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Temperature-Temperature cascade control of Binary Batch Distillation Columns

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***Abstract*—In this paper it is addressed the joint operation and control design for binary batch distillation columns, based on temperature measurements. The combination of nonlinear constructive control theory with passivation and observability notions and existing batch distillation concepts yields: (a) a simple methodology for temperature sensor locations, (b) the on-line generation of temperature policies that ensure constant distillate purity, and (c) a temperature-to-temperature cascade tracking controller to force the prescribed constant distillate product purity policy. The methodology is tested with representative binary systems.**

I. Introduction

Distillation is one of the most common operations in the chemical industry, as it is a required operation for the production of many intermediate and final products in the chemical and petrochemical industries. Distillation columns are energy intensive equipments [1], in which the regulation of distillate and/or bottoms compositions is required. The related feedback control design for continuous distillation units has been extensively studied with a diversity of linear and nonlinear approaches [2]-[3] .

Batch distillation columns (BDC) offer advantages over continuous columns, especially, their flexibility and the possibility of separating multicomponent mixtures using the same unit [4]-[5]. The resulting operation and control problems are complex due to the resulting transient nonlinear and finite time dynamics, and have been addressed in sequential manner. First, the operation is designed via open-loop optimization [6], yielding a nominal output signal to be followed by means of a feedback control that is designed in a second stage, using either linear gain scheduled [7], nonlinear model predictive techniques [8], or geometric techniques [9].

Other works have studied the constant distillate composition problem, i.e., the design of a control scheme that computes a variable reflux rate policy that keeps constant distillate purity, by means of modeling error compensation techniques ([10] and references therein), driven by possibly uncertain and dead-time affected composition measurements. On the other hand, an observer-based optimal closed-loop operation technique for binary batch distillation columns has been proposed [11]. In this work, three main ingredients are needed: (i) the solution of the state-feedback (SF) control problem for constant distillate purity that sets the time-varying reflux flow rate, (ii) a temperature measurement-driven nonlinear observer to reconstruct the column state, and (iii) an on-line computed objective function that decides the duration of the production period on the basis of maximum profit. This technique is robust in the presence of initial condition uncertainty (i.e., mass and composition of the batch feed). The main drawbacks of this technique are its high model dependency, the complexity of the required nonlinear observer, and noise propagation. Additionally, the existing cascade composition-to-temperature control design methodologies for continuous columns ([12]-[13] and references therein) have been applied to batch distillation columns [14]. Given that cascade composition-to-temperature (cascade CT) control schemes depend heavily on composition measurements, and that batch distillation columns operate at (rather short) finite times, dead times and uncertainty can affect severely their functioning. Therefore, there is still a need for efficient control designs with reduced model dependency and improved robustness.

In previous works [15]-[16], the existing cascade CT control design methodology for continuous distillation columns was modified for the case in which additional temperatures replace composition measurements. Here, the decentralized control loops are designed using constructive control arguments (i.e., passivation via backstepping), according to the following rationale: (i) the constructive nonlinear control-based cascade composition-to-temperature linear-decentralized control scheme is recalled [13]), and (ii) its primary slow composition loop pair is replaced by a faster temperature driven counterpart, with measurement locations that represent a suitable tradeoff between sensitivity and closeness to the column end. The resulting closed-loop (CL) behavior yielded significant improvement of the transient response, i.e., smaller effluent purity deviations and regulation times.

Motivated by: (i) the previous observer-based optimal closed-loop operation of BDC [11], and (ii) the above mentioned idea of improving the effluent regulation capability of existing temperature control schemes for continuous binary distillation columns by employing additional temperature measurements [15]-[16], and by the need of assessing the general-purpose feasibility of the idea and systematizing its control design, in this work the problem of controlling the distillate composition in a binary batch distillation column by manipulating the reflux flow

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rate on the basis of two temperature measurements is addressed:

