Advanced Control Solution for a Refinery Gas Sweetening Process

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Gas sweetening unit, also known as acid gas removal unit, is a wide used unit in refineries and petrochemical plants and refers to a group of processes that use different amines such DiEthanolAmine (DEA) to remove the hydrogen sulfide (H_2S) from gases, in order to use them as fuel to different furnaces from other processing units. If the hydrogen sulfide is not well removed, will lead to furnaces tubular corrosion and therefore will generate loss in refractory properties. Also, the emission values of the exhaust gases will not be in accordance with the environmental regulations. The paper outlines the main results obtained by authors concerning the advanced control solution for a refinery gas sweetening unit and contains three parts. The first part outlines the results of the dynamic gas sweetening process investigation using HYSYS® simulation software. The second part presents the proposed structure of the advanced control system and the third part outlines the results obtained using the control structure in order to remove as good as possible the H_2S content from residual gases.

Keywords: gas sweetening process, advanced control, internal model control

One of the objectives of a gas sweetening process is the compliance with the emission standards in accordance with environmental regulations. Also, on the proper functioning of this system are based all process units that use as fuel the sweetened gas because it is more profitable to process the gasses from other units and to use them as fuel, than to use methane gas from the national network. Because the sweetened gas is input for furnaces, if the hydrogen sulfide (H_2S) from gases is not well removed, this will lead to corrosion and loss in furnaces refractory properties.

In a refinery process, in terms of weight of energy consumption, the raw materials preheating furnaces or reboilers stands out. If the refinery has its own Gas Turbine Cogeneration Plant it is more economically advantageous that the refinery gas network to be reconfigured taking into account this consumer. An alternative to the refinery gases used as fuel for furnaces is the methane gas from the national networks [1]. However, this alternative is not advantageous in terms of the involved costs, so the optimal configuration of the refinery gas networks must include secondary flows resulting from the purge gas or waste streams.

The waste gas streams that feed the refinery network can have the following origin:

• vents from the separator vessels of the hydrofining and hydrocracking units;

• the hydrogen-rich gas excess from catalytic reforming plants;

• gas flows from the reflux vessels of the stripping and distillation columns etc;

In order to reduce the corrosive effect on tubular furnaces and emissions, a pretreatment by absorption in amine solutions, of these gas flows is indicated, before sending them to the refinery fuel gas network.

In crude oil streams sulfur may appear as compounds (such mercaptans, sulfides, disulfides), as hydrogen sulfide (H_2S) or as elemental sulfur. Between these, hydrogen sulfide is a primary contributor to corrosion on the iron and steel used in refinery process equipment, piping and tanks within refinery processing units. Also, the combustion of petroleum streams containing sulfur compounds produces sulfuric acid and sulfur dioxide and undesirable affects the environment [2].

Having in mind the above aspects, the authors concerning was to find a control solution for the sweetening gas process in order to maintain the hydrogen sulphide concentration close to specifications.

Gas sweetening process dynamic investigation using HYSYS® simulation software

The sweetening process structure is presented in (fig. 1). From this figure can be seen that the most important equipments are the two columns: the absorption column (ABSORBER) of the impurities (typically hydrogen sulfide) using a solution of DiEthanolAmine (DEA) and the regeneration column (REGENERATOR) which is designed to regenerate the absorbent (DEA solution) [3].

Because both, the overhead stream of the absorption column and the bottom stream of the regeneration column are main streams, the optimal operation of these two columns is important, the technical specifications achievement depending on the accuracy of their functioning [4, 5].

