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MPC of distillation column with side reactors for methyl acetate[☆]Cuimei Bo¹, Jun Li^{1,*}, Lei Yang¹, Hui Yi¹, Jihai Tang², Xu Qiao²¹ College of Automation and Electrical Engineering, Nanjing Tech University, Nanjing 210009, China² State Key Laboratory of Materials-oriented Chemical Engineering, Nanjing Tech University, Nanjing 210009, China

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ABSTRACT

This paper focuses on the dynamic control of distillation column with side reactors (SRC) for methyl acetate production. To obtain the optimum integrated structure and steady state simulation, the systematic design approach based on the concept of independent reaction amount is applied to the process of SRC for methyl acetate production. In addition to the basic control loops, multi-variable model predictive control modular with methyl acetate concentration and temperature of sensitive plate is designed. Then, based on process simulation software Aspen Plus, dynamic simulation of SRC for methyl acetate production is used to verify the effectiveness of the control scheme.

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1. Introduction

Traditional reactive distillation is the reaction coupled with distillation in the same equipment. The process has advantages of reaction heat full use, device integration enhancing and costs saving, etc. [1–3]. However, traditional reactive distillation must meet the strict demands of application systems, for example, the condition of reaction and distillation should keep consistent, with the limitation of reaction by the distillation column. The new technology of distillation column with side reactors (SRC) can not only increase the conversion rate and selectivity, but also overcome the limitations of traditional reactive distillation process [4].

SRC operates more flexibly, increasing the flexibility of control system design. Due to different structures and control objectives, 8 different control structures proposed by Luyben *et al.* may be used for reference, Al-Arfaj and Luyben [5,6], Kaymak and Luyben [7], Kaymak and Luyben [8]. Based on choosing controlled variables and manipulated variables reasonably and designing different control structures, the best control scheme can be determined with the analysis of system disturbance rejection performance. The dynamic model of SRC was built by Kaymak and Luyben [9], and CS5 and CS7 were studied by them as well. Balasubramhanya and Doyle [10] proved that tight control of the process may be obtained with the use of reduced model in a model predictive control algorithm. Recently, Sudibyo *et al.* [11] compared MIMO MPC

and PI Decoupling in Controlling Methyl Tert-butyl Ether Process. Sharma and Singh [12] used PID, MPC and NNPC for controlling a TAME reactive distillation column, and it is found that NNPC and MPC give better control performance than the PID controller. Due to strong nonlinearity and coupling of SRC, conventional linear control methods cannot meet the control requirements of the special working conditions [13,14]. Thus, more and more applications of advanced control in special reaction distillation process are used. Particularly, MPC is suitable for industry, to keep the system running smoothly with disturbance rejection.

In this paper, a distillation column with side reactors (SRC) for methyl acetate production is used as an example. Optimal steady-state is designed by the systematic design approach based on the concept of independent reaction amount, and a conventional multi-variable control scheme is studied with model predictive control (MPC). Finally, the control performance is verified by dynamic simulation using Aspen Plus.

2. Distillation Column with Side Reactors (SRC) Process for Methyl Acetate Production

2.1. Methyl acetate production

Methyl acetate reaction is the production process which belongs to the typical reactive distillation process. Methyl acetate process uses methanol and acetic acid as the production of raw materials, the production is methyl acetate. Reaction equations are as follows:



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It is a reversible reaction. And by using a rate control model, the kinetic equation for benzene chloride is as follows [15]:

$$r = m_{\text{cat}}(k_+ \alpha_a \alpha_b - k_- \alpha_c \alpha_d) \quad (2)$$

$$k_+ = 2.961 \times 10^4 \exp\left(\frac{-49190}{RT}\right) \quad (3)$$

$$k_- = 1.384 \times 10^6 \exp\left(\frac{-69230}{RT}\right) \quad (4)$$

where r is the reaction rate, m_{cat} is the catalyst's quality, and k_+ and k_- are positive and reverse reaction rate constants respectively; α_a , α_b , α_c , and α_d are component activities respectively. The physical properties of methanol (MeOH), acetic acid (HAc), methyl acetate (MeAc) and water (H₂O) are shown in Table 1.

Table 1
Material properties

	HAc	MeOH	MeAc	H ₂ O
Boiling point/°C	117.9	65.4	57.05	100
Molar mass/g·mol ⁻¹	60.05	32.04	74.07	18.01

2.2. Steady-state optimization

The optimal steady state configuration parameters and operating parameters will be used to design the control system. And the steady-state optimal parameters, such as product quality, reaction conversion rate and production cost will be the control objectives of the control scheme, which can be obtained on the basis of an optimal economic design to minimize the total annual cost [16]. The influence of different numbers of reactors, space of trays, feed locations and reflux ratio are studied separately, as shown in Fig. 1, to get a reasonable model.

Based on the concept of independent reaction amount, a systematic design approach, the optimal steady-state is obtained [17]. The optimal configuration of the reactive distillation column is given in Fig. 2. It is assumed that the number of separation stages between each reactor is the same. Therefore, there are mainly five independent design parameters to be optimized with the given boil up rate: the number of stripping stages in the separation zone (NS), the number of stages between each reactor in the reaction zone (NRS), the number of reactors (NR), the feed flow rate of methanol (MeOH), and the feed flow rate of acetic acid (HAc). In the configuration, the light product, methyl acetate (MeAc), is removed from the top, while the heavy product, water (H₂O) is taken out from the bottom. The feed streams of acetic acid and methanol are fed on the fourth tray and 25th tray, respectively. The column pressure is at atmospheric, the feed flow rates of acetic acid and methanol are both 50 kmol·h⁻¹, and the flow rate of produced methyl acetate is 50 kmol·h⁻¹. The reflux ratio of the column is $R = 1.5$. There are five side reactors of the methyl acetate production; the first reactor is connected with distillation in the 26th tray, the second in the 22th tray, and so on.

