

# Optimization control of LNG regasification plant using Model Predictive Control

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**Abstract.** Optimization of liquified natural gas (LNG) regasification plant is important to minimize costs, especially operational costs. Therefore, it is important to choose optimum LNG regasification plant design and maintaining the optimum operating conditions through the implementation of model predictive control (MPC). Optimal tuning parameter for MPC such as P (prediction horizon), M (control of the horizon) and T (sampling time) are achieved by using fine-tuning method. The optimal criterion for design is the minimum amount of energy used and for control is integral of square error (ISE). As a result, the optimum design is scheme 2 which is developed by Devold with an energy savings of 40%. To maintain the optimum conditions, required MPC with P, M and T as follows: tank storage pressure: 90, 2, 1; product pressure: 95, 2, 1; temperature vaporizer: 65, 2, 2; and temperature heater: 35, 6, 5, with ISE value at set point tracking respectively 0.99, 1792.78, 34.89 and 7.54, or improvement of control performance respectively 4.6%, 63.5%, 3.1% and 58.2% compared to PI controller performance. The energy savings that MPC controllers can make when there is a disturbance in temperature rise 1°C of sea water is 0.02 MW.

## 1. Introduction

Several ways have been done to transport gas, one of which is by building gas pipes or liquefying natural gas into LNG, to reduce its volume and make the gas in liquid phase so that transportation is easier. Gas pipe is very limited utilization. Because the installation of gas pipelines has a high difficulty, as well as very expensive costs, especially for long distances and through the oceans. For LNG itself, in West Java has been built units FSRU (Floating Storage and Regasification Unit). This unit is useful for returning LNG back into gas phase to be utilized as industrial fuel.

LNG regasification plant should be in optimum design, both the design of the equipment and the control design in order to minimize the energy used. Therefore, in this research we will select the optimal LNG regasification plant design and set the control parameters used to maintain optimum operating conditions in case of process disturbances. The controller used is MPC because MPC is one of the advanced process control that implements the optimization method (cost function) in its control and has been proven to have better control performance compared to conventional controller [1].



## 2. Methodology Experimental

### 2.1 FOPDT Model

Figure 1 shows the research flow diagram. The research begins by creating a steady state model developed by Wahid and Ryan [2] and Devold [3]. Then for the selection of models based on energy use and optimization of operating conditions. After that, change the simulation to dynamic mode and install the control system on each equipment according to the control objectives. After that, the test model is performed on each controller. This model test is performed to obtain the PRC curve as shown in Figure 2 to obtain an empirical model.

Empirical model used is first order plus dead-time model (FOPDT) with equation:

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

with  $K_p = \Delta/\delta$  ( $\Delta$ = magnitude of change of output variable and  $\delta$  = amount of input variable change affecting output variable),  $\tau = 1,5 (t_{63\%} - t_{28\%})$  with  $t_{63\%}$  dan  $t_{28\%}$  respectively are output response times when it reaches 63% and 28% of the final value of the output variable, and  $\theta = t_{63\%} - \tau$  [4].

### 2.2 Tuning Controller

**2.2.1 Proportional-Integral Controller.** Based on the parameters of this FOPDT empirical model, the PI controller is adjusted using the Ziegler-Nichols method [5]. The transfer function PI is used equation:

$$G_c(s) = K_c \left( 1 + \frac{1}{T_i s} \right) \quad (2)$$

With  $K_c$  is the controlling gain and  $T_i$  is the integral time. For Ziegler-Nichols method used equation:

$$K_c = \frac{0.9}{K_p} \left( \frac{\theta}{\tau} \right)^{-1} \text{ dan } T_i = 3.33 \theta \quad (3)$$

**2.2.2 Model Predictive Control.** In MPC a dynamic model is used to predict the future output over the prediction horizon based on a set of control changes. The desired output is generated as a setpoint that may vary as a function of time; the prediction error is the difference between the setpoint trajectory and the model prediction. A model predictive controller is based on minimizing a quadratic objective function in Equation 4 over a specific time horizon based on the sum of the square of the prediction errors plus a penalty related to the square of the changes in the control variable(s) [6].