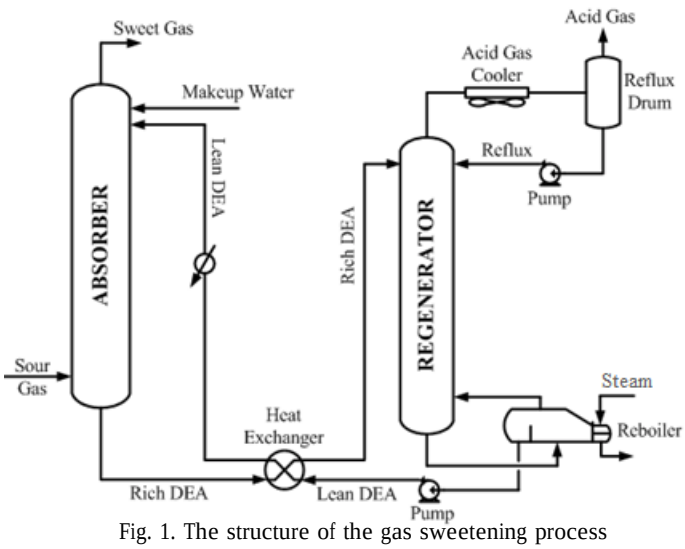

Fig. 1. The structure of the gas sweetening process

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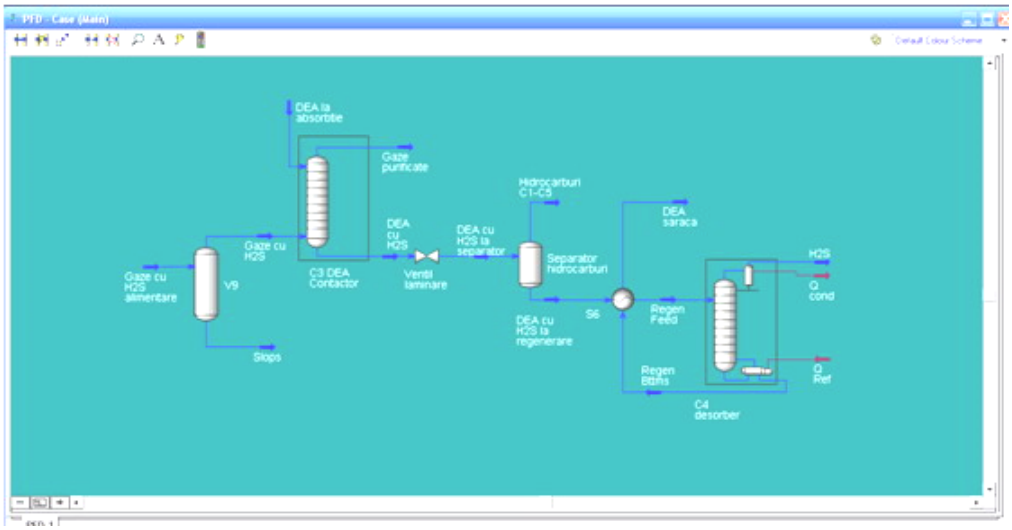


Fig. 2. The gas sweetening process structure, implemented with HYSYS®

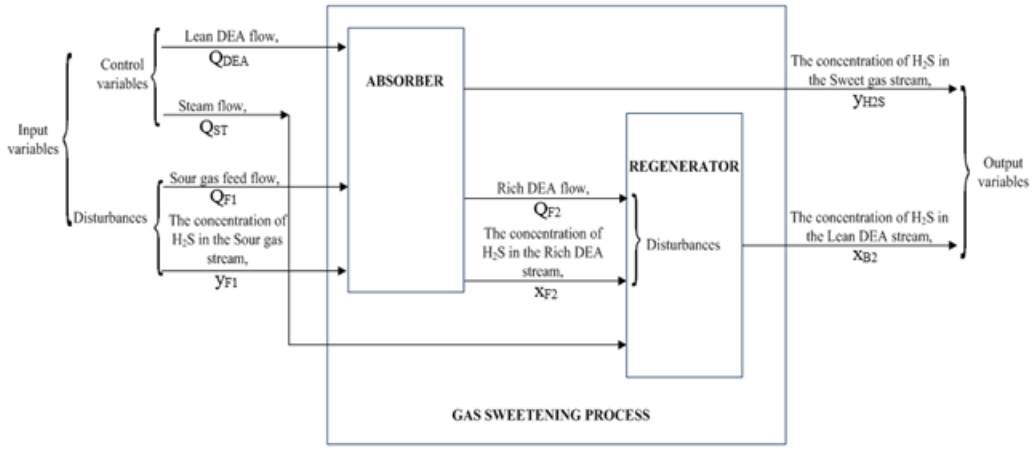


Fig. 3. Gas sweetening process block diagram

A classical control structure for the two columns involved in the sweetening process has the Lean DEA flow rate as control variable for controlling the concentration of H_2S in the top (Sweet gas) stream, for ABSORBER, and the Steam flow rate for controlling the concentration of H_2S in the bottom (Lean DEA) stream, for REGENERATOR (fig.1).

Taking into account the final use and amounts of use, the specifications represented by the concentration of H_2S in the Sweet gas stream from the absorber column, and in the regenerated absorbent (Lean DEA) from the regenerator column, have different characteristics. Thus, the concentration of H_2S in the Sweet gas stream must be 100 ppm; this specification is a hard restriction. The concentration of H_2S in the stream of regenerated absorbent specification corresponds to the interval 50 ... 70 ppm. This specification is a soft one for the control system.

The dynamic behaviour of the gas sweetening process was studied using the HYSYS® simulation software (fig. 2) [6, 7].

As systemic approach (fig. 3), the ABSORBER from the gas sweetening process has as output the concentration of H_2S in the top (Sweet gas) stream (y_{H2S}) and as inputs the Lean DEA flow (Q_{DEA}) as control variable and the Sour gas feed flow (Q_{F1}) and concentration (y_{F1}) as disturbances. The REGENERATOR has as output the concentration of H_2S in the bottom (Lean DEA) stream (x_{B2}) and as inputs the steam flow (Q_{ST}) as control variable and the Rich DEA feed flow (Q_{F2}) and concentration (x_{F2}) as disturbances.

The control variables from (fig. 3) are named also *manipulated variables* and the output variables, are named *controlled variables*.

The gas sweetening process dynamic behavior analysis using the HYSYS® simulation software consisted in observing the dynamic evolution of the output variables for each of the two columns to a step change in the input variables (control variables or disturbances).