Fig. 3 shows the vapor–liquid phase flow rate profiles for such optimally designed column. It is observed that the reactants can be fully reacted and slowly separated in the reaction section, to promote the separation by reaction. At the same time, separation can be fully performed when the reactants are beyond the reaction section with the large consumption of reactants.

3. Controller Scheme Design of SRC for Methyl Acetate Production

3.1. Traditional control scheme design of SRC

In order to guarantee the smooth running of the SRC for methyl acetate production, controlled variables and the corresponding

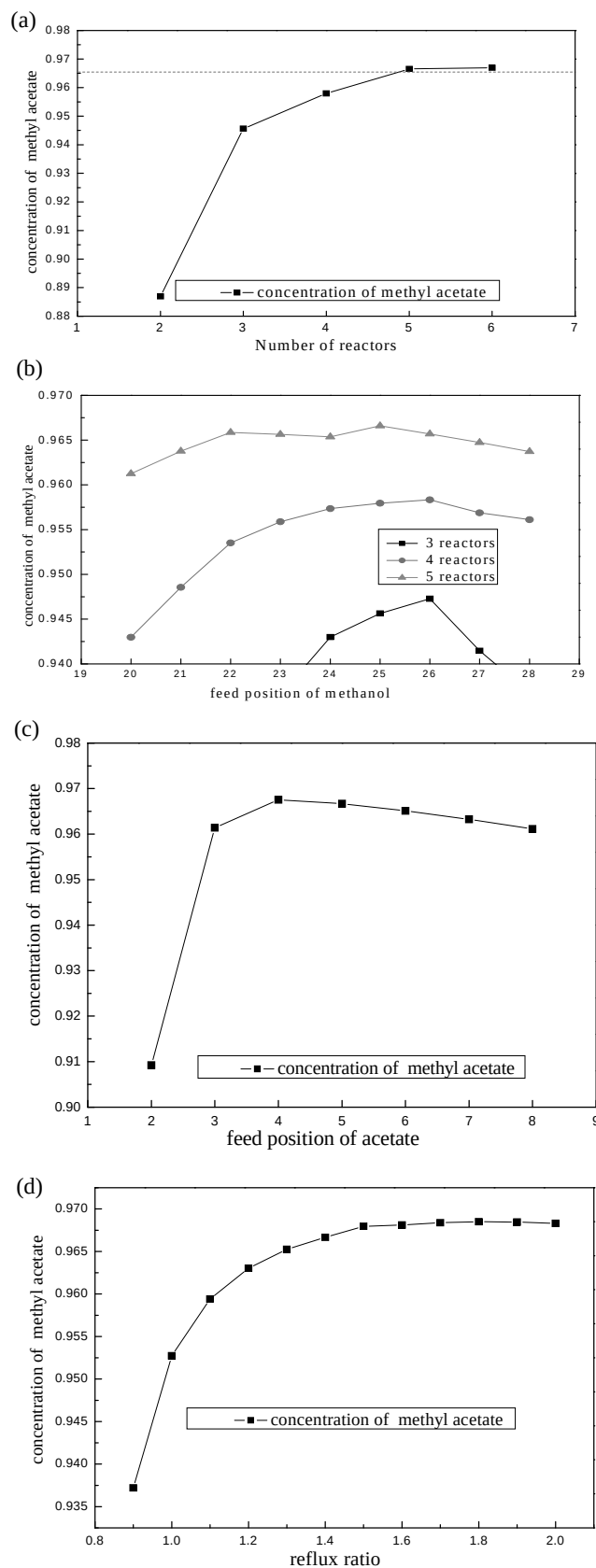


Fig. 1. Optimal design and analysis of SRC for methyl acetate production. (a) Effects of different numbers of reactors. (b) Effects of different feed positions of methanol. (c) Effects of different feed positions of acetate. (d) Effects of different reflux ratios.

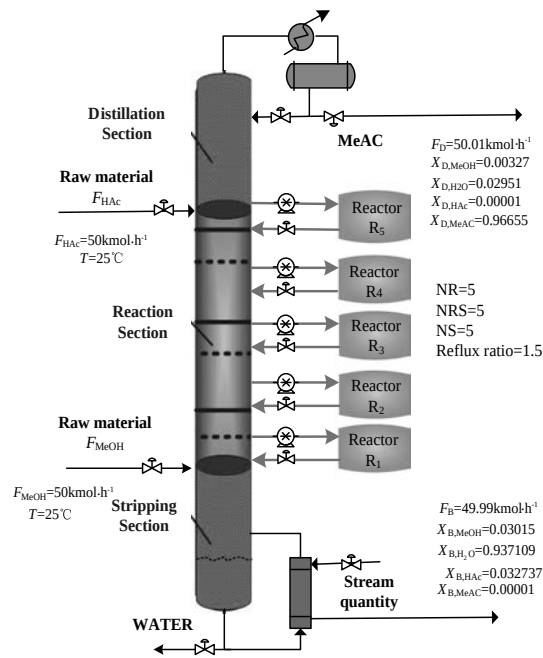


Fig. 2. Optimum design of SRC for methyl acetate production.

