Robust PID Control Design in CPS-based Batch Distillation Column

Wirenda Sekar Ayu
Control and Computer Systems Research Group
School of Electrical Engineering and Informatics
Institut Teknologi Bandung
Indonesia
wirendasekarayu@students.itb.ac.id

Pranoto Hidaya Rusmin
Control and Computer Systems Research Group
School of Electrical Engineering and Informatics
Institut Teknologi Bandung
Indonesia
pranoto@lskk.ee.itb.ac.id

Egi M. I. Hidayat
Control and Computer Systems Research Group
School of Electrical Engineering and Informatics
Institut Teknologi Bandung
Indonesia
egi.hidayat@lskk.ee.itb.ac.id

***Abstract*— Interconnected system between computation and physical process (Cyber-Physical Systems) has been widely used in industrial processes. In CPS-based industrial process, sensors, controllers, and actuators are connected into a communication network. The communication network may introduce delay time uncertainties due to shared resources and load traffic in the network. Furthermore, the nonlinear time-varying characteristic of batch distillation column may causes another uncertainties to take into account in control system design. Parameter model and delay process uncertainty is introduced due to linearized system approximation that unmodeled high-frequency dynamics. The dynamic uncertainty on both I/O channel are also introduced to the system uncertainty. In this paper, robust PI and PID controller using AMIGO method with appropriate weighting function is designed to guarantee robust stability spesification of batch distillation column. The impact of system uncertainties to closed-loop system performances such as peak overshoot and integral error is investigated. MATLAB/Simulink simulation is used to validate the methods before its implementation in CPS-based batch distillation column. Based on simulation, the proposed robust PI/PID controller can guarantee robust stability of system compared to conventional PID controller. Furthermore, the robust PI/PID controller can improve closed-loop system performances compared to conventional PID.**

Keywords—Cyber-Physical Systems, Distillation Column, Robust PID, Robust Stability, Integral Error

I. INTRODUCTION

Recent technological advance in industrial control system is integrating computation, communication, and control technology through a network. The integration of computation with physical process in a large-scale interconnected systems is commonly called as cyber-physical systems (CPS). CPS tightly integrate components for sensing, actuating, and computing into distributed feedback loops to directly control physical processes [1]. Several issues regarding definition, realization, and challenges of CPS has been discussed recently [2, 3, 4].

Implementation of CPS is a new areas to realize systems that exchange data through a low-power wireless communication medium. The system to be controlled is known as networked control systems (NCSs), whose operation of actuators, sensors, and controllers is coordinated through some form of communication network [4]. Fig.1 shows block diagram of network-based closed loop control systems. Such systems offer benefits such as maintenance and installation simplicity, greater flexibility, and cost implementation efficiency. NCSs give better communication within industry to fulfill the requirement of production planning and control integration, quality management, traceability, maintenance systems and many others [5].

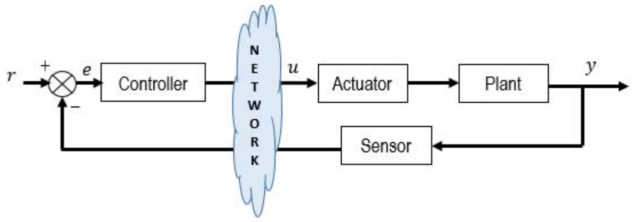

Fig. 1 Block Diagram of Network-based Control Systems

Networked control systems use sensor to monitor production process of the plant. The measured variables from sensor is sent to the controller through communication network, whose control signal is computed. Actuator receives and executes signal produced by the controller as an input of the plant to achieve the specified goals. These component elements are spatially distributed, may operate asynchronously, but have cooperated operation to achieve system objectives.

Cyber-physical systems put emphasis on meeting real-time demands in controlling data flow in the network. During data exchange among component elements in CPS, time delay will occur due to constraint network resources, or even varying network load. This delay, either constant or time-varying, can degrade control system performances, both in time domain and frequency domain. It can even destabilize the control system [6]. The amount of each data sent may not have a big size, but when a lot of data is sent through the network at the same time, there will be a network congestion. In particular, this paper gives attention on network-induced delay caused by varying load in the process of data transmission as one of parameter uncertanty to design control system.

