Multivariable control with adjustment by decoupling using a distributed action approach in a distillation column

Cintia Marangoni[1]. Joel G. Teleken[2]. Leandro O. Werle[3]. Ricardo A. F. Machado[4]. Ariovaldo Bolzan[5]

Federal University of Santa Catarina State – UFSC . Chemical Engineering Department.
P.O. Box 476 – Trindade, Florianópolis, SC – Zip Code 88010-970 – Brazil. Fone/ Fax: +55 (48) 3721-9554
[1](e-mail:sissi@enq.ufsc.br). [2](e-mail: joel_teleken@yahoo.com.br). [3](e-mail:leandro@enq.ufsc.br).
[4](e-mail:ricardo@enq.ufsc.br) [5](e-mail:abolzan@enq.ufsc.br).

Abstract: This paper presents an approach which uses a multivariable control strategy with distributed action in order to minimize operation transients in distillation columns when a disturbance in the temperature feed is introduced. Experiments were carried out adjusting multivariable PID controllers by static decoupling of the temperature loops of the bottom and distillate trays, characterizing a 2 x 2 system. The dynamics was compared to the distributed approach (same controllers on the bottom and top and an additional control loop on a tray). The controllers adjustment of this new system (3 x 3) was carried out considering the temperature control loop stage in two different ways: decentralized and coupled with the bottom and top temperature control loops. The minimization of transients was verified in both distributed approaches.

Keywords: control schemes, distillation columns, dynamics, multivariable systems, transient analysis.

1. INTRODUCTION

A well designed and adjusted control system is not sufficient to eliminate operation transients of a distillation process. One aspect that contributes to this situation, besides the column operation stage, is the centralization of the control system in the bottom and top column variables. In this way there is the propagation of the corrective control action through the whole unit, generating a production period out of the desired specification. The formation of transients in a distillation column occurs when the process is disturbed and its characteristics reduce the control system efficiency or when an external factor induces the modification of the unit operation point. In the first case there are factors such as variable coupling, nonlinearities, deadtime, high time constants and process constrains. In the second case there are aspects such as the mixture to be distillated, feed composition changes and operation transitions that are necessary due to changes in the market. In both cases, the process dynamics influences the way the transient operation will be generated and what the final result will be.

The current proposals to minimize the transient time of distillation columns use control techniques which consider process dynamics only to study the efficiency of these algorithms, without changing the process conception or evaluating the minimization of transients (Zhu and Liu, 2005). Stricter product specifications and greater demands in terms of environmental control, together with the design of more and more integrated units, require a better performance of these systems. Thus, economic incentives for the development and application of high performance control systems in industrial plants have grown considerably.

The proposal here developed, which is the object of this study, consists of the distribution of the control action throughout the column stages aiming at the minimization of the transient operation. This approach is based on the study of diabatic distillation columns (Koeijer, Rosjorde and Kjestrup, 2005), where intermediate heating points are used instead of only one heat input (reboiler) and one heat remover (condenser). These additional points keep a certain desired temperature profile throughout the column. Previous research (Marangoni and Machado, 2007) has demonstrated the feasibility of this proposal with the use of classic controllers (PID). The unit dynamics was evaluated and the results showed a reduction in the operation transition time when feed disturbances are introduced into the distillation column. Although 90% of industrial processes use classic controllers (Astrom and Hagglun, 2001), it was also necessary to evaluate the use of advanced controllers (model-based) which consider the process dynamics. Multivariable and predictive control seem to be the most used techniques due to their great flexibility. Here, it is worth mentioning the studies carried out with model-based controllers: Model Predictive Control (MPC) (Bezzo et al., 2005); Dynamic Matrix Control (DMC) (Jana et al., 2005); and Generalized Predictive Control (GPC) (Karanca, 2003). On the other hand, some studies have been carried out with proportional-integral-derivative (PID) controllers, aiming at a more flexible adjustment considering the distillation characteristics (Zhu and Liu, 2005). However, even in these recent studies, the controllers used to obtain the quality profile are implemented with the control action only in the bottom and top column stages.

Thus, aiming at the application of easy implementation strategies, the objective of this study was to evaluate the use of a 2 x 2 control system (controllers of the temperature loops of the bottom and distillate trays) and compare it to a new distributed approach (same controllers on the bottom and top and an additional temperature control loop on a tray) implemented in two different ways: the first considering the additional control stage without interaction with the other two control loops (decentralized), and the second considering the system as 3x3 multivariable.

