Chemical Engineering Journal 99 (2004) 187–201

Chemical Engineering Journal

www.elsevier.com/locate/cej

Multivariable control configurations for composition regulation in a fluid catalytic cracking unit

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Received 27 July 2003; accepted 15 October 2003

Abstract

The control problem of fluid catalytic cracking (FCC) units is a challenging task due to its model complexity, non-linear dynamics, constrained variables and cross-coupling interaction between inputs and outputs. This paper is concerned with the design of multivariable feedback control configurations for composition control at the riser output for FCC units. A linear cascade (master/slave) control configuration is proposed, which leads to asymptotic regulation of the riser output composition (e.g. gasoline yield) about a feasible set point. Sufficient conditions to achieve regulation in terms of the steady-state gain matrix and response time-constants are provided, allowing to obtain a systematic procedure for analyzing multivariable control configurations of complex and interacting processes. Some tuning guidelines issues are discussed. Numerical simulations on a non-linear dynamical model operating in the partial-combustion mode are used to show the effectiveness of several multivariable control configurations under disturbances and uncertainty parameters.

Keywords: Chemical reactors; Composition control; Multivariable control; Fluid catalytic cracking

1. Introduction

Petroleum refiners use fluid catalytic cracking (FCC) technology to convert crude oil to blending stocks for use in gasoline, diesel, and heating oil. About 45% of worldwide gasoline production comes from FCC processes and its ancillary units. Due to its large throughput, the high product-feed upgrade and commercial importance, the overall economic benefits of a refinery could be considerably increased if proper control and optimization strategies on operating (e.g. temperatures, flow rates, etc.) and quality (e.g. composition) variables are implemented on FCC units [1,2].

There have been many studies in the literature addressing the problem of controlling FCC unit. For instance, non-linear controllers [3,4] and more complex mode predictive strategies [5,6] have been proposed. Multivariable control of FCC units has been considered for instance by Balchen et al. [7] and Grosididier et al. [8]. Balchen et al. [7] have used state–space predictive control to regulate temperature in FCC units. The controller is obtained by solving on-line a non-linear programming problem and simulations showed good closed-loop performance. However, the complex interaction among the process variables and the constraints on the manipulated and controlled variables can cause the computing cost to be high and time-consuming. Besides, the robustness analysis of the feedback controller becomes quite involved because of the model used to compute the compensator. Grosididier et al. [8] have described the advanced computer controls installed on the reactor and the regenerator on a real-life FCC unit, concluding that advanced multivariable control can lead to a good dynamic performance with a margin of robustness. Application of non-linear controllers with uncertainty estimation for the temperature regulation of FCC processes can be found in Alvarez-Ramirez et al. [3] and Aguilar et al. [4]. These approaches produce practical controllers where the closed-loop temperature trajectories are forced to remain in a neighborhood close to the set point. In an important contribution, Hovd and Skogestad [9] addressed the problem of control structure (strategy) selection based on linear models. They concluded that a favorable selection of controlled variables is critical for good control of FCC units. Overall, these studies have shown that FCC units are non-linear, multivariable and complex dynamic control systems. Complexity often is caused

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doi:10.1016/j.cej.2003.10.003

by the strong interactions existing between the control loops.

Most FCC control designs have dealt with the problem of stabilizing the process temperatures at a given set point. From an operating viewpoint, temperature regulation is a basic control objective imposed to guarantee a safe process operation. Nevertheless, the main task in the operation and control of FCC units is the regulation of gasoline quality (e.g. composition) at the output of the riser [1,2]. In practice, gasoline composition regulation is approached via indirect methodologies where a specified riser outlet temperature is regulated at a given set point which, in principle, corresponds to the desired composition value. However, model uncertainty and load disturbances (e.g. feed flow, temperature and composition) may unmatch the prescribed steady-state temperature corresponding to the composition set point, which in turns induces a steady-state offset in composition. This problem converts the manual control by the operator into a trial-and-error one to set the "correct" temperature set point. Despite operational experience with this class of units has been satisfactory, the lack of the possibility of a direct control of yields and compositions is a matter of concern. Thus, from industrial practice viewpoint, it would be desirable to dispose of controllers with servo responses (zero steady-state error in gasoline composition) and robustness against modeling errors and disturbances. On the other hand, the FCC unit is operated against one or more constraints. The most common constraint is a coke-burning limit. In this way, a controller design strategy for FCC units should therefore be capable of interactions between input–output pairs, and must be robust to modeling errors and non-linearities over the operating region.

In this paper, a combined multivariable cascade control structure has been developed for composition regulation in FCC units. The primary or master control loop is a pure integral control, which uses composition measurements to provide servo responses to the secondary or slave control loops. The secondary control loop corresponds to a linear multivariable PI temperature regulator, which uses both riser exit and regenerator cyclone temperature measurements. For control design, we have employed simple linear input–output models and we have exploited open-loop gain matrix properties in order to construct different multivariable control configurations. The analysis and design are straightforward for refining industrial applications, and take full advantage of the true interactive nature of a multidimensional system to achieve good controller performance. Some tuning issues are also discussed and their performance is illustrated by means of numerical simulations.

This work is organized as follows. Section 2 describes the FCC process. Section 3 presents the corresponding model identification for control purposes. Section 4 presents the analysis and design of different multivariable cascade control configurations. Numerical simulations results on a FCC unit operating in partial combustion mode are then given in Section 5. Finally, some conclusions are drawn in Section 6.

2. The FCC process

Several authors have made substantial efforts to model the behavior of FCC units. A detailed review of recent work on FCC modeling can be found in a paper by Arbel et al. [10]. Since FCC feedstocks consist of thousands of components, the estimation of intrinsic kinetic constants is very difficult; thus, the lumping of components according to the boiling range is generally accepted. Contributions to the modeling of FCC units vary from regenerator models over kinetic models for the reactions taking place in the reactor riser [1,2]. The model used for this case study is one developed by Lee and Groves [11] with slight modifications introduced by Hovd and Skogestad [9]. It is based on the three-lump reactor model, which comprises the main components in a FCC unit. The cracking is then described by the following three reactions:

$$\begin{aligned}&\text{feed (F)} \xrightarrow{1} \text{gasoline (G)}\\&\text{gasoline (G)} \xrightarrow{2} \text{coke} + \text{gas (L)}\\&\text{feed (F)} \xrightarrow{3} \text{coke} + \text{gas (L)}\end{aligned} \tag{1}$$

In general, FCC processes are highly reactive in the sense that almost every molecule in the feed undergoes some change, but overall conversion as used here is typically 30–40 wt.%. The gasoline yield increases with conversion up to a maximum and then decreases as the second reaction in Eq. (1) predominates, and gasoline cracks to lighter products and to coke. A FCC model with capability to describe the main dynamical aspects (e.g. interactions, convergence rates, etc.) for a feedback control study can be found in Hovd and Skogestad's paper [9]. Control design results and numerical simulation tests described in subsequent sections will be based on such a non-linear FCC model.

2.1. FCC unit description

A schematic overview of the FCC process is shown in Fig. 1. There are two main stages in the process: the cracking where pertinent reactions take place, and the regeneration where the catalyst is regenerated by burning off the coke deposited on the bed. Feed oil is contacted with hot catalyst at the bottom of the riser, causing the feed to vaporize. The cracking reactions occur while the oil vapor and catalyst flow up the riser. The residence time of the catalyst and hydrocarbon vapors in the riser is typically in the range 5–8 s. The riser top temperature is typically between 750 and 820 K and is usually controlled by regulating the flow of hot regenerated catalyst to the riser. As a by-product of the cracking reactions, coke is formed and deposited on the catalyst, thereby reducing catalyst activity. The catalyst and products are separated in the reactor. Steam is supplied to the stripper in order to remove volatile hydrocarbons from the catalyst. In the regenerator, which is operated in the fluidization regime, the coke is burnt off the catalyst surface by the air blown into the bed. This combustion reaction serves to reactivate the catalyst activity and to maintain the bed

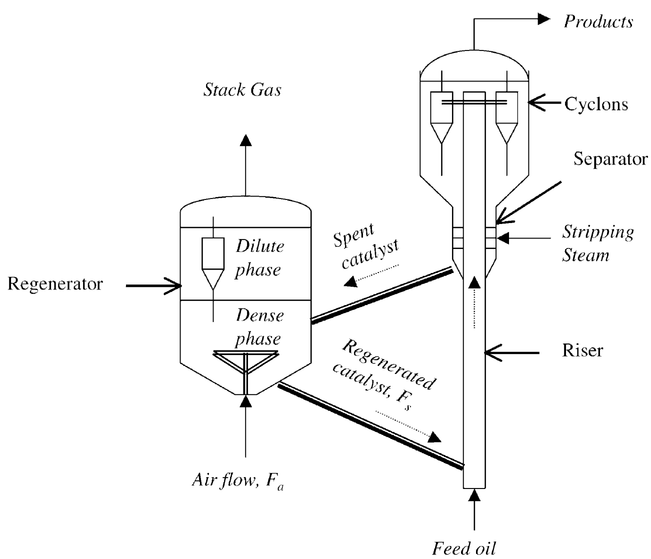

Fig. 1. Schematic diagram of a FCC unit.

temperature (950–980 K for a gas oil cracker and 980–1080 K for a resid cracker) high enough to supply the heat required for the vaporization and cracking reactions of the feed in the reactor [2].

