Preprints of the 19th World Congress
The International Federation of Automatic Control
Cape Town, South Africa. August 24-29, 2014

Feedforward-feedback control of an industrial multicomponent distillation column

Marcella Porru*, Roberto Baratti*
Jesus Alvarez **

*Dip. Ingegneria Meccanica, Chimica e dei Materiali,
via Marengo, 2 09123 Cagliari, Italy.
** Dep. de Ingeniería de Procesos e Hidráulica,
UAM, Apdo. 55534, 09340 México, D.F. Mexico

Abstract: The problem of regulating the distillate impurity content in a two-feed (paraffin-olefin) industrial multicomponent distillation column (iso-normal butane splitter) by manipulating the heat duty according to olefin-to-total feedflow load disturbance and tray temperature measurements is addressed. First, the detailed model-based nonlinear state-feedback robust control problem is addressed, identifying the solvability conditions and attainable closed-loop behavior. Then, the behavior of this controller is recovered with an output feedback PI temperature controller with dynamic feedforward setpoint compensation and antiwindup protection. The proposed approach is illustrated and tested through numerical simulations with a detailed model calibrated with industrial data, establishing the feasibility of improving the existing control scheme with PI temperature control plus manual setpoint compensation.

Keywords: Distillation column control, PI control, feedforward control, inventory control, passive control, robust control.

1. INTRODUCTION

The aim of the present study is to assess the feasibility of improving the closed-loop behavior of a two-stream feed industrial heptacomponent distillation column (butane splitter) which is: (i) subjected to significant feed flow load disturbances, and (ii) controlled with a PI temperature controller plus manual setpoint adjustment. The idea is to replace the manual setpoint compensator with a feedforward setpoint compensator, and to retune the PI temperaure controller.

Due to their strong nonlinearity and interaction, the intensive energy consuming columns, with multivalued relationship between regulated composition and temperature measurement, are controlled by manipulating the vapor (or reflux) flow rate on the basis of a single-point temperature PI control loop with manual or feedforward setpoint adjustment (Shinskey, 1977; Skogestad, 1997; Fruehauf and Mahoney, 1994). In the case of significant load disturbances with manual setpoint adjustment, like the butane splitter, improvements in composition effluent regulation capability and energy efficiency can be attained with feedforward control (Shinskey, 1988).

From an industrial control practice perspective, the most effective way to control a processes susceptible to load disturbances is the combination of feedforward (FF) with a output feedback (OF) control (Shinskey, 1977). FF performs most of the disturbance rejection on the basis of a reversed model of the process, and OF achieves regulation by compensating the modeling error of the FF. According to control theory, such reversed model is the dynamical inverse of the process (Hirschorn, 1979), or equivalenlty, its zero-dynamics (Isidori, 1989). The associated process passivity is key condition for robustness (Castellanos-Sahagún et al., 2005), and the design, in the light of passivity and observability properties, of a simplified model for control design is a fundamental issue in the development of application-oriented OF controllers (Castellanos-Sahagún et al., 2005; Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013). These considerations motivate the present study within a constructive control framework (Sepulchre et al., 1977), by blending industrial and advanced control notions and tools in the light of the process characteristics.

Technically speaking, our problem consists in designing a feedforward-output feedback (FF-OF) control scheme to regulate (with quick response and reduced offset) the unmeasured distillate NC4 impurity by manipulating the heat injection rate according to feed stream flow ratio load disturbance and sensitive tray measurements. A central concern is the development of an application-oriented reliable control scheme, as simple as possible in terms of nonlinearity, coupling, model dependency, implementation, and tuning.

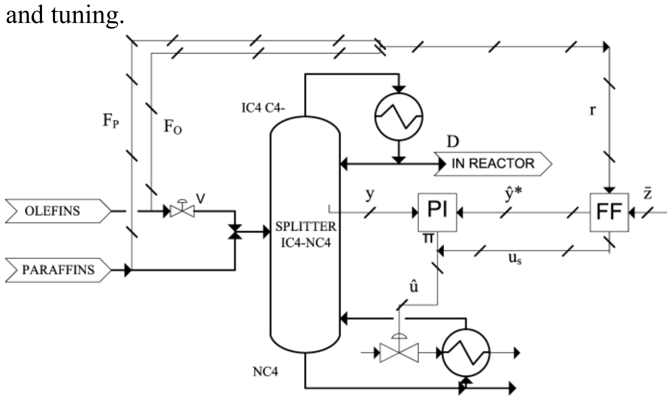

Fig. 1 Column with proposed FF-OF control scheme.

