Binary Distillation Column Control Techniques: a Comparative Study

By

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14692

Dissertation

Submitted in partial fulfilment of the requirements

for the Bachelor of Engineering (Hons) Program

Chemical Engineering

FYP II

January 2015

Universiti Teknologi PETRONAS

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CERTIFICATION OF APPROVAL

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(CHEMICAL ENGINEERING)

Approved by,

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April 2015

CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

Mohammed Abobakr Basaar

Abstract

The purpose of this study is to propose the best control strategy for the binary distillation column. Woods & Berry model is used to represent the distillation column process. The control process is simulated on Matlab Simulink. Traditional controller settings including P, PI and PID are put to comparison. PI is found to result in a control superior to P and PID. PI is then tuned using different tuning method including Ziegler Nichols, Cohen Coon, ITAE, IMC and Symmetric Optimum. The study finds that IMC tuning parameters relatively improves the PI controller response and robustness. It is suggested to compare IMC-tuned PI controller with an advance Model Predicative Controller to ultimately conclude a superior control technique for the binary distillation column.

Acknowledgement

First and foremost I thank God for His continual blessings in my daily life, for the health in body and mind. I feel blessed and pleased with the completion of my final year project. Now that I have come to the end of my project, I feel grateful to some people who have stood beside me and provided guidance, attention, advise, support and concern.

First of all, I would like to express my gratitude to my supervisor Mr. Nasser Ramli for giving me the chance to learn from his experience in the field of process control guiding me throughout the project period.

My gratitude and appreciation extends to the Final Year Project coordinator Dr Asna Binti Mohd Zain for her constant attention and follow ups regarding the research project besides arranging for seminars to help polishing the students' research skills

Apart from those directly related to my Final Year Project, I extend my thanks to all of the Chemical Engineering Department's professors and lecturers for their valuable information throughout my degree program.

Last but not least, my thanks and appreciation to my family for the support and love which have always been the elixir to inspire me for success. My friends who have always been a gateway for me to escape from the stressful life and homesickness.

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Chapter 1. Introduction

1.1 Background

In this chapter, an introductory will define and briefly explain the distillation process and control, and the topic control strategies.

1.1.1 Distillation process

Among the technologies available for separation, distillation continues to the most commonly applied technology due to the simplicity and applicability of its principle of operation besides the high viability and low cost compared to other alternative separation process (Olujiæ, Jansen, Rietfort, Zich, & Frey, 2003). 95% of industrial separation systems implies distillation according to (Humphrey & Koort, 1991).

Distillation processes industrially take place in distillation columns where components of a mixture are separated based on the difference in volatilities. Distillation columns are said to be the least costly equipment for liquid separation as long as the ratio of volatilities of the feed composing components is at least 1.1 (Douglas, 1988). These columns can be classified according to the process operation, feed mixture nature, internal configuration as well as some other criteria.

- Batch or continuous process,
- Binary or multi-component feed mixture,
- And tray or packed column.

The distillation column operates at a specific temperature and pressure and separates the two components of the mixture (Feed) such that the concentration of the light key is increased in the top product (Distillate) and decreased in the bottom product (Bottoms) whereas the opposite for the heavy key. A simple common example is a continuous binary distillation column separating a mixture of Methanol and Water. Methanol in this example is termed the “light key” because of its higher volatility as it boils at 64.7 °C compared to Water “heavy key” which boils at 100 °C in atmospheric pressure.

The following notations are commonly used, and will be used throughout this work, to describe the streams and compositions around a distillation column:

F: The molar flow rate of the feed stream;

D: The molar flow rate of the distillate (top product);

L: The molar flow rate of the reflux;

B: The molar flow rate of the bottoms (bottom product);

V: The molar flow rate of the boil-up;

Z_L: The mole fraction of the light key in the feed stream;

Z_H: The mole fraction of the heavy key in the feed stream;

Y_L: The mole fraction of the light key in the top vapor stream;

Y_H: The mole fraction of the heavy key in the top vapor stream;

X_L: The mole fraction of the light key in the bottom liquid stream;

X_H: The mole fraction of the heavy key in the bottom liquid stream.

A typical binary distillation column is illustrated in Figure 1.The column is utilized with a total condenser which liquefy the overhead vapor stream into a receiving drum. The condensed stream is then partially drawn as distillate (D) while part of the liquid is sent back to the distillation column as reflux (L) for control and purity enhancement purposes.

Similarly, a reboiler vaporizes part of the liquid bottom steam to provide the boilup (V) flowing up through the distillation column and the rest of the liquid is drawn as bottoms product (B).

