ADVANCED CONTROL FOR AN ETHYLENE REACTOR

Dumitru POPESCU[1],
Catalin DIMON[2], Severus OLTEANU[3], Ciprian LUPU[4]

Rezumat. *Obiectivul central al acestei lucrări este de a dezvolta si implementa o soluție de control pentru procese petrochimice și anume controlul si optimizarea unui reactor de piroliza, instalație cheie in industria petrochimică. Sunt prezentate caracteristicile tehnologice ale acestui proces petrochimic și unele aspecte despre sistemul de control propus pentru instalația de etilenă. În cele din urmă, o soluție optimală este găsită, considerând că procesul are o structură neliniară multivariabilă. Rezultatele au fost implementate pe un ansamblu de reactoare de piroliză pe o platformă petrochimică din România.*

Abstract. *The main objective of this paper is the design and implementation of control solutions for petrochemical processes, namely the control and optimization of a pyrolysis reactor, the key-installation in the petrochemical industry. The authors present the technological characteristics of this petrochemical process and some aspects about the proposed control system solution for the ethylene plant. Finally, an optimal operating point for the reactor is found, considering that the process has a nonlinear multivariable structure. The results have been implemented on an assembly of pyrolysis reactors on a petrochemical platform from Romania*.

Keywords: *ethylene pyrolysis, numerical control system design, robust control, optimization*

1. Introduction

The petrochemical industry is still a fertile field from the perspective of the automatic control of technological processes and therefore, some of the most representative applications find their place in this area. In recent decades, the petrochemical industry has experienced an unprecedented development by upgrading their equipment and the production lines and also, by expanding their production capacity. The petrochemical industry is an important provider of products for population and industry (food, pharmaceutics, mechanics, electronics, textiles, transportation) and remains a priority for the Romanian economy.

[1]Prof. Eng., Faculty of Automatic Control and Computer Science, University "*Politehnica*", Bucharest, Romania, full/corresponding member of the Romanian Academy (popescu_upb@yahoo.com).
[2]Prof. Eng., PhD, Faculty of Automatic Control and Computer Science, University "*Politehnica*", Bucharest, Romania.
[3]Researcher, "Control Engineering & Energy" SRL, Bucharest, Romania.
[4]Prof. Eng., Faculty of Automatic Control and Computer Science, University "*Politehnica*", Bucharest, Romania.

Today, the demand for ethylene is over 125 million tons per year with a growth rate of 3.5 % per year. The average capacity of production plant has risen from 300 KTA in the 1980's to over 1000 KTA today and is expected to grow beyond 2000 KTA in the near future ([1], [2]).

Since ethylene is one of the raw materials in the chemical industry, and the product market situation, as well as its utility are rapidly changing, the optimal operation and control of the plant is important. Ethylene is produced mainly by thermal cracking of hydrocarbons in the presence of steam, and by recovery from refinery cracked gas [7].

In steam cracking, a gaseous or liquid hydrocarbon feed like naphtha, LPG or ethane is diluted with steam and briefly heated in a reactor without the presence of oxygen.

Typically, the reaction temperature is very high, at around 850 °C, the pressure is low - 4 bar, but the reaction is only allowed to take place very briefly.

In modern cracking furnaces, the residence time is even reduced to milliseconds, resulting in gas velocities faster than the speed of sound, to improve yield. After the cracking temperature has been reached, the gas is quickly quenched to stop the reaction in a transfer line heat [3]. As a general rule, a steam cracking reactor uses radiant heat delivered by burners such as conventional gas or oil burners as its source of heat. The burners are often placed on the floor and/or the walls of the reactor and they define a high temperature zone in the furnace, also referred to as the "radiation" zone of the furnace. Immediately above said zone, there is a convection zone through which the hot combustion gases escape from the radiation zone; this convection zone is generally used for preheating the mixture of hydrocarbons to be cracked, also known as the "feedstock" to be cracked. Thus, the mixture is usually preheated to about 500 ^{o}C in the convection zone of a reactor, after which it enters the radiation zone of the furnace where it reaches the reaction temperature, in particular a temperature in the range 700 ^{o}C to 900 ^{o}C.