$$\sum_{i=1}^P w_i e^2(k+i) + \lambda \sum_{i=1}^m \Delta u^2(k+i-1) \quad (4)$$

where  $e(k+i)$  denotes the predicted error at time  $(k+i)$ ,  $i = 1, \dots, p$ ,

$$e(k+i) = r(k+i) - \hat{y}(k+i) \quad (5)$$

$r(k+i)$  is the reference value or setpoint at time  $k+i$ , and  $\Delta u(k)$  denotes the vector of current and future control moves over the next  $m$  sampling instants:

$$\Delta u(k) = [\Delta u(k), \Delta u(k+1), \dots, \Delta u(k+m-1)]^T \quad (6)$$

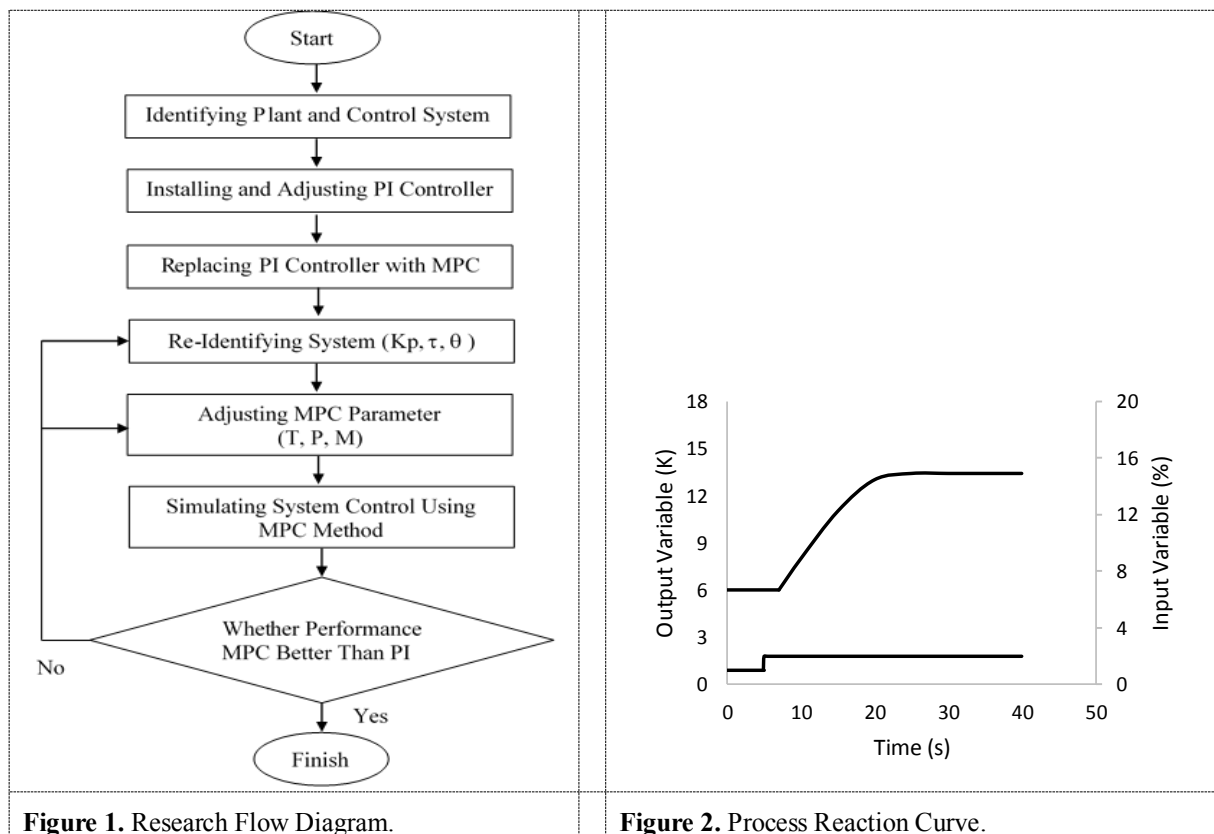
Equation 4 contains design parameters that can be used to tune the controller, that is, you vary the parameters (P and M) until the desired shape of the response that tracks the setpoint trajectory is achieved [7].

