Modeling and model predictive control of dividing wall column for separation of Benzene–Toluene–o-Xylene

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In this paper, dividing wall column (DWC) has been chosen for a BTX (Benzene–Toluene–o-Xylene) system. A MATLAB® program has been written for nonlinear unsteady-state DWC, which is used in Simulink environment for control of the system by Model Predictive Control (MPC). Compositions of the three products (benzene, toluene, and o-xylene) are indirectly controlled by controlling the corresponding temperatures of the respective tray due to requirement of online analyzer. The temperature of uppermost tray in the rectifying section, stage temperature in the main column corresponding to the side stream withdrawn, and the bottom stage temperature in the stripping section have been chosen in order to maintain the compositions of the three products. The manipulated variables are reflux rate (L0), side-stream flow rate (SSRF), and reboiler heat duty (Qb). It has been observed that MPC shows good performance even in the presence of ±10% change in the feed flow rate, feed composition, and liquid split factor in comparison with conventional controllers. The MPC has less settling time (almost 1.5 h) compared with the PI controller (approximately 3–4 h).

Keywords: BTX separation; dividing wall column; multivariable control; model predictive control

1. Introduction

Distillation is the most desirable process for the separation of multi-component liquid mixture. However, when separation of high-purity products is required, a single distillation column is not sufficient; for the separation of n-component mixture, n −1 columns are required (Sotudeh & Shahraki, 2007). An important effort has been focused on the development of a new design, optimization, and control method for thermally coupled distillation columns, which provide saving up to 30–40% of the total annual cost for the separation of some multi-component liquid mixtures as compared with classical distillation sequences (Triantafyllou & Smith, 1992). The use of fully thermally coupled distillation arrangements, such as the dividing wall column (DWC), leads to significant reductions in both energy and capital costs when compared with conventional two-column arrangements in the separation of ternary mixtures.

Although some authors have studied control of DWC to improve the process operation, yet model-based control finds its good applicability on this type of the distillation column. DWC is a nonlinear system due to its inherent complex dynamics because of the middle wall. Therefore, this system is challenging to control at its optimum operating parameters. Rewagad and Kiss (2012) showed the model predictive control (MPC) results for composition control; however, in this study, we have focused the control of temperatures. Besides disturbance in feed flow rate and feed composition, the disturbance in liquid split factor has also been considered. The parameter-based controllers, like proportional-integral-derivative controllers, are characterized by a short development time and smaller development efforts. The development of MPC for a DWC is a promising approach as the advantages of this control methodology can be intensified in view of the operational difficulties of the column. For maintaining the product purity at its desirable conditions, proper control is required. Therefore, model predictive control was selected to analyze its suitability.

The huge impact of distillation processes in both operation and investment costs has motivated the development of various types of fully thermally coupled distillation columns that can be used in saving energy and capital cost (Halvorsen & Skogestad, 2004). Conventionally, a ternary mixture can be separated via direct sequence (most volatile component is separated first), indirect sequence (heaviest component is separated first), or distributed sequence (mid-split) consisting of two to three distillation columns (Van Diggelen, Kiss, & Heemink, 2010). Eventually, this led to the concept known today as DWC that integrates in fact the two columns of a Petlyuk system into one column shell (Kaibel, 1987; Kolbe & Wenzel, 2004; Schultz et al.,

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In DWC, a single column is divided into four sections by inserting a wall in it. DWC is especially advantageous for separating ternary mixtures. The schematic diagram of the DWC is shown in Figure 1. It is divided into four sections: top section of the column is known as a rectifying section, left section as prefractionator, right section as main column, and the bottom section as stripping section. Feed consisting of 30 mol% benzene, 30 mol% toluene, and 40 mol% o-xylene is introduced on the 12th stage in the prefractionator. Almost pure toluene is withdrawn from the 11th stage of the main column. Benzene and o-xylene are obtained as top and bottom products, respectively.