2. MATERIALS AND METHODS

Experiments were carried out in a pilot unit processing an ethanol-water mixture. The conditions used are summarized in Table 1. Composition measurements were carried out during the experiments using a densimeter for alcohol.

Table 1. Operation conditions used in the experiments.

Variable	Value
Ethanol feed volumetric fraction	0.15
Feed Temperature	92^{o}C
Volumetric feed flow	300 L.h^{-1}
Column top pressure	1.25 bar
Drop pressure	0.25 bar
Reflux ratio (Reflux stream/Distillate)	5
Bottom Holdup	4 L
Accumulator Holdup	5 L

2.1 The pilot unit

The unit, illustrated in Figure 1, represents a tray distillation process. It operates in a continuous way and thus there is a main tank responsible for the feed.

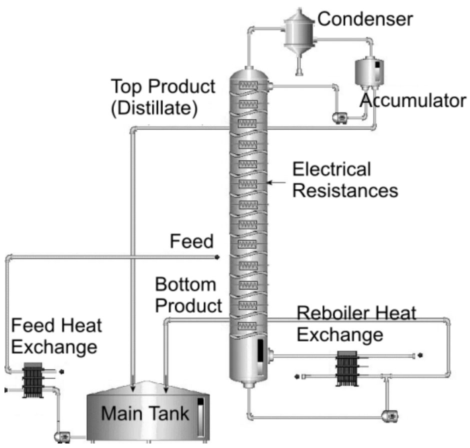

Fig. 1. Schematic illustration of the experimental unit's.

The column has 13 equilibrium stages and each module has one point for temperature measurement, one for sample collection and a third for the distributed heating adaptation. The latter was carried out by means of electrical resistances designed with up to 3.5kW power each. Temperature sensors (Pt-100) were used to monitor this variable in all equilibrium stages, as well as the main tank and the reflux accumulator. The feed was carried out on the fourth tray, with the reboiler as the zero stage.

The control configuration of the distillation column was formulated based on Nooraii et al. (1999), and is illustrated in Figure 2. The following control loops were defined: (1) bottom level control through the bottom product flow rate adjustment; (2) reflux accumulator level control by manipulating the top product flow rate; (3) feed flow rate control as a function of the adjustment of the same stream flow rate; (4) feed temperature control through the fluid flow rate adjustment in the heat exchanger of this stage; (5) last tray (distillate) temperature control by means of the manipulation of the reflux flow rate; (6) reboiler temperature control through the vapor flow rate in the heat exchanger of this stage; and (7) temperature control of pre-defined stages of the column through the adjustment of the dissipated power in the tray electrical resistance.

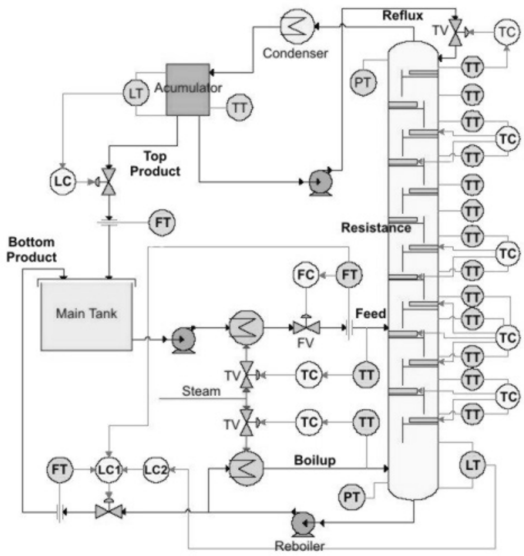

Fig. 2. Control configuration of the distillation unit.

The first, second and third loops represent the column mass balance (inventory) control. The fifth and sixth loops comprise the quality control – in this case represented by the temperature. The use of these two loops in combination is referred to herein as conventional control. When these two loops are combined with the seventh loop mentioned above, it is considered herein as the distributed strategy.

All control loops are instrumented with fieldbus protocols, along with the acquisition and indication of the bottom and distillate stream flows and the pressures at the same stages. The temperatures of all the trays, reboiler, accumulator and feed are monitored by a programmable logic controller and used in the dynamic study of the distributed control. The pressures were monitored in order to assure the proper functioning of the equipment and the process.