The mixture is then cooled and compressed for recovering a certain amount of heat and for purifying the waste products.

After that, it follows a complex sequence of separation techniques and different chemical treatments.

2. Numerical control design

A standard automation solution for the process plant should be able to handle two major aspects concerning the proper operation of the process.

On the one hand, it should offer the possibility to maintain a good proportion between the quantities of reactants that are fed to the installation and, on the other

hand, it should provide the means to maintain the temperature and pressure conditions within the limits imposed by the nature of the chemical reactions that should take place inside the reactor [4], [5].

Therefore, we have chosen the corresponding automatic control solution presented in Fig.1 (*(FRC-1) - control loop for gasoline flow; (FRC-2) - control loop for steam flow; (TRC-3) - control loop for temperature; (PRC-4) - control loop for pressure*).

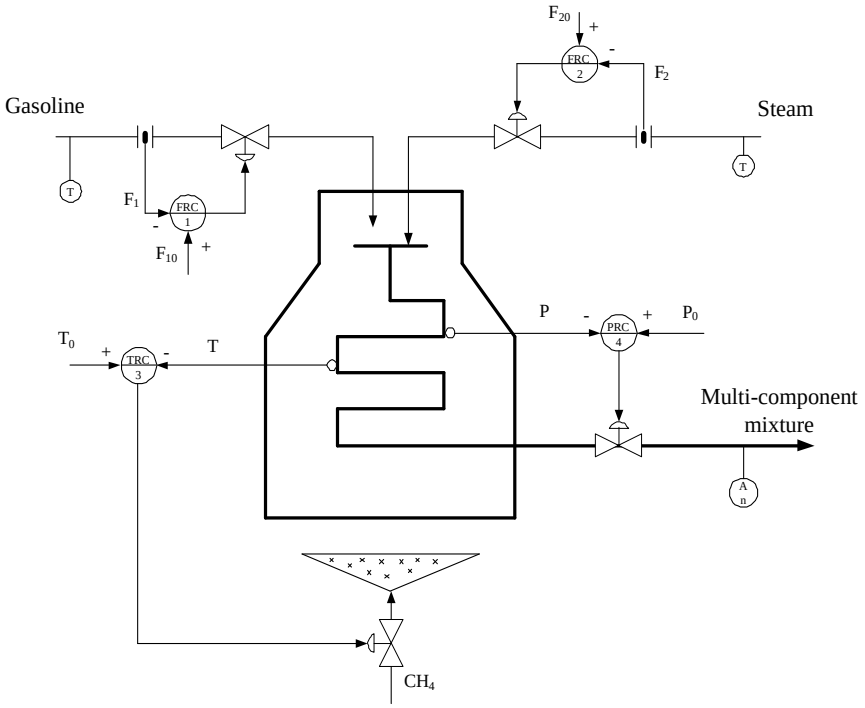

Fig. 1. Automation solution for the pyrolysis reactor.

The first two control loops ensure a good ratio of gasoline flow and steam flow to the ethylene reactor (1500 m^3/h: 500 m^3/h). Besides the gasoline and the steam flow, another controlled technological parameter is the pressure inside the reactor. It is very important that the pressure value is kept within the boundaries of the admissible operating range so that the process will function correctly and the reactor walls will not be submitted to any risk of deterioration.

In this case, the pressure inside the reactor should be of at least 3.3 bars and of at most 4.5 bars.

Finally, the temperature within the median section of the reactor is regulated by the control of the quantity of CH_4 used in the combustion process that heats the plant. The automation solution should be able to provide the possibility of maintaining the temperature values within 820 ^oC and 860 ^oC.

2.1. Nominal Control Systems Solution

The control level has to fulfil two important tasks: the processing of data acquired from the physical plant and the regulation of the major parameters of the process.

The physical data is used to identify the mathematical models, based on which the controls associated to the various regulation systems are computed.

2.1.1. Gasoline and steam flow control

In this subchapter we will briefly present the steps that were followed in order to determine the algorithms suited for the control of the quantities of gasoline and steam that are pumped into the reactor.

First of all, we have analytically determined the mathematical models of the two processes based on a series of technological data provided by the plant's owners.