Van Diggelen et al. (2010) proposed a model of DWC with the assumptions of constant pressure, no vapor flow dynamics, linearized liquid dynamics, and neglecting energy balance and changes in enthalpy. The researchers used this model to compare various control strategies. Hiller, Buck, Ehlers, and Fieg (2011) developed a non-equilibrium stage model by assuming heat and mass transfer between the liquid and vapor phases for the ideal component system. Ignat and Woinaroschy (2011) used a dynamic model for minimizing the distillation start-up time for the separation of an ideal benzene–toluene–o-xylene ternary mixture and the separation of a non-ideal methanol–ethanol–1-propanol mixture. The effect of the liquid split ratio and vapor split ratio is responsible for internal disturbance and the dynamics of the DWC. Ignat and Woinaroschy (2011) also assumed the hold-up volume in the column. The vapor split ratio is assumed to be constant and liquid split ratio as a load change in this study.

Recently, Buck, Hiller, and Fieg (2011) also reported experimentally, temperature profile of the column by implementing MPC. Their study proves the real-life practicability of MPC; they did not provide an analysis of the transient behavior of DWC under desired disturbances. Kvernland, Halvorsen, and Skogestad (2010) applied MPC on the Kaibel column that separates a feed stream into four product streams using only a single column. The main objective for optimal operation was to minimize the total impurity flow. Finally, they concluded that MPC obtained typically less total impurity flow as compared with conventional decentralized control.

In this work, a mathematical model of DWC has been developed assuming non-constant volatility. Several simulation runs of the model have been used to investigate the effects of several parameters and dynamics of the system. BTX (Benzene–Toluene–Xylene) has been chosen as a component system. A MATLAB® program was written for unsteady-state DWC for investigating the MPC methodology. The controlled variables selected are: temperature of uppermost tray in the rectifying section, stage temperature corresponding to the side stream withdrawn, and the bottom stage temperature in the stripping section in order to maintain the compositions of the three products (Ling & Luyben, 2009). The manipulated variables are reflux rate, side-stream flow rate (SSRF), and reboiler heat duty. The performance of MPC controller has been verified by giving load changes in the feed flow rate and feed composition. The performance of MPC in the presence of disturbances variables has been compared with that of PI controller.

2. Mathematical model

The following assumptions have been taken into account for model development:

1. Constant volume holdup of condenser/reflux
2. Fast energy dynamics on trays (molar enthalpy change on trays with respect to time is negligible)
3. Vapor split ratio is constant as it is fixed by column design.
4. Raoult’s law is constant as BTX is close to ideal solution.

For the rectifying section, prefractionator, main column, and stripping section:

Mass balance for component $i$ at tray $j$,

$$\frac{d(M_j x_{j,i})}{dt} = V_{j+1} y_{j+1,i} + L_{j-1} x_{j-1,i} - V_j y_{j,i} - L_j x_{j,i} + F_j z_{j,i} - S_j x_{j,i}, \tag{1}$$
where \( n_c \) denotes the number of components and \( M_j \) denotes the total liquid holdup on the \( j \)th stage.

**Summation Equations:**
\[
\sum_j x_{j,i} = 1; \quad \sum_j y_{j,i} = 1, \tag{2}
\]
where \( x_{j,i} \) and \( y_{j,i} \) represent the mole fraction of the \( i \)th component in liquid phase at the \( j \)th stage, respectively.

**Energy Balance at tray \( j \):**
\[
\frac{d(H_{L,j} M_j)}{dt} = V_{j+1} H_{V_{j+1}} + L_{j-1} H_{L_{j-1}} - V_j H_{V_j} - L_j H_{L_j} + F_j H_{F_j} - S_j H_{L_j}, \tag{3}
\]
where \( H_{L_j} \) and \( H_{V_j} \) represent the liquid and vapor enthalpy at the \( j \)th stage, respectively.