Furthermore, nonlinear time-varying characteristic of batch distillation column gives more concern on system uncertainties. The linearized model of distillation column that used in this reaserch may cause error model due to unmmodelled high-frequency dynamics. The incompatibility of system model in controller design process with the actual system model lead to inappropriate controller performance. The model uncertainty is not the only concern when it comes to robustness. Other considerations include time-varying process delay, unmodelled disturbance, and measurement sensor error also give an important rule to the system uncertainties. In this paper, Robust PID controller with AMIGO method is designed to handle system uncertainties in guarantee robust stability and closed-loop system performances of CPS-based batch distillation column.

II. CPS-Based Distillation Column

Spatially distributed components as in network-based control systems has been widely used not only in industrial process, but also in chemical process, aircraft, automobiles, refineries, healthcare, and power plant. Distillation is one of chemical process for components separation, purification, and waste product disposal by utilizing the boiling points of each chemical component in a product. Distillation is classified into 2 categories, they are batch and continous distillation. In batch distillation, the material is loaded into boiler just for once distillation process at the beginning. On the contrary, in continous distillation, the material is fed into feeding tray, so the distillation process occurs continously [7]. In this paper, mini batch distillation column of Honeywell Laboratory ITB is analyzed in the term of CPS-based industrial process. The distillation column and its parts is shown in Fig. 2.

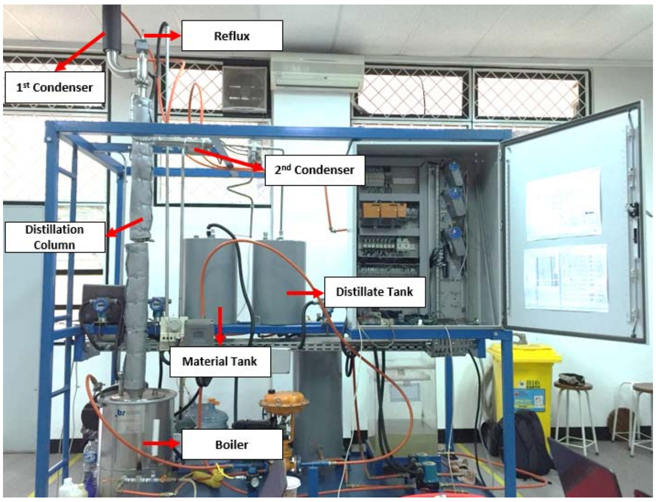

Fig. 2 Mini Batch Distillation Column

The distillation process of binary mixtures between ethanol and water occurs in a packed column distillation plant. The plant is about 3 meters tall and it has a heater pot with 22 liter capacity. It comprises two columns in series with 21 mm diameter and 90 mm tall each. A reflux valve is located at the top part of the column and accessible for on-off control. Among other installed sensors for monitoring and feedback purposes, temperature sensors in the heater pod, top column, and cooling water are used to monitor [7].

Recently, the distillation column in Honeywell Laboratory ITB has been developed for CPS-based automation industrial process purposes. There are 3 main components of CPS implementation, namely physical part (sensors, controllers, and actuators), communication network part, and cyber part as shown in Fig. 3. The MQTT protocol is used as a communication network, whose integrate physical system and cyber system. There are 2 control loop process in the distillation column, namely boiler process and column process. The variable measured from these processes are boiler temperature and distilate concentration, respectively. Sensor PT100 is used to measure temperature, while smart concentration sensor from previous research is used to measure distillate concentration [8].

The process offered by this plant is the MIMO case, but for the sake of simplicity, this paper only analyzed the SISO case of column process. The actuator for this process is a reflux valve that controls distillate flow into distillate tank. Both sensor and actuator use smart transmitter from previous research [8] to communicate in term of sending and receiving data through communication network.

