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Steady-state energy optimization and transition assessment in a process of CO_2 absorption from natural gas

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ABSTRACT

An industrial-scale MDEA-based CO_2 absorption process is considered in this work. Natural gas plant of Aguaragüe Argentina is selected to perform the studies of optimization and dynamics. With the steady-state of the process simulated, the reboiler energy demand of the plant is optimized in Aspen Hysys v8.8. In this optimization study, we analyze the base case (current) conditions of the reference plant and also six disturbance scenarios. According to the results, the main energy requirement can be reduced in more than a 48%, with the current process design.

In the second part of our work, the plant is simulated in dynamic mode with the aim of characterizing the transition towards the found optimal conditions. It is proved that the dynamic model can reproduce the specified values after a short-time. In addition, the absence of undesirable states during the stabilization proves that the current control scheme is appropriate to minimize the energy consume in the plant under study.

1. Introduction

Different studies have been performed to optimize the acid gas absorption process, to evaluate the dynamic performance, and to model a control scheme [1]. For instance, Panahi et al. [2] design and optimize a post-combustion plant for CO_2 capture using monoethanolamine (MEA). Lastari et al. [3] consider the energy demand as objective function and the use of the simulator Aspen Hysys to optimize the Ethane-CO_2 extractive distillation process. Also Abbas et al. [4] design and optimize the acid gas removal, particularly the absorber design in steady and dynamic modes. The developed model demonstrate that the absorption of CO_2 remains constant throughout the process, thus they generate proper process design and control loops. Furthermore, Abdulrahman and Sebastine [5], Øi et al. [6], Ghanbarabadi and Gohari [7], Al-Lagtah et al. [8], and Jassim [9] develop the steady-state of the acid gas absorption by using process simulators. Taking the capability of simulators, Mechleri et al. [10] present a dynamic simulation and control strategies applied to the post-combustion process for CO_2 capture by using MEA. In their proposed control diagram, several controllers are comprised to maintain the specified level of greenhouse gases emissions.

Zahid et al. [11] analyze the effect of different parameters on the reboiler duty and the purity of the product, in a process to capture CO_2 with diethanolamine (DEA). In parallel, Pouladi et al. [12] reveal the optimal conditions for the inlet acid gas in respect to the conditions of the solvent, in a comparable process.

Kvamsdal et al. [13] present a dynamic model of an absorption column that is capable of representing the behavior during the course of the dynamic operation, with varying loads. Prölss et al. [14] develop a dynamic model of a MEA absorption process to minimize the energy demand of the system. Starting from a rather detailed model of the process, the reduction is performed based on physical insight. Studying the lean loading in the stripper as a control variable, simulation results indicate local minima in the reboiler energy consumption. Nittaya et al. [15] present and analyze the dynamic model of an industrial-scale CO_2 capture with MEA, considering multiple scenarios. The proposed control scheme shows smooth responses with small oscillations during the time, with controlled variables at the desired set points.

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Similar to Sahraei and Ricardez-Sandoval [16] and Nittaya et al. [15], Gaspar et al. [17] evaluate the control response of the CO_2 capture process for sudden and sustained changes in the plant's load. However, they also investigate the behavior of the carbon capture plant for large load decreases, manipulating the flow rate of the lean amine, and the steam to the reboiler.

With reference to previous works, extended bibliography exists regarding the topic of the optimization of a CO_2 absorption plant. Conversely, less extended works asses the application of the results in an existing plant and characterize the transition to the optimal conditions. To the author's knowledge, the decrease of the lean amine flow, the decrease of the reboiler duty, and the time for the system to reproduce the optimal conditions have been barely discussed. This study performs simulation, optimization, and dynamic analysis of a commercial gas sweetening plant and serve as a very good example of the use of advanced computational tools for enhancing the efficiency of a process plant.

In a preceding work [18], the steady-state of the CO_2 absorption plant from Aguaragüe Argentina was simulated in Aspen Hysys v8.8 and the model validated. In the present work, the ratio between the reboiler duty and the absorbed CO_2 is optimized, due to the high value observed currently in the existing plant. The nonlinear problem (NLP) is solved with the simulator Aspen Hysys by using a combination of the BOX algorithm and the Sequential Quadratic Programming (SQP). From the results, it can be seen that not only the specific heat of the reboiler is reduced, but also the solvent to be cycled, from 53.18 to 11.43 m^3/h. Additionally, the decrease of the energy requirement is proved to be reduced when disturbances in the inlet gas conditions are considered.

After the steady-state cases (Base and Optimal) are simulated, the dynamic mode is formulated to evaluate the transition from the nominal conditions to the optimal state. In this stage, rigorous considerations, modifications to the flowsheets, and equipment and valves sizing are implemented. Reference plant piping and instrument diagrams (P&ID) are carefully evaluated to reproduce each controller's location. The capacity to reproduce the optimal values, from the nominal conditions, proves the validation of the model and the proper selection of the controllers' parameters.