Depending on the coke producing tendency of the feed, the FCC process can be operated in two distinct modes: partial combustion and the complete combustion modes. In the partial combustion mode the conversion of coke to CO_2 is not complete, which means that relatively large amounts of both CO are formed (this CO-rich regenerator flue gas can be sent to a CO boiler for further combustion to produce high pressure steam). It is not always possible operate a FCC unit in the complete combustion mode, specially if the feed has a large coke production tendency and there exists also an economic incentive operating in the partial combustion mode, as the heat recovered in the CO boiler is valuable.

3. FCC identification and control structure selection

Research on the dynamic characteristics of FCC units have revealed that the critical operating dynamics of FCC processes consist essentially of a MIMO system with two inputs and two outputs [9]. As in previous works [3,9,11], in this paper the independent variables that will be used for control are the catalyst circulation rate F_s, and the air flow to the regenerator F_a. Although several manipulated input-regulated output pairings are possible, in this paper we shall constraint ourselves to the riser-regenerator cyclone (Hicks) control structure, which uses the regenerator cyclone temperature T_{cy}, and the riser exit temperature T_{ri} as controlled variables [9]. Hovd and Skogestad [9] have reported that Hicks (T_{ri}, T_{cy})-structure is able to provide good controllability properties in terms of right half-plane (RHP) transmission zeros and the relative gain array (RGA). Moreover, controlling T_{cy} avoids exceeding the metallurgical temperature limit in the regenerator cyclones and controlling T_{ri} directly affects the amount of gas products and therefore helps ensuring that the wet gas compressor operating limits are not exceeded. For this choice of pairing, there is no right half-plane (RHP) transmission zeros (which limit the achievable bandwidth) with a consequent potential for high-performance control. It must be pointed out that RHP transmission zeros are not desirable in a control structure because any controller cannot invert the plant and perfect control is not possible [9]. Moreover, RHP transmission zeros may lead to unstable dynamic compensation (non-minimum-phase systems), regardless of the design of the controller.

3.1. Input–output temperature dynamics model

Once the Hicks control structure has been chosen, the lower level control objective is to regulate the FCC process temperatures T_{cy} and T_{ri}. Regulation of T_{cy} is required to guarantee a reliable catalyst reactivation (via coke burning). On the other hand, regulation of the riser outlet temperature T_{ri} is needed to ensure product (e.g. gasoline) quality. The feedback control design aimed to achieve such temperature regulation will be based on a (simple) dynamical model of the FCC process.

In the process industries, where a high degree of uncertainty about process behavior are commonly, the input–output (transfer function) model approach is generally adequate for control design analysis and design purposes. An input–output model must retain the main dynamics characteristics of the processes dynamics, including dominant time-constants and control input interactions (reflected in the steady-state gain matrix). This section therefore considers only the structures of input–output models of multivariable processes used in control systems design and analysis.

Temperature (open-loop) step-responses of a FCC unit are shown in Fig. 2. Dynamic simulation of the FCC process was performed according to the non-linear model and parameters given in Hovd and Skogestad [9]. All simulation runs started from a steady-state condition corresponding to the base case study and the subsequent transient response was obtained as each simulation variable went through a series of step changes as shown in Fig. 2. The corresponding input–output model was determined from the reaction curve of the process obtained by giving step disturbances to catalyst circulation rate F_s and the air flow to the regenerator F_a. It can be seen that the step response in the regenerator unit is smooth, almost monotonous, and convergent, such that, it is reasonable to model the input–output response with simple stable first-order models. On the other hand, the output response in the riser for a catalyst flow rate disturbance exhibits a large peak before it settles at the new steady-state value. We can use a filtered output response in order to put the transfer function in a more appropriate form. Physically, this can be interpreted as accommodating a storage mixing

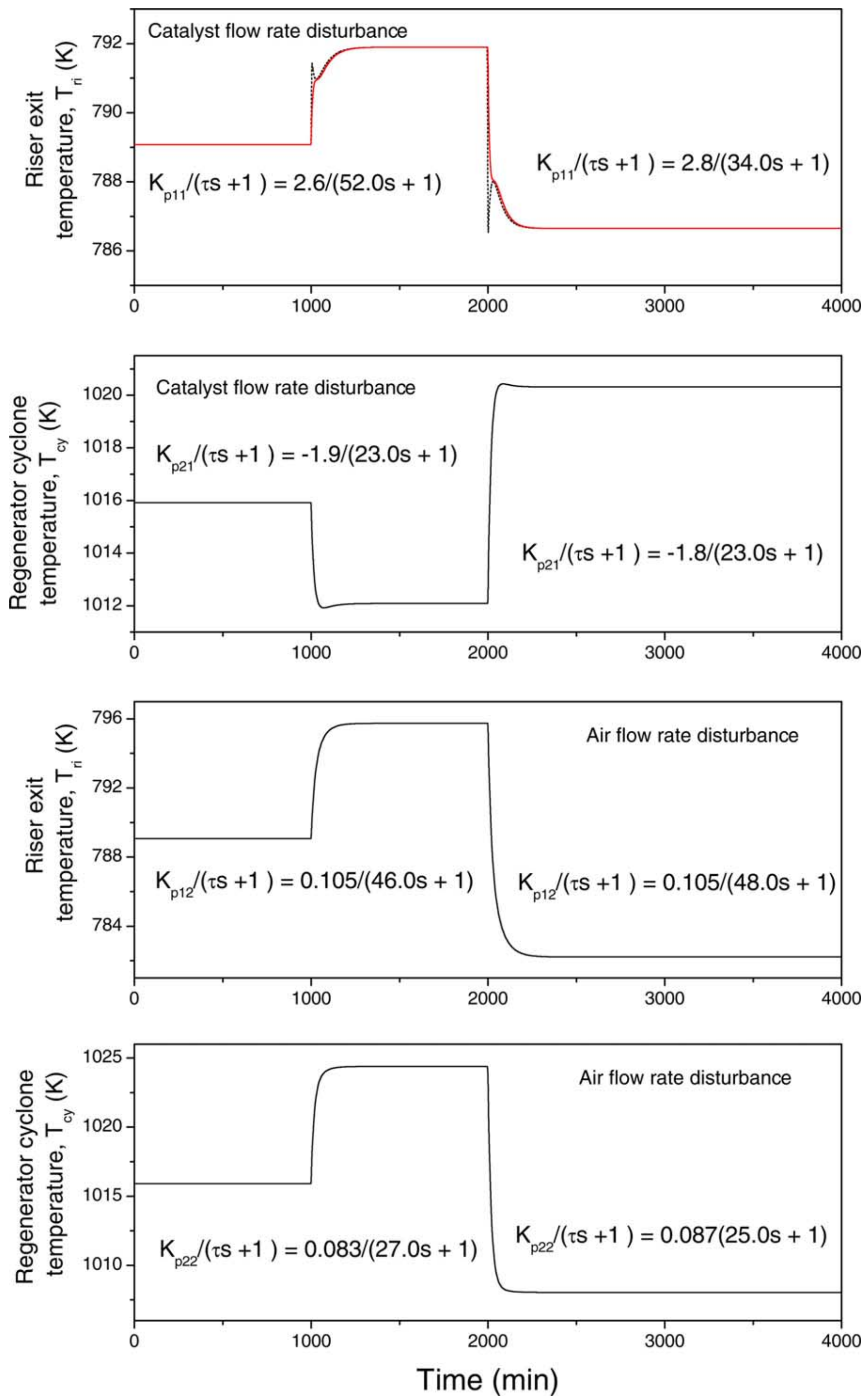

Fig. 2. Step identification for a FCC unit.

tank between riser and regenerator units, where the filtered time-constant corresponds to the residence time of this virtual tank.