2. CONTROL PROBLEM

Consider the industrial heptacomponent ($n_c = 7$) distillation column (depicted in Fig. 1) located at SARLUX refinery (Sarroch, Italy), where iso-butane (IC4) and normal-butane (NC4) splitting occurs. The total feed (with molar flow rate *F*) is the sum of paraffin and olefin streams, with molar flow rates F_p and F_o and nominal composition $\bar{x}_P$ and $\bar{x}_O$, respectively. Significant changes in the olefin-to-total feed ratio *r* occur,

$$0 \le r = F_o/F \le 0.5, \quad F = F_o + F_p \tag{1}$$

When $r = 0$ the feed is said to be *saturated* (with negligible amount of olefins), and when $r > 0$ the feed is called *mixed* (the olefin content is significant). In Table 1 are listed the typical saturated and mixed feed concentrations as well as the normal boiling points of the associated components. The changes in the total feed (*F*) and distillate (*D*) flow rates, through the heat injection duty (*Q*) are comparatively milder. Since the distillate is fed to a subsequent alkylation reactor to produce high-octane gasoline, it must contain mostly IC4 accompanied by a small amount of NC4. The column has N = 57 stages, three kettle reboilers (stage 1), three total condensers (stage 57), temperature measurement at stage $n_m = 49$, and feed at stage $n_f = 33$-th. In Fig. 2 are presented typical feed stream and distillate flows over a 100 h period, (provided by SARLUX refinery), and the corresponding feed ration r (1) is plotted in Fig. 3.

Table 1 Nominal hydrocarbon compositions in the saturated and mixed feeds and component normal boiling points

HIDROCARBONS	SATURATED FEED concentration [molfrac]	MIXED FEED concentration [molfrac]	NORMAL BOILING POINT [K]
PROPANE	0.025	0.008	231.1
I-BUTANE	0.400	0.394	261.4
I-BUTENE	0.003	0.032	266.2
N-BUTENE	0.001	0.031	266.9
N-BUTANE	0.569	0.467	272.7
2-BUTENE TRANS	0.001	0.039	274.0
2-BUTENE CIS	0.001	0.029	276.9

The NC4 distillate composition ($z = c_N^{NC4}$) was maintained within a certain band by manipulating the heat duty (Q) through an automatic PI temperature loop (with temperature measurement $y = T_{n_m}$ at stage n_m) with manual setpoint adjustment based on the feed and distillate load disturbances (r, F, and D). The aim of the present study is to assess the feasibility of improving the control scheme by means of: (i) the systematization through FF control of the setpoint compensation procedure, (ii) the verification of the sensor location of the PI temperature controller, and (iii) the revision of the PI control gains.

From standard assumptions (material balances, energy balance neglected on each tray, feed variation due to feed and reflux subcoolings only, tight controller and condenser level control, and ideal vapor-liquid equilibrium) the N-stage n_c-component column dynamics are described by the n-dimensional open-loop dynamical column system (Luyben, 1990; Baratti et al., 1998)

$$\dot{\boldsymbol{x}} = \boldsymbol{f}(\boldsymbol{x}, u, d, \boldsymbol{w}), \ \boldsymbol{x}(0) = \boldsymbol{x}_o, \ y = h(\boldsymbol{x}), \ z = C\boldsymbol{x} \tag{2}$$

where

$$dim\, \boldsymbol{x} = n = N(n_C - 1) = 392, \ z = c_N^{NC4}$$
$$y = T_{n_m}, \ u = Q, \ d = r, \ \boldsymbol{w} = (F, D)'$$

x is the n-composition state, z is the regulated output (distillate NC4 composition c_N^{NC4}), y is the measured output (temperature T_{n_m} at stage n_m), u is the manipulated input, d is the significant load disturbance, and $\boldsymbol{w}$ (total feed-distillate flowrate pair) is the comparatively milder load disturbance. It must be pointed out that: (i) system (2) (run with Aspen simulation package), that will be referred to as *detailed model*, will be used for control scheme development, analysis, and testing, and (ii) a *simplified model* will be designed to construct the FF-OF controller.

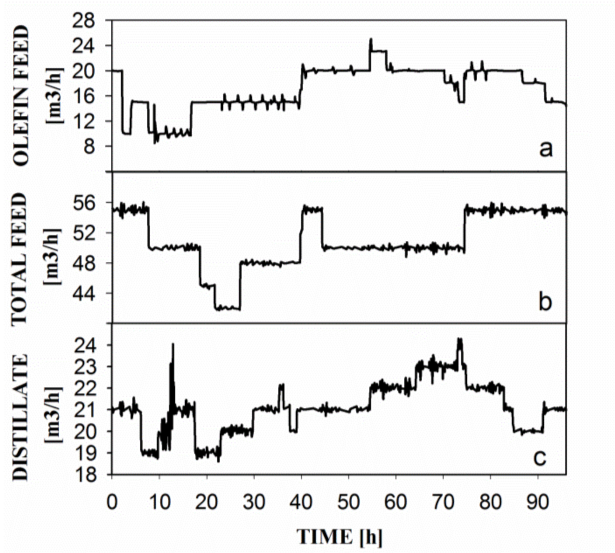

Fig. 2 Olefin feed (F_o), total feed (F), and distillate (D) molar flowrates of the industrial column.