For the MPC controller, the empirical model parameters obtained from PRC can be directly used on UniSim software R390.1 as a process model. Then re-identified the system, by retesting the model with the close loop method using the default MPC adjustment constants and then the results are compared with the MPC performance by using the previous PRC. The re-identification process is completed if the MPC performance using the previous PRC is better than the performance of MPC with the PRC used. The performance parameter used is ISE. The system identification is performed to obtain the optimum

empirical model. After that done, tuning parameters MPC Prediction Horizon (P), Control Horizon (M), and Sampling Time (T) by using the fine-tuning method. MPC fine-tuning adjustment is done by trial error and then view system behavior to change adjustment parameters (P, M, T) to obtain stable settings.

### 2.3 Performance of Controller

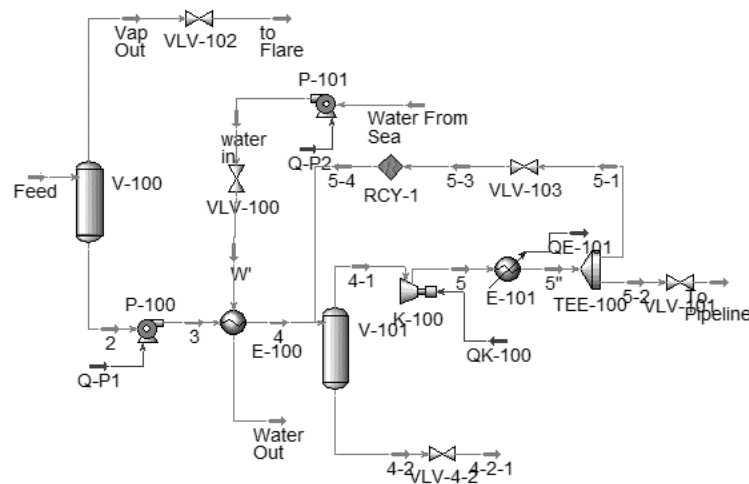
Based on these two controller tuning methods, the control performance was calculated using ISE (Integral Square Error) [8]. Setting methods that have smaller ISE values are the best tuning method.



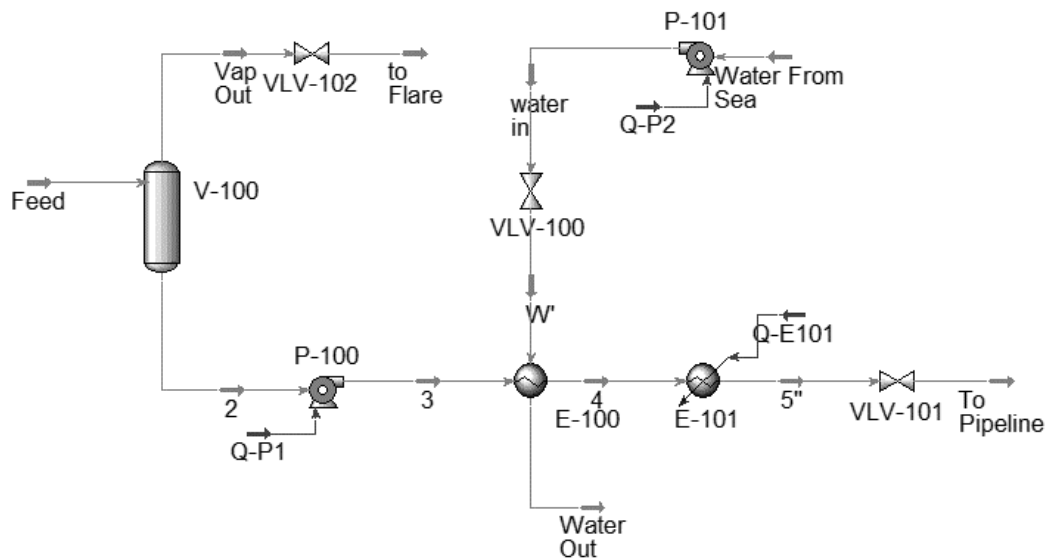
## 3. Result and Discussion

### 3.1 Optimization of LNG Regasification Scheme

To determine the most optimum regasification process scheme between the two schemes in figure 3 and figure 4, a simulation is performed that can calculate the energy requirements for each process scheme.



