COMBINED GAIN-SCHEDULING AND MULTIMODEL CONTROL OF A REACTIVE DISTILLATION COLUMN

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Abstract: Reactive distillation (RD) is a favourable alternative to conventional series of reaction-separation processes. Control of RD is challenging due to its integrated functionality and complex dynamics. Linear PID algorithm is not satisfactory and needs because of the need for adequate retuning over a wide range of operating conditions. Combined gain-scheduling and multimodel control scheme is proposed to handle the nonlinearities of the process. Simulation results show the superior performance of the proposed method to that of a standard PI control.
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Keywords: reactive distillation, gain-scheduling, multimodel, switching, tuning.

1. INTRODUCTION

The RD column is gradually becoming an important unit operation in chemical process industry. It offers reduction in both investment as well as operational costs. Tight control of product composition and conversion is fundamental for an economically optimal operation. Unfortunately, both composition and conversion cannot be economically and reliably measured on-line and in real time. Moreover, the relationship between the product composition and the potential manipulated variable (eg. reboiler duty) may reveal multiplicity. Inferential control via stage temperatures, which have monotonic relationship with the manipulated variable, is commonly adopted.

Directionality of a chemical process means that a vector of inputs (eg. manipulated variables) is differently amplified according to its direction. It has been known to create considerable complex problem in control system design for multivariable processes such as in conventional distillation (Sågfors and Waller, 1995). Standard PID with fixed parameters is not satisfactory because of the need for adequate retuning over a wide range of operating conditions.

Inferential control of RD, which has directionality in the process gain, is investigated in this study. Limited number of reports has discussed control aspects of RD. Control strategies of batch RD (Sorensen and Skogestad, 1994) and its structure for optimisation (Wajge and Reklaitis, 1999) have been investigated. Recently, nonlinear control of batch RD has been proposed (Balasubramhanya and Doyle III, 2000).

For continuous RD, a nonlinear input-output linearizing controller and nonlinear controller have been designed for ethylene glycol system (Kumar and Daoutidis, 1999). A robust PI control scheme has been proposed for the same system (Loperena et al., 2000). Linear and nonlinear control strategies have been applied for an ethyl acetate system (Vora and Daoutidis, 2001). A variety of control structures have also been explored for two product RD (Al-Arfaj and Luyben, 2000).

For ethyl tert-butyl ether (ETBE) RD, which is the focus of this study, general control considerations have been presented (Sneesby et al., 1997). Combined composition and conversion control have been discussed (Sneesby et al., 1999). Control performance of a variety of one-point control schemes has been compared (Bisowarno and Tadé, 2002). Pattern-based predictive control has recently been proposed for controlling the product composition (Tian et al., 2003). Effectiveness of control schemes has been compared for single and double-feed RD (Al-Arfaj and Luyben, 2002). Standard PI algorithms, which were employed for all cases, indicated more advanced controller is required to improve the control performance.

In this study, combined gain-scheduling and multimodel control will be implemented on one-point control of an ETBE RD. The models cover directionality of the process gain and a switching scheme will be employed to integrate them. Its performance will be compared to that of a standard PI controller.

2. REACTIVE DISTILLATION

A pilot scale packed RD column for ETBE production serves as an example for a typical single-feed two-products RD process. The column consists of 1 rectifying stage, 3 reactive stages, 4 stripping stages, a total condenser, and an electric partial reboiler, respectively, as shown in Figure 1. The feed is a mixture of isobutylene, ethanol, ETBE, and n-butane, resulting from a pre-reactor, which converts most of isobutylene to ETBE. Typical operating data including the operating range are summarised in Table 1. The primary and secondary manipulated variables are reboiler duty (Q_R) and reflux rate (L_R), respectively. LV control scheme, which outperforms other control schemes for this column (Bisowarno and Tadé, 2002), is employed.