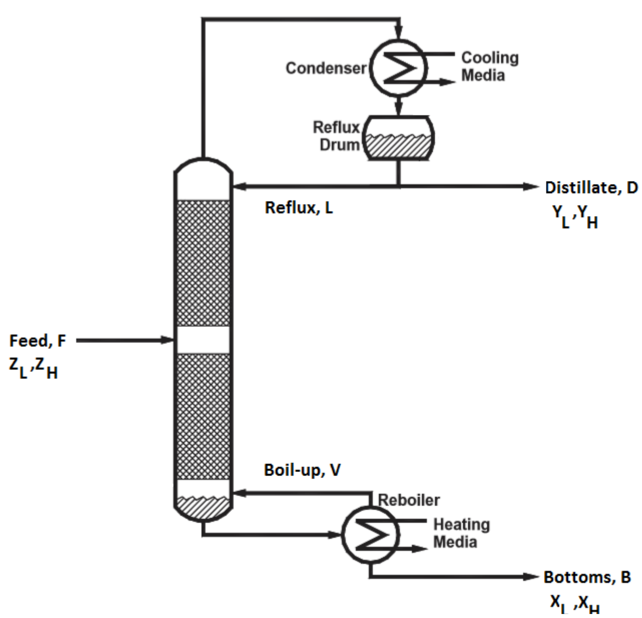

Figure 1: Basic Diagram of Distillation

1.1.2 Distillation Control

A distillation process aims to produce products of an acceptable purity with regard to the plant requirement. Thus, control strategy must be well designed and tailored for any particular column. In contrast to the high viability of this technology, its control is quite a complex task mainly because of the inherent nonlinear behavior of distillation being a MIMO, multiple-input-multiple-output, process. Interaction between controlled variable which requires presence of decouplers especially in the case of dual composition control. Moreover, severity of disturbances adds up to the complexity of distillation columns control problems.

In practice, essential variables for the operation, such as pressure and level, are entertained prior to quality variables which are product compositions and flow rates. (M. Willis, 2000). Nevertheless, product quality carries high economic importance. In (Smith, 2012), it was suggested that for dual-composition control, one of the products shall be controlled by manipulating its respective energy term while the other product shall be controlled by its draw flow rate. In other words, either the distillate or the bottoms composition is controlled by manipulating the reflux or the boilup rate respectively. Whereas the other composition is controlled by manipulating its draw flow rate. Hence, the degree of interaction in the control problem shall be reduced.

Control configuration can be refer to as "configuration [L V]" indicating that reflux and boilup flows are the manipulated variable. Configuration [D V] or [L B] means that distillate and boilup or reflux and bottom product flow rates are the controlled variables.

1.1.3 Control Strategies.

Complexity of industrial processes and the demand of enhanced safety of operation and optimal quality of product have increased the significance of development in process control (Seborg, Edgar, & Mellichamp, 2004). Various concepts define different control strategies that have been evolving since the past century. Process control strategies ca be categorized widely into conventional and advance process control

A controller receives an input signal of measured variable from a sensor and calculate the error, which is the difference between a set point and measured controlled variable, and then correlates it to an output signal sent to the final-control element which adjust the manipulated variable. Different types of controllers utilize different mathematical correlation of input to output.

a) Conventional PID Controller

They are the most commonly used controllers in the industry with a dominance of 90%. These controllers correlate the error to the corrective action signals in a proportional, integral or/and derivative terms.

Proportional term: $p(t) = \bar{p} + K_c\, e(t)$ (1)

Integral term: $p(t) = \bar{p} + \frac{1}{\tau_I} \int e(t)$ (2)

Derivative term: $p(t) = \bar{p} + \frac{1}{\tau_D} \frac{d}{dt} e(t)$ (3)

Where :

$p(t)$: $controller\ output$

$\bar{p}$: $Bias\ (steady - state) value$

K_c : $gain$

$e(t)$: $Error\ signal$

τ_I : $Integral\ time - constant$

τ_D : $Differential\ time - constant$

In practice, proportional, integral and derivative control are combined together for optimal control actions. Integral is added to the proportional control in PI controller in order to eliminate the offset. However, the integral term introduce oscillatory behavior in the response and hence, derivative term is commonly introduced in the controller along with the proportional and integral to form PID.

b) Advanced Model-Predictive Control (MPC).

Advanced process control (APC) came to emergence in the late 1970’s to compete with conventional controller and overcome its weaknesses especially in nonlinear behavior process and when process variables are tightly coupled (M. J. Willis & Tham, 1994).

MPC is the most commonly used class of advanced process control in the industry (Al-Shammari, Faqir, & Binous, 2014). It utilizes algorithms to predict the future behavior of a process based on a process model obtained from sufficient data coming from the real process that are usually identified at the commissioning stage (Badwe, Gudi, Patwardhan, Shah, & Patwardhan, 2009; Qin & Badgwell, 2003). It then solves the control problem optimally according to the predicted future response with a finite horizon at each sampling instant.

1.2 Problem Statement

The complexity of dual products binary distillation column control arises due to four reasons listed by (Hurowitz, Anderson, Duvall, & Riggs, 2003); inherent nonlinearity, non-stationary behavior, coupling interaction of controlled variables, and severity of disturbance.

Conventional PID controllers are designed to control SISO processes with high efficiency. For multivariable control (MIMO), multi-loop PID controllers are used widely with utilization of decouplers to minimize the interaction. The performance of such technique is doubtful in high purity column. The main drawback is the relatively late response of the corrective action especially that controlled variables (compositions) alters vigorously with the main disturbances (feed flow and composition). As a result, quality of product is affected and consequently, economical loss to plant is likely to occur. Conventional PID controllers on its own have variety of tuning methods.

However, MPC offers an alternative solution for the multivariable control problems using a single loop. Little studies in the open literature have compared the two control strategies on the distillation column. In this work, optimal performance of each control strategy are to be developed, investigated and compared to support or defy other existing studies.