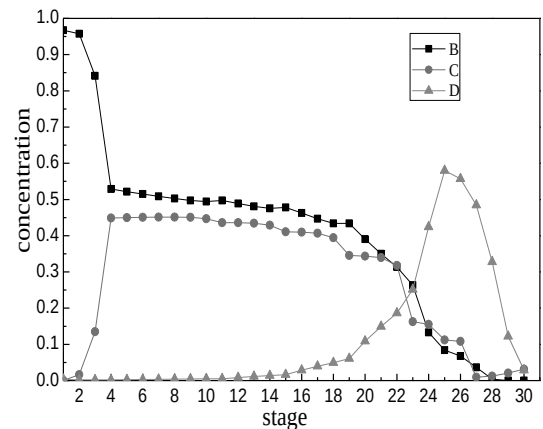


Fig. 3. Composition profiles of the optimal steady state configuration.

manipulated variable may be chosen at first. Basic control loops of the SRC process are shown in Table 2, including LIC100 (level of condenser controlled by overhead circulatory flow), LIC101 (level of reboiler controlled by flow-out of column bottom), PIC100 (overhead pressure controlled by condensation amount), FIC100 and FIC101 (feed rate of

Table 2

Control loops of scheme

Control loops	Controlled variables	Control objectives	Manipulated variables
LIC-100	Condenser level	50%	Overhead circulatory flow
LIC-101	Reboiler level	50%	Flow-out of column bottom
PIC-100	Overhead pressure	101.3 kPa	Condensation amount
FIC-100 (sub-loop)	Feed rate of MeOH	50 kmol · h ⁻¹	Feed rate of HAc
FIC-101 (main-loop)	Feed rate of MeOH	50 kmol · h ⁻¹	Feed rate of MeOH
LIC201–LIC205	Level of reactors	50%	Flow-out of the reactors
TIC201–TIC205	Temperature of reactors	65 °C	Calefaction heat quantity

MeOH controlled by the ratio of feed rate of HAc and MeOH), LIC201–LIC205 (level of reactors controlled by flow-out of the reactors) and TIC201–TIC205 (temperature of reactors controlled by calefaction heat quantity). And the multi-variable basic control structure is shown in Fig. 4.

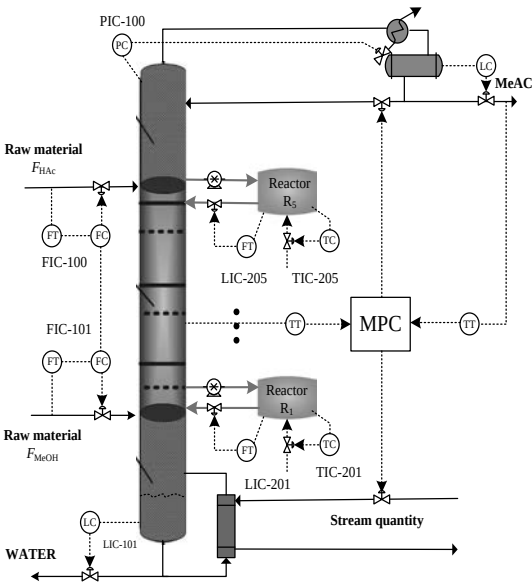


Fig. 4. The structure scheme of the decentralized control.

3.2. Prediction model identification

The SRC process for methyl acetate production is constructed by the distillation column coupled with 5 reactors. Due to the strong coupling of reaction and distillation as well as materials and energy, conventional control is hard to make accurate control. Based on the basic PID control loops, a 2 × 2 multi-variable MPC controller is designed, with quantity of reflux and calefaction heat quantity of reboiler being manipulated variables and overhead methyl acetate concentration and temperature of sensitive plate being controlled variables, as shown in Fig. 4.

Continuous output response can be obtained by introduction of persistent excitation to the input [18]. Sampling period of the system is 0.01 h, and 10 hours' data (1000 groups) are collected to identify the model, including 800 groups training data and 200 groups extrapolated calibration data. Input signal is shown in Fig. 5 and system response under excitation is shown in Fig. 6.

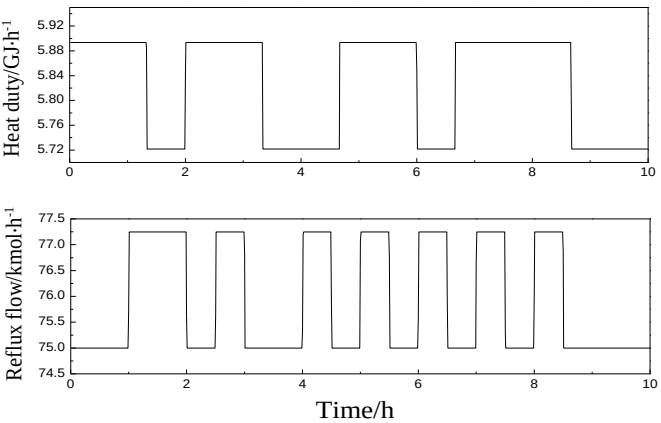


Fig. 5. M sequence test signal.