Inferential control is adopted to control the ETBE purity. The relationship between the purity and the reboiler duty reveals input multiplicity phenomena as shown in Figure 2. Based on the sensitivity analysis, stage 7 temperature is found to be the most appropriate measured variable to infer the ETBE purity (Tian and Tadé, 2000). Figure 2 also shows the relationship between the stage 7 temperature and the reboiler duty. The nonlinear process gain ($\Delta T_7/\Delta Q_r$) is large around the nominal operating condition and becomes small outside this range.
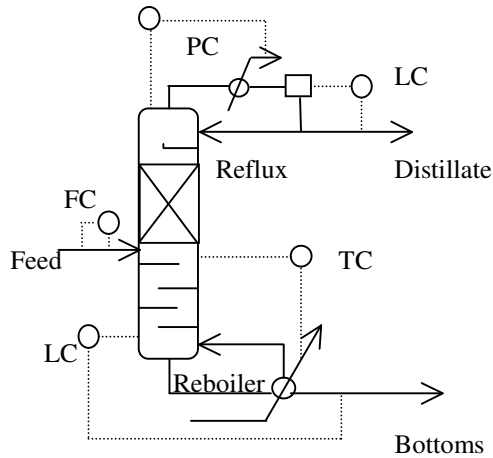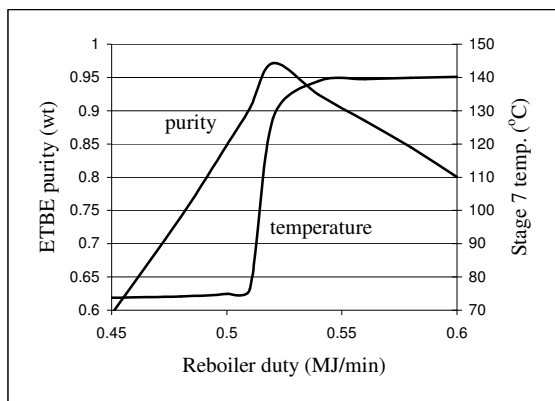

Fig. 1 ETBE Column with the controllers

Fig.2 Stage 7 temp./purity vs. Q_R (L_R = 2.2 l/min)

Table 1 ETBE RD column characteristics and Inputs

Column Specifications:		
Nre/Nrx/Nst	3/3/5	
Feed stage	6	
Overhead pressure	950 kPa	
Feed Conditions:	Range	Nominal*
Temperature	30°C	30°C
Rate (l/min)	0.684-0.836	0.76
Comp. (mol)	70–80 mol% conversion in the pre-reactor	0.291 ETBE, 0.091 EtOH, 0.073 iBut, 0.545 nBut
Man. Variables:	Range	Nominal*
L_R (l/min)	2.0 – 2.4	2.2
Q_R (MJ/min)	0.4825-0.555	0.520

* Nominal (optimum) operating condition for designing the control system

3. CONTROL OBJECTIVES

The main objective of the control system is to keep the controlled stage 7 temperature close to the set-points despite the presence of disturbances. The most significant disturbances are changes in the feed flow rate and in the feed composition. The second objective is a sufficiently fast set-point tracking. These two objectives must be achieved for the entire operating range of the reactive distillation column.

4. CONTROL DESIGN METHOD

Adaptive control with multimodel was introduced in (Narendra and Balakrishnan, 1997). The basic idea is to choose the best model for the column from an a priori known set of models at every instant, and then apply the output of the corresponding controller to the column. Firstly, the process identification is performed by rapidly choosing the smallest error with respect to a criterion (switching). Unlike the previous work, the controller parameters are then adjusted using a parameter-adaptation algorithm in this study (gain-scheduling).

4.1 Multimodel

Although a single highly nonlinear and/or adaptive model may be used to represent the process dynamics, several simple fixed multimodels are employed. They are chosen to cope with nonlinear and time varying characteristics of each operating condition point. A proper switching scheme is needed to integrate the models. As a result, process identification and rapid control action can be satisfied.