1.3 Objective

The objective of this work is to investigate different control strategies on a binary distillation column. Conventional PID and advanced MPC control approaches will be optimized, analyzed and then compared. Performance indices for comparison include the overshoot, stability, speed of response (settling time) and steady-state error. Finally, this work is aimed to determine which control strategy is superior to maintain highly accurate purity products of top and bottom streams.

In points, the objective of the study is broken down into:

1. Determining the better traditional control setting; P, PI or PID.
2. Selecting the optimal tuning method.
3. Comparing the performance of conventional controller to advanced predictive model controller

1.4 Scope of Study

This study focuses on the performance of different control strategies named PID and MPC. An experiment based on binary distillation column will be simulated. The purpose of the control loop is to maintain the overhead and bottom product composition against disturbances. Step change in the required product purity will be introduced to investigate the control response.

Mathematical representation of the distillation column process is adapted from literature, Wood and Berry model, (Wood & Berry, 1973). Control loop is designed and controllers are tuned to optimize performance. References for the design and tuning procedures for PID and MPC are explained in (Bemporad, Morari, Ricker, & MathWorks, 2004; O'Dwyer, 2009; Wang, 2009).

Matlab Simulink® is utilized to simulate the process and test the controllers.

Chapter 2. Literature Review

Performance of conventional feedback PID control strategies are doubtful in high purity distillation column. The main drawback is the late response of the corrective action especially that controlled variables (compositions) alters vigorously with the main disturbances (feed flow and composition) according to (Sigurd Skogestad, 1997). As a result, quality of product is affected and consequently, economical loss to plant is likely. It was also reported by (Shinskey, 1984; S. Skogestad, 2007) that the top and bottom product compositions tight interaction which make the process sensitive to small changes are a big concern in the industry.

Several studies comparing PID and MPC controllers in wide range of process were released. An enriching work by (Haitao Huang & James B Riggs, 2002) compared PI with MPC implemented in a gas recovery unit of consisted of three distillation columns with constraints on the production rate. The decentralized PI controller compromised the composition set point when constraint control took action while MPC succeeded to maintain both. Consequently, it was concluded the superiority of MPC in the simultaneous adjustment of multiple manipulated variables.

In a separate work, (Haitao Huang & James B. Riggs, 2002) examined including the column level control in the MPC in comparison to leaving it to regulatory PI controller. Results showed that no coupling integration between level and compositions control when [L V] control configuration was used. However for other control configurations, significant performance improvement was noticed when level control was included in the MPC for fast responding distillation columns.

MPC adapts with dynamic changes and captures dynamic properties based on the process model while PID need its parameters to be adjusted eventually. (Li, 2010) have carried out a similar comparative study but implemented on series water tank level control. MPC showed strong robustness towards multiple changes in the system dynamic and varying time delay with an acceptable steady-state error compared to PID which exhibited overshoot and a high steady state error. However, Li's experiment favored the response of the PID controller when a change in the reference level was made as it had no steady-state error.

Additionally, (Šehić & Šehić-Galić, 2012) tested control of FOPDT & SOPDT processes with PID controller versus an MPC controller based on different orthonormal functions; Laguerre and Kautz. Their work, which was aimed to achieve response without overshoot, supported the distinction of MPC over PID in both cases with and without errors in the model of the process. However, with an error in the time constant, the response was slower than in MPC than PID.

Likewise, in the paper by (Alpbaz, Karacan, Cabbar, & Hapoğlu, 2002), MPC performed better than conventional PID. Simulation of dynamic models of binary distillation columns by (Heathcock, 1988) and (Luyben, 1989) was also accompanied by experimental validation on a pilot-scale methanol-water packed column to evaluate MPC versus PID. Dynamic Matrix Control (DMC) algorithm was used in MPC strategy to maintain the temperature of distillate which relates to composition of product.

Chapter 3. Methodology

This study will adopt the flow of methodology shown in Figure 2 to achieve objectives:

1. *Simulate the adapted process model on Simulink.*
2. *Implement and compare traditional controller settings; P, PI, PID.*
3. *Select the most appropriate controller setting.*
4. *Tune the selected controller's parameters using different methods*
5. *Compare and determine the optimal tuning parameters.*
6. *Compare traditional controller to advanced MPC controller.*
7. *Suggest the superior control techniques for this process.*

Figure 2: Methodology

The steps above will be carried out for PID controller at first and different P, PI and PID will be compared so that the best performing PID combinations will be then compared with MPC.

3.1 Gantt Chart

The Gantt chart for activities scheduling throughout this project is shown in Figure 3 and Figure 4. The first shows the flow of activities for FYP I (September 2014) while the second is for FYP II (January 2015).

No.	Activity/Week	1	2	3	4	5	6	7	8	9	10	11	12	13	14
1	Briefing on Final Year Research Project Background	■													
2	Project Titles Assignment		●												
3	Preliminary Research Work/Literature survey		■	■	■	■									
4	FYP seminar "Research Methodology"			■											
5	Preparation of Extended Proposal				■	■	■								
6	Submission of Extended Proposal							●							
7	Proposal Defence								●						
8	Secondary Research Work								■	■	■	■	■		
9	Developing, tuning and testing of PID Controller										■	■	■		
10	Submission of Interim Draft Report													●	
11	Submission of Interim Report														●