Analyzing the results indicated in [6, 7] it was observed that the process has a nonlinear behavior, characterized by different gains and transient times for different process operating ranges and process channels (fig. 3).

In [6, 7] models of the process were determined for every process channel (fig. 3) and different operating ranges. In this paper will be used only the models for *manipulated variable – controlled variable* pairs from each column.

The model is represented by a first order transfer function, for the ABSORBER and by a second order transfer function, for the REGENERATOR, as follows:

$$G_{Q_{DEA}-y_{H_2S}} = \frac{k_{p1}}{T_{p1} \cdot s + 1}, \quad (1)$$

$$G_{Q_{ST}-x_{B_2}} = \frac{k_{p2}}{T_{p22}^2 \cdot s^2 + T_{p21} \cdot s + 1}, \quad (2)$$

where k_{P1} is the absorber process gain, T_{p1} is the absorber process time constant and k_{P2} is the regenerator process gain, T_{p22} and T_{p21} are the regenerator process time constants, having the values from tables 1 and 2.

The minus sign that accompanies the two process gains (k_{P1} and k_{P2}) signifies that a control variable (Q_{DEA} or Q_{ST}) increase will lead to an output variable (y_{H2S} or x_{B2}) value decrease.

The proposed control structure

Because the process model was already found and presented in [6, 7], this paper will consider the internal

Table 1
THE PARAMETER VALUES FOR THE ABSORBER COLUMN PROCESS MODEL VALID FOR DIFFERENT PROCESS OPERATING RANGES [6, 7]

y_{H2S}* [mass fr.]	k_{p1}	T_{p1} [min]
0.001 - 0.003	-0.026	9.4
0.003 - 0.005	-0.028	7.5
0.005 - 0.007	-0.031	6.4
0.007 - 0.009	-0.033	5.7
0.009 - 0.011	-0.036	4.4
0.011 – 0.013	-0.037	3.5
0.013 – 0.015	-0.038	2.5
Mean values	-0.033	6.0

y_{H2S} is the concentration of H_2S in the top (Sweet gas) stream of the absorber column

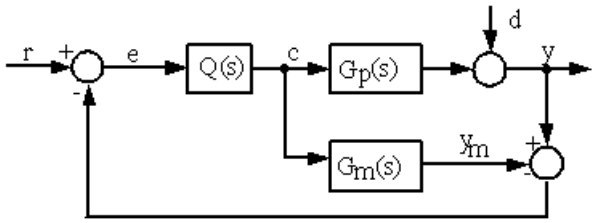


Fig. 4. Internal Model Control structure [8]:
Q(s) – the primary controller transfer function, $G_p(s)$ – the process transfer function, $G_m(s)$ – the process model transfer function, r – setpoint, e – error, c – control variable, d – disturbance, y_m – model output, y –process output.

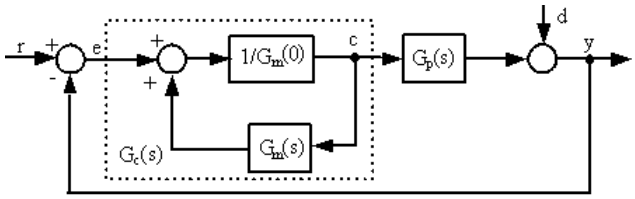


Fig. 5. Internal Model Control alternative structure [10]

Table 2
THE PARAMETER VALUES FOR THE REGENERATOR COLUMN PROCESS MODEL VALID FOR DIFFERENT PROCESS OPERATING RANGES [6, 7]

x_{B2}* [mass fr.]	k_{p2}	T_{p22} [min]	T_{p21} [min]
0.001 - 0.003	-0.132	25.8	94.2
0.003 - 0.005	-0.131	23	74.4
0.005 - 0.007	-0.122	22	67.4
0.007 - 0.009	-0.117	21.4	59.2
0.009 - 0.011	-0.114	20.8	55.2
Mean values	-0.123	22.6	58.4

x_{B2} is the concentration of H_2S in the top (lean DEA) stream of the regenerator column

model control structure, which is an advanced control structure based on the process model and distinguishes by its simplicity in implementation and use.

This control structure was implemented with good results for propylene/propane distillation column [8, 9]. Because there are some similitudes between the two processes (propylene/propane column and gas sweetening process) regarding the dynamic behavior, the authors concern was to see if this control structure offers also good results for gas sweetening process.

The Internal Model Control system has the structure illustrated in (fig. 4).

In order to have a null steady-state error for a step change in setpoint or a step change in disturbance it is required that the control system to be stable and the controller static gain to be equal to the inverse of the model static gain [8]:

$$Q(0) = \frac{1}{G_m(0)}. \quad (3)$$

The simplest form for internal model strategy is the one in which the transfer function Q(s) is chosen as a zero order transfer function, equal to the inverse of the model gain:

$$Q(s) = \frac{1}{G_m(0)}. \quad (4)$$

The controller transfer function, which contains the primary controller *Q* and the model G_m, is:

$$G_C(s) = \frac{Q(s)}{1 - Q(s) \cdot G_m(s)} = \frac{1}{G_m(0) - G_m(s)}. \quad (5)$$

A more suggestive equivalent structure, is represented in (fig. 5).

