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Use of Stripping or Rectifying Trays for a Distributed Control Strategy in Distillation Column

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In distillation, transient behavior cannot be completely eliminated, even with the application of advanced controllers, because the operation in stages imposes the propagation of corrective action throughout whole unit. In previous studies, a control strategy was proposed that uses a combined action between reboiler and heating tray in stripping section of the column aiming to minimize transient operation in distillation. In this work, using the commercial simulator Aspen HYSYS Dynamics, the application of this control strategy was evaluated and it was compared with one testing a distributed cooling action applied to a tray of the rectifying section. Both strategies were also compared with a conventional dual temperature control system when disturbances in the feed temperature were performed. Results had demonstrated a reduction in transition time in bottom and top temperature control loop when distributed strategy was use, either when heating or cooling actions were taken. Internal variables were analyzed in order to verify alterations in flows. Steady state profiles of temperature and ethanol composition were not modified in relation to that obtained with conventional actions, indicating that the proposed distributed strategy only influences transition time.

1. Introduction

Distillation control systems are developed with varied objectives including increasing production, minimizing operational transients, energy consumption or products out of specifications, reducing process costs, eliminating risks inherent to the process and improving final product quality (Enagandula and Riggs, 2006). When the system is disturbed either by alterations in the operational points or by a desired set point change the rapid achievement of steady state minimizes the time necessary to meet the required product specifications. Numerous difficulties in the control of distillations units are due to inherent process characteristics that represent a challenge for the reduction of transient time. Among these are non-linear behavior, which is associated with the coupling of variables; operational restrictions; elevated time constants; and a delay in the response (Skogestad, 2007). This behavior results from the configuration of the column in stages, requiring successive heating and cooling actions for heat and mass transfer to occur in the stages; therefore, there is a propagation of a control action throughout the unit to establish the product quality. The result is a transient behavior that is difficult to eliminate, even with well-adjusted control systems.
The solution most used to reduce operation time involves the implementation of control techniques, more specifically, advanced systems that consider the unit dynamics in their structure. However, these tend to be difficult to implement, and simple industrial controllers, such as PID controllers (proportional-integral-derivative) are still widely used (Astrom and Hägglund, 2001). On the other hand, proposals for intensified units have been made with the aim of miniaturization and better energy use. Diabatic distillations are one example, where heat provided to the reboiler is distributed in the stripping section and the heat removed by the condenser is distributed between the trays in the rectifying section (Koeijer et al., 2005).

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Connecting this concept of heat distribution in the trays with the objective of minimizing operational transients, a control scheme with distributed corrective action to minimize the effect of perturbations in feed was previously proposed and evaluated (Marangoni et al., 2009). In this approach, the objective of controlling product quality in the bottom and top streams was performed by associating the conventional temperature dual control with an internal heating stage in stripping section of the unit. Results indicated that the approach was a valid option for reducing transient behavior. However, in all of the tests performed, only heating points were considered. Due to the experimental nature of the constructed unit, the analysis of heat removal and internal variables profiles could not be performed. Thus, the objective of this study is to evaluate the distributed corrective action proposal applying heating and cooling actions. For that, PID controllers were implemented to the bottom and top stages temperatures and associated with one stage of the stripping/rectifying section aiming to minimize perturbations in the feed temperature. It is important to emphasize that this study is focused on process dynamics and not on the design or tuning of the control system. The results presented in this work are only simulated once the experimental unit does not allow the evaluation of cooling actions working with a tray in rectifying section as proposed. Thus, the simulations representing the unit dynamics were validated by reproducing the experimental results of previous studies (Marangoni et al., 2013), where the distributed action was evaluated applied to a tray in stripping section (heating action). From these simulation cases, the proposed work presented here was evaluated, considering only disturbances in feed temperature (and not feed flow or composition disturbances).