Equilibrium relationship:
\[
y_{j,i} = K_{j,i} y_{j,i} \text{ where } K_{j,i} = \frac{y_{j,i} \gamma_{j,i}^{\text{nat}}}{y_j}, \tag{4}
\]
where \( y_{j,i} \) denotes the activity coefficient of the \( i \)th component in liquid phase at the \( j \)th stage.

For the condenser,

**Material Balance:**
\[
\frac{d(M_0 x_{D,j})}{dt} = V_1 x_{1,j} + L_0 x_{D,0} - D x_{D,j}. \tag{5}
\]

**Energy Balance:**
\[
\frac{d(M_0 H_D)}{dt} = V_1 H_{V,1} - L_0 H_{L,0} - D H_D - q_e. \tag{6}
\]

Summation equation:
\[
\sum_i x_{D,i} = 1. \tag{7}
\]

For the reboiler,

**Material Balance:**
\[
\frac{d(M_R x_{W,j})}{dt} = L_{n_1} x_{n_1,i} - V_{n_1+1} y_{n_1+1,i} - w x_{n_1+1,i}, \tag{8}
\]
where \( M_R \) denotes liquid holdup in reboiler.

**Energy Balance:**
\[
\frac{d(M_{n_1+1} H_{L,n_1+1})}{dt} = L_{n_1} H_{L,n_1} - H_{V,n_1+1} V_{n_1+1} - w H_{L,n_1+1} + q_R. \tag{9}
\]

Equilibrium Relationship:
\[
y_{n_1+1,i} = K_{n_1+1,i} x_{w,i}. \tag{10}
\]

Summation Equations:
\[
\begin{align*}
\sum_i x_{n_1+1,i} & = 1 \quad \text{and} \quad \sum_i y_{n_1+1,i} = 1. \tag{11}
\end{align*}
\]

At the intersection of rectifying section (Section 1) with prefractionator (Section 2) and main column (Section 3):

**Vapor Mixing:**
\[
\begin{align*}
V_{n_1+1}^{(1)} &= V_1^{(2)} + V_1^{(3)}, \tag{12} \\
V_{n_1+1}^{(1)} y_{n_1+1,i} &= V_1^{(2)} y_{1,i} + V_1^{(3)} y_{1,i} \tag{13}.
\end{align*}
\]

**Liquid Splitting:**
\[
\begin{align*}
L_0^{(2)} &= \alpha L_1^{(1)}, \tag{14} \\
L_0^{(3)} &= (1 - \alpha) L_1^{(1)},
\end{align*}
\]
where \( \alpha \) is liquid split factor.

\[
\begin{align*}
x_{n_1,j}^{(2)} &= x_{n_1,j}^{(1)}, \quad x_{n_1,j}^{(3)} = x_{n_1,j}^{(1)},
\end{align*}
\]

At the intersection of Sections 2 and 3 with Section 4 (Stripping section):

**Vapor splitting:**
\[
\begin{align*}
V_{n_2+1}^{(2)} &= \beta V_1^{(4)}, \\
V_{n_2+1}^{(3)} &= (1 - \beta) V_1^{(4)}, \tag{15}
\end{align*}
\]
where \( \beta \) is a vapor splitting factor.

\[
\begin{align*}
y_{n_2+1,j}^{(2)} &= y_{n_2+1,j}^{(4)}, \\
y_{n_2+1,j}^{(3)} &= y_{n_2+1,j}^{(4)}.
\end{align*}
\]

**Liquid mixing:**
\[
\begin{align*}
L_0^{(4)} &= L_0^{(2)} + L_0^{(3)}, \tag{16} \\
L_0^{(4)} x_{n_2,j}^{(2)} + L_0^{(3)} x_{n_2,j}^{(3)} &= L_0^{(4)} x_{n_2,j}^{(4)}. \tag{17}
\end{align*}
\]

These model equations are set of ordinary differential equations-initial value problems, which were solved by ode15s solver (an inbuilt function in MATLAB®) to solve ordinary differential equations. For simulation, physical properties and the nominal operating conditions of the investigated system (Benzene, Toluene and o-Xylene) are given in Table 1.