It should be observed that temperature measurements in industrial practice are obtained continuously and without time-delay, while composition measurements are commonly not available for on-line control, or the measurements are delayed. Although higher-order models can be considered, the above suggest that first-order models without time-delay are more suitable model for the model the temperature dynamics of both riser and regenerator units. Besides, from simulations we have observed that, regardless the type of disturbance introduced, the time-constants are nearly the same. This means that the dynamics in both units is governed by the characteristic capacitance of each unit. In this way, one can approximate the temperature input–output dynamics as follows:

$$Y(s) = G(s)U(s)$$

where $Y(s) = (\Delta T_{\rm ri}, \Delta T_{\rm cy})$ is the two-dimensional vector of regulated output temperatures, $U(s) = (\Delta F_{\rm s}, \Delta F_{\rm a})$ is the vector of manipulated inputs and

$$G(s) = \begin{pmatrix} \dfrac{k_{\rm p11}}{\tau_{\rm a}s+1} & \dfrac{k_{\rm p12}}{\tau_{\rm a}s+1} \\ \dfrac{k_{\rm p21}}{\tau_{\rm b}s+1} & \dfrac{k_{\rm p22}}{\tau_{\rm b}s+1} \end{pmatrix}$$

The elements within the blocks in the transfer function matrix define the relationship between the respective input–output pairs. In the time-domain, the above first-order model can be written as

$$\dot{Y} = -GY + HU \tag{2}$$

where

$$G = -\begin{pmatrix} \tau_{\rm a}^{-1} & 0 \\ 0 & \tau_{\rm b}^{-1} \end{pmatrix} \tag{3}$$

and

$$H = \begin{pmatrix} \tau_{\rm a}^{-1} & 0 \\ 0 & \tau_{\rm b}^{-1} \end{pmatrix} \begin{pmatrix} k_{\rm p11} & k_{\rm p12} \\ k_{\rm p21} & k_{\rm p22} \end{pmatrix} \tag{4}$$

The corresponding steady-state gain matrix of the FCC multivariable system is then given by

$$K_{\rm p} = \begin{pmatrix} k_{\rm p11} & k_{\rm p12} \\ k_{\rm p21} & k_{\rm p22} \end{pmatrix} \tag{5}$$

The optimal operating point for a FCC unit usually lies at one or several constraints [2]. Common constraints include: (i) maximum regenerator cyclone temperature ($T_{\rm cy} \leq$ 1000 K) constraint. This constraint is usually important in the complete combustion mode and is determined by the metallurgical properties of the cyclones, (ii) maximum and minimum air blower capacity 0.0 kg /($s \leq F_{\rm a} \leq 60$ kg/s). The air blower provides the air needed for the combustion in the regenerator, and (iii) maximum and minimum catalyst circulation rate 100 $\rm kg_{cat}$/($s \leq F_{\rm s} \leq 400\,\rm kg_{cat}$/s). Indeed, these constraint should be accounted for during the control design procedure.

3.2. Input–output steady-state map

In principle, temperature regulation will suffice for a safe FCC operation. However, from an economics viewpoint product (gasoline) quality is of prime importance. Quality (composition) regulation is an upper level control problem whose implementation shall be based, within a cascade control configuration, on an existing temperature regulator. In our cascade control design methodology, to be described in later sections, the input–output steady-state map for product composition with respect to riser operating temperature, will be required.

Fig. 5 show the riser exit temperature–gasoline yield composition input–output map at steady-state conditions. The gasoline yield increases with conversion up to a maximum at about 43 wt.% (riser exit temperature $\approx$ 785 K) and then decreases as the second reaction in Eq. (1) predominates, and gasoline cracks to lighter products and to coke. Thus, because of competing effects of the two consecutive reactions [12], at $T_{\rm ri} > 785$ K exist two steady-state values of the riser exit temperature corresponding to a given value of the gasoline yield composition ($y_{\rm g}$). If the range is constrained to be in the interval $I_{\rm T} = [750, 785]$ K, which is also desirable for safe and operational restrictions (regenerated catalyst slide valve saturation), the map $T_{\rm ri} \rightarrow y_{\rm g}$ is strictly increasing (i.e. no input multiplicities).

4. Multivariable control configurations

Due to operational and safe restrictions, it is important to maintain both the regenerator cyclone and riser exit temperatures within a given operating range. Maintaining the cyclone temperature, $T_{\rm cy}$, under a maximum limit provides safe thermal operation for the regenerator (avoiding catalyst deactivation or destruction) and for the downstream units (piping and CO boiler). On the other hand, the riser (reactor) temperature, $T_{\rm ri}$, has a direct relation with the amount of heavy hydrocarbons that are converted to more valuable products. Thus, *the control objective is to design multivariable controllers with servo responses* (*zero steady-state error in gasoline composition*) *through measurements of composition with manipulation of the riser exit temperature and with simultaneously regulation of the regenerator cyclone temperature*. Moreover, since significant uncertainties in kinetics, model parameters and feedstock of FCC units can be present, the control design must be robust against modeling errors and disturbances.

For the control design, firstly the temperature stabilization multivariable loop is designed and then the composition regulation loop is constructed within a cascade framework (see Fig. 3). Cascade control is a common control configuration in several processes, which can be thought of as partial state feedback. Besides, most cascade control designs are based on P and PI control laws as the basic (master and slave) feedback compensations [13–15]. In the stabilization loops the variables to be controlled are both the riser exit temperature and the regenerator cyclone temperature, and the corresponding manipulated variables are the catalyst flow coming from the regenerator and the air flow to regenerator, as done in industrial practice. For the composition regulation loop the variable to be controlled is the gasoline yield composition with manipulations of the riser exit temperature. In this way, the temperature control loop is posed as a signal tracking problem where the output of the composition

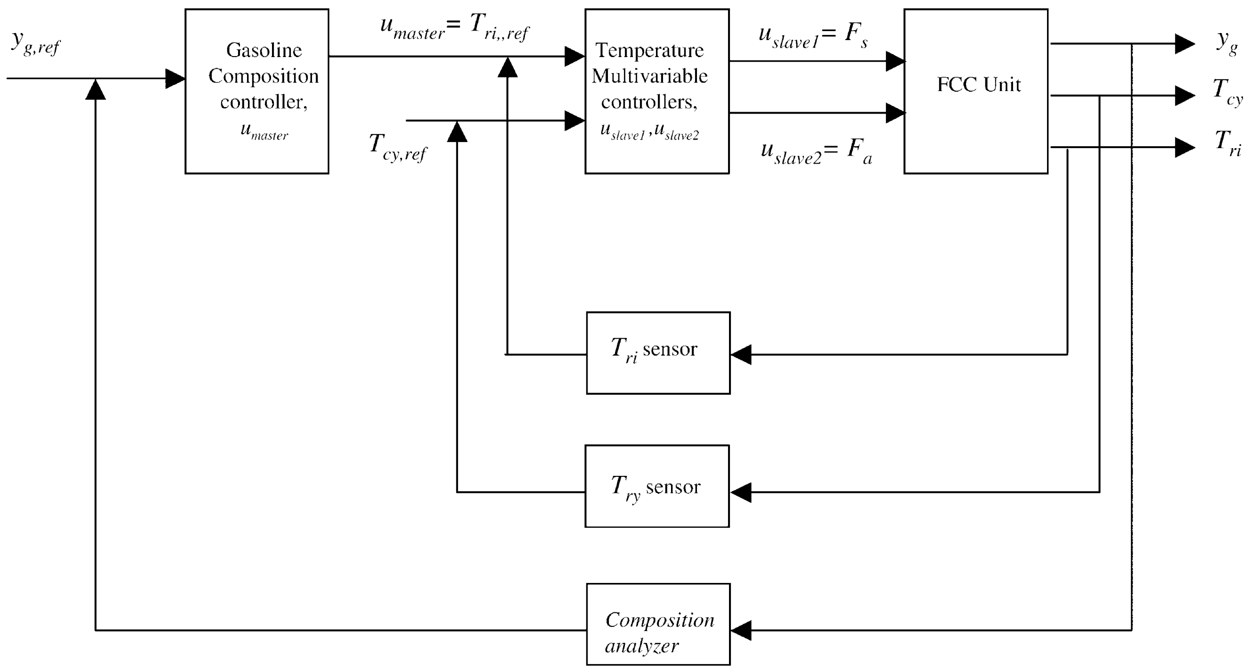

Fig. 3. Schematic diagram of the cascade control scheme for composition control.

regulation control loop is one of the reference signals to be tracked.

4.1. Temperature stabilization control design

Consider the nominal plant of the FCC unit given by Eq. (3). By introducing the regulation error vector $E = Y - Y_{\text{ref}}$, where $Y^{\text{T}} = (T_{\text{ri}}, T_{\text{cy}})$ and $Y_{\text{ref}}^{\text{T}} = (T_{\text{ri,ref}}, T_{\text{cy,ref}})$, one has that the error dynamics can be written as

$$\dot{E} = GE + HU$$

where, without lost of generality and for simplicity in algebraical manipulations, we have taken $Y_{\text{ref}}G = 0$ (in fact, the feedback controller is intended to be equipped with integral action). The following PI compensation structure is proposed for temperature regulation purposes:

$$U = -Q\left(\alpha_{\text{p}}E + \alpha_{\text{i}}\int E\,\mathrm{d}t\right) \tag{6}$$

where $\alpha_{\text{p}}, \alpha_{\text{i}} > 0$ are closed-loop tuning parameters, and Q is a *control structure matrix* to be determined later. Let $Z = \int E\,\mathrm{d}t$ and $X = (E, Z)$. Then, the corresponding closed-loop equations are then given as follows:

$$\dot{X} = A_{\text{c}}X \tag{7}$$

where

$$A_{\text{c}} = \begin{pmatrix} G - \alpha_{\text{p}}HQ & I \\ -\alpha_{\text{i}}HQ & 0 \end{pmatrix} \tag{8}$$

In this way, the stability of the temperature regulation loop is determined by the stability of the matrix A_{c}. Stability conditions for the stability of the matrix A_{c} are given in the following result.