2. Methodology

This simulation study presented in this work was based on the continuous experimental unit used by the research group in previous studies (Werle et al., 2009) which contains 13 perforated trays, with the reboiler being stage zero and the accumulator being stage 14. The feed (ethanol and water) is added to tray 4. Simulations were performed employing the Aspen HYSYS Dynamics software version 7.3 (AspenTechnology Inc, 2014), with the UNIQUAC thermodynamic package (Billal et al., 2014). The conditions used are listed in Table 1 and are the same as tested experimentally.
Two different control strategies were evaluated: a conventional and a distributed strategy, in which PID type controllers were implemented in the following loops for the conventional system: (1) bottom temperature control, manipulating the reboiler heat and (2) top stage temperature control manipulated by the reflux flow rate. The distributed approach considers these two loops with the addition of another intermediate temperature control that was activated separately. For heating actions, tray 3 was used and for cooling strategy, tray 11 was selected – both determined by sensitivity analysis (Mello et al., 2013). For this control loop, the manipulated variable was the heat removed from or added to the tray. Parameters used for the controllers are listed in Table 2 and were determined from fine-tuning based on experimental values.
It is important to emphasize that from the experimental viewpoint, the addition and removal of heat must be promoted with devices coupled to the trays of the unit. In the simulations, the software only enables one simple energy stream to be inserted into the stage and to be defined in terms of heat exchange. The range of 0 to 3.5 kW was defined once this was the range applied experimentally (Werle et al., 2009). For cooling, 100% corresponds to zero heat removal (minimum value) and 0% indicates that the total cooling value was applied (in this case, -3.5 kW).

Table 1: Operational conditions and parameters of the distillation column

Variable	Value
Feed temperature	Sub cooled (~ 80°C)
Volumetric feed flow rate	300 L/h
Volumetric fraction of ethanol in the feed	0.2
Pressure at the top of the column	1.2 bar
Pressure drop along the column	0.15 bar
Reflux ratio	6

Table 2: Parameters used for the PID controllers

Parameter	Reboiler	Tray 13	Tray 3 or 11
K_c	3.90*	0.70**	7.22*
τ_i (s)	7.45×10^{-2}	9.67×10^{-2}	2.57×10^{-2}
τ_d (s)	9.17×10^{-3}	1.00×10^{-2}	5.33×10^{-3}

* (℃/%heat transferred), ** (℃/%valve opening)

The conventional strategy was validated based on the experimental results of the transition time when the unit was perturbed in the feed temperature. In this study, when distributed system was tested for heating actions, -14 °C was applied in feed temperature (same used experimentally). For cooling action, +14°C was applied in same variable.

3. Results and Discussion

Figure 1(a) shows the derivative of the bottom temperature with respect to time when tray 3 (stripping section) is used in the distributed strategy. Figure 1(b) shows the same behavior, but when the distributed strategy is applied to tray 11 (rectifying section). Derivative profile was chosen once represents the time that the system is operating out of steady state (zero line, in this case), i.e., the transient time. In both cases, it could be observed that the distributed strategy has lower transition time compared with the conventional one, that is, the distributed control is more efficient compared with the conventional. To minimize feed perturbation, when tray 3 was used for the application of the distributed strategy, 0.09 h (5.4 min) were necessary with the conventional approach to reach the set point value and this time was reduced to 0.07 h (4.2 min) with the distributed control. On the contrary, when tray 11 was used, 0.10 h (6 min) were needed for the conventional strategy, and 0.08 h (4.8 min) for the distributed strategy. Either for heating or cooling actions, the distributed system allows to obtain a 20 % reduction in transition time of the bottom temperature control.
The derivative of top stage temperature is presented in Figure 2(a) for distributed strategy using to a tray 3 and in Figure 2(b) for the same strategy applied to the rectifying section (tray 11). In both cases for this control loop it is confirmed that the distributed strategy achieves stability faster than the conventional strategy. The transition time for the conventional approach was 0.5 h (30 minutes) and for the distributed approach 0.18 h (10.8 minutes), yielding a reduction of approximately 64 % when tray 3 was used. On the other hand, the transition time for the conventional strategy and distributed strategy was 0.44 h (26.4 minutes) and 0.12 h (7.2 minutes), respectively, resulting in a reduction of 72.72 % when the tray 11 was used.
These results can be corroborated through the analysis of the control actions (manipulated variables), that is, through the profile of the heat transferred to the reboiler and of the reflux flow rate (Figure 3).

