed in UniSim [3].

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doi:10.1016/j.cep.2011.11.004

Table 1
Plantwide control procedure [2].

I. Top-down part (focus on steady-state economics)
Step 1. Define operational objectives (economic cost J to be minimized) and constraints
Step 2. Identify degrees of freedom (MVs) and optimize the operation for important disturbances (offline analysis) to identify regions of active constraints
Step 3. Each region of active constraints: select primary (economic) controlled variables CV1:
-Active constrains
-"Self-optimizing" CV1s for the remaining unconstrained degrees of freedom
Step 4. Select location of throughput manipulator (TPM)
II. Bottom-up part (focus on dynamics)
Step 5. Choose structure of regulatory control layer (including inventory control)
a. Select "stabilizing" controlled variables CV2
b. Select inputs (valves) and "pairings" for controlling CV2
Step 6. Select structure of supervisory control layer
Step 7. Select structure of (or need for) optimization layer (RTO)

There are some other works that study the dynamic behavior of the CO_2 capturing processes ([4–7]) but the current study is the first to consider different operational regions and synthesizes a simple control structure that works in the different regions.

In another study [8], we designed a control structure for a CO_2 capturing process where the objective function was to minimize energy requirement with fixed CO_2 recovery (90%). In the current study, an economic penalty on the CO_2 released to the air is further imposed, which makes it optimal to remove higher amounts of CO_2 (~95%). However, at higher flue gas rates, when the capacity constraint for the stripper is reached, the CO_2 recovery will drop, and at the capacity bottleneck it will reach the minimum allowed recovery which is set to 80%. The details about the objective function, optimization and selection of the best self-optimizing controlled variables are given in part I of this study [1].

The article is organized as follows. In Section 2, design of the control loops for primary (CV1) and secondary (CV2) controlled variables in regulatory and supervisory control layers is presented. In Section 3, alternative control structures to handle larger throughputs are introduced and discussed. The dynamic performance of the alternative control structures is evaluated for large disturbances in Section 4 and the best control structure for (near) optimal operation over the entire throughput range is recommended.

2. Design of the control layers

In general, the control system can be divided in two main layers (see Fig. 2).

- *Regulatory layer*. Control of secondary (stabilizing) controlled variables (CV2). This layer usually involves the use of single loop PID controller.
- *Supervisory (economic) control layer*. Control of the primary (economic) controlled variables (CV1) using as manipulated variables (MVs) the setpoints to the regulatory layer or "unused" valves (from the original MVs). This layer is usually about a factor 10 or more slower than the regulatory layer. Since interactions are more important at longer time scales, multivariable control may be considered in this layer.

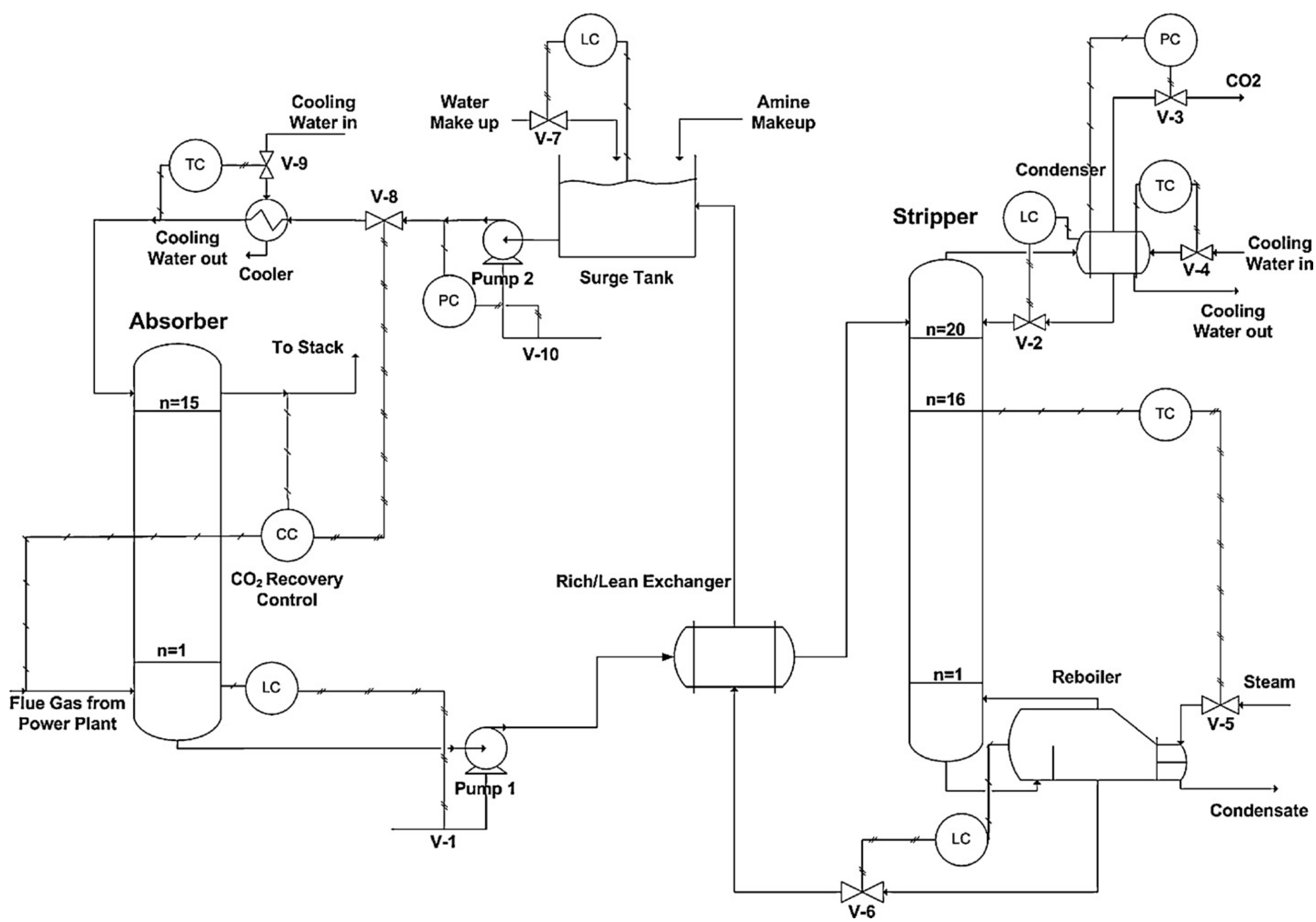

Fig. 1. Alternative 1, proposed decentralized control structure for region I with given flue gas flowrate [1].

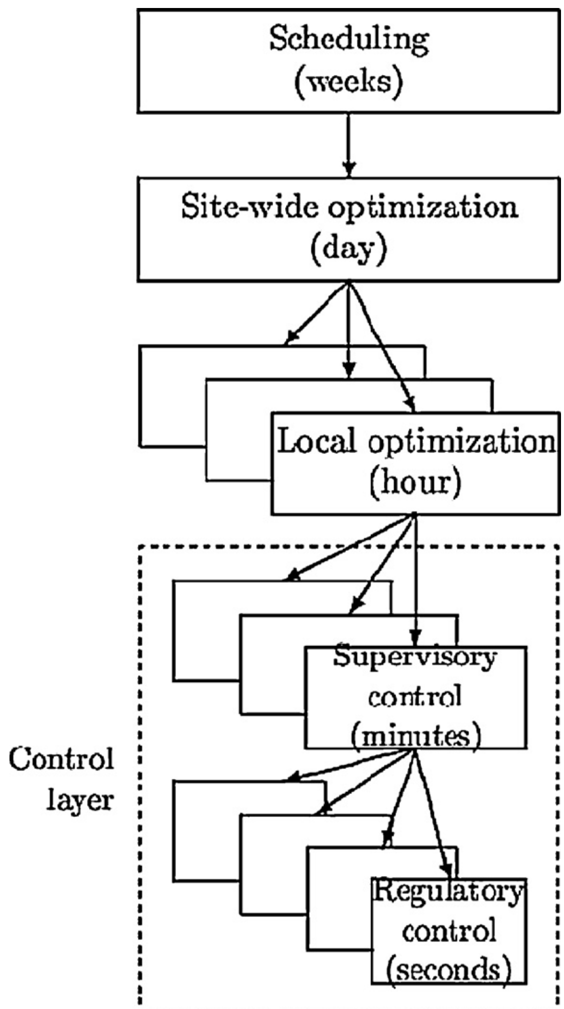

Fig. 2. Typical control hierarchy in a chemical plant [2].