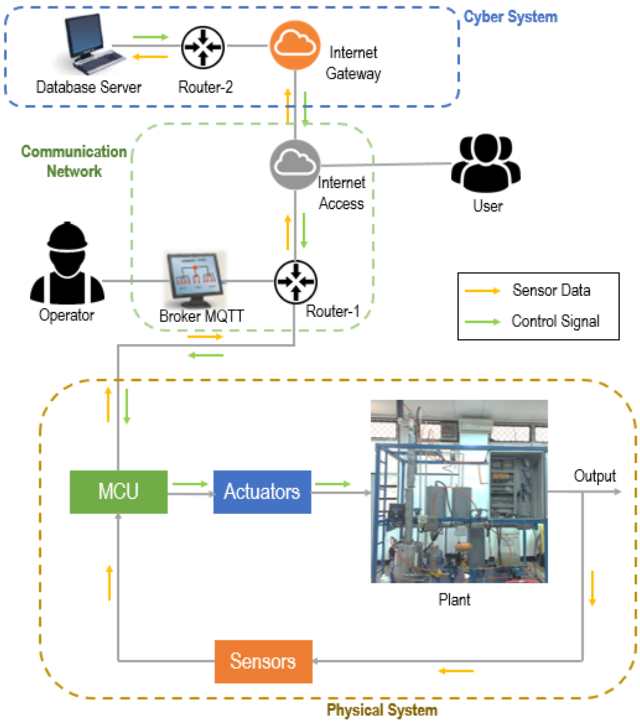

Fig. 3 Block Diagram of CPS-based Batch Distillation Column

III. Modelling of Distillation Column and System Uncertainties

The modelling of distillation column system is approximated with FOPDT (First Order Plus Dead Time) model based on its characteristic as shown in (1). In this FOPDT model, K, T and L denote gain constant, time constant, and time delay, respectively. The data of distillate concentration with various PWM value of reflux valve is collected to identify the system characteristic. In modelling process, PWM reflux value is set as the input control to the system. This value indicates the percentage of how long the reflux will be open and close in a period of time for a batch of distillation process. The identification of collected data is done by black-box method using System and Identification toolbox in Matlab. In this process, the percentage of PWM reflux and distillate concentration is set as the input and output data, respectively.

$$G(s) = \frac{K}{Ts+1} e^{-Ls} \quad \cdots (1)$$

The FOPDT model obtained for the distillation column is shown in (2). This transfer function with 70% fits to validation data is set as a nominal transfer function in controller design process. The nominal transfer function contains ±30% of error modelling uncertainty from linearized system approximation. The delay component in transfer function consists of time-varying process delay and communication delay. The communication delay occurs due to data transmission from sensor-to-controller, and from controller-to-actuator. The computation delay is assumed to be included in communication delay. The uncertainty due to unmodelled system dynamics also contained in both of the delay component. Several experiments of distillation process conducted by authors show the delay value and its uncertainty as in Table I.

$$G_0(s) = \frac{35{,}289}{211{,}6069s + 1} e^{-240{,}6967s} \quad \cdots (2)$$

TABLE I. DELAY VALUE OF DISTILLATION PROCESS

PWM Reflux (%)	Process Delay (s)	Communication Delay (s)
70	174	0.45
	146	0.31
	181	0.33
80	180	0.38
	169	0.40
	155	0.43
90	213	0.38
	170	0.41
	338	0.31
Range Uncertainty	$146 \leq \theta_1 \leq 338$	$0.31 \leq \theta_2 \leq 0.45$

The dynamic uncertainty on both input and output channel is also known as actuator and sensor failure, respectively. The difference resolution between controller and actuator induce uncertainty in the term of computed control signal. This actuator failure is also known as disturbance input to the system that occurs in low frequency region. In this research, control signal is computed in database server computer with resolution 1001 bit. The computed control signal is sent to actuator through communication network with resolution 10 bit. Rounding method of control signal value to the nearest place is taken into account to compensate the difference resolution. This method induce input uncertainty of about ±0.5%.