The mathematical models are:

$$H_{F1}(s) = \frac{0.24}{(8s+1)(0.67s+1)} \tag{1}$$

in case of the gasoline flow process and:

$$H_{F2}(s) = \frac{0.21}{(4s+1)(0.2s+1)} \tag{2}$$

in case of the steam flow process.

The controllers were designed for the two nominated systems by using a PI control algorithm and the Poles Placement Method in order to determine the controller parameters that would ensure the imposed performances for the closed loop systems [6].

The two designed PI controllers are described by the transfer functions:

$$\begin{aligned} H_{R1}(s) &= 11.11\left(1+\frac{1}{8s}\right) \\ H_{R2}(s) &= 6{,}34\left(1+\frac{1}{4s}\right) \end{aligned} \tag{3}$$

In order to validate the performances ensured by these control algorithms, we used Matlab-Simulink simulation environment.

The results are shown in Fig. 2. The disturbances are considered as deterministic variations of the imposed reference values.

It can be noticed that the tracking performances are met and the controller provides the rejection of the step type disturbance signals applied to the output of the system.

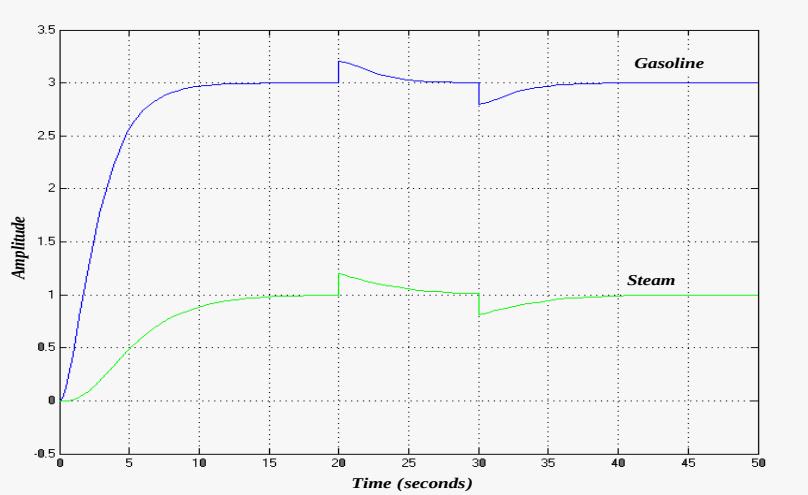

Fig. 2. Step response of the gasoline and steam flow control systems.

In order to be able to implement these control algorithms on a physical processing device, we need to obtain their corresponding numerical representation. Using the Euler approximation method for a sampling period of 0.1 seconds, we obtained the following discrete representations of the two control algorithms, prepared for real time implementation:

$$H_{R1}(z^{-1}) = \frac{11.24 - 11.11z^{-1}}{1 - z^{-1}}$$
$$H_{R2}(z^{-1}) = \frac{7.92 - 6.34z^{-1}}{1 - z^{-1}} \quad (4)$$

2.1.2. Temperature and pressure control

For temperature and pressure control systems we directly designed the numerical versions of the associated controllers following the steps briefly described below.

First of all, the process models were obtained through an experimental identification method using the data acquired from the physical plant, and the common parametric adaptation technique, Recursive Least Squares Method, as [10], [11]. After identification, the models have to be validated using different data sets. For the model validation we used the Whiteness of the Residuals Test, [8], [9].

For the temperature control system, we have identified and validated the model in (5) by using a sampling period of 5 seconds. As the dynamics of the process is slow, we can afford to consider a larger sampling period:

$$H_{P3}(z^{-1}) = \frac{0.04711z^{-1}}{1 - 1.61402z^{-1} + 0.65344z^{-2}} \quad (5)$$

In the case of the pressure control system, the model that was identified and validated for a sampling period of 2 seconds is the following:

$$H_{P4}(z^{-1}) = \frac{0.00597z^{-1}}{1-1.68364z^{-1}+0.70730z^{-2}} \tag{6}$$

We designed two controllers that ensure a series of desired tracking and regulation performances by using an RST control algorithm [6], [9], whose structure is shown in Fig. 3.