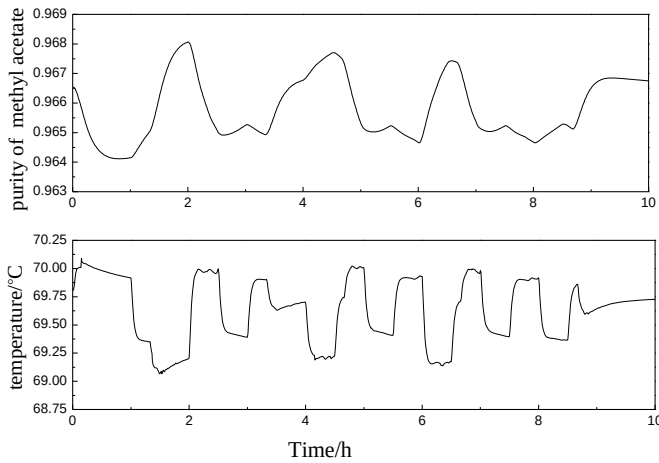


Fig. 6. System output response.

The state-space model of the system can be obtained by identifying the sampling data using improved least squares algorithm. Assume that the above system is a linear time-invariant system as follows:

$$\mathbf{x}(k+1) = \mathbf{A}\mathbf{x}(k) + \mathbf{B}\mathbf{u}(k) + \mathbf{K}\mathbf{e}(k) \quad (5)$$

$$\mathbf{y}(k) = \mathbf{C}\mathbf{x}(k) + \mathbf{D}\mathbf{u}(k) + \mathbf{e}(k) \quad (6)$$

where $\mathbf{u}(t) \in \mathcal{R}^m$ is the system input measurement, $\mathbf{y}(t) \in \mathcal{R}^l$ is the system output measurement, $\mathbf{x}(t) \in \mathcal{R}^n$ is the system process state, $\mathbf{e}(t) \in \mathcal{R}^l$ is the output measurement noise, $\mathbf{K}\mathbf{e}(t) \in \mathcal{R}^n$ is the system process noise, and $(\mathbf{A}, \mathbf{B}, \mathbf{C}, \mathbf{D})$ are the corresponding coefficient matrices.

We can make $\mathbf{K}\mathbf{e}(k) = \mathbf{w}(k)$, $\mathbf{e}(k) = \mathbf{v}(k)$.

By changing, the equation as follows can be available:

$$\mathbf{Y}_M(k) = \mathbf{\Gamma}_M \mathbf{X}(k) + \mathbf{H}_M \mathbf{U}_M(k) + \mathbf{H}_M^w \mathbf{W}_M(k) + \mathbf{V}_M(k) \quad (7)$$

$$\begin{aligned} \mathbf{Y}_M(k) &= [y(0) \ y(1) \ \dots y(M)]^T \\ \mathbf{U}_M(k) &= [u(0) \ u(1) \ \dots u(M)]^T \\ \mathbf{W}_M(k) &= [w(0) \ w(1) \ \dots w(M)]^T \\ \mathbf{V}_M(k) &= [v(0) \ v(1) \ \dots v(M)]^T \end{aligned}$$

$\mathbf{\Gamma}_M$ is generalized observability matrix, M is the block line number, and $\mathbf{H}_M, \mathbf{H}_M^w$ denote generalized controllability matrices respectively.

$$\begin{aligned} \mathbf{\Gamma}_M &= \begin{bmatrix} \mathbf{C} \\ \mathbf{C}\mathbf{A} \\ \vdots \\ \mathbf{C}\mathbf{A}^{M-1} \end{bmatrix} \\ \mathbf{H}_M &= \begin{bmatrix} \mathbf{D} & 0 & \mathbf{I} & 0 \\ \mathbf{C}\mathbf{B} & \mathbf{D} & \mathbf{I} & 0 \\ \vdots & \vdots & \vdots & \vdots \\ \mathbf{C}\mathbf{A}^{M-2}\mathbf{B} & \mathbf{C}\mathbf{A}^{M-3}\mathbf{B} & \mathbf{I} & \mathbf{D} \end{bmatrix} \\ \mathbf{H}_M^w &= \begin{bmatrix} 0 & 0 & \mathbf{I} & 0 \\ \mathbf{C} & \mathbf{D} & \mathbf{I} & 0 \\ \vdots & \vdots & \vdots & \vdots \\ \mathbf{C}\mathbf{A}^{M-2} & \mathbf{C}\mathbf{A}^{M-3} & \mathbf{I} & 0 \end{bmatrix} \end{aligned}$$

As is known, the sampling number of the system is N , the Hankel matrix can be constructed as follows:

$$\begin{aligned} \mathbf{Y}_M(k) &= [\mathbf{Y}_M(1) \ \mathbf{Y}_M(2) \ \dots \ \mathbf{Y}_M(N)] \\ \mathbf{U}_M(k) &= [\mathbf{U}_M(1) \ \mathbf{U}_M(2) \ \dots \ \mathbf{U}_M(N)] \\ \mathbf{W}_M(k) &= [\mathbf{W}_M(1) \ \mathbf{W}_M(2) \ \dots \ \mathbf{W}_M(N)] \\ \mathbf{V}_M(k) &= [\mathbf{V}_M(1) \ \mathbf{V}_M(2) \ \dots \ \mathbf{V}_M(N)] \\ \mathbf{X}_M(k) &= [\mathbf{x}(1) \ \mathbf{x}(2) \ \dots \ \mathbf{x}(N)] \end{aligned} \quad (8)$$