Measurement error in output channel mainly caused by uncallibrated sensor for certain time. The difference between sensor measurement and its actual value is also known as noise to the system that occurs in high frequency region. In this research, smart concentration sensor is used to measure distillate concentration. This sensor transmit the measured data to the controller through communication network. Several experiments are obtained by authors to observe how deviate the sensor value from the callibrated one. This results in output uncertainty of about ±2%. The structured and unstructured system uncertainty is modelled by multiplicative uncertainty model as shown in Fig. 4.

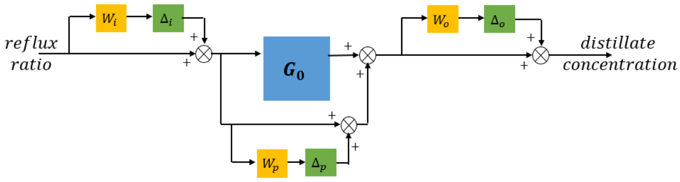

Fig. 4 Block Diagram of Uncertainty Model

The utilized multiplicative model is described by (3), where $G(s)$ represents an uncertain (perturbed) model, $G_0(s)$ is a nominal model, $W_M(s)$ is stable weighting function of dynamic uncertainty, and $\Delta_M(s)$ is the uncertainty itself. In this paper, the stable weighting function consists of input, parametric and output weighting function which denotes by $W_i(s)$, $W_p(s)$ and $W_o(s)$, respectively.

$$G(s) = G_0(s)[1 + W_M(s)\Delta_M(s)] \quad \cdots (3)$$

Each weighting function corresponds to input $\Delta_i(s)$, parametric $\Delta_p(s)$, and output $\Delta_o(s)$ uncertainty. These uncertainties can be represented by stable function fulfilling the inequality in (4).

$$\|\Delta_M(s)\|_\infty \leq 1 \rightarrow |\Delta_M(j\omega)| \leq 1 \quad \forall\omega \quad \cdots (4)$$

The choice of suitable weighting function in modelling system uncertainty is the important part of process design. In [16], the weighting function must fulfill inequality (5), where the left-hand side represents normalized perturbation.

$$\left|\frac{G(j\omega)}{G_0(j\omega)} - 1\right| \leq |W_M(j\omega)| \ \forall\omega \quad \cdots (5)$$

The chosen weighting function should cover the upper side of normalized perturbation, or even the worst possible case of system uncertainty in nominal model. Several approaches of weighting function computation is used in this paper. For parametric uncertainty, as stated in [9], the rule of choosing its weighting function is determined by equation in (6). Variables with 'min' and 'max' in its subscription denote the minimal and maximal possible value of uncertain parameter in nominal model, respectively. While '0'-subscription denotes the variable of system nominal model.

$$W_p(s) = \frac{K_{max}}{K_0} \cdot \frac{T_0 s + 1}{T_{min} s + 1} \cdot \frac{\tau s + 1}{-\tau s + 1} - 1; \ \ \tau = \frac{\theta_{max} - \theta_{min}}{4} \quad \cdots (6)$$

For unmodelled dynamics uncertainty, such as input and output channel uncertainty, a simple multiplicative weight in (7) is usually used to describe its unknown dynamics as stated in [10]. Variable r_0 denotes the relative uncertainty at steady-state, $1/\tau$ is approximately the frequency at which the relative uncertainty reaches 100%, and r_∞ is the magnitude of the weight at high frequency.

$$W_i(s) = \frac{\tau s + r_0}{(\tau/r_\infty)s + 1} \quad \cdots (7)$$

In SISO systems, it doesn't matter about the perturbation in input or output channel, since it fulfill the rule in (8).

$$G(1 + W_i\Delta_i) = (1 + W_o\Delta_o)G \ \ with \ \Delta_i(s) = \Delta_o(s) \ and \ W_i(s) = W_o(s) \quad \cdots (8)$$

IV. ROBUST PID CONTROL DESIGN

PID controllers have been widely used in industries since 1940s and remain the most often implemented controller today. A convenient feature of the PID controller is its compatibility with enhancement that provides easy constant tuning and implementation in many application areas. The algorithm of PID controller is simple, single equation but it can provide good control performance for many different processes. This flexibility is achieved through several adjustable parameters whose values can be selected to modify the behaviour of the closed loop system. The good closed-loop system performance can be achieved by a proper choice of tuning constant values, but poor performance and even instability can result from a poor determination of PID parameter.