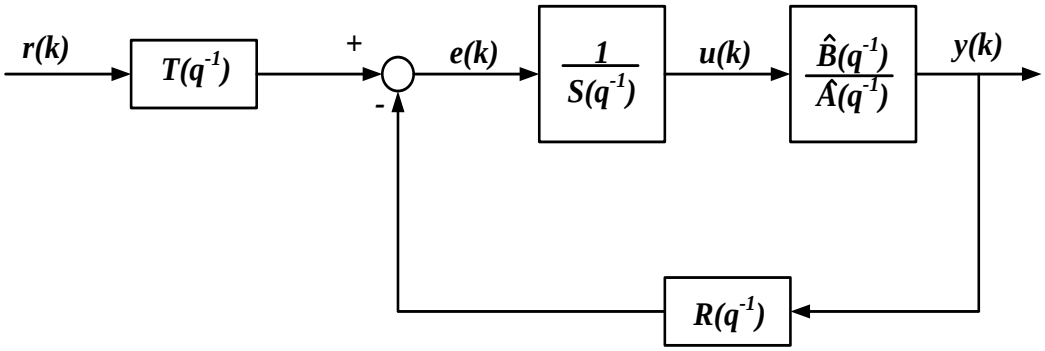

Fig. 3. Polynomial RST control structure.

For the temperature control system, we have obtained the polynomials in (7) and for the pressure control system we have obtained the polynomials in (8).

$$\begin{cases} R_3(z^{-1}) = 36.45 - 52.01z^{-1} + 18.98z^{-2} \\ S_3(z^{-1}) = 1 + 0.36z^{-1} - 1.36z^{-2} \\ T_3(z^{-1}) = 21.22 - 26.42z^{-1} + 8.63z^{-2} \end{cases} \tag{7}$$

$$\begin{cases} R_4(z^{-1}) = 241.9 - 372.9z^{-1} + 145.3z^{-2} \\ S_4(z^{-1}) = 1 + 0.22z^{-1} - 1.22z^{-2} \\ T_4(z^{-1}) = 167.5 - 243.9z^{-1} + 90.8z^{-2} \end{cases} \tag{8}$$

Fig. 4 shows the step response of the closed loop system with the process model given by (5) and the controller described by (7), as well as the evolution of the computed command signal.

Similarly, Fig. 5 shows the step response of the closed loop system with the process model given by (6) and the controller described by (8), as well as the evolution of the computed command signal.

The RST controllers ensure that the desired tracking and regulation performances for the closed loop systems are met.

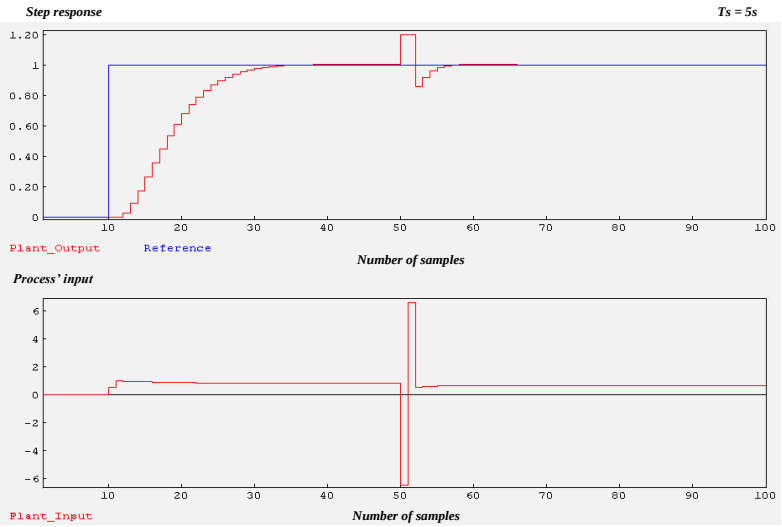

Fig. 4. Closed loop temperature nominal control system step response and process input signal profile.