Theorem 1. *Choose the control structure matrix Q such that HQ is anti-stable (i.e. all its eigenvalues have positive real part). Then, there exist a constant $\alpha_{\text{p,min}}$ and a positive constant $\alpha_{\text{i,max}}$ such that the matrix A_{c} is stable for all $\alpha_{\text{p}} > \alpha_{\text{p,min}}$ and all $0 < \alpha_{\text{i}} < \alpha_{\text{i,max}}$.*

Proof. First, consider the matrix $-G + \alpha_{\text{p}}HQ$. We can rewrite such matrix as $\alpha_{\text{p}}(-\varepsilon_{\text{p}}G + HQ)$, where $\varepsilon_{\text{p}} = \alpha_{\text{p}}^{-1}$. We can see the matrix $\varepsilon_{\text{p}}G$ as a (singular) perturbation to the anti-stable matrix HQ. By continuity arguments, there exists a positive number $\varepsilon_{\text{p,max}}$ such that $-\varepsilon_{\text{p}}G + HQ$ is anti-stable for all $0 < \varepsilon_{\text{p}} < \varepsilon_{\text{p,max}}$. Consequently, the matrix $G - \alpha_{\text{p}}HQ$ is stable for all $\alpha_{\text{p}} > \alpha_{\text{p,min}}$, where $\alpha_{\text{p,min}} = 1/\varepsilon_{\text{p,max}}$. Similar singular perturbation arguments can be used to show the existence of a positive number $\alpha_{\text{i,max}}$ such that the matrix A_{c} is stable for all $0 < \alpha_{\text{i}} < \alpha_{\text{i,max}}$. □

Notice that the key condition to ensure stability of the (multivariable) temperature regulation loop is that the matrix HQ be anti-stable. Given the fulfillment of this condition, the control structure matrix Q allows the selection of several control configurations (e.g. input–output pairings) of practical interest. Namely, these control configuration are the following:

- *Decoupling compensation*: This control structure corresponds to the selection $Q = H^{-1}$, such that $HQ = I$. The main objective with the use of a decoupling controller is to compensate for the effect of interactions brought about by cross couplings of the process variables, thus leading to a decoupled closed-loop response of the regulated outputs. This control structure can be useful for highly interacting processes where a decoupled output response is desired.

- *Transpose compensation*: In this case, one chooses $Q = H^{\mathrm{Tr}}$, where tr denotes transposition. One has that, as required by Theorem 1, $HQ = HH^{\mathrm{Tr}}$ is an anti-stable matrix as long as H is non-singular. As in the case of decoupling compensation, the transpose compensation uses all the control directionality information contained in the high-frequency matrix H^{Tr}. However, contrary to decoupling compensation, the resulting controlled outputs can display a significant interacting and sluggish response.
- *Partially decentralized compensation*: If Q is selected as an either upper or lower triangular matrix, the corresponding control structure is called as partially decentralized. This structure produces a single control loop and a second loop driven by the second one. This type of control structures can be attractive for processes with material and energy recycling. In fact, using a partial decentralization allows the feedback controller to compensate the possibly unstabilizing effects introduced by recycling flows (e.g. catalyst recycling in FCC units). In this way, the non-diagonal element in Q can be chosen to induce partial decoupling (via a type of feedforward action) in the recycling part of the processes.
- *Fully decentralized compensation*: In this control structure, one chooses $Q = \mathrm{diag}(\beta_1, \beta_2)$, where β_1 and β_2 are constants guaranteeing that HQ is an anti-stable matrix. In this case, one obtains two single control loops with non-decoupled closed-loop regulated temperatures response. Decentralized feedback controller is widely used in practice and, in general, is intended for modestly interacting processes. The main advantage of fully decentralized structures is that it leads to easy design and tune single control loops.

4.1.1. Some robustness issues

The stability of the temperature regulation (multivariable) control loop depends strongly on the stability of the matrix HQ. If the matrix H was exactly known, one only has to check that the matrix HQ is anti-stable. Testing of this condition is straighforward for the decoupling and the transpose compensation structures. However, this is not the case for partially and fully decentralized compensation structures, although the computations are not so involved since one has to check only that the determinant and the trace of the matrix HQ be positive. However, if the matrix H is not exactly known or has significant variations because of, e.g. changing operating conditions, some care must be taken in order to ensure fulfillment of the condition HQ. That is, if $\bar{H}$ is an available estimate of $\bar{H}$ and ΔH is its perturbation belonging to an uncertainty set Ω_H, then one has to guarantee that $(\bar{H} + \Delta H)Q$ is stable for all $\Delta H \in \Omega_H$. This corresponds to checking the robust stability of the matrix $-(\bar{H} + \Delta H)Q$ with respect to the uncertainty set Ω_H. There exists several methods to study the stability of the matrix $-(\bar{H} + \Delta H)Q$. In particular, the parametric approach developed by Bhattacharyya et al. [16] seems to be appropriate and computationally efficient. However, given that the FCC process under study corresponds to the simplest two-dimensional case, some particular and practical results can be provided:

(i) Decoupling and transpose compensation structures rely on the full directionality information contained in the interaction matrix H. Essentially, the role of decoupling is to compensate a multivariable process into a series of independent single-loop sub-systems. In particular, decoupling compensation requires inversion of the matrix H. To guarantee control stability, one has to ensure that the matrices $(\bar{H}+\Delta H)\bar{H}^{-1} = I+\Delta H\bar{H}^{-1}$ (for decoupling control) and $(\bar{H} + \Delta H)\bar{H}^{\mathrm{Tr}} = \overline{HH}^{\mathrm{Tr}} + \Delta H\bar{H}^{\mathrm{Tr}}$ (for transpose control) are anti-stable for all $\Delta H \in \Omega_H$. One can see the matrices $\Delta H\bar{H}^{-1}$ and $\Delta H\bar{H}^{\mathrm{Tr}}$ as perturbations of the anti-stable matrices I and $\overline{HH}^{\mathrm{Tr}}$, respectively. For small departures ΔH from $\bar{H}$, anti-stability can be proven via continuity arguments [16]. In general, the effects of the "errors" $\Delta H\bar{H}^{-1}$ and $\Delta H\bar{H}^{\mathrm{Tr}}$ on the stability of the control loop can be evaluated by means of the condition number κ of the matrix H. In fact, the condition number provides some information on the propagation of the estimation errors $\Delta H\bar{H}^{-1}$ and $\Delta H\bar{H}^{\mathrm{Tr}}$ in the feedback control system. That is, the larger the condition number, the more less the robustness margin of the control system. In this way, decoupling and transpose control configurations are not recommended if the uncertainty ΔH is non-negligible.

(ii) Fully decentralized control is very important in process applications because of its several advantages over a fully multivariable design. Including flexibility in operation, failure tolerance, simplified design, and simplified tuning. In this way, it is interesting to provide stability conditions for the case of FCC units. In can shown that, for our case study described in Section 2, the interaction matrix H has the following structure:

$$\begin{pmatrix} h_{11} & h_{12} \\ -h_{21} & h_{22} \end{pmatrix}$$

where all the h_{ij}'s are positive numbers. If $Q = \mathrm{diag}(\beta_1, \beta_2)$, then

$$HQ = \begin{pmatrix} \beta_1 h_{11} & \beta_2 h_{12} \\ -\beta_1 h_{21} & \beta_2 h_{22} \end{pmatrix}$$

The determinant and the trace of HQ are given by

$$\mathrm{Det} = \beta_1\beta_2(h_{11}h_{22} + h_{12}h_{21}),$$
$$\mathrm{Trace} = \beta_1 h_{11} + \beta_2 h_{22}$$

Choose $\beta_1 > 0$ and $\beta_2 > 0$. Then, $\mathrm{Det} > 0$ and $\mathrm{Trace} > 0$, implying that the matrix HQ is anti-stable. In this way, *unconditional* stability of the control loop is guaranteed if one chooses $Q = \mathrm{diag}(\beta_1, \beta_2)$ with β_1

and β_2 being any positive parameters. In particular, as done commonly in practice, one can choose $\beta_1 = \bar{h}_{11}^{-1}$ and $\beta_2 = \bar{h}_{22}^{-1}$, where $\bar{h}_{11}$ and $\bar{h}_{22}$ are available estimates of the actual values h_{11} and h_{22}, respectively.