2. Process description

In natural gas industrial plants, the specific process to remove the acid gases from the sour natural gas is known as sweetening process. Particularly, an amine-based sweetening process can be applied to remove the CO_2 from the natural gas and adjust its composition for sale. An amine-based CO_2 absorption process consist of two main sectors [19]. In the first, the acid gases react with an amine aqueous solution removing the acid gases from the natural gas, which migrate to the liquid phase. The reaction is of electrolytic basis and exothermic type, favored by large pressures [20,21].

In the second stage, the rich amine solution is flashed, heated, and sent to a regenerator tower, where the acid gases are distillated [22]. In this unit, the stripping vapor comes from the reboiler where the combustion of hydrocarbon fuels takes place. Afterwards, the regenerated amine is sent back to a heat exchanger, a cooling unit, a pumping system, and recycled to the absorber. Amine and water makeup tanks serve to compensate the liquid leaks throughout the

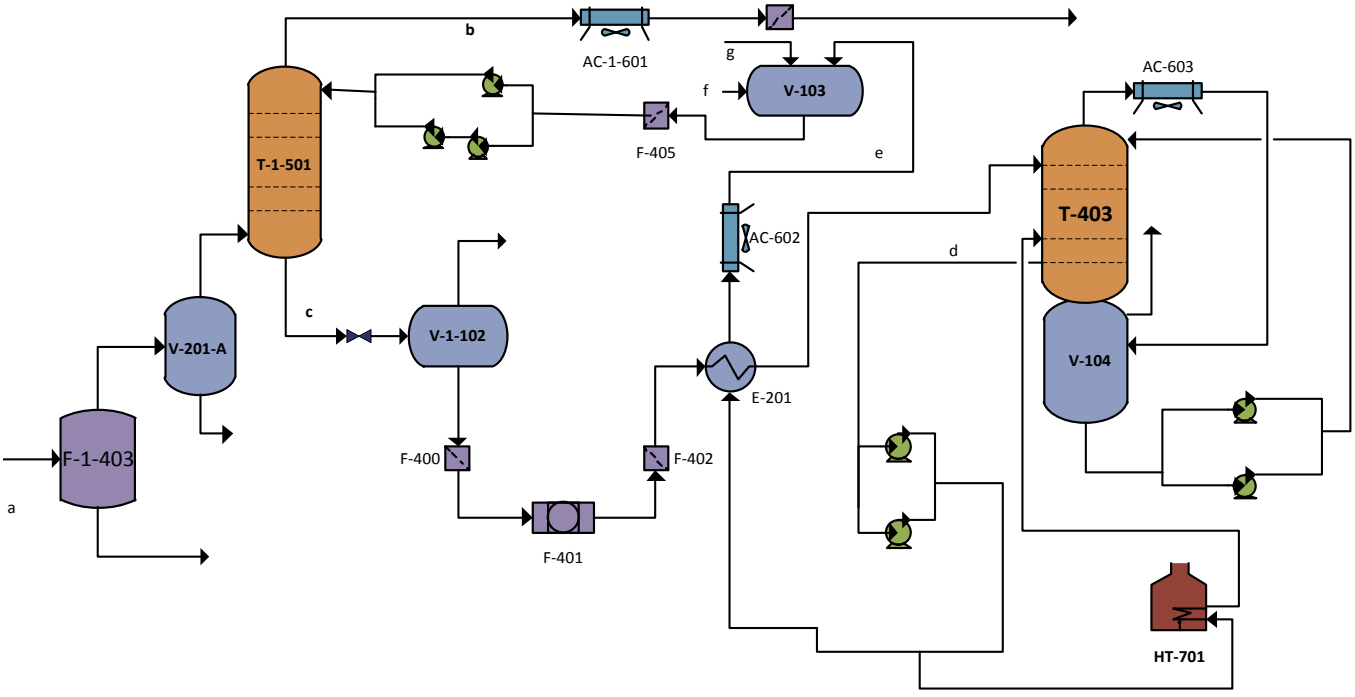

Fig. 1. Design of the reference CO_2 absorption plant.

Where:
a: Sour natural gas
b: Sweet natural gas
c: Rich amine
d: Lean amine
e: Cooled lean amine
f: Makeup water
g: Makeup amine

V-201-A: Phase separator
T-1-501: Absorption tower
V-1-102: Flash separator
E-201: Rich-Lean amine heat exchanger
T-403: Regenerator or stripping tower
V-104: Reflux accumulator
HT-701: Reboiler
V-103: Makeup

continuous operation (Fig. 1) [23–25].

As it can be anticipated, this process is integrated in terms of material and energy balances. However, the energy requirement for the regeneration of the solvent is considerably high and thus optimization strategies can be applied to minimize it [11,26].

3. Methodology

In this section, the reference plant is introduced, the optimization problem is described, and the development of the control scheme is exposed.

3.1. Reference plant

For this study, authors use data coming from the CO_2-absorption plant of Aguaragüe, Argentina [18]. Material streams, flowrates, compositions, and conditions were defined according to the measures of the plant under study. The parameters of the simulation were determined by measurements made in the plant. The output of simulator is obtained by means of the employment of the Acid Gas fluid package, provided by Aspen Technology Inc [27]. This property package contains the eNRTL model parameters and other transport property model parameters identified from regression of extensive thermodynamic and physical property data for aqueous amine solutions. The thermodynamic fluid package includes the chemical reactions system without the necessity of specifying them into the simulation software [28–30].

