

Comparative Study of Biggest Log Modulus and Direct Synthesis Tuning Methods for Multiloop PI Controllers in a Distillation System

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Abstract

Most chemical processes exhibit multivariable characteristics with complex interactions between control loops; therefore, this study evaluates the performance of the Biggest Log Modulus (BLT) and Direct Synthesis (DS) tuning methods on multiloop Proportional-Integral (PI) controllers within a binary distillation system to determine optimal tuning parameters. Through simulations of the set point tracking scenario (Y_1), the DS method proved superior, yielding an Integral Absolute Error (IAE) of 10.32 (compared to 37.18 for BLT) and settling times of 119 seconds for Y_1 and 111 seconds for Y_2 . Similarly, in disturbance rejection, the DS method demonstrated a more responsive performance with an IAE of 3.392 and settling times of 71 seconds for Y_1 and 52 seconds for Y_2 , confirming that fine-tuning techniques are crucial for minimizing overshoot and maintaining overall system stability. The advantage of the DS method in dampening oscillations provides higher operational certainty for distillation columns, which are highly sensitive to thermal fluctuations and sudden changes in feed composition. Implementing precise parameters through this approach is expected to reduce operational costs resulting from energy waste during control transitions. The results of this research provide a significant contribution to the development of adaptive control strategies in the dynamic chemical process industry in selecting the most effective tuning method to achieve maximum efficiency, workplace safety, and optimal long-term operational stability.

Keywords: *multivariable system, tuning, biggest log modulus, direct synthesis, pi controller, distillation system*

Abstrak

Sebagian besar proses kimia memiliki karakteristik multivariabel dengan interaksi antar loop kontrol yang kompleks, sehingga penelitian ini mengevaluasi kinerja metode *Biggest Log Modulus* (BLT) dan *Direct Synthesis* (DS) pada pengendali *Proportional-Integral* (PI) multiloop dalam sistem distilasi biner untuk menemukan parameter tuning yang optimal. Melalui simulasi pada skenario pelacakan *set point* (Y_1), metode DS terbukti lebih unggul dengan nilai *Integral Absolute Error* (IAE) sebesar 10,32 (dibandingkan BLT sebesar 37,18) serta waktu menetap (*settling time*) 119 detik untuk Y_1 dan 111 detik untuk Y_2 . Begitu pula dalam penanganan gangguan, metode DS menunjukkan performa yang lebih responsif dengan nilai IAE 3,392 serta waktu menetap 71 detik untuk Y_1 dan 52 detik untuk Y_2 , yang menegaskan bahwa teknik *fine-tuning* sangat krusial dalam meminimalkan *overshoot* dan menjaga stabilitas sistem secara menyeluruh. Keunggulan metode DS dalam meredam osilasi memberikan kepastian operasional yang lebih tinggi pada kolom distilasi yang sangat sensitif terhadap fluktuasi termal dan perubahan komposisi umpan secara mendadak. Implementasi parameter yang tepat melalui pendekatan ini diharapkan mampu menekan biaya operasional akibat pemborosan energi selama masa transisi kontrol. Hasil penelitian ini memberikan kontribusi signifikan terhadap pengembangan strategi pengendalian adaptif di industri proses kimia yang dinamis dalam memilih metode tuning yang paling efektif demi mencapai efisiensi maksimal, keamanan kerja, serta stabilitas operasional yang optimal dalam jangka panjang.

Kata Kunci: *sistem multivariabel, penyetelan, modulus logaritmik terbesar, sintesis langsung, pengendali pi, sistem distilasi*

1. Introduction

The chemical industry stands as a dynamic and vital pillar of modern economies, constantly adapting to contemporary shifts. Its robust expansion significantly fuels national economic progress, with sectors like petroleum refining, natural gas processing, fertilizer production, and food manufacturing acting as substantial earners of foreign exchange [1]. Beyond mere economic contributions, these industries serve as

catalysts for local employment, champion environmental stewardship, and drive innovation that elevates the well-being of neighboring communities. At its core, a chemical plant represents a meticulously designed and logically structured assembly of diverse processing units [2-3]. These units, encompassing reactors where chemical transformations occur, pumps for fluid transport, distillation columns for separation, absorbers for gas-liquid contact, evaporators for concentration, and storage tanks, work in concert [4]. The primary objective of an integrated facility of this nature is to facilitate the efficient and economically viable transformation of designated raw materials into valuable final products, thereby maximizing the effective use of available energy resources [5-6].

In this complex operational landscape, effective process control emerges as a critical methodology for achieving these objectives. Through the adoption of advanced control methodologies, chemical manufacturing facilities can guarantee uniformity in product quality, optimize resource utilization, bolster safety measures, and reduce ecological repercussions [7]. The capacity to accurately govern process variables including temperature, pressure, flow rate, and composition is critical for attaining superior operational efficacy and financial success within the chemical sector. Consequently, the study and advancement of process control techniques, including the tuning of multivariable controllers as previously discussed, are essential for the continued progress and sustainability of the chemical sector [8-9].

Despite advancements in control theory, Proportional-Integral (PI) and Proportional-Integral-Derivative (PID) controllers remain the industry standard for chemical processes [10]. This is because these controller types offer various advantages due to their combination of basic control actions. The capability of PI and PID controllers to yield practical industrial processes has led to their widespread acceptance in the industrial world [11]. Koivo and Tanttú [12] estimate that at least only 5-10% of control loops cannot be controlled by PI or PID types. In fact, these two controller types provide good performance on process dynamics [13]. Due to this achieved performance, it is not surprising that even now, more than 95% of process control applications use PI or PID controllers [14]. Complex chemical processes involving multivariables require special attention to the interactions between control loops when adjusting parameters.³ This study will explore and evaluate various methods designed to improve the control of such complex systems [15].