(iii) Assume only partially decentralized control with an upper triangular matrix Q as follows:

$$Q = \begin{pmatrix} \beta_1 & \gamma \\ 0 & \beta_2 \end{pmatrix}$$

In this case,

$$HQ = \begin{pmatrix} \beta_1 h_{11} & \gamma h_{11} + \beta_2 h_{12} \\ -\beta_1 h_{21} & -\gamma h_{21} + \beta_2 h_{22} \end{pmatrix}$$

so that

$$\text{Det} = \beta_1\beta_2(h_{11}h_{22} + h_{12}h_{21}),$$
$$\text{Trace} = \beta_1 h_{11} + \beta_2 h_{22} - \gamma h_{21}$$

Analogous to the fully decentralized case, stability of the control loop is guaranteed if $\gamma < 0$. In particular, one can choose either $\gamma = -\bar{h}_{21}^{-1}$ or $\gamma = -\bar{h}_{21}$. Similar stability conditions can be derived for the lower triangular case.

4.2. Regulation cascade control design

Let $y_{g,\text{ref}}$ be the composition reference. The temperature regulation control described above ensures that $T_{\text{ri}} \to T_{\text{ri,ref}}$ asymptotically. That is, the temperature at the riser exit, $T_{\text{ri}}(t)$, approaches the prescribed reference $T_{\text{ri,ref}}$. The temperature set point $T_{\text{ri,ref}}$ is an exogenously specified command. so it can be manipulated to achieve additional control task. Similar in nature to the operator-based procedure used in practice, we will use the set point $T_{\text{ri,ref}}$ as a manipulated control input to regulate the riser exit composition $y_{\text{g}}(t)$ at a feasible set point $y_{g,\text{ref}}$. A control structure designed in this form leads, in turn, to a cascade control configurations where a feedback control loop for the map $T_{\text{ri,ref}} \to y_{\text{g}}$ is the master controller, and the previously described (multivariable) temperature regulator plays the role of the slave controller.

It is expected that, once the temperature regulation loop has been closed, the dynamics of FCC unit displays acceptable damping. In this way, as in classical linear control, this suggests to use a simple integral action to ensure that $y_{\text{g}} \to y_{g,\text{ref}}$,

$$T_{\text{ri}} = \bar{T}_{\text{ri}} + K_{\text{I}} \int (y_{g,\text{ref}} - y_{\text{g}}(t-\theta))\,\text{d}t$$

where K_{I} is the master integral gain. The parameter θ represents a time-delay introduced by composition measurement devices. The sign of the integral gain K_{I} is defined by the sign of the steady-state map $y_{\text{g}} = g(T_{\text{ri}})$ in the domain $I_{\text{T}} = [750, 785]$. That is, the derivative $D_{\text{g}}(T_{\text{ri}})$ is the steady-state gain of the FCC unit. In this way, one has that $\text{sign}(K_{\text{I}}) = \text{sign}(D_{\text{g}}(T_{\text{ri}}))$. The stability results reported by Desoer and Lin [17] implies that the master feedback function given by T_{ri} provides asymptotic regulation of the output composition y_{g} about any prescribed set point $y_{g,\text{ref}} = g(T_{\text{ri}})$, for any $T_{\text{ri}} \in I_{\text{T}}$ and sufficiently small $|K_{\text{I}}|$. Basically, the master controller is a low-gain integral feedback compensator to cope with non-linearities and output composition measurement delays. Saturation protection can be provided by the slave controllers when the desired reactor temperature violates certain limit. In this way, unacceptable reactor temperature overshoots can be avoided because the temperature set point is limited within certain security operation domain.

Summarizing, the actual control input is a virtual input reference for the riser output temperature, T_{ri}. Thus, the proposed multivariable-cascade structure is given by

$$T_{\text{ri,ref}} = \bar{T}_{\text{ri,ref}} + K_{\text{I}} \int (y_{g,\text{ref}} - y_{\text{g}}(t-\theta))\,\text{d}t$$

and

$$F_{\text{s}} = \bar{F}_{\text{s}} + q_{11}\alpha_{\text{p}}(T_{\text{ri}} - T_{\text{ri,ref}}) + \alpha_{\text{i}} \int (T_{\text{ri}} - T_{\text{ri,ref}})\,\text{d}t + q_{12}\alpha_{\text{p}}(T_{\text{cy}} - T_{\text{cyref}}) + \alpha_{\text{i}} \int (T_{\text{cy}} - T_{\text{cyref}})\,\text{d}t$$

and

$$F_{\text{a}} = \bar{F}_{\text{a}} + q_{21}\alpha_{\text{p}}(T_{\text{ri}} - T_{\text{ri,ref}}) + \alpha_{\text{i}} \int (T_{\text{ri}} - T_{\text{ri,ref}})\,\text{d}t + q_{22}\alpha_{\text{p}}(T_{\text{cy}} - T_{\text{cyref}}) + \alpha_{\text{i}} \int (T_{\text{cy}} - T_{\text{cyref}})\,\text{d}t$$

where the q_{ij}'s are the elements of the matrix Q. In other words, the composition control loop provides the riser exit temperature reference values $T_{\text{ri,ref}}$ to the temperature stabilization control loops. In this way, the riser exit temperature controller forces T_{ri} to reach $T_{\text{ri,ref}}$ specified by the composition control loop.

4.3. Tuning guidelines

The real advantage of the proposed MIMO cascade control configurations over traditional and non-linear control MIMO designs is that easy design and tuning methodologies can be provided. In fact, if the control structure matrix Q is chosen adequately such that HQ is anti-stable, then the FCC unit can be robustly stabilized via simple multivariable PI control configurations.

Tuning rules should be well motived, and preferably model-based and analytically derived. Besides, they should be simple and easy to memorize, and should work well on a wide range of processes. In the context of multivariable processes there are some important contributions in the literature. Luyben [18] describe a tuning procedure for multiloop PI and PID controllers operating in a multivariable environment. These controllers contain up to three tuning constants per loop, and therefore, tuning then could present difficulties to the operator. Luyben [18] presented the biggest log modulus tuning (BLT) procedure for tuning multiloop PI controllers. Wang et al. [19] proposed a fully cross-coupled multivariable PID controller tuning method. Two successful PI and PID tuning methods for multivariable processes have been proposed by Loh et al. [20], and Halevi et al. [21], in which the relay feedback developed by Astrom and Hagglund [22] is used. The proposed methods appear to work well, but have the disadvantage that not all classes of multivariable processes can exhibit sustained and near-sinusoidal oscillations under multiloop relay feedback.

In the spirit of the available tuning methods for multivariable control, and exploiting the particular cascade structure of the proposed feedback controller, tuning and design rules can be formulated sequentially. First, the stabilization controllers are tuned. Second, the tuning of the regulation controller is accomplished by regarding the secondary loop as the system to be controlled. This procedure can be made in the following way:

1. *Temperature control loop*. Determine the matrix Q such that HQ is an anti-stable matrix. Once that the matrix Q has been chosen, the control tuning depends only on two parameters, namely, $\alpha_p, \alpha_i > 0$. These parameters α_i and α_p are adjustable tuning parameters for the speed of process response, so that they should be chosen to give a good performance and robustness. Smaller values of the constants α_i and α_p are selected to sacrifice the performance for robustness as the uncertain between the plant and model increase. We propose to choose α_p in a way that the dominant time-constant of the linear system $\dot{x} = (G - \alpha_p HQ)x$ be about 0.5–0.75 times the dominant open-loop time-constant. The rationale behind this tuning guideline is that enhanced convergence rate is induced by means of temperature feedback control, but without an excessive control effort that could lead to instabilities due to uncertainties and non-modeled (high-frequency) dynamics [23]. On the other hand, α_i must be chosen to provide good disturbance rejection response. By noticing that α_i has frequency units, similar to IMC tuning guidelines [23], we propose to use α_i values of the order of the inverse of the dominant open-loop time-constant.
2. *Composition regulation loop*. The steady-state input–output map can be used to tune the composition regulation loop based on simple IMC rules. Some care must be taken to avoid excitation of high-frequency dynamics, and so unstabilizing effects, due to composition measurement delays. In fact, the master controller is basically a low-gain feedback. Suppose that K_{ss} is an estimate of the input–output steady-state gain of the FCC unit at a nominal operating point. The estimate K_{ss} can be obtained from either a model or experimental data. Following ideas from linear internal model control [23,24], and considering the input–output steady-state map as approximately a first-order system, we propose the following tuning:

$$K_I = (K_{ss}\tau^*)^{-1}$$

where τ^* is about 0.5–0.75 times $\max\{\theta, \tau_d\}$ and τ_d is the dominant open-loop time-constant of the FCC unit.

