Controlling a Complex Propylene-Propane Distillation Column using a Robust Method Suitable for Simple Processes

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In order to reveal the unused potential of a simple and well known control method, like Proportional-Integral-Derivative (PID) algorithm, or other variants derived from this such P, PI or PD, which is suitable for simple processes, was considered the case of controlling the industrial propylene-propane distillation column. Because the distillation process is a nonlinear one, in order to control the top and bottom propylene concentration, the proposed PI algorithm is adapting to the process nonlinearities by changing the tuning parameters depending on the process operating range. The results obtained with the proposed control structure were compared with the ones obtained in case of considering a standard PI algorithm, with nonadaptable tuning parameters to the process operating ranges, observing that in case of adaptable method the process steady-state and dynamic performance are better.

Keywords: propylene-propane distillation, complex process simulation, PID algorithm

The Proportional-Integral-Derivative (PID) is the oldest and widely used feedback control method for different industrial processes. This strategy, in comparison with the advanced control methods, has the big advantage of having the same structure for each particular process.

A PID controller computes an error value as the difference between a measured process output value and a desired output value (setpoint). The controller tries to bring the error close or equal to zero, by adjusting the process control inputs, the control variable value.

The PID algorithm control variable is computed as the sum of three control terms namely: the proportional term (P), which depends on the present error, the integral term (I), which depends on the accumulation of past errors and the derivative term (D) which is a prediction of future errors, based on current rate of change [1].

The earliest concerns regarding the PID controller are known from the beginning of past century, when, in 1911, the first PID controller example was developed by Elmer Sperry [2] and the first theoretical analysis of a PID controller was made and published by Nicolas Minorsky in 1922 [3, 4].

The PID control algorithm is used to control almost all loops in the chemical process industry, and it is also the basis for many advanced control algorithms and strategies.

Like any other standard control algorithm, PID offers good results in case of linear processes or in the case of operating the process in the vicinity of an operating range, but when the process is a nonlinear one, it is wise to use a method in which the tuning parameters are adapting according to the process operating range.

This paper will present a comparison between the results that can be obtained when a variant of the PID algorithm, namely the PI algorithm is used for controlling a propylene-propane distillation column, in case of adaptable tuning parameters method and of a standard, nonadaptable one.

The propylene-propane distillation column

The industrial process used for the study is represented by the distillation of propylene-propane mixture having a feed concentration of 0.67 C_3H_6 mol. fr. and 0.33 C_3H_8 mol. fr.. The considered distillation column is presented in (fig. 1). From this figure can be seen the L-B control structure (Shinskey approach) [5] with the reflux flow used to control the top propylene concentration and the bottom product flow used to control the bottom propylene concentration.

Although were many concerns in controlling this complex process, using the advanced control techniques [6 - 9], the only problem is that a control structure efficiency depends on a high level on the internal control structure efficiency, because the advanced controller gives only the setpoint to the internal classical controller, flow controller for example. If this internal control loop does not have the adequate tuning parameters, then the advanced control structure will not work properly.

As systemic approach, the process (fig. 2) has two outputs (the propylene top and bottom concentrations) and four main inputs, namely two control variables (the reflux and bottom product flows) and two disturbances (the feed flow and feed concentration) [10].

The control variables from (fig. 2) are named also manipulated variables and the output variables, are named controlled variables.

Using data from the industrial propylene-propane distillation column, the process was simulated with ASPEN HYSYS® software, observing that it has a nonlinear behaviour, characterized by different gains and transient times for different process operating ranges and each process channel (fig. 2) [10].

Using the Markov parameters identification procedure [11], models of the process were determined for every process channel (fig. 2) and different operating ranges. This procedure was presented in [10].

A way to express the model of a process [10] is by a second order transfer function with dead time:

$$G = \frac{k_m e^{-\tau s}}{T_2 s^2 + T_1 s + 1}, \qquad (1)$$

where:

k_m is the process gain;

τ - the dead time;

T_2 and T_1 are time constants.