The overarching aim of this investigation is to evaluate and contrast the efficacy of diverse control methodologies that are extensively utilized in the chemical industry, especially within the framework of multivariable systems. Specifically, this research aims to identify the most effective tuning method for addressing interactions between loops in multi-loop control systems [16]. In particular, this study compared the performance of PI controllers tuned using the Biggest Log-Modulus Tuning (BLT) and Direct Synthesis (DS) methods on a distillation column [17-18]. By analyzing settling time, Integral Absolute Error (IAE) values, and the level of oscillation, this research evaluated the impact of both tuning methods on the control system's performance. Consequently, it is expected that this research can provide recommendations for the optimal tuning method to be applied to similar multi-loop control systems.

2. Methodology

This research was conducted at the Chemical Engineering Computer Center (Puskomteka) of Syiah Kuala University using Simulink and Matlab R2013a software. Simulations were performed on a standard specification Asus K45DR notebook suitable for running the software. The research variables investigated were the performance of PI controllers in a multi-loop system tuned using three different methods: Biggest Log-Modulus Tuning (BLT), Direct Synthesis and Fine Tuning.

2.1 Process Model Establishment

To analyze the system's response to set point changes, this research employs a first-order plus dead time (FOPDT) model that has been adjusted with experimental data, specifically:

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

This model is based on a binary distillation column [19]. The input and output system of the column is as shown in **Figure 1**. In this distillation column control system, X_D (distillate composition) and X_B (bottoms composition) act as the controlled variables, F_R (reflux flow rate) and F_V (vapor boil-up rate) are the manipulated variables, while X_F (feed composition) serves as the disturbance. The identified process model, a first-order plus dead time (FOPDT) system, is simulated using Simulink in MATLAB to analyze the system's stability against parameter changes.

The transfer function model used in this study is a development of the basic model proposed by Rajapandiyan and Chidambaram [20], which has been calibrated using experimental data from process reaction curves.

The transfer function for this system is given in Eq. 1.

$$\begin{bmatrix} X_D \\ X_B \end{bmatrix} = \begin{bmatrix} \frac{0.0747e^{-3s}}{12s+1} & \frac{-0.0667e^{-2s}}{15s+1} \\ \frac{0.1173e^{-3.3s}}{11.7s+1} & \frac{-0.1253e^{-2s}}{10.2s+1} \end{bmatrix} \begin{bmatrix} F_R \\ F_V \end{bmatrix} + \begin{bmatrix} \frac{0.70e^{-5s}}{14.4s+1} \\ \frac{1.3e^{-3s}}{12s+1} \end{bmatrix} X_F \quad (1)$$

This model represents the dynamic relationship between input variables—namely the reboiler flow rate (F_R), reflux flow rate (F_V), and feed composition (X_F) and output variables, which are the top product composition (X_D) and bottom product composition (X_B) in a distillation system. Each relationship between variables is modeled using a first-order transfer function with time delay, in the form of $\frac{K_p e^{-\theta s}}{\tau s + 1}$, indicating the process dynamics such as gain, time constant, and time delay. The main 2×2 transfer function matrix reflects the influence of each input (F_R and F_V) on the two main outputs, while the additional transfer function vector shows the direct effect of X_F on the system outputs. This model is developed to more accurately describe the dynamic behavior of the system and can be used in the design of multivariable control systems that take into account the interactions between variables as well as the time-delay characteristics of the distillation process.

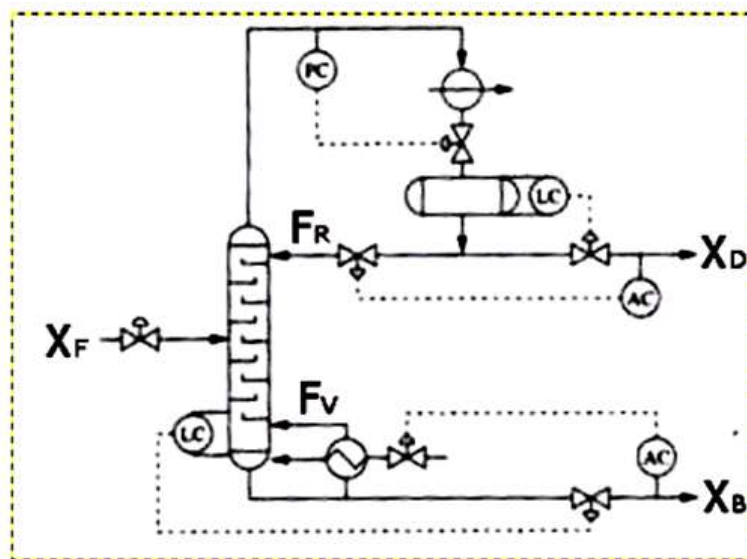


Fig. 1: Distillation Column Control System (Atsushi, et al., 1995).

