Distillation Column Hierarchical Control

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This paper treats the problem of hierarchical control of a butylene-butane distillation column (BBDC). A hierarchical control system with three levels is proposed for BBDC. First hierarchical level, which is represented by the column with its conventional automation, is simulated and the results led to the idea of using a decoupler for the process crossed channels. Thus, it is proposed a nonlinear decoupler which determines automatically its parameters and type. At the second hierarchical level are implemented two internal model controllers for the top and bottom concentrations. Finally, the third hierarchical level consists of an optimal control system for the butylene-butane separation process.

Keywords: distillation column, nonlinear model, hierarchical control, decoupling, internal model control, optimal control

Distillation is one of the most important processes for separating large multicomponent streams into high purity products. These processes are large consumers of energy with about 3% of the total energy consumption in the world [1]. This amount of energy is introduced in the bottom of the column and approximately the same amount is removed in the top, but at significantly lower temperature, which makes this process one of the most effective for the separation of mixtures [2].

Operation methods of distillation columns directly affect the product quality, production rates and utility usage. Hence suitable control of the distillation tower is very important from an economic viewpoint [3]. One of the most important objective for a distillation column is to improve performance by minimizing costs to obtain a product at a specified quality [4], which is an optimization problem that usually refers to the activation of the purity constraint for the most valuable product. This means that as much as possible of the valuable product should be produced, or in other words, the selling of this product at low prices should be avoided [5].

Distillation processes complexity given by their dimensions, the multiple objectives which they have to satisfy, nonlinearity, restrictions etc., implies difficulties in modeling and controlling these processes. In most cases a hierarchical approach can be the solution for this problem. Hierarchical control permits a decomposition of the complex systems control problem into a series of subproblems with lower complexity, which are easier to solve and can be organized as a hierarchical structure [6].

The distillation process analyzed in this paper is the butylene-butane separation process whose purpose is to separate a mixture composed of isobutane, isobutylene, n-butane, cis- and trans-butylene in two products consisting mainly of isobutane and isobutylene at the top and mainly of the other three components at the bottom of the column in which the process takes place.

The paper proposes for the butylene-butane distillation column a hierarchical control system with three levels: first level is associated with conventional automation, second level is the advanced control level, and the third level is dedicated to optimal control. In the following will be presented the characteristics of each hierarchical level.

First hierarchical level

According to [7], the most suitable control configuration for the butylene-butane distillation column is *LB* structure (presented in fig. 1) which represents the first level of the hierarchical control system. This structure assumes that the top pressure and levels in the reflux drum and in the column bottom are controlled. What remains is a composition control problem with two variables, namely the concentrations x_D (at the top) and x_B (at the bottom) for whose control are used the reflux flowrate (L) and bottom product flowrate (B) respectively.

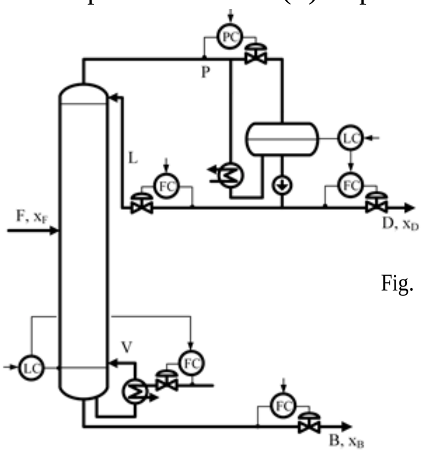

Fig. 1. LB control structure

The column was simulated in SIMULINK®, the mathematical model for the column incorporated in the simulator being a nonlinear one [8, 9] based on equations of liquid-vapor equilibrium and total and component material balance, similar to the ones described in the following.

For the tray K of the column, the material balance can be written

$$\frac{dM_K}{dt} = LF_K + L_{K+1} - L_K + VF_K + V_{K-1} - V_K \, . \quad (1)$$

A component material balance for tray K is given by relation

$$\frac{d}{dt}(M_K x_K) = LF_K \cdot x_{F\_K} + L_{K+1} \cdot x_{K+1} -$$

$$- L_K \cdot x_K + VF_K \cdot y_{F\_K} + V_{K-1} \cdot y_{K-1} - V_K \cdot y_K \, . \quad (2)$$