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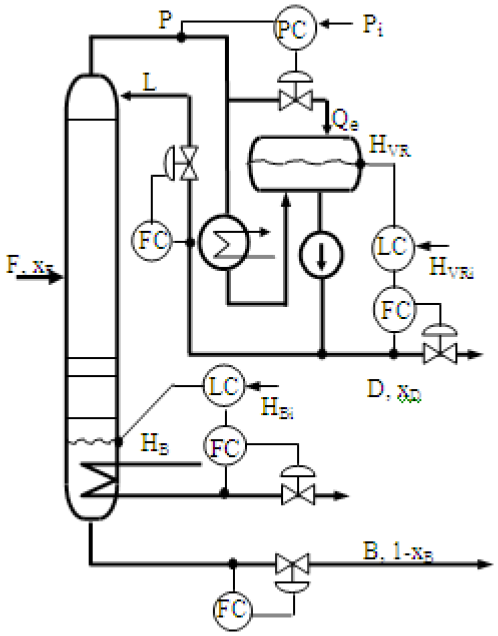

Fig. 1. Propylene-propane distillation column:
PC – pressure controller, FC – flow controller,
LC – level controller, L – reflux flow, P - pressure,
Pi – pressure setpoint, B – bottom product flow,
H_{VR} – reflux tank level, H_{VRi} – reflux tank level setpoint,
H_B – bottom column level, H_{Bi} – bottom column level setpoint,
F – feed flow, x_F – feed C_3H_6 concentration, x_B – bottom C_3H_6 concentration, x_D – top C_3H_6 concentration

For L-x_D and B-(1-x_B) process channel (fig. 2) the values of the model parameters k_m, T_2 and T_1, determined using the procedure from [10], are presented, for different operating ranges, in table 1 and 2.

Table 1
MODEL PARAMETERS FROM TRANSFER FUNCTION (1) FOR CHANNEL L –x_D VALID FOR DIFFERENT PROCESS OPERATING RANGES

x_D* [mol.fr.]	k_m	T_2[min^2]	T_1[min]
0.8-0.85	1.09	0.69	30.58
0.85-0.88	0. 85	0.61	28.10
0.88-0.9	0. 65	0.58	27.00
0.9-0.91	0. 58	0.56	26.45
0.91-0.92	0. 48	0.54	25.68
0.92-0.93	0. 42	0.53	25.37
0.93-0.95	0. 36	0.51	24.78
0.95-0.98	0. 30	0.48	23.72
0.98-1.00	0. 28	0.44	22.55

* x_D is the propylene molar fraction in the top of the column

Table 2
MODEL PARAMETERS FROM TRANSFER FUNCTION (1) FOR CHANNEL B –(1-x_B) VALID FOR DIFFERENT PROCESS OPERATING RANGES

1-x_B* [mol.fr.]	k_m[mol.fr./%]	T_2[min^2]	T_1[min]
0.79-0.81	-0.69	0.56	50.50
0.81-0.83	-0.74	1.33	74.00
0.83-0.9	-0.78	1.80	84.33
0.9-0.92	-0.98	3.96	144.00
0.92-0.94	-1.04	4.56	160.43
0.94-0.96	-1.10	5.17	177.10
0.96-0.98	-1.14	6.00	181.98
0.98-1.00	-1.20	6.38	210.18

*x_B is the propylene molar fraction and 1-x_B is the propane molar fraction in the bottom of the column

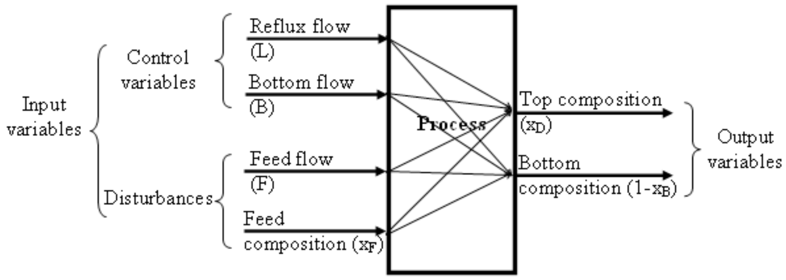

Fig. 2. Propylene-propane distillation process block diagram

Accordingly, the process model will be adapted, depending of the operating range by choosing the adequate model parameters.

The distillation column behaviour, characterized by the transfer function (1) with the model parameters from the table 1 and 2 was simulated, for this study, using Matlab®.