2.2 Process Dynamics Testing

To comprehend the essential attributes of the process, preliminary simulations were performed within an open-loop framework. Changes in the setpoint or disturbances were introduced to observe the process response without any feedback. Subsequent to this modification, the system was reconfigured to a closed-loop architecture through the integration of a Proportional-Integral (PI) controller. The efficacy of the controller was subsequently assessed by analyzing the system's response to a range of setpoint variations and external disturbances. **Figure 2** illustrates the research process flow diagram. It begins with the process model that has been implemented in Simulink. An open-loop analysis is then performed to understand the system's behavior without control. This is followed by evaluating the performance of a multiloop control system using the Ziegler-Nichols tuning method. Based on this, PI controller tuning values are determined using three methods: Biggest Log-Modulus Tuning, Direct Synthesis Tuning, and Fine Tuning. Subsequently, the controller performance is tested using each of these tuning methods. Finally, the process performance results are analyzed to assess the effectiveness of each tuning approach. The schematic of the process dynamics testing can be seen in **Figure 2**, as follows:

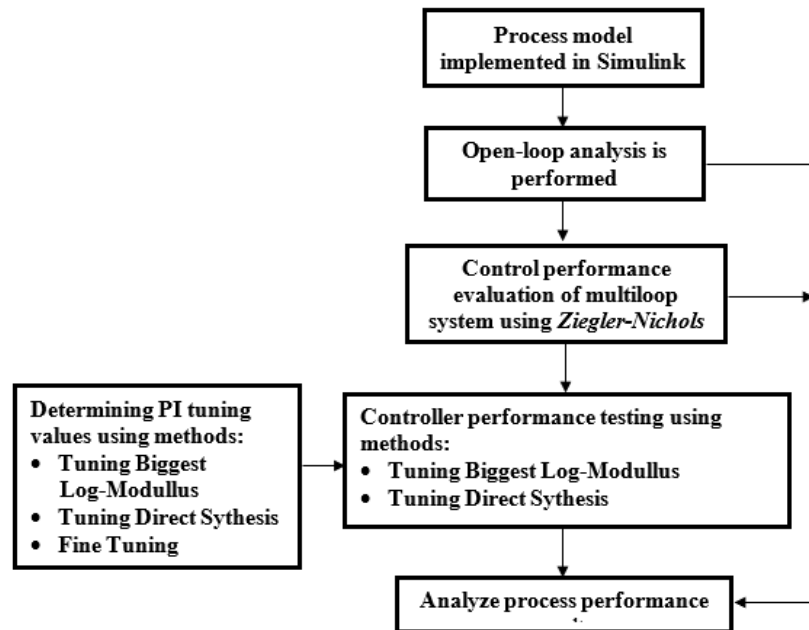


Fig. 2: Research Process Flow Diagram

2.3 Data Analysis

The IAE value can be determined in two ways: using Simulink and using response value data over time. In the IAE calculation with Simulink, an IAE block is added to the block diagram, and then the errors from loop 1 and loop 2 are summed into the IAE block. Meanwhile, for the IAE calculation from the response data, to obtain the response data from Simulink, synchronization between Simulink and the MATLAB window needs to be performed by adding a Time block and a Workspace block as data readers for the response results. After Simulink and the window are synchronized, the time and response data can be called through the MATLAB window. It is from this data that the IAE value is calculated using Integration Equation 2.

$$IAE = \int_a^b f(x) dx \quad \text{where, } f(x) = f(a) + \frac{f(b)-f(a)}{b-a}(x-a) \quad (2)$$

The function $f(x) = \int_0^{\sim} |CV - SP|$, where the notation CV represents the control variable or the response over time, and SP represents the set point. The settling time value is obtained using the process response data. From this data, the difference in the response value to a change in the set point of $\pm 5\%$ from $t = 0$ until the difference becomes constant is observed.

3. Results and Discussion

3.1 Open Loop System

An open-loop system operates without any feedback mechanism to correct errors. The simulation in **Figure 3** illustrates the basic structure of a binary distillation system in open-loop mode. **Figure 3** illustrates the open-loop structure of a binary distillation column, as presented by Athanassios [21], depicting the interconnectedness of its primary input and output variables through transfer function blocks. The diagram shows two manipulated variables, reflux flow rate (FR) and boil-up rate (FV), acting as inputs to the system. These inputs dynamically influence the two controlled output variables, top product composition (Y1) and bottom product composition (Y2), through a 2x2 transfer function matrix represented by the blocks G11, G12, G21, and G22. Additionally, the diagram incorporates a disturbance variable, feed composition (XF), which affects both output compositions through the disturbance transfer function blocks Gd1 and Gd2. The summing junctions indicate how the effects of the manipulated variables and the disturbance are combined to determine the final values of the top and bottom product compositions in this open-loop configuration, where no feedback control is implemented to regulate the outputs against setpoint deviations or the influence of the disturbance.

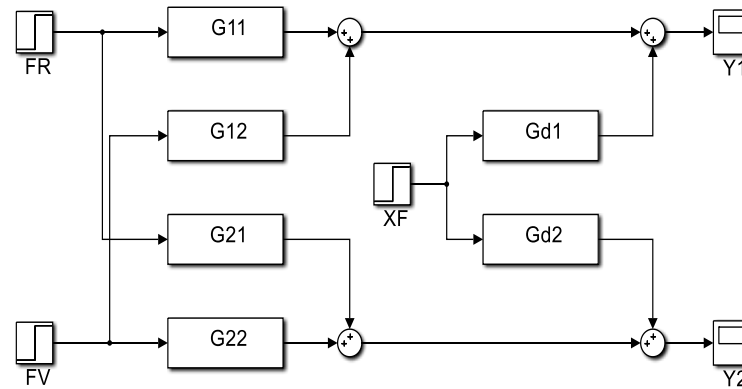


Fig. 3: Interconnection of distillation components in block diagram form [21]

To analyze the dynamic behavior of the distillation column, an open-loop simulation was conducted by introducing a step change in key operating variables as a disturbance. **Figures 4, 5 and 6** illustrate the distillate (Y1) and bottom (Y2) concentration responses to a step change in reflux flow rate (F_R), reboiler flow rate (F_V), and feed composition (X_F) in the open-loop simulation.