Then formula (7) can be written in the form of the Hankel matrix:

$$\mathbf{Y}_M(k) = \mathbf{\Gamma}_M \mathbf{X}(k) + \mathbf{H}_M \mathbf{U}_M(k) + \mathbf{H}_M^w \mathbf{W}_M(k) + \mathbf{V}(k) \quad (9)$$

By using oblique projection, noise $\mathbf{W}(t), \mathbf{V}(t)$ and future input $\mathbf{U}(t)$ can be eliminated, then, the equation can be obtained:

$$\mathbf{O}_M = \mathbf{\Gamma}_M \hat{\mathbf{X}}(k) \quad (10)$$

In this equation, $\hat{\mathbf{X}}(k)$ denotes estimated state sequence.

On the condition that $\mathbf{\Gamma}_M$ is full rank, we can execute SVD for the matrix $\mathbf{O}_M$:

$$\mathbf{O}_M = [\mathbf{U}_1 \ \mathbf{U}_2] \begin{bmatrix} \mathbf{S}_1 & 0 \\ 0 & \mathbf{S}_2 \end{bmatrix} \begin{bmatrix} \mathbf{V}_1^T \\ \mathbf{V}_2^T \end{bmatrix} \quad (11)$$

where $\mathbf{S}_1 \in \mathcal{R}^{n \times n}$, and $\mathbf{\Gamma}_M = \mathbf{U}_1 \mathbf{S}_1^{1/2}$. Then the state sequence can be obtained:

$$\hat{\mathbf{X}}(k) = \mathbf{\Gamma}_M^\dagger \mathbf{O}_M \quad (12)$$

where the symbol “ $\dagger$ ” denotes Moore–Penrose inverse. According to the state sequence, we can use the regression method to solve the system matrix, and solve the linear equation as follows:

$$\begin{bmatrix} \hat{\mathbf{X}}(k+1) \\ \mathbf{Y}(k) \end{bmatrix} = \begin{bmatrix} \mathbf{A} & \mathbf{B} \\ \mathbf{C} & \mathbf{D} \end{bmatrix} \begin{bmatrix} \hat{\mathbf{X}}(k) \\ \mathbf{U}(k) \end{bmatrix} + \begin{bmatrix} \rho_w \\ \rho_v \end{bmatrix} \quad (13)$$

by using least squares algorithm, the result can be obtained:

$$\begin{bmatrix} \hat{\mathbf{A}} & \hat{\mathbf{B}} \\ \hat{\mathbf{C}} & \hat{\mathbf{D}} \end{bmatrix} = \arg \min_{\mathbf{A}, \mathbf{B}, \mathbf{C}, \mathbf{D}} \left\| \begin{bmatrix} \hat{\mathbf{X}}(k+1) \\ \mathbf{Y}(k) \end{bmatrix} - \begin{bmatrix} \mathbf{A} & \mathbf{B} \\ \mathbf{C} & \mathbf{D} \end{bmatrix} \begin{bmatrix} \hat{\mathbf{X}}(k) \\ \mathbf{U}(k) \end{bmatrix} \right\| \quad (14)$$

In conclusion, state space matrix $(\mathbf{A}, \mathbf{B}, \mathbf{C}, \mathbf{D})$ of the system can be estimated. As for a further comment, a system state space model can be obtained, which is essential in follow-up work.

Fitting conditions of the model output and the actual output under the disturbances of temperature of sensitive plate and overhead purity of methyl acetate are shown in Figs. 7 and 8. We can see that the fitting degree of sensitive plate's temperature reaches 94.95%, and the fitting degree of the methyl acetate concentration is lower, which is still in the range of control requirements.

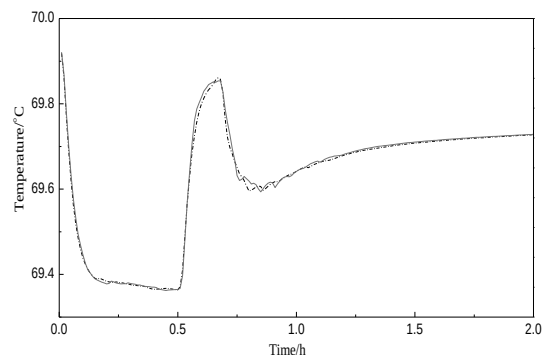


Fig. 7. Sensitive plate temperature alignment.

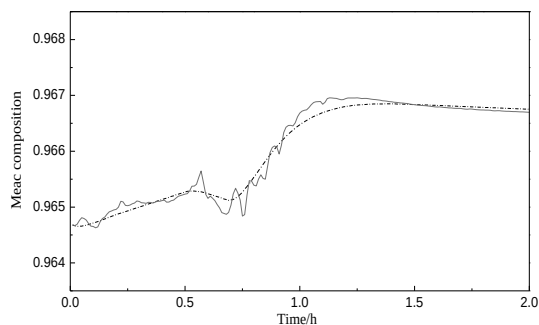


Fig. 8. Component alignment of methyl acetate.