2.2 The control strategies tested

For this study, the experiments were carried out with three different control strategies: (1) conventional 2 x 2 – with multivariable control applied to the reboiler and distillate temperatures; (2) distributed 2 x 2 – with multivariable control applied to the reboiler and distillate temperatures and decentralized control in only one stage (PID without interaction with the other loops); and (3) distributed 3 x 3 – with multivariable control applied to the reboiler, second stage and distillate temperatures.

2.3 Controller's tuning

PID controllers are used in the three strategies tested. This kind of controller was employed since it is the most widely used (Astrom and Hagglund, 2001). Multivariable tuning was applied as the experiments consider both loops (reboiler and distillate temperatures) coupled to control the process.

Thus, for strategies 1 and 3 a static decoupler was used (Lee et al, 2005) to cancel the undesired effects of the interaction and adjustment of the multivariable controllers. This procedure can be designed from steady-state process gains, which are easier to obtain and can be tuned in the field. As an initial estimate the PID controller parameters were calculated using the criterion of the integral absolute error (ITAE). A fine adjustment was then made in the plant.

For strategy 2, the same controllers obtained in strategy 1 were used for the reboiler and distillate temperature loops. For the tray temperature controller the ITAE criterion was used to estimate the parameters followed by a fine adjustment (considering this loop decentralized from others, i.e., with weak interaction).

2.4 Stage selection

To identify the most sensitive stage for the consequent application of the distributed control, three different methods were applied (Luyben, 2006). In the first method, the difference between the temperatures of two successive trays was calculated throughout the column and the most sensitive tray was that which presented the greatest difference in relation to its adjacent tray. In the second method, a temperature profile for a given value of the manipulated variable (in this case, the reflux flow and the reboiler heat) is obtained. The most sensitive tray gives a symmetrical response to positive and negative equally variations. Finally, the third method analyzes the tray with the highest derivative of the temperature in relation to the stage when the process is disturbed. To analyze the first method, the temperature profile for three different conditions of ethanol feed composition (15, 25 and 35%) was observed. In the second and third methods it was necessary to disturb the process and evaluate its behavior. The feed flow used was 400L.h^{-1} as the standard condition, which was increased to 600L.h^{-1} and also decreased to 200L.h^{-1}.

It is important to emphasize that the different methods can produce different answers. The definition was based on this analysis together with the characteristics of the plant.

2.5 Disturbances

To analyze the control strategies, changes in the temperature feed were introduced, decreasing this variable by around 15^{o}C (from 92^{o}C to 77^{o}C). This was achieved by controlled cooling of this stream.

This study aimed to interfere in the column temperature profile in order to minimize the response time when some disturbance occurs. Thus, it was not tested for set point tracking.

3. RESULTS

The first step of this study was to determine the stage were distributed heating could be applied. As cited before, this was achieved through a sensitivity analysis employing three different methods. The results obtained with the first method (successive trays) using three feed ethanol composition conditions demonstrated the possibility of using trays 1, 2, 3, 5 and 7. As the fifth and seventh trays are located in the rectifying section, they were discarded. It was assumed, following diabatic studies upon which this proposal was based, that in this section it is better to remove heat than supply it.

In addition, as this is an initial study, it was defined that only one tray will be used to test the proposal. To define this stage, since method 1 was not conclusive, the analysis of symmetrical response and maximum derivative (methods 2 and 3) was used. The derivative method again pointed to stages 5 and 7, which were previously discarded, but the symmetrical response method indicated tray 2 as the most appropriate for this study. Figure 3 shows this analysis, where it can be observed that tray 2 is almost the same distance from steady state when the process is disturbed with positive and negative perturbations in the feed flow.

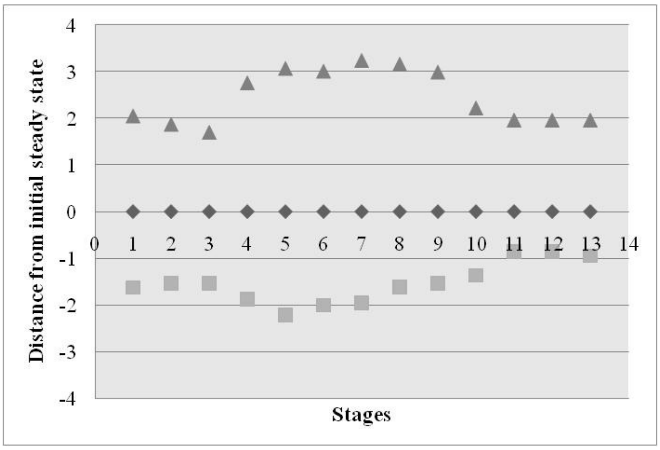

Fig. 3. Results of sensitivity analysis using symmetrical response method (■ negative disturbance, ◆ steady state, ▲ positive disturbance).