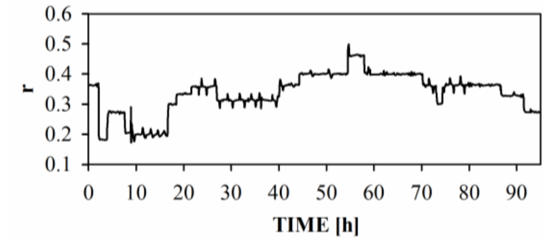

Fig. 3 Olefin-to-total feed ratio r (1).

Our *problem* consists in designing a *FF-OF controller* for the distillation column (2). The controller must regulate (within an admissible tolerance about a prescribed value $\bar{z}$) the unmeasured output z (NC4 distillate concentration) by manipulating the control input *u* (heat injection rate *Q*) on the basis of the load input d (feed ratio) and output y (temperature) measurements. We are interested in drawing an application-oriented reliable control scheme as simple (linear and dynamically decoupled) and model independent as possible, and with simple (conventional-like) tuning guidelines underlain by robust closed-loop stability conditions.

3. FEEDFORWARD STATE-FEEDBACK CONTROL

In this section the detailed model-based nonlinear (NL) state-feedback (SF) robust control problem of the column (2) is addressed. The purposes are: (i) the setting of the methodological point of departure for the development (in the next section) of the FF-OF control scheme, (ii) the identification of solvability conditions, and (iii) the assessment of the attainable closed-loop behavior.

3.1 FF composition regulation controller

The aim of the FF composition controller is to keep the distillate NC4 composition z at its prescribed setpoint $\bar{z}$,

$$z(t) = \bar{z} \tag{3}$$

by adjusting the heat injection rate u(t) against the value of the feed ratio load disturbance $d(t)$. According to industrial control practitioners (Shinskey, 1988) the corresponding FF controller is a the "reversed model" of the process which results from solving for the input $u(t)$ the dynamic model written in terms of $\bar{z}$ and $d(t)$. For this aim, set the secondary load disturbance $\boldsymbol{w}$ as its nominal value $\bar{\boldsymbol{w}}$, rewrite the column model (2) in the partitioned form

$$\dot{\boldsymbol{c}}_\zeta = \boldsymbol{f}_\zeta(\boldsymbol{c}_\zeta, c_z, d, u), \boldsymbol{c}_\zeta(0) = \boldsymbol{c}_{\zeta o}\,, \quad \boldsymbol{x} = \boldsymbol{I}_\zeta(\boldsymbol{c}_\zeta^T, c_z)^T \tag{4a}$$

$$\dot{c}_z = f_z(\boldsymbol{c}_\zeta, c_z, u), \quad c_z(0) = c_{zo}, \quad z = c_z := c_N^{NC4} \tag{4b}$$

and enforce condition (3) upon this system to draw the *dynamical inverse* (Hirschorn, 1979), or equivalently, the *NL FF dynamic composition controller*

$$\dot{\boldsymbol{c}}_\zeta^* = \boldsymbol{f}_\zeta^*[\boldsymbol{c}_\zeta^*, \bar{z}, d(t)],\ \boldsymbol{c}_\zeta^*(0) = \boldsymbol{c}_{\zeta o};\ u^* = \mu_z(\boldsymbol{c}_\zeta^*, \bar{z}) \tag{5a, b}$$

where

$$f_z\text{: u-monotonic} \tag{6}$$

$$\boldsymbol{f}_\zeta^*(\boldsymbol{c}_\zeta^*, \bar{z}, \boldsymbol{d}) = \boldsymbol{f}_\zeta[\boldsymbol{c}_\zeta^*, \bar{z}, d, \mu_z(\boldsymbol{c}_\zeta^*, \bar{z})]$$

$$\boldsymbol{c}_\zeta^*(t) = \boldsymbol{\tau}_\zeta^*[t, 0, c_{\zeta o}, \bar{z}, d(\cdot)],\ u^*(t) = \mu_z[\boldsymbol{c}_\zeta^*(t), \bar{z}], \tag{7a, b}$$

$\boldsymbol{c}_\zeta^*(t)$ is the *robustly state solution motion* of system (5), $u^*(t)$ is the associated control, and $\mu_z(\boldsymbol{c}_\zeta^*, \bar{z})$ denotes the unique solution for u of eq. (4b) with $c_z = \bar{z}$. Condition (6) (i.e., f_z depends monotonically on u) is the related solvability condition. This condition is met because f_z (4b) is u-antitonic.