Once defined involved components and thermodynamic package, the next step is to add the operating units. The logical operator 'recycle' is introduced in order to properly run the simulation. The absorber tower includes 24 internal trays; tray 1 is at the top of the tower and tray 20 at the bottom. The gas to treat is fed in the tray 20 while the amine aqueous in tray 1. The stripper tower has 20 trays, a condenser and a reboiler. The amine solution to regenerate enters the column in tray number 8.

In Table 1, we reproduce the conditions of the base case (B.C.) model. In this case, the sweetening process must reduce the CO_2 molar content of the gas to the value fixed by regulation, 2 mol% in sale conditions [19].

For the validation of the output, authors calculated the relative error percentage between model prediction and measured values (Table 2).

By comparison with the results of Behroozsarand and Zamaniyan [23], the model properly reproduce the process and it is suitable to perform the study under the scope of this article.

Table 1
Input data for Aspen Hysys v8.8 Base Case simulation [18]. For the validation of the output, authors calculated the relative error percentage between model prediction and measured values (Table 2).

Parameter	Value
Feed Gas Temperature	35 °C
Feed Gas Pressure	6178 kPa
Feed Gas Flow	250 Mm^3/d
CO_2 in Feed Gas	4 mol%
CH_4 in Feed Gas	93 mol%
Lean Amine Temperature	40 °C
Lean Amine Pressure	9610 kPa
Lean Amine Rate	2100 kmol/h
MDEA in Lean Amine	38 wt%
Flash Pressure	441 kPa
Stripper Feed Temperature	90 °C
Top of the Stripper Temperature	87.78 °C

Table 2
Operating data, simulation obtained data and relative error percentage [18].

Parameters	Operating Data	Simulation Data	R. E. (%) (*)
Sweet gas Temperature (°C)	40	42	4.95
Sweet gas Pressure (kPa)	6000	5884	1.93
Sweet gas Mole Flow (m^3/d)	240,000	239,308	0.29
Rich amine Pressure (kPa)	6000	6080	1.32
Rich amine Temperature (°C)	110	106.2	3.62
Acid gas Mole Flow (m^3/d)	9755	9325	4.60
Rich Amine to Exchanger Temperature (°C)	50	47	7.33
Stripper Bottoms Temperature (°C)	110	106	3.66
Stripper Bottoms Pressure (kPa)	132	120	10.00
Lean Amine Temperature (°C)	48	41	17.18
Lean Amine Pressure (kPa)	9800	9611	1.97

(*) Relative Error (R.E.) (%) = Absolute ((Operating data-Simulation data)/Operating data)×100.

3.2. Formulation of the steady-state optimization

With the current design and conditions of the plant, the energy consumption of the reboiler unit is 7.0453 MJ/kg CO_2, and then we applied a strategy to reduce it. In order to define the limits where the process can be optimized in terms of energy consumption, the flexibility of the system is discussed. Initially, the natural gas to be treated (250 Mm^3/d or 441 kmol/h) contains 4 mol% of CO_2, reduced by countercurrent contact with the lean amine at 0.35 mol %. Subsequently, the sweet gas is mixed with a sour natural gas stream coming from the well, which is not CO_2 conditioned. As consequence, the final natural gas contains 1.667 mol% of CO_2, below the CO_2 specification.

Aim of the optimization problem is to minimize the reboiler demand (Q_{Reb}) per absorbed CO_2 mass unit in the contactor tower ($AbsCO_2$). According to Muhammad and GadelHak [31], the most important variables affecting the flow sheet are solvent flow rate and the absorber temperature, through the cooling of the lean amine stream. In this regard, the standard ideal volumetric flow of the lean amine to the absorption tower (u_1), the temperature (u_2), and, additionally, the pressure of the flash separation tank (u_3) are selected as decision variables. In Table 3, nominal values of the base case, for each decision variable are observed.

Similar to the objective function formulated by Mores et al. [32], here we minimize the reboiler energy ratio, subject to the variables u_1, u_2 and u_3 (Equation (1)). As mentioned, the molar fraction of CO_2 in the sales gas (y_{CO_2}) must be lower than 0.02 [23]. In

Table 3
Nominal values of the decision variables.

Stream	Decision Variable	Base Case
Lean Amine	u_1 : Std Ideal Liq Vol Flow (m^3/h)	53.18
Lean Amine	u_2: Temperature (°C)	41.32
To Separator	u_3: Pressure (kPa)	441.3

Table 4
Analyzed (steady-state) disturbance scenarios.