### 3. Model predictive control of a DWC

The controllability indicator such as relative gain array (RGA) is a useful method to understand the behavior of the system (Segovia-Hernandez, Hernandez-Vargas, & Marquez-Munoz, 2007; Skogestad & Postlethwaite, 2005). The RGA provides information about the interactions among the controlled and manipulated variables. The RGA element has been calculated as the ratio of open-loop gain to the closed-loop gain for a pair of variables.

For a selected pair of variables, values of the RGA element close to unity are preferred as the best combination. Due to complex interactions among the manipulated and controlled variables, decentralized controller’s performance is not very good.
Table 1. Properties of the system and nominal operating parameters.

| Properties of the system | Benzene (C₆H₆) | Toluene (C₇H₈) | Xylene (C₈H₁₀) |
|--------------------------|----------------|----------------|----------------|
| Molecular formula        | C₆H₆          | C₇H₈          | C₈H₁₀          |
| Molecular weight         | 78.11         | 92.14         | 106.17         |
| Boiling point            | 80.2 °C       | 110.7 °C      | 144.5 °C       |
| Density (mol/cm³)        | 0.01133       | 0.00941       | 0.00829        |
| Critical temperature     | 289.1 °C      | 318.7 °C      | 357.2 °C       |

Antoine equation and its constants

\[ T_{sat} = \frac{B}{(A - \log P) - C}, \text{ units of } T \text{ in K and } P \text{ in atm} \]

| Components | A    | B       | C       |
|------------|------|---------|---------|
| Benzene    | 15.9008 | 2788.51 | 220.79  |
| Toluene    | 16.0137 | 3096.52 | 219.48  |
| Xylene     | 16.1156 | 3395.57 | 213.69  |

Nominal operating parameters

| Value       |
|-------------|
| Feed flow rate | 1 kmol/s   |
| Feed temperature | 85 °C      |
| Feed composition (mol%) | 30% B, 30% T, 40% X |
| Reflux rate   | 0.860 kmol/s |
| Vapor split factor | 0.627      |
| Liquid split factor | 0.353     |
| Side stream flow rate | 0.296 kmol/s |
| Reboiler heat duty | 40.544 MW  |
| Bottoms flow rate | 0.401 kmol/s |
| Reflux ratio  | 2.84        |

The DWC is a multivariable process that gives motivation for use of a MPC. The MPC offers a large number of operational advantages for diverse processes like reaction, separation, etc. (Backx, Bosgra, & Marquardt, 2000; De Temmerman, Dufour, Nicolaï, & Ramon, 2009). In view of the multi-input–multi-output system, the MPC is characterized by many manipulated and controlled variables and it may be performed for a complex system (Buck et al., 2011). It is an appropriate controller to control a multivariable process with considerable interactions. The MPC shows a good result in eliminating loop interaction when it is compared with a decentralized control scheme. Model predictive controller is good for creating a linear model at any operating point. Total operating cost and settling time can be reduced by controlling the temperature of the different sections in spite of the compositions of the products (Rodríguez Hernández & Chinea-Herranz, 2012). The manipulated variables, reflux ratio, side stream flow rate, and reboiler heat duty are used to control the temperatures (and therefore product purities). For investigating the controllability of MPCs, MPC Toolbox in MATLAB® was used. The cost function in MPC methodology is

Min \[ F = \sum_{j=1}^{N_y} \sum_{i=1}^{P} (y_{sp,i} - y_{ij})^2 w_{ij} + \sum_{j=1}^{N_y} \sum_{i=1}^{M} (\Delta u_{ij})^2 w_{u_{ij}}, \]

subject to:

\[ \Delta u_{LB} \leq \Delta u \leq \Delta u_{UB}, \]  \[ u_{LB} \leq u \leq u_{UB}, \]

where \( u_{LB} \) and \( u_{UB} \) represent the lower and upper bounds to control the oscillations of the manipulated variables.

