Optimization of control performance on CO_2 removal in subang field using model predictive control

Abdul Wahid[1,*], and *Yoga* Wiranoto[1]

[1]Chemical Engineering Department, Universitas Indonesia, Depok, Indonesia

Abstract. A model predictive control (MPC) is used to optimize the control performance on CO_2 removal in Subang Field. MPC is implemented to control the feed gas pressure (PIC-1101), amine flow rate (FIC-1102), and makeup water flowrate (FIC-1103) to maintain CO_2 concentration in sweet gas. MPC is built using the first-order plus dead time (FOPDT) models. The control performance tests are used set point (SP) tracking and disturbance rejection with the performance indicator is the integral of square error (ISE). The result show that the optimum setting of prediction horizon (P), horizon (M) and Time Sampling (T) in MPC are 9 1, 32 and 1 on PIC-1101; 34, 10 and 5 on FIC-1102 and 40, 10 and 5 on FIC-1103. Based on ISE values, the use of MPC can improve performance for set point tracking by 14.02% in PIC-1101, 76.74% in FIC-1102, and 16.31% in FIC-1103, the use of MPC can improve performance for disturbance rejection by 19.32% in FIC-1102, and 91.57% in FIC-1103, compared with the proportional-integral (PI) controller that used in the field.

1 Introduction

Indonesia as the largest country in Southeast Asia with econoomic growth per capita was recorded at 5.0% in 2016, up to 4.9% from the year earlier [1]. The increasing number of econoomic growth leads to an increase in the need for energy, the importance of energy mix in meeting those needs is very important one of which is the use of natural gas. Natural gas is widely used as fuel for power generation activities, both for household and office and industrial needs, besides the use of natural gas can also be used as raw material for the petrochemical industry [2].

Natural gas coming out from wells have the main content of CH_4 (Methane), C_2H_6 (Ethane), C_3H_8 (Propane), C_4H_{10} (Butane), C_5H_{12} (Pentane), CO_2 (Carbon dioxide), H_2S (Hydrogen Sulfide), and N_2 (Nitrogen) as well as small amounts of H_2 (Hydrogen), He (Helium), O_2 (Oxygen), and C_6H_{14} (Hexane and other heavy hydrocarbons) (ISO 6974-1: 2000 - Natural gas). Sales gas have a requirement for a maximum of 5% -vol CO_2 and 8 ppmv H_2S to increase the calorific value and consider to the environment [3].

To fulfill the process is required process of removal of acid gas in order to reach the desired gas specification. The principle of the removal process is by absorbing the acid gas, but there is often a problem of the loss of a number of hydrocarbons in the process of the release of acid gas as an example of CO_2 removal unit at "X" Company in Subang Field. To overcome these problems, optimization of operating conditions and the addition of equipment required to achieve these operating conditions and the use of advance control.

Recently studies that discussed the control of the CO_2 removal unit is very limited to certain controlled variables (CV). The configuration used in the control is a single control loop with the PI (Propotional - Integral) method, which consists of one transmitter, one controller, and a final control element. The goal is only to control the level on the absorber column [4], while controlling for pressure variable, previous study used single control loop with PI method, then for final control element controlled with control valve at output of KO drum sweet gas with the intention that pressure on absorber column can be maintained at optimum condition [5], whereas another study discuss about temperature control to maintain amine circulation temperature [6] which is not available in field because it is set by manual controlling. Therefore, in this research, the research is conducted to see the whole factory control system supported by the use of predictive control model (MPC) method to get control parameters and improve process performance in terms of control.