The liquid-vapor equilibrium is given by Fenske's equation for binary mixtures with constant relative volatility

$$y_K = \frac{\alpha \cdot x_K}{1 + (\alpha - 1) \cdot x_K} \, . \quad (3)$$

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Using the relation

$$\frac{d}{dt}(M_K x_K) = M_K \frac{dx_K}{dt} + x_K \frac{dM_K}{dt}, \quad (4)$$

and taking into account relations (1) and (2) the following equation is obtained

$$M_K \frac{dx_K}{dt} = LF_K(x_{F\_K} - x_K) + L_{K+1}(x_{K+1} - x_K) + VF_K(y_{F\_K} - x_K) + \\ + V_{K-1}(y_{K-1} - x_K) - V_K \cdot (y_K - x_K). \quad (5)$$

Relation (5) is the most used differential equation associated to mass transfer and is used to calculate the concentration x_K.

In order to utilize the above mentioned model, the column was considered a pseudo-binary one with x_D the concentration of the mixture isobutane + isobutylene in the top light product and x_B the concentration of the same mixture in the bottom heavy product. The simulator was configured and validated based on industrial data (flowrates, levels, compositions, design data of the column etc.) obtained from a refinery. The initial data used for the simulation of this column is presented in table 1.

Table 1
INITIAL DATA FOR BBDC SIMULATOR

Parameter	Value
NT	82
NF	52
F [kmole/min]	4.34
x_F [mole fr.]	0.6761
q_F	1
L [kmole/min]	16.65
V [kmole/min]	19.60
M0_K [kmole]	1.7
M0_1 [kmole]	55
M0_NT [kmole]	90
α	1.24

The simulation of this column consisted in most part in changes of the control agents L and B and recordings of the top and bottom concentrations (x_D and x_B) time evolutions. Figure 2 shows the evolution of x_B to a change in reflux flowrate and figure 3 presents the evolution of x_D to a change in bottom product flowrate.

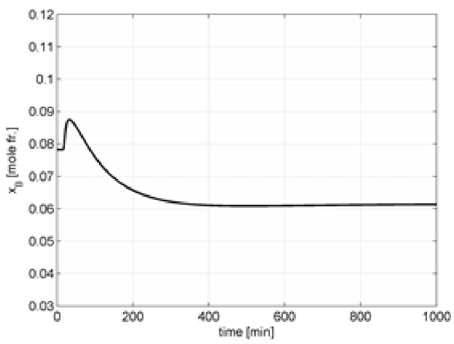

Fig. 2. Time evolution of concentration x_B to a 3% change of L

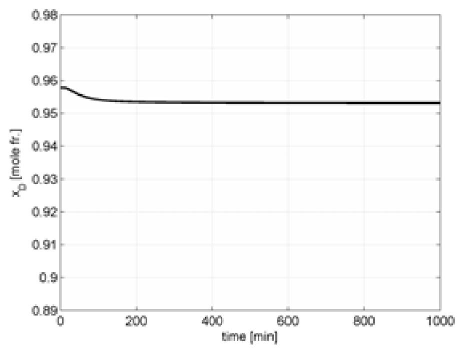

Fig. 3. Time evolution of concentration x_D to a -3% change of B

The most important aspect of the results from figures 2 and 3 is that the process presents interactions on the crossed channels, namely the concentration x_B is influenced by control agent L and also control agent B has an effect on concentration x_D. Consequently, before the implementation of the second hierarchical level, a decoupler is designed in order to reduce or eliminate these interactions.

Control loops decoupling for BBDC

After closing pressure and levels control loops (fig. 1) the butylene-butane distillation column becomes a 2x2 multivariable system (with two control agents-L and B - available for control of the two concentrations - x_D and x_B). As stated above, a decoupler must be designed to diminish the influence of L on x_B and the influence of B on x_D.

The system consisting of the column and the decoupler is presented in figure 4 where it is considered that control signal c_1 has an effect only on x_D and control signal c_2 influences only x_B.

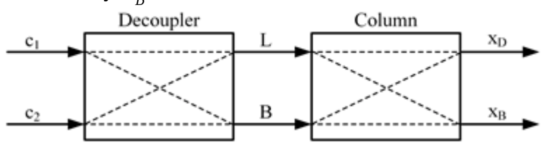

Fig. 4. Decoupling scheme

The proposed method of decoupling is based on a decoupler with standard structure implementable in four versions depending on process dynamic characteristics. The decoupler has two inputs and two outputs with unit gains on direct channels and at least two static channels out of four.