Ziegler-Nichols (Z-N) tuning rules is one of the most popular method that commonly used in PID controller tuning. Recently, some different and new approaches of tuning rules have been developed. Some of the developed methods mainly concerned only on the system performance, such minimizing integrated error criteria as developed by Murrill et al. (1967) [11] or Rovira et al. (1969) [12], or the more recent work by Awouda and Mamat (2010) [13]. There are also tuning methods concidered the unstable FOPDT process, ranging from relatively simple analytic tuning formula [14], to more complex techniques algorithm [15]. A drawback of those tuning rules is that such rules do not consider load disturbance, model uncertainty, and measurement noise, since tuning for high performance is always accompanied by low robustness. Nevertheless, there are other developed tuning rules that mainly concerned the robustness, such as AMIGO (Approximate M-constraint Integral Gain Optimization) developed by Åström dan Hägglund [16, 17], or those developed by et al. [18].

As shown in the previous part, the use of network will cause latency which produces significant variation in the time required to transmit data from sensor-to-controller, and controller-to-actuator. Furthermore, the nonlinearity characteristic of distillation column also cause variation in process delay time that inherent in distillation process. The variation in response time is called time delay uncertainty. In control system designs, time delay has potential to destabilize the closed loop control systems. It is also difficult to control a system which has dominant time delay, i.e. when the ratio of the time delay and the time constant is bigger than one. Beside as a source of instability, time delay can also lead to poor system performances.

Error modelling caused by linearized approximation of system may lead to differences between actual model and system model which was used to design the controller. The imperfections of model used is known as parameter uncertainty. If parameter uncertainty is not taken into account in designing the controller, it will lead to different system performance between design process and implementation. Furthermore, the error sensor measurement and disturbance input can also affect the system performance. Hence, this paper proposed PID control design method which mainly concerned on system robustness. The AMIGO tunning rules is used in this research to compute PID gain controller. That is, an alternative method to optimize the setting of a PID controller in order to achieve robust stability and performance of closed-loop system.

The suggested AMIGO tuning rules for PID controller of FOPDT model as mentioned in [17] are,

$$\begin{aligned} K_p &= \frac{1}{K}(0.2 + 0.45\frac{T}{L}) \\ T_i &= \frac{0.4L + 0.8T}{L + 0.1T}L \qquad \cdots (9) \\ T_d &= \frac{0.5LT}{0.3L + T} \end{aligned}$$

Meanwhile, AMIGO tuning rules for PI controller as shown in [19] are,

$$\begin{aligned} K_p &= \frac{0.15}{K} + \left(0.35 - \frac{LT}{(L+T)^2}\right)\frac{T}{KL} \\ T_i &= 0.35L + \frac{13LT^2}{T^2 + 12LT + 7L^2} \end{aligned} \qquad \cdots (10)$$

The determinant of choosing PI or PID controller is depend on performance characteristic in design process [19]. If the Derivative-term (D) is present in controller, the high frequency noise is highly amplified, hence the closed-loop system will sensitive to noise measurement. On the other hand, if the Integral-term (I) is present, a windup behavior can occurs in the system. The condition occurs when a large following error is present in the system. Therefore, robust PI and robust PID is designed to figure out which one gives better performance to the system. The controller constants obtained from (9) and (10) are,

Robust PI controller

$$K_p = \frac{0.15}{K} + \left(0.35 - \frac{LT}{(L+T)^2}\right)\frac{T}{KL} = 0{,}006768$$

$$T_i = 0.35L + \frac{13LT^2}{T^2 + 12LT + 7L^2} = 216{,}2352$$

Robust PID controller

$$K_p = \frac{1}{K}\left(0.2 + 0.45\frac{T}{L}\right) = 0{,}016878$$

$$T_i = \frac{0.4L + 0.8T}{L + 0.1T}L = 244{,}104$$

$$T_d = \frac{0.5LT}{0.3L + T} = 89{,}72908$$

In this research, robust stability and closed-loop system performance of both controller above is analyzed through simulation and implementation as shown later. The closed-loop system is robustly stable if and only if [10, 21] :