Results and discussions

An advanced internal model control (IMC) solution was implemented for controlling the concentration of H_2S in the Sweet gas stream for the absorption column and the concentration of H_2S in the Lean DEA stream for the regeneration column, from a gas sweetening process.

As it has been previously pointed out, the refineries sour gases pretreatment by absorption in amine solutions must be used, in order to reduce the corrosive effect on tubular furnaces, before sending them to the refinery fuel gas network. As a result of gases absorption in amine solutions process, the content of H_2S from gases must be substantially reduced, H_2S being the main contributor for tubular furnaces corrosion and a source of pollution. Maintaining the H_2S specifications is an important task of the control structure. Based on the control system performance relies on the entire process of gases recovery and further use as fuel, in order to reduce the refineries energy consumption. This action has also a great impact

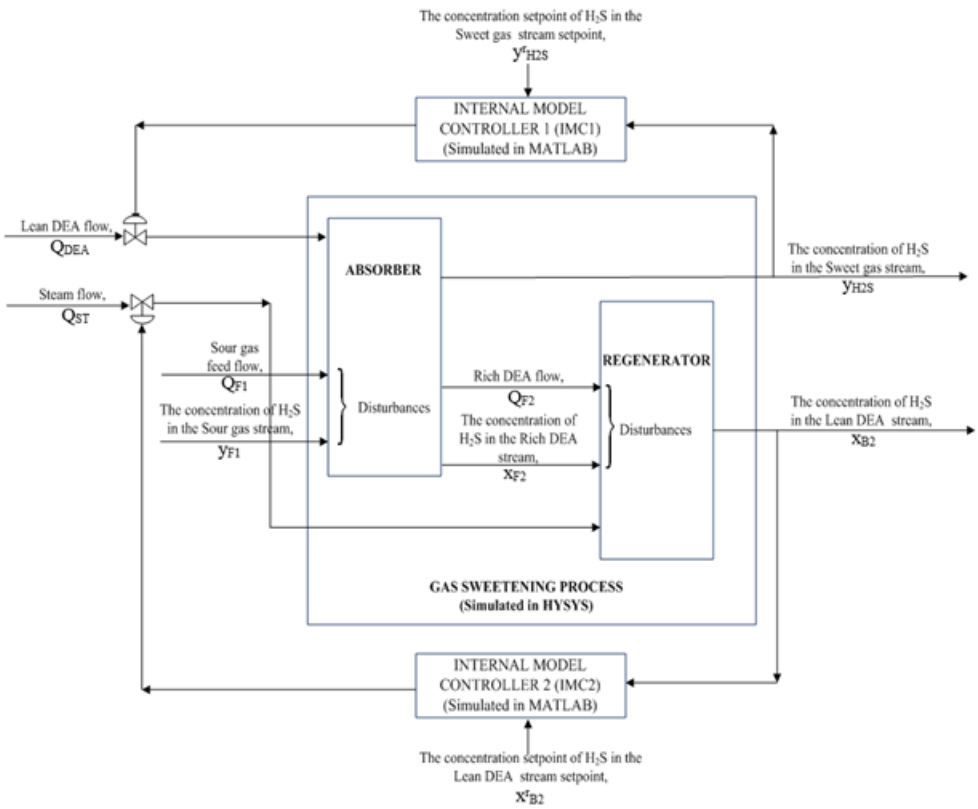


Fig. 6. The proposed Internal Model Control structure

on the environment by reducing the dangerous gas emissions.

Both objectives, the process efficiency and environment security are important aspects that should be taken in consideration for a good control system operation [11].

The proposed control system structure is presented in (fig. 6).

The two internal model controllers IMC 1 and IMC 2 (fig. 6) are used in order to control the concentration of H_2S in the Sweet gas stream, from the ABSORBER and the concentration of H_2S in the Lean DEA stream from the REGENERATOR, respectively.

The IMC controllers have the structure from (fig. 5) with $G_m(s)$ given by relations (1 or 2) and with the model parameter values from tables 1 or 2.

Because the gas sweetening process, represented by the two columns, is a nonlinear one, but was linearized for different operating ranges (relations 1 and 2 and tables 1 and 2), we have to use an IMC algorithm with model parameters adaptable to the process operating range.

For example, in case of changing the concentration of H_2S in the Sweet gas stream controller setpoint from 0.002 to 0.004 H_2S mass fr., the system automatically senses when the operating range changes and loads the adequate model parameters for IMC 1, accordingly with table 1: for the process operating range from 0.001 to 0.003 H_2S mass fr., the IMC 1 model parameters are: k_{P1}=-0.026, T_{P1}=9.4 min and for the process operating range from 0.003 to 0.005 H_2S mass fr., the IMC 1 model parameters are: k_{P1}=-0.028, T_{P1}=7.5 min.

If the control system is a classical one, having the controller with nonadaptable tuning parameters (which do not adapt to the current operating range), whatever the process operating range, the model parameter values for IMC 1 will be equal to the mean values k_{P1}=-0.033, T_{P1}=5.7 min.