The application of the FF controller (5) to the column (4) yields closed-loop robustly stable state motions and output trajectories, according to the expressions

$$c_{zo} = \bar{z},\ u(t) = u^*(t) \Rightarrow z(t) = \bar{z},\ \ \boldsymbol{c}_\zeta(t) = \boldsymbol{c}_\zeta^*(t) \tag{8a}$$

$$c_{zo} \approx \bar{z},\ u(t) = u^*(t) \Rightarrow z(t) \overset{\lambda_*}{\to} \bar{z},\ \ \boldsymbol{c}_\zeta^*(t) \overset{\lambda_*}{\to} \boldsymbol{c}_\zeta(t) \tag{8b}$$

Eq. (8a) says that when the column output composition is initially at its setpoint $\bar{z}$ the output remains at $\bar{z}$ and the FF controller ($\boldsymbol{c}_\zeta^*$) and column ($\boldsymbol{c}_\zeta$) states coincide. Eq. (8b) says that when the column output is initially deviated from its setpoint the output (or control state) reaches asymptotically, with exponential rate λ_*, the output setpoint (or the column state).

The FF controller (5) is an inventory one (Shinskey, 1988), in the sense that the output z is kept constant by exactly balancing the heat u delivered to the column against the demand of the load d. The *solvability* of the FF control problem is ensured by the column *passivity* for the input-output pair (*u, z*) (Sepulchre et al., 1977; Luyben, 1990; Isidori, 1989): (i) the column has relative degree equal to one (*RD = 1*) for (*u, z*), and (ii) the corresponding zero-dynamics (5a) have robustly stable state motions (8) (Wolfgang, 1967; Sontag, 2001).

3.2 SF temperature tracking controller

Here the control task is to allow the temperature y to track the setpoint $y^*(t)$ generated by the static maps

$$y^* = \beta(\boldsymbol{c}_\zeta^*),\ \dot{y}^* := v^* = f_y(\boldsymbol{c}_\zeta^*, u^*) \tag{9a,b}$$

$$f_y(\boldsymbol{c}_\zeta, u) = \partial_{\boldsymbol{c}_\zeta}\beta(\boldsymbol{c}_\zeta)\boldsymbol{f}_{n_m}(\boldsymbol{c}_\zeta, u)$$

of the NL FF composition controller (5), according to the prescribed closed-loop linear dynamics

$$\dot{e}_y = -ke_y,\ e_y = y - y^*(t),\ k > \lambda_* \tag{10}$$

with adjustable gain k. The enforcement of these dynamics upon the column (4) yields the equation

$$f_y(\boldsymbol{c}_\zeta^*, u^*) = v^* - ke_y,\ f_y\text{: u-monotonic} \tag{11a,b}$$

and the solution for u^* of this equation yields the *NL SF temperature tracking controller*

$$u = \mu_y(\boldsymbol{c}_y, y^*, v^*) \tag{12}$$

The corresponding solvability condition (11b) [the column (2) must have relative *RD = 1* for $(u,\ y)$] is met if the temperature measurement is located at a sufficiently sensitive tray (Luyben, 2006).

3.3 FF-SF cascade composition regulation controller

The concatenation of the FF composition controller (5), the setpoint generator (9b), and the NLSF controller (11) yields the NL *FF-SF composition cascade dynamic controller*

$$\dot{\boldsymbol{c}}_\zeta^* = \boldsymbol{f}_\zeta^*[\boldsymbol{c}_\zeta^*, \bar{z}, d(t)],\ \boldsymbol{c}_\zeta^*(0) = \boldsymbol{c}_{\zeta o}^* \approx \boldsymbol{c}_{\zeta o} \quad \text{(primary FF)} \tag{13a}$$

$$u^* = \mu_z(\boldsymbol{c}_\zeta^*, \bar{z}),\quad y^* = \beta(\boldsymbol{c}_\zeta^*),\ v^* = f_y(\boldsymbol{c}_\zeta^*, u^*) \tag{13b-d}$$

$$u = \mu_y(\boldsymbol{c}_y, y^*, v^*) \quad \text{(secondary SF)} \tag{13e}$$

with closed-loop robustly stable state motions and output convergent trajectories, according to the expressions

$$c_{zo} = \bar{z} \Rightarrow (\boldsymbol{c}_\zeta, u, y)(t) = (\boldsymbol{c}_\zeta^*, u^*, y^*)(t), z(t) = \boldsymbol{c}_z^* = \bar{z} \tag{14a}$$

$$c_{zo} \approx \bar{z} \Rightarrow (\boldsymbol{c}_\zeta, u, y)(t) \overset{k}{\to} (\boldsymbol{c}_\zeta^*, u^*, y^*)(t),\ z(t) \overset{k}{\to} \boldsymbol{c}_z^* \overset{\lambda_*}{\to} \bar{z} \tag{14b}$$

The NL FF-SF cascade controller (13): (i) has better disturbance rejection capability (quicker response and smaller composition offset) in the presence of (initial condition, function pair f-β, and measurement-actuation) modeling errors than its FF counterpart (5), (ii) represents the behavior (14) attainable with any robust FF-SF controller, and (iii) constitutes the behavior recovery target for the FF-OF control design problem addressed in the next section.

4. FEED-FORWARD (FF) OUTPUT-FEEDBACK (OF) CONTROLLER

Here the behavior (14) of the detailed model-based NL FF-SF cascade composition controller (13) is recovered with a dynamic FF-OF controller built with a simplified model tailored according to: (i) the passivity property of the detailed model (2), and (ii) the observability property of the simplified model.