PID control algorithm

The PID control law has the expression [12]:

$$c(t) = c_0(t) + k_R \cdot (e(t) + \frac{1}{T_i}\int_0^t e(t) + T_d \frac{de}{dt}), \qquad (2)$$

where:

$c_0(t)$ is the initial control value;
$e(t)$ - error value;
k_R - controller gain;
T_i - integral time;
T_d is the derivative time.

The PID controller transfer function has the expression

$$G_{PID}(s) = k_R \cdot (1 + \frac{1}{T_i \cdot s} + T_d \cdot s). \qquad (3)$$

Each term from the PID algorithm has its own property [12]:

-the proportional term speeds up the control system response, minimizes but does not eliminate the steady state error;

-the integral term eliminates the steady state error and introduces oscillations in the system response;

- the derivative term is used for an excess of the control action; it has to be used in case of processes with high inertia because can lead to instability problems.

By choosing adequate values for the controller gain (k_R), integral time (T_i) and derivative time (T_d) constants, the PID algorithm can be reduced only to the proportional term (P), having $T_i=\infty$ and $T_d=0$, the proportional-integral term (PI), having $T_d=0$, or proportional-derivative term (PD), having $T_i=\infty$.

In the case of the considered process are used only the proportional and integral terms, leading to a PI algorithm, because the derivative term it is not necessary due to the process characteristic.

The PID controller tuning parameters (k_R, T_i and T_d) can be found using some formulas based on the tuning parameters values which bring the process to its limit of stability or using formulas based on the process model parameters (gain, time constants, dead time) [1].

In case of using the second method, the tests can be made online, in closed loop, with the control loop running. Some of these tuning methods are Ziegler-Nichols, Seborg, Offereins [13].

In case of using the process model identification methods, the model parameters are computed from the process step response and then, using some formulae, depending on method, the PID tuning parameters can be found. Some of these methods are Ziegler-Nichols, Oppelt, Cohen-Coon, Cirtoaje [12, 14 - 16].

In this paper was used a PI tuning parameter method based on the process model parameter values, because these values were already computed, see table 1 and 2.

Generally, if the process step response is

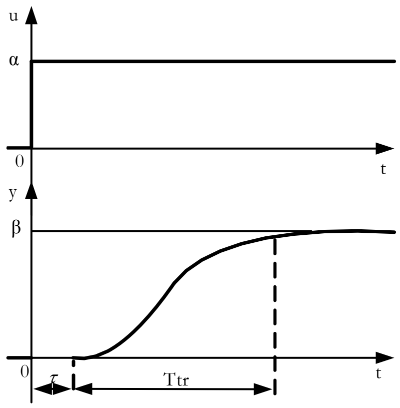

Fig. 3. Process step response: u – the system input variable which has the steady-state value α, y – the system output variable which has the steady-state value β, T_{tr} – the system transient time and τ – the dead time

then, the controller tuning parameters can be found using the following relations [12]:

$$k_R = \frac{0.9}{k_P \cdot (1 + \frac{6 \cdot \tau}{T_{tr}})}, \tag{4}$$

$$T_i = \frac{T_{tr}}{3.5}. \tag{5}$$

In case of using also the derivative term, the derivative time constant value is chosen experimentally, so that we have the best dynamic and steady-state process response.

Results and discussions

The PI control method was implemented for controlling the C_3H_6 top and bottom concentrations of the industrial propylene-propane distillation column.

Because the process is a nonlinear one, but was linearized for different operating ranges, see relation (1) and tables 1 and 2, we have to use a PI algorithm with tuning parameters adaptable to the process operating range.

Table 3
TUNING PARAMETERS VALUES FOR THE PI TOP CONCENTRATION CONTROLLER, WITH THE PROCESS OPERATING RANGE

x_D[mol.fr.]	k_R	T_i[min]
0.8-0.85	0.52	26
0.85-0.88	0.65	24
0.88-0.9	0.83	23
0.9-0.91	0.92	23
0.91-0.92	1.10	22
0.92-0.93	1.25	21
0.93-0.95	1.45	21
0.95-0.98	1.70	20
0.98-1.00	1.78	19

Table 4
TUNING PARAMETERS VALUES FOR THE PI BOTTOM CONCENTRATION CONTROLLER, WITH THE PROCESS OPERATING RANGE

$1-x_B$[mol.fr.]	k_R	T_i[min]
0.79-0.81	-0.96	43
0.81-0.83	-0.98	63
0.83-0.9	-0.95	72
0.9-0.92	-0.82	123
0.92-0.94	-0.77	138
0.94-0.96	-0.74	152
0.96-0.98	-0.71	156
0.98-1.00	-0.69	180

The PI controller has as input the system error - the difference between the controlled output value and the process output desired value (setpoint).