$$\|W_M(s)T_0(s)\|_\infty < 1 \quad \longrightarrow \quad |T_0(j\omega)| < \frac{1}{|W_M(j\omega)|} \quad \forall\omega \qquad \cdots (12)$$

where $T_0(s)$ represents a complementary sensitivity function defined by :

$$T_0(s) = \frac{L_0(s)}{1 + L_0(s)} \qquad \cdots (13)$$

and $L_0(s)$ is the open-loop frequency transfer function, defined by :

$$L_0(s) = C(s)G_0(s) \qquad \cdots (14)$$

V. SIMULATION

Matlab Simulation is used to analyzed does robust stability and closed-loop system performance achieved, before it implemented in CPS-based batch distillation column. In this research, system uncertainties are modelled by multiplicative uncertainty as shown in Fig. 4. Weighting function for input/output chanel obtained from eq. (7) and (8) is,

$$W_O(s) = W_I(s) = \frac{10s + 0{,}005}{500s + 1} \quad \cdots (15)$$

Meanwhile, the weighting function for parameter uncertainty as shown in eq.(6) is,

$$W_p(s) = \frac{-2.66e05s^2 - 3105s - 0.5386}{6.137e04s^2 - 1230s - 1} \quad \cdots (16)$$

The parametric weighting function is plotted in bode diagram in common with frequency response of perturbed model. This plot is important to show is the weighting function convenient with perturbed model or not as shown in Fig. 5. Basically, the appropriate weighting function has to fulfill inequality (4), and cover from the upper side the normalized perturbation of even the worst possible case of uncertainty in the model (2).

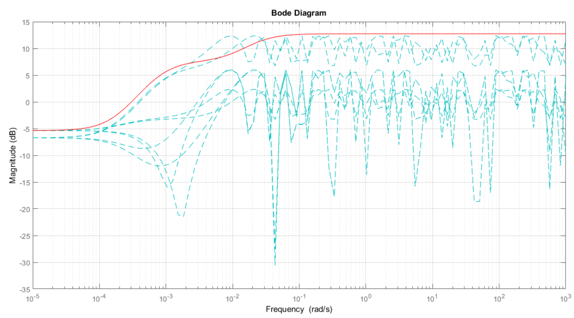

Fig. 5 Parameteric Weighting Function

From fig. 5, it is obvious that $W_p(s)$ (16) does not cover the normalized perturbations in the upper low frequencies, so it is inapplicable and would have to be modified. The frequency response of modified parametric weighting function $W_p(s)$ is shown in Fig. 6, and its mathematical representation in (17).

$$W_p(s) = \frac{-4.257e05s^2 - 4968s - 0.8618}{6.137e04s^2 - 1230s - 1} \quad \cdots (17)$$

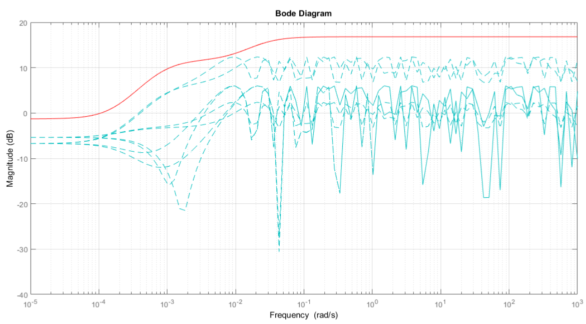

Fig. 6 Modified Parametric Weighting Function

The step response of robust PI, robust PID, and conventional PID control is shown in Fig. 7.