4.1 OF temperature tracking controller

Following previous reactor (Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013) and distillation column (Castellanos-Sahagún et al., 2005) constructive control studies, here the behavior of the NL SF tracking temperature controller (13e) is recovered (up to observer convergence) using a PI temperature controller with antiwindup protection. For this aim, write the measured output derivative in the form

$$\dot{y} = -au + \iota, \quad \iota = f_y(\boldsymbol{c}_\zeta, u) + au, \tag{15a,b}$$

$$a \approx \partial(f_y(\bar{\boldsymbol{c}}_\zeta, \bar{u}))/\partial u > 0 \tag{15c}$$

with *RD = 1* with respect to (u, y) and (u, ι), and input ι observable from the measured signal-pair (y, u). The enforcement of the prescribed closed-loop tracking dynamics (10) upon system (15a) yields the next realization (in terms of the observable input ι) of the NL SF controller (12):

$$u = [-v^* + k(y - y^*) + \iota]/a \tag{16}$$

By virtue of its observability property, the input ι can be quickly reconstructed (with adjustable speed ω, up to measurement noise) by the reduced-order observer (Diaz-Salgado et al., 2011) (17a). The combination of observer (17a) with the SF controller (16) yields the dynamic-linear *OF tracking temperature controller* in IMC form (Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013; Alvarez-Ramírez et al., 2002)

$$\dot{\chi} = -\omega\chi - \omega(\omega y - au), \quad \hat{\iota} = \chi + \omega y \tag{17a}$$

$$\hat{u} = [-v^* + k(y - y^*) + \hat{\iota}]/a \tag{17b}$$

In PI form this controller is written as follows (Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013)

$$\hat{u} = (-v^*/a) + \pi(y - y^*) \tag{18a}$$

where

$$\pi(e_y) = k_y[e_y + (1/\tau_y)\textstyle\int_0^t e_y dt], \quad e_y = y - y^* \tag{18b}$$

$$\kappa_y = (k + \omega)/a, \quad \tau_y = 1/k + 1/\omega_s$$

and k_y (or τ_y) is the proportional gain (or integral time). Thus, the OF tracking controller (17) is an observer-based realization of the PI controller (18) with antiwindup protection.

4.2 FF temperature setpoint (SP) compensator

Here the FF composition controller (9,12) is approximated by means of a dynamic temperature SP compensator with (pre-computed) static-nonlinear and (on-line) dynamic-linear components.

For this aim, introduce the stationary version

$$0 = \boldsymbol{f}_\zeta^*[\boldsymbol{c}_\zeta^s, \bar{z}, \boldsymbol{d}(t)], \quad u_s = \mu_z(\boldsymbol{c}_\zeta^s, \bar{z}), \quad y_s = \beta(\boldsymbol{c}_\zeta^s) \tag{19a-c}$$

of the primary NL FF composition controller (13). The solution $\boldsymbol{c}_\zeta^s = \boldsymbol{s}_\zeta(\bar{z}, \boldsymbol{d})$ of eq. (19a) followed by its substitution in (19b,c) yields the *static FF composition controller*

$$u_s = \mu_s(\bar{z}, d), \; y_s = h_s(\bar{z}, d) \tag{20a-b}$$

where

$$\mu_s(\bar{z}, d) = \mu_z[\boldsymbol{s}_\zeta(\bar{z}, d), \bar{z}], \quad h_s(\bar{z}, d) = \beta[s_\zeta(\bar{z}, d)]$$

$$f_s(\bar{z}, d) = f_y[\boldsymbol{s}_\zeta(\bar{z}, d), \mu_s(\bar{z}, d)]$$

For given composition setpoint $\bar{z}$, the μ_s and h_s versus load disturbance (d) plots are: pre-computed using the detailed model (4), fitted with suitable analytic scalar functions correlations, and stored for on-line use.

Following standard response testing-assessment identification techniques, the actual (13) minus stationary (19) transient output behavior e_s of the NN FF controller (13) can be reasonably approximated with the linear m-order dynamical matching model

$$e_s^{(m)} + k_{m-1}^* e_s^{(m-1)} + \cdots + k_1^* e_s = 0, \quad e_s = \hat{y}^* - y_s$$

where y_s is the SP generated by the precomputed static FF compensator (20), and $\hat{y}^*$ is the transient approximation of the SP y^* generated by the detailed model-based NL FF controller (13). Without restricting the approach, for the control improvement feasibility assesment scope of the present study, it suffices to use the first-order model (m = 1, with adjustable gain k_s)

$$\dot{\hat{y}}^* = \dot{y}_s - k_s(\hat{y}^* - y_s), \quad k_s \approx \lambda_* \tag{21}$$