Using this error, and the tuning parameters (k_R, T_i and T_d), based on the PI algorithm (2) a control variable (c) is computed.

For the distillation column, using relations (4) and (5) and process model parameters values from table 1 and 2, the PI controller tuning parameters have the values from tables 3 and 4. According to the experimental results, the dead time value is 3 min for both process channels.

The proposed control system structure is presented in (fig. 4).

The proposed control system is running closely to a classical one, with nonadaptable tuning parameters to the operating range, the only difference appears when the process operating range changes because other tuning parameters are selected from a set of already computed (see tables 3 and 4), and updated in the controller.

For example, in case of changing the top concentration controller setpoint from 0.89 to 0.92 C_3H_6 mol. fr., the system automatically senses when the operating range changes and loads the adequate tuning parameters for the controller, from table 3: for the process operating range from 0.89 to 0.9 C_3H_6 mol. fr., the tuning parameters are: k_R=0.83, T_i=23 min, for the process operating range from

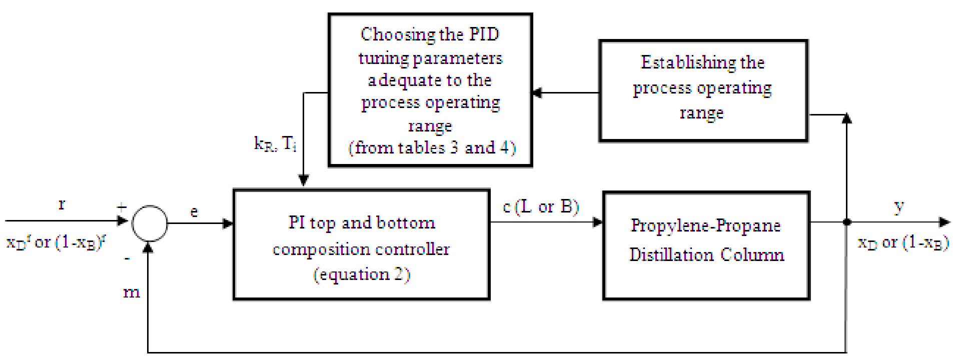

Fig. 4. The proposed PI control structure with adaptable tuning parameters: r – setpoint (for top (x_D^r) or bottom $(1-x_B)^r$ concentration), e – error, c – control variable (reflux (L) or bottom (B) flows), y – system output (top (x_D) or bottom ($1-x_B$) concentration current value), k_R – PID controller gain, T_i – PID controller integral time

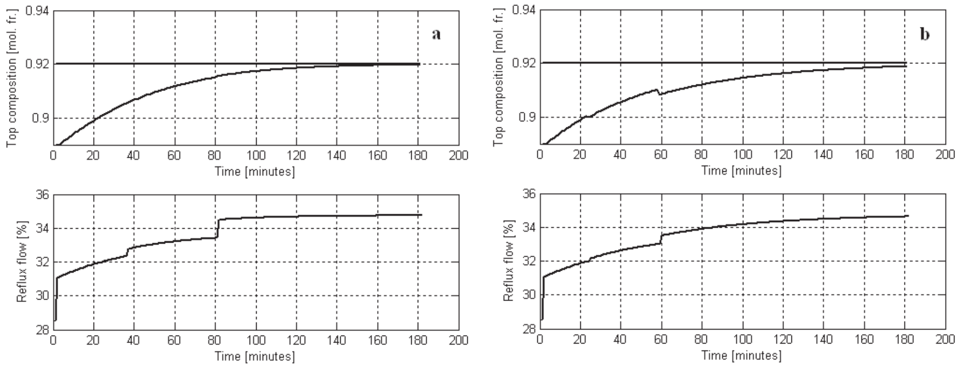

Fig. 5. Top concentration trend when the controller setpoint increases from 0.89 to 0.92 C_3H_6 mol. fr., using a PI controller:
a - with tuning parameters adaptable to the process operating range (k_R=0.83, T_i=23 min, k_R=0.92, T_i=23 min, k_R=1.1, T_i=22 min),
b - with nonadaptable tuning parameters (k_R=0.83 T_i=23 min)