Based on this sensitivity analysis, the distributed action of the proposal was used only in tray 2. The simultaneous action of the trays was not tested because the main objective was to analyze the distributed proposal with a multivariable system, and its behavior, through the coupling of control loops.

In sequence, the process transfer functions were determined and the relative gain array matrix was evaluated.

Multivariable control algorithms were studied since the process has multiple inputs and outputs and it is characterized by the high coupling degree among the variables. These kinds of controllers require process models for which an approximated model was used, obtained by the transfer functions of the reboiler, second stage and distillate temperature control loops. In fact, a multivariable system can be easily modeled through transfer functions. These functions associate the system outputs (Y) with the disturbances (L) and the inputs (U), and integrate the transfer function matrix with the disturbance (G_L) and inputs (G).

The transfer functions were obtained by means of experimental tests, through input and output data collection and later numeric treatment of this information, disturbing the variables that are used for the manipulation of the control loops. The equations obtained are presented below (time values expressed in seconds and deadtime obtained by Taylor series approximation).

Equation (1) presents the matrix which represents the 3 x 3 system, where Tb is the reboiler temperature, Td the distillate temperature and T_2 the second stage temperature, corresponding to the system outputs. Q_b (steam valve opening at the reboiler entrance), R (reflux flow valve opening) and Q_2 (dissipated power at the electrical resistance stage), represent the inputs.

$$\begin{bmatrix} Y_1 = Tb(s) \\ Y_2 = Td(s) \\ Y_3 = T_2(s) \end{bmatrix} = \begin{bmatrix} G_{11}(s) & G_{12}(s) & G_{13}(s) \\ G_{21}(s) & G_{22}(s) & G_{23}(s) \\ G_{31}(s) & G_{32}(s) & G_{33}(s) \end{bmatrix} \begin{bmatrix} U_1 = Qb(s) \\ U_2 = R(s) \\ U_3 = Q_2(s) \end{bmatrix} \tag{1}$$

Equations (2) to (10) present the transfer functions obtained, which consist of the input/output relations presented in (1). For the calculations, the deadtime of these functions were expressed using a simple first-order Taylor series approximation.

$$G_{11} = \frac{0.69}{112s+1} \tag{2}$$

$$G_{12} = \frac{3.12e^{-13s}}{203s+1} \tag{3}$$

$$G_{13} = \frac{0.56e^{-1s}}{145s+1} \tag{4}$$

$$G_{21} = \frac{-0.08e^{-26s}}{172s+1} \tag{5}$$

$$G_{22} = \frac{-0.04e^{-4s}}{364s+1} \tag{6}$$

$$G_{23} = \frac{-0.06e^{-17s}}{135s+1} \tag{7}$$

$$G_{31} = \frac{0.02e^{-1s}}{74s+1} \tag{8}$$

$$G_{32} = \frac{0.14e^{-2s}}{407s+1} \tag{9}$$

$$G_{33} = \frac{0.02}{85s+1} \tag{10}$$

Experimental tests were carried out, data were evaluated and with the process equations the existing interactions were verified by controlling the process with and without the proposed approach. Experiments were carried out in order to construct the relative gain array matrix (RGA) (Shinskey, 1996) for the 2 x 2 system (reboiler and distillate temperature control loops) and for the 3 x 3 systems (reboiler, second stage and distillate temperature control loops). In this case, the cited method was used to identify the degree of coupling among the proposed systems and not to define the control structure, which is the usual purpose. This evaluation is important since the intermediate column stages also influence the temperature profile and the process composition.

Equation (11) presents the matrix obtained for the 2 x 2 system and (12) the matrix for the 3 x 3 system.

$$\Lambda = \begin{bmatrix} 0.89 & 0.11 \\ 0.11 & 0.89 \end{bmatrix} \tag{11}$$

$$\Lambda = \begin{bmatrix} 0.77 & 6.80 & -6.57 \\ -4.92 & -0.07 & 6.03 \\ 5.50 & -5.69 & 1.54 \end{bmatrix} \tag{12}$$

Although the selection of the best control structure is not the main objective of this study, the 2 x 2 system is adequate as shown in (11), where the sum of the matrix columns and lines are 1 (one).