This is the case of a test presented in (fig. 7).

The proposed control structure was simulated in MATLAB®, while the process remained simulated in HYSYS®.

Further are presented the results obtained using the proposed IMC structure.

The tests consisted of changing the setpoint values for the two controllers in case of using the IMC algorithm with adaptable model parameters to the process operating range and in case of using the nonadaptable method, when the controller has as parameters the mean values.

As we can observe from (fig. 7a), the concentration becomes equal with its setpoint with good dynamic performance, because another model parameters are considered when the process operating range is changing,

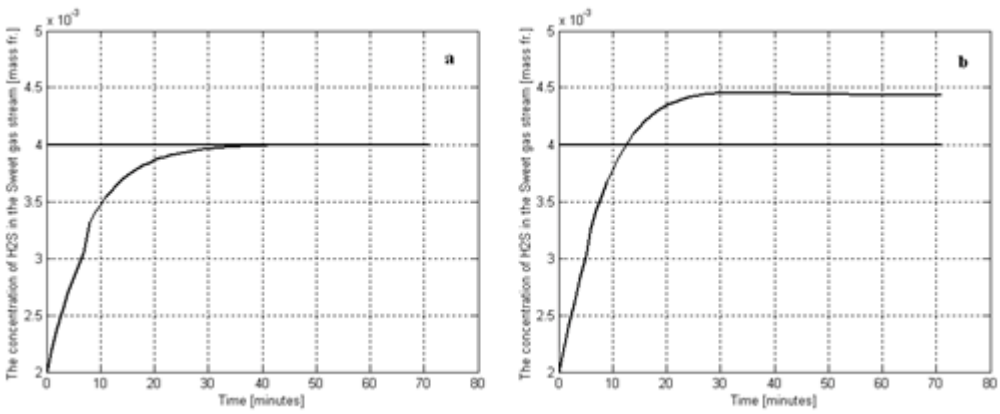


Fig. 7. The concentration of H_2S in the Sweet gas stream trend when the controller setpoint increases from 0.002 to 0.004 H_2S mass fr., using an IMC: **a** - with model parameters adaptable to the process operating range (k_{P1}=-0.026, T_{P1}=9.4 min, k_{P1}=-0.028, T_{P1}=7.5 min), **b** - with nonadaptable model parameters (k_{P1}=-0.033, T_{P1}=6 min).

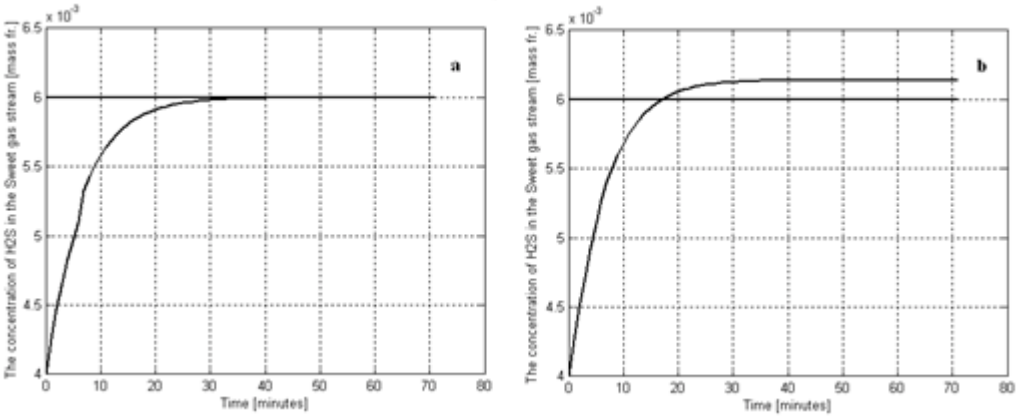


Fig. 8. The concentration of H_2S in the Sweet gas stream trend when the controller setpoint ncreases from 0.004 to 0.006 H_2S mass fr., using an IMC: **a** - with model parameters adaptable to the process operating range (k_{P1}=-0.028, T_{P1}=7.5 min, k_{P1}=-0.031, T_{P1}=6.4 min), **b** - with ıonadaptable model parameters (k_{P1}=-0.033, T_{P1}=6 min)

so that the process nonlinearities are compensated. In case of using the IMC with nonadaptable model parameters (fig. 7b), we have an important steady-state error because the model from the controller is not the right one, and the process nonlinearities remain uncompensated.

In the situation illustrated in (fig. 8), if the IMC model parameters are adapting to the process operating range (fig. 8a), the steady-state error is zero and the dynamic performance are good (no output overshoots and a small transient time). If the IMC model parameters do not adapt to the process operating range (fig. 8b), and have the mean values, we have steady-state error but smaller in comparison with the previous case (fig. 7b) because the values of the process model are closer to the mean values considered for the model from controller.