Let us start with the control objectives for the supervisory control layer, which generally may change depending on the disturbances and active constraints. As a result of step 3 in the procedure (Table 1), in region I (low feed rates) the primary (economic) controlled variables were identified as [1]:

$CV1_1 = CO_2$ recovery in the absorber,
$CV1_2$ = Temperature at tray 16 in the stripper.

In region II with higher feedrates, the stripper reboiler duty is at maximum, so there is one less degree of freedom and the best CV was identified as [1]:

CV1 = Temperature at tray 13 in the stripper.

Furthermore, before starting the bottom-up part of the procedure, which starts with the choice of the regulatory layer (step 5), one generally needs to locate the throughput manipulator (step 4). However, the CO_2 capture plant is a part of the overall power plant, and the throughput manipulator is located further upstream in the power plant. This means that the feed (flue gas) to the CO_2 capture plant is given and will act as a disturbance, and the control system must be set up to handle this disturbance. However, the CO_2 capture contains a closed amine/water system and one must set the amine flow between the columns. The location of the amine recycle flow manipulator is an important decision.

The details of the regulatory layer design (step 5) are given in Table 2. In addition to the comments given in Table 2, one would generally like to combine control tasks in order to simplify the control system. For example, one may select to control the same stripper tray temperature in both regions I and II although there may be a small economic penalty. Furthermore, stabilizing the stripper temperature profile is necessary to maintain the CO_2 inventory circulating around the amine recycle loop. The stripper temperature control loop may therefore be moved to the regulatory layer.

Let us go back to the decisions involved in designing the control layer (Table 2).

(a) The first issue is to identify the variables that need to be controlled to "stabilize" the operation. Here "stabilization" means that the process does not "drift" too far away from the designed operational point when there are disturbances. For our process, we identify the following seven "stabilizing" CVs which need to be controlled (CV2):

1. Absorber bottom level,
2. Stripper (distillation column) temperature,
3. Stripper bottom level,
4. Stripper top level,
5. Stripper pressure,
6 and 7. Recycle surge tank: inventories of water and amine.

Note that there is no need to control the absorber temperature for the purpose of stabilization its profile. However, the temperature inside the absorber needs to be kept at a given value for good operation. We select to do this by controlling the absorber liquid feed temperature at 51 °C [1] in the regulatory layer.

The absorber pressure is not controlled ("floating") because a fixed value would require a valve which would give an undesired loss. It is set indirectly by the ambient pressure.

In addition, the inventories of water and amine in the recycle system must be controlled to make up for losses. However, these are small so even manual control may be possible.

(b) The next decision is to select the inputs "pairings" to control these variables (CV2).

Let us consider operation in region I (Fig. 1). We start with the inventory control system, that is, control of levels and pressure. The feed flow is given so the inventory control system needs to be in the direction of flow ([2] and [9]). However, for this particular process this does not really fix any loops, except for the pressure control of stripper using the CO_2 outflow (V-3). Next, we consider the closed recycle system of amine and water where we, as mentioned, need to decide on where to set the recycle flowrate. We choose to set it using the recycle liquid flow to the absorber column (V-8). The choice of V-8 as the flow manipulator in the recycle loop is an important decision (we will reconsider it later), because the "radiation rule" (Table 2) implies that inventory control in the recycle loop "downstream" of this location must be in the direction of flow. Thus, the bottom levels in the absorber and stripper must be controlled by their outflows (V-1 and V-6 respectively).

The pump V-10 in the recycle controls the pump outlet pressure but this is mainly for simulation purposes and the pump could also be set to run on constant power. However, when the pump controls pressure, the pressure measurement must be after the pump (in the opposite direction of flow). Finally, the stripper has a partial condenser, and the condenser level is controlled by the reflux flow (V-2).

Next, with the inventory control system fixed, we consider the stabilizing temperature loop for the stripper. We choose to use the reboiler duty (steam V-5) as the MV which is the only remaining option. To combine regulatory and supervisory control, the temperature sensor is located at tray 16, which was identified as a

Table 2
Details on step 5. Structure of regulatory (stabilizing) layer [2].

The purpose of the regulatory layer is to "stabilize" the plant, preferably using a simple control structure with single-loop PID controllers. "Stabilize" here means that the process does not "drift" away from acceptable operating conditions when there are disturbances. In addition, the regulatory layer should follow the setpoints given by the "supervisory layer".
Reassignments (logic) in the regulatory layer should be avoided: preferably, the regulatory layer should be independent of the economic control objectives (regions of steady-state active constraints), which may change depending on disturbances, prices and market conditions.

The main decisions are:
(a) Identify CV2s for the regulatory layer, these include "stabilizing" CVs, which are typically levels, pressures, reactor temperature and temperature profile in distillation column. In addition, active constraints (CV1) that require tight control (small back-off) may be assigned to the regulatory layer.
(b) Identify pairings (MVs to be used to control CV2), taking into account:
–Want "local consistency" for the inventory control [9]. This implies that the inventory control system is radiating around a given flow.
–Avoid selecting as MVs in the regulatory layer, variables that may optimally saturate (steady-state), because this would require either
◦ Reassignment of regulatory loop (complication penalty), or
◦ Back-off for the MV variable (economic penalty)
– Want tight control of important active constraints (to avoid back-off).
– The general pairing rule is to "pair close" to achieve a small effective delay from input (MV) to outputs (CV2).

self-optimizing variable in region I. Finally, the temperature of the absorber liquid feed is kept at 51 °C using the cooler duty (V-9).

Let us finally consider the supervisory layer which operates at a slower time scale. The stripper condenser temperature, which should be low to reduce the required work for CO_2 compression in the downstream process, should clearly be controlled using the cooling water flow (V-4). The remaining "economic" variable to be controlled is the CO_2 recovery and as a MV we may use the recycle amine flowrate to the absorber column (V-8), which was not used in the regulatory layer. This will be a relatively slow loop. The final control structure for region I is shown in Fig. 1.

3. Alternative control structures to handle larger throughputs

In region II, the stripper reboiler duty (steam) is at its maximum so the stabilizing temperature loop used for region I (Fig. 1) will no longer work. This is to be expected, because we have not followed the recommendation in Table 2 that says "Avoid selecting as MVs in the regulatory layer, variables that may optimally saturate". The problem is that there are no obvious alternative for controlling the column temperature. The reflux is often used for distillation columns, but this is a stripper where the reflux of the key component (CO_2) is very small and the reflux has no effect on column temperature. In addition, from the steady state optimization, it is optimal with maximum cooling to minimize the CO_2 compressor work in the down stream process [1]. Therefore, reflux must be used to control condenser level. Alternatively, the column federate could be used, but this is already used for controlling the absorber sump level.

3.1. Alternative 2 ("reverse pairing")

To find alternative solutions, let us start with region I. We consider the problem of controlling the two self-optimizing CVs.

y_1: CO_2 recovery
y_2: temperature on tray no.16 in the stripper

using the two available MVs

u_1: recycle amine flowrate (V-8)
u_2: reboiler duty (V-5).

The pairing in Fig. 1, where y_1 is controlled using u_1 and y_2 is controlled using u_2 is referred to as the "diagonal" pairing (Alternative 1). We will also consider the reverse "off-diagonal" pairing (Alternative 2; see Fig. 3).

Pairing issues

Alternative 1 may seem to be an "obvious" pairing choice but a more careful analysis shows that this is not so clear. A useful tool for selecting pairings is the relative gain array (RGA) and two main pairing rules are [10]:

RGA-rule 1. Prefer pairings such that the rearranged system, with the selected pairings along the diagonal, has an RGA matrix close to identity at frequencies around the closed-loop bandwidth. This may be quantified by selecting pairings with a small RGA number:

$$\text{RGA number} = \left\| \text{RGA}(G) - I \right\|_{\text{sum}} \tag{1}$$

RGA-rule 2. Avoid pairing on negative steady-state RGA elements.