First, the solution to the constant distillate state feedback (SF) control problem for binary distillation column is recalled, and the corresponding zero dynamic are invoked and reinterpreted as the time-varying temperature policies that must be tracked in order to yield constant distillate purity. Then, in a way that is analogous to the development of a previous SI-2O (single input-two output) linear-decentralized cascade control scheme [15-16], a two-state linear-decentralized model with reconstructible load inputs, built on the basis of the column relative degree structure, is set. Then, following previous polymer reactor [17-18] and distillation column [11,19] and constructive control ideas [21-22], a (passivated by backstepping) output feedback (OF) controller is built within a nonlinear constructive control framework. The result is a linear temperature-driven cascade controller with reduced model dependency and straightforward construction and tuning, with: (i) a primary loop that, driven by the top tray temperature regulation error, generates the time-varying setpoint for the (ii) secondary controller, that manipulates the reflux flow rate to track the time varying temperature setpoint computed by the primary loop, that yields constant distillate purity. The resulting methodology is applied to a representative example through simulations, yielding behaviors that recover the performance of previous nonlinear schemes.

II. Statement of the Problem

Consider the N-tray binary batch distillation column (depicted in Figure 1), with reboiler, condenser and accumulator vessel to collect the distillate product. From standard assumptions [5,20] (constant pressure; equilibrium in all trays, equimolal overflows), the batch column dynamics are given by:

$$\dot{c}_i = [L(m_{i+1})\Delta^+ c_i - V\Delta^- E(c_i)]/m_i, \qquad 0 \le i \le N-1$$

$$\dot{c}_N = [R\Delta^+ c_N - V\Delta^- E(c_N)]/m_N, \qquad \dot{c}_{N+1} = V[E(c_N) - c_{N+1}]/m_D$$

$$\dot{m}_i = L(m_{i+1}) - L(m_i), \qquad 1 \le i \le N-1;$$

$$\dot{m}_N = R - L(m_N) \qquad \text{(1a-e)}$$

where:

$\Delta^+ c_i := c_{i+1} - c_i$, $\quad \Delta^- E(c_i) := E(c_i) - E(c_{i-1})$

$E(c_{-1}) := c_0$, $\quad c_{N+1} = c_D$, $\quad c_B = c_0$

$T_I = \beta(c_I)$, $\quad T_{II} = \beta(c_{II})$

$L(m_i) = \hat{R} + (m_i - \hat{m}_i)/\tau_i$

c_i (or m_i) is the light component mole fraction (or holdup) at the i-th tray, E, β and L are the nonlinear liquid-vapor equilibrium, bubble point and hydraulic functions, respectively, T_I, T_{II} are two temperature measurements (at locations to be determined) and $(\hat{\cdot})$ denotes the steady-state value of $(\cdot)$ ***at total reflux***. In compact notation, the column dynamics (1) are written as follows:

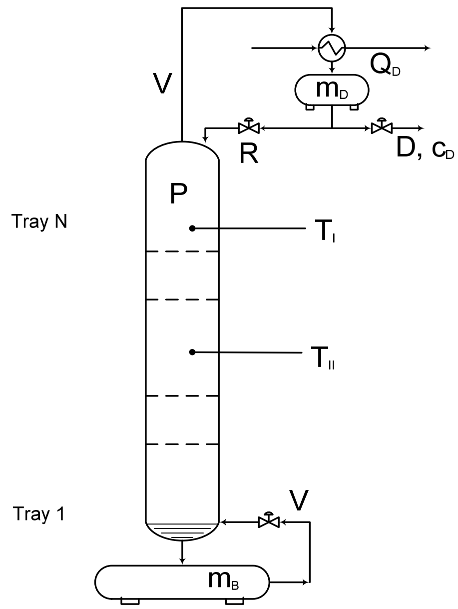

Fig. 1. Batch distillation column

$$\dot{c} = F_c(c, m, R), \qquad \dot{m} = F_m(m, R), \qquad \psi = h(c) \quad \text{(2a-b)}$$

where:

$c = (c_0,..., c_{N+1})'$, $\quad m = (m_1,..., m_N)'$

$\psi = (T_I, T_{II})'$, $\quad h(c) = [\beta(c_I), \beta(c_{II})]'$

For batch distillation operation (i.e., during startup and production periods) the vapor flow rate V is kept constant. During the startup period, the column is operated at total reflux, i.e., R = V [5,20], until time t_s, when the distillate composition $c_D(t)$ reaches the prescribed value $\bar{c}_D$ [11]. Then, the extraction period begins, with distillate product withdrawn at rate V-R, where the product composition must be maintained at the prescribed value $\bar{c}_D$ by manipulating the reflux flow rate R, on the basis of two temperature measurements T_I and T_{II} (whose locations and setpoints are to be determined). Usually this constant distillate purity operation is maintained until a utility (i.e., profit) functional reaches its maximum [11].