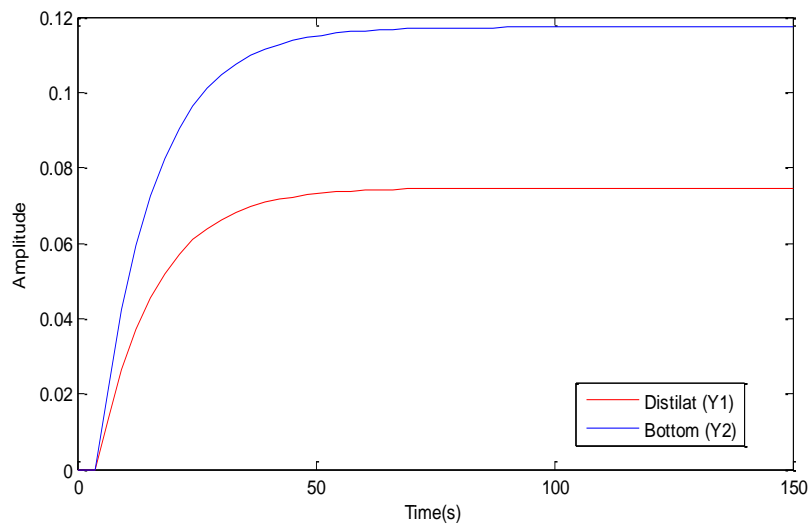


Fig. 4: Effect of a sudden 1unit step increase in F_R on the values of Y1 and Y2

Figure 4 presents the open-loop dynamic response of the distillate (Y1) and bottom (Y2) product compositions of the binary distillation column following a step change in the reflux flow rate (F_R). The x-axis represents time in seconds, while the y-axis displays the amplitude of the concentration response. The red curve illustrates the temporal evolution of the distillate composition (Y1), exhibiting a gradual increase from its initial steady-state value and eventually settling at a new, higher steady-state concentration of approximately 0.075. Concurrently, the blue curve depicts the response of the bottom product composition (Y2) to the same step change in reflux flow rate. This response also shows a gradual increase, reaching a different and higher steady-state concentration of approximately 0.118. The distinct dynamic characteristics and final steady-state values observed for both Y1 and Y2 highlight the influence of reflux flow rate on the separation process within the distillation column under open-loop conditions, indicating the system's inherent dynamic behavior and the eventual equilibrium shifts resulting from the input perturbation.

Figure 5 illustrates the open-loop dynamic response of the distillate (Y1) and bottom (Y2) product compositions following a step change in the reboiler flow rate (F_V). The x-axis represents time in seconds, while the y-axis displays the amplitude of the concentration response. The red curve shows the temporal evolution of the distillate composition (Y1), exhibiting a rapid initial decrease from its initial steady-state value and gradually settling at a new, lower steady-state concentration of approximately -0.065.

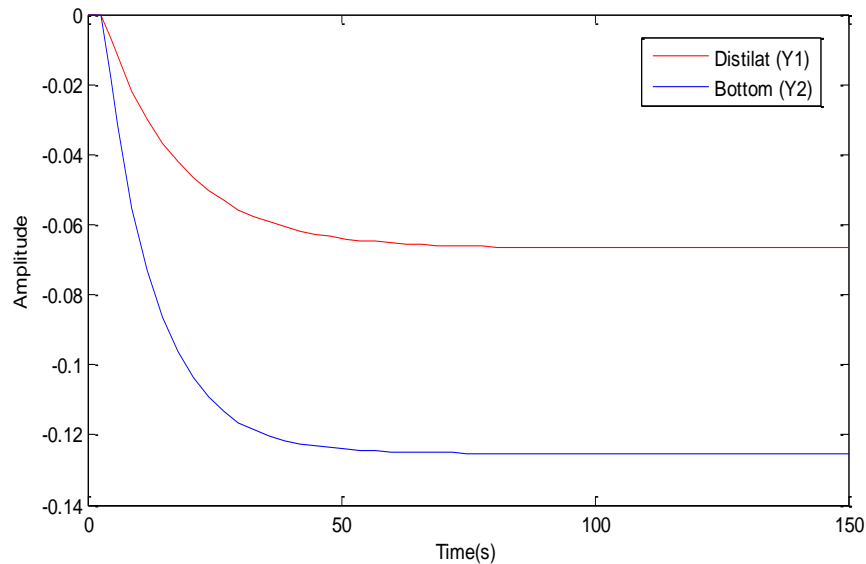


Fig. 5: Response curves of Y1 and Y2 to a significant 1 unit step change in F_V .

Simultaneously, the blue curve depicts the response of the bottom product composition (Y2) to the same step change in reboiler flow rate. This response also shows a rapid initial decrease, eventually stabilizing at a significantly lower steady-state concentration of approximately -0.125. The distinct and substantial negative shifts in the steady-state values of both Y1 and Y2 underscore the considerable influence of the reboiler flow rate on the separation efficiency within the distillation column under open-loop conditions, highlighting the system's sensitivity to this particular manipulated variable.

Figure 6 illustrates the open-loop dynamic response of the distillate (Y1) and bottom (Y2) product compositions following a step change in the feed composition (X_F). The x-axis represents time in seconds, while the y-axis displays the amplitude of the concentration response. The red curve depicts the temporal evolution of the distillate composition (Y1), showing a gradual increase from its initial steady-state value and eventually settling at a new, higher steady-state concentration of approximately 0.69. Simultaneously, the blue curve illustrates the response of the bottom product composition (Y2) to the same step change in feed composition. This response also exhibits a gradual increase, reaching a different and higher steady-state concentration of approximately 1.30. The distinct dynamic characteristics and the resulting elevated steady-state values for both Y1 and Y2 demonstrate the significant influence of the feed composition on the separation outcomes within the distillation column under open-loop conditions, highlighting the system's sensitivity to changes in this crucial input variable.