The second rule states that pairing on a negative RGA provides a potential unstable response when individual loops are malfunctioning, as it is the case of input saturation. To compute the RGA we need a dynamic model. Using the dynamic UniSim simulator and "Profit Design Studio" (PDS) [11], we identified the following linear model:

$$G_{dyn.}(s) = \begin{bmatrix} \dfrac{6.85s + 1.74}{19.7s^2 + 11.4s + 1} & \dfrac{-0.76s + 0.038}{2400s^2 + 107s + 1} \\ \dfrac{(-9.51s - 1.02)e^{-2s}}{218s^2 + 17.3s + 1} & \dfrac{0.45s + 0.0754}{205s^2 + 18.8s + 1} \end{bmatrix} \tag{2}$$

The steady-state RGA computed from this model is

$$\text{RGA}_{dyn.}(s = 0) = \begin{bmatrix} 0.77 & 0.23 \\ 0.23 & 0.77 \end{bmatrix} \tag{3}$$

Since all elements are positive one cannot eliminate any of the pairings using RGA-rule 2. The RGA-number as a function of frequencies is plotted for the two alternative pairings in Fig. 4. As expected, we find that the diagonal pairing is the best with the RGA-number close to 0 at all frequencies.

To check the model obtained from the dynamic simulator, we compute the steady-state gains using the steady-state UniSim model and this gave a very surprising result. By making 5% perturbations in the inputs, the following steady-state model was identified:

$$G_{SS} = 10^{-2} \times \begin{bmatrix} -0.5232 & 1.48 \\ -8.47 & 5.17 \end{bmatrix} \tag{4}$$

with the corresponding RGA matrix:

$$\text{RGA}_{SS} = \begin{bmatrix} -0.27 & +1.27 \\ +1.27 & -0.27 \end{bmatrix} \tag{5}$$

Note that steady-state RGA for the diagonal pairing is negative. The reason is that the sign of the 1,1-element of the gain matrix

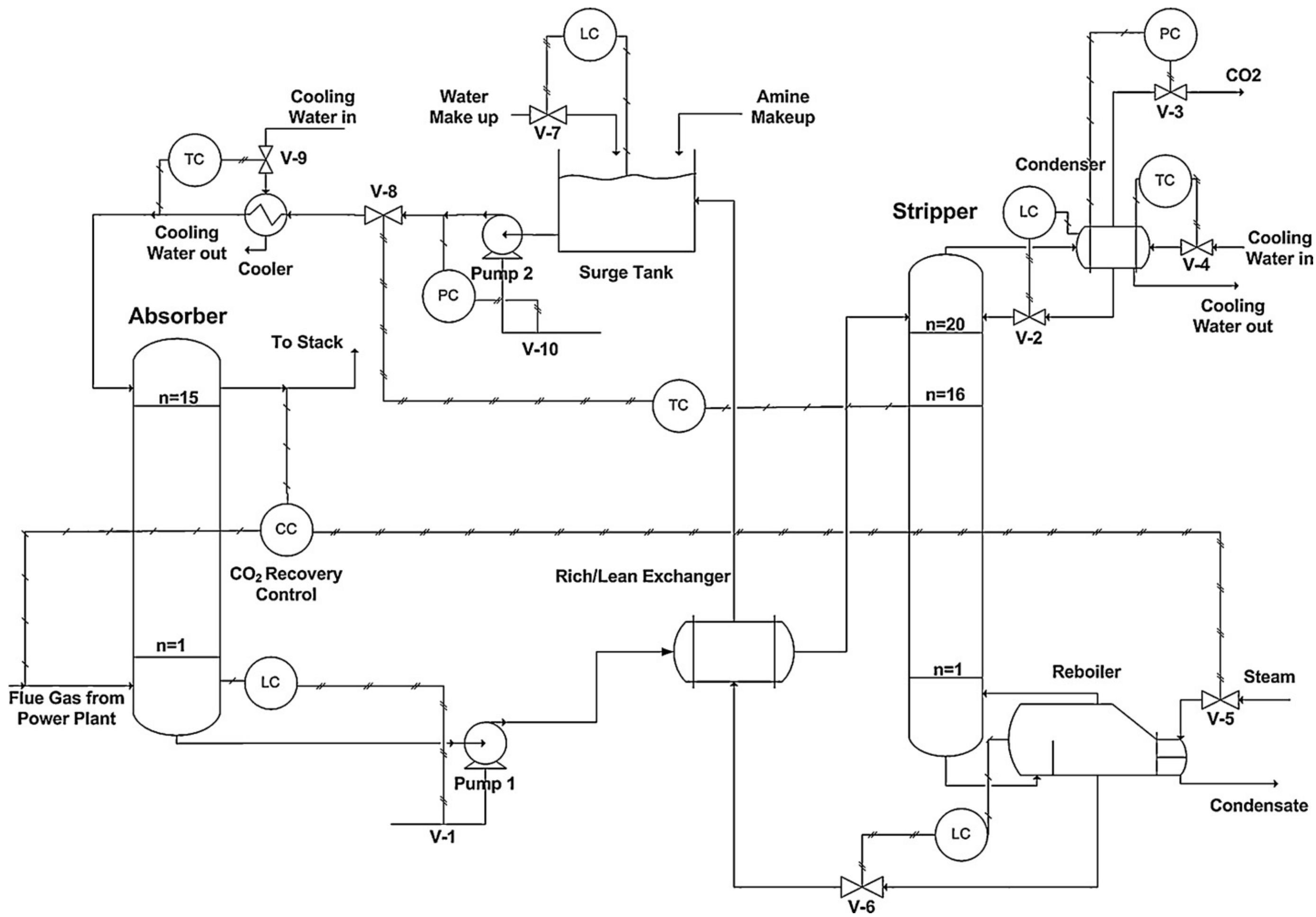

Fig. 3. Alternative 2, reverse pairing for region I, and also close to the optimal structure for region II.

(4) is negative, whereas we found from the dynamic model in (2) that the diagonal elements are dominant at higher frequencies correspond to having a positive gain at steady state. The sign change can be explained by Fig. 5, where we see that y_1 (CO_2 recovery) initially increases in response to a step increase in u_1 (recycle amine flowrate). This agrees with the dynamic model $G_{dyn.}$. The reason for the initial increase in the CO_2 recovery is an initial decrease in the CO_2 vapor mole fraction in the top of the absorber, which is expected. However, on a longer time scale, with an increased federate to the stripper column (u_1) and a constant reboiler duty (u_2), the CO_2 concentration in the bottom of the stripper increases and starts "filling up" the absorption column with CO_2. This takes a long time because of the large holdup in the absorber (see Fig. 6). Finally, the CO_2 "breaks through" at the top of the absorber column and after almost 170 min the CO_2 recovery (y_1) starts decreasing and keeps decreasing until about 250 min. There is a second smaller decrease in y_1 about 70 min later. To actually get a negative gain for y_1, we need the change in recycle amine flowrate (u_1) to be sufficiently large, so this is partly a nonlinear effect. The dynamic model $G_{dyn.}$ is based on simulations on an intermediate time scale and this is why the reversal of the sign of the gain was not identified.

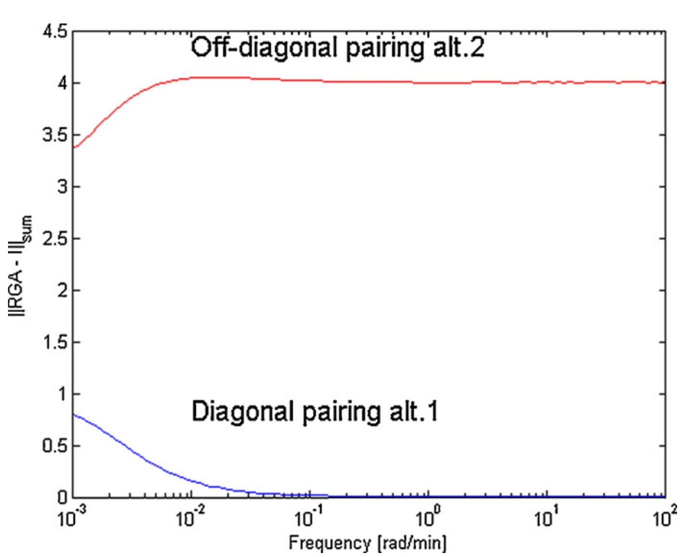

Fig. 4. RGA number of two different alternatives in pairing.

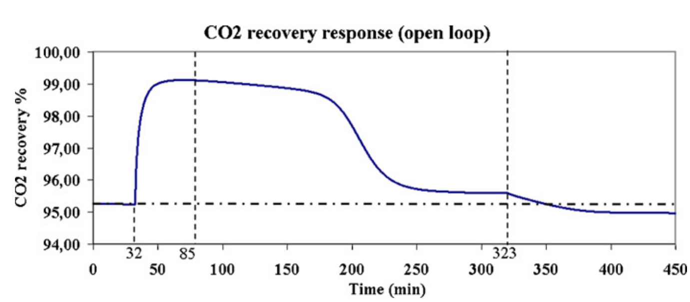

Fig. 5. Response in CO_2 recovery (y_1) to step change in recycle amine flowrate (u_1 by 5%).