Our batch distillation control problem consists in the following: after the initial total reflux operation period (i.e. at time $t = t_s$) is over, the reflux flow rate must be manipulated on the basis of two temperature measurements (T_I, T_{II}) at locations to be determined (see Figure 1), in such a way that the distillate effluent composition (c_D) is maintained at its prescribed value during the production period. The control scheme must be linear, with reduced model-dependency and simple construction and tuning guidelines. The behavior of the proposed temperature driven cascade control scheme must be compared with the ones of its previously studied counterparts based on composition and/or temperature measurements.

III. Control Design

BDC operation requires two operation periods: (i) a Total reflux period, whose duration must be set by an event controller (e.g., when the distillate composition/temperature has reached the desired value), and (ii) an extraction period, with constant distillate purity and time varying reflux ratio. Here we study the extraction period only, given that for binary systems the duration of the total reflux period can be established by measuring the distillate temperature.

From the nonlinear control theory we know that optimal nonlinear SF controllers [21-22]: (i) are inherently robust and passive (i.e., minimum phase with relative degrees equal to one), (ii) cannot in general be constructed in analytic form via direct optimality, (iii) can be constructed in analytic form via inverse optimality by starting with a passive controller and verifying for which objective function the controller is optimal, and (iv) can be constructed, on the basis of non-passive models, with a passivation via backstepping procedure. From previous distillation column [2, 15-16] and chemical reactor [17-18] control studies, we know that: (i) the behaviour of a passive nonlinear SF controller can be recovered with a linear OF controller made of conventional proportional and integral components, and (ii) that a linear-decentralized model for control can be set according to the system relative-degree structure and observability property.

Following the above mentioned ideas, in Section 3.1 (i) a nonlinear passive model is drawn for nonlinear constructive control analysis, and (ii) the corresponding SF controller for constant distillate purity is recalled in Section 3.2, and reinterpreted as a material balance-like controller that can be used to generate a temperature tray policy. Then, In Section 3.3 the nonlinear passive model of Section 3.1 is realized in terms of a single input-two output (SI-2O) model, that is used in Section 3.4 to draw the proposed OF cascade control based on temperature measurements

3.1 Nonlinear passive model

Here, a passive model for control design purposes is drawn. As it is known in distillation column control, the hydraulic dynamics are faster than composition dynamics [2-3, 23], so that they can be assumed in quasi-steady state in the design stage, and their effect must be accounted for in the tuning stage. Then, Eq. (2b) can be set as (3) with liquid flows given by (4):

$$\dot{m} = F_m(m, R) \approx 0 \qquad (3)$$

$$L(m_i) \approx R, \quad 1 \leq i \leq N \qquad (4)$$

The unique root of (4) for a given R is given by:

$$m_i^* = G_i(R) = (R - \hat{R})\tau_i + \hat{m}_i, \quad 1 \leq i \leq i \leq N \qquad (5)$$

where τ_i is the tray hydraulic time constant. Substituting Eq. (4) and (5) in Eq. (1a) yields the *reduced–order passive model*:

$$\dot{c}_0 = [R\,(c_1 - c_0) - V\,\Delta^- E(c_0)]/\bar{m}_B := f_0(c, R) \qquad (6a)$$

$$\dot{c}_i = [R\,\Delta^+ c_i - V\,\Delta^- E(c_i)]/[(R - \hat{R})\tau_h + \hat{m}_i] := f_i(c, R) \quad 1 \leq i \leq N \qquad (6d)$$

$$\dot{c}_D = = V[E(c_N) - c_D] / \bar{m}_D := f_{N+1}(c, R) \qquad (6e)$$

$$\psi_I = \beta(c_I),\ \psi_{II} = \beta(c_{II}) \qquad (6f)$$

This model is used for analysis purposes.