In (12) we can observe some values above one for seven of the nine possible combinations of control loops, which indicates strong interactions in these combinations. It is well known that the closer the element is to +1, the weaker the interaction between the loops, and the elements with high modulus values indicate strong interactions between the loops, or it could be that the system is sensitive to parameter changes (less robustness).

With this phase completed, the studies were followed by experimental tests using multivariable control algorithms. As mentioned above, the common technique of controller adjustment by decoupling was used. This tuning was defined since it provides good results (Waller, et al. 2003, Liu et al., 2006). It is important to note that the objective is to improve the control of a new column and the operation approach, and therefore the application of techniques used industrially is the aim.

In decoupled control, it is implicit that the design objective is to obtain a system that reduces the interaction between the loops through specific additional controllers, called decouplers. These are used to improve the performance of multivariable control systems through interaction compensation, though they are sensitive to changes in the process and require detailed process models, which are often difficult to obtain. These disadvantages often limit the use of multivariable controllers industrially. However, the static decouplers approach can be designed from the gains in the process in steady state, which are easy to obtain and can be adjusted in the field (Lee et al. 2005). Because of these advantages, the static decouplers approach was used, based on the gain of each control loop.

Besides the objective of the implementation and study of advanced techniques using the distributed approach, these studies were carried out to evaluate whether or not it is possible to work with the hypothesis that the interactions caused by the tray can be eliminated when the reboiler interactions are reduced or eliminated.

Figures 4 and 5 show the reboiler and distillate temperature profiles for the strategies applied. In these experiments, the disturbance was applied by decreasing the feed temperature.

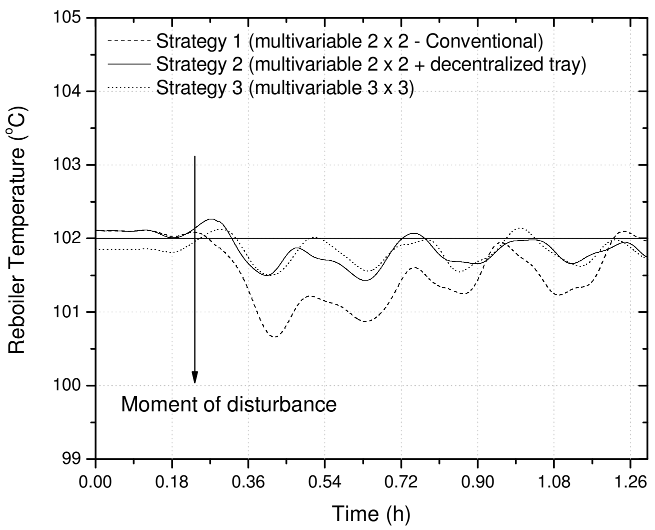

Fig. 4. Effect of the disturbance on reboiler temperature control loop response in relation to setpoint.

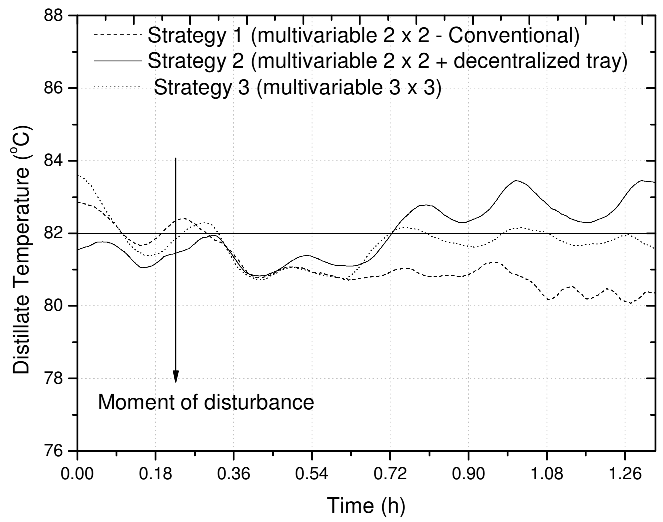

Fig. 5. Effect of the disturbance on distillate temperature control loop response in relation to setpoint.