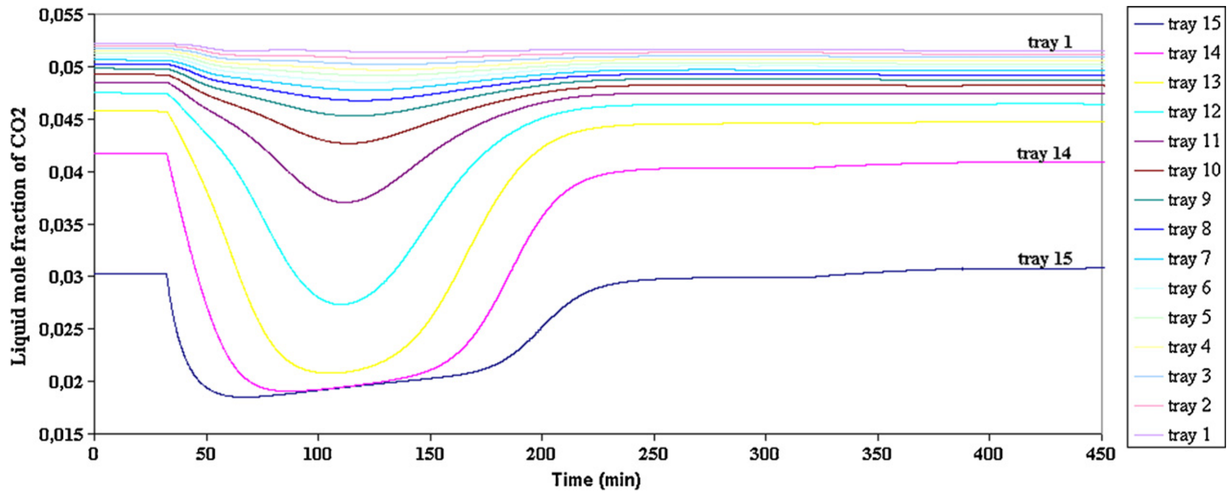

Fig. 6. Sluggish dynamic response of the CO_2 mole fraction due to large holdup on absorber trays.

The steady-state RGA suggests "reverse pairing" which is referred to as Alternative 2.

Based on pairing rule 2 and the model G_{ss}, we would conclude that the diagonal pairing (Alternative 1) should not be used. However, recall that the reason for RGA rule 2 is to avoid instability if one of the loops is no longer working, which in this case will occur when the reboiler duty saturates. In summary, the diagonal pairing ("Alternative 1") should be used in region I because the CVs and the MVs are close, but we need to reconfigure the loops if the reboiler duty reaches its maximum (region II). Alternative 2, on the other hand would work in both regions without any reconfiguration. However compared to Alternative 1 the dynamic performance in region I will be poorer as the associated MVs are not close-by. This is confirmed by dynamic simulations later.

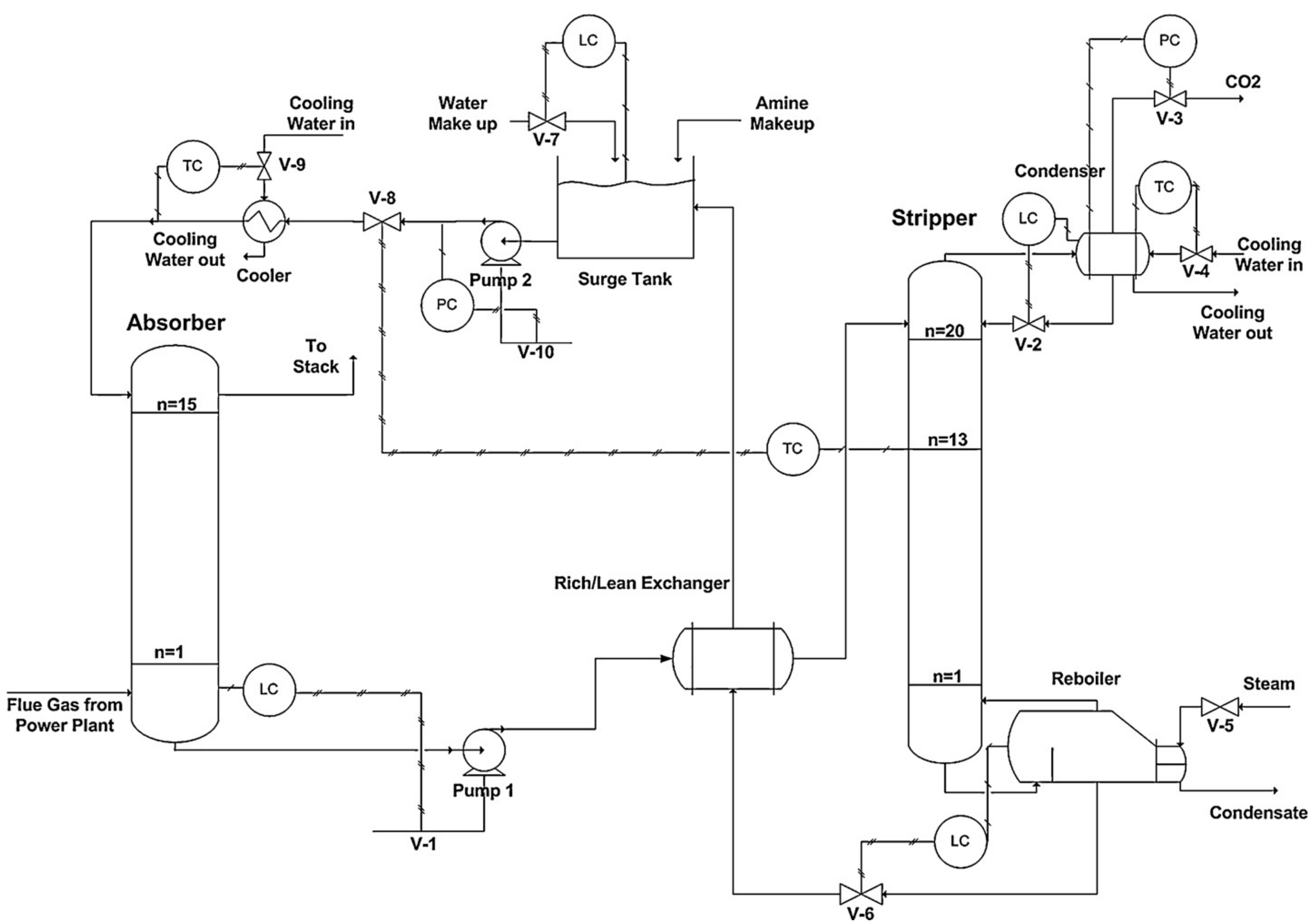

Fig. 7. Alternative 3, proposed decentralized control structure for region II [1].