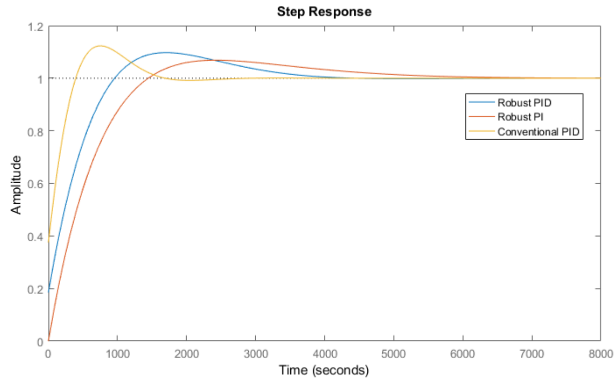

Fig. 7 Step Response of Designed Controller

The robust stability of designed controller is shown in Fig. 8. It can be seen that the complementary sensitivity function is fulfill requirement (12), hence all the designed controller are robustly stable. The performance of each controller is summarized in Table II.

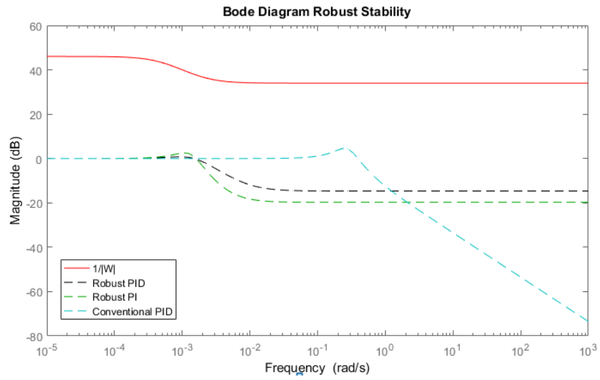

Fig. 8 Robust Stability of Designed Controller

TABLE II. PERFORMANCE COMPARISONS OF THE DESIGNED CONTROLLER

Controller	%OS	Rise Time (s)	Settling Time (s)	Integral Error				Robust Stability	Robust Performance
				IAE	ISE	ITAE	ITSE		
Robust PI	6,84	999,894	4,61e+03	7,0337	2,7483	86,2287	17,4214	YES	YES
Robust PID	9,73	703,876	3,47e+03	7,6476	2,6118	126,6629	16,0519	YES	YES
Conv. PID	12,32	301,340	1,53e+03	9,0454	3,4227	146,9273	25,0743	YES	NO

VI. IMPLEMENTATION ROBUST PID IN DISTILLATION COLUMN

The designed robust PI and PID controller above is implemented in CPS-based distillation column to investigate its system performance. In this research, the distillate reference concentration is 90%. The resulting implementation of both controller is shown in Fig.9. Furthermore, the system performance (integral error and distillate mixture ratio) of both controller is shown in Table III.

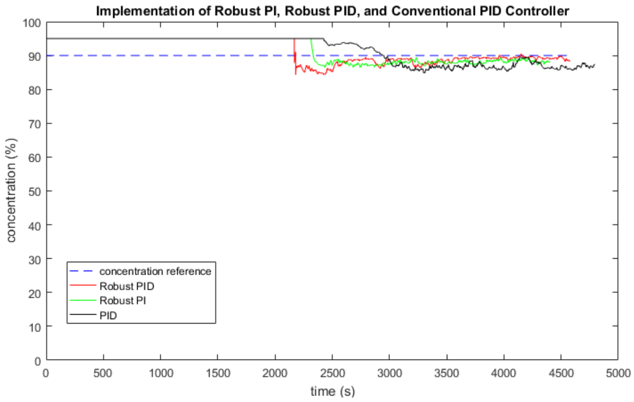

Fig. 9 Implementation of Robust PI, Robust PID and Conv. PID Controller

TABLE III. SYSTEM PERFORMANCES OF IMPLEMENTED CONTROLLER

Controller	Integral Error				Material	Distillate	Distillate Mixture Ratio
	IAE	ISE	ITAE	ITSE			
Robust PI	4,29e+03	9,56e+03	4,15e+06	8,51e+06	30% 3L	88% 0,663L	0,648
Robust PID	6,93e+03	3,57e+04	8,43e+06	3,00e+07		88% 0,71L	0,694
Conv. PID	9,13e+03	5,32e+04	1,08e+07	4,08e+07		87% 0,59L	0,570