Simplified input-output dynamic models of the manipulation and disturbance paths are identified

Table 2 Multimodels based on open-loop tests

L_R (l/min)	Q_R, min	Q_R, nom	Q_R, max
2.0	$\dfrac{4709.5}{238.4Ti\,s + 1}$	$\dfrac{4679.5}{78.5Ti\,s + 1}$	$\dfrac{498}{21\,Ti\,s + 1}$
2.2	$\dfrac{6442.75}{197.4Ti\,s + 1}$	$\dfrac{4675}{73.2\,Ti\,s + 1}$	$\dfrac{493}{23.2\,Ti\,s + 1}$
2.4	$\dfrac{960.5}{122.4Ti\,s + 1}$	$\dfrac{9043.5}{126.1Ti\,s + 1}$	$\dfrac{1472}{54.9Ti\,s + 1}$
Disturbances at L_R = 2.2 l/min			
Feed rate	$\dfrac{0.412}{23.75Ti\,s + 1}$		
Feed comp.	$\dfrac{0.163}{6.75Ti\,s + 1}$		

around the optimum reboiler duty at constant reflux rate. Referring to Figure 2, three simplified models are derived to capture the nonlinearity of the process gain for each constant reflux rate. At each of the reboiler duty of 0.4825, 0.520, and 0.555 MJ/min, respectively, the models are derived at the reflux rate of 2.0, 2.2, and 2.4 l/min, respectively. The models of disturbance patch are derived at the optimum reboiler duty of 0.520 MJ/min. Table 2 shows the models formulated as first order transfer functions.

4.2 Switching scheme

The switching scheme involves firstly monitoring a performance index based on the identification error for each model and then switching to the model with smallest index. A small identification error leads to a small tracking error (Narendra and Balakrishnan, 1997). The performance index (IE) is formulated in equation 1,

$$IE = \alpha\,\varepsilon_i^2 + \beta \int \varepsilon^{-\lambda(t-\tau)}\,\varepsilon_i^2\,dt,\ \alpha \geq 0 \text{ and } \beta > 0 \qquad (1)$$

where α and β are the weighting factors on the instantaneous measures and the long-term accuracy, respectively. These two free design parameters provide smooth transition between different process models. ε_i is the difference between the outputs of the model and the real plant.

4.3 Gain Scheduling

Gain scheduling is based on linear time-invariance of the process at a number of operating points. A linear controller is then designed for each operating point. Therefore, the parameters of the controller should be interpolated or scheduled (Rugh and Shamma, 2000). The controller gain is commonly scheduled due to the process nonlinearity with constant dynamics.

Switching between local linear controllers is a conventional way for gain scheduling. A function of a scheduling variable can also be employed to interpolate the gain. Measured ouput or set-point may be used as a scheduling variable (Bequette, 2000).
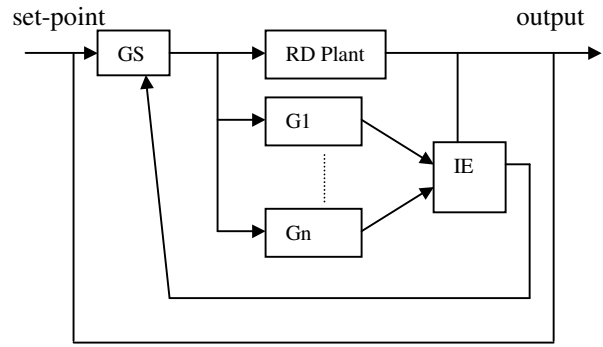

Fig. 3 Combined gain-scheduling and multimodel control scheme

In this study, the scheduling controller gain (Kc) is formulated in equation 2 (Ogunnaike and Ray, 1994),

$$Kc = \frac{Kco\,Kpo}{Kp} \qquad (2)$$

where Kco and Kpo are the reference values of the controller gain and process gains, respectively. The time varying process gain (Kp) is identified and computed on-line from the inferred variable and the manipulated variable.

