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Model Predictive Control Design of Pressure Swing Distillation for Methanol-C5 Separation

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Abstract

The azeotropic mixture that is greatly influenced by pressure is usually separated by pressure-swing distillation (PSD). This paper investigates the control selection of the PSD system. The PSD scheme was developed by Luyben. Model predictive control (MPC) is used as an alternative control method because it had a better response compared to the proportional-integral (PI) controller. MPC utilizes a dynamic model to make a control action toward the process. Generating the model is a crucial step for the MPC to work accurately. In this work, the model representing the process was approached by the first order plus dead time (FOPDT) model. The steady-state and dynamic simulations were carried out with HYSYS V11. The fine-tuning method was used to find the optimum tuning parameter of MPC meanwhile, the PI controller was tuned by auto-tuner. The results show that MPC was successfully implemented to the process and the control performance has increased by 21-52% in the set-point tracking test and 5-42% in disturbance rejection test compared to the PI controller.

Keywords*: methanol recovery, model predictive control, PI controller, pressure swing distillation.*

1. Introduction

Pressure swing distillation is a common unit that is used for separating azeotropic mixture [1]. The binary homogeneous azeotrope can be separated using pressure-swing distillation (PSD) [2]. PSD is used when the azeotropic separation is affected by pressure [3]. Tert amyl methyl ether (TAME) production is one of the chemical processes that can use both PSD or extractive distillation for separating the excess reactants. The comparison between them was studied in many aspects by Luyben [4]. TAME is produced by reacting isoamylenes (2M1B, 2M2B) and methanol in a reactive distillation. In that column, TAME is a heavy component therefore it comes as the bottom product. The overhead product D1 consisting of methanol and C5 mixture will be then separated by PSD [5-6]. To achieve the desired specification product from the separation process, a controller is needed to control the operating condition of the units.

PSD design used in this case is developed by Luyben [4] and will be referred to as a case study in this paper. The results of steady-state are close enough to the case study. The process is shown in figure 1. The control system in this unit was PI controller. In this paper, model predictive control (MPC), which is one of the advanced controls, was used to improve the control performance when handling any kind of disturbances.

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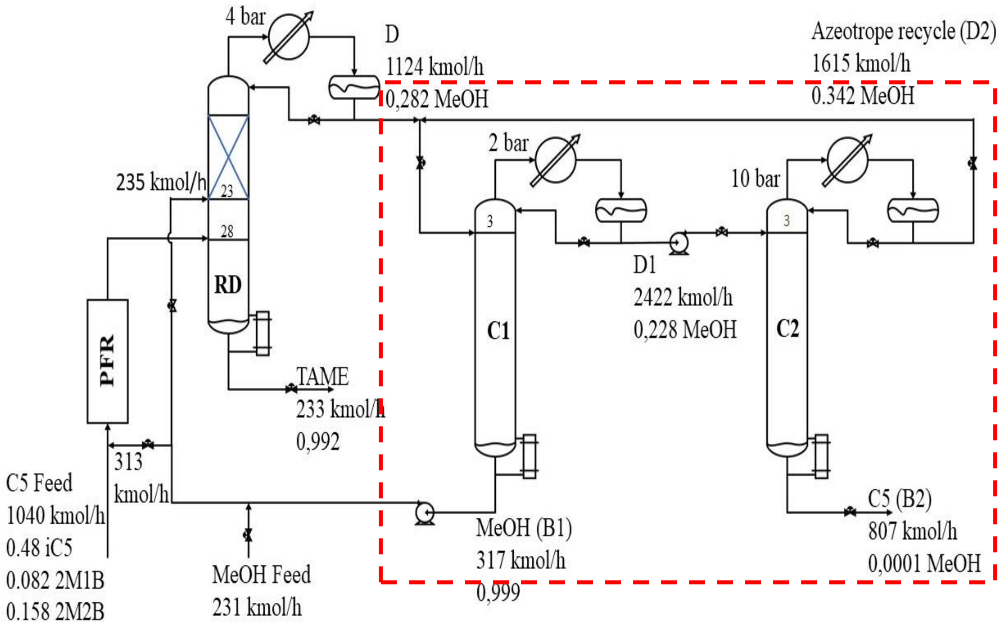

Figure 1. Pressure Swing Distillation Design in TAME Production Scheme

MPC is a control strategy that uses a dynamic model to predict the dynamic behavior of the plant. The controller will predict the process variable to minimize the future error between the predicted process variable and a desired set-point along a prediction horizon [7]. Since MPC has the ability to predict the output of the system, the controller is expected to have better performance than the PI controller [8-9].