Figure 3: Gantt Chart of Activities for FYP I, semester September 2014

No.	Activity/Week	1	2	3	4	5	6	7	8	9	10	11	12	13	14
1	Developing the controller model	■	■												
2	Validation of controller on the simulation environment			■	■										
3	Modifications and Improvement of Design				■	■									
4	Preparation of Progress Report				■	■	■								
5	Testing of controller performance						■								
6	Submission of Progress Report							●							
7	Analysis of results of MPC							■	■						
8	Comparison of PID vs MPC									■					
9	Pre-SEDEX									●					
10	Preparation of Final Report									■	■	■			
11	Submission of Draft Final Report											●			
12	Submission of Dissertation / Technical Paper												●		
13	Viva													●	
14	Final Hardcopy Submission														●

Figure 4: Gantt Chart of Activities for FYP II, semester January 2015

3.2 Simulation Procedure

For this term, the process model developed by (Wood & Berry, 1973) was simulated on Simulink as appears in Figure 5. The process dynamic model is shown in Appendix I.

Controllers were initially tuned utilizing Matlab's Auto Tuning (see Appendix II). Table 1 shows the obtained parameters in order to compare which of P, PI and PID is a better option.

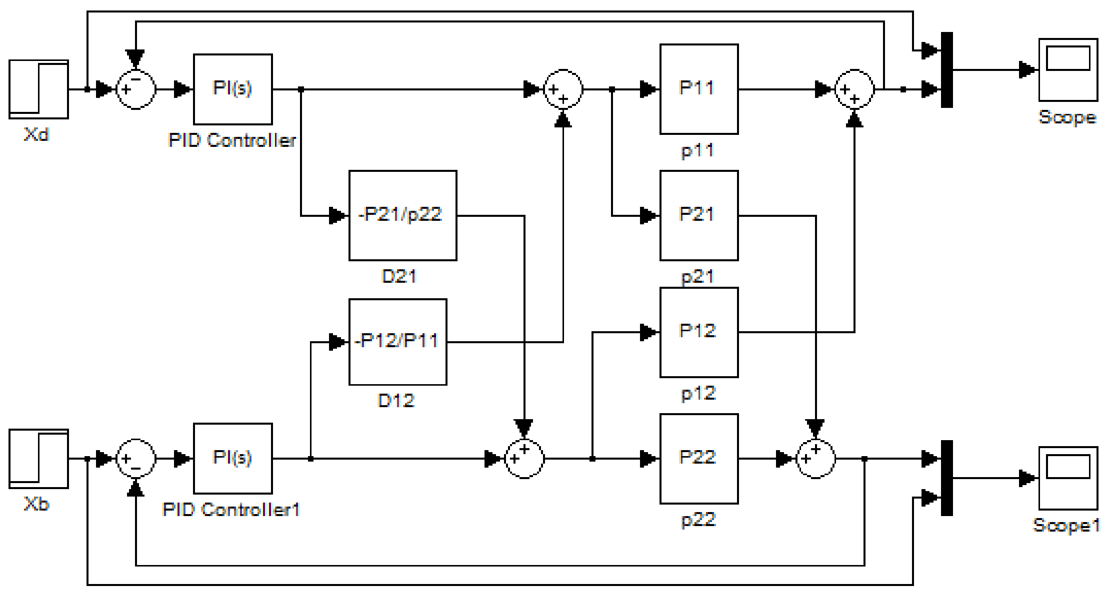

Figure 5: Simulink Block Diagram of the Distillation Process

A step change in the overhead composition X_d from 0 to 10 was introduced to take place at time 10 seconds. Results are discussed in the next subsection.

After the best controller was identified, the controller was tuned using different methods available in the literature such as Ziegler Nichols, Cohen Coon, Internal Model Control (IMC), Integral of Time Absolute Error (ITAE) and Symmetric Optimum. Table 2 & Table 3 summarize the calculated tuned parameters.

Note that the calculated parameters are ought to resemble the Ideal form of a PI controller equation: $K = K_P(1 + \frac{1}{T_i s})$, while Simulink controller settings refers to an equivalent form: $K = P + I\frac{1}{s}$.

Table 1: Auto Tuned Parameters of the Overhead and Bottom Controller

Controller / Parameter	Overhead product controller			Bottom product controller		
	P	**PI**	**PID**	**P**	**PI**	**PID**
Proportional	0.4929983	0.1392848	0.4620595	-0.216033	-0.095715	-0.182426
Integral	-	0.0153928	0.043542	-	-0.01361	-0.006269
Derivative	-	-	0.2356692	-	-	-0.21862
Filter Coefficient	-	-	5.6844125	-	-	0.511546

Table 2: PI Controller parameters for Top Product Controller from Different Tuning Methods

Form of Equation	Ideal		Matlab	
Parameter / Method	**Ki**	**Ti**	**P**	**I**
Zieglar Nichols	1.029339	3.5	1.029339	0.294097
Cohen Coon	1.180729	2.959483	1.180729	0.398965
IMC	0.488647	16.7	0.488647	0.02926
ITAE	0.603526	16.37063	0.603526	0.036866
Symmetric Optimum	0.326172	32	0.326172	0.010193

Table 3: PI Controller parameters for Bottom Product Controller from Different Tuning Methods

Form of Equation	Ideal		Matlab	
Parameter / Method	**Ki**	**Ti**	**P**	**I**
Zieglar Nichols	-0.19409	7.5	-0.19409	-0.02588
Cohen Coon	-0.22698	6.977848	-0.22698	-0.03253
IMC	-0.13746	14.4	-0.13746	-0.00955
ITAE	-0.12709	14.46328	-0.12709	-0.00879
Symmetric Optimum	-0.09278	64	-0.09278	-0.00145

Chapter 4. Results & Discussion

4.1 Comparison of Controllers Settings

As result of the step change in the top composition, the bottom composition was also altered. Hence, both controllers functioned to bring back the measurement to set points. The plots of responses by different controllers' settings were obtained as shown in Figure 6 & Figure 7.