3.3. Incremental prediction model

To obtain the prediction model of the system based on increment, operator Δ is defined as $\Delta x = f(x) - f(x-1)$, and Δ is applied into formula (5) as follows:

$$\Delta \mathbf{X}(k+1) = \mathbf{A} \Delta \mathbf{X}(k) + \mathbf{B} \Delta \mathbf{u}(k) + \mathbf{K} \Delta \mathbf{e}(k) \quad (15)$$

$\mathbf{Z}(k)$ is submitted into formulas (5) and (6) with the definition of $\mathbf{Z}(k) = \begin{bmatrix} \Delta \mathbf{X}(k) \\ \mathbf{y}(k-1) \end{bmatrix}$:

$$\begin{aligned} \mathbf{Z}(k+1) &= \mathbf{G} \mathbf{Z}(k) + \mathbf{H} \Delta \mathbf{u}(k) + \mathbf{L} \Delta \mathbf{e}(k) \\ \mathbf{y}(k) &= \mathbf{Q} \mathbf{Z}(k) + \mathbf{D} \Delta \mathbf{u}(k) + \Delta \mathbf{e}(k) \end{aligned} \quad (16)$$

where $\mathbf{G} = \begin{bmatrix} \mathbf{A} & \mathbf{0} \\ \mathbf{C} & \mathbf{E} \end{bmatrix}$, $\mathbf{H} = \begin{bmatrix} \mathbf{B} \\ \mathbf{D} \end{bmatrix}$, $\mathbf{L} = \begin{bmatrix} \mathbf{K} \\ \mathbf{1} \end{bmatrix}$, $\mathbf{Q} = [\mathbf{C} \quad \mathbf{E}]$, and the prediction model can be obtained by the above formulas:

$$\begin{aligned} \mathbf{y}(k+p|k) &= \mathbf{G} \mathbf{Q}^T \mathbf{Z}(k) + \sum_{j=1}^M \mathbf{G} \mathbf{Q}^{j-1} \mathbf{H} \Delta \mathbf{u}(k+j-1) + \\ &\sum_{j=1}^P \mathbf{G} \mathbf{L}^{j-1} \mathbf{H} \Delta \mathbf{e}(k+j-1) + \mathbf{D} \Delta \mathbf{u}(k+p) + \Delta \mathbf{e}(k+p) \end{aligned} \quad (17)$$

where P is the prediction horizon, $P = 1, \dots, p$; $k+p|k$ represent the prediction value; M is the control horizon, $P > M$, $P = 12$, $M = 5$. Formula (17) can be expressed by the following:

$$\mathbf{Y}(k) = \mathbf{G} \phi \mathbf{Z}(k) + \mathbf{y} \Delta \mathbf{u}(k) + \mathbf{G} \mathbf{Z} \Delta \mathbf{E}_1(k) \quad (18)$$

3.4. Optimization of prediction controller

Deviation between reference trajectory $\mathbf{y}_{\text{ref}}(k+j|k)$ and the predictive output is penalized by performance index function J :

$$\begin{aligned} J_k[\mathbf{x}(k, \mathbf{u})] &= w \times \sum_{j=1}^N \|\mathbf{y}(k+j|k) - \mathbf{y}_{\text{ref}}(k+j|k)\| + r \\ &\times \sum_{j=1}^{N_c-1} \|\Delta \mathbf{u}(k+j|k)\| \end{aligned} \quad (19)$$

where w is the error and r is the weight coefficient, $w = 1.3$, $r = 1$. Formula (19) is given in a matrix form:

$$J_k = \|\mathbf{Y}(k) - \mathbf{Y}_{\text{ref}}(k)\|_{\mathbf{Q}}^2 + \|\Delta \mathbf{U}(k)\|_{\mathbf{R}}^2 \quad (20)$$

The matrices $\mathbf{Y}(k)$, $\mathbf{Y}_{\text{ref}}(k)$, $\Delta \mathbf{U}$, $\mathbf{Q}$ and $\mathbf{R}$ are given by

$$\mathbf{Y}(k) = \begin{bmatrix} \mathbf{y}(k+1|k) \\ \vdots \\ \mathbf{y}(k+N|k) \end{bmatrix}, \mathbf{Y}_{\text{ref}}(k) = \begin{bmatrix} \mathbf{y}_{\text{ref}}(k+1|k) \\ \vdots \\ \mathbf{y}_{\text{ref}}(k+N|k) \end{bmatrix}, \Delta \mathbf{U}(k) = \begin{bmatrix} \Delta \mathbf{u}(k|k) \\ \vdots \\ \Delta \mathbf{u}(k+N_c-1|k) \end{bmatrix},$$

$$\mathbf{Q} = \mathbf{w} \times \mathbf{E}, \mathbf{R} = \mathbf{r} \times \mathbf{E}.$$

The matrix $\mathbf{E}(k)$ is denoted by $\mathbf{E}(k) = \mathbf{Y}_{\text{ref}}(k) - \phi \mathbf{Z}(k) - \mathbf{D} \Delta \mathbf{U}(k+5)$, which is similar to the trajectory error, that is the error between the reference trajectory and the free response. Formula (13) can be modified into the following form:

$$\begin{aligned} J_k &= \|\mathbf{G}_y(k) \Delta \mathbf{U}(k) - \mathbf{E}(k)\|_{\mathbf{Q}}^2 + \|\Delta \mathbf{U}(k)\|_{\mathbf{R}}^2 \\ &= [\Delta \mathbf{U}^T(k) \mathbf{G}_y^T - \mathbf{E}^T(k)] \mathbf{Q} [\mathbf{G}_y \Delta \mathbf{U}(k) - \mathbf{E}(k)] + \Delta \mathbf{U}^T(k) \mathbf{R} \Delta \mathbf{U}(k) \\ &= \Delta \mathbf{U}^T(k) [\mathbf{G}_y^T \mathbf{Q} \mathbf{G}_y + \mathbf{R}] \Delta \mathbf{U}(k) - 2 \mathbf{E}^T(k) \mathbf{Q} \mathbf{G}_y \Delta \mathbf{U}(k) + \mathbf{E}^T(k) \mathbf{Q} \mathbf{E}(k) \end{aligned} \quad (21)$$