2 CO_2 Removal Unit

2.1 Process Control Target

General process flow for CO_2 Gas Removal Plant at "X" in Subang Field is shown in the scheme of the process, can be seen as in Fig.1. The image shows from the absorption process to the regeneration part. The scheme is slightly different from typical acid gas removal in general [7], in the regeneration section where there are only LP (Low Pressure) Flash columns with tray and packed configurations therein. Methyl diethanolamine solution with piperazine acceleration agent is used in this process due to its flexibility and availability, to maintain its strength (amine strength) in the absorption process, it is

* Corresponding author: wahid@che.ui.ac.id

necessary to add water or makeup water in absorber column, the location of the inlet flow is also designed above the inlet solution amine or in the upper tray in order to capture the volatile amine solution to be excluded from the sweet gas stream. The main tools of concern are the absorber columns and LP Flash columns as well as associated pipelines, heat exchangers, and knock out drums.

2.2 Control Target

In this project, limitation of the variables that need to be maintained are the amine flow rate, the makeup water flow rate, and the pressure of feed gas to the absorber column. The control is intended to maintain the quality of the final product i.e the CO_2 content in the sweet gas, in addition some variables also need to be controlled for reasons of process safety [8].

Control system in CO_2 Removal Plant that used in this research shown in Fig 2 and tag number from each controllers shown in Table 1 .The PIC-1101 controller is the process controller for controlling the pressure difference between the bottom and top of the absorber column and maintaining sweet gas pressure. The pressure difference is maintained at about 3 psig.

The FIC - 1102 controller is the process controller for controlling the level in absorber and LP flash columns and amine strength, but in this study more focused on controlling amine strength. The amine solution used is a mixture of aMDEA and piperazine with water. Due to the amount of water that evaporates during the regeneration process, it is necessary to makeup water to keep the amine strength remains in the range of 45% [9]. It takes 2 – 4 m^3/h makeup water to keep it up.

The FIC - 1103 controller is the process controller for controlling amine strength. The amine solution used is a mixture of aMDEA and piperazine with water. Due to its effect on the CO_2 removal process, the flow rate should be maintained at 650 – 700 m^3/h to maintain a 45% amine strength concentration.

The controller currently used in the plant uses a P-I (Propotional - Integral) control mechanism. Then to compare it is done by using MPC (Model Predictive Control) in SISO system (Single Input Single Output).

Table 1. Control system on CO_2 Removal Unit

Type of Controller	Controlled Variable	Manipulated Variable
Pressure Control (PIC – 1101)	Feed gas to the absorber column	Sweet gas flow rate of KO drum output
Flow Control (FIC – 1102)	Makeup water flow rate to absorber column	Makeup water flow rate from utility
Flow Control (FIC – 1103)	Amine flow rate to the absorber column	Amine flow rate from amine pump