The goal of this paper is to implement MPC in the PSD process in order to improve the control performance. In this study, the performance of MPC and PI will be compared based on the integral square of error (ISE) of MPC and PI against set-point change and disturbance, respectively.

2. Methodology

2.1. Simulation Environment

The process was simulated in the HYSYS V11 simulator. The UNIFAC model is selected for the vapor-liquid equilibrium (VLE) activity coefficient calculation. For exporting the steady-state simulation into dynamics, all equipment was sized based on the reference paper. The tray controlled is chosen by finding the location where the temperature profile is steep [4]. The tray controlled in column C1 is different from the reference paper. In that paper, tray 9 was selected to maintain the temperature profile inside the column, but in this paper, tray 3 is used to control the column C1. The only reason we can give for this difference is the use of different simulator and version of the software. The PI controllers are applied in every controller, except all liquid levels are controlled by P controllers with gains of 2. The PI controllers were tuned by auto-tuner that is available in HYSYS. The control structure of the PSD can be seen in figure 2.

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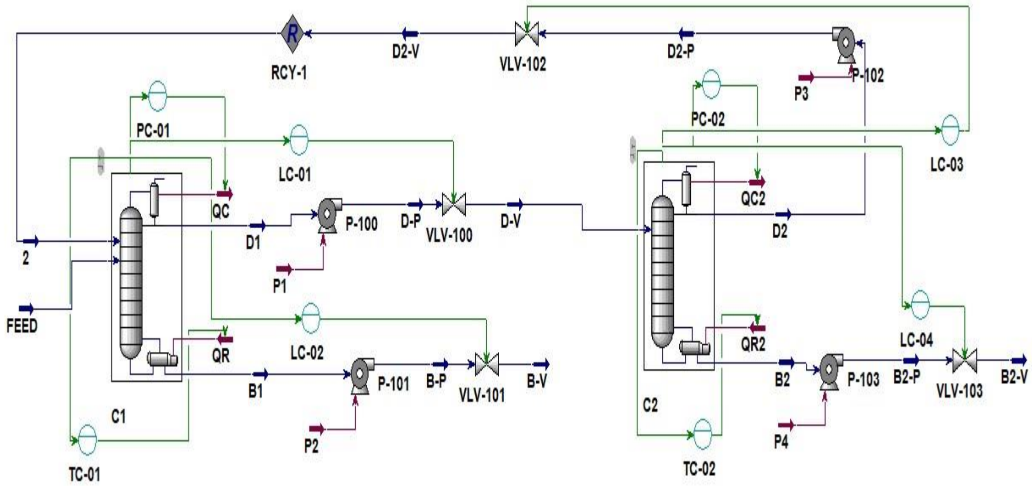

Figure 2. Structure of PSD Simulation

2.2. Model Predictive Control

In this study, the first order plus dead time (FOPDT) model was used as a dynamic model of the PSD process. FOPDT model can be generated by doing model testing in each controller to get a process reaction curve (PRC). The empirical model of FOPDT is shown by equation (1)

$$G_p(s) = \frac{K_p\, e^{-\theta s}}{\tau s + 1} \tag{1}$$

Where G_p (s) is process transfer function, $K_p = \Delta/_{\delta}$ (Δ = the magnitude of the steady-state change in the output and δ = the magnitude of the input change) , $\tau = 1{,}5\ (t_{63\%} - t_{28\%})$ with $t_{63\%}$ and $t_{28\%}$ respectively are the times when the controlled variable reaches 63% and 28% of the final steady-state , and $\theta = t_{63\%} - \tau$ [10].

As stated earlier, MPC can use a dynamic model to control the process by predicting future output over the prediction horizon. The FOPDT model generated from PRC can be directly used in HYSYS as a dynamic model. After that, MPC tuning parameters (P, M, T) were tuned by using the fine-tuning method. Table 1 summarizes the controller parameters. Furthermore, the performance of the controller was tested by doing set-point (SP) tracking and disturbance rejection. ISE values were used to see which controller giving more optimum performance. The smaller ISE values the better the performance of the controller [11].