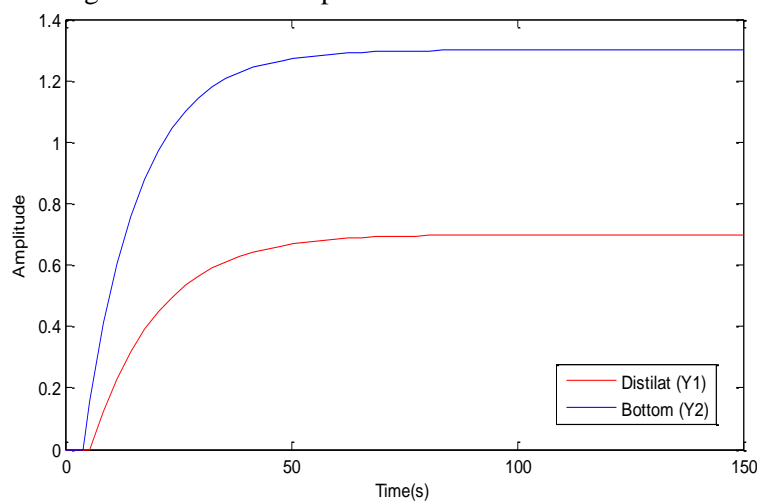


Fig. 6: Response curves of Y1 and Y2 to a 1 unit step change in feed composition (X_F).

The simulation results demonstrate a complex interaction between the distillation process variables. Changes in the reflux flow rate (F_R) not only impact the distillate concentration (Y1) but also the bottom concentration (Y2). Similarly, alterations in the reboiler vapor flow rate (F_V) and feed composition (X_F)

will influence the process equilibrium. It is important to note that changes in FV often produce a response contrary to the desired set point, thus requiring careful tuning.

3.2 Closed Loop System

Figure 7 illustrates the closed-loop control structure of the binary distillation column, incorporating Proportional-Integral (PI) controllers for regulating both the top (Y1) and bottom (Y2) product compositions. The diagram shows two control loops, each with a setpoint (SP), a PI controller (Gc1 and Gc2), and the distillation process represented by the transfer function matrix (G11, G12, G21, G22). Feedback signals from the outputs (Y1 and Y2) are compared to their respective setpoints, generating error signals that are then processed by the PI controllers to manipulate the reflux flow rate (F_R) and boil-up rate (F_V), respectively, aiming to minimize the deviations. The disturbance variable, feed composition (X_F), also influences the outputs through disturbance transfer functions (Gd1 and Gd2), highlighting the closed-loop system's role in mitigating the impact of such disturbances on the controlled variables.

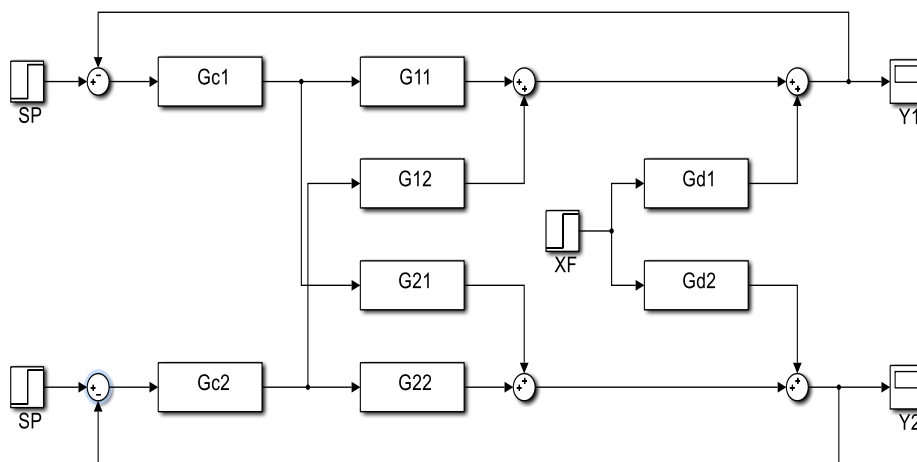


Fig. 7: Block diagram of a closed-loop control system with PI controllers

(1) Ziegler-Nichols Methods

Table 1 presents the initial Proportional Gain (K_c) and Integral Time (T_i) values obtained from the Ziegler-Nichols tuning method for two control loops (Loop I and Loop II) within the distillation column control system. For Loop I, the Ziegler-Nichols method yielded a proportional gain (K_c) of 41.9351 and an integral time (T_i) of 7.8266. Similarly, for Loop II, the method provided a proportional gain (K_c) of 87.6148 and a shorter integral time (T_i) of 2.2093. These values represent the initial controller settings derived from the Ziegler-Nichols heuristic, intended as a starting point for achieving satisfactory closed-loop performance for each individual control loop.

Table 1: K_c and T_i values obtained from the Ziegler Nichols tuning method for pi controllers.

Loop	K_c	T_i
I	41,9351	7,8266
II	87,6148	2,2093

Figure 8 illustrates the closed-loop system response of the top (Y1) and bottom (Y2) product compositions to a 1unit step change in their respective setpoints, utilizing the PI controller parameters derived from the Ziegler-Nichols tuning method. The top subplot shows the response of Y1, exhibiting an oscillatory behavior with an initial delay before rising, overshooting the new setpoint, and then oscillating with increasing amplitude, suggesting instability or poor damping. Similarly, the bottom subplot displays the response of Y2, which also demonstrates an oscillatory pattern with a delayed initial response, overshoot, and subsequent growing oscillations. The unstable or poorly damped responses for both Y1 and Y2 indicate that the Ziegler-Nichols tuning method, in this specific multivariable distillation column control application, results in unsatisfactory closed-loop performance, necessitating further controller tuning or a more sophisticated control strategy.