The general form of the decoupler [10] is

$$G_D(s) = \begin{bmatrix} \frac{1}{T_{d\_11}s+1} & \frac{K_{d\_12}}{T_{d\_12}s+1} \\ \frac{K_{d\_21}}{T_{d\_21}s+1} & \frac{1}{T_{d\_22}s+1} \end{bmatrix}. \quad (6)$$

Depending on the input-output channels dynamics there are four types of the decoupler [11] obtained from the general form (6) by analyzing the dynamic of the direct channels in relation to the crossed channels.

The decoupler gains (K_d) are obtained from the steady-state decoupling condition as ratios of corresponding process gains, and the decoupler time constants (T_d) are calculated as differences between process time constants on the appropriate channels for each of the four decoupler types.

A first step in the design of the decoupler for the butylene-butane distillation column was the identification of the process gains (K_p) and transient times (T_{tr}) at step variations of the control agents L and B. Using relation $T_p = T_{tr} / 4$ are obtained the process time constants (T_p) necessary for the decoupler time constants calculation.

A selection of the variations of the process gains K_p and time constants T_p on different channels according to control agents (L and B) variations is shown in figures 5-6.

This phase of the decoupler design consisted in approximation of the process gains and time constants dependencies on control agents' variation by determining the degrees of polynomial regression functions so that the best possible approximation is achieved, as it can be seen in figures 5-6. Using these regression functions, process gains and time constants can be calculated for variations of L and B other than the considered ones [12].

Using a function developed in MATLAB® by the author, the values of K_p and T_p are calculated automatically

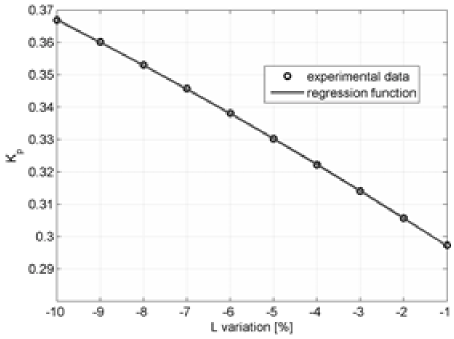

Fig. 5. Variation of K_p depending on L variation on L - x_D channel

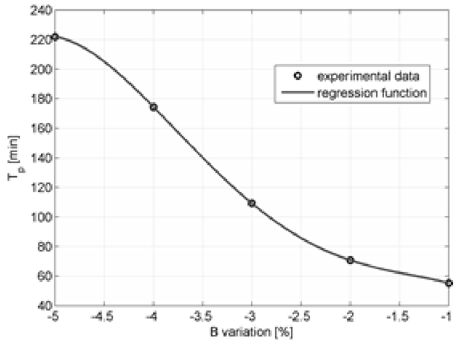

Fig. 6. Variation of T_p depending on B variation on B - x_D channel

depending on control signals variations, by interpolation (when c_1 and c_{-2} are within the initial considered ranges) or using the regression functions (when c_1 and c_{-2} are outside the initial considered ranges). Also, this function automatically determines the decoupler gains (K_d) and time constants (T_d) and its type according to process input-output channels dynamics. Depending on the decoupler type, four decoupling schemes are implemented. Each of this scheme is chosen automatically and uses the corresponding values of the parameters K_d and T_d [12].

The system composed of the decoupler and the butylene-butane distillation column was simulated in SIMULINK® to verify the effects of the decoupling on crossed channels L- x_B and B -x_D.