It can be observed, in both figures, that the disturbance is quickly rejected when the distributed approach is used. Both strategies 2 (2 x 2 multivariable adjustment and decentralized adjustment at stage 2) and 3 (3 x 3 multivariable adjustment) showed that the steady state was reached faster than with strategy 1 (conventional multivariable control – 2 x 2 system). This result indicates that the distributed control action maintains the temperature profile in the column and thus it allows the reduction of the transients generated.

However, strategy 2, which assumes that the interactions between the tray temperature control loops and the other quality controllers is weak, leads to a value slightly higher than that desired. This case is better observed in relation to the distillate temperature. For this same variable, the disturbance applied was not completely rejected using the conventional control, and the distillate temperature stabilized at a lower value.

When reboiler and distillate temperature loops are evaluated together, the 3 x 3 distributed approach allows a better performance. It is possible that this adjustment, considering the interactions between the three control loops, made the system a bit slower, although it is still faster and less oscillatory than the conventional approach.

Figure 6 gives the second stage temperature profile, where the distributed control was implemented. As expected, the performance of the 2 x 2 multivariable adjustment strategy with decentralized adjustment at stage 2 leads to a value closer to the set point. In fact, using a decentralized PID controller at this stage, leads to faster dynamics than applying a multivariable 3 x 3 system which consider all interactions of this tray with the reboiler and distillate temperature control loops. However, strategy 3 showed a slight overshoot and rejected the disturbance quickly, in contrast to the conventional strategy.

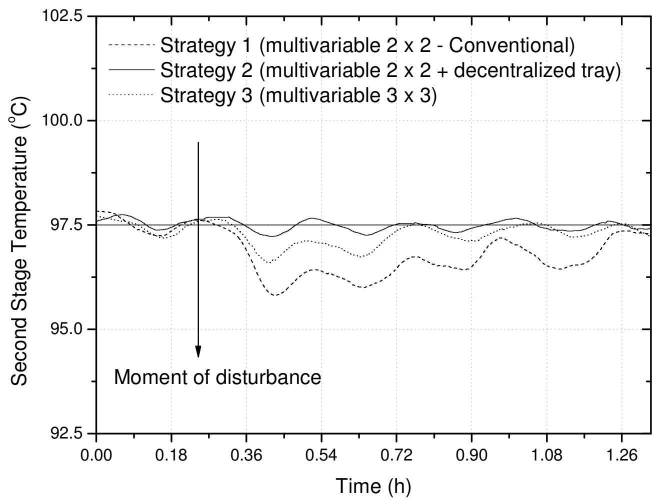

Fig. 6. Effect of the disturbance on the tray 2 temperature control loop response in relation to the setpoint.

In order to carry out a final evaluation regarding which strategy leads to the best performance, the effect of the disturbance on the temperature of the accumulator tank was studied. Since this is the last unit stage, it is the one with the highest transition time. Therefore, its behavior was observed by analyzing the temperature derivative in relation to the time required for the disturbance rejection, as illustrated in Figure 7.

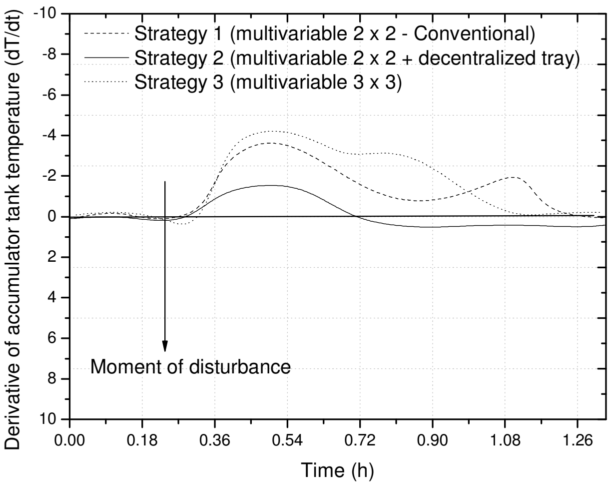

Fig. 7. Derivative of the temperature of the reflux accumulator tank in relation to the setpoint.