**Figure 3.** Scheme 1 Regasification Process.



**Figure 4.** Scheme 2 Regasification Process.

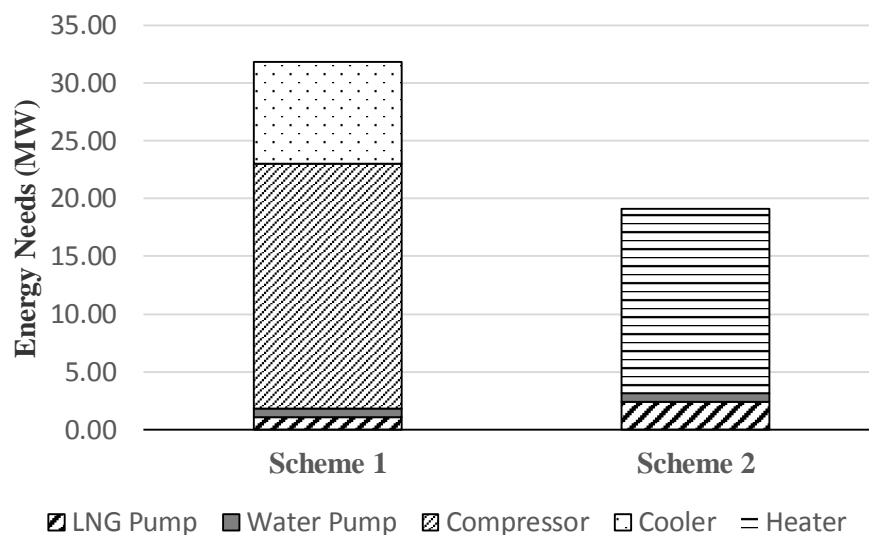
**Table 1.** Use of Energy at Regasification Plant (MW).

Process Equipment	Scheme 1	Scheme 2
LNG Pump	1.04	2.37
Water Pump	0.79	0.81
Compressor	21.15	-
Cooler	8.87	-
Heater	-	15.97
Total	31.85	19.15

From the simulation results in table 1 and figure 5, it can be seen that the energy in scheme 1 is very large for the compressor process equipment, and small enough for the LNG pump energy needs compared to scheme 2. In scheme 2, the amount of energy required is in the heater unit that functions to increase the gas temperature to achieve the gas pipe specification. In scheme 2, there is no compressor to increase pressure but is replaced by a high pressure LNG pump which requires considerable magnitude compared to scheme 1. If viewed as a whole, scheme 2 requires less energy than scheme 1. Energy savings can be done by replacing the process in order to increase in pressure from the compressor unit to the high pressure LNG pump unit reaches 40%.

### 3.2 Optimization of LNG Regasification Operating Conditions

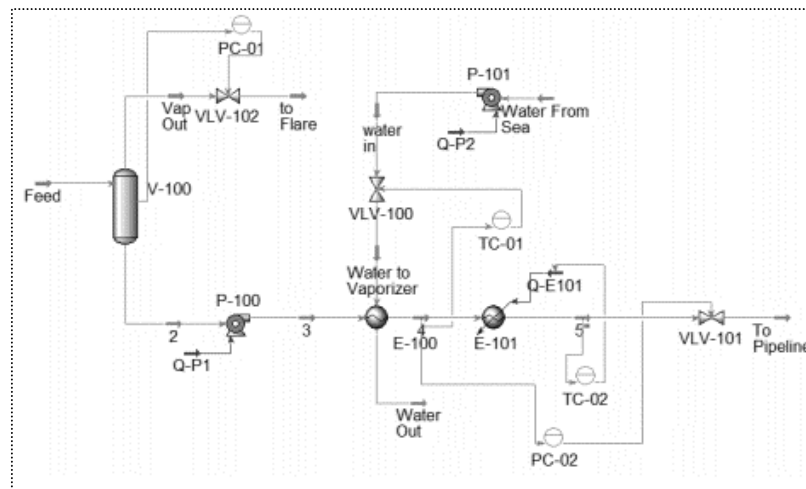
At the LNG regasification plant the unit that has an important role is the vaporizer unit. In this unit, there is a phase change from liquid LNG to gas which is the main objective of the regasification process. Therefore, an optimum operating condition is required for this unit to make the plant run efficiently. To obtain optimum condition simulations are performed using Unisim R390.1, to see the effect of vaporizer output temperature with required seawater pump power and duty of heater needed to achieve gas pipe specification.