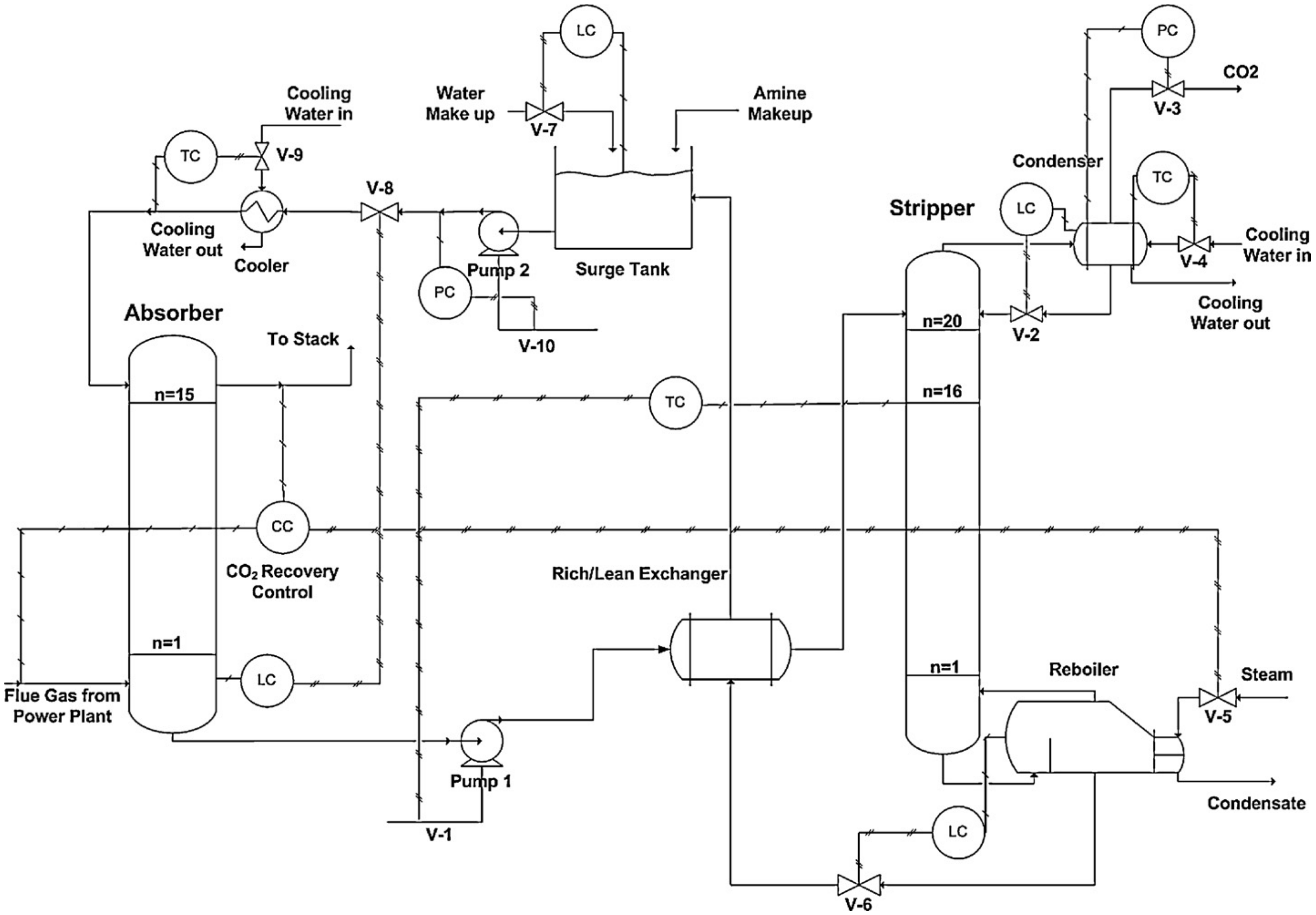

Fig. 8. Alternative 4, proposed structure for combined regions I and II (modified of Alternative 2).

3.2. Operation in region II (Alternative 3)

Let us next consider operation in region II. The "optimal" solution for region II, is to let the reboiler duty stay at its maximum, "give up" controlling the CO_2 recovery, and move the stripper temperature from tray 16 to tray 13 [1] (Alternative 3, Fig. 7). Similar to Alternative 2, the stripper temperature is controlled using the recycle flow which is the only available "free" MV in region II.

For strictly optimal operation in both regions I and II, one could use same supervisory logic that switches between Alternatives 1 and 3. It would switch from Alternative 1 to Alternative 3 when the reboiler duty saturates and it would switch back from Alternative 3 to Alternative 1 when CO_2 recovery passes 95.26%.

However, note that Alternative 3 (region II) is actually very close to the "reverse pairing" (Alternative 2) for region I. This suggests that Alternative 2 could also be used to handle region II, although it would involve a small loss because the stripper temperature is not at its optimal location in region II. With Alternative 2, switching between regions I and II would simply be to give up control of CO_2 recovery when the reboiler duty saturates.

3.3. Alternative 4 (regions I and II)

So far Alternative 2 is the best structure for the combined regions I and II, and we want to improve on it. In Alternative 2, the recycle amine flowrate (V-8) is manipulated to control the stripper temperature. However, this control loop has a large effective delay so performance is relatively poor (see dynamic simulations later). In Alternative 2, the liquid inflow to the stripper, which actually has a much more direct effect on the stripper temperature, is manipulated to control the level of the absorber. A modification (Alternative 4) is to change the pairings for these two loops, which can be viewed as moving the location of the given flow in the recycle loop from the inlet to the outlet of the absorber (Fig. 8).

4. Performance of alternative control structures

In this section we analyse the four alternatives, as well as multivariable control (MPC), using dynamic simulation. We observed some discrepancy in the nominal steady-state in dynamic mode versus the steady-state mode. For example, for the same nominal steady state CV values, the reboiler duty in the dynamic mode is 1074 kW while in the steady state mode the reboiler duty is 1161 kW. This is partly due to the dynamic simulation being pressure driven so that the column pressure profiles are slightly different. In addition, the UniSim dynamics solver may be using approximate thermodynamic models. Fortunately, from a control structure synthesis perspective, the difference in the nominal steady states is not very important as the first constraint to become active (reboiler duty) as throughput increases, remains unchanged in both modes.

As in part I, there are maximum capacities compared to nominal values for the reboiler duty (+20%), cooler (+50%) and pumps (+40%). These constraints are even more important in the present dynamic study.

Table 3
Tuning parameters (Alternative1).

Control loop	Closed loop time constant, τ_c (min)	K_c	τ_I (min)
CO_2 recovery (y_1) with recycle amine flowrate (u_1)	0.38	0.315	3.04
Temp. 16 in the stripper (y_2) with reboiler duty (u_2)	0.30	10.53	2.4

4.1. Alternative 1 (region I)

We first consider the diagonal pairing for region I (Fig. 1) and look at the performance (dynamic behavior) when there are large disturbances. All the controllers were tuned using SIMC method [12] with tuning parameters in Table 3. The flowrate of flue gas is increased gradually in steps of 5% from 0% to +25% compared to the nominal flowrate (Fig. 9a). From Fig. 9 we see that the control structure behaves very well in region I (up to +20%) with tight control of the CO_2 recovery and stripper temperature. When the flowrate of flue gas increases +20% (at $t = 269$ min), the reboiler duty saturates at $t = 273$ min (Fig. 9b), signifying a transition to region II. Initially, it seems that the system is stable although the temperature of tray no. 16 (Fig. 9c) in the stripper is not controlled in its setpoint, but with a further increase in the flue gas flowrate (at $t = 358$ min), the stripper temperature drops further ($t = 361$ min) resulting in insufficient CO_2 stripping which builds up in the amine recirculation loop. This necessitates the more amine flow in the absorber to meet the desired CO_2 recovery which saturates the pump 1 (Fig. 9d) at $t = 437$ min. At $t = 600$ min the recycle valve V-8 (u_1) becomes fully open (Fig. 9e) and we get an unstable system where the CO_2 recovery can no longer be maintained at 95.26% (Fig. 9f) resulting in large objective function value (Fig. 9g). This is as expected from the earlier RGA analysis. In summary, pairing Alternative 1 can only handle increased flue gas flowrates of up to +20% (region I), at which point the reboiler duty (u_2) saturates.