The figure demonstrates that the time required to reduce the effects on feed temperature disturbance is shorter when the distributed control approach is applied, considering the second stage temperature control loop not interacting with the others. This hypothesis will be true if the interaction at this stage can be eliminated by the reboiler temperature control loop decoupling. However, it is important to note that the value of the accumulator tank temperature did not return to the same steady state present before the disturbance. It is possible that the use of a PID decentralized controller at stage 2 allowed the production of a greater vapor phase inside the column as the temperature of the last stage was higher in this case. If this occurred, the condenser would produce more distillate and the accumulator temperature tank would stabilize at a different value, as was in fact observed.

4. CONCLUSIONS

The evaluation of the conventional and distributed approach, for a feed temperature disturbance, allowed a reduction in the column transition time and in the oscillations of the controlled variable when the strategy with control at stage 2 was used (independent of the tuning of this loop – decentralized or not).

It is also necessary to consider that the decentralized utilization of the temperature control loop which comprises the distributed approach gives better results than the conventional control system. This is an important result since most industrially implemented controllers are considered decentralized (Garelli et al. 2006).

The comparison between the use of a distributed control loop, decentralized or not with the reboiler and distillate temperature loops, shows that the hypothesis of weak interaction of an intermediate stage can be assumed. When a decoupler was used to tune the quality controllers of the base and top it is possible that the interactions with the trays were reduced. Thus, with the application of an advanced control algorithm, it is observed that the introduction of heat to one of the column stages allows a reduction in the operation time out of the desired conditions. As with classic controllers tested in previous research studies, the introduction of distributed heat throughout the column was shown to be a valid option for the reduction of transients, enabling faster dynamics and lower volumes of products processed out of the pre-defined quality parameters.

ACKNOWLEDGMENTS

The authors are grateful for the financial support of the National Agency of the Petroleum - ANP - and of the funding agency - FINEP - through the Program of Human Resources of ANP for the Petroleum and Gas Sector - PRH-34-ANP/MCT and to the National Council of Research - CNPq.

REFERENCES

Astrom, K.J.; Hägglund, T. (2001) The future of PID control. *Control Engineering Practice*, v. 9, p. 1163-1175.

Bezzo, F; Micheletti, F.; Muradore, R. Barolo, M. (2005) Using MPC to control middle-vessel continuous distillation columns. *Journal of Process Control.* v.15, p. 925-930.

Garelli, F., Mantz, R. J., Battista, H. De. (2006) Limiting interactions in decentralized control of MIMO processes. *Journal of Process Control.* v. 5, p. 473-483.

Jana, A. K.; Samantha, A. N.; Ganguly, S. (2005) Globally linearized control system design of a constrained multivariable distillation column. *Journal of Process Control.* v. 15, p., 169-181.

Karaca, S. (2003) Application of a non-linear long range predictive control to a packed distillation column. *Chemical Engineering and Processing*, v. 42, p. 943-953.

Koeijer, G.; Røsjorde, A.; Kjelstrup, S. (2005) Distribution of heat exchange in optimum diabatic distillation columns. *Energy*. v. 29, p. 2425-2440.

Lee, J., Dong, H. K.; Edgar, T. F. (2005) Static decouplers for control of multivariable processes. *AICHE Journal.* v. 51, p. 2712-2720.

Liu, T., Zang, W., Gao, F. (2006) Analytical decoupling control strategy using unit feedback control structure for MIMO processes with time delays. *Journal of Process Control***.** v. 17, p. 173-186.

Luyben, W.L. (2006) Evaluation of criteria for selecting temperature control trays in distillation columns. *Journal of Process Control* v. 16, p. 115-134.

Marangoni, C. Machado, R. A. F. (2007) Distillation tower with distributed control strategy: Feed temperature loads. *Chemical Engineering an. Technoly.* v 30, p 1292-1297.

Nooraii, A.; Romagnoli, J. A.; Figueroa, J. (1999) Process, identification, uncertainty characterization and robustness analysis of a pilot scale distillation column. *Journal of Process Control*, v. 9, n. 3, p. 247-264.

Shinskey, F. G. (1996) Process Control Systems. McGraw-Hill, 4a edition, New York, 439p.

Waller, M., Waller, J. B., Waller, K. V., (2003) Decoupling revisited. *Industrial & Engineering Chemical Research.* v. 42, p. 4575-4577.

Zhu, Y. & Liu, X. G. (2005) Dynamics and control of high purity heat integrated distillation columns. *Industrial & Engineering Chemical Research.* v. 44, p.8806-8814.