Disturbed variable	Scenario	Disturbance range
Feed Gas Mole Flow	d_1	+150 kmol/h (from 441 to 591 kmol/h)
	d_2	−150 kmol/h (from 441 to 291 kmol/h)
Feed Gas Temperature	d_3	+15 °C (from 35 °C to 50 °C)
	d_4	−10 °C (from 35 °C to 25 °C)
Feed Gas Pressure	d_5	+1000 kPa (from 6178 kPa to 7178 kPa)
	d_6	−1000 kPa (from 6178 kPa to 5178 kPa)

Table 5
Connections and discharge pressures of the main control valves.

Name	Charge	Discharge	Discharge pressure (kPa)
Valve	*Rich Amine*	*To Separator*	441.3
VLV-100	*Regen Feed*	*Regen Feed2*	200.0
VLV-101	*To pump*	*Lean Amine 2*	160.0
VLV-102	*Sweet Gas2*	*Sweet Gas3*	5688.0
VLV-103	*Acid Gas*	*Acid gas2*	12.4
VLV-Reflux	*Condenser out*	*Reflux*	110.0

accordance to Torres-Ortega et al. [33], we perform a sensitivity analysis to find the suitable ranges of variation for the decision variables along the optimization.

$$\begin{aligned} &\min_{u_1,u_2,u_3}\left(\frac{Qreb}{AbsCO_2}\right)\\ &s.t.:\\ &10 \le u_1 \le 70;\\ &25 \le u_2 \le 45;\\ &200 \le u_3 \le 1,000;\\ &y_{CO_2} < 0.02 \end{aligned} \tag{1}$$

In addition to the current conditions (B. C.), we also consider, for the steady-state optimization, scenarios related to disturbances in the conditions of the gas to treat, $d_1, d_2, d_3, d_4, d_5, d_6$ (Table 4).

The proposed nonlinear optimization (NLP) problem is solved using the optimization solver from Aspen Hysys v8.8. Among the methods available for its resolution, we use a mixed method that combines the BOX algorithm for the global convergence and the efficiency of the SQP. Using this hybrid method, we guarantee the global convergence through the SQP local solver, performing repetitive simulations changing successively the starting point of the simulation. The method initiates the minimization with the BOX method using a very loose tolerance, then the SQP method locates the final solution, according to the desired tolerance.

Although the number of equations required for the BOX convergence is restricted, the method offers an adequate robustness for process systems. On the other hand, SQP has demonstrated acceptable efficiency to solve problems with a reduced number of variables [34].

In spite of being a simple optimization expression, the minimum reboiler energy requirement involves a complex, rigorous, and integral resolution of the system (inner loop), for each iteration of the optimizer (outer loop) through the feasible-path strategy. The parameters in each iteration are solution accuracy, 0.001, iterations number, 100, and a maximum step equal to 0.05.

3.3. Cost estimation

As Kazemi et al. [35], we employ Aspen Economic Analyser

Table 6
Process Variable Source Object (OP), Process Variable (PV), Set Point (SP) and PID parameters of each controller.

Controller	OP	PV	SP	Kc	Ti (min)	Td (min)
FIC-101	VLV-101	*Lean Amine 2*	2088 kmol/h	0.5	0.2	–
TIC-100	Q_{ac1}	*Sweet gas 2*	28 °C	5	10	2
TIC-101	$Q_{Condenser}$	*Condenser Reflux*	53.48 °C	5	10	2
TIC-102	Q_{ac2}	*To Tank*	40 °C	5	10	2
PIC-102	VLV-103	*Acid Gas*	12 kPa	2	2	–
LIC-100	VLV-100	*Separator*	35%	2	1	–
LIC-101	VLV-Reflux	*Condenser*	45%	2	1	–