Model predictive control works on prediction horizon, control horizon and robustness factor parameters. Prediction horizon is the number of sampling intervals over which the cost function is minimized. Control horizon is the number of sampling intervals over which the control moves are estimated by optimization routine. The best suitable MPC parameters based on several runs by selecting different values are given in Table 2. Robustness factor controls the sluggishness and the speed of the controller. The working scheme of MPC is shown in Figure 2.

These parameters have a high impact on the total controller performance. The lower and upper limits on the manipulated variables (reflux rate, side stream flow rate,

The working scheme of MPC is shown in Figure 2.

| Parameter             | Value |
|-----------------------|-------|
| Control interval (sampling interval) | 10 s   |
| Prediction horizon    | 20    |
| Control horizon       | 6     |
| Robustness factor     | 0.25  |

Table 2. Best suitable model predictive control parameters.
and reboiler heat duty) have been set as [360.5 96 20.54] and [2000 496 60.54], respectively. The prediction horizon is coupled with the system’s time constants and the chosen sampling rate (Agachi, Nagy, Cristea, & Imre-Lucaci, 2006).

In this study, temperature of tray 6 in Section 1 (Tsec1 (6)), temperature of tray 18 in Section 3 (Tsec3 (18)), and temperature of tray 6 in Section 4 (Tsec4 (6)) have been controlled by the reflux rate ($L_0$), side stream flow rate (SSFR) and reboiler heat duty ($Q_B$).

4. Effect of parameters on temperature and composition

Some important parameters of DWC such as feed composition of benzene, toluene, and xylene, feed flow rate, reflux rate, SSFR, and reboiler heat duty were selected for investigation of the effect on temperatures and the resulting product compositions.

4.1. Effect of benzene composition in the feed

The effect of benzene composition of the feed is shown in Figure 3. While changing the benzene composition, the other two components were assumed to be in the same proportion. As can be seen, on increasing benzene composition in the feed, there is no significant change in $o$-xylene composition in the bottoms product. Toluene composition in the middle product does not change much until 30 mol% of benzene composition in the feed after which there is a declining trend because extra benzene is mixed up in the middle product. The benzene composition of the top product increases up to 30 mol% of benzene in the feed, beyond which, there is no significant change. The temperatures also follow the trend accordingly. Therefore, it is evident that 30 mol% composition of benzene in feed is an optimum value giving the maximum purity of all the products.

4.2. Effect of toluene composition in the feed

The effect of toluene composition of the feed is shown in Figure 4. On increasing the feed composition of toluene, the benzene composition in the top product and $o$-xylene composition of the bottom product do not show significant effect up to 30 mol% of toluene composition in the feed but decreases afterwards. However, toluene composition in the middle product increases and becomes almost constant after 30 mol% of toluene composition in the feed. Therefore, the toluene composition of 30% seems to be an optimum feed composition as it gives the highest purity of all the products. The temperatures also follow the trend accordingly in equilibrium with the product composition.

4.3. Effect of $o$-xylene composition in the feed

The effect of $o$-xylene composition in the feed is shown in Figure 5. While changing the $o$-xylene composition, the other two components are supposed to be in the same proportion. Purity of benzene is constant up to 40 mol% $o$-xylene in the feed after which, it suddenly decreases. Moreover, $o$-xylene purity of the product increases up to 40% and then it becomes constant on further increasing xylene composition. Toluene purity decreases beyond 40
Figure 4. Effect of toluene composition in feed on temperature and product composition.

Figure 5. Effect of \( \omega \)-xylene composition in feed on temperature and product composition.

Figure 6. Effect of feed flow rate on temperature and product composition.

mol\% composition of the \( \omega \)-xylene in the feed. Therefore, the temperature in the bottom section (stripping section) increases and the temperatures in the top section (rectifying section) and middle section decrease slightly.