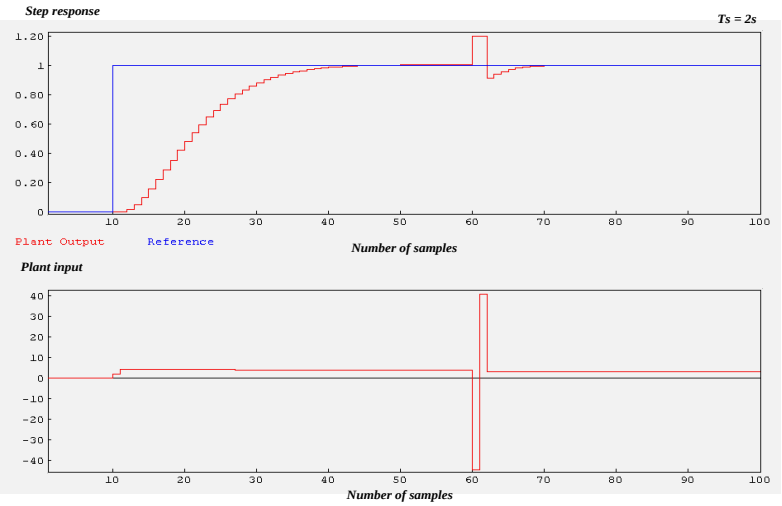

Fig. 5. Step response of the closed loop pressure nominal control system and plant input signal profile.

2.2. Robust control system design

By taking a closer look at Fig. 4 and 5, one can easily see that even the presence of a deterministic disturbance causes high amplitude variations of the control signal within a short period of time. This kind of shocks can lead to the malfunctioning of the actuator and has a strong negative influence as far as the physical process is concerned. Since the control algorithm is implemented on a hardware processing unit, such as a microcontroller and the process model is not identical with the identified one, the performances obtained for the nominal system might not be preserved in the real case.

Thus, our goal is to design robust controllers that can provide a way of preserving the performances obtained in simulation when it comes to the physical system as well as to ensure that the closed loop control system is stable in the presence of disturbances, nonlinearities or model uncertainties.

In order to assess the robustness of a controller, we will use as indicators the output sensitivity function and the modulus margin.

The output sensitivity function represents the influence of the disturbance on the output:

$$S_{py}(\omega) = \frac{\hat{a}(z^{-1})S(z^{-1})}{\hat{a}(z^{-1})S(z^{-1}) + (\hat{b}\,z^{-1})R(z^{-1})} \tag{9}$$

in where $\hat{a}(z^{-1})$ and $\hat{b}(z^{-1})$ are the polynomials which define the identified process model and R(z-1) and S(z-1) are the computed control polynomials.

The modulus margin ΔM represents the minimal distance of the open loop Nyquist plot from the control system to the critical point [*-1, 0j*]. The output sensitivity function and the modulus margin are related through the following formula:

$$\Delta M = \left|S_{py}^{-1}\right|_{\min} = \left|S_{py}\right|_{\max}^{-1} \tag{10}$$

For a good robustness of the system, it is necessary to have a modulus margin greater or equal to 0.5, which implies a maximum of 6 dB for the output sensitivity function, [6]. The output sensitivity function of the previously designed nominal temperature control system does not meet this requirement, as shown in Fig. 6.

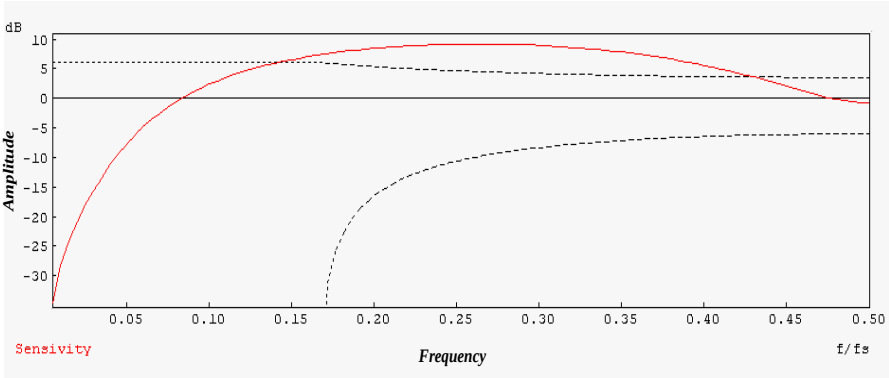

Fig. 6. Frequency/Amplitude representation of the output sensitivity function for the nominal temperature control system.