Then formula (21) can be expressed in the standard form:

$$J_k = \frac{1}{2} \Delta \mathbf{U}^T(k) \mathbf{H} \Delta \mathbf{U}(k) + \mathbf{f}^T \Delta \mathbf{U}(k) + \text{const} \quad (22)$$

where

$$\mathbf{H} = 2(\mathbf{G}_y^T \mathbf{Q} \mathbf{G}_y + \mathbf{R}), \quad \mathbf{f} = -2 \mathbf{G}_y^T \mathbf{Q} \mathbf{E}(k) \quad (23)$$

The minimizing equation of formula (23) is as follows:

$$\min \frac{1}{2} \Delta \mathbf{U}^T(k) \mathbf{H} \Delta \mathbf{U}(k) + \mathbf{f}^T \Delta \mathbf{U}(k) + \text{const} \quad (24)$$

The optimal predictive input increments are obtained by zeros of the performance index function:

$$\Delta \mathbf{U}_{\text{opt}}(k) = \frac{1}{2} \mathbf{H}^{-1} \mathbf{f} \quad (25)$$

However, only the first column of the optimal solution can be used as the control increment in the rolling optimization:

$$\frac{1}{2} \begin{bmatrix} \mathbf{I}_m & \mathbf{0} & \cdots & \mathbf{0} \\ & \ddots & & \\ & & \mathbf{I}_{N_c-1} & \end{bmatrix} \Delta \mathbf{u}_{\text{opt}} \quad (26)$$

4. Simulation and Analysis

In order to test the performance of the model predictive controller, the reference trajectory of the SRC is changed. And the changing trends for calefaction heat quantity of reboiler and quantity of reflux are list out respectively, with disturbance resistance in terms of changes in sensitive plate temperature and overhead methyl acetate concentration. Meanwhile, measurement disturbances are introduced into the system.

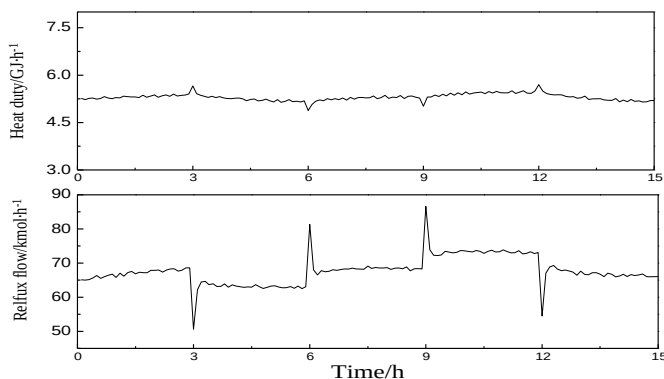


Fig. 9. Input response of sensitive plate temperature setpoint change.

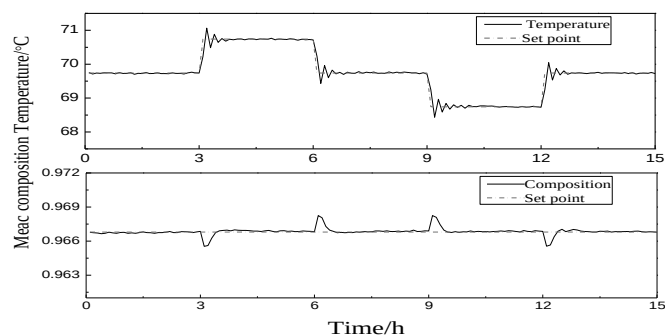


Fig. 10. Output response of sensitive plate temperature setpoint change.

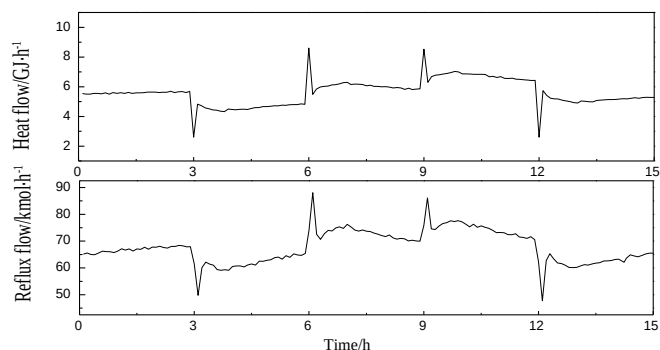


Fig. 11. Input response of purity setpoint change.