Table 1. Tuning Parameters of MPC

Controlled variable	P	M	T
C1-Top stage pressure	370	1	1
C1-Tray 3 temperature	65	5	1
C1-Condensor liquid level	350	1	1
C1-Reboiler liquid level	350	2	1
C2-Top stage pressure	350	1	1
C2-Tray 7 temperature	345	1	1

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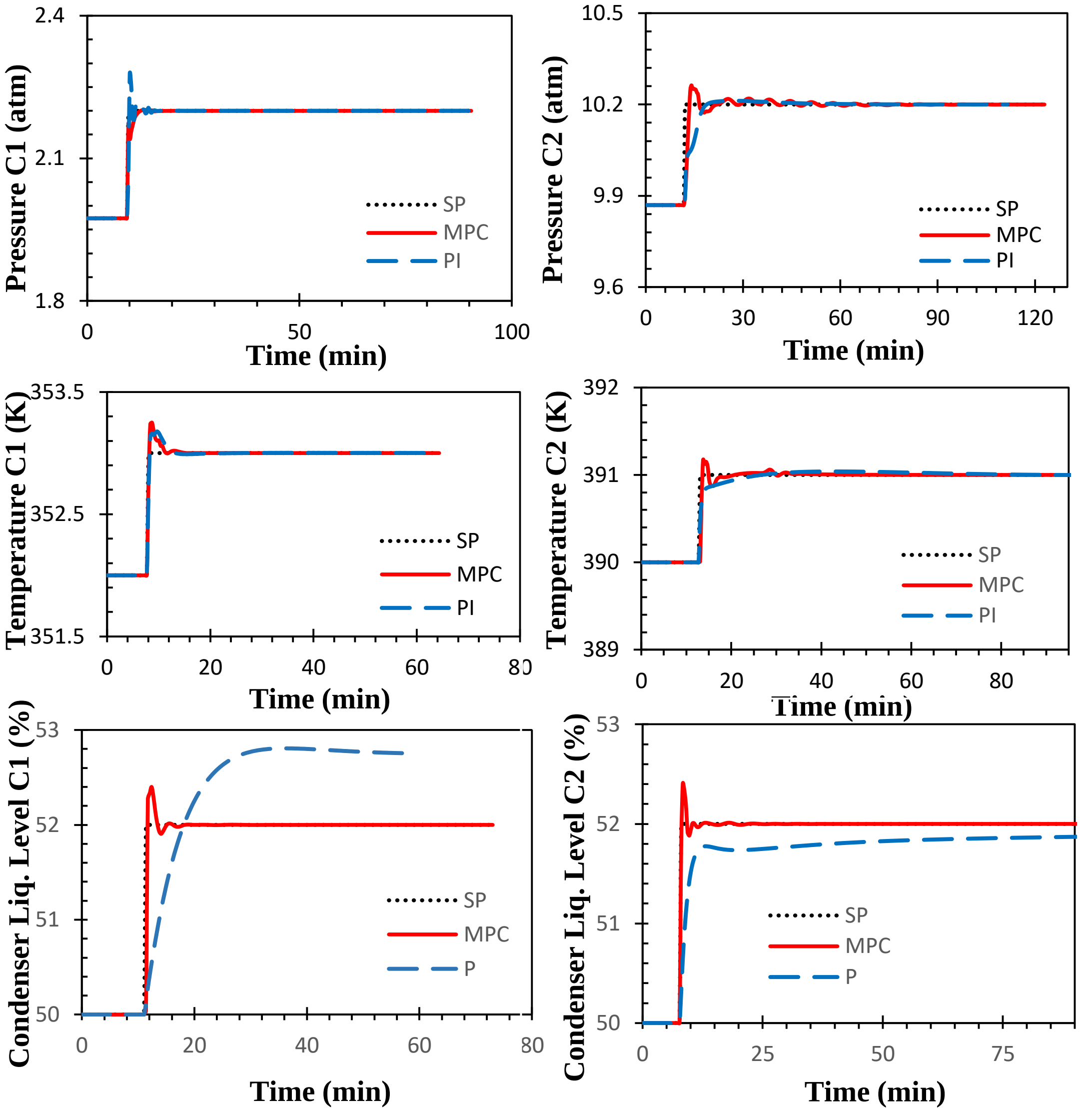
C2-Condensor liquid level	350	2	1
C2-Reboiler liquid level	350	2	4

3. Results and Discussion

3.1. Set-point tracking

A set-point tracking is performed to see the dynamic responses when the controllers track a new set-point. The results are represented in figure 3.