4.4. Effect of feed flow rate
As the feed flow rate increases up to 1000 mol/s, benzene and toluene compositions in the product increase, but the xylene composition remains maximum at almost 100\%. On increasing the feed flow rate further, the benzene composition remains constant; however, other two product compositions decrease as shown in Figure 6. Accordingly, there is a small decrease in the temperature of the rectifying section, while the temperatures of the main column and stripping section decrease significantly.

4.5. Effect of reflux rate
The reflux rate plays an important role in the product purities as shown in Figure 7. Below reflux rate of 860 mol/s, benzene and toluene compositions increase and \( \omega \)-xylene composition in the bottom product remains constant at almost 100\%. For more than reflux rate of 860 mol/s, benzene composition of the top product remains
constant, but o-xylene and toluene compositions decrease sharply. Accordingly, the temperature in the rectifying section (Tsec1) initially decreases and then becomes constant after 860 mol/s of reflux rate, while the temperature in the stripping section (Tsec4) decreases after 860 mol/s of reflux rate.

4.6. Effect of side stream flow rate

For side stream flow rate less than 296 mol/s, benzene composition of the top product does not change much for a change in the side stream flow rate; however, toluene composition in the middle product decreases after 296 mol/s as shown in Figure 8. The o-xylene composition in the bottom product increases and becomes constant after 296 mol/s of side stream flow rate. Therefore, side stream flow rate of 296 mol/s is the optimum flow rate as it gives the highest purity of all the products.

4.7. Effect of reboiler heat duty

The effect of the reboiler heat duty is shown in Figure 9. As can be seen, benzene composition of the top product does not show significant change until 41 MW, beyond which it starts decreasing. Toluene composition first increases and then decreases, and o-xylene in the bottom product increases up to 41 MW after which it remains almost constant. Therefore, 41 MW is the optimum reboiler heat duty.
5. Results for model predictive control of DWC
5.1. Load change in benzene composition of feed

±10% load changes were given in the feed flow rate to observe the performance of the MPC as shown in Figure 10. Temperature overshoots in Section 1 are approximately 0.3 °C and 0.2 °C for −10% and +10% change, respectively. Similarly, there is only very small overshoot (only 0.5%) in the product composition. Moreover, the offset is negligibly very small in both cases.

To compare the performance of MPC with conventional controller, Ling and Luyben’s (2010) results have been considered for the study. The PI controller stabilized the controlled variables in about 3–4 h; however, time taken by MPC controller is one-fourth time of that by PI controller. Temperature overshoot in Section 1 is 0.3 °C and 0.4 °C for −10% and +10% change, respectively, in case of PI controller; in case of MPC, this is marginally lower, that is, 0.3 °C and 0.2 °C, respectively. Similarly, in Section 3, temperature overshoot is 0.6 in PI control; however, in case of MPC, there is very small overshoot, that is, 0.13 °C and 0.12 °C for −10% and +10% load change, respectively. In Section 4 also, the overshoot is marginally better for MPC performance.

5.2. Load change in toluene composition of feed

The MPC performance for ±10% load change in toluene feed composition is also studied as shown in Figure 11. The maximum overshoot was 0.1 °C in temperature. The controller brought back the temperature to their set points within approximately half an hour. Accordingly, the composition was also controlled with negligible offset.

PI control achieved stabilization in 4–5 h as reported by Ling and Luyben (2010); however, MPC stabilized in 0.6 h upon ±10% disturbance in toluene feed composition. Temperature overshoot in Section 1 is 0.5 °C in PI and 0.1 °C in MPC Controller. Temperature overshoot in Section 3 is 0.7 in case of PI controller and 0.1 in MPC. Similarly, in Section 4, the overshoot is marginally better.

5.3. Load change in o-xylene composition of feed

The MPC performance for load change of ±10% in o-xylene composition of feed is shown in Figure 12. The maximum overshoot is approximately 0.2 °C in temperature with almost no offset. The product compositions were also controlled according to their initial values. PI control achieved stabilization in 4–5 h as reported by Ling and Luyben (2010); however, MPC stabilized in 1.5 h upon ±10% disturbance in the o-xylene feed composition. Temperature overshoot in Sections 1 and 3 are a little bit better in the case of MPC; however, in Section 4, MPC shows more overshoot than PI control (a difference of 0.5 °C).
Figure 11. MPC performance for ± 10% disturbance in toluene feed composition.