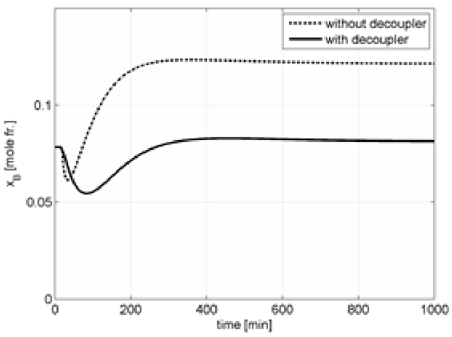

Fig. 7. x_B time evolution to a -6% change of c_1

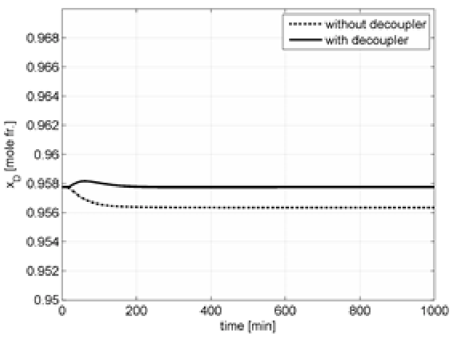

Fig. 8. x_D time evolution to a -1% change of c_2

Figures 7 and 8 show the time evolution of concentration x_D to changes of the second control signal and time evolution of concentration x_B to changes of the first control signal, with and without decoupler, where can be observed the important effect of the decoupler on process crossed channels. The decoupling offers the possibility of treating the concentration multivariable control system as two independent monovariable control systems, implemented at the second level of the hierarchical system.

Second hierarchical level

The hierarchical control system with two levels associated to BBDC is illustrated in figure 9.

At the second hierarchical level (ACS – Advanced Control System) are implemented two feedback internal model controllers having as inputs the current values of the two concentrations and their set-points, and as outputs the set-points for the reflux flowrate and bottom product flowrate control systems. In this representation the decoupler proposed in section 3 is included in ACS.

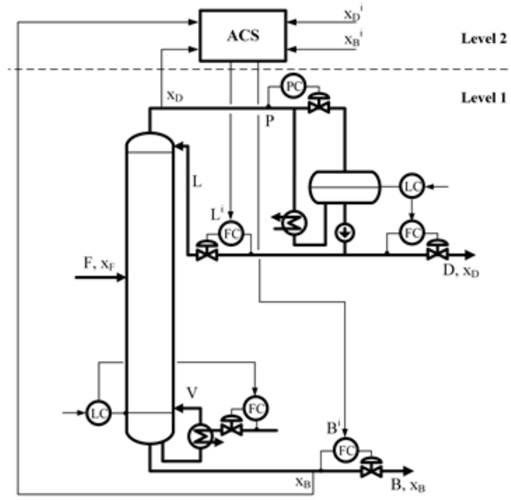

Fig. 9. Hierarchical control system with 2 levels

The block diagram of the hierarchical control system with two levels is shown in figure 10.

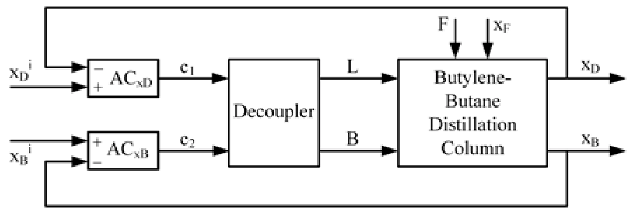

Fig. 10. Block diagram of the concentration control system

As stated before the two controllers from figure 10 are based on internal model control (IMC) method which will be presented briefly in the following.

The main motivation of internal model control is based on the principle that good control involves the inclusion in the control system of a representation (implicit or explicit) of the process to be controlled [13].

The standard internal model control method used in this research has as main characteristic the fact that at step set-point the control signal has also a step form [14]. For the system to be tunable, in the controller structure was included a proportional element with gain K_C whose standard value is 1.

A stable and overdamped process, characterized by gain K_p, transient time T_{tr} and dead time T_m, can have associated the model [14]

$$G_M(s) = \frac{K_M e^{-T_m s}}{(T_M s + 1)^2}. \qquad (7)$$

In practical applications model gain K_M can be considered equal to process gain K_p, and the model time constant T_M H≈T_{tr}/6 [14].

The standard internal model controller has four main parameters, namely: three process parameters (K_p, T_{tr}, and T_m) and a tuning parameter (K_C).

The above described method can be used for processes with overdamped response. As it can be seen from figure 11 the process response on the direct channel c_1 - x_D is underdamped with overshoot.

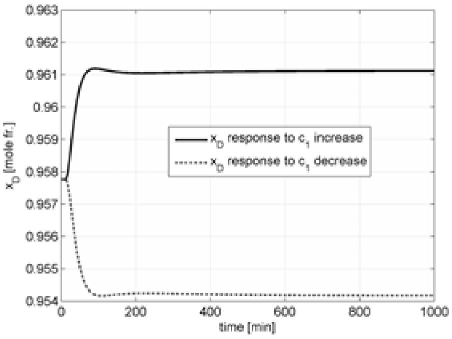

Fig. 11. Time evolution of x_D to c_1 change

In order to obtain an overdamped response on this channel in series with the process a filter will be used [14]

$$G_f(s) = \frac{1}{T_f s + 1}. \qquad (8)$$

The filter time constant T_f is determined so that the process response to be overdamped (without overshoot).