3.2 Nonlinear SF passive controller

For the sake of analysis [2, 24], consider the dynamics of the distillate composition, Eq (1b), and observe that, for constant distillate composition c_D at the desired value $\bar{c}_D$, it is required that $E(c_N) = \bar{c}_D$, which yields the constant value of the N-th tray composition

$$\bar{c}_N = E^{-1}(\bar{c}_D) \qquad (7)$$

Consequently, the regulation of the N-th tray composition (or temperature) around the desired value $\bar{c}_N$ (or $\bar{T}_N$) implies constant distillate purity. Then, the SF control problem for constant distillate purity can be solved in this way, i.e., by using backstepping [22]. The resulting SF control problem resumes to the one of keeping constant the N-th tray composition by regulating the reflux flow rate, and the corresponding control law has been drawn previously [4], and is given by:

$$R = \{[-k_I\,(c_N - \bar{c}_N)(\hat{m}_N - \hat{R}\tau_h)] + [V\,(E(c_N) - E(c_{N-1})]\}/(c_D - c_N) \qquad (8a)$$

that results from the enforcement of the stable first order dynamics for the composition error regulation

$$\dot{c}_N = -\,k_I\,(c_N - \bar{c}_N) \qquad (8b)$$

where k_I is a positive control gain. The corresponding zero dynamics controller is given by

$$R_\zeta = [V\,(E(\bar{c}_N) - E(c_{N-1})]/(c_D - \bar{c}_N) \qquad (9)$$

that can be reinterpreted as a material balance controller [11, 25], i.e., the controller input R_ζ exactly compensates the material balance so that the N-th tray composition is kept constant. Observe that the term $[E(\bar{c}_N) - E(c_{N-1})]/(c_D - \bar{c}_N)$ is the time varying version of an operating line slope in a continuous column McCabe-Thiele diagram; here, this term must be different from zero, which can be guaranteed after a total reflux period. Additionally, the column (6) under the zero dynamics controller (9) and the restriction (7) yield the system zero dynamics that are assumed stable [11].

The preceding nonlinear passive controller is robust, but requires the states of the reduced system (6). This controller is to be reinterpreted in the following sections as a linear decentralized cascade controller driven by two temperature measurements.

3.3 Linear passive model

To simplify the control design, first consider the following coordinate change

$$x_I = \beta(c_I), \qquad x_{II} = \beta(c_{II}) \qquad (10)$$

and rewrite the preceding reduced passive model (6) as follows:

$$\dot{x}_I = g_I(x_I, x_{II}, x_z, R), \qquad y_I = x_I \tag{11a-b}$$

$$\dot{x}_{II} = g_{II}(x_I, x_{II}, x_z, R), \qquad y_{II} = x_{II} \tag{11c-d}$$

$$\dot{x}_z = f_z(x_I, x_{II}, x_z, R) \tag{11e}$$

where

$g_I(x_I, x_{II}, x_z, R) = \beta'(c_I)f_I(c, R)$
$g_{II}(x_I, x_{II}, x_z, R) = \beta'(c_{II})f_{II}(c, R)$
$x_z = (c_1, c_2,\ldots, c_{II-1}, c_{II+1},\ldots, ,\ldots, c_{N-1}, c_D)$
$f_z = (f_1, f_2,\ldots, f_{II-1}, f_{II+1},\ldots, , f_{N-1}, f_D)'$

x_I and x_{II} are the two temperature at the measurement trays after a bubble point function-based coordinate change, and x_z are the remaining deviated compositions, and R is the reflux flow rate control input.

Observe that the model (6) or (11) has relative degrees (RD´s) equal to one for both inputs and any choice of measured temperatures, excepting the distillate. Assuming that the CL column forces a unique material balance [11], then the zero dynamics of the resulting system is stable. Thus, system (6) or (7) is *passive*, implying that related robust nonlinear SF control problem is solvable.

3.4 Linear-decentralized model for OF Control

Next, a linear-decentralized model with reconstructible load inputs is set for OF control design purposes. Following previous developments in two point temperature and composition-temperature cascade control designs [13, 15-16, 26], on the basis of the preceding RD structure and the linearity-decentralization feature specifications for our OF control design, rewrite the passive model (11) as follows:

$$\dot{x}_I = a_I R + b_I, \qquad b_I = \varphi_I(x_I, x_{II}, x_z, R), \qquad y_I = x_I \tag{12a-c}$$