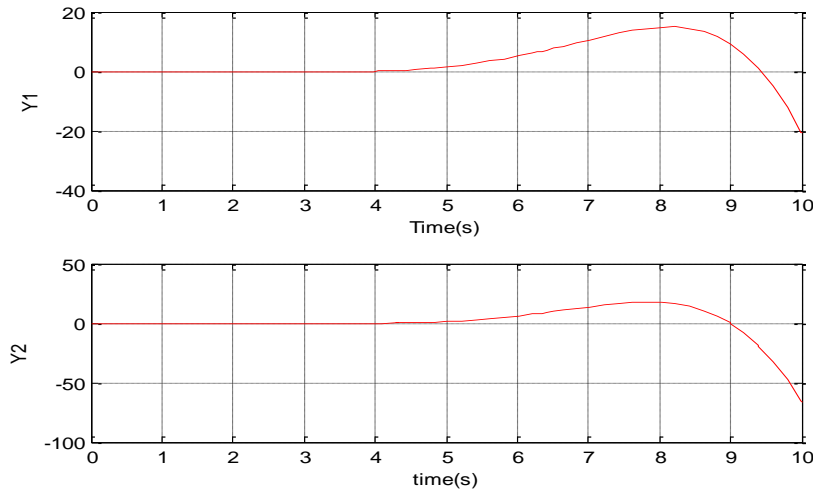


Fig. 8: Response of variables Y1 and Y2 to a 1unit step change in set point using the Ziegler-Nichols tuning method

(2) BLT And DS Methods

The determination of the PI controller parameters, namely the K_c and T_i values, using the BLT and DS tuning methods can be seen in the following table. **Table 2** shows the K_c and T_i values obtained from the BLT and DS methods for controllers Gc1 and Gc2.

Table 2: Calculation results of Pi controller parameter values using the BLT and DS methods

Methods	Loop	K_c	K_i
BLT (Biggest Log Modulus)	I	19,0614	0,0581
	II	-39,8249	0,2057
DS (Direct Synthesis)	I	4.3428	1.4312
	II	-49.4237	0.1427

Figure 9 compares the closed-loop responses of top (Y1) and bottom (Y2) compositions to a step change in the Y1 setpoint, with PI controllers tuned using BLT and DS methods (parameters in **Table 2**). For Y1, the DS-tuned controller ($K_c = 4.3428$ for Loop I) exhibits a faster rise time and less overshoot compared to the BLT-tuned controller ($K_c = 19.0614$ for Loop I), which shows a slower approach to the new setpoint.

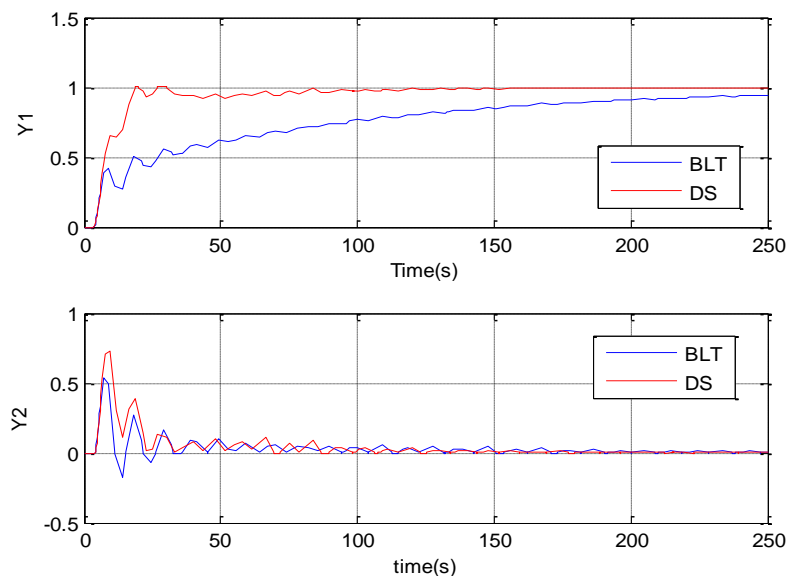


Fig. 9: Comparison of system responses Y1 and Y2 when the Y1 set point is increased by 1 unit step, using the BLT and DS methods

Observing Y2, the DS-tuned controller ($K_c = -49.4237$ for Loop II) induces more pronounced and rapid oscillations initially compared to the BLT-tuned controller ($K_c = -39.8249$ for Loop II), which results in a smoother but slower initial change in Y2 due to the setpoint change in Y1. The integral gains (K_i) also contribute to the steady-state behavior and the speed of error elimination in both loops for each tuning method.

Figure 10 shows how a step change in the Y2 setpoint affects Y1 and Y2 under BLT and DS control. For Y1, DS causes larger oscillations than BLT. For Y2, DS leads to a faster but more oscillatory response to the new setpoint compared to the smoother, slower response of BLT. This illustrates the different dynamic characteristics imparted by each tuning method on the coupled control loops.