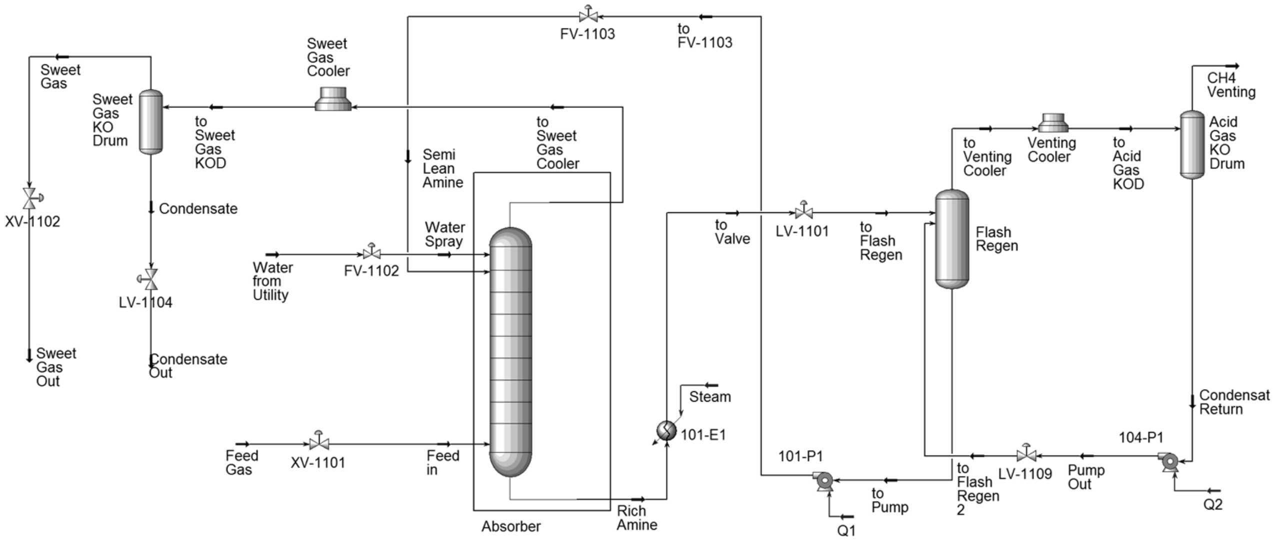

Fig. 1. CO_2 Removal Process Scheme

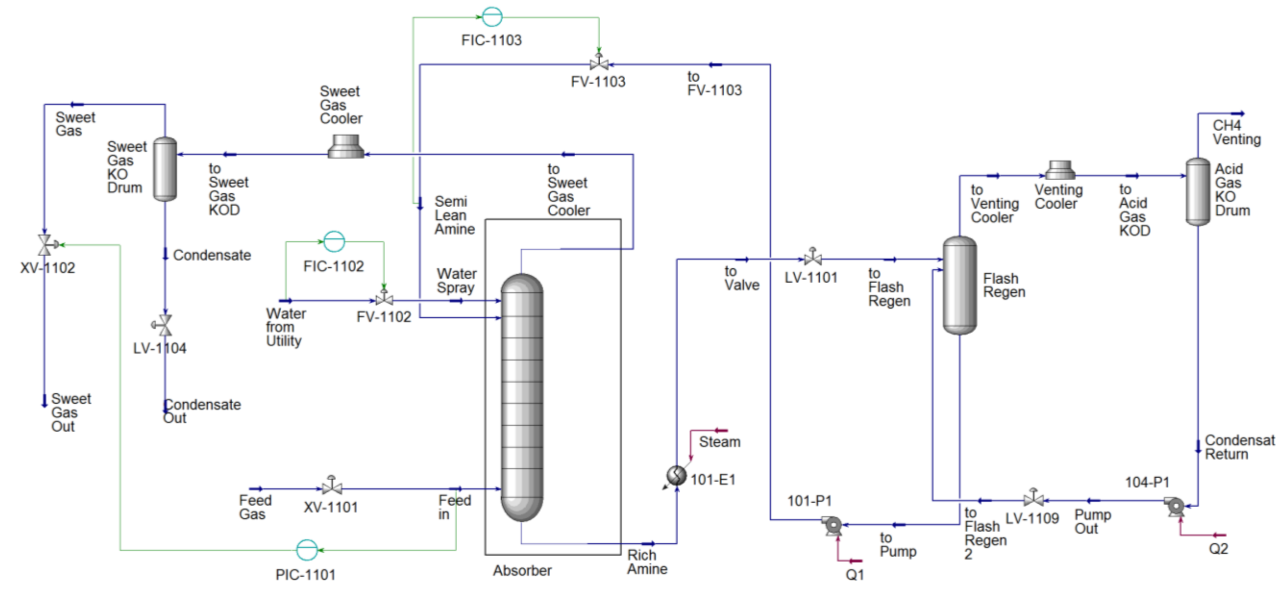

Fig. 2. Control System in CO_2 Removal Plant

3 Empirical Model of Process Control

The model used in obtaining the parameters of the empirical model is used Fisrt Order Plus Dead Time (FOPDT) approach in this research [10]. Eq. (1) shows FOPDT model that has dead time (θ), this number shows how much time the system needs to respond to step change. The equation also shows that the speed of the system in response to the interference (τ) . While gain (Kp) shows the amount of system sensitivity.

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

The result of FOPDT will be obtained a Process Reaction Curve (PRC) through model testing, empirical models were developed as shown in Table 2.