$$\dot{x}_{II} = a_{II} R + b_{II}, \qquad b_{II} = \varphi_{II}(x_I, x_{II}, x_z, R), \qquad y_{II} = x_{II} \tag{12d-f}$$

$$\dot{x}_z = f_z(x_I, x_{II}, x_z, R) \tag{12g}$$

where $\Delta^+\tilde{T}_k$ is an average temperature gradient at the k-th stage during the extraction period, and

$$\varphi_I(x_I, x_{II}, x_z, R) = f_I(x_I, x_{II}, u, d) - a_I R, \qquad a_I = (\Delta^+\tilde{T}_I/\tilde{m}_I)$$

$$\varphi_{II}(x_I, x_{II}, x_z, R) = f_{II}(x_I, x_{II}, u, d) - a_{II} R, \qquad a_{II} = (\Delta^+\tilde{T}_2^s/\tilde{m}_{II})$$

The inputs (b_I, b_{II}) satisfy the matching condition [21], as they enter in the same channels as the control inputs. Since the temperature states x_I and x_{II} are measured, the load disturbances (b_I, b_{II}) are instantaneously observable [27], as they can be reconstructed from the inputs and the measured output derivatives, according to the expressions:

$$b_I = \dot{y}_I - a_I R, \qquad b_{II} = \dot{y}_{II} - a_{II} R, \tag{13}$$

These load inputs can be quickly reconstructed with linear-decentralized reduced-order observers, one per measurement-load pair. Therefore, the dynamics (12g) is not necessary. Accordingly, our *model for OF control design* is given by:

$$\dot{x}_I = a_I R + b_I, \qquad \dot{b}_I \approx 0, \qquad y_I = x_I \tag{14a}$$

$$\dot{x}_{II} = a_{II} R + b_{II}, \qquad \dot{b}_{II} \approx 0, \qquad y_{II} = x_{II} \tag{14b}$$

where (b_I, b_{II}) are unknown-reconstructible load inputs.

3.5 Linear-decentralized OF cascade Control

To build the controller, let us consider the dynamics of the primary temperature T_I (i.e., the one associated to the N-th tray). For this aim, impose the closed-loop first order regulation dynamics (15) ($k_I > 0$ is the primary temperature loop gain) in (14a), and solve for R to obtain the "virtual" controller (16), which in turn is applied to the dynamics (14b) to obtain the rectifying section secondary tray temperature setpoint generator, Eq. (17):

$$\dot{T}_I = - k_I (T_I - \bar{T}_I), \qquad k_I > 0 \tag{15}$$

$$R^* = [- k_I(T_I - \bar{T}_I) - b_I]/a_I \tag{16}$$

$$\dot{T}_{II}^* = - (a_{II}/a_I) [k_I (T_I - \bar{T}_I) + b_I] + b_{II}, \qquad T_{II}^*(0) = T_{IIo}^* \tag{17}$$

Now, let us recall (14b), enforce the first order temperature dynamics (18) (k_{II} is the secondary temperature loop gain):

$$\dot{T}_{II} - \dot{T}_{II}^* = - k_{II}(T_{II} - T_{II}^*), \qquad k_{II} > 0 \tag{18}$$

and solve for R to obtain the controller:

$$R = [\dot{T}_{II}^* - k_{II}(T_{II} - T_{II}^*) - b_{II}]/a_{II} \tag{19}$$

The combination of these components with a pair of first order filters (20a-d) to estimate the loads (b_I, b_{II}) yields the temperature driven BDC cascade controller in internal model control (IMC) form [28]:

$$\dot{w}_I = - \omega_o(w_I + \omega_o T_I + a_I R), \qquad \hat{b}_I = w_I + \omega_o T_I \tag{20a-b}$$

$$\dot{w}_{II} = - \omega_o(w_{II} + \omega_o T_{II} + a_{II} R), \qquad \hat{b}_{II} = w_{II} + \omega_o T_{II} \tag{20c-d}$$

$$\dot{T}_{II}^* = - (a_{II}/a_I) [k_I (T_I - \bar{T}_I) + \hat{b}_I] + \hat{b}_{II}, \qquad T_{II}^*(0) = T_{IIo}^* \tag{20e}$$

$$R = [\dot{T}_{II}^* - k_{II}(T_{II} - T_{II}^*) - \hat{b}_{II}]/a_{II} \tag{20f}$$

where ω_o is an observer gain.