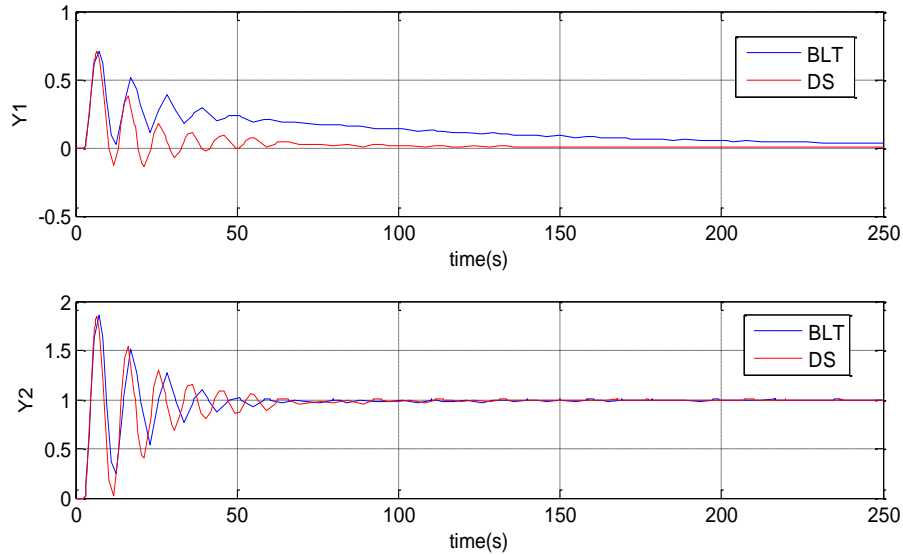


Fig. 10: Comparative analysis of the response of variables Y1 and Y2 to a 1 unit step change in the Y2 set point, applying the BLT and DS tuning methods.

Figure 11 compares how the BLT and DS-tuned controllers handle a step disturbance affecting the system. For Y1 (top), both methods show initial oscillations, but the DS method (red) settles back to the original value faster with less sustained oscillation compared to the BLT method (blue). Similarly, for Y2 (bottom), the DS method exhibits a quicker return to the initial steady state after the disturbance, though with some initial larger oscillations, while the BLT method shows a more gradual recovery. This suggests that under a step disturbance, the DS-tuned controllers provide a faster disturbance rejection for both output variables compared to the BLT-tuned controllers.

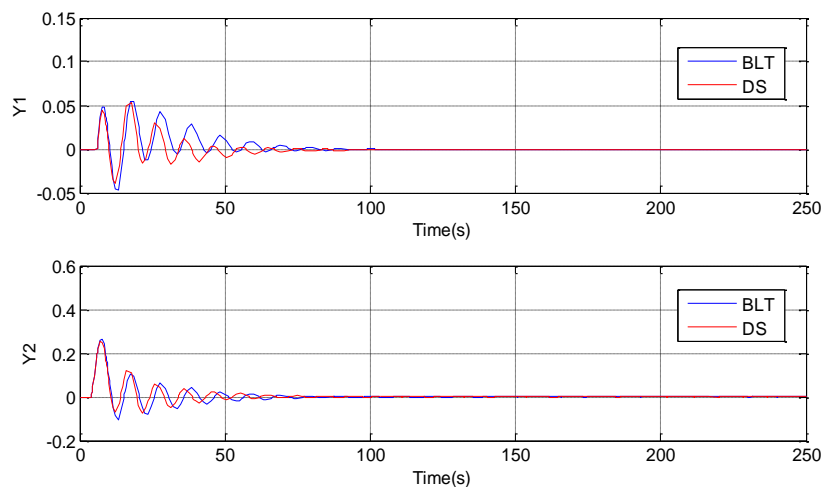


Fig. 11. Comparative study of the system response of Y1 and Y2 to a unit step disturbance signal, using the BLT and DS tuning methods

Based on the observed responses in **Figures 9, 10, and 11**, a quantitative analysis of the control performance reveals notable differences between the BLT and DS tuning methods. For setpoint tracking of Y1 (**Figure 9**), the DS method demonstrates superior performance, exhibiting a significantly lower integrated absolute error (IAE) of 10.32 compared to the BLT method's 37.18, and achieving settling times of 119s for Y1 and 111s for Y2, both considerably faster than the BLT method's 250s and 124s, respectively. Similarly, when the Y2 setpoint is changed (**10**), the DS method generally leads to a lower overall error (IAE of 22.57 vs. BLT's 30.5) and quicker stabilization, with settling times of 95s for Y1 and 75s for Y2, again outperforming BLT's 228s and 69s. Furthermore, when the system is subjected to a disturbance (**Figure 11**), the DS-tuned controller effectively minimizes the impact, resulting in a lower IAE of 3.392 compared to BLT's 4.03, and achieving faster settling times of 71s for Y1 and 52s for Y2, while BLT requires 92s and 71s, respectively. This suggests that the DS tuning method provides better integrated error and faster settling characteristics across different operating scenarios compared to the BLT tuning method, as evidenced by the system's dynamic responses.

4. Conclusion

The analysis of the binary distillation system through open-loop and closed-loop simulations confirms the profound influence of manipulated variables—such as reflux and reboiler flow rates—on system dynamics. While open-loop configurations fail to maintain steady-state under fluctuations, closed-loop control using the Ziegler-Nichols method yields unsatisfactory results marked by large oscillations and instability. In contrast, the Direct Synthesis (DS) method consistently outperforms the BLT method. In set-point tracking for Y₁, DS achieved an IAE of 10.32 compared to BLT's 37.18, with a faster settling time of 119 seconds. Similarly, in disturbance rejection, DS produced a superior IAE of 3.392. Furthermore, this study demonstrates that an iterative fine-tuning approach effectively refines controller parameters beyond conventional limits, achieving minimal overshoot and enhanced stability compared to standalone DS applications. Despite these gains, this research is limited to a linear model that does not account for extreme nonlinear thermal dynamics. Future research should explore intelligent control algorithms, such as Fuzzy Logic or Model Predictive Control (MPC), to enhance robustness against complex feed fluctuations in nonlinear multivariable systems. Selecting such advanced tuning strategies is critical for ensuring stable, responsive, and energy-efficient industrial distillation processes.