Table 2. FOPDT Model of CO_2 Removal Unit

Controller	Model
PIC – 1101 Controller	$\frac{-0.199e^{-1.5s}}{31.5s+1}$
FIC – 1102 Controller	$\frac{0.044e^{-0.543s}}{2.084s+1}$
FIC – 1103 Controller	$\frac{3.330e^{-3.106s}}{0.528s+1}$

4 Controller Tuning

As resulted empirical model, we have to find controller parameters with some tuning procedures [10]. Based on empirical model from FOPDT, we can calculate the control parameters of PI controller by using Eq. (2) and Eq. (3). The controllers is tuned using Ziegler-Nichols method [8]. However, the control parameters of PI controller from actual is different than using Eq. (2) and Eq. (3) to get that paramters, so this step could be neglected.

$$K_c = \frac{0.9}{K_p}\left(\frac{\theta}{\tau}\right)^{-1} \tag{2}$$

$$T_I = 3.33\,\theta \tag{3}$$

The modeling results from the previous step then tuned to get the control parameters. As in SISO system, the tuning of MPC Controller is based upon [11]. Approximate the process dynamics of all controller output to measured process variable pairs with FOPDT models, select the sample time (T) as close as possible with Eq. (4), compute the prediction horizon (P) with Eq. (5), and compute a control horizon (M) with Eq. (6).

$$\begin{Bmatrix} T_{rs} = Max(0{,}1\,\tau_{rs}\,;0{,}5\theta_{rs}) \\ T = Min(T_{rs}) \end{Bmatrix} \tag{4}$$

$$P = Max\left(\frac{5\tau_{rs}}{T} + k_{rs}\right) \text{ where } k_{rs} = \left(\frac{\theta_{rs}}{T} + 1\right) \tag{5}$$

$$M = Max\left(\frac{\tau_{rs}}{T} + k_{rs}\right) \tag{6}$$

Tuned values of MPC and control parameter from actual i.e PI Controllers are shown in Table 3.

Table 3. Tuned values of CO_2 Removal Controller

Parameter	MPC Fine Tuning	MPC Dougherty & Cooper	Proportional-Integral
PIC – 1101			
P	91	54	-
M	32	12	-
T	1	3	-
Kc	-	-	7.2
Ti	-	-	46.1
FIC – 1102			
P	34	51	-
M	10	11	-
T	5	5	-
Kc	-	-	0.5
Ti	-	-	10
FIC - 1103			
P	40	51	-
M	10	11	-
T	5	5	-
Kc	-	-	0.37
Ti	-	-	10.6

5 Performance of Controller

Set point tracking and disturbance rejection are the way to test controller performance, reliable controller must be aggressive by set point changing and be able the regulator from disturbances [12].

5.1 Set Point Tracking

The first comparison is between Propotional – Integral, MPC – Dougherty & Cooper, MPC – Fine Tuning for set point (SP) tracking. Fig. 3 (a) shows PIC – 1101 performance from SP changing about 2 psig, controlling using the resultant constants from tuning using the MPC-fine tuning and proportional-integral method provides a fast response when compared to the constants of the MPC-Dougherty & Cooper tuning constants. The MPC-fine tuning method responds more slowly than proportional integral but the MPC-fine tuning method has an advantage over the proportional-integral in the process to achieve a more stability.

Fig. 3 (b) presents FIC – 1102 performance from set point change about 0.5 m^3/h, and Fig. 3 (c) FIC – 1103 performance from set point change about 30 m^3/h. We can see that controlling by using the resultant constants of the tuning using the MPC-fine tuning and Dougherty & Cooper method provides a fast response when compared to the proportional-integral. However, the MPC-fine tuning method has an advantage compared to MPC-Dougherty & Cooper in the process speed to achieve stability. So that the tuned constant using MPC-fine tuning method is better to overcome set point changes compared to the constants of propotional-integral and MPC-Dougherty & Cooper.