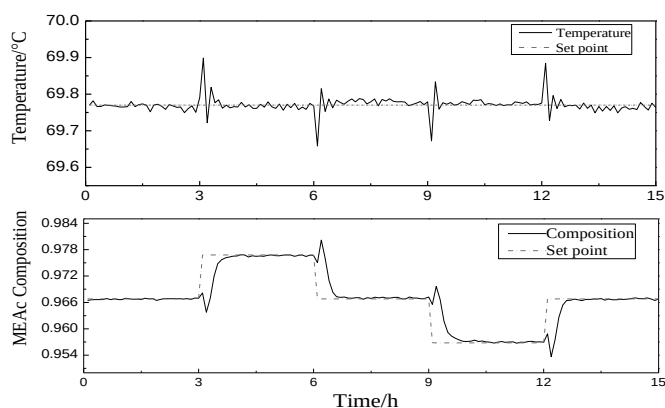


Fig. 12. Output response of purity setpoint change.

4.1. Dynamic response of the sensitive plate temperature fluctuation

Inputs and outputs of the system with $\pm 1\text{ }^{\circ}\text{C}$ disturbance of the sensitive plate's temperature introduced are shown in Figs. 9 and 10, respectively. And the set point is changed by $1\text{ }^{\circ}\text{C}$ every 3 h. Fig. 9 shows the changing trends of the manipulated variables including quantity of reflux and calefaction heat quantity of reboiler with

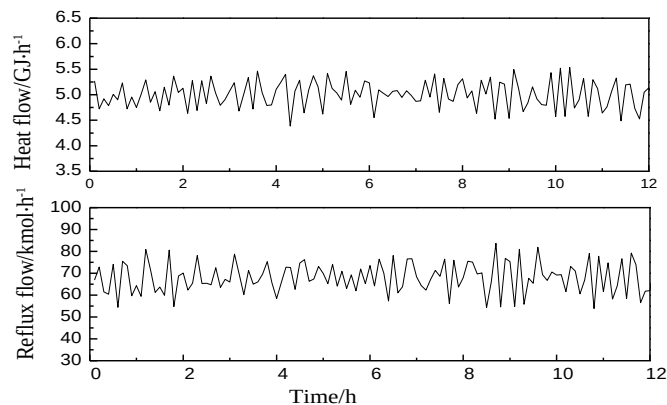


Fig. 13. Input response of feed random fluctuation.

perturbation, which changes along the trends of temperature. And it can be seen from Fig. 10 that the tracking performance control system is satisfactory. The fluctuation of sensitive plate temperature is in a small range, and the overhead methyl acetate concentration is back to the steady state quickly after a disturbance.

4.2. Dynamic response of methyl acetate concentration fluctuation

Inputs and outputs of the system with 1% disturbance of methyl acetate concentration introduced are shown in Figs. 11 and 12, respectively. And the set point is changed every 3 h. Fig. 11 shows the changing trends of the manipulated variables. And the manipulated variables are influenced by the perturbation of methyl acetate concentration significantly, with similar changing trend, that is, perturbation of methyl acetate concentration brings greater difficulty and stronger coupling. It can be seen from Fig. 11 that the perturbation caused an obvious lag of the response. Because the adjustment of concentration is a more complex process, related to the separation of the distillation column and the coupling of the reactors.

Performance index of the system with the above disturbances is shown in Table 3.

4.3. Dynamic response of feed rate random fluctuation

Inputs and outputs of the system with random perturbation introduced are shown in Figs. 13 and 14, respectively. Fig. 13 shows the changing trends of the manipulated variables. Fig. 14 shows that the fluctuation of sensitive plate's temperature is within $1\text{ }^{\circ}\text{C}$, and the fluctuation of methyl acetate concentration is within 1%.

It can be seen from the dynamic responses in different conditions that the model predictive control system of the SRC producing methyl acetate works timely and effectively, with satisfactory tracking performance. And compared to the several hours adjustment of most multi-variable control system, about only 1 h of the model predictive control system is more effective.

Table 3
Performance index of control schemes

Category	T_s/h		$e(\infty)$		IAE	
	Temperature	Composition	Temperature	Composition	Temperature	Composition
Increase T	0.7	0.5	2×10^{-4}	4×10^{-6}	1.6618	0.0058
Cut T	1	0.6	2.1×10^{-4}	4.6×10^{-6}	1.8295	0.0075
Increase C	0.9	1.2	8×10^{-5}	4×10^{-5}	0.4206	0.0447
Cut C	0.5	1.1	8×10^{-5}	3×10^{-5}	0.3168	0.0481

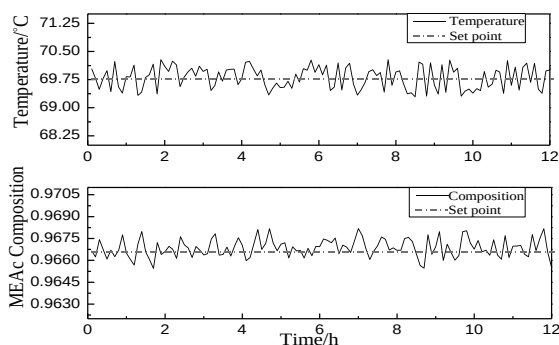


Fig. 14. Output response of feed random fluctuation.

5. Conclusions

Distillation column with side reactors (SRC) for methyl acetate production is constructed in order to overcome the disadvantages of the traditional reactive distillation. Methyl acetate production simulation experimental platform is established by Aspen Plus to obtain the optimum integrated structure and operating parameters. Meanwhile, based on the conventional control loops, multi-variable model predictive control is applied to the process to meet the control requirements. The effectiveness of control structure is demonstrated by fluctuations in methyl acetate concentration and temperature of sensitive plate. The simulation results show good control precision, robustness and dynamic follow performance of the control scheme.

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