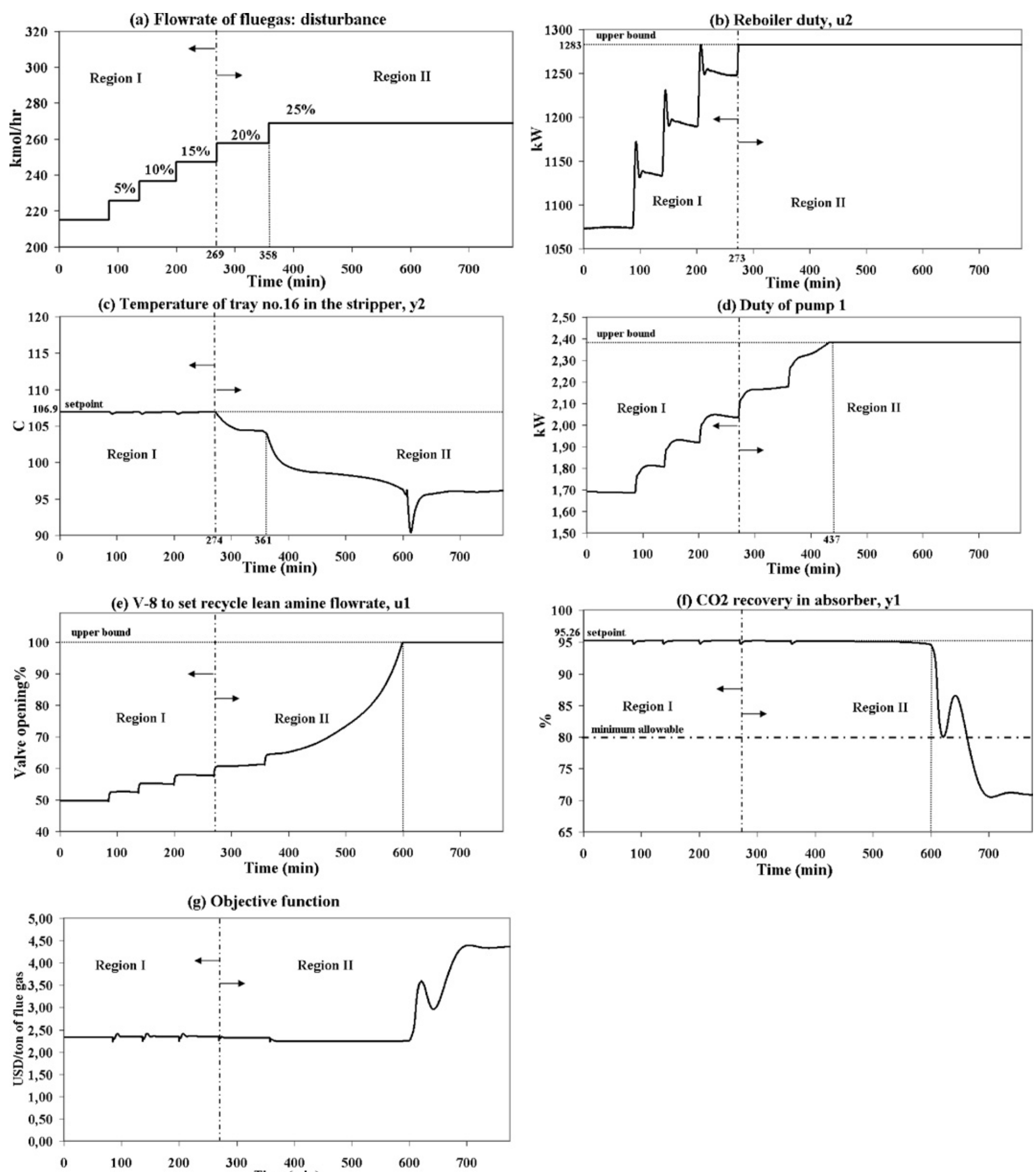

Fig. 9. Dynamic simulation of pairing Alternative 1 (Fig. 1). The structure handles increase in flue gas flowrates of up to +20% (region I).

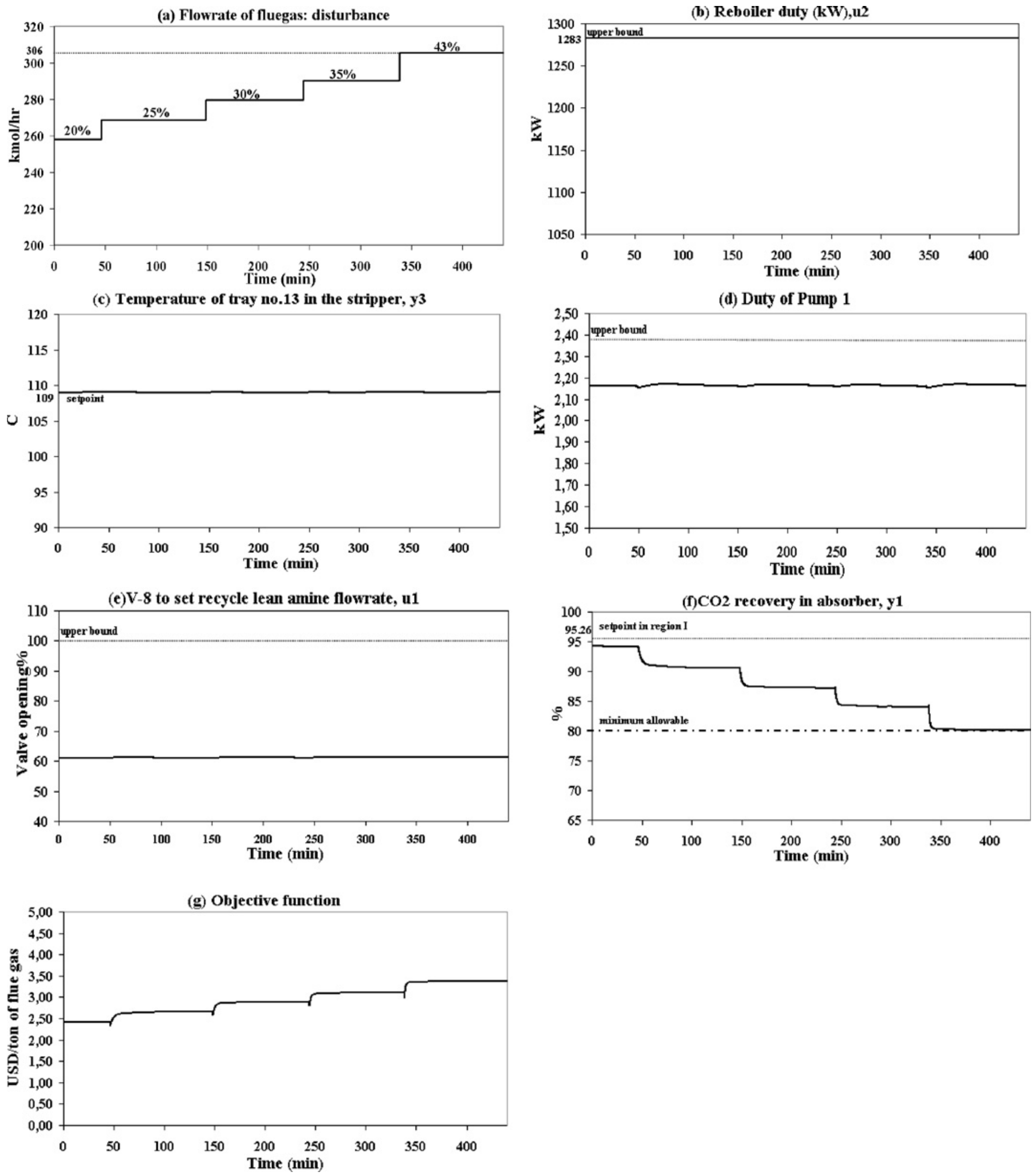

Fig. 10. Region II: Alternative 3 (Fig. 7) handles increase in flue gas flowrate to +43%, but it does not cover region I.

4.2. Alternative 3 (region II)

We here consider the "optimal" structure for region II. ("Alternative 3", Fig. 7). In the simulation, for a closed loop $\tau_c = 2.33$ min, SIMC method gives $K_c = 0.314$, $\tau_I = 14$ min as PI tuning parameters. Fig. 10 shows how this structure handles feed flowrate changes with operation within region II, starting from 20% above the nominal (where the process enters region II) with steps of +5%. In this case, we do not attempt to control CO_2 recovery and it drops gradually from 95.26% to 80%, which is the minimum allowable CO_2 recovery. This happens when the feed flowrate increase is +43% (Fig. 10a). As shown in Fig. 10c, the stripper temperature is well controlled and hardly affected by the feed flow disturbance.

In general, any active constraint should be considered as a disturbance [2], thus we changed the fixed (saturated) reboiler duty by ±10% and the dynamic responses in Fig. 11 shows that also this disturbance is handled well with stripper temperature deviation of less than 3 °C.

4.3. Alternative 2 (regions I and II)

The simulations in Fig. 12 show that this alternative can handle increase of the flue gas starting from the nominal value (0%) and up to +42% (Fig. 12a). The reboiler duty (u_2) saturates (Fig. 12b) when the flue gas flowrate increase is +20%, at which point the CO_2 recovery control is given up, and at +42% the minimum CO_2 recovery of 80% (Fig. 12f) is met. Note that we in both regions control y_2 (temperature of tray no. 16 in the stripper). This is the best choice in region I, and in region II it is close to the best self-optimizing variable which is y_3 (temperature of tray no. 13 in the stripper). Thus, it is not surprising that if Alternative 2 can handle changes of +42% which is close to the region II optimal at +43% (using Alternative 3).