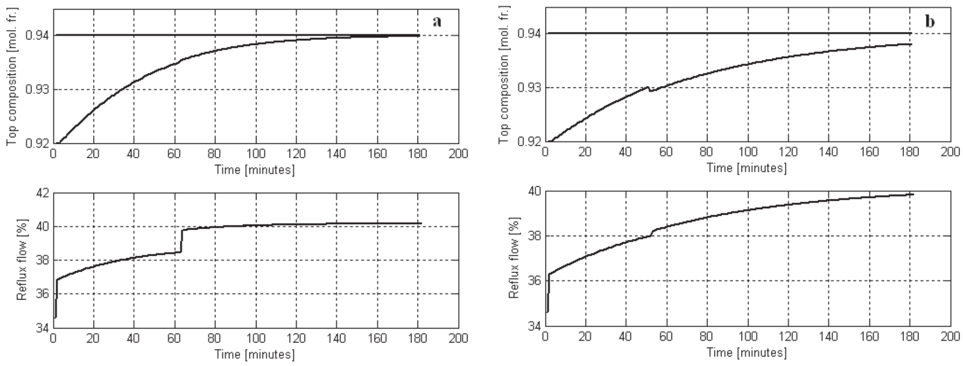

Fig. 6. Top concentration trend when the controller setpoint increases from 0.92 to 0.94 C_3H_6 mol. fr., using a PI controller:
a - with tuning parameters adaptable to the process operating range (k_R=1.1, T_i=22 min, k_R=1.25, T_i=21 min, k_R=1.45, T_i=21 min),
b - with nonadaptable tuning parameters (k_R=0.83, T_i=23 min)

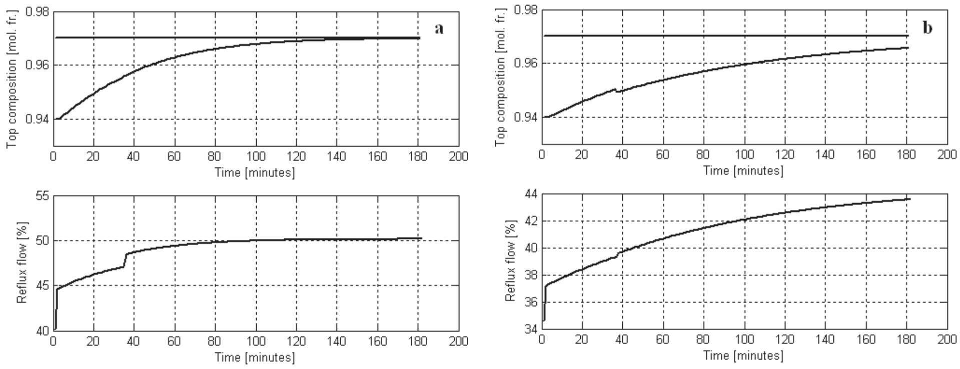

Fig. 7. Top concentration trend when the controller setpoint increases from 0.94 to 0.97 C_3H_6 mol. fr., using a PI controller: **a** - with tuning parameters adaptable to the process operating range (k_R=1.45, T_i=21 min, k_R=1.7, T_i=20 min), **b** - with nonadaptable tuning parameters (k_R=0.83, T_i=23 min)

0.9 to 0.91 C_3H_6 mol. fr.: k_R=0.92, T_i=23 min and for the process operating range from 0.91 to 0.92 C_3H_6 mol. fr.: k_R=1.1, T_i=22 min.

If the control system is a classical one, having the controller with nonadaptable tuning parameters, no matter what is the process operating point the tuning parameters values will be k_R=0.83 T_i=23 min.

This is the case of a test presented in (fig. 5).