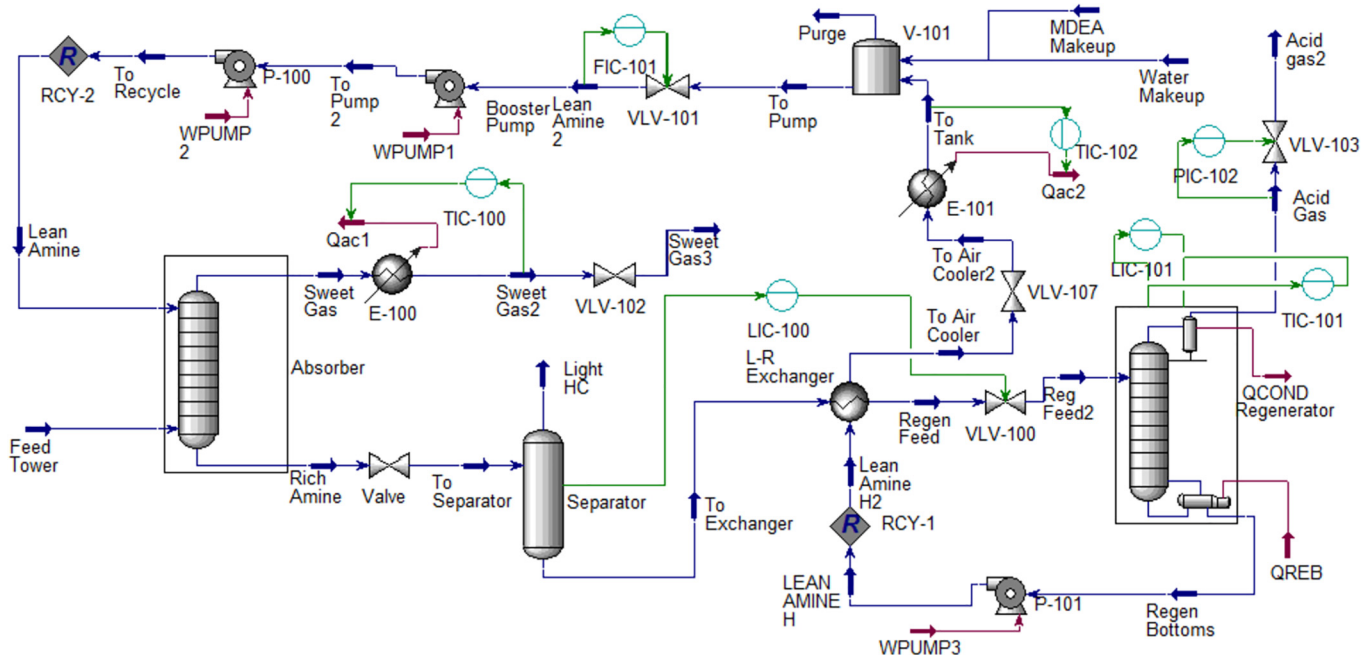

Fig. 2. Dynamic flowsheet modeled in Aspen Hysys v8.8.

Table 7
Nominal values of CO_2 mol fraction in the sales gas stream, the reboiler duty, and absorbed CO_2.

	B.C.	d_1	d_2	d_3	d_4	d_5	d_6
y_{CO_2}	0.0167	0.0154	0.0196	0.0167	0.01617	0.01671	0.0167
Q_{Reb} (MJ/h)	5096.8	5641.3	4519.7	5093.1	5093.7	5093.6	5093.0
$AbsCO_2$ (kg/h)	723.43	931.46	499.17	723.25	722.92	722.99	723.06

Table 8
Calculated optimal values for each decision variable.

	B.C.	d_1	d_2	d_3	d_4	d_5	d_6
u_1(m^3/h)	11.43	13.58	18.01	11.29	11.30	10.97	10.31
u_2 (°C)	37.79	38.73	49.92	39.87	42.59	44.16	42.63
u_3 (kPa)	532.8	628.3	528.4	600.2	559.4	578.2	585.4

Table 9
Optimal values of CO_2 molar fraction in the sales gas, the reboiler duty, absorbed CO_2 and objective function.

	B.C.	d_1	d_2	d_3	d_4	d_5	d_6
y_{CO_2}	0.0197	0.0200	0.0199	0.0199	0.0199	0.0199	0.0199
Q_{Reb} (MJ/h)	2316.5	2785.8	2397.5	2305.2	2325.9	2264.1	2144.1
$AbsCO_2$ (kg/h)	631.94	773.22	498.46	627.96	632.94	617.62	587.84

(AEA) to estimate operating and utility costs of the process and its optimization. Also Gutierrez et al. [18] calculate the operating expenses for the process of natural gas sweetening.

3.4. Control valves and controllers' configuration

In order to run the dynamic integrator correctly, the degrees of freedom of the entire flowsheet must be reduced to zero by specifying flows and pressures. For this, the specifications of the system input and output are set. In turn, the flow and pressure specifications of all internal material streams are released, as they are calculated by transitive considerations.

With regard to the absorber and regenerator dimensioning, design values are taken from the design of the gas plant, and verified by using the Tray Sizing tool of the simulator. On the other hand, an empirical method is used for estimating the size of the condenser, reboiler, makeup tank, and flash separator [27]. A residence time of 10 min and a 50% liquid level are considered for each unit. After the equipment sizing, the main control valves and proportional-integral-derivative (PID) controllers are located in the flowsheet of the process. Table 5 shows control valves placed in the simulation, and their connections and discharge pressures.

Linear type control valves are assumed, where the flow through the fitting is directly proportional to the opening percentage of the valve. In each of them, 50% valve opening is considered while the ANSI/ISA S75.01 standard method is applied to calculate the flow coefficients by using the pressure drop and the molar flow of the fluid as data [36].

In Table 6, the Source Object (OP), Process Variable (PV), Set Point (SP), and PID parameters are established for each PID controller. More precisely, it lists the controllers of lean amine flow (FIC-101), sweet gas temperature (TIC-100), reflux temperature (TIC-101), amine to makeup tank temperature (TIC-102), acid gas pressure (PIC-102), flash tank liquid level (LIC-100), and reflux accumulator liquid level (LIC-101).

Dimensionless gains (Kc) are considered in each case, as they were defined using the PID controller equation indicated by Marlin [37]. All the controllers are modeled in automatic mode. For the temperature (TIC-100, 101 and 102), pressure (PIC-102), and level (LIC-100 and 101) controllers, direct action mode is configured. In the case of the FIC-101 flow controller, a reverse configuration.fig.