As a result of identification of the compensated process (filter + process) on channel $c_1 - x_D$ the following values for the model parameters were obtained: K_{M1} = 0.3, T_{M1} = 43 min and T_m = 10 min. These values will be used as starting parameters in the tuning of the controller associated to top concentration.

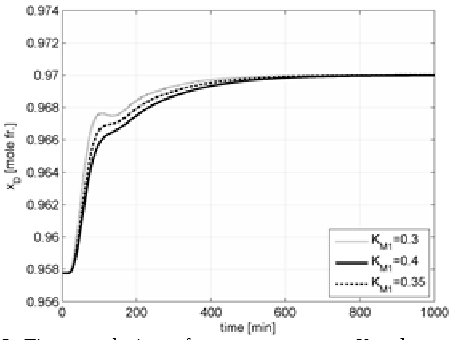

Fig. 12. Time evolution of x_D to parameter K_{M1} changes

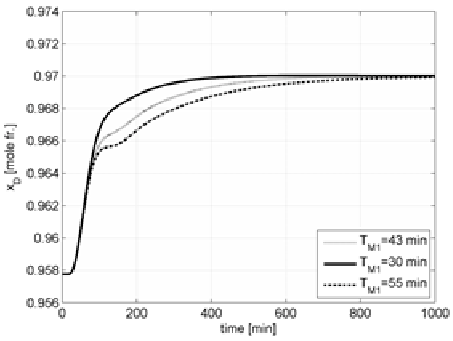

Fig. 13. Time evolution of x_D to parameter T_{M1} changes

Taking into account the time responses of concentration x_D obtained at step changes of set-point, illustrated in figures 12 and 13, the best tuning parameters for top concentration controller are: K_C = 1, K_{M1} = 0.4, T_{M1} = 30 min.

On c_2-x_B channel the process response is suitable for the application of the standard internal model control method. The process parameters identified for this channel and used as starting parameters in the tuning of the bottom concentration controller are: K_{M2} = 1.83, T_{M2} = 75 min, and T_m = 10 min.

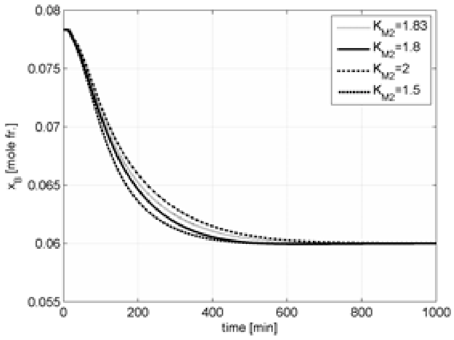

Fig. 14. Time evolution of x_B to parameter K_{M2} changes

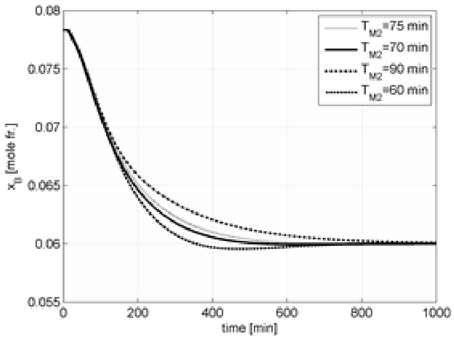

Fig. 15. Time evolution of x_B to parameter T_{M2} changes

Having as purpose to obtain a response without overshoot and a proper transient time, as it can be seen in figures 14 and 15, the best tuning parameters obtained for the bottom concentration controller are: K_C = 1, K_{M2} = 1.8, T_{M2} = 70 min.

In the following figures are presented the evolutions of the two concentrations at disturbances (F and x_F) changes.

Analyzing figures 16 and 17 it can be observed that the two internal model controllers associated to the concentrations x_D and x_B eliminate the errors caused by disturbances modifications.