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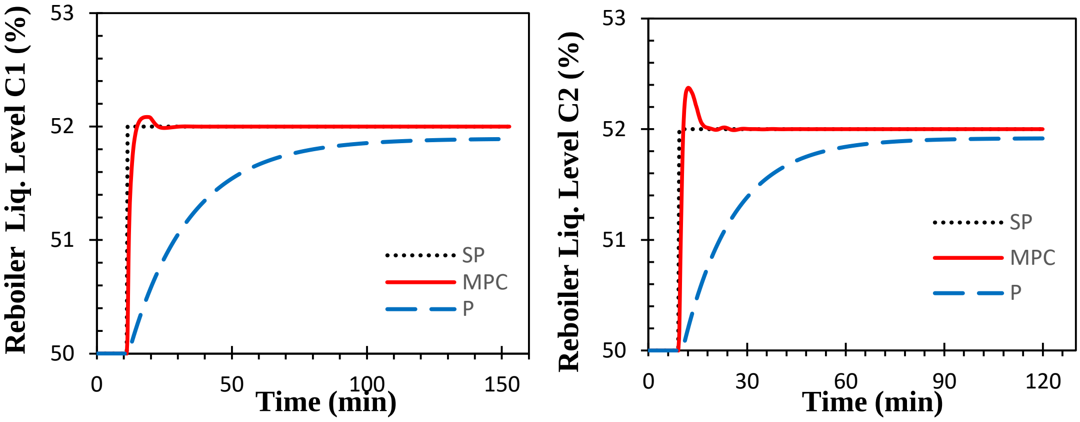


Figure 3. Controller Responses due to Set-Point Change

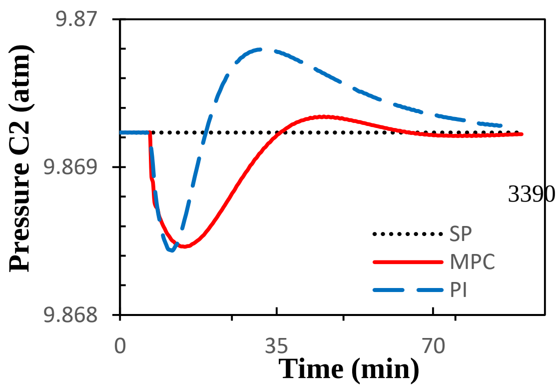
A Set-point tracking was done by making a step change in set-point value. For pressure and temperature controllers, both MPC and PI controller can reach the desired set-point. However, the rise time of MPC responses is slightly faster than PI controller responses although MPC generally makes overshoot in any controller responses. In figure 3, all level controllers in P mode cannot reach the desired value because of the characteristic of the P controller which cannot produce zero offset. To find the error value of each controller, the ISE values are calculated and the results are shown in Table 2. MPC gives an improvement in each controller by 21-52% compared to PI controllers and 93-99% compared to P controllers.

Table 2. ISE values due to set-point change

Controlled variable	ISE		Improvement (%)
	PI / P	MPC	
C1-Top stage pressure	7.49	4.76	36.4
C1-Tray 3 temperature	41.79	29.30	30
C1-Condensor liq. level (P)	2055	77.48	96.2
C1-Reboiler liq. level (P)	3536	143.82	96
C2-Top stage pressure	60.32	47.38	21.4
C2-Tray 7 temperature	158.83	76.37	52
C2-Condensor liq. level (P)	3444	35.86	99
C2-Reboiler liq. level (P)	2450	168.88	93

3.2. Disturbance Rejection

Figure 4 gives the responses to 10% increases in the feed flow rate of column C1. Increased feed flow rate will affect the process variable values to move away from the set-point. The controllers will take actions to make the process variable values back to the desired value.