By virtue of the u-antitonicity of f_y (11) the derivative of the stationary output map (20b) can be written as follows

$$\dot{y}_s = -au_s + \iota_s, \quad \iota_s = f_s(\bar{z}, d) + \mu_s(\bar{z}, d) \tag{22}$$

where the input-output pairs (u_s, y_s) and (u_s, ι) have *RDs = 1*, ι_s is observable from (y_s, u_s) (20), and consequently can be quickly estimated (with adjustable speed ω_s) with the reduce-order observer (23a) (Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013). The combination of the observer (23a) with the dynamic SP model (21) yields the *dynamic temperature SP compensator*

$$\dot{\chi}_s = -\omega_s\chi_s - \omega_s(\omega_s y_s - au_s), \hat{\iota}_s = \chi_s + \omega_s y_s \tag{23a}$$

$$\dot{\hat{y}}^* = -k_s(\hat{y}^* - y_s) + \hat{\iota}_s - au_s \tag{23b}$$

4.3 FF-OF controller

The concatenation of the OF tracking temperature controller (17) with the static (20) and dynamic (23) SP compensators yields the *dynamic FF-OF composition controller* (depicted in Fig. 1):

• FF temperature SP-control compensator

$$u_s = \mu_s(\bar{z}, d), \qquad y_s = h_s(\bar{z}, d) \tag{24a}$$

$$\dot{\chi}_s = -\omega_s\chi_s - \omega_s(\omega_s y_s - au_s), \quad \hat{\iota}_s = \chi_s + \omega_s y_s \tag{24b}$$

$$\dot{\hat{y}}^* = -k_s(\hat{y}^* - y_s) + \hat{\iota}_s - au_s \tag{24c}$$

$$\hat{u}^* = u_s + [k_s(\hat{y}^* - y_s) - \hat{\iota}_s]/a \tag{24d}$$

• OF temperature tracking controller

$$\dot{\chi} = -\omega\chi - \omega(\omega y - a\hat{u}), \quad \hat{\iota} = \chi + \omega y \tag{24e}$$

$$\hat{u} = \hat{u}^* + [k(y - \hat{y}^*) + \hat{\iota}]/a \tag{24f}$$

In PI form this controller is written as follows

$$u_s = \mu_s(\bar{z}, d), \quad y_s = h_s(\bar{z}, d) \tag{25a}$$

$$\dot{\hat{y}}^* = -a[\pi_s(\hat{y}^* - y_s) + u_s], \quad \hat{u}^* = u_s + \pi_s(\hat{y}^* - y_s) \tag{25b}$$

$$\hat{u} = \hat{u}^* + \pi(y - \hat{y}^*) \tag{25c}$$

where π is the PI temperature tracking controller (18b) and

$\pi_s(e_s) = \kappa_s[e_s + (1/\tau_s)\int_0^t e_s dt]$, $e_s = y - \hat{y}^*$
$\kappa_s = k_s/a$, $\tau_s = 1/\omega_s$
is a PI component of the dynamic SP compensator (25b) with proportional gain κ_s and integral time τ_s.

The simplified model-based FF-OF controller in IMC form (24): (i) recovers (up to observer convergence and transient SP approximation) the behavior of the detailed model-based robust NL FF-SF controller, (ii) is an observer-based version of its PI-based counterpart (18) with antiwindup protection, (iii) can be tuned with conventional-like tuning guidelines underlain by closed-loop stability conditions (Diaz-Salgado et al., 2011; Schaum and Alvarez, 2013), and (iv) executes in automatic mode the manual setpoint compensation procedure of the actual industrial control scheme.

From an industrial perspective the FF-OF controller constitutes an upgrade of the existing controller, including: (i) the replacement of the manual setpoint adjustment with a dynamic FF controller (24) with static-nonlinear (24a,b) and dynamic-linear (24b-d) components, (ii) the retuning of the existing PI temperature loop, (iii) the implementation of the PI controller in IMC form (24e-f) with antiwindup protection, (iv) a systematic construction-implementation procedure, (v) guarantee of robust closed-loop behavior, and (vi) an implementation with a gradual transition from the old to the new control scheme. The OF component (24) is constructed on the basis of the rather simple column model (15), and the FF component is off-line calculated on the basis of the detailed column model (2).

5. CONTROL TESTING

5.1 Static FF component

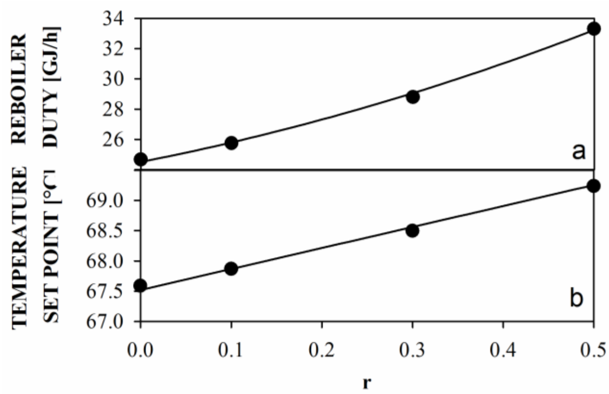

Fig. 4 FF static SP compensator (24a) over four load values (circles), and polynomial fitting (continuous curve).