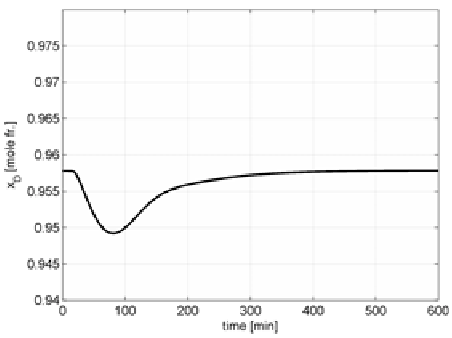

Fig. 16. Time evolution of x_D to a 3% change in F

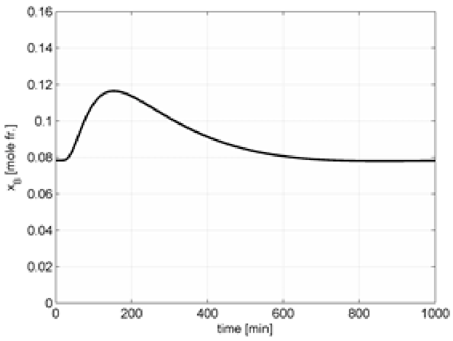

Fig. 17. Time evolution of x_B to a 3% change in x_F

The results from this section lead to the conclusion that the used internal model control method is robust for both set-point and disturbance changes. The obtained results are also confirmed by [15-17].

Third hierarchical level

The hierarchical control system with three levels for the butylene-butane distillation column is presented in figure 18.

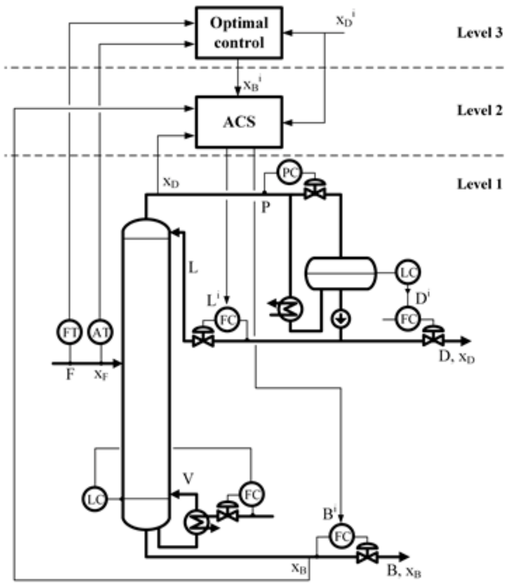

Fig. 18. Hierarchical control system with 3 levels

The control system from the third hierarchical level is an optimal control system for the butylene-butane separation process. The quality specification for the top product refers to the concentration of the isobutane + isobutylene mixture in the overhead product and is stiff with value $x_D^i = 0.96$ mole fr., and the specification for the bottom product which refers to the concentration of the isobutane + isobutylene mixture in the bottom product is flexible, $x_B^i \in [0.01 \ldots 0.09]$ mole fr.

The objective function at this level aims the optimal recovery of the isobutane + isobutylene mixture in the conditions of an energy effort as low as possible:

$$F_{obj}(x_B^i) = MM_B \cdot \Delta p \cdot B \cdot x_B^i + price_{steam} \cdot r \cdot (F + L - B), \quad (9)$$

where: $price_{steam}$ - price of the steam [euro/kg]; Δp – difference between the price of isobutane + isobutylene mixture and the price of n-butane + (cis- + trans)-butylene

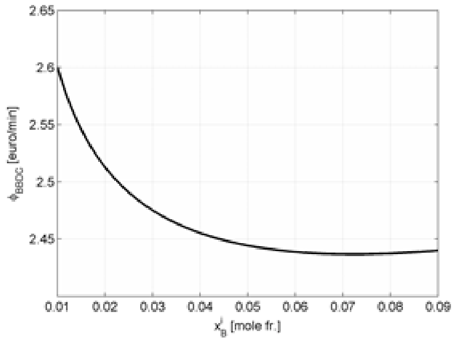

Fig. 19. Objective function representation

mixture [euro/kg]; B – bottom product flowrate [kmole/min]; MM_B – molar mass of n-butane + (cis- + trans)-butylene mixture [kg/kmole].

The term $r \cdot (F+L-B)$ represents (practically) the steam flowrate. Parameter r [kg/kmole] is the ratio between the latent heat of vaporization of the mixture from the bottom of the column and the latent heat of condensation of the steam.

The first term from relation (9) is associated to the recovery of the valuable product and represents the financial loss generated by the non-recovery of the valuable component from the bottom product. The second term defines the operating effort which refers to the steam used in the reboiler. The optimal value of the set-point x_B^i is obtained by minimizing the objective function.