5. Numerical simulations

Numerical simulations were carried out considering set point changes and typical disturbances to FCC reactor, namely, flow rate and temperature disturbances in the feedstock. The nominal values 25.35 and 294 kg/s are assigned to the control inputs F_a and F_s, respectively. Numerical values for other parameters are found in Hovd and Skogestad [9]. To account for actuator saturation, in all the simulation we have used the a saturated version of the temperature controller with control limits $U^{max} = [400\,kg_{cat}/s,\ 60\,kg/s]$ and $U^{min} = [100\,kg_{cat}/s,\ 0.0\,kg/s]$.

From numerical simulations about the nominal operating point, one obtain that the steady-state gain matrix is given by

$$K_p = \begin{pmatrix} 2.7 & 0.105 \\ -1.85 & 0.085 \end{pmatrix}$$

and the time-constants are $\tau_a = 45$ and $\tau_b = 23$ min, so that $G = \mathrm{diag}(0.022, 0.0435)$. In this way, the high-frequency matrix $H = GK_p$ is given by

$$H = \begin{pmatrix} 0.0594 & 0.0023 \\ -0.081 & 0.0037 \end{pmatrix}$$

The condition number of the interaction matrix H is $\kappa = 24.852$, showing that the FCC units is only a modestly interacting process. This lack of strong interaction is may be due to the large difference in capacities between the riser and the regenerator. In fact, the residence time of the regenerator is about 100 times the residence time of catalyst in the riser, so that the regenerator acts as a damping tank with certain decoupling capability.

Numerical simulations based on the non-linear dynamical model reported by Hovd and Skogestad [9] were used to evaluate the performance of the different control structures described in the above section.

5.1. Temperatures stabilization with multivariable control configurations

Fig. 4 show the closed-loop performance by considering the stabilization loop as derived in Section 4.1. The control action is at $t = 500$ min. Tuning parameters (α_p, α_i) were adjusted considering firstly a pure proportional action (α_p), such that control actions are free of input-saturation. Then, the integral parameter α_i was adjusted according to the tuning guidelines described in Section 4.3 in order to eliminate steady-state off set and to achieve better control

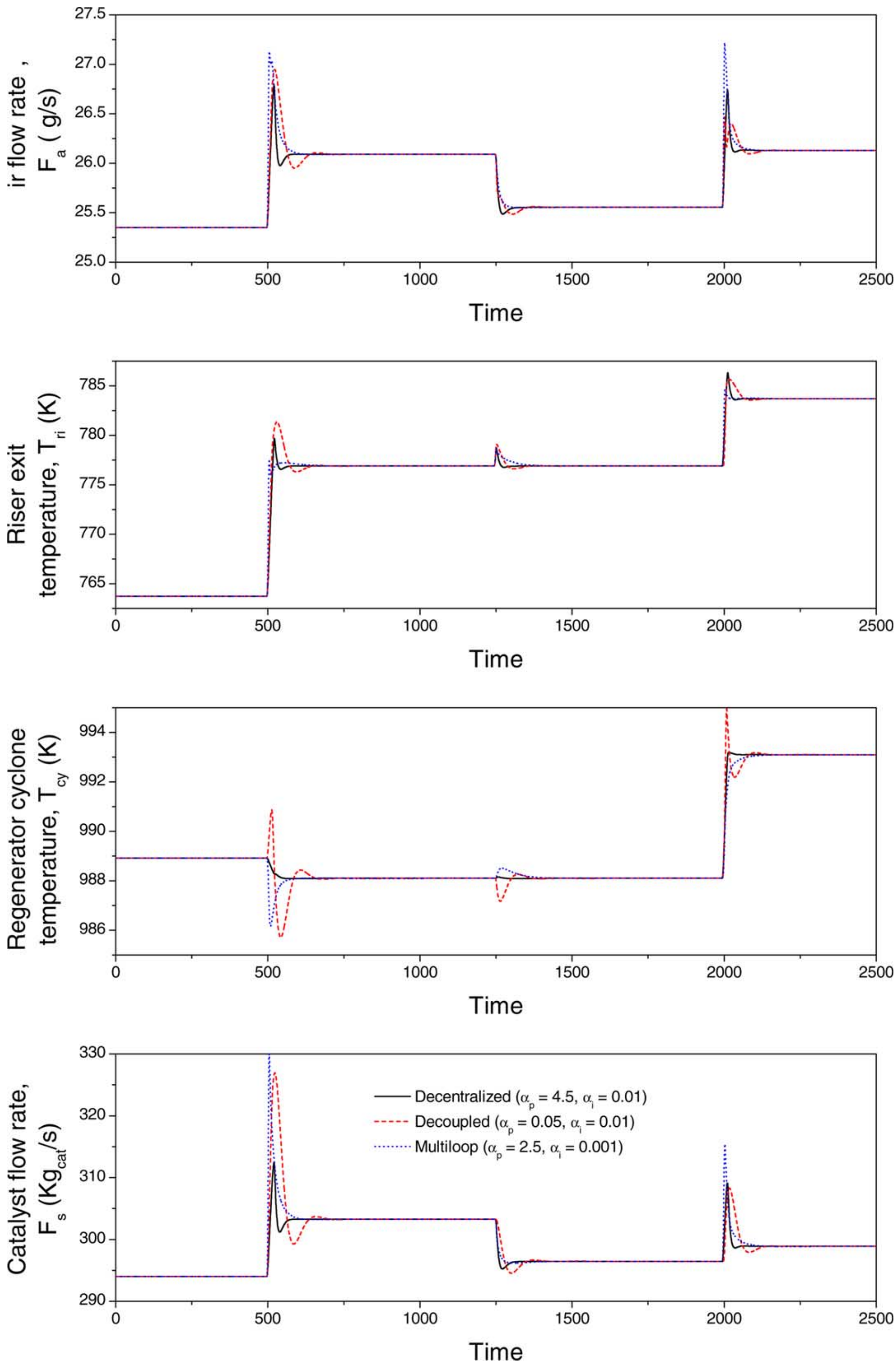

Fig. 4. Closed-loop performance by considering only the temperature stabilization loop.

performance. Fig. 4 presents the dynamics of the controlled FCC unit for fully decentralized, decoupling, and transpose compensation configurations. The disturbances entering the process are +5% in the feed flow rate and +8% in temperature of feed flow at $t = 1250$ min. The set-point values used in simulations were $T_{ri,ref} = 776.9$ K and $T_{cy,ref} = 988.1$ K at $t = 500$ min to $T_{ri,ref} = 783.7$ K and $T_{cy,ref} = 993.1$ K at $t = 2000$ min, which are about the set-point values used in previous works [3,9]. This class of step disturbances is expected to be present in industrial FCC units. For the partially decentralized compensation configurations cases, we have found similar control performance to that displayed by the complete decentralized compensation case by using the same tuning parameters. This seems to indicate that, for regulation control purposes, partial decoupling does not add significant performance (e.g. convergence rate) into the control loop. On the other hand, worst performance was observed for the decoupling and transpose compensations control structures. In the decoupling compensation case, such a performance degradation is produced because such control structure makes use of the inverse of $\bar{H}$. As discussed in Section 4, the stability of the temperature regulation is conditioned to the fulfillment of the condition that the matrix $H\bar{H}^{-1}$ be anti-stable. In this way, the larger the departure of $\bar{H}$ from H, the important the uncertainties propagation through the control loop. As a consequence, contrary to decentralized structures, which are unconditional stable, the decoupling compensation structure can display conditional stability as disturbances and set point changes induce variations in the entries of the high-frequency matrix H. Regarding the transpose compensation configuration, one founds that, although the closed-loop is stable, the response is very sluggish, which may not be acceptable in practice.

From Fig. 5 it can be seen that the value of $T_{ri,ref} = 783.7$ K corresponds to the maximum gasoline yield composition in the $T_{ri} - y_g$ steady-state map. It can be seen from Fig. 4 that the better control performance is obtained with the decentralized compensation configuration which is corroborated with the ISE values shown in Table 1. Simulations results in Fig. 4 show that the effects of disturbances on the control performance are relatively small. Nevertheless, Fig. 4 show that the set point change in both riser and cyclone regenerator temperatures combines to produce an overshoot in the controlled variables. In the decentralized compensation configuration, the riser exit temperature controller will increase the catalyst flow to attend the new set point and the regenerator cyclone controller will increase the air flow due to an increase in the catalyst flow in the riser-regenerator cycle. On the other hand, an increment in the temperature and flow rate of the feed will increase the riser exit temperature and then the riser exit temperature will decrease the catalyst flow and corresponding the regenerator cyclone controller will decrease the air flow rate to maintain both the riser exit and the regenerator cyclone temperatures at the prescribed set points. The decoupling and transpose compensations achieve the set points in similar settling times than decentralized compensation configurations, but with a slight degradation in the output responses and some oscillations in the input responses. In general, an increase in the tuning parameter α_i will result in faster but oscillatory responses, whereas an increase in the tuning parameter α_p leads to saturation of the control inputs.