The reference control parameters are tuned by using Abbas method (Abbas, 1997). This method relates the controller parameters to the characteristics of a first-order plus time delay model and the desired over-shoot of the closed loop system (Alexander and Trahan, 2001). For a PI controller, the controller parameters are formulated in equation 3,

$$Kc = \frac{T + \theta/2}{Kpo\,(\lambda + \theta)} \qquad (3a)$$

$$Ti = T + \theta/2 \qquad (3b)$$

where Kpo, T, and θ are the open-loop process gain, time constant, and time delay, respectively, and λ is the desired closed-loop time constant.

The general structure of the combined gain-scheduling and multimodel control scheme is shown in Figure 3. G1, Gn, GS, and IE are the simplified model 1, the simplified model n, the gain-scheduling controller, and the performance index, respectively.

5. CONTROL PERFORMANCE

The control performance of the proposed method is compared to that of a standard PI controller. The desired closed-loop time constant is chosen to be 5 min. Applying the Abbas method, the controller gain and time constant of the PI controller in the range of operating conditions are 0.00622 ^{o}C/(MJ/min) and

84.53 min, respectively. For the proposed method, the time constant is kept constant at 84.53 min while the controller gain is computed on-line as a function of the scheduling variable.

Figures 4 and 5 show the dynamic responses when the feed flow rates (F_f) increase or decrease steeply by using either standard PI or the proposed method, respectively. The Figure shows that the disturbance rejection of the proposed method is superior to the standard PI controller.

Figure 6 shows the dynamic response resulting from step change in the feed composition (F_c). The changes were represented by changes in the pre-reacted feed from 80 to 70 mol% of the isobutylene conversion. Figure 6 shows the proposed controller, which can tightly control the stage 7 temperature, does not keep the purity at the set-point value. This results from the model mismatch. Although more models can be employed to enhance the control performance, the intrinsic problem remains. This problem results from the difficulties to infer composition from VLE temperature measurements in multi-component mixture.

Figures 7 and 8 show the dynamic responses for step changes in the set-point value (T_7). The proposed method clearly has better set-point tracking as shown by shorter settling time.

The corresponding values of the integral absolute error (IAE) and integral of time-weighted absolute error (ITAE) criteria are shown in Table 3. The criteria confirm the previous analysis that the control performance can be improved by using the proposed method.
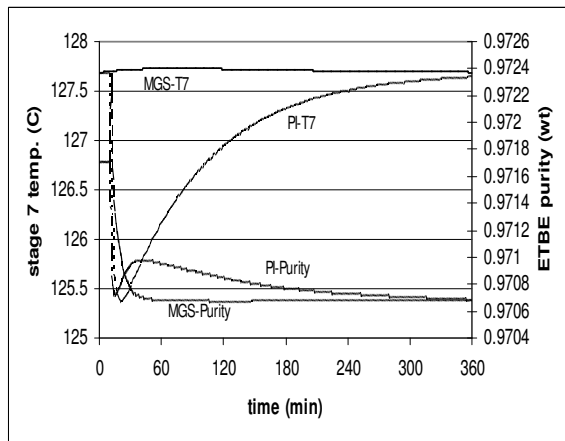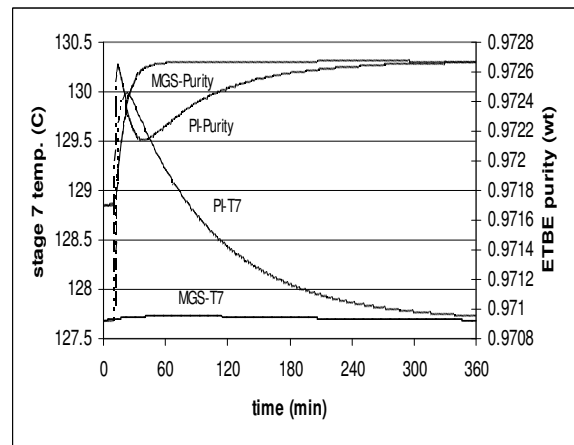