**Figure 5.** Proportion of Energy Requirement of LNG Regasification Plant.

### 3.3 Empirical Model and Adjustment of Control

Control system at LNG regasification plant that used in this research shown in Figure 6. The empirical model is obtained from PRC (process reaction curve) resulting from system identification for each controller. From the PRC, FOPDT empirical models were developed as shown in Table 2, which is computed by equation. (1).



**Figure 6.** Control System at LNG Regasification Plant.

**Table 2.** Process Model of Regasification LNG Plant.

Controller	FOPDT
Pressure Control V-100	$\frac{0.069 e^{-0.0108s}}{0.103s + 1}$
Pressure Control Stream 4	$\frac{0.121 e^{-0.003s}}{0.017s + 1}$
Temperature Control E-100	$\frac{0.045 e^{-0.0182s}}{0.642s + 1}$
Temperature Control E-101	$\frac{0.381 e^{-0.0088s}}{0.0438s + 1}$

### 3.4 Performance of Control

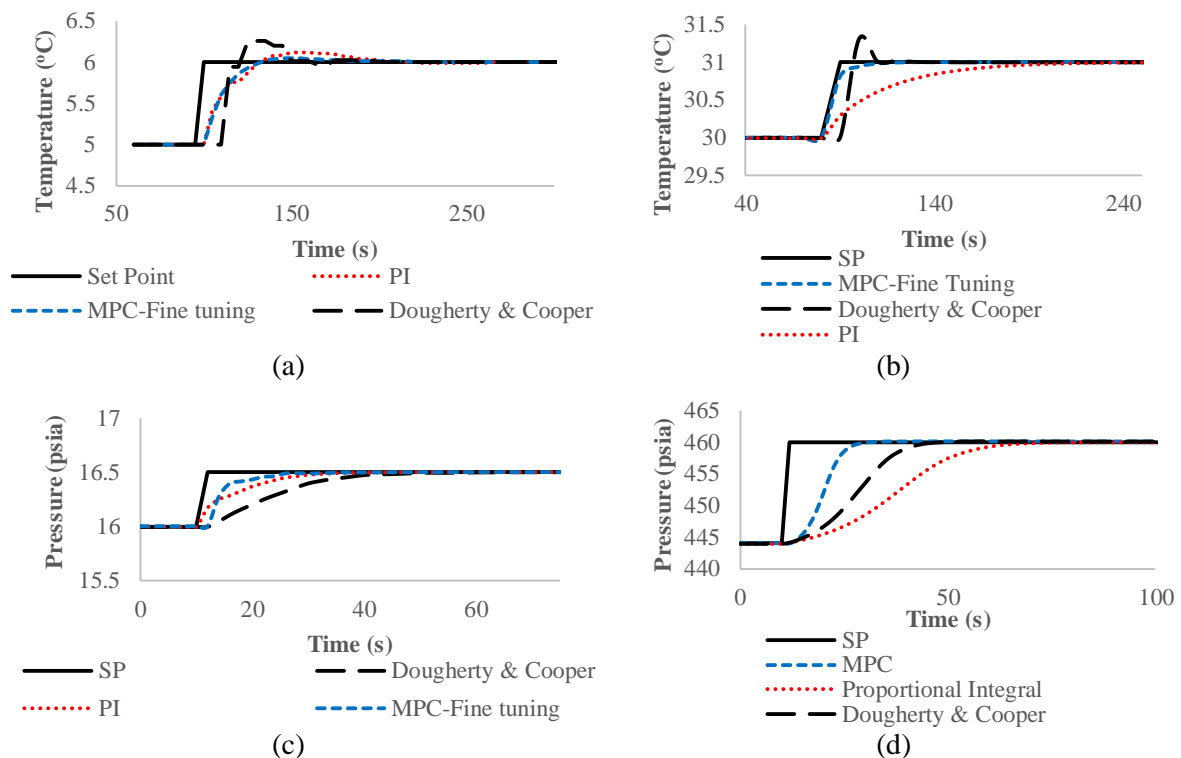
The way to find out the performance of each controller, set point tracking and disturbance rejection are performed and evaluated with the value of ISE. Here are the results of both methods:

**3.4.1 Set Point Tracking.** In this method, a set point change is made to see how the controller response to achieve process stability. The controller that has the lowest ISE value is the best controller. The result for the set point tracking can be represented by Figure 7. From Figure 7, the tuning with MPC-Fine tuning method produces the best performing control with the total ISE scores in sequence for the vaporizer, heater, storage tank, and stream 4 controllers are 34.88, 7.54, 0.98 and 1792.78.