The advantage of Alternative 2 compared to optimal structure (which is to use Alternative 1 in region I and Alternative 3 in region II) is that we do not need switching of CVs. The main disadvantage is that we have more interactions than Alternative 1 in region I. This is seen by comparing the simulations in Fig. 9

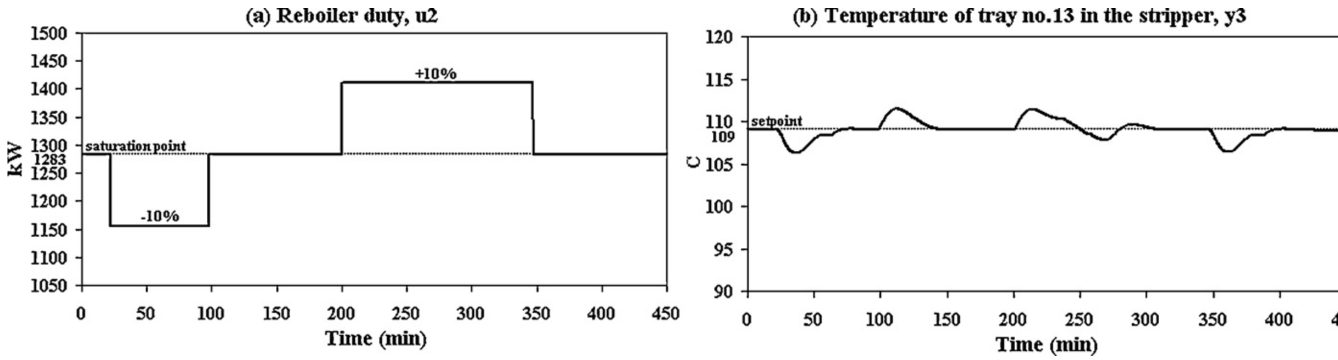

Fig. 11. Region II: the structure in Fig. 7 also handles disturbance in reboiler duty which is the active constraint.

(Alternative 1) and 12 (Alternative 2) in region I. The CO_2 recovery is not as tightly controlled (compare the max. deviation of 0.4% in Fig. 9f with 3% in Fig. 12f) and also the stripper temperature shows larger variations (compare max. deviation of 0.3 °C in Fig. 9c with 1 °C in Fig. 12c). In region II, using Alternative 2, there is a minor loss compared to Alternative 3 (compare Figs. 10g and 12g) and also here the temperature control is comparable (compare Figs. 10c and 12c).

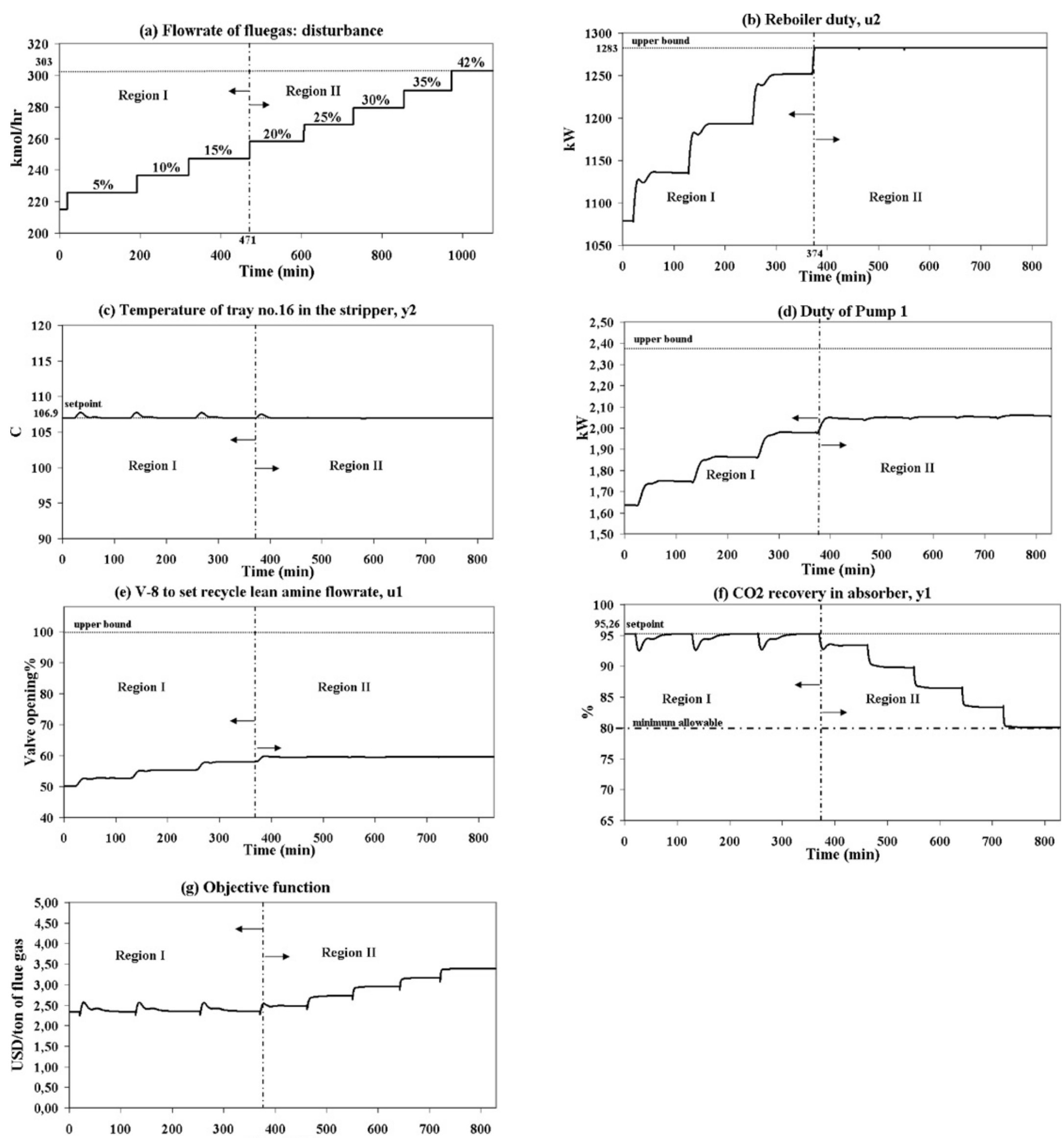

Fig. 12. Dynamic simulation of alternative 2 (reverse pairings) in combined regions I and II. The structure handles the increase in flowrates of flue gas to +42%.