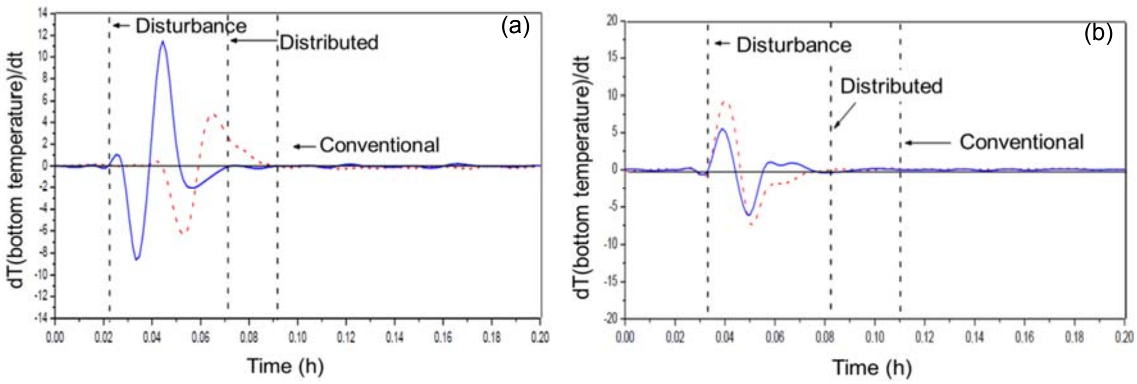

Figure 1: Derivative of bottom temperature in relation to time comparing the conventional strategy (- - -) with the distributed control (—) applied to tray 3 (a) and tray 11 (b) relative to the set point value

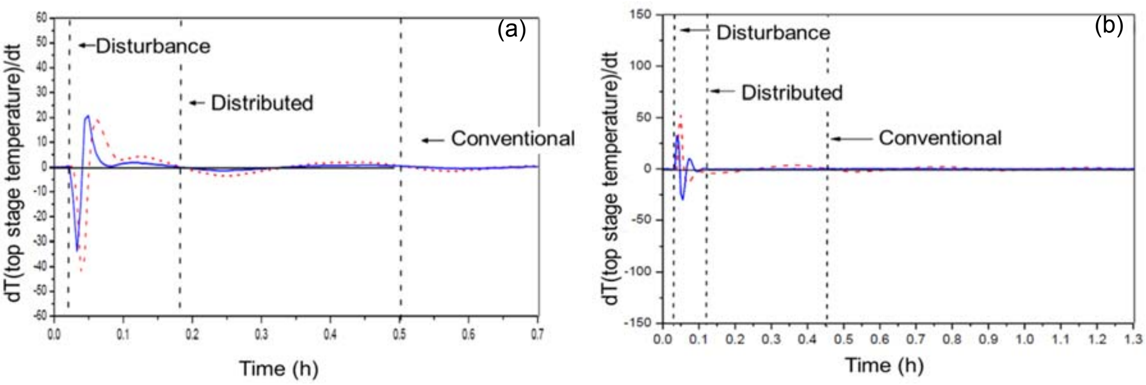

Figure 2: Derivative of tray 13 temperature in relation to time comparing the conventional strategy (- - -) with the distributed control (—) applied to tray 3 (a) and tray 11 (b) relative to the set point value