1. P Controller

The proportional only controller shows the response settling at 50 seconds but with an offset of -2.3. Moreover, Figure 7 indicates the behavior of the bottom product response. It was brought to the set point in 140 seconds with and overshoot of 0.22.

2. PI Controller

The proportional-integral controller has a settling time of around 80 seconds with no overshoot for the overhead product composition. Likewise, the bottom product required 100 seconds to settle due to interaction between variable. It is notable that the bottom product response of the PI controller is the least vigorous.

3. PID Controller

The proportional-integral-derivative controller response plot in Figure 6 oscillates at a fast rise time and have a settling time of 70 seconds for the top product. Overshoot is almost negligible after 30 seconds. In the other hand, the bottom product response to the interaction is quite oscillatory with an overshoot of 0.2 and settling time slightly beyond 200 seconds.

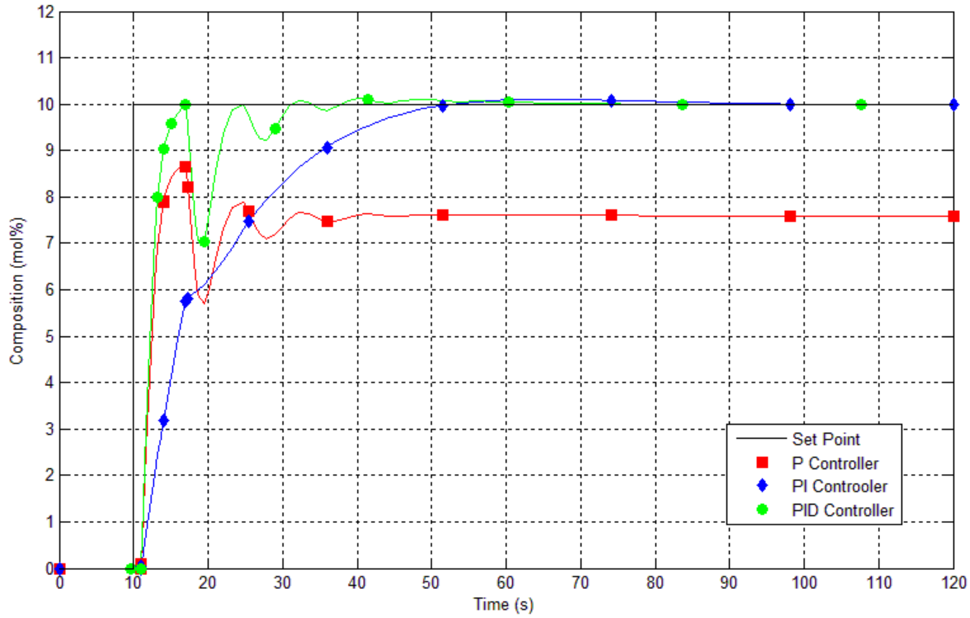

Figure 6: Response of the Top Product to a Step Change in its Controlled Variable by Different Controller Settings

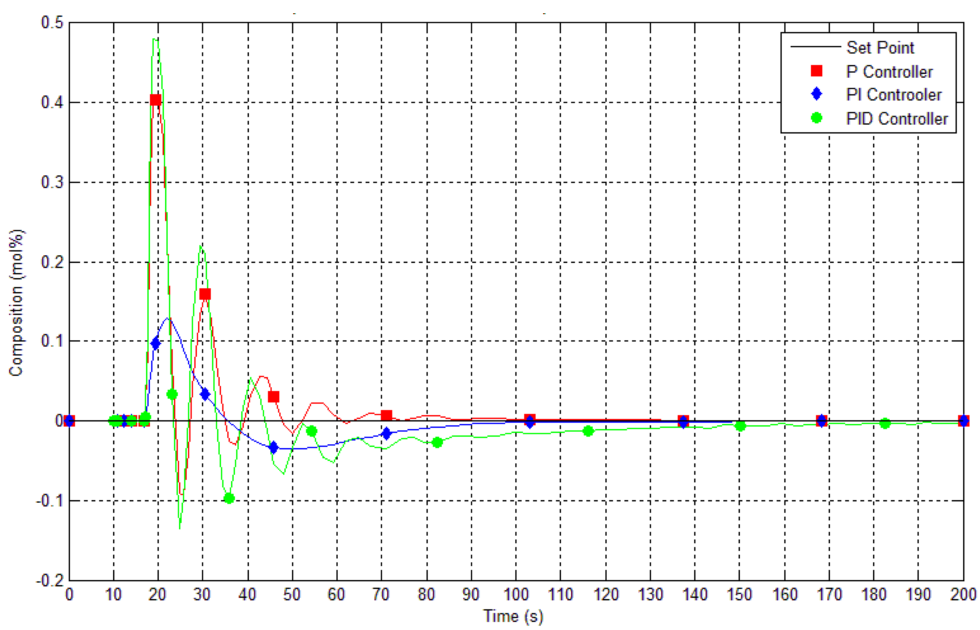

Figure 7: Response of the Bottom Product to a Change in its Disturbance Variable by Different Controller Settings

Hence, based on the results obtained from the simulation which are summarized in Table 4, it is clear that the P controller is unable to maintain stability of control for this problem. In the other hand, in the case of the PI controller, the settling time was higher for the overhead product, 80 seconds compared to 50 seconds, but the main objective of the control was achieved and offset was completely eliminated. Likewise, the PID achieved the set point with even shorter settling time of 70 seconds. However, PI's response rise was steep while PID's was oscillatory.