In order to improve the controller, we have added a pair of complex auxiliary poles to the characteristic polynomial given by the denominator of the sensitivity function in (9). The new control polynomials are:

$$\begin{cases} R_3'(z^{-1}) = 3.42 - 5.53z^{-1} + 2.24z^{-2} \\ S_3'(z^{-1}) = 1 - 1.24z^{-1} + 0.24z^{-2} \\ T_3'(z^{-1}) = 21.22 - 60.68z^{-1} + 65.15z^{-2} - \\ -31.19z^{-3} + 5.63z^{-4} \end{cases} \quad (11)$$

The output sensitivity function of the new closed loop system meets the imposed 6 dB limit (see Fig. 7), which indicates that the new controller is robust.

The tracking and the regulation performances from the nominal case are preserved, but the control signal does not undergo such drastic variations in the presence of a disturbance, as can be seen from Fig. 8.

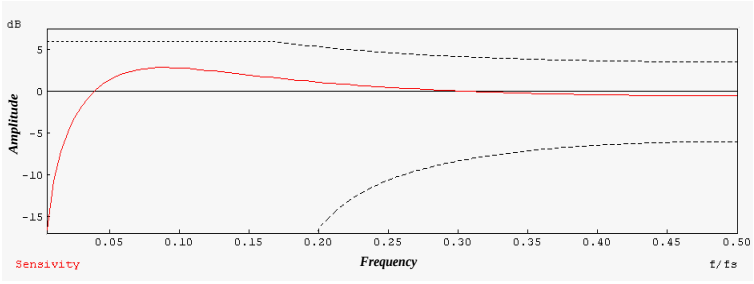

Fig.7. Frequency – amplitude representation of the output sensitivity function for the robust temperature control system.

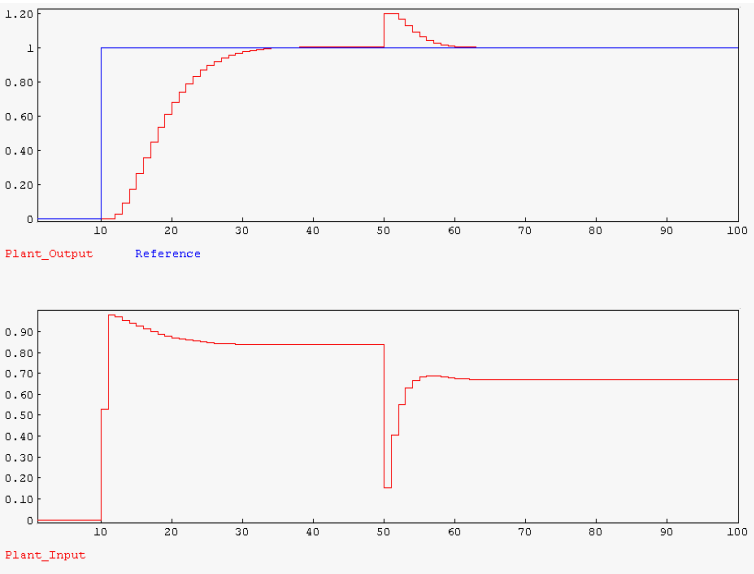

Fig. 8. Step response and process input for the robust temperature control system.

As far as the pressure control system is concerned, we have encountered a similar situation (see Fig. 9). We added two real poles to the characteristic polynomial and we obtained the following control polynomials:

$$\begin{cases} R_4'(z^{-1}) = 14.34 - 24.15z^{-1} + 10.14z^{-2} \\ S_4'(z^{-1}) = 1 - 1.45z^{-1} + 0.45z^{-2} \\ T_4'(z^{-1}) = 167.5 - 526.6z^{-1} + 620.1z^{-2} - \\ -325.5z^{-3} + 64.24z^{-4} \end{cases} \quad (12)$$

The obtained RST controller is a robust one, as can be seen from Fig.10.