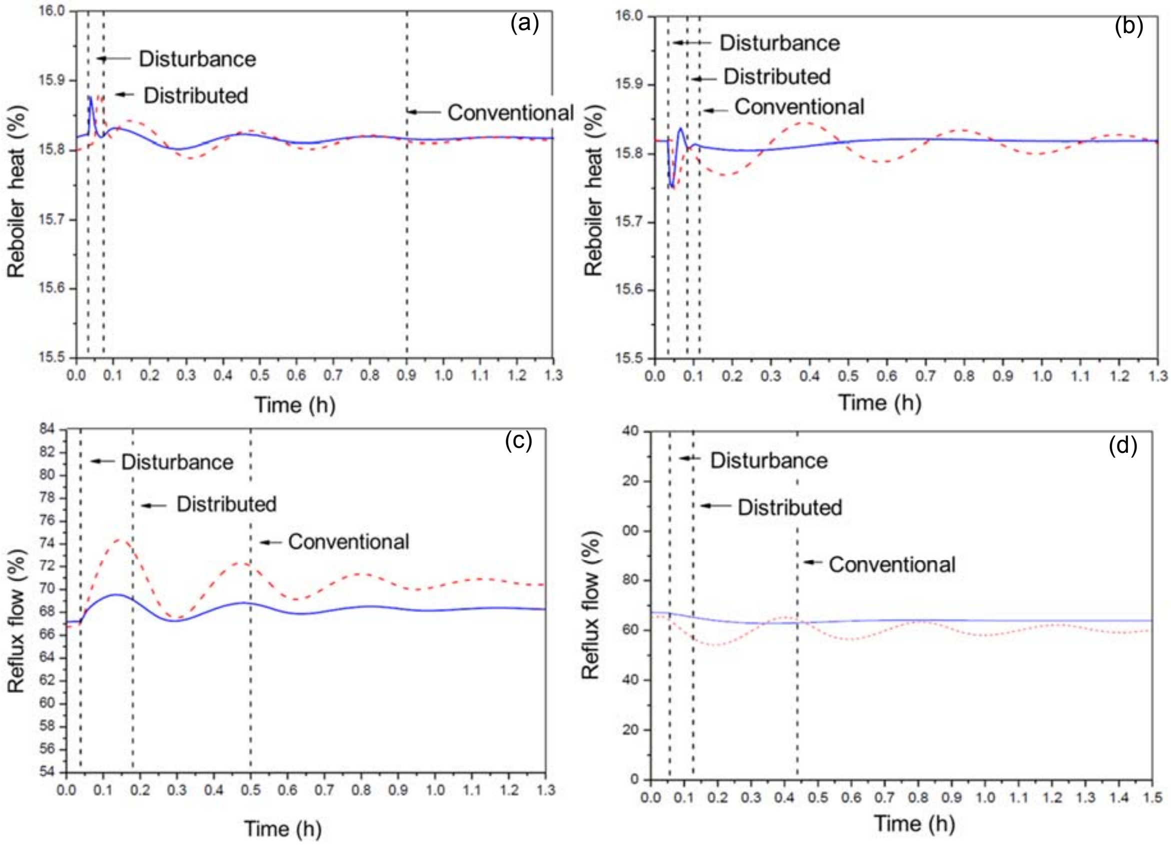

Figure 3: Control action profile of the distributed strategy (—) compared to the conventional (- - -): reboiler heat profile for distributed applied to tray 3 (a) and tray 11 (c) and reflux flow rate for distributed applied to tray 3 (b) and tray 11 (d)

Figure 3 shows the reduction in oscillations in manipulated variables profiles when the strategy implementing the distribution of heat exchange is applied. Because heat is being added to tray 3 coupled with the reboiler heating and removed from the column at tray 11 in conjunction with the action performed in the reflux flow rate, more rapid temperature stabilization occurs through the stages. This behavior is more evident in the reflux flow rate profile, demonstrating that less valve opening is required from this variable when tray is activated. For the heat transferred to the reboiler, the conventional approach is slower to initiate the correction process due to the need for propagation of the action from the reflux to the reboiler. This action is anticipated by the tray controller, making the distributed control process faster.

The temperature of tray 3 and 11 (stages used in distributed control) was also analyzed. It is verified in Figure 4(b) that the conventional control strategy achieved the reference value for tray 11 even without the controller on this stage but with a constant oscillation around the set point. This behavior, as expected, is different when the distributed strategy is used, once a local control is implemented the transition time for the conventional strategy and the distributed strategy applied to tray 11 is 0.47 h (28.2 min) and 0.14 h (8.4 min), respectively. As shown in Figure 4(a), for distributed control using tray 3 a more interesting situation could be observed. In addition to the reduction in transition time (0.12 h for conventional and 0.08 h for distributed one), it could be visualized that in stripping section, the conventional strategy is not able to reach the steady state, stabilizing at a temperature value lower than the set point by 0.6 ºC. This could be explained by the application of an instantaneous energy supply to the tray with the distributed approach that causes the temperature to remain at the desired value.

Because the quantity of product produced out of specifications also depends on the time the process takes to return to the desired steady state after a perturbation occurs, an estimate of this quantity was obtained and is listed in Table 3. The data listed in Table 3 demonstrate that the volume of product out of specification in the distillate stream when the distributed control strategy is used corresponds to 38 % of the total volume produced testing the conventional approach when tray 3 was used and to only 3.5 % when tray 11 was employed. That is, the reduction in the oscillations and transition time promoted by the proposed approach represent an elevated gain in product quality when considering the productive scenario.