At this level, the controller contains a control mathematical model of the process, the objective function (9) and an algorithm to calculate the optimal value of x_B^i.

The representation of the objective function is shown in figure 19.

By solving the optimization problem it was obtained xB^i_{opt} = 0.0721 mole fr., value which is sent as set-point for the control system from level 2 (ACS), and to be more precise for the internal model controller associated to concentration x_B.

The block diagram of the hierarchical control system is presented in figure 20.

The control system with three hierarchical levels was simulated in SIMULINK® the results being presented in figure 21.

Figure 21 shows that when the set-point x_B^i is modified the concentration x_D practically is not influenced (due to the presence of the decoupler), and the control system associated to the bottom concentration manages to bring x_B to the optimal set-point value received from the third hierarchical level (represented by the optimal controller).

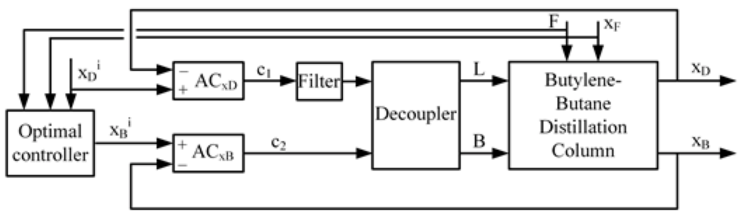

Fig. 20. Block diagram of the hierarchical control system

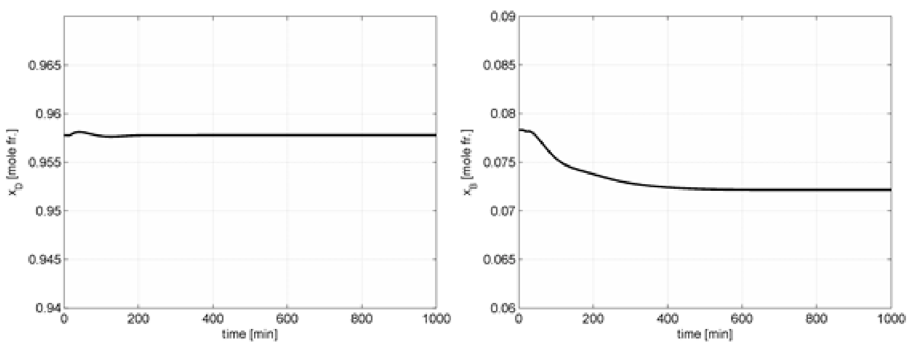

Fig. 21. Time evolutions of concentrations x_D and x_B to x_B^i change

Conclusions

The purpose of this paper was to present the design of the hierarchical control system for the butylene-butane distillation column. The proposed hierarchical system has three levels: first level represented by the column with its conventional automation, second level is the advanced control level and the third level has implemented an optimal control system.

The column with the conventional automation was simulated and the results confirmed the fact that there are crossed interactions between process channels. In order to diminish these interactions a nonlinear decoupler was designed. This offers the possibility of treating the concentration multivariable control system as two independent monovariable control systems.

At the second hierarchical level two internal model controllers were proposed for control of the top and bottom concentrations (x_D and x_B). The internal model controllers proved their robustness to set-point and also to disturbances modifications.

The third hierarchical level was dedicated to optimal control. At this level was proposed an objective function which aims the optimal recovery of the isobutane + isobutylene mixture in the conditions of an energy effort as low as possible. The controller from this hierarchical level contains a control mathematical model of the process, the objective function and an algorithm to calculate the optimal value of x_B^i.

Future work could include the design of a hierarchical control system for the whole plant from which the butylene-butane distillation column takes place.

Notations

M_K - tray K liquid holdup
LF_K, VF_K - external liquid and vapor feed flowrates of tray K
L_K, V_K -flowrates of liquid and vapor which leave tray K
x_{F_K} -concentration of the light component in the external liquid feed of tray K
y_{F_K} -concentration of the light component in the external vapor feed of tray K
x_K - concentration of the light component in the liquid phase on tray K
y_K - concentration of the light component in the vapor phase on tray K
α - relative volatility
NT - number of theoretical trays
NF-feed tray
F, x_F – feed flowrate and composition
L, V – reflux and boilup flowrates
q_F – feed liquid fraction
K – theoretical tray number (1 – bottom, NF - feed, NT - total condenser)

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