If the IMC model parameters are adapting to the process operating range (fig. 9a), the process nonlinearities are compensated and the control system has good steady-state and dynamic performance. If the IMC model parameter values are the mean ones (fig. 9b), due to the fact that the model parameters values are even closer to the values of the process, we do not have steady-state error, in comparison with the results from (fig. 7b and 8b), only the transient time is a little bit greater than the one from (fig. 9a).

The results obtained with the proposed control structure were compared with the ones that were obtained using a PI controller (fig. 10), that has also tuning parameters that adapt to the operating range, having $k_R=0.9/k_{P1}$ and $T_i=T_{p1}$.

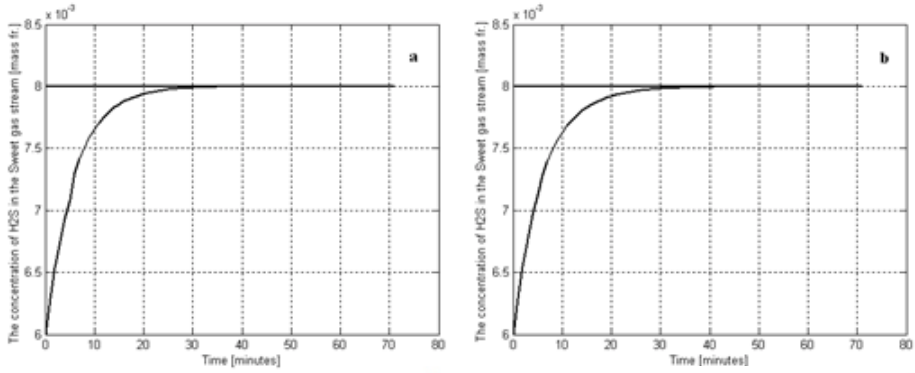
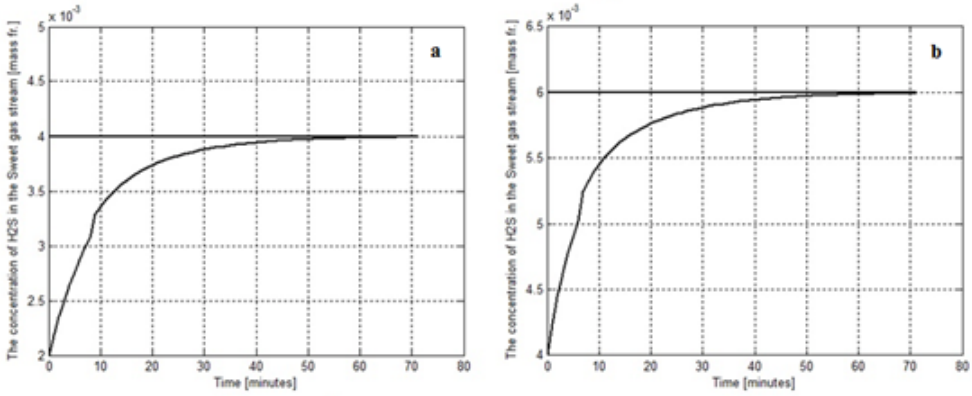


Fig. 9. The concentration of H_2S in the Sweet gas stream trend when the controller setpoint increases from 0.006 to 0.008 H_2S mass fr., using an IMC: **a** - with model parameters adaptable to the process operating range (k_{P1}=-0.031, T_{P1}=6.4 min, k_{P1}=-0.033, T_{P1}=5.7 min), **b** - with nonadaptable model parameters (k_{P1}=-0.033, T_{P1}=6 min)



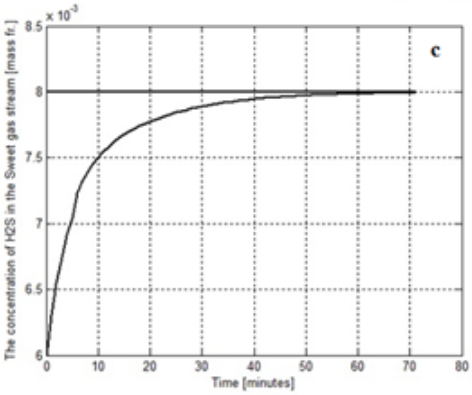


Fig. 10. The concentration of H_2S in the Sweet gas stream trend when a PI controller with adaptable tuning parameters is used: a – H_2S setpoint increases from 0.002 to 0.004 mass fr., b - H_2S setpoint increases from 0.004 to 0.006 mass fr., c - H_2S setpoint increases from 0.006 to 0.008 mass fr..