5. Acknowledgment

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6. References

- [1] V. Kanchana, S. Jagannathan, and S.A. Shanmuga, "The recent developments of green and sustainable chemistry in multidimensional way: current trends and challenges. *Green Chemistry Letters and Reviews*, vol. 17, no. 1, 2024. <https://doi.org/10.1080/17518253.2024.2312848>.
- [2] Ogodo, P. Uchenna, and Abosede, O. Olufunso, "The Role of Chemistry in Achieving Sustainable Development Goals: Green Chemistry Perspective", *International Research Journal of Pure and Applied Chemistry*, vol. 26 (1), pp. 1-8, 2025. <https://doi.org/10.9734/irjpac/2025/v26i1893>.
- [3] Minglu Xia, "Research on Measurement of Manufacturing Industry Chain Resilience Based on Index Contribution Model Driven by Digital Economy," *De Gruyter, Economics*, vol. 18, pp. 20220123, 2024.
- [4] J. Pang, Y. Zhang, & F. Jiao, "The impact of the digital economy on transformation and upgrading of industrial structure: A perspective based on the "Poverty Trap", *Sustainability*, vol. 15(20), pp. 15125, 2023.
- [5] H. F. Rahmani, and H.R. Teymori, "The effect of economic uncertainty on the supply chain risks of industrial plants," *Journal of Applied Research on Industrial Engineering*, vol. 11(3), pp. 487–496, 2024.
- [6] A. Rezaei, & L. Qiong, "Robust supply chain network design with resilient supplier selection under disruption risks", *Journal of Applied Research on Industrial Engineering*, vol. 11(3), pp. 398–422, 2024.
- [7] N. Nuria, P. Adela, B. Francesc, E. Lluís, and D. Xavier, "Reinforcement Learning for Autonomous Process Control in Industry 4.0: Advantages and Challenges", *Applied Artificial Intelligence*, vol. 38, no. 1, 2024. <https://doi.org/10.1080/08839514.2024.2383101>.

- [8] M. K. Chakravarthi, Y. V. Pavan Kumar, G.P. Reddy, “Potential Technological Advancements in the Future of Process Control and Automation”, *2024 IEEE Open Conference of Electrical, Electronic and Information Sciences (eStream)*. doi:10.1109/eStream61684.2024.10542581.
- [9] B.R. Douglas, C. Yugender, M.P. Kalpesh, H.A. Abdullah, S.S. Hussain, S.B. Ammar, S.P. Rohit, “Machine learning & conventional approaches to process control & optimization: Industrial applications & perspectives”, *Computers & Chemical Engineering*. Vol. 189, pp. 108789, 2024. <https://doi.org/10.1016/j.compchemeng.2024.108789>.
- [10] K.J. Åström, and T. Hägglund, “PID control”, *IEEE Control Syst. Mag.*, 1066, 033X/06, 2006.
- [11] J.G. Ziegler, N.B. Nichols and N.Y. Rochester, “Optimum settings for automatic controllers”, *J. Dyn. Syst. Meas. Control*, vol. 115, 2B, pp. 220-222, 1993.
- [12] H.N. Koivo, and J.T. Tanttu, “Tuning of PID controllers: Survey of SISO and MIMO techniques”, *IFAC Symposia Series*, vol. 7, pp. 75-80, 1991.
- [13] T.N.L. Vu, and M. Lee, M. (2010), “Multi-loop PI controller design based on the direct synthesis for interacting multi-time delay processes”, *ISA Transactions*, vol. 49, 1, pp. 79-86, 2010.
- [14] A. O’Dwyer, “Handbook of PI and PID Controller Tuning Rules”, 3rd edn. *Imperial College Press*, London, 2009.
- [15] A. Wahid, and N.S. Maro, “Multi Input Multi Output (MIMO) Control 2x2 at Vacuum Distillation Unit for LVGO, MVGO, and HVGO Production”, in *E3S Web of Conferences, EDP Sciences*, 3012, 2018.
- [16] F. Zhi-Yong, G. Huiru, S. Jinhua, and X. Li, “Weighted sensitivity design of multivariable PID controllers via a new iterative LMI approach”, *Journal of Process Control*, vol. 110, pp. 24-34, 2022.
- [17] D. Chen and D.E. Seborg, “PI/PID Controller Design Based on Direct Synthesis and Disturbance Rejection”, *Ind. Eng. Chem. Res.*, vol. 41, pp. 4807-4822, 2002.
- [18] R. Dittmar, S. Gill, H. Singh, and M. Darby, “Robust optimization based multiloop PID controller tuning: A new tool and its industrial application”, *Control Engineering Practice*, vol. 20(4), pp. 355–370, 2012.
- [19] M. Atsushi, T.E. Marlins, and N.H. Andrew, “Operations optimization and control design for a petroleum distillation process”, *The Canadian Journal of Chemical Engineering*, vol.73 (6), pp. 896, 1995.
- [20] C. Rajapandiyan, and M. Chidambaram, “Controller Design for MIMO Processes Based on Simple Decoupled Equivalent Transfer Functions and Simplified Decoupler”, *Industrial & Engineering Chemistry Research*, vol. 51 (38), pp. 12398–12410, 2012.
- [21] K. Athanassios, P. Jasmin, T.E. Marlin, “Integrating process and controller models for the design of self-optimizing control”, *Computers and Chemical Engineering*, vol. 24, pp. 2589–2602, 2000.