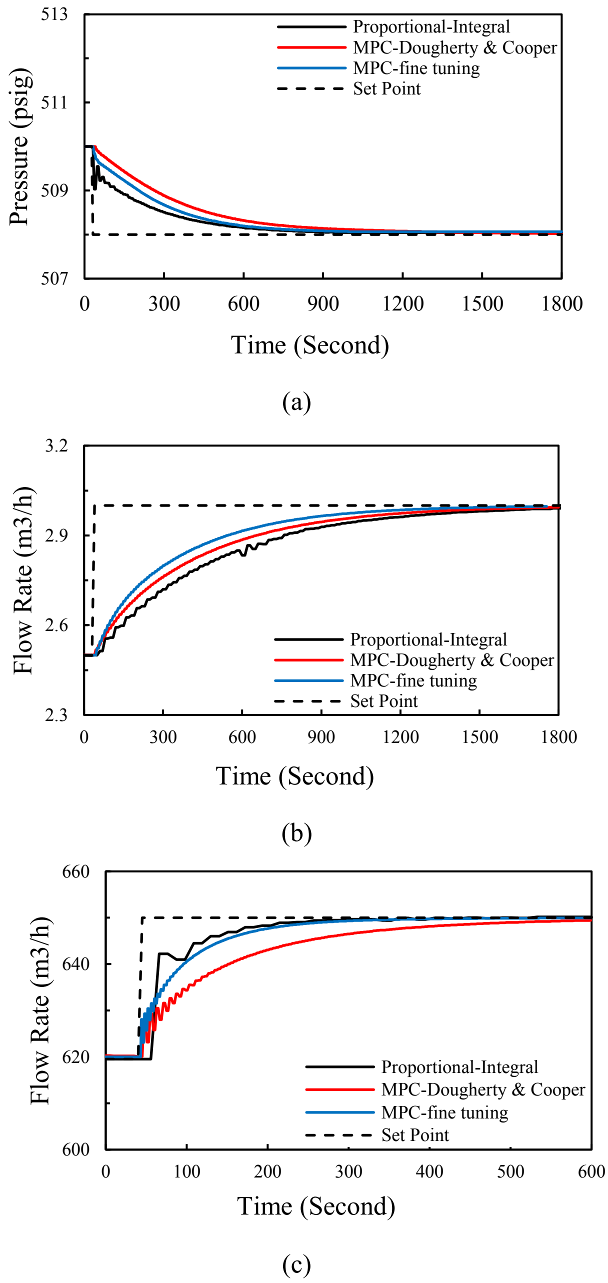

Fig. 3. Controllers response to set point changes: (a) PIC – 1101 (b) FIC – 1102 (c) FIC – 1103

ISE is the integral of the square error of controller. In the context of process control, the meaning of error is the difference between the value of the actual CV of the targeted CV value or so-called Set Point (SP) [13].

$$ISE = \int_0^{\infty} [SP(t) - CV(t)]^2 dt \quad (7)$$

https://doi.org/10.1051/e3sconf/20186701028

ISE is one of the parameters that shows the performance of control. The small value of error, which also means the smaller value of Eq (7), then this indicates a good controller performance. ISE values from three controllers above are summarized in Table 4.

Table 4. Performance of PI controller and MPC controller by ISE calculation for SP tracking

Controller	Proportional-Integral	MPC-fine tuning	MPC-Dougherty & Cooper
PIC – 1101 Controller	8.946	7.692	11.377
FIC – 1102 Controller	0.834	0.194	0.346
FIC – 1103 Controller	380	318	665

5.2 Disturbance Rejection

Controllers response to the effect of interference changes the flow rate of feed gas comparisons shown in Fig. 4. One controller to maintain pressure and two controlller to maintain flow rate. The disturbance to test reliable controller is feed gas flow rate decreasing about 5 MMSCD. Fig. 4 (a) shows PIC – 1101 to reject disturbance, the tuning method uses proportional-integral and MPC-fine tuning has the same ability to overcome the interference. Seen from the duration of the MPC-Dougherty & Cooper method of control response to be able to stabilize then this adjustment is no better than the setting of the previous two methods.