3.5. Dynamic model

Fig. 2 shows the final flowsheet of the reference plant, in the simulation environment. It can be appreciated plant equipment, control valves, PID controllers and control loops.

The key part in Fig. 2 is the implementation of the controllers and the loops, especially we are interested in following the behavior of the decision variables and consequently we are interested in controllers FIC-101 and TIC-102.

4. Results

4.1. Steady-state optimization results

Table 7 lists the uncontrolled responses of the system −CO_2 mol

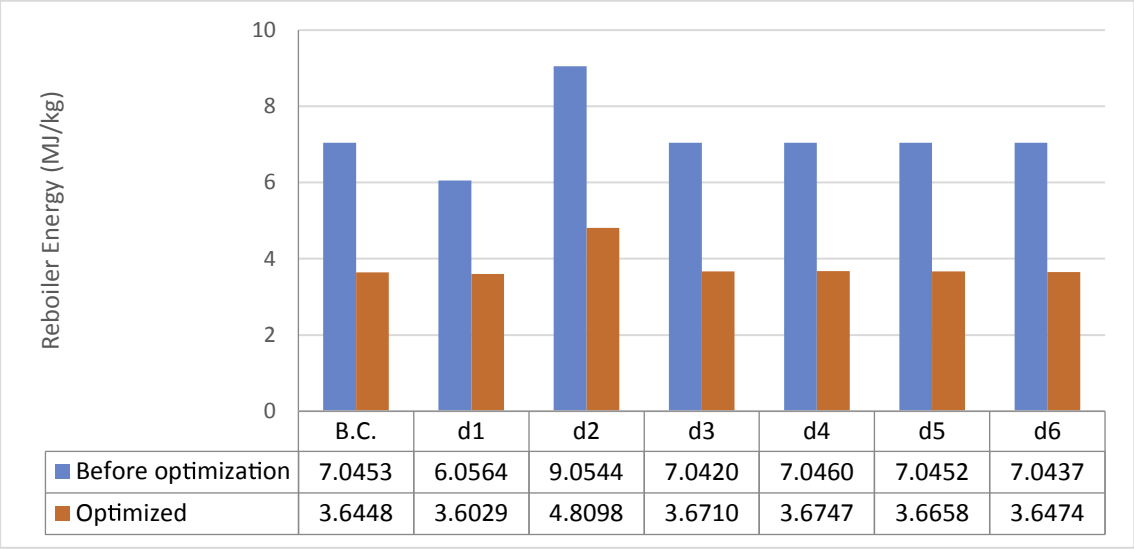

Fig. 3. Objective function results before and after minimization, for each scenario (MJ/kg).

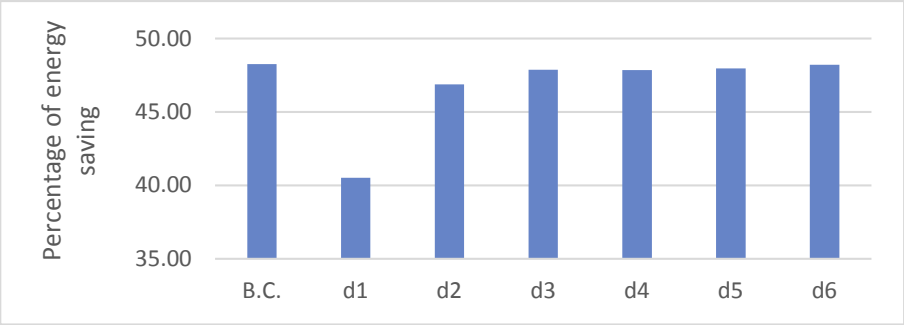

Fig. 4. Percentage of energy saving, for each scenario.

fraction in the sales gas stream (restriction), the reboiler duty and absorbed CO_2 in the absorber tower, before the optimization.

Table 8 reports the values of the decision variables after optimization. For the base case and for each disturbance scenario, it shows the modified values to maintain the objective function at a minimum point.

Table 9 summarizes the CO_2 restriction, the reboiler duty and the absorbed CO_2 after the optimization, for the B.C. and each disturbance scenario.

Fig. 3 shows the objective function results, the specific energy requirement $(Qreb/AbsCO_2)$ is presented for the base case and for each scenario under study.

Fig. 4 evidences the percentage of energy saving, for each scenario under consideration.

4.2. Operating and utility costs

As reported by Gutierrez et al. [18], the total operating cost for the plant under analysis was equal to 1,750,790 USD/y and the utility costs of 408,766 USD/y. In the case of the optimized case and following the methodology of AEA, we estimate an operating cost equal to 1,513,800 USD/y and a total utility cost of 196,393 USD/y.

4.3. Dynamic mode control response

This subsection is divided in two points. The first expose the ability of the model to reproduce the nominal values. While, the second exposes the transition from the base case to reach the optimal conditions.

4.3.1. Nominal values

Next Figure (Fig. 5) presents the evolution of each followed variable.