Fig. 5 Dynamic responses due to –10% Feed rate step change

Fig. 6 Dynamic responses due to feed concentration step change

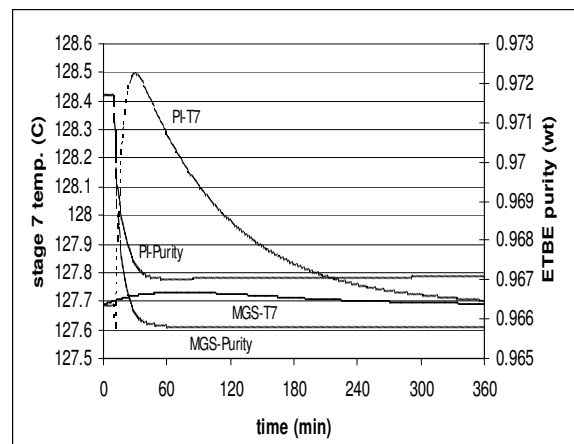
Fig. 4 Dynamic responses due to +10% Feed rate step change
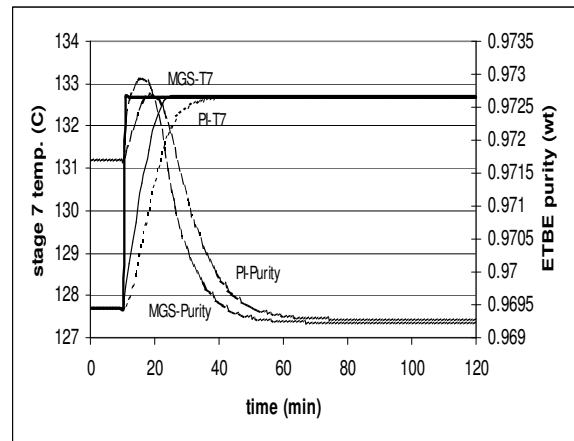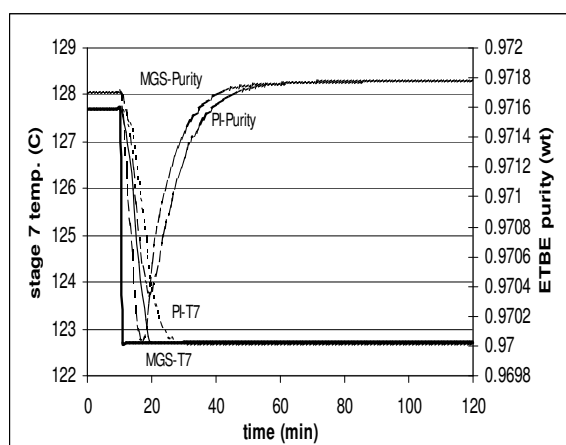

Fig. 7 Dynamic responses due to +5 set-point step change

Fig. 8 Dynamic responses due to –5% set-point step change

Table 3 Comparison of IAE and ITAE indices

Magnitude	IAE		ITAE	
	PI	GS*	PI	GS*
+10% F_f	219	8.3	19990	1120
-10% F_f	218	8.3	19877	1113
80-70%conv. (F_c)	81	8.3	7741	1117
+5°C T_7	19	4.1	389	288
-5°C T_7	15	4.3	334	294

* The proposed method using gain-scheduling and multimodel control scheme

6. CONCLUSIONS

The combined gain-scheduling and multimodel control has been applied to a RD column for ETBE production having nonlinearity in the process gain. Several input-output first order models were derived. Gain-scheduling control was then implemented employing the multimodel for model identification and scheduling the controller gain. The proposed controller is superior to standard PI control with fixed parameters for RD columns. However, the effectiveness is reduced for feed composition disturbances. This work clearly demonstrates that nonlinear processes can be controlled successfully with a linear multimodel concept.

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