Table 1
Computed ISE values for the temperatures stabilization loop

	Decentralized	Decoupled	Multiloop
$ISE_1 = \int (T_{ri} - T_{ri,ref})\, d\sigma$	14148.3	37115.1	56030.4
$ISE_2 = \int (T_{cy} - T_{cy,ref})\, d\sigma$	131.8	4263.1	16130.5

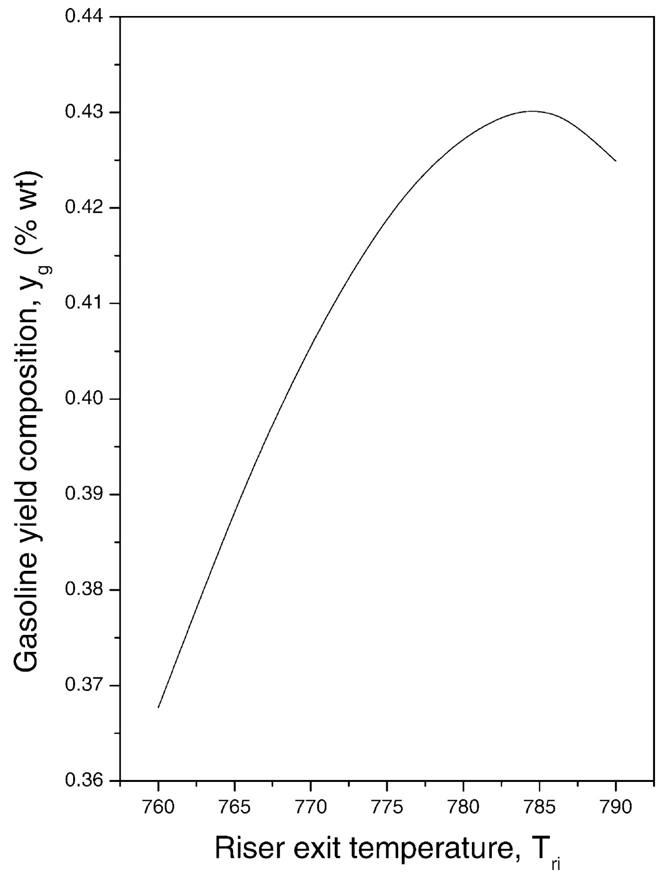

Fig. 5. Input–output riser exit temperature-gasoline yield composition steady-state map.

5.2. Gasoline regulation with multivariable cascade control configurations

We consider both regulation of regenerator cyclone temperature at $T_{cy,ref} = 997.4$ K and gasoline yield composition at $y_{g,ref} = 0.43$ wt.%. It can be seen from Fig. 5 that above of this value of $y_{g,ref}$ there is a sign change in the steady-state input–output process gain, which leads to an unstable compensation. At $t = 1250$ min, we consider the same disturbances as in the case of temperature stabilization. At $t = 2000$ min we consider set point changes in the

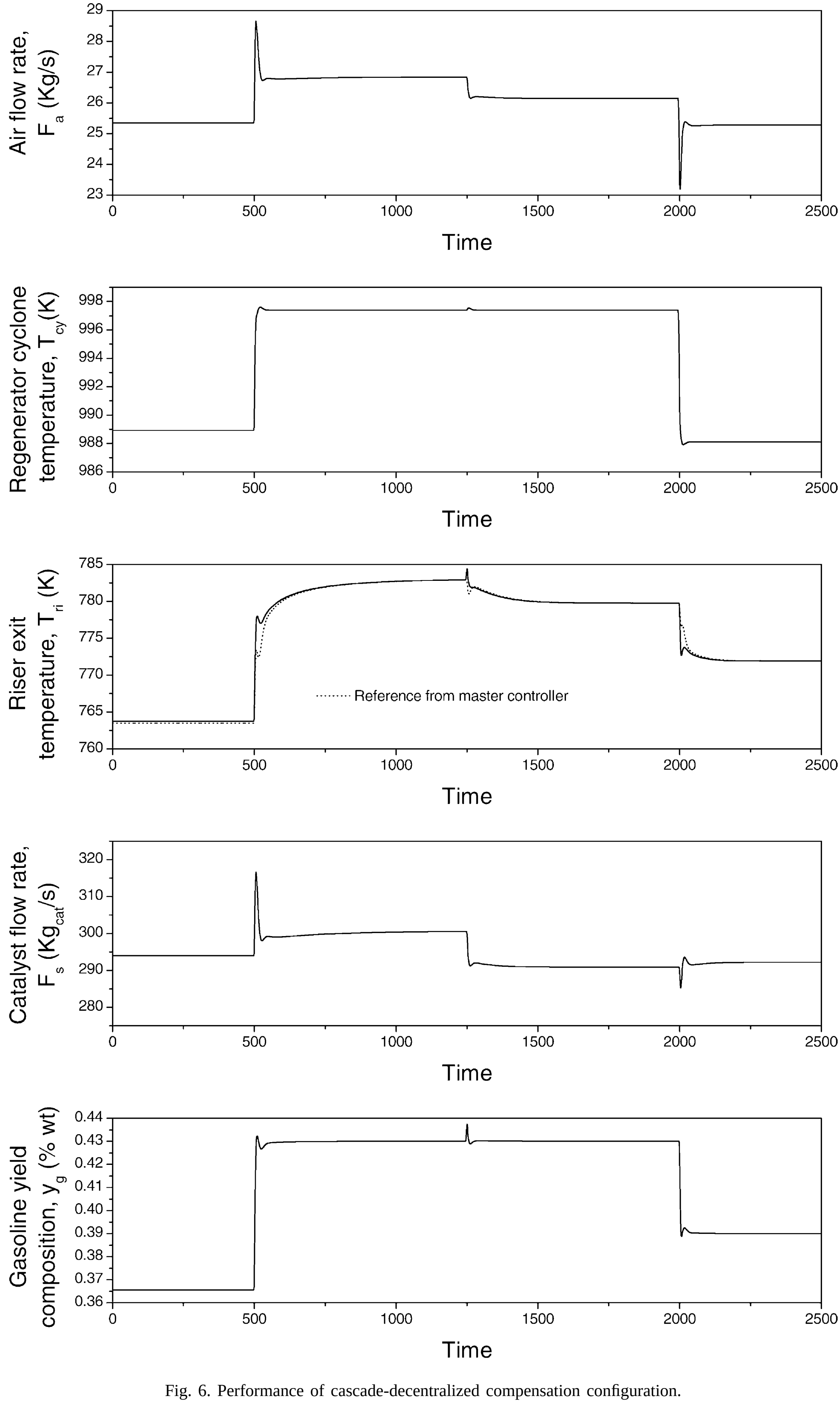

Fig. 6. Performance of cascade-decentralized compensation configuration.