Observe that the generated temperature setpoint T_{II}^* given by Eq (20e) is time-varying, while the setpoint $\bar{T}_I$ for the primary loop (15) is constant. In this way, the distillate purity can be kept constant, because the computation of the time varying secondary loop temperature policy and the resulting reflux flow rate can be performed on-line, implying that the proposed controller has reduced model dependency when compared to the previous nonlinear observer based SF controllers.

3.6 Controller implementation

1. The scheme is based only on temperature measurements, i.e., it is not affected by measurement delays nor dead times. Consequently, the filter-based estimation of the load

disturbances ($\hat{b}_I$, $\hat{b}_{II}$) will be limited only by the high frequency holdup dynamics, and the measurement noise.

2. The calculation of the virtual controller R^* requires the inverse of the coefficient a_I. In high-purity columns, this can be a very large number, and consequently, this control structure can be prone to amplify the measurement noise. For this reason, the signal to noise ratio should be characterized, and the controller (k_I, k_{II}) and observer (ω_o) gains must be detuned accordingly.

3. The controller (5-20) are expressed in the IMC form [28] and consequently, it is equipped naturally with an anti-windup control scheme, i.e., they can tolerate actuator saturations without significant performance degradation.

4. The proposed technique requires a secondary temperature sensor. We recommend to locate this sensor after performing some simulation work, and observing the tray-to-tray temperature gradients of the column during the production period.

5. In the previous developments, this gain ω_o is chosen equal for all the filters. In addition, this assumption can be relaxed to assign different values to each filter's gain.

6. The controller and filter tuning can be executed in a systematic manner by following the same arguments used in the tuning of the temperature-to-temperature controller, described in references [15-16].

4. APPLICATION EXAMPLE

The studied BDC separates a methanol-water mixture. There are N = 8 trays, the initial load is m_{Bo} = 12 Kmol at composition c_F = 0.15, the vapor flow is V = 5400 mol/h, the tray, hydraulic parameter set is ($1/\tau_m$, a, b) = (1000 1/h, 5400, 30), the condenser holdup is m_D = 250 mol, and the nominal product composition is $\bar{c}_D$ = 0.985.

Total reflux period. At the beginning of the batch operation, total reflux policy is required. The switching time (i.e., the duration of the total reflux period) was chosen as follows: when the distillate temperature (i.e., composition) T_D reached the desired value, i.e., $T_D = \beta(\bar{c}_N)$. In this case, the switching time was 22.5 min. After that, the cascade controller Eq (20) was applied.

Proposed Cascade Control structure. As stated previously, the primary temperature measurement corresponds to the one in the N-th tray. To place the secondary temperature measurement, some simulation work was required, in which the SF controller (8) was applied to the BDC after the total reflux period. The resulting temperature profiles for each tray are depicted in Fig. 2A, and the corresponding temperature gradients are shown in Fig. 2B. It can be seen from Fig. 2B that the trays with the maximum gradients over the batch column operation are trays 2, 3 and 4. Here we chose tray 4 as the secondary measurement for the cascade control scheme.

Controller tuning. Following the tuning guidelines given in Section 3.6, the following filter and controller gains were used: (ω_o, k_{II}, k_I) = (120, 40, 20) 1/hr.

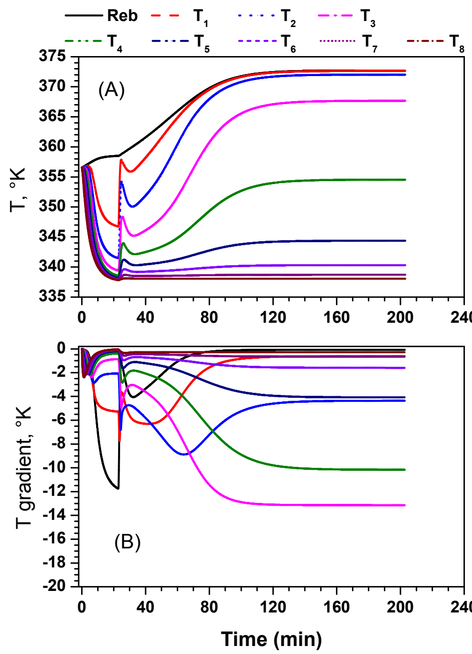

Fig. 2. (A) Temperature profiles and (B) Temperature gradients for the BDC with the SF controller (8)