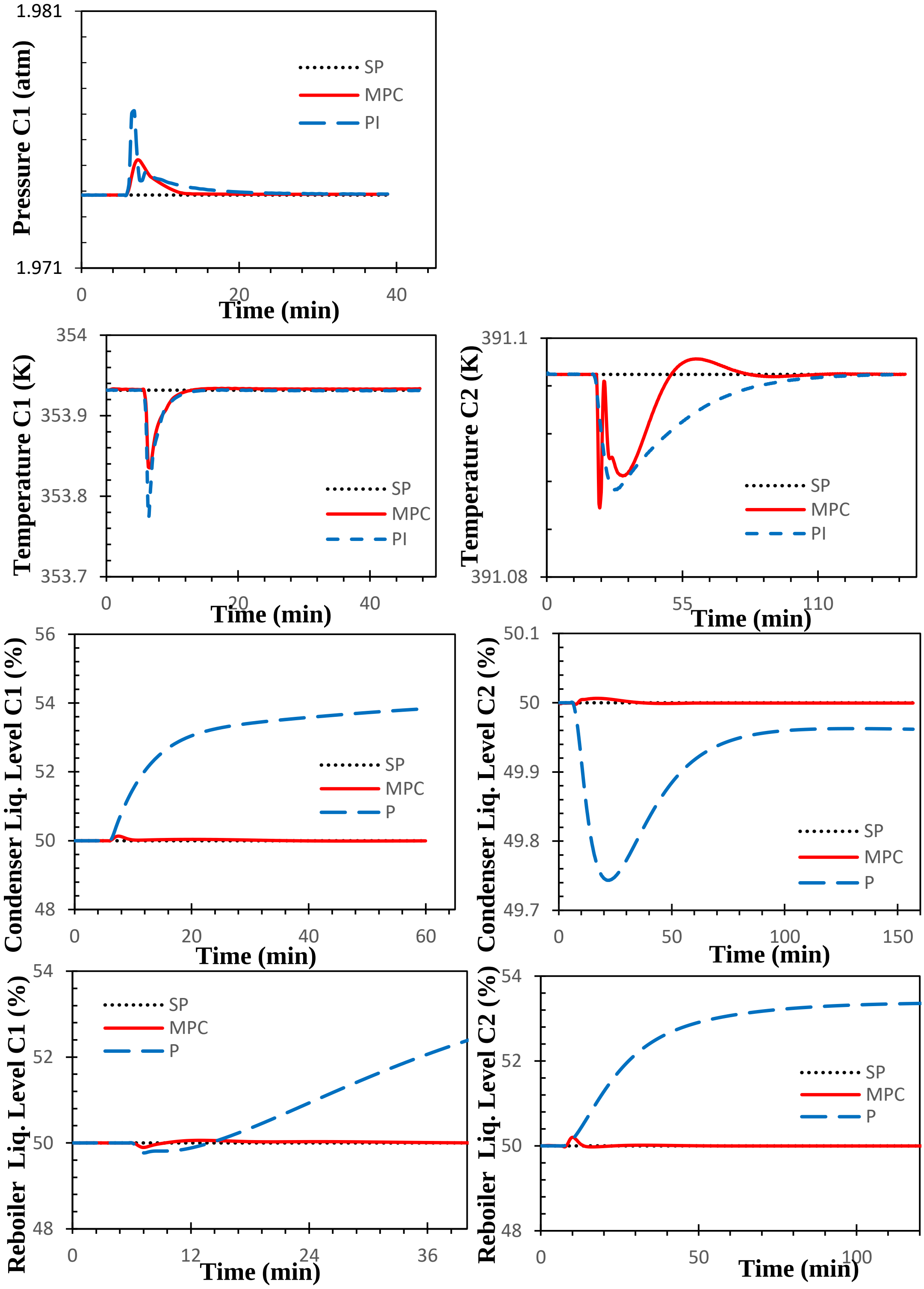


Figure 4. Controller Responses due to 10% Step Changes in Feed Flow Rate

The results shown in figure 4 indicate that MPC provides better responses than PI to overcome the disturbance because it is more stable to bring the responses back to the set-point value. The ISE values and improvements for each controller are calculated and shown in table 3.

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Table 3. ISE Values due to 10% Step Changes in Feed Flow Rate

Controlled variable	ISE		Improvement (%)
	PI / P	MPC	
C1-Top stage pressure	0.5	0.3	42.3
C1-Tray 3 temperature	16.4	15.5	5.2
C1-Condensor liq. level (P)	10215	67.5	99.3
C1-Reboiler liq. level (P)	3135	61.1	98.05
C2-Top stage pressure	1.5	0.9	37.4
C2-Tray 7 temperature	19.1	11.2	41.5
C2-Condensor liq. level (P)	792.6	10.3	98.7
C2-Reboiler liq. level (P)	2026	71	96.5

4. Conclusion

In this study, an alternative control type has been studied. The comparison between MPC and PI clearly shows that the MPC has better responses than the PI controller. MPC provided satisfactory control when dealing with set-point change and disturbance tests. MPC produced around 21-52% improvement in set-point change tests compared to PI controllers. A disturbance of 10% increases in the feed flow rate of column C1 was done to see the controller responses to reduce the deviation between the process variable and set-point value. Around 5-42% improvement was generated by MPC compared to the PI controller.

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