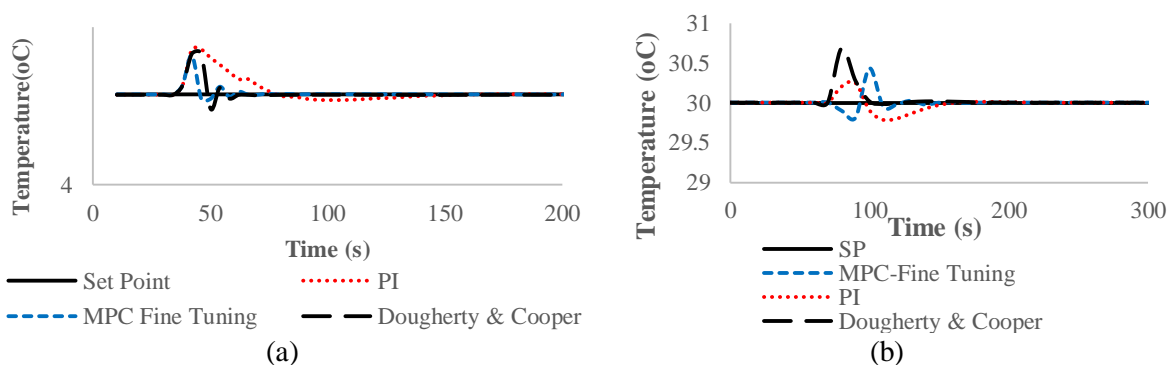
**3.4.2 Disturbance Rejection.** In this method, disturbance is given to the system to see the controller response in overcoming the disturbance on the system. The results of the test for disturbance rejection can be seen in Figure 8. From Figure 8, the tuning with MPC-Fine tuning method produces the best performing control with the total ISE scores in sequence for the vaporizer, heater, storage tank, and stream 4 controllers are 7.4, 7.54, 0.35 and 0.016.

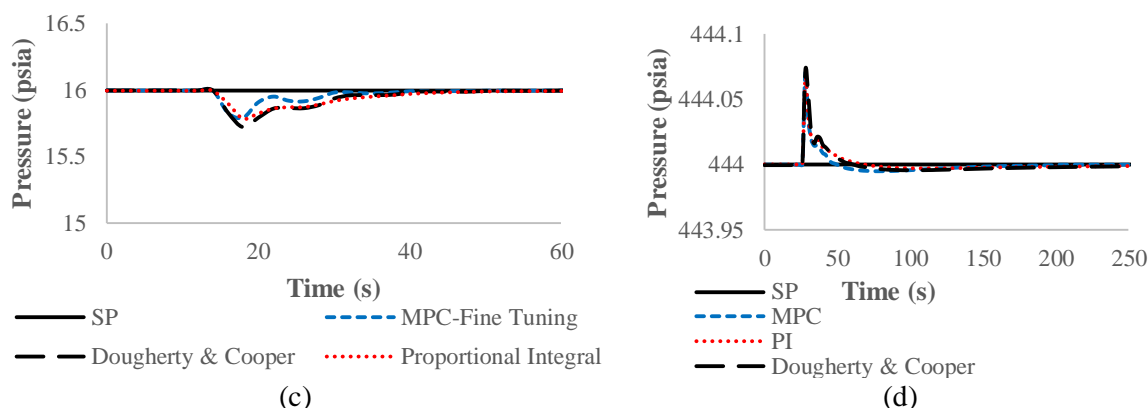
### 3.5 Energy Efficiency

To determine the effect of errors on both controllers on energy to be used can be seen in figure 9 and figure 10. Based on the calculation of energy deviation, the use of control with the MPC-fine tuning method has fewer errors than the proportional-integral method controller. The use of MPC-fine tuning controller can reduce 0.02 MW and MPC controller also reduces error to product quality by 34.25% compared to the PI controller.