Control behavior. The application of the total reflux period until t = 22.5 min and the latter application of the proposed cascade temperature-temperature controller Eq (20) is depicted in Fig. 3A and 3B. To illustrate the advantages of the proposed TT cascade controller over conventional cascade schemes, the BDC was also tested with a previous cascade composition-to-temperature controller [e.g., see ref. 13] based on the N-th tray composition measurement, that is affected by 1 min dead time. The controller gain set for this controller is (ω_o, k_{II}, k_I) = (30, 20, 15) 1/hr. Because of the presence of dead time, in this case [13] the controller requires significant detuning to attenuate the oscillatory response The regulation of the distillate composition is depicted in Fig, 3A, showing that: at the beginning of the production period (between t= 22.5 and t =45 min) both controllers yield an overshoot in the distillate purity, caused by the sudden operation switching (see Fig 3B, where the reflux flow rate changes from the saturation value 5.4 kmol/hr to lower values) and by the time required for the observer to converge. For the proposed temperature measurement -driven cascade controller, at t=45 min (i.e., ≈ 23 min after the switching time), the distillate composition is properly regulated around the desired constant value 98.5 mol%, and a reasonable control effort (Fig 3B). The conventional cascade CT controller requires a rather oscillatory control effort (Fig. 3B) at the beginning of the production period, that translates in an overpurification of the distillate (Fig. 3A).

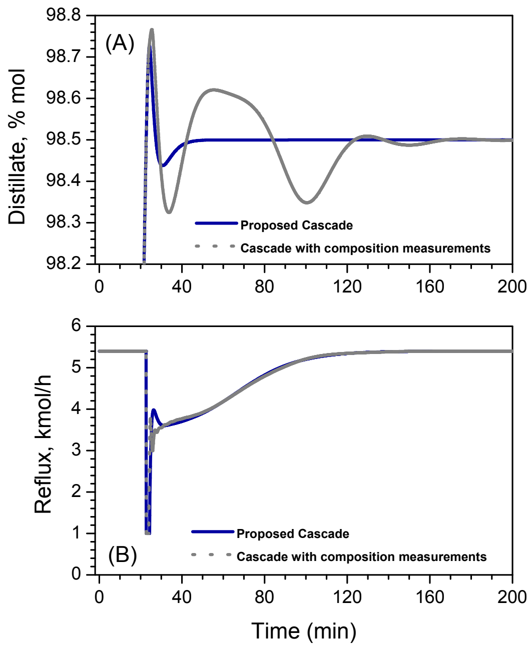

Fig. 3. CL behavior of the BDC: (A) Distillate (B) Control effort

5. CONCLUSIONS

The problem of designing a cascade temperature-to-temperature driven robust controller for binary batch distillation columns has been addressed, by combining optimality, passivity and detectability notions and tools with conventional cascade control tools. The approach recalls the previously established solvability conditions for the nonlinear SF control problem for the regulation of the N-th tray composition, and inherits the basic feature of yielding constant distillate product purity, translated in the regulation of the N-th tray temperature by means of a cascade control scheme driven by temperature measurements. The rather simple resulting controller requires few column parameters and can be implemented with the usual PLC software. The proposed approach was successfully applied to an 8-tray methanol –water BDC, finding that the proposed control scheme effectively yields the desired behavior, i.e., constant distillate purity, matching the behavior of its more complex observer based nonlinear counterparts, and outperforming conventional CT cascade control schemes based in possibly delayed composition measurements. The extension of this methodology for multicomponent mixtures is a matter of future research.

REFERENCES

[1] Humphrey, J. L., Seibert, A. F., Koort, R. A. *Separation Technologies-Advances and Priorities*. OE Contract AC07-901D12920, Feb. 1991.

[2] Castellanos-Sahagún, E., Alvarez, J. (2006). *Chem. Eng. Comm*. Vol. 193, pp. 1-27 (2006).

[3] Skogestad, S. Dynamics and control of distillation columns: A critical survey. Mod. Ident. & Cont. 18(3), pp. 177-217 (1997).