Figure 12. MPC performance for ± 10% disturbance in o-xylene feed composition.
Figure 13. MPC performance for ±10% disturbance in feed flow rate.

Figure 14. MPC performance for ±10% disturbance in liquid split factor.
5.4. Load change in feed flow rate
The MPC response for ±10% load change in feed flow rate is shown in Figure 13. The temperature of all the three sections was controlled at their respective set points after maximum overshoot of 0.4 °C. The compositions were also maintained at their initial values with maximum deviation of 0.5%. PI controller stabilized in about 4–5 h (Ling and Luyben, 2010); however, MPC achieved the control performance in 1.5 h.

5.5. Load change in liquid split factor
The response for ±10% load change in liquid split factor is shown in Figure 14. The temperatures in Sections 1, 3, and 4 are controlled at their respective set points with maximum overshoot of 0.1, 0.4, and 0.4, respectively. Toluene composition shows a maximum overshoot of 0.02; other two compositions have a smaller overshoot in comparison with the toluene.

6. Conclusions
A mathematical model was developed for DWC, which was used as a case study for studying the control behavior of model predictive control. The controlled variables were selected as the temperatures of the 6th tray in the rectifying section, the 18th tray in the main column, and the 6th tray in the stripping section to ensure the maximum purity of benzene, toluene, and xylene, respectively. The manipulated variables selected are reflux rate, side stream flow rate, and reboiler heat duty. Model predictive controller was able to control the temperatures (and therefore product purities) of all the three sections in the presence of ±10% load changes in feed composition, flow rate, and liquid split factor. The performance of MPC was also compared with conventional controller and it is concluded that being a multivariable controller, MPC performs better than PI controller.

Nomenclature

- $M_j$: Total liquid holdup on the $j$th stage, moles
- $x_{j,i}$: Mole fraction of the $i$th component in the liquid phase at the $j$th stage
- $V_j$: Vapor flow rate from the $j$th stage, mole/s
- $L_j$: Liquid flow rate from the $j$th stage, mole/s
- $y_{j,i}$: Mole fraction of the $i$th component in the vapor phase at the $j$th stage
- $F_j$: Feed flow rate on the $j$th stage, mole/s
- $Z_{j,i}$: Mole fraction of the $i$th component in feed at the $j$th stage
- $S_j$: Side stream flow rate at the $j$th stage, mole/s
- $n_k$: Number of stages in the $k$th section ($k = 1, 2, 3, 4$)
- $n_c$: Number of components = 3
- $H_{V,j}$: Vapor enthalpy on the $j$th stage, J/mole
- $P_{sat,j}$: Saturation pressure of the $j$th stage component at the $j$th stage, Pa
- $P_j$: Pressure on the $j$th stage, Pa
- $M_0$: Liquid holdup in condenser, moles
- $x_{D,i}$: Mole fraction of the $i$th component in distillate
- $L_0$: reflux rate, mole/s
- $D$: Distillate rate, mole/s
- $H_{D,i}$: Liquid enthalpy of distillate, joule/mol
- $q_C$: Condenser duty, J/s
- $M$: Control horizon
- $M_R$: Liquid holdup in reboiler, mole
- $N_u$: Number of manipulated variables
- $N_y$: Number of controlled variables
- $p$: Prediction horizon
- $x_{u,i}$: Mole fraction of the $i$th component in bottom product
- $w$: Bottom product flow rate, mole/s
- $w_{u,i}$: Weight on control move
- $w_{v,i}$: Weight on controlled variable deviation
- $y_{sp}$: Set point of controlled variables
- $\Delta u$: control move of manipulated variables
- $y_{ref}$: Reference value of controlled variables

Disclosure statement
No potential conflict of interest was reported by the authors.

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