Fig. 4 (b) presents FIC – 1102 to reject disturbance, and Fig. 4 (c) shows FIC – 1103 to reject disturbance. From both images it is seen that MPC is better to overcome the interference because it is more stable to reach its original condition, using fine tuning method and Dougharty & Cooper method. While the use of PI controllers shows fluctuations to achieve stable conditions.

ISE values from three controllers above are summarized in Table 5.

Table 5. Performance of PI controller and MPC controller by ISE calculation

Controller	Proportional-Integral	MPC-fine tuning	MPC-Dougherty & Cooper
PIC – 1101 Controller	0.947	0.947	1.378
FIC – 1102 Controller	0.295	0.238	0.238
FIC – 1103 Controller	4.306	0.363	0.623

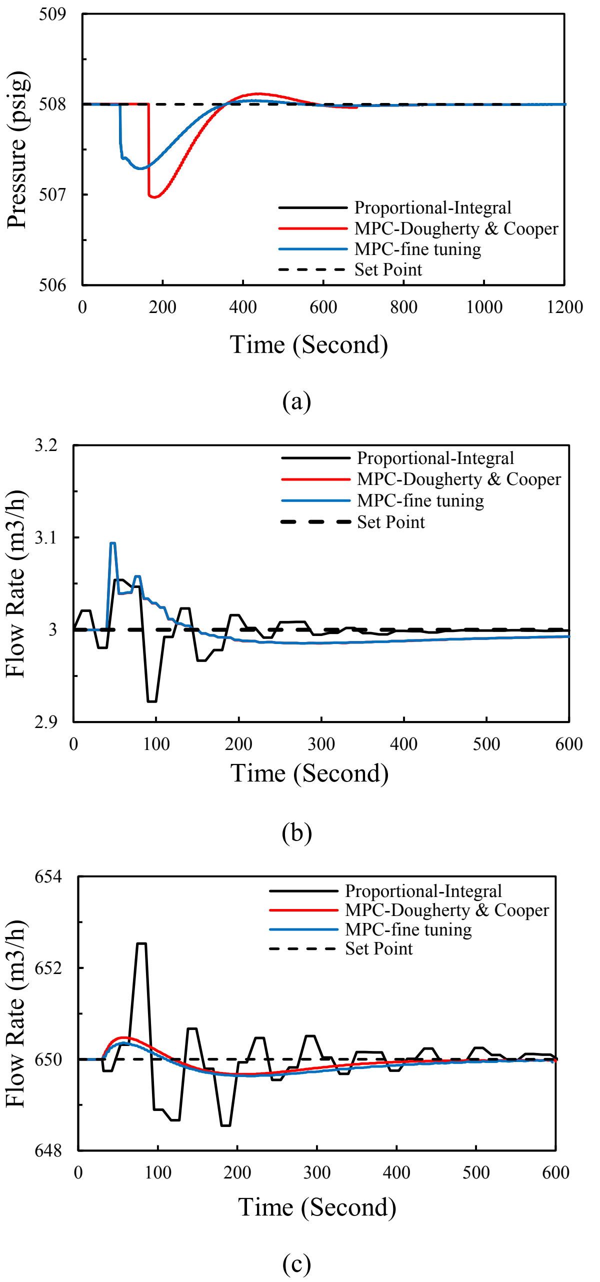

Fig. 4 Controllers response to the effect of interference changes the flow rate of feed gas: (a) PIC – 1101 (b) FIC – 1102 (c) FIC – 1103

6 Conclusion

Replacement of controller from Proportional-Integral to Model Predictive Control is very influential on the speed of process back to its stability. For that use of MPC can be an option in controlling the process in CO_2 Removal at"X" coompany in Subang field. The control response is based on the research that has been done by testing the reliability of the controller using the change of set point, the use of MPC can improve the control performance by 14.02% in PIC-1101 for control of absorber column pressure and keep pressure difference the top and bottom of the absorber column, 76.74% in FIC-1102 to maintain the flow rate of makeup water in order to maintain the desired strength of amine, and 16.31% in FIC-1103 to maintain the semi-lean amine circulation. While on disturbance by changing of feed gas flow rate, the use of MPC can improve the control performance by 19.32% in FIC-1102 to maintain the flow rate of makeup water in order to maintain the desired strength of amine, and 91.57% in FIC-1103 to maintain the semi-lean amine circulation against the effect of rate change disturbances gas feed flow.