For the propylene-propane distillation process presented above, was used the PI controller without the derivative term. The control structure was also simulated in Matlab®.

Further are presented the results obtained using the proposed PI algorithm for controlling the top and bottom concentrations of the distillation column.

The tests consisted of changing the setpoint values for the top and bottom concentration controllers in case of using the PI algorithm with adaptable tuning parameters to the process operating range and in case of using the nonadaptable method, with the tuning parameters from the first considered operating range.

As we can observe from (fig. 5a), the top concentration becomes equal with its setpoint with good dynamic performance; the control variable (reflux flow) value modifies when the process operating range is changing, because another tuning parameters are computed so that the process nonlinearities to be compensated.

In case of using a standard PI controller with nonadaptable tuning parameters (fig. 5b), the control system transient time increases, in comparison with the trend from (fig. 5a) leading to worse dynamic performance. Also, we can observe that the process nonlinearities are not compensated.

In this case, if the PI controller tuning parameters are adapting to the process operating range (fig. 6a), the steady-state error is zero and the dynamic performance is good (no output overshoots and a small transient time).

If the PI tuning parameters do not adapt to the process operating range (fig. 6b), and are the ones from the first operating range, considered in these simulations, adequate for x_D= 0.88-0.9 C_3H_6 mol. fr., the control system transient time increases substantially, in comparison with the previous case (fig. 6a). The process nonlinearities remain uncompensated and we have steady-state error.

If the PI controller tuning parameters are adapting to the process operating range (fig. 7a), the process nonlinearities are compensated and the control system has good steady-state and dynamic performance.

If the PI controller tuning parameters are the ones from the first process operating range (fig. 7b), considered from these tests, adequate for x_D= 0.88-0.9 C_3H_6 mol. fr., the

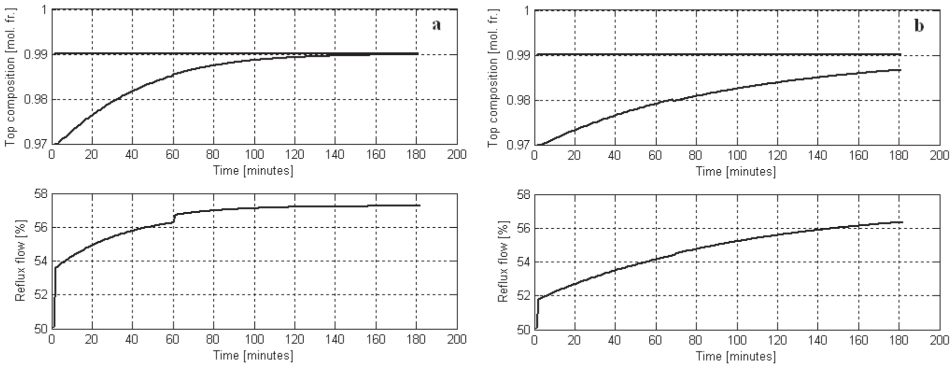

Fig. 8. Top concentration trend when the controller setpoint increases from 0.97 to 0.99 C_3H_6 mol. fr., using a PI controller: **a** - with tuning parameters adaptable to the process operating range (k_R=1.7, T_i=20 min, k_R=1.78, T_i=19 min), **b** - with nonadaptable tuning parameters (k_R=0.83, T_i=23 min)