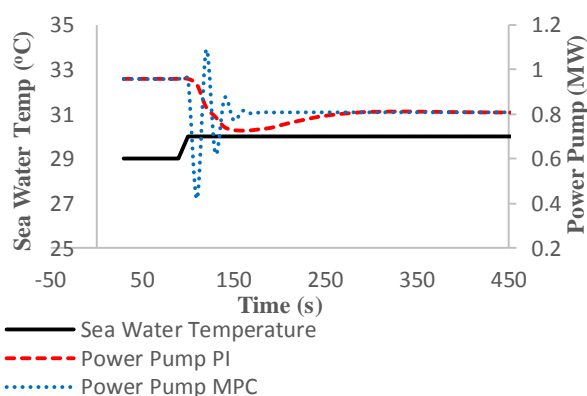


**Figure 7.** Comparison of Three Controller (Set Point Tracking): (a) Vaporizer (b) Heater (c) Storage Tank (d) Stream 4.

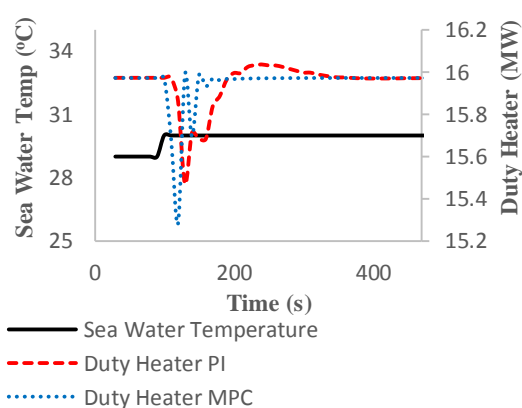




**Figure 8.** Comparison of Three Controller (Disturbance Rejection): (a) Vaporizer (b) Heater (c) Storage Tank (d) Stream 4.



**Figure 9.** Relationship needs of pump power to disturbance on Sea water temperature.



**Figure 10.** Relationship needs of Duty Heater to disturbance on Sea water temperature.

#### 4. Conclusion

Scheme 2 LNG regasification plant requires less energy than scheme 1, with energy savings reaching 40% from scheme 1, so scheme 2 is chosen. Tuning MPC (P, M and T) that is required as follows: pressure tank control (90, 2, 1), pressure on stream 4 (95, 2, 1), vaporizer temperature (65, 2, 2) and heater temperature: (35, 6, 5) with ISE value in set point tracking respectively 0.99, 1792.78, 34.89 and 7.54, or improved control performance respectively 4.6%, 63.5%, 3.1% and 58.2%, compared to PI controller performance. The energy savings that MPC controllers can perform when a disturbance in temperature rise 1°C of sea water is 0.02 MW.

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#### References

- [1] Wahid, A and Afdal, A 2016 Model predictive control based on system re-identification for methanol and dymethyl ether synthesis control. *Proc. Conf. Nasional Teknik Kimia "Kejuangan"*. (Yogyakarta)
- [2] Wahid, A and Tanuwijaya, R 2015 Pemilihan Metode Penyetelan Pengendali PI pada Pengendalian



- Pabrik Regasifikasi LNG Menggunakan Metode Skor. *Proc. Conf. Nasional Teknik Kimia UNPAR*. (UNPAR: Bandung)
- [3] Devold, H 2013 *Oil and Gas Production Handbook: An Introduction to Oil and Gas Production, Transport, Refining and Petrochemical Industry*. (Oslo : ABB Oil and Gas)
- [4] Marlin, T 2000 *Process Control: Designing Processes and Control Systems for Dynamic Performance*. (United States: McGraw-Hill Higher Education)
- [5] Smith, C and Corripio, A 1997 *Principles and Practice of Automatic Process Control*. (United States: John Wiley & Sons Inc.)
- [6] Himmelblau D M, Edgar T F and Lasdon L S 2001 *Optimization of Chemical Processes, Second Edition*. (New York: McGraw-Hill)
- [7] Altman, W 2005 *Practical Process Control for Engineers and Technicians*. (IDC Technologies)