We express our gratitude to the Universitas Indonesia which has funded this research through the scheme of Hibah Publikasi Internasional Terindeks untuk Tugas Akhir Mahasiswa (PITTA) No.2359/UN2.R3.1/HKP.05.00/2018.

References

1. Bank Indonesia, 2016 Economic Report on Indonesia (2017)
2. H. Cabalu, Indicators of security of natural gas supply in Asia. *Energy Policy*. **38**, 218-225 (2010)
3. M. Stewart, K. Arnold, *Gas sweetening and processing field manua*l (Gulf Professional Publishing, 2011)
4. M.A. Darmiyanto, "Analisis Kontrol Level Pada Absorber di CO_2 Removal Plant Subang" (Universitas Diponegoro, 2011)
5. R.D. Imam, "Analisis Pressure Control Pada Absorber di CO_2 Removal Field Subang" (Universitas Diponegoro, 2011)
6. L. Kurniawan, "Kontrol Temperatur Pada Rich Solution Heater di CO_2 Removal Plant Subang" (Universitas Diponegoro, 2011)
7. A.L. Kohl, R.B. Nielsen, R, *Gas purification* (Gulf Professional Publishing, 1997).
8. C.A. Smith, A.B. Corripio, Principles and practice of automatic process control. *Automatica*. **23**, 1, (1987)
9. R. S. Alvis, N. A. Hatcher, R. H. Weiland, CO 2 Removal from Syngas Using Piperazine - Activated MDEA and Potassium Dimethyl Glycinate. *International Conference & Exibition Nitrogen+ Syngas*, 20–23 (2012)
10. R. Kandulna, *Design of Pid Controller for FODPT and IPDT System* (Thesis, 2015)
11. D. Dougherty, D.J. Cooper, Tuning guidelines of a dynamic matrix controller for integrating (non-self-regulating) processes. *Industrial and Engineering Chemistry Research.* **42**, 1739-1752 (2003)
12. W. Tan, H.J. Marquez, T. Chen, Performance assessment of PID controllers. *Control and Intelligent Systems*. **32**, 158-166 (2004)
13. T. Kealy, A. O'Dwyer, Analytical ISE calculation and optimum control system design. *Irish Signals and Systems Conference*, (University of Limerick, Ireland 2003).
14. A. Ahmad, A. Wahid, Application of model predictive control (MPC) tuning strategy in multivariable control of distillation column. *Reaktor*, **11**, 66-70 (2007)
15. A. Wahid and A. Ahmad, Min-max controller output configuration to improve multi-model predictive control when dealing with disturbance rejection. *International Journal of Technology*, **6**, 3, 504-515. (2015)
16. A. Wahid and A. Ahmad, Improved multi-model predictive control to reject very large disturbances on a distillation column. *International Journal of Technology*, **6**, 962-971. (2016).
17. A. Wahid and A.P. Prasetyo, A Comparative study between MPC and PI controller to control vacuum distillation unit for producing LVGO, MVGO, and HVGO. IOP Conf. Ser.: Mater. Sci. Eng. **334**, 012020 (2018)
18. A. Wahid and I.G.E.P Putra, Multivariable model predictive control design of reactive distillation column for Dimethyl Ether production. IOP Conf. Ser.: Mater. Sci. Eng. **334**, 012018 (2018)
19. A. Wahid and F.F. Adicandra, Optimization control of LNG regasification plant using Model Predictive Control. IOP Conf. Ser.: Mater. Sci. Eng. **334**, 012022 (2018)