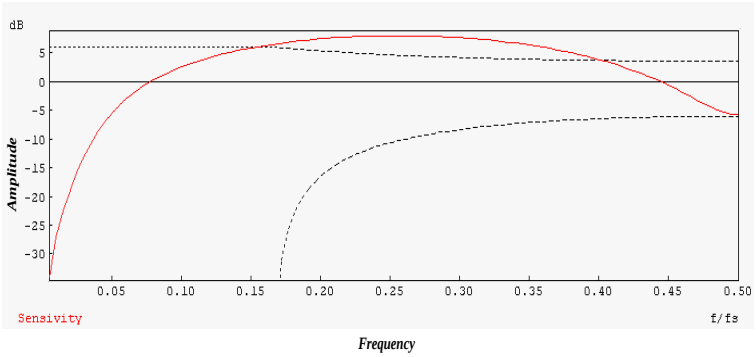

Fig. 9. Frequency/Amplitude of the output sensitivity function for the nominal pressure control.

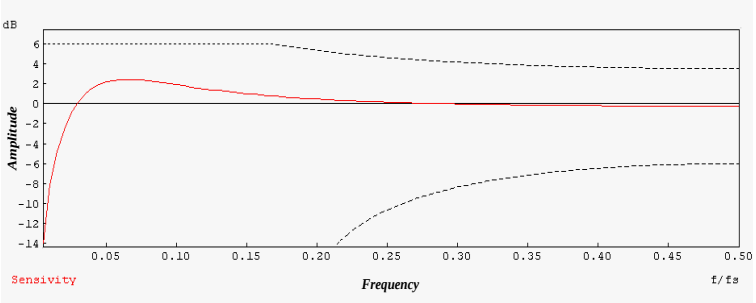

Fig.10. Frequency/Amplitude of the output sensitivity function for the robust pressure control.

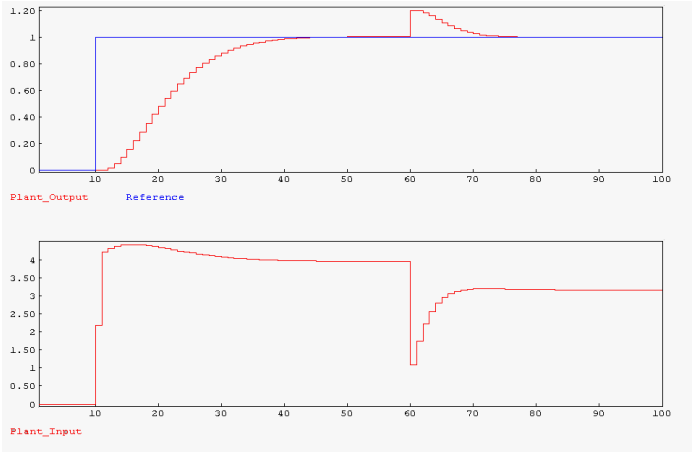

Fig. 11. Step response and process input for the robust pressure control system.

Fig. 11 shows that the tracking and the regulation performances imposed for the nominal system are preserved and that the disturbance does not have such an important influence over the process input signal anymore. After the robustness correction, the shaping of the sensitivity function was clearly improved.

3. Pyrolysis process optimization

The optimization problem of the operating conditions aims to determine the point where the process performances are optimal [12], [13], [14], [15], [16], [17]. In this case study, the optimization objective is to maximize the concentration of ethylene from the plant output (z), by computing the best choice for the set-points of the closed-loop systems. For solving this problem, we use a number of experimental data collected directly from the plant, data representing values for the gas flow ($D_{mp}=y_1$), the steam flow ($D_{ab}=y_2$), the reaction pressure ($P=y_3$) and the reactor temperature ($T=y_4$), and ethylene concentration (C_2H_4=z), Table 1.