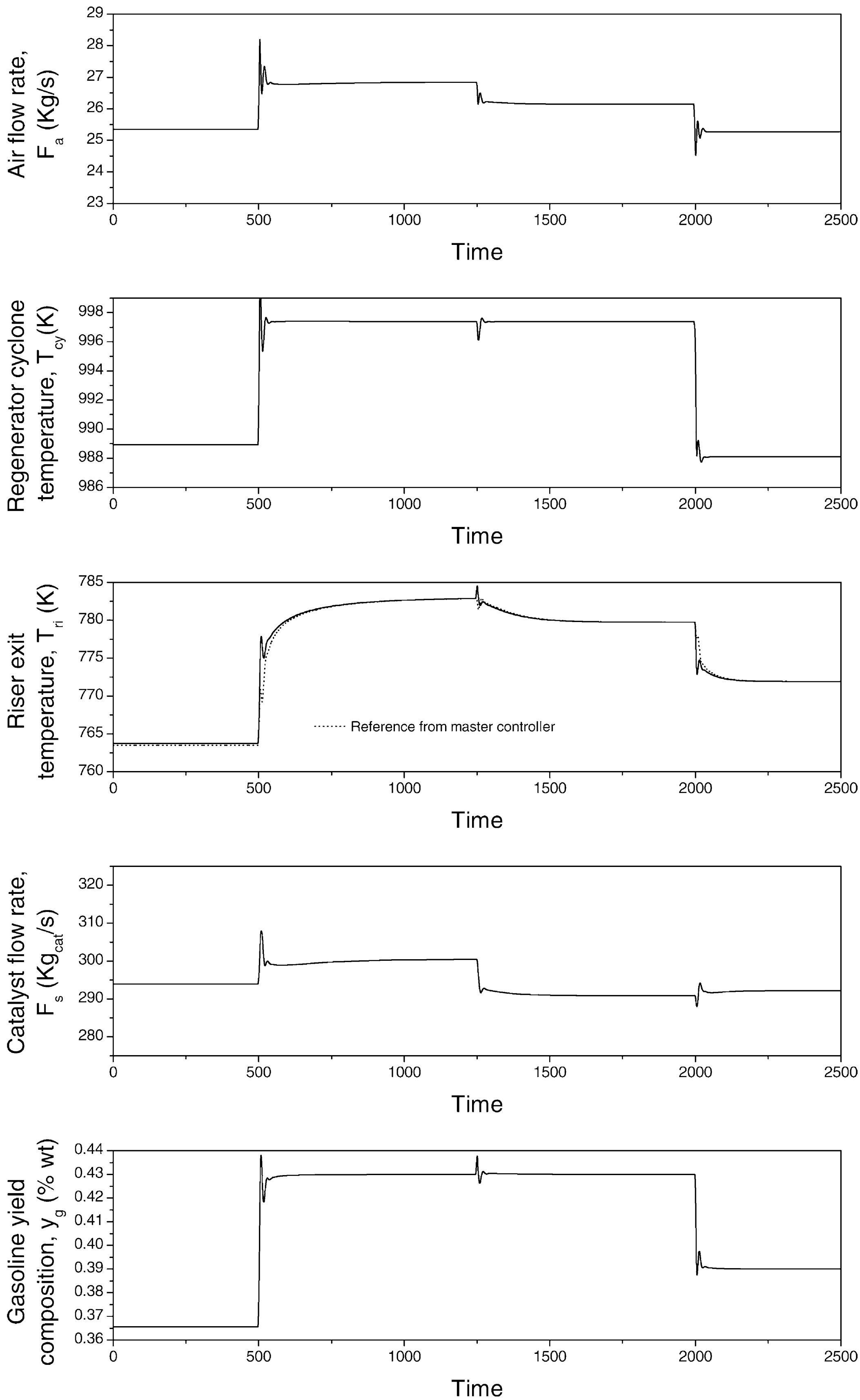

Fig. 7. Performance of cascade-decoupled compensation configuration.

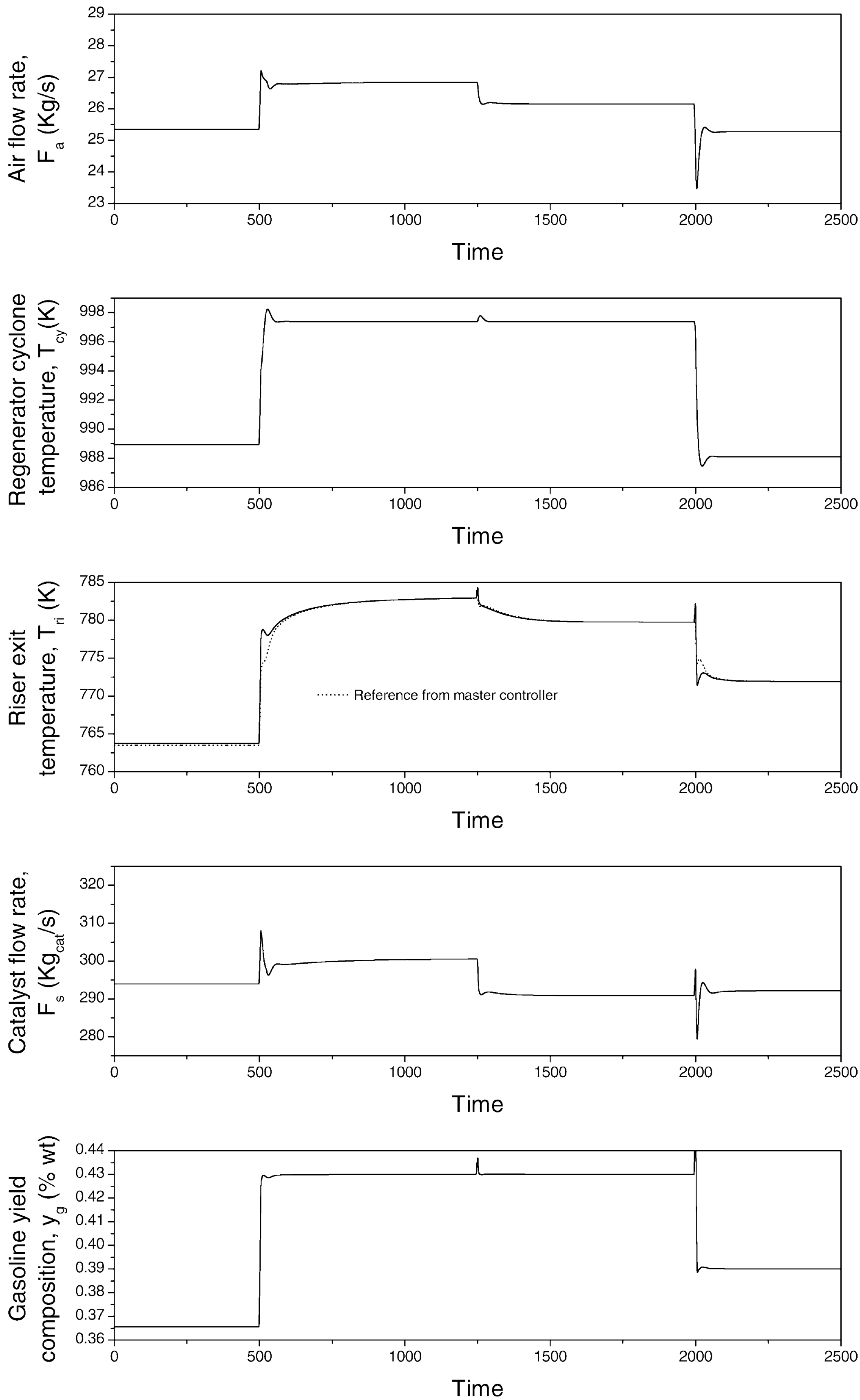

Fig. 8. Performance of cascade-transpose compensation configuration.

Table 2
Computed ISE values for the gasoline regulation loop

	Decentralized	Decoupled	Multiloop
$ISE_1 = \int (y_g - y_{g,ref})\, d\sigma$	0.2825	0.5030	0.2644

gasoline yield composition to $y_{g,ref} = 0.39$ wt.% and in the regenerator cyclone temperature to $T_{cy,ref} = 988.1$ K.

In the following simulations, it was assumed that composition measurements were available. Despite important advances in composition analyzer to provide near-real time composition measurements, still often such measurements are not available for on-line control, or the measurements are delayed. In some situations the composition can be inferred from available measurements (e.g. pressure, flows, etc.), which, however, is beyond the scope of this paper. In all simulations below, the measurement delay was assumed to be equal to 4 min.

Figs. 6–8 show the control performance for the cascade-decentralized compensation, cascade-decoupling compensation and cascade-transpose compensation configurations cases respectively. The tuning parameters α_p, α_i for the three control configurations are given as in Fig. 4. The integral gain in the master controller loop is tuned with IMC rules based on the steady-state gain obtained form the input–output map shown in Fig. 5. The estimated steady-state gain is $K_{ss} = 1240.0$ and the dominant time-constant $\tau_d = 45$ min, which corresponds to the slower time-constant obtained from the step responses from Fig. 2. We have set the $\bar{T}_{ri} = 763.0$, which was chosen arbitrarily with the only constraint of begin contained in the operating domain $I_T = [750, 785]$.

As can be seen from Figs. 6–8, the slave controller tracks the output signal from the master controller within a very good performance. Moreover, the control configurations shows good disturbance rejection and set-point following capabilities. Table 2 shows the ISE computed values for gasoline composition regulation. It is clear that, although the performance of the three configurations is comparable in the set point changes and disturbance rejection, the cascade-decentralized compensation outperforms both the cascade-decoupled and cascade-transpose compensation cases in controlling the gasoline yield composition.

6. Conclusions

In this paper, we have presented a methodology for multivariable cascade composition control for FCC units. In a first step, a multivariable temperature regulation loop was designed on the basis of simple linear models, and the stability of the control loop is guaranteed in terms of a control structure matrix. In this way, we have found that decentralized control configurations presents advantages over decoupling and transpose ones because unconditional stability of the control loop is found for the former configurations. This stability property of decentralized structures induces interesting robustness properties, such as no accurate knowledge of steady-state gains and time-constants requirements. As a consequence, the control design and tuning effort are simplified. These findings have been corroborated with numerical simulations on a non-linear model of the FCC unit. In a second step, the previously designed temperature regulation loop is used as a framework to provide feedback regulation composition to the FCC unit. This is done by designing a simple integral control that uses delayed composition measurements to provide temperature references to the temperature regulation loop. Departing from IMC ideas, practical tuning guidelines are provided and tested with numerical simulations. Overlay, our results shown that automatic composition regulation at the riser exit can be implemented with linear control configurations derived from simple input–output response models.

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