Given the prescribed composition setpoint $\bar{z} = 0.018$, the control (μ_s) and output functions (h_s) of the static SP compensator (24a) were computed with the detailed model (4) over four load values (circles in Fig. 4), and fitted with the polynomials

$u_s = c_1 + c_2 d + c_3 d^2$, $(c_1, c_2, c_3) = (24.51, 11.76, 11.36)$
$y_s = b_1 + b_2 d$, $(b_1, b_2) = (67.52, 3.47)$

plotted in Fig. 4 (continuous curves), showing that: the static function y_s (or u_s) of the SP compensator (24) is linear (or quadratic, nearly-linear).

5.2 Test with multi-load disturbance data

First, the simulated closed column driven by the industrial multi-load disturbances (presented in Fig. 2 and 3) was run with (i) the existing PI temperature control (without manual SP compensation), (ii) its retuned version in IMC form (24b), and (iii) the proposed single-load FF-OF control (24). The resulting NC4 distillate composition responses are presented in Fig. 5, showing that, as expected: (i) the FF-OF yields the best behavior followed first by the retuned OF controller and then by the existing PI controller without retuning. Comparing with the actual industrial behavior (Fig. 5a), the retuned PI improved the closed-loop behavior, and further improvement was attained by the incorporation of the FF controller. The control responses of the retuned PI and proposed FF-OF control are shown in Fig. 6, showing that the behavior improvement is accomplished with reasonable control action.

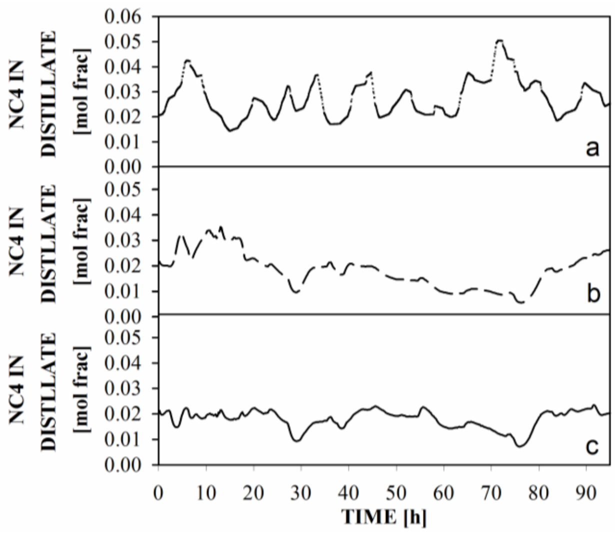

Fig. 5 Closed-loop NC4 distillate composition responses with industrial multi-load disturbance data (Figs 1 and 2) using: (a) the existing PI temperature control, (b) its retuned version (b), and (c) single-load FF-OF control (24).

These results (Fig. 5 and 6) constitute a robustness test of the proposed single-load FF-OF controller, as the simulated column was subjected to the actual multi-load disturbances of the industrial column.

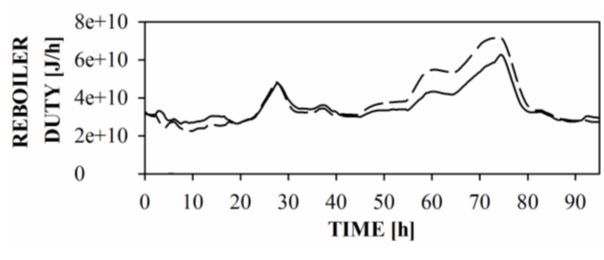

Fig. 6 Control effort for the closed-loop column subjected to multi-load disturbance (Fig. 1 and 2) with retuned temperature OF (dotted line) and single-load FF-OF control (continue line).

5.3 Test with single-load disturbance data

Having in mind that the ultimate goal of the industrial control problem is the development of a multi-load FF-OF control scheme, in this study we circumscribe ourselves to a first step towards that goal: the assessment of the feasibility improvement through single-load FF and OF retuning in the presence of the simulated ratio load disturbance step change series r shown in Fig. 7 (keeping the total feed F and distillate D flow rates at their nominal values). The results are presented in Fig. 8, showing that, as expected: in terms of impurity (transient and asymptotic) regulation, and temperature setpoint tracking, the behavior of the proposed FF-OF controller (triangles) outperforms the behavior of the OF controller (circles).

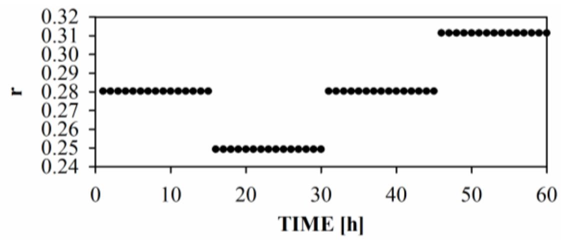

Fig. 7 Simulated sequence of feed ratio step changes.