VII. CONCLUSION

In this paper, robust PI and robust PID controller for FOPDT model of CPS-based batch distillation column is designed using AMIGO Method with appropriate weighting function chosen. This method is mainly concerned on system robustness to system uncertainties. Error modelling, input/output channel, and delay system is taken into account in control design process as system uncertainties. Matlab simulation is used to validate the method before its implementation. Based on simulation, the proposed Robust PI and Robust PID controller give better system performance on integral error and DMR compared to conventional PID. The simulation result is appropriate to its implementation, where as Robust PI gives better performance on system's integral error, but gives lower DMR value compared to Robust PID controller. For future work, it is better to control another loop control of distillation column, hence become multi-loop control problem. Furthermore, it is also better to compare the designed controller performance with conventional robust control performance.

ACKNOWLEDGMENT

This work was supported by Honeywell Laboratory ITB.

REFERENCES

[1] R. Jacob, M. Zimmerling, P. Huang, J. Beutel, and L. Thiele, "End-to-end Realtime Guarantees in Wireless Cyber Physical Systems," in IEEE Realtime Systems Symposium, 2016.

[2] H. Ghailani, and S. Al Mesmari, "State-of-the Art and Trends in Cyber-Physical Systems (CPSs)," in International Journal of Liberal Arts and Social Science, Vol. 6 No. 5, 2018.

[3] J. Baillieul and P. J. Antsaklis, "Control and Communication Challenges in Networked Real-Time Systems," in Proceedings of the IEEE, Vol. 95 No. 1, 2007.

[4] Y. Wang, M. C. Vuran, and S. Goddard, "Cyber-physical Systems in Industrial Process Control,"

[5] E. Joelianto, "Networked Control Systems : Time Delays and Robust Control Design Issues," in 2nd International Conference on Instrumentation Control and Automation, 2011.

[6] M. S. Mahmoud and A. Ismail, "Role of Delays in Networked Control Systems," 2013.

[7] A. S. Rohman, P. H. Rusmin, R. Maulidda, et al, "Modelling of the Mini Batch Distillation Column," in International Journal on Electrical Engineering and Informatics Vol. 10 No. 2, 2018.

[8] I. Budiawan, "Cyber Physical System-Based Automation : Studi Kasus Kendali PD Berbasis Kejadian pada Kolom Distilasi Mini," in Magister Thesis, 2018.

[9] Laughlin DL, Morari M., "Graphical Stability Analysis for Control Systems with Model Parameter Uncertainties," Computer Cision, Graphics, and Image Processing. 1989; 47(1): 59-76.

[10] Skogestad S, Postlethwaite I., "Multivariable Feedback Control: Analysis and Design," Chichester, UK: John Wiley and Sons; 2005.

[11] Murill, P.W., "Automatic Control of Process," International Textbook, Co: Scranton, PA, USA, 1967.

[12] Rovira A.A, Murril, P.W, Smith C.L, "Tuning Controllers for Stepoint Changes," Instrum. Control Syste. 1969.

[13] Awouda, A.E.A; Bin Mamat, R. "Refine PID Tuning Rule using ITAE Criteria," in International Conference on Computer and Automation Engineering, 2010.

[14] Cho, W.; Lee, J.; Edgar, T.F. "Simple Analytic Proportional-integral-derivative (PID) Controller Tuning Rules for Unstable Process," 2013.

[15] Rajinikanth, V.; Latha, K. "Tuning and Retuning of PID Controller for Unstable Systems using Evolutionary Algorithm," 2012.

[16] Hagglund, T.; Astrom, K.J. "Revisiting the Ziegler-Nichols Tuning Rules for PI Control," 2002.

[17] Astrom, K.J.; Hagglund, T. "Revisiting the Ziegler-Nichols Step Response Method for PID Control," 2004.

[18] Ho, M-T.; Datta, A; Bhattacharyya, S.P. "Robust and non-fragile PID Controller Design," 2001.

[19] M. L. Ruz, J. Garrido, F. Vazquez, and F. Morilla. "Interactive Tuning Tool of Proportional-Integral Controllers for First Order Plus Time Delay Process," 2018.