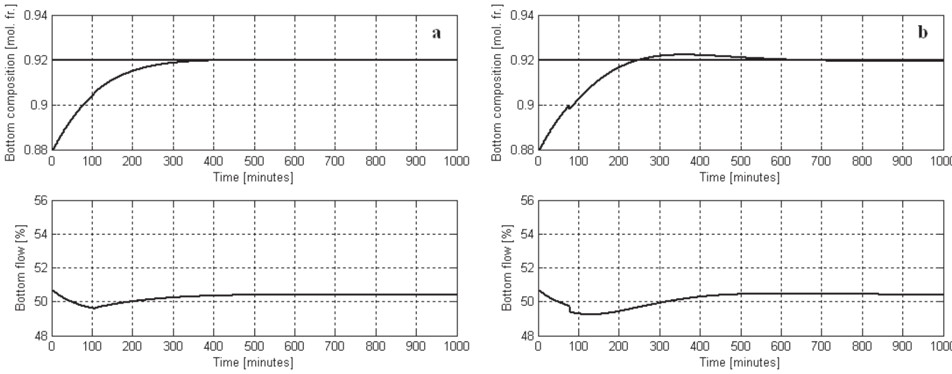

Fig. 9. Bottom concentration trend when the controller setpoint increases from 0.88 to 0.92 C_3H_8 mol. fr., using a PI controller: **a** - with tuning parameters adaptable to the process operating range (k_R=-0.95, T_i=72 min, k_R=-0.82, T_i=123 min), **b** - with nonadaptable tuning parameters (k_R=-0.95, T_i=72 min)

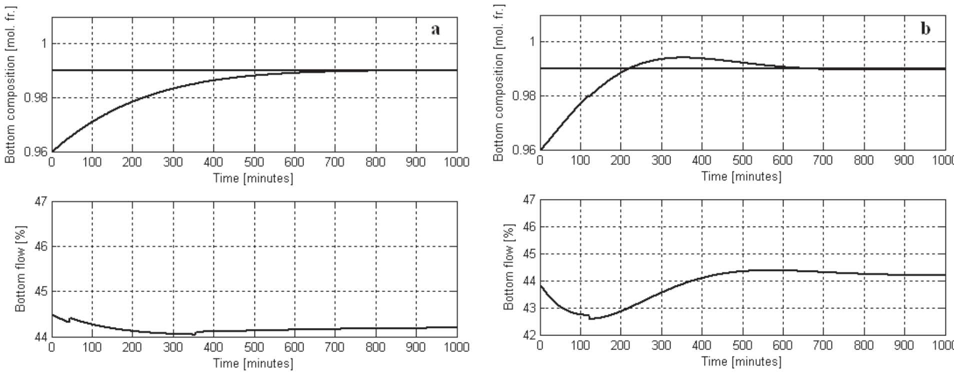

Fig. 10. Bottom concentration trend when the controller setpoint increases from 0.96 to 0.99 C_3H_8 mol. fr., using a PI controller: **a** - with tuning parameters adaptable to the process operating range (k_R=-0.74, T_i=152 min, k_R=-0.71, T_i=156 min, k_R=-0.69, T_i=180 min), **b** - with nonadaptable tuning parameters (k_R=-0.95, T_i=72 min)

control system transient time increases and the steady-state error differs from zero, in comparison with the results from (fig. 7a).

In case of using the PI controller with adaptable tuning parameters (fig. 8a), the control action is done with good steady-state and dynamic performance.

If the PI controller tuning parameters are the ones from the first process operating range (fig. 8b), considered for these tests, adequate for x_D= 0.88-0.9 C_3H_6 mol. fr., the control system transient time increases substantially and the steady-state error differs from zero, in comparison with the results from (fig. 8a).

As we can observe from (fig. 9a), the bottom concentration becomes equal to its setpoint with good dynamic performance; because other tuning parameters are computed at each process operating range change the process nonlinearities are compensated.

When a PI controller with nonadaptable tuning parameters is used (fig. 9b), the process output has overshoot and the control system transient time increases.

If the PI controller tuning parameters are adapting to the process operating range (fig. 10a), the bottom concentration becomes equal with its setpoint value without overshoot.

If the PI controller tuning parameters are the ones from the first process operating range (fig. 10b), considered for these tests, adequate for 1-x_B= 0.83-0.9 C_3H_8 mol. fr., the control system transient time increases and the controlled variable (bottom propane concentration) has overshoot.

Conclusions

A simple Proportional-Integral (PI) controller, which is a variant of the Proportional-Integral-Derivative (PID) controller without the derivative term was considered in order to control the propylene concentration on the top and bottom of a propylene-propane distillation column.

Two cases were considered: the first when the PI controller tuning parameters are adapting to the process operating range and the second when the PI tuning parameters are not adapting to the process operating range. From the simulation results there was observed that the first proposed control structure offers good steady-state and dynamic performance because there is not a steady-state error, there is no overshoot and the process transient time is small. In the second case, there is a steady state error (between 5-20%), there is overshoot and the process transient time increases.

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