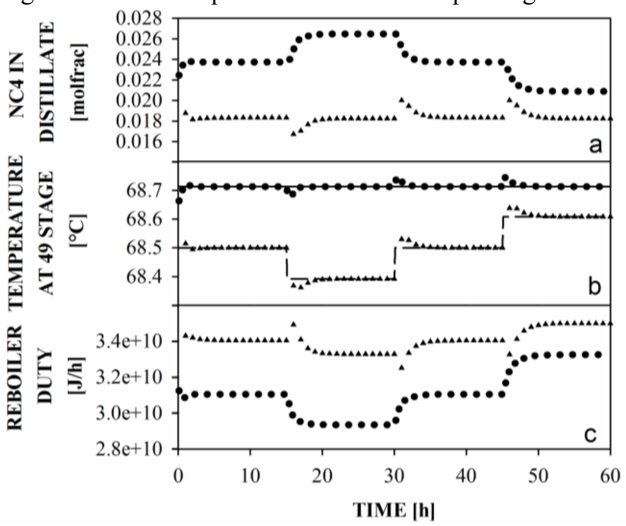

Fig. 8. Closed-loop behavior with FF-OF control (24) (triangles) and OF control (circles).

6. CONCLUSIONS

The feasibility of improving the performance of the industrial multicomponent distillation column by upgrading its actual OF temperature control (with manual setpoint compensation) has been established. The upgrade consists in: (i) retuning the existing control loop and realizing it in IMC form with antiwindup protection, (ii) adding a nearly-linear single-load dynamic temperature setpoint compensator which amounts to a PI loop with antiwindup protection plus a linear first-order lag, (iii) accompanying the design with conventional-like tuning guidelines underlain by a robust stability criterion drawn from advanced control theory, and (iv) enabling a gradual transition from the old to the new control scheme. While the development of the robust FF-OF control scheme required the combination of notions and tools from advanced nonlinear robust control theory, the actual implementation of the controller amounts to the combination of conventional-like elements. The derivation of the controller within conventional (Shinskey, 1988, Luyben, 1990) or linear-advanced (Peng et al., 2013) control frameworks does not seem to be a straightforward task. Work is under way to derive the multi-load version of the proposed single-load FF OF controller.

ACKNOWLEDGEMENTS

The authors thankfully acknowledges SARLUX refinery for M. Porru's doctorate fellowship support as well as for sharing valuable industrial experience and insight.

REFERENCES

Alvarez-Ramírez, J. J., Alvarez, J., Morales, A. (2002). An Adaptive Cascade Control for a Class of Chemical Reactors. *International journal of Adaptive Control and Signal Processing*, 16, 681.

Baratti, R., Bertucco, A., Da Rold, A., Morbidelli M. (1998). A Composition Estimator for Multicomponent Distillation Columns - Development and Experimental Test on Ternary Mixtures, *Chemical Engineering Science*, 53, 20, 3601-3612.

Castellanos-Sahagún, E., Alvarez-Ramírez, J., Alvarez J. (2005). Two point temperature control and structure design for binary distillation columns. *Industrial and Engineering Chemistry Research*, 44, 142-152.

Diaz-Salgado, J., Alvarez, J., Schaum A., Moreno, J.A. (2011). Feedforward output-feedback control for continuous exothermic reactors with isotonic kinetics, *Journal of Process Control*, 22 (1), 303-320.

Fruehauf, P. S. and Mahoney, D. P. (1994). Improve distillation control design. *Chemical Engineering Progress*, 90(3), 75-83.

Hirschorn, R. M. (1979). Invertibility of Nonlinear Control Systems. *SIAM Journal of Control and Optimization*. 17(2), 289–297.

Isidori, A. (1989). *Nonlinear Control Systems*, Springer-Verlag, Berlin.

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

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

Peng, C., Zhang, Z., Zou, J. et al. (2013) *Journal of Process Control*, 23 (10), 1415-1425.

Schaum, A.. and Alvarez, J. (2013). Saturated linear dynamic output-feedback control for a class of three-state continuous bioreactors with inhibited kinetics. *Journal of Process Control*. 23 (3), 332–350.

Sepulchre, R., Jankovic, M., Kokotovic, P. (1977). Constructive Nonlinear Control, Springer-Verlag, Berlin

Shinskey, F. G. (1977). Distillation Control for Productivity and Energy Conservation, Mc GrawHill New York.

Shinskey, F. G. (1988). Process Control Systems-Application, Design, and Tuning. 3rd Ed. McGraw-Hill.

Skogestad, S. (1997). Dynamics and control of distillation columns: A critical survey. *Modeling, Identification and Control*, 18(3), 177-217.

Sontag, E. D. (2001), The ISS philosophy as a unifying framework for stability-like behavior, *Lecture Notes in Control and Information Sciences*, 259, 443-467

Wolfgang, H. (1967). *Stability of Motion Springer*, Verlag New York.