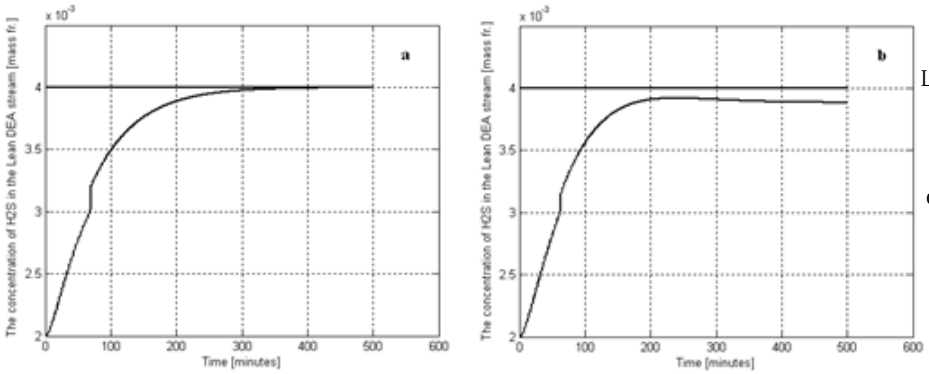


Fig. 11. The concentration of H_2S in the Lean DEA stream trend when the controller setpoint increases from 0.002 to 0.004 H_2S mass fr., using an IMC: **a** - with model parameters adaptable to the process operating range (k_{p2}=-0.132, T_{p22}=25.8 min, T_{p21}=94.2 min, k_{p2}=-0.131, T_{p22}=23 min, T_{p21}=74.4 min), **b** - with nonadaptable model parameters (k_{p2}=-0.123, T_{p22}=22.6 min, T_{p21}=58.4 min)

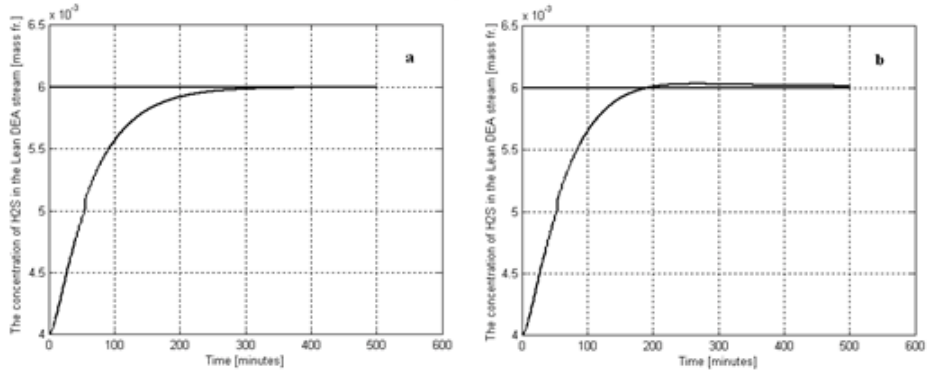


Fig. 12. The concentration of H_2S in the Lean DEA stream trend when the controller setpoint increases from 0.004 to 0.006 H_2S mass fr., using an IMC: **a** - with model parameters adaptable to the process operating range (k_{p2}=-0.131, T_{p22}=23 min, T_{p21}=74.4 min, k_{p2}=-0.122, T_{p22}=22 min, T_{p21}=67.4 min), **b** - with nonadaptable model parameters (k_{p2}=-0.123, T_{p22}=22.6 min, T_{p21}=58.4 min)

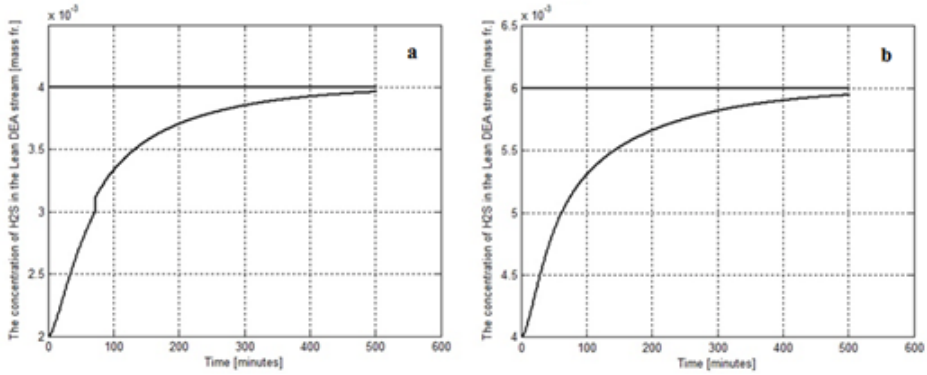


Fig. 13. The concentration of H_2S in the Lean DEA stream trend when a PI controller with adaptable tuning parameters is used: a – H_2S setpoint increases from 0.002 to 0.004 mass fr., b - H2S setpoint increases from 0.004 to 0.006 mass fr..

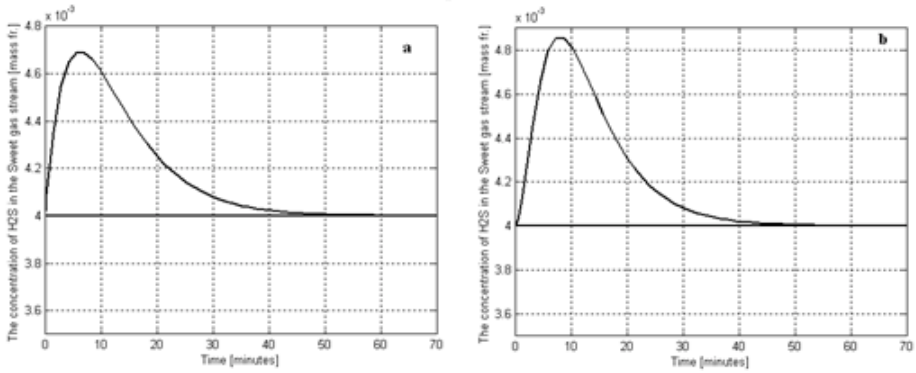


Fig. 14. The concentration of H_2S in the Sweet gas stream trend when the IMC setpoint is set at 0.004 H_2S mass fr., when: **a** – the Sour gas feed flow (Q_{F1}) increases with 50 kmol/h, **b** – the Sour gas H_2S concentration (y_{F1}) increases with 0.002 mass fr..