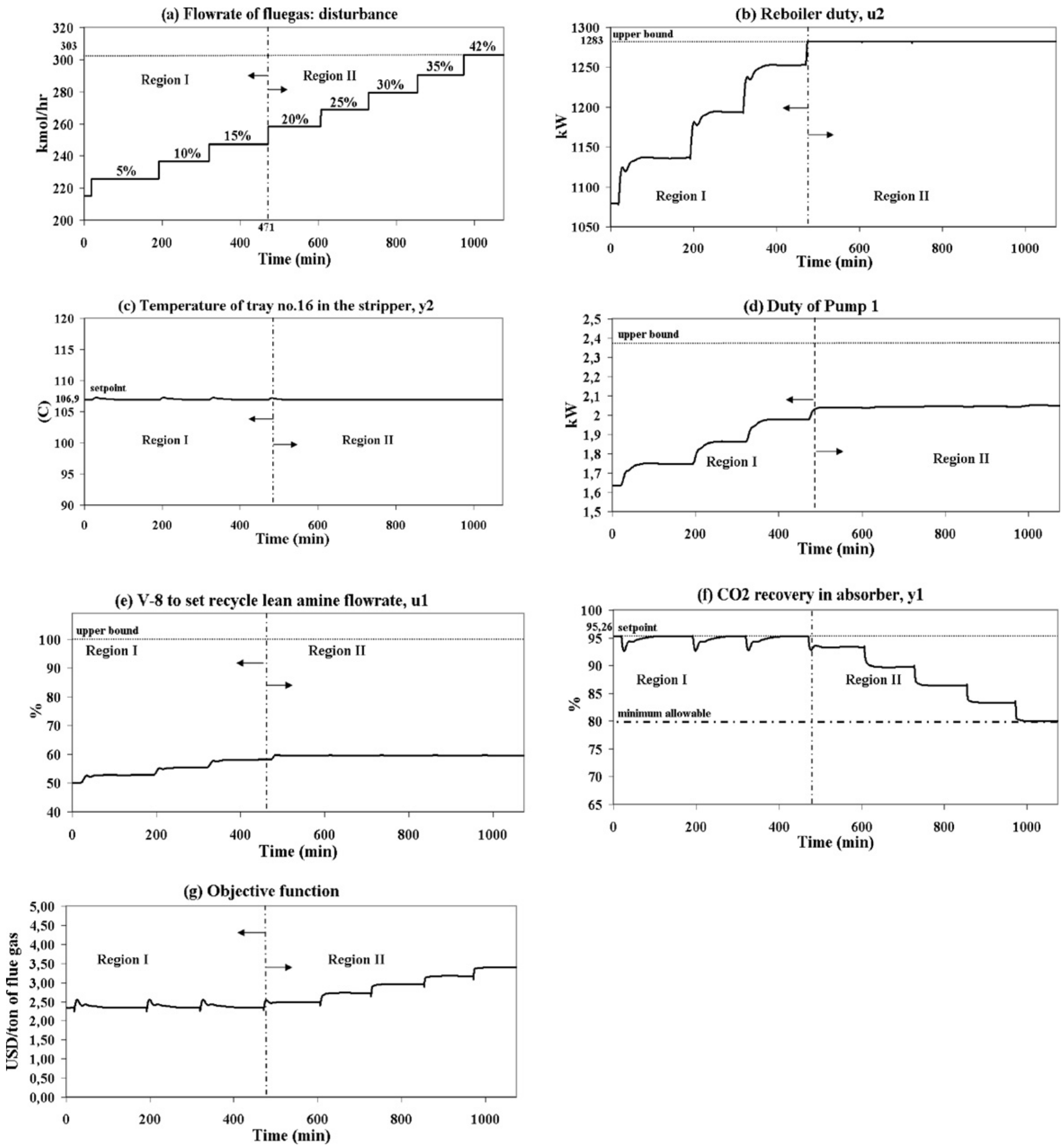

Fig. 13. Dynamic simulation of pairings alternative 4 (modified reverse pairings) in combined regions I and II.

4.4. Alternative 4 (regions I and II)

Alternative 4 (Fig. 8) involves repairing the absorber level control. As shown in Fig. 13, this structure handles increase in flue gas flowrate up to +42% like in Alternative 2 but with tighter control of stripper temperature. By comparing Figs. 12c and 13c, we see that the stripper temperature deviation decreases from 1 °C to 0.2 °C while control of CO_2 recovery remains the same. Therefore, Alternative 4 may be the best structure for practical implementation.

4.5. Multivariable controller (regions I and II)

Finally, we consider a multivariable controller (MPC) obtained using the RMPCT (Robust model predictive controller) from Honeywell [3]. The aim is to compare its performance with the previous decentralized PI controllers. This MPC includes 2 CVs (y_1 and y_2), 2 MVs (recycle lean amine flowrate, u_1 and reboiler duty, u_2) and 2 disturbances (flowrate of flue gas and its CO_2 composition). We used Honeywell's Profit Design Studio to identify the dynamic models and the final responses with MPC are shown in Fig. 14. For feed values greater than +20%, when the reboiler duty saturates (region II), instead of controlling the CO_2 recovery at its setpoint value (95.26%), we put less emphasis on controlling the CO_2 recovery by introducing a range with a lower bound of 80% for this CV. The result is that RMPCT controls the temperature of tray no. 16 at its setpoint and gives up controlling the CO_2 recovery. When the flue gas flowrate reaches +42% the minimum allowable CO_2 recovery of 80% is met and there are no degrees of freedom left and further increase of flue gas flowrate is infeasible.

We finally compare the economical performance of all the structures. From the objective function in Figs. 10g (Alternative 3), 12g (Alternative 2), 13g (Alternative 4) and 14g (MPC) we find that there is only a small difference. At the final steady state, the objective function values are 3.38, 3.39, 3.39 and 3.39 (USD/ton of flue gas) respectively. This implies that we can use both Alternative 2, Alternative 4 or MPC to control the system even in the presence of large flowrates of flue gas. Note that the design and implementation of reverse pairings with decentralized controllers are simpler and

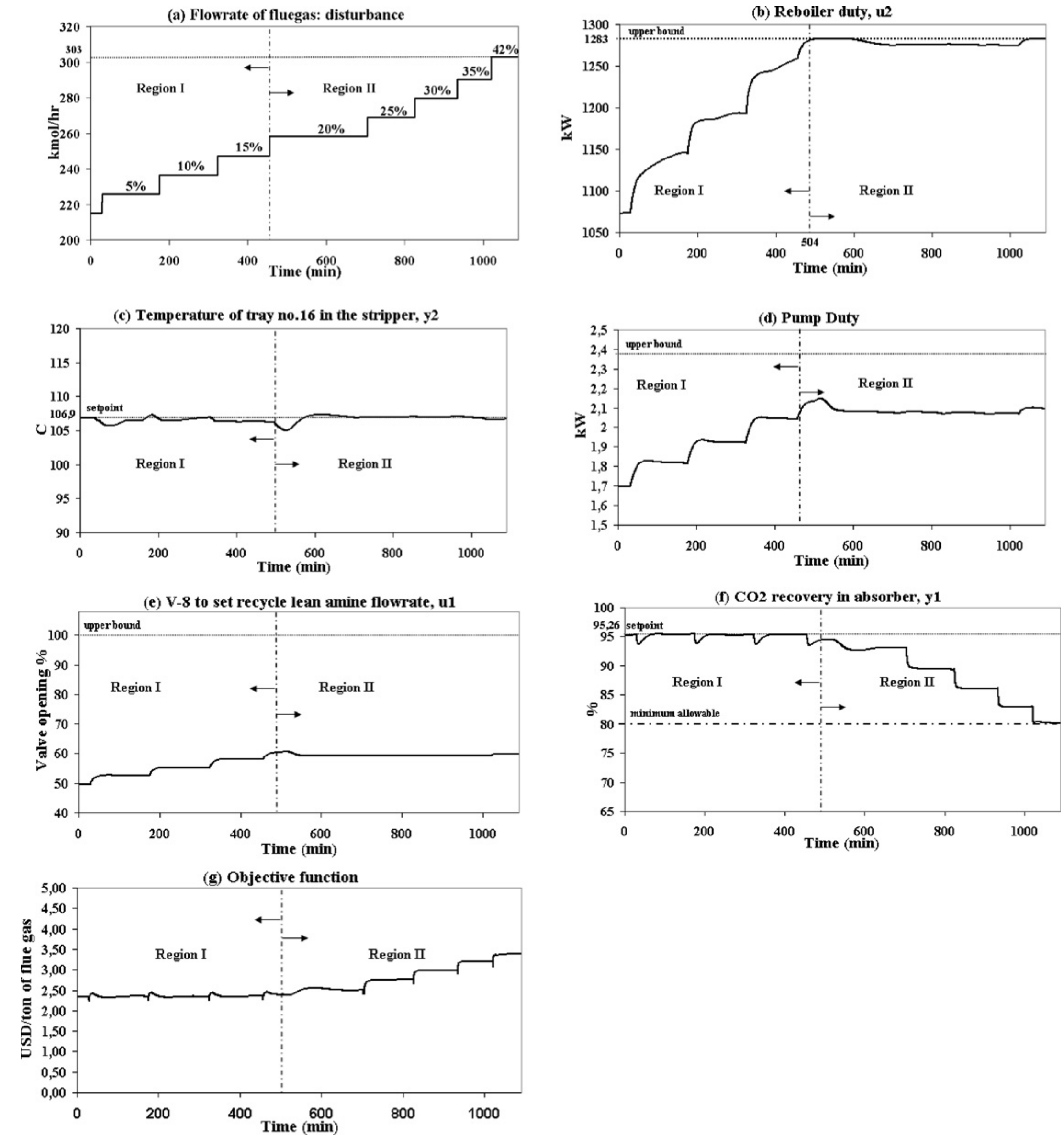

Fig. 14. Performance of RMPCT tuned in region I, when extended to region II, the structure handles increase in flue gas flowrates of +42%.

cheaper than MPC which needs model identification etc. although responses time are comparable.

Of all the alternatives considered, Alternative 4 is proposed as the best structure for the CO_2 capturing process, studied here.

5. Conclusions

Using the bottom-up part of the plantwide control procedure, we obtained alternative control structures that implement the optimal controlled variables (CO_2 recovery in the absorber and the temperature of tray 16 in the stripper) from part I. Alternative 1 (diagonal pairing of the best CVs found in region I with close-by MVs: recycle amine flowrate and reboiler duty) can handle flue gas flowrates of up to +20% (region I). The reverse pairings (Alternative 2) or MPC handle flow values up to +42%. Due to the large delay between the paired MVs and CVs in Alternative 2, a modified structure (Alternative 4) is proposed where the recycle amine flow manipulator is moved from the inflow to the outflow of the absorber. This simple structure (Alternative 4) has dynamic performance comparable to MPC, and is proposed as the best alternative because of its much simpler implementation.

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