Fig. 5(a) shows the evolution of the lean amine (nominal) molar flow. It is observed that the output variable stabilizes in SP = 2088 kmol/h, after 113 min approximately. In this case (FIC-101), the controller parameters are $PV_{min} = 1,788 kmol/h$ and $PV_{max} = 2,388 kmol/h$.

Fig. 5(b) presents the evolution of the lean amine temperature. With the TIC-102 tuning values, a stable response is achieved before the 45 min of the dynamic simulation. $PV_{min} = 36\,^{\circ}C$ and $PV_{max} = 44\,^{\circ}C$ are selected for the controller, while the control valve supports cooling heat between 0 and 9.4E6 KJ/h.

In the same sense, the pressure of the gases leaving the regenerator shows a constant evolution over time. As shows Fig. 5(c), the evolution curve does not show fluctuations regarding the stabilization capacity. In Fig. 5(d), the temperature of the sales gas stream after cooling is shown. It is observed that it does not reach the value tuned until the first 130 min of simulation.

4.3.2. Optimal values

Finally, the dynamic response of the base case when the optimal values are implemented is studied. For the purpose, the B.C. optimal value of 11.43 m^3/h (283.4 kmol/h) is tuned in the FIC-101 while 36.27 °C is tuned in the TIC-102 controller.

In Fig. 6(a), the evolution of the lean amine (optimal) molar flow is presented. It is observed that the output variable stabilizes in SP = 283.4 kmol/h, after 33 min approximately. In this case (FIC-101), the controller parameters are $PV_{min} = 263.4\ kmol/h$ and $PV_{max} = 303.4 kmol/h$.

In Fig. 6(b), the evolution of the (optimal) lean amine temperature is observed. A stable response before the 23 min of the dynamic simulation is achieved. $PV_{min} = 32.27\,^{\circ}C$ and $PV_{max} = 40.27^{\circ}C$ are selected for the controller while the control valve supports cooling heat between 0 and 1.08E6 KJ/h.

5. Discussion

In Tables 2 and 7, it is observed that the optimal lean amine flow rate (u_1) is reduced from 53.18 to 11.43 m^3/h. Regardless of the scenario considered, the decrease of the recycled solvent implies an opportunity for cost saving [24].

Particularly in cases d_2, d_4, d_5 and d_6, the optimal temperatures of the lean amine (u_2) are higher than that of the B.C. (41.32 °C). This temperature increment reduces the heat to be removed from the air-cooler E-101 and consequently a lower utility cost might be expected.

It was found that the flash pressure (u_3) raising from 441.3 to 532.8 kPa can imply a lower amount of the fuel to be burned in the reboiler unit, reducing the Q_{Reb} from 5096.76 to 2316.50 MJ/h (Tables 6 and 8).

In Table 8, it can be appreciated that the molar fraction (y_{CO_2}) does not exceed the restriction of 2%, in the analyzed scenarios. Therefore, the modification of the decision variables preserves the O.F. at the lowest point without exceeding the sale specification.

In all the cases, a considerable reduction of the O.F. with respect to the nominal values is observed. The graphed value of 3.64 MJ/kg in the B.C. agrees with those reported by Cousins et al. [38] and Gervasi et al. [39]. In all optimization scenarios, reductions between 47% and 49% are observed, with the exception of scenario d_1 where the specific heat reduction is 40% (Fig. 4).

When considering the operating costs of the BC, an economic save of 236,990 USD/y is observed respect to the operating cost after the optimization. As regards the utility cost a significant decrease of 52%, respect to the BC in current conditions.

As Gaspar et al. [17] stated, various disturbances appear in practice due to the complexity of the absorption process. In Fig. 5, it can be noticed how the simulator initializes the state variables which explains the peaks during the transformation. However, it is