The comparison is between PI and PID. Considering only the top product control where the step change was introduced, the analysis would favor PID over PI as it required less settling time. Nevertheless, considering the process as a whole, the PI managed to maintain the bottom product more efficiently than PID as latter went beyond 200 second for slow settling time in addition to the vigorous oscillation upon the moment of interaction. PI controller showed an overshoot five times less than that of the PID.

Table 4: Summary of Response Analysis for Controller's Setting Comparison

	Controller / Criteria	P	PI	PID
Top	**Settling time (s)**	50	80	70
	Offset	-2.3	0	0
	Overshoot	1	0	0
	Oscillation	Slight	None	Sligh
Bottom	**Settling time (s)**	140	100	200
	Offset	0	0	0.01
	Overshoot	0.4	0.47	0.2
	Oscillation	Moderate	Slight	Aggressive

4.2 Comparison of Tuning Methods

Following to the decision of the best performaing controller setting which is PI controller, different tuning methods were examined. Valuese from from Table 2 & Table 3 resulted in the varying responses as indicated in Figure 8 &Figure 9.

a. Ziegler Nichols

The Z-N tuning parameters worked well for controlling the top product composition with fastest settling time compared to the other four tuning methods. Overshoot was relatively high with 6.2 mol% change which resembles 62% of the step change introduced.

In the bottom product, the disturbance unsettled the composition for 49 seconds with an overshoot of 0.29. The analysis of Ziegler Nichols response had a sligh oscilation however it was only significant in the bottom product which had a notable overshoot compared to other methods.

b. Cohen Coon

The C-C tuning parameters produce a similar response in handling the step change introduced to the controlled variable as it settled in 42 seconds and had an overshoot of 82% of the change introduced.

Response of the bottom to the coupled disturbance was the least satisfactory. It had the slowest respoinse time, 80 seconds, and largest overshoot, 0.41 mol% change. It also failed to compete the other tuning methods taking into acount the relatively moderate oscilatory response in the bottom product composition control.

c. Integral Model Control

The IMC tuned parameters gave slightly slower response in manage the step change in the top product composition compared to the methods mentioned earlier. However, it was the steepiest and did not record any overshoot whivh gives it a plus point.

Moreover, the bottoms control showed to be superior to the other methods as it elimiated disturbance in 37 seconds with minimal overshoot of 0.03 and negligible oscillation.

d. ITAE

ITAE methodgave a reasonable response yet slower than Z-N, C-C and IMC. Overshoot is 30% and relatively moderate aggressiveness in the response. Bottoms control is also satisfactory with competitive settling time of 41 and small overshoot of 0.05.

e. Symmetric Optimum

Last but not least, the SO method response analysis was the least stable in controlling the top product. It had an offset of -0.2 which is 2% of the step change. The response curve was the smoothest with no oscilation of overshoot at all. However, the settling time was the slowerst going slightly beyong 120 second.

For the bottom product disturbance control, SO had the best control with shortest time, 29 seconds and negligible oscillation and overshoot of 0.015.