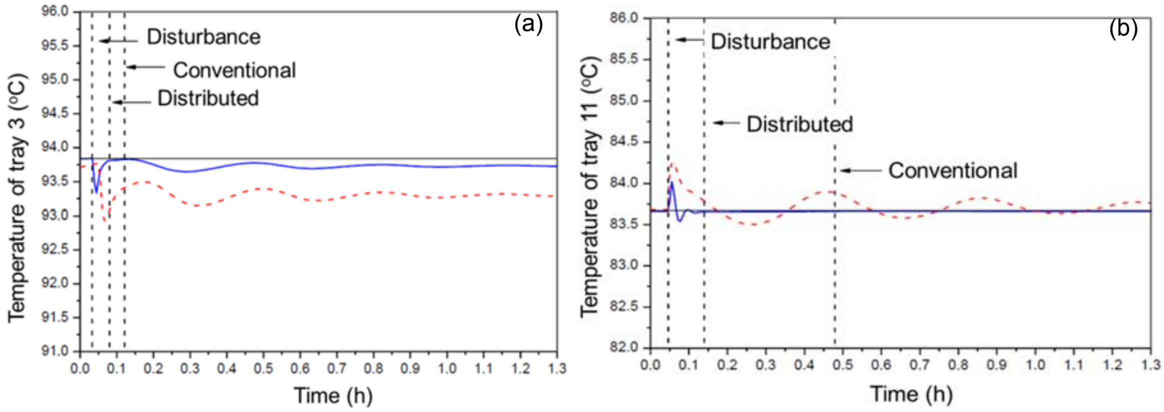

Figure 4: Temperature profile on tray 3 (a) and on tray 11 (b), comparing the conventional strategy (- - -) and the distributed strategy (—) relative to the set point value (—)

Table 3: Quantity of product out of specifications (in volume) during the transition time for the conventional and distributed strategies

Flow rate (m^3/h)	Conventional	Distributed with tray 3	Distributed with tray 11
Distillate	5.45×10^{-5}	2.09×10^{-5}	1.95×10^{-6}

Finally, the temperature, pressure and composition profiles along the stages of the distillation unit (Figure 5) were evaluated, comparing their values before and after the perturbation with the application of the two evaluated strategies.

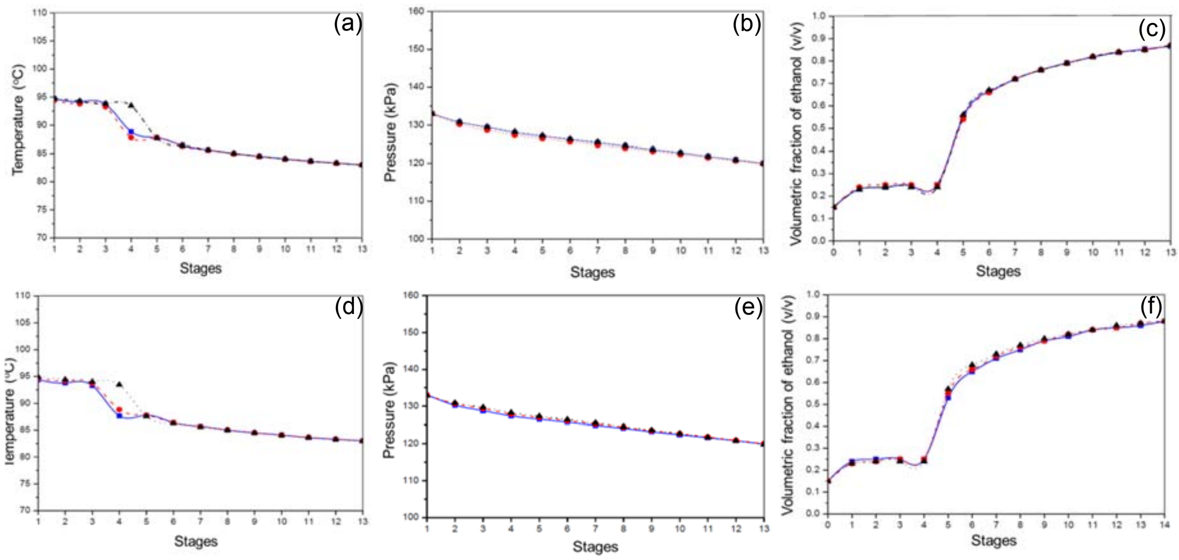

Figure 5: Profiles of temperature, pressure and volumetric fraction of ethanol along the stages of the column at steady state (-▲-) for the conventional strategy (-●-) and for the distributed strategy (-■-) applied to tray 3 (a,b,c) and tray 11 (d,e,f)

The insertion of an intermediate control loop, which composed the distributed strategy, did not affect the final steady state after the perturbation, demonstrating that this proposal did not alter the product quality, independent of the applied strategy. That is, the distributed action makes the control faster, without promoting alterations in the profiles of variables in the column stages.

4. Conclusions

In general, distributed control exhibited better performance compared with conventional control for the analysis of the distributed action either in stripping or rectifying section. The behavior of the variables became less oscillatory, and the controllers maintained the process variables closer to the desired values after the disturbance for all of the control loops that were affected by the perturbation.

The results presented consolidate the control approach with distributed action that has been proposed because the heating actions (proven experimentally by our group in previous studies and corroborated here) or cooling actions can be used with classical, PID-type controllers to reduce the transient behavior observed when a distillation unit is perturbed.

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