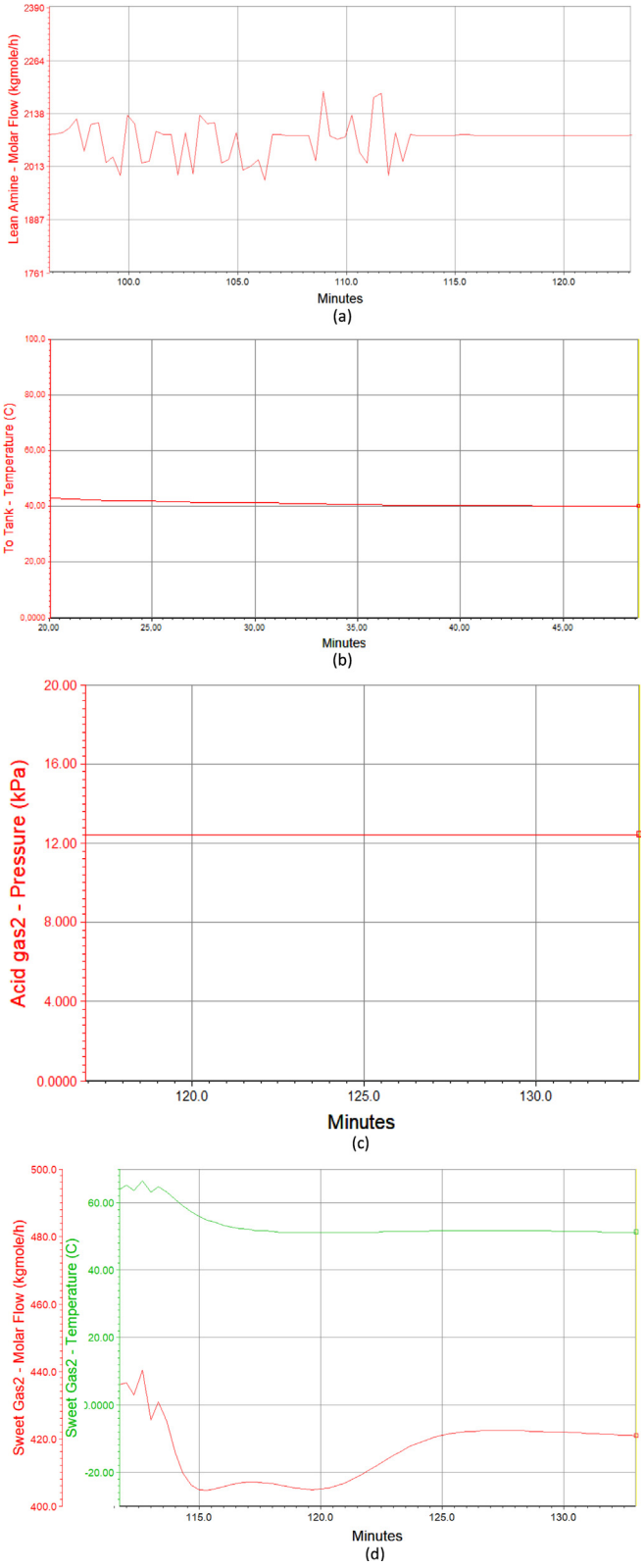

Fig. 5. B.C. controllers' response of (a) FIC-101; (b) TIC-102; (c) PIC-102; (d) T-100.

demonstrated that the dynamic model can reach the set points and the nominal values of the B.C.

In addition, the dynamic response of the B.C. controllers is studied when the optimum values are replaced as set points (Fig. 6).

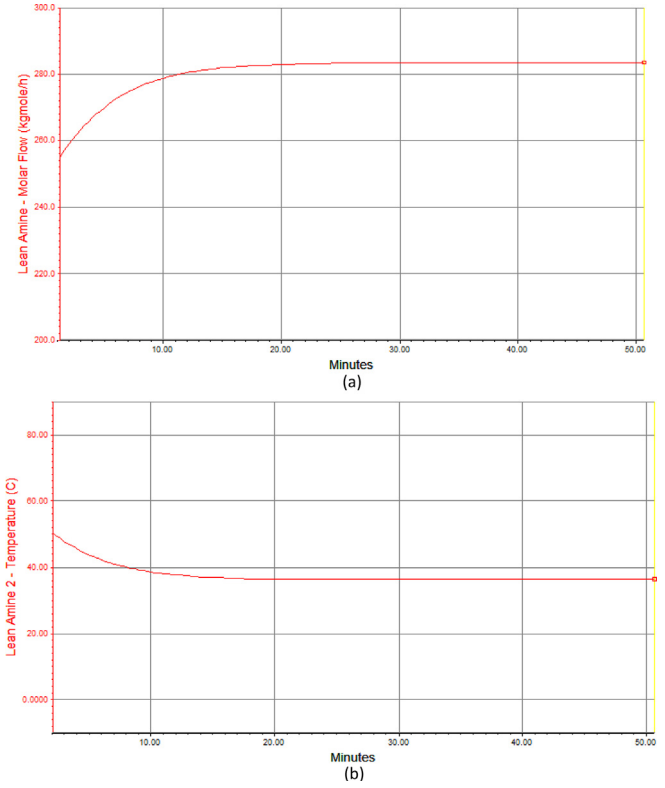

Fig. 6. Controllers' response towards the optimal conditions: (a) FIC-101; (b) TIC-102.

In respect to the transition, the stabilization of system is obtained faster than that of the B.C. and this implies that the optimal values might be implemented in the reference plant.

Suleiman et al. [25] state that a reboiler energy decrease translates directly into utility cost reduction, with positive impact on the process economy. This work proves that the modification of the analyzed variables reduces the main energy requirement of the plant in a 48%, in current conditions.

6. Conclusions

In the simulated model, it is possible to establish optimal conditions for the process in the reference plant. In addition, the optimization of the system is proposed under different scenarios of disturbances, related to the inlet natural gas conditions.

The system stability is proved: the optimized case stabilizes faster than the base case. The introduced controllers serve to establish an appropriate set of variables and manage to keep their values in the SP. Then, the calculated optimal values can be implemented in the existing plant using the current control scheme.

As this work demonstrates, almost a 50% decrease is achieved in the main energy requirement, for current conditions. Even when inlet gas disturbances, the performed model is able to reduce the main energy demand of the process under study. In this regard, a significant utility cost reduction maybe conducted with the found modifications.

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