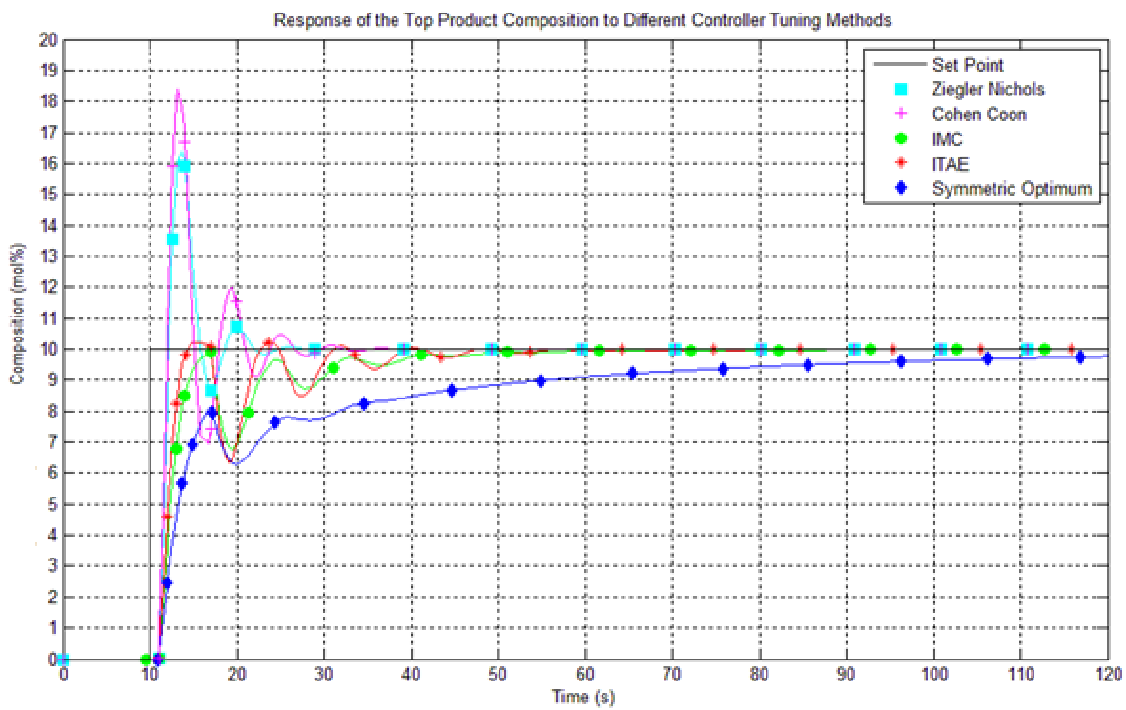

Figure 8: Response of the Top Product to a Step Change in its Controlled Variable by Different Tuning Methods

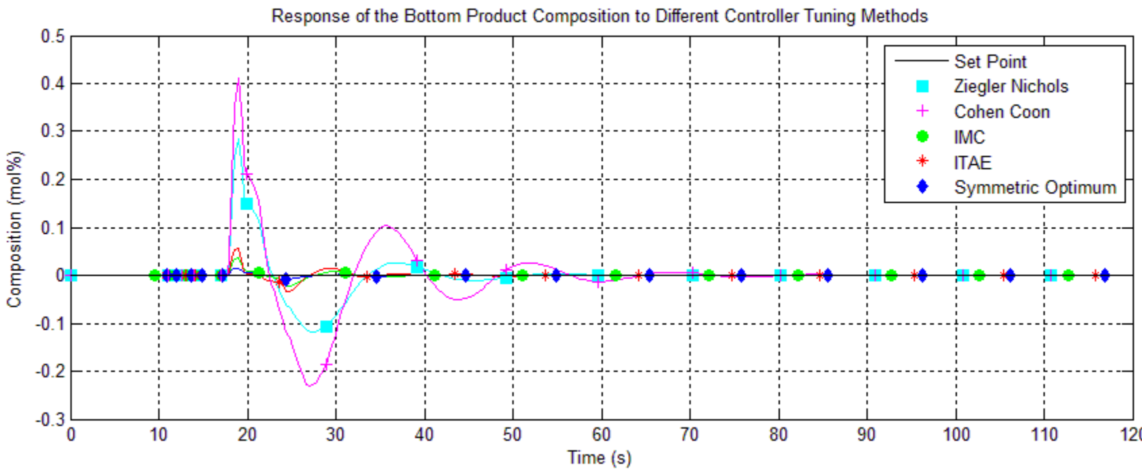

Figure 9: Response of the Bottom Product to a Step Change in its Disturbance Variable by Different Tuning Methods

Consequently, the tuning methods comparison nominates IMC and ITAE for overall superiority to other methods with a slight preference to the IMC. The response is plotted in Figure 8Figure 9 and summurized quantitaively in Table 5. It is notable Symmetric Optimum method gives a better result rejecting the indirect disturbance to bottoms composition 8 and 12 seconds faster than IMC and ITAE respectively but lacks that fast response in controlling the top. Ziegler-Nichols and Cohen Coon methods lacked stability with high overshoot in the top product composition when step change is introduced.

Table 5: Summary of Response Analysis for Tuning Methods Comparison

	Controller / Criteria	ZN	CC	IMC	ITAE	SO
Top	**Settling time (s)**	40	42	49	64	120
	Offset	0	0	0	0	-0.2
	Overshoot	6.2	8.2	0	0.3	0
	Oscillation	Slight	Slight	Slight	Moderate	None
Bottom	**Settling time (s)**	49	80	37	41	29
	Offset	0	0	0	0	0
	Overshoot	0.29	0.41	0.03	0.05	0.015
	Oscillation	Slight	Moderate	Negligible	Slight	Negligible

Chapter 5. Conclusion & Recommendation

In conclusion, the binary distillation column process model was simulated on Simulink. Traditional controllers, P, PI and PID, were set up and tuned using Matlab Auto Tuning Tool. Step change was introduce to the top product. Consequently, the inherent interaction affected stability of the bottom product as well. Response plots were obtained and different controllers were evaluated based on settling time, overshoot and stability of response.

The outcome of the response analysis carried out using the simulation environment favored PI slightly over P and PID. PI controller was then tuned using several tuning methods. IMC tuning parameters gave the best result compared to ITAE, Ziegler Nichol, Cohen Coon and Symmetric Optimum method.

As a result of this study, a PI controller tuned using IMC method is the best representative for the class of traditional controllers.

To improve this study, it is recommended to test more tuning methods to select an ideal traditional controller. Moreover, the decouplers may also be worked out in a different technique for thorough comparison. Lastly. MPC controller is a more advanced class of controllers that is claimed to be superior to traditional controller. It is suggested to be put in comparison against ideally tuned PI controller for the binary distillation column.

All in all, the study has achieved two objectives; the better traditional control setting which is found to be PI, and the better tuning method which is the IMC method. More thorough knowledge in the subject of advanced process control is required to compare the proposed PI controller to MPC.