As we can see, when a PI controller is used (fig. 10) the control system dynamic performace is worse than in case of using the proposed IMC control structure (fig. 7a, 8a, 9a). In case of using the PI controller the control system transient time is higher than in case of using the IMC controller.

In case of using IMC with adaptable model parameters (fig. 11a), the control action is done with good steady-state and dynamic performance. If the IMC model parameters are the mean computed values (fig. 11b) the steady-state error differs from zero, in comparison with the results from (fig. 11a).

As we can observe from (fig. 12a), the concentration becomes equal to its setpoint with good dynamic performance, because other model parameters are considered for each process operating range change and the process nonlinearities are compensated. When an IMC with nonadaptable model parameters is used (fig. 12b), the process output has overshoot and the control system transient time increases.

In (fig. 13) are presented the results obtained in case of using a PI controller that has tuning parameters that adapt to the operating range, having $k_R=0.9/k_{p2}$ and $T_i=T_{p21}$.

As we can see, when the PI controller is used (fig. 13) the control system dynamic performance is worse than in case of using the proposed IMC (fig. 11, 12), because the control system transient time is higher.

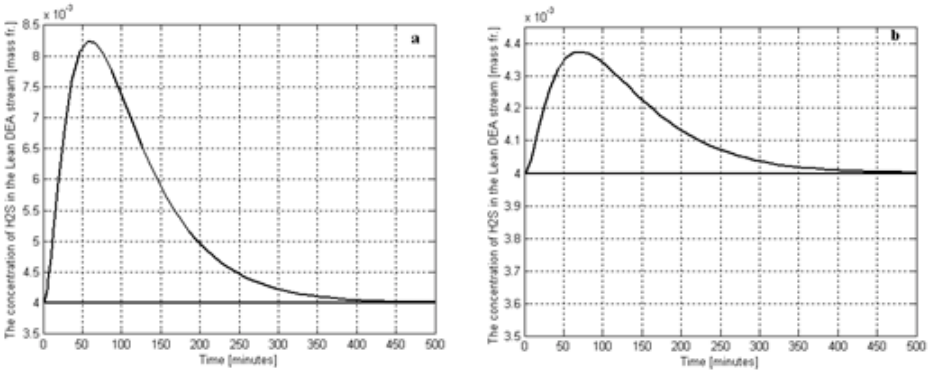


Fig. 15. The concentration of H_2S in the Lean DEA stream trend when the IMC setpoint is set at 0.004 H_2S mass fr., when: **a** – the Rich DEA feed flow (Q_{F2}) increases with 50 kmol/h, **b** –the Rich DEA H_2S concentration (x_{B2}) increases with 0.002 mass fr

For the tests was also considered the case when we have a step change of the disturbances Sour gas feed flow (fig. 14a) or H2S concentration (fig. 14b) for the ABSORBER and Rich DEA feed flow (fig. 15a) or H2S concentration (fig. 15b) for the REGENERATOR.

As we can observe from the above test results (fig. 14 and 15) when a disturbance appears, its effect on the process output is eliminated by the proposed Internal Model Control structure.

Conclusions

An advanced control structure, having simple Internal Model Controllers (IMC), was considered to control the concentration of H_2S in the Sweet gas stream and in the Lean DEA stream, for the ABSORBER and REGENERATOR units from a gas sweetening plant, in order to remove the hydrogen sulfide from the gases which will further be used as fuel to furnaces from others plants. This is done to prevent tubular corrosion and loss in refractory properties and the obtained emission values to be in accordance with the environmental regulations.

Two cases were considered: the first is when the IMC model parameters are adapting to the process operating range and the second when the IMC model parameters are the mean values of the process parameters and the same values are used on the whole operating range.

From the simulation results it was observed that the first proposed control structure offers good steady-state and dynamic performance because the responses do not have steady-state error, there is no overshoot and the process transient time is relatively small. In the second case, we have considerable steady state error or overshoot and the control system transient time increases.

The simulation results were also compared with the results that were obtained with PI controllers that use adaptable tuning parameters, showing that in case of using the advanced control structure the control system has a better dynamic performance, characterized by shorter transient times.

In conclusion, using the proposed advanced control solution the concentration of the hydrogen sulphide is maintained close to specifications, with good static and dynamic results, in order to prevent the corrosive effect on the tubular furnaces that uses the sweetened gas as fuel, and protecting the environment from undesirable emissions.

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