Table 1. Experimental data

	D_{mp} [m^3/h]	D_{ab} [m^3/h]	P [atm]	T [degC]	C_2H_4 [%]
1	1120	473	3.50	850	35.8850
2	1270	450	3.80	836	29.4924
3	1330	465	4.20	851	34.1172
4	1363	485	4.30	852	35.7660
5	1363	474	3.38	850	32.6069
6	1387	468	3.40	852	33.2400
7	1330	486	4.00	852	32.3200
8	1406	453	3.40	852	34.1172
9	1320	463	3.80	852	30.6440
10	1283	465	3.70	852	30.2000
11	1215	475	3.50	851	29.2485
12	1035	486	3.30	839	23.9480
13	1100	474	3.30	835	31.4000
14	1280	442	3.50	850	31.4134
15	1175	444	3.80	840	28.9828
16	1368	468	3.40	842	27.9044
17	1340	450	4.50	858	32.1534
18	1440	482	4.70	852	31.7095
19	1570	540	4.00	850	27.3292
20	1380	465	3.70	853	31.5422
21	1330	540	4.10	828	28.4544
22	1450	480	3.17	825	34.3804
23	1160	515	3.30	832	31.9108
24	1370	432	3.70	846	29.6825
25	1295	475	3.73	848	32.5520
26	1306	476	3.60	849	32.7630
27	1293	479	3.80	846	32.4781
28	1290	473	3.70	848	32.4536
29	1308	473	3.85	847	32.4780
30	1310	470	3.80	847	32.5051
Mean values	**1306**	**473**	**3.73**	**846**	**31.37**

We estimate a global model in which the quality variable *z* represents the concentration of ethylene in the final multi-component mixture.

By analysing experimental data sets, the structure of the multivariable nonlinear model was proposed:

$$\hat{z} = \hat{a}_0 + \hat{a}_1 y_1 + \hat{a}_2 \frac{1}{y_2} + \hat{a}_3 y_3 + \hat{a}_4 y_4^2 \tag{13}$$

in where $\hat{a}^* = [\hat{a}_0^*, \hat{a}_1^*, \hat{a}_2^*, \hat{a}_3^*, \hat{a}_4^*]$, represents the model parameters estimation, which were determined by using the Least Squares Method (LSM) on a set of experimental data from Table 1.

$$\begin{aligned} \hat{a}_0^* &= 0.138549 \cdot 10^2 \\ \hat{a}_1^* &= 0.573751 \cdot 10^{-1} \\ \hat{a}_2^* &= 0.129318 \\ \hat{a}_3^* &= -0.365534 \cdot 10^{-2} \\ \hat{a}_4^* &= 0.196089 \cdot 10^{-4} \end{aligned} \tag{14}$$

The optimization problem is formulated as:

$$\max\left\{ I = \hat{z}^*(y) \right\} \tag{15}$$

with the technological restrictions:

$$\begin{aligned} &1000 \le y_1 \le 1600 [m^3/h] \\ &430 \le y_2 \le 540 [m^3/h] \\ &3.3 \le y_3 \le 4.5 [barr] \\ &820 \le y_4 \le 860 [^0 C] \end{aligned} \tag{16}$$

The results obtained from a Matlab optimization routine (BOX) follow the limitations imposed, having this form: y^*_1=1599.6 [m^3/h], y^*_2=430.03 [m^3/h], y^*_3=3.3 [*bar*] and y^*_4=859.85 [^{0}C]. The maximum quantity of ethylene produced as part from the total of final products resulted from the pyrolysis reaction is given by the optimal value $I(y^*)$=34.16%.

The optimal decision (y^*_1, y^*_2, y^*_3, y^*_4) from this supervisory level is automatically transferred as set-points to the plant's control systems.

After implementing this result in real time, the concentration of obtained ethylene (34.16%) has been increased with approximately 2.79% compared to the computed mean value (31.37%) and with 3.716m3/h ethylene production respectively.

Conclusions

This paper proposes a numerical solution for the control and optimization of an ethylene reactor on a petrochemical industrial platform.

Our proposed solution is based on a hierarchical control architecture comprised of two inter-connected levels, a control level and an optimization level.

The control level ensures a good ratio between the gasoline and steam flow at optimal reaction conditions, through a set of robust control systems.

At the decision level, an optimal operating point of the reactor is evaluated by maximizing the ethylene concentration and the ethylene production augmented with 3.716 m^3/h (2,79% gain) compared with the previous behavior.

The results have been implemented in real time on an assembly of ethylene reactors at one of the important petrochemical platforms from Romania.

In some situations the reactor will reach an unstable thermal operating point and additional measures must be taken, consisting in the introduction of an innovative strategy that will ensure the closed loop heat-stable operation. This scenario represents a good starting point for further research.

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