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Appendix I

Dynamic Model of the Woods & Berry distillation column

$$\begin{bmatrix} X_D(s) \\ X_B(s) \end{bmatrix} = \begin{bmatrix} P_{11}(s) & P_{12}(s) \\ P_{21}(s) & P_{22}(s) \end{bmatrix} \begin{bmatrix} R(s) \\ S(s) \end{bmatrix}$$

Where:

$$P_{11}(s) = \frac{12.8\, e^{-1s}}{16.7s+1} \qquad P_{12}(s) = \frac{-18.9\, e^{-3s}}{21.0b\, s+1}$$

$$P_{21}(s) = \frac{6.6\, e^{-7s}}{10.9\, s+1} \qquad P_{22}(s) = \frac{-19.4\, e^{-1s}}{14.4\, s+1}$$

Decouplers are given by:

$$D_{12}(s)\, P_{11}(s) + P_{12}(s) = 0$$

$$D_{21}(s)\, P_{22}(s) + P_{21}(s) = 0$$

Hence,

$$D_{12}(s) = -\frac{P_{12}(s)}{P_{11}(s)} = \frac{24.6586s + 1.4766}{21.0s + 1}\, e^{-2}$$

$$D_{21}(s) = -\frac{P_{21}(s)}{P_{22}(s)} = \frac{4.8990s + 0.3402}{10.9000s + 1}\, e^{-4}$$

Appendix II

Matlab PID auto tuner is an efficient tool for tuning conventional controllers. The user-friendly interface in the extended design mode enables manipulating the bandwidth and phase margin to achieve an optimal control with respects to the user's prioritized criteria. In this problem, the response was aimed to achieve fastest response with minimal overshoot.

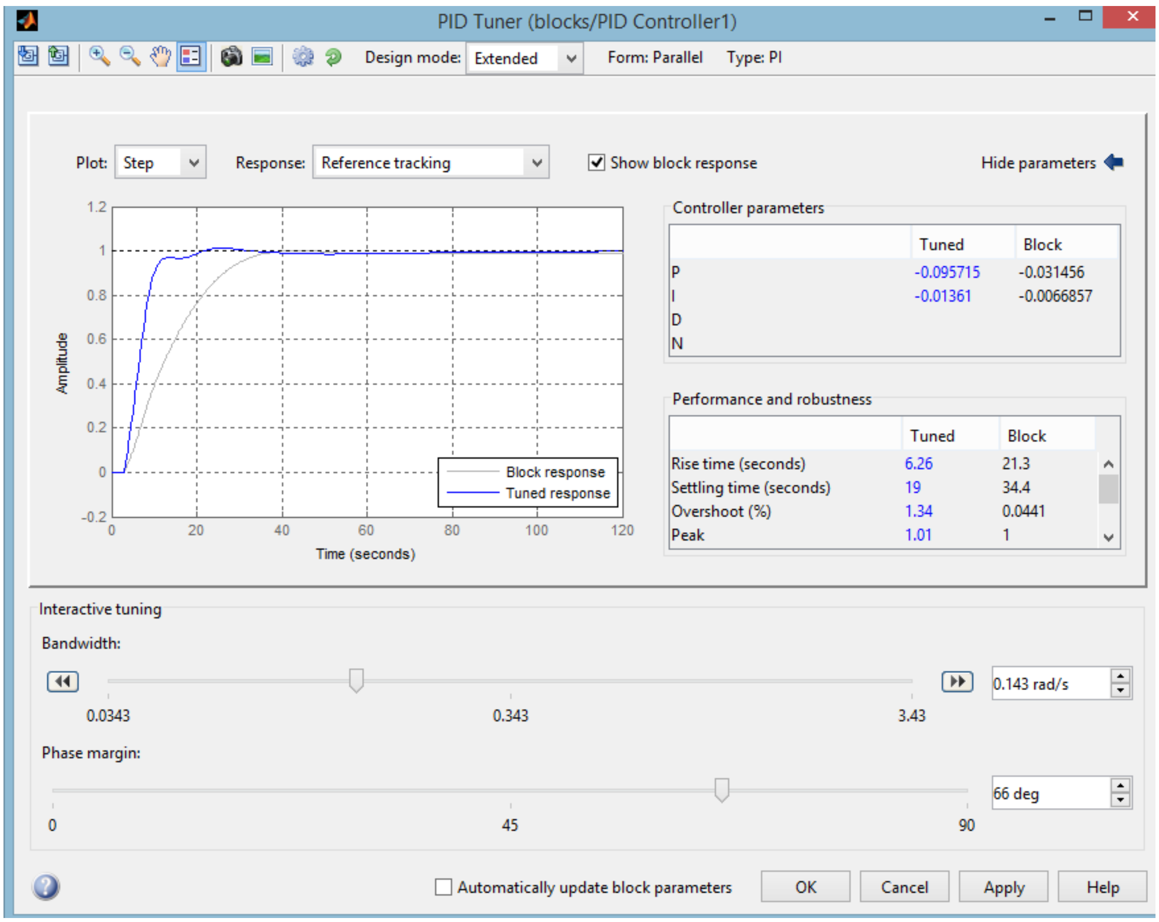

Figure 10: Interface of the PID Auto Tuner Tool in MATLAB