[4] Diwekar, U. Batch Distillation: Simulation, Optimal Design and Control. Taylor & Francis, 1995(Bristol, PA).

[5] Muhrer, C. A. Luyben, W. (1992). Batch Distillation. In W. L. Luyben (Ed.), Practical Distillation Control, pp. 508-528. NY, Van Nostrand Reinhold.

[6] Mujtaba, I. M., Macchieto, S. (1996). J. Proc. Cont. 6(1), pp. 27-36.

[7] Finefrock, Q. B., Bosley, J. R. Edgar, T. (1994). Gain scheduled PID control of batch distillation to overcome changing system dynamics. AIChE Annual Mtng, San Francisco, 1994.

[8] Bosley, J. R. Edgar, T. (1992). Application of nonlinear model predictive control to optimal batch distillation. 3rd IFAC Symposium DYCORD'92, pp. 271-276.

[9] Barolo, M., Berto, F. (1998). Ind. Eng. Chem. Res., 37, pp. 4689-4698.

[10] Monroy-Loperena, R., Alvarez-Ramírez, J. Chem. Eng. Sci., Vol. 58, pp. 4729-4737 (2003).

[11] Alvarez, J., Castellanos-Sahagún, E., Fernández, C., Aguirre, S. Optimal Closed-Loop Operation of Binary Batch Distillation Columns IFAC World Congress, Prague (2005).

[12] Alvarez-Ramírez, J., Monroy-Loperena, R., Alvarez, J. *AIChE J*. 48 (8), pp. 1705-1718 (2002).

[13] Castellanos-Sahagún, E, Alvarez-Ramírez J., Alvarez, J. *Ind. Eng. Chem. Res*. Vol. 45, 9010-9023 (2006).

[14] Monroy-Loperena, R., Alvarez-Ramírez, J. AIChE J, Vol. 50(9), pp 2113-2129 (2004).

[15] Castellanos-Sahagún, E., Alvarez, J. Dual Effluent Composition Control of Binary Distillation Columns with a Four Temperature Measurement-Based Control Scheme Proc. Symp. On Advanced Control of Industrial Process *ADCONIP 2008, Jasper, Canada (2008).*

[16] Castellanos-Sahagún, E., Alvarez, J., Frau, A., Baratti, R.. Temperature control of binary distillation columns. Submitted to I&ECR (2012).

[17] Alvarez, J., González, P. *J. Proc. Cont*. Vol. 18, pp. 896-905 (2008).

[18] Alvarez, J., Zaldo, F., Oaxaca, G. (2004b), "Towards a joint process and control design for batch processes: application to semibatch polymer reactors". In: Integration of Process and Control (Ed. Georgiadis, M. Seferlis, P.), Elsevier.

[19] Alvarez, J., Castellanos-Sahagún, E., Fernández, C., Aguirre, S. (2004a). Joint operation and control design for binary batch distillation columns. Symp. on Knowledge Driven Batch Processes, BATCHPRO 2004, Greece, pp. 211-218.

[20] Luyben, W. L. *Process Modeling, Simulation and Control for Chemical Engineers, 2nd Ed*. McGraw-Hill, Singapore (1990).

[21] Freeman, R. Kokotović, P. *Robust Nonlinear Control Design*, Birkhauser (1996).

[22] Sepulchre, R., Jankovic, M, Kokotović. *Constructive Nonlinear Control*. Springer-Verlag (1997).

[23] Levy, R. E., Foss, A. S., Grens II, E. A. Ind. *Eng. Chem. Fundam*., Vol. 8, No. 4, pp. 765-776 (1969).

[24] Castro, R., Alvarez, Ja., Alvarez, Jo. *Automatica* Vol. 26, pp. 567-572 (1990).

[25] Shinskey, F. G., Process Control Systems, 2nd Ed. McGraw-Hill, 1977.

[26] Castellanos-Sahagún, E., Alvarez, J. (2005a). Ind. Eng. Chem. Res. (44), pp. 142-152.

[27] Hermann, R., Krener, A. J. *IEEE Trans. Autom. Contr.* Vol. AC-22, pp. 728-740 (1977).

[28] Kothare, M. V., Campo, P. J., Morari, M., Nett, C. N. A unified framework for the study of anti-windup designs. Automatica, Vol. 30, No. 12, pp. 1869-1883 (1994).