MONITORING AND CONTROLLING OF TEMPERATURE IN A GAS PLANT VIA CASCADE ARCHITECTURE

By

MOHAMAD FASYAN BIN MOHAMAD SABRI

FINAL PROJECT REPORT

Submitted to the Electrical & Electronics Engineering Programme
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Universiti Teknologi Petronas
Bandar Seri Iskandar
31750 Tronoh
Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

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Approved by,

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(DR NORDIN SAAD)

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

JUNE 2009

CERTIFICATION OF ORIGINALITY

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

MOHAMAD FASYAN BIN MOHAMAD SABRI

ABSTRACT

This report discusses the development and implementation of computer control on an industrial process plant. The objectives of the project is to design and tune two different PID controller for the control of temperature in a gaseous pilot plant. The gaseous pilot plant, located at Universiti Teknologi PETRONAS, is used in the case study. The focus of the project is on the control and monitoring of the temperature of gas in the pilot plant.

The PID controller will be designed and simulated via MATLAB/Simulink. The work involves two main stages, modeling and simulation, and real-time implementation. Once the PID controller has been designed and simulated via MATLAB/Simulink, the model will be interfaced to the plant via an xPC target card for real-time analysis.

The result of this investigation shows that the cascade control architecture would be a viable method to be used in plant process control. The cascade configuration that indicates the better performance can specifically be defines to use the Ziegler Nichols closed loop tuning method for the primary loop, while for the secondary loop the Cohen Coon tuning method is preferable.

ACKNOWLEDGEMENT

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Special thanks to my family members for the moral support and strength, to which has fueled me through out. To all students and lecturers of Universiti Teknologi PETRONAS, who directly or indirectly contributed to the project, the help given is highly appreciated.

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CHAPTER 1

INTRODUCTION

1.1 BACKGROUND STUDY

The use of electronics and computers for the control of automated processes has been widely used over the past decades. The advancement of computers saw the control system used in the industry to move forward in tandem, such examples being the introduction of the Supervisory Control and Data Acquisition (SCADA) System. However, the SCADA system is still not a full individual control system as it is only overlays the hardware, focusing on the supervisory level. With its main function being to monitor and logs process data, the SCADA system is still at a software level, where it only interfaces with the programmable logic controller (PLC). With this in view, the need to have a more direct form of monitoring was recognized, a form of monitoring and control directly affiliated to the transmitters/transducer.

1.2 PROBLEM STATEMENT

Among some of the critical issues in the plant is on the controlling and monitoring of temperature and. In previous approaches, as explained in previous published papers, the temperature of a gas medium can be controlled using its own individual proportional-integral-differential (PID) controller in a single closed loop system. However, the control performance when using a single loop can sometimes be unsatisfactory. Cascade control is a method that could dramatically improve the performance of a single loop control by utilizing additional measurement of a process variable to assist in the control system.

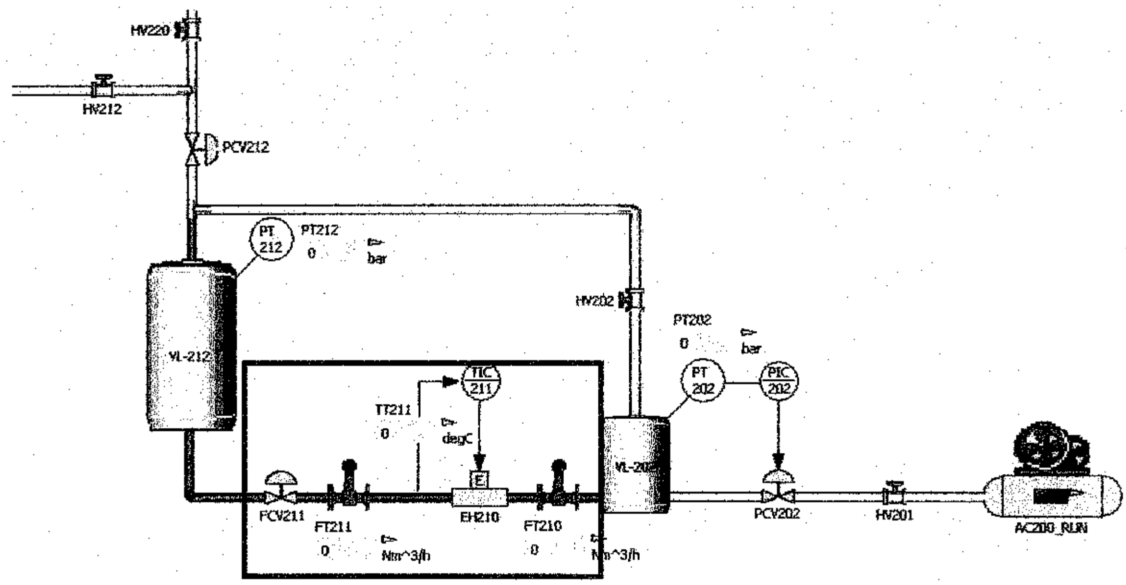

Figure 1: Block diagram of test rig

Figure 1 shows the block diagram of the test rig used for the implementation of the control strategy under study in this project. Here, we target to control the temperature of the gas medium along the pipe that goes through the flow transmitter and also the temperature transmitter. The reading measured by TT211 will provide as the additional secondary process input variable as required by the cascade architecture. The PID controllers, tagged with TIC211 and FIC211, will be used in the cascade architecture and will be modeled using the MATLAB/Simulink software. The controller should perform well between the operation range and at the desired set point despite of the abnormalities. The stability of the system is also taken into consideration.

This project will attempt to address the issue on computer control of the said variables by designing and implementing PID control. This project will involve designing and implementation of two different PID controllers and an attempt to bind both controllers through cascade architecture.

1.3 OBJECTIVE AND SCOPE OF STUDY

The case study revolves around two main control variables: the flow and temperature of the gaseous medium in the pilot plant. These two control variables are to maintain one process variable, which is the temperature of the gas medium within the pipe of the plant. Hence, the focus of the project is investigating ways on monitoring and controlling the control variables.

With the above stated, the main objectives of this project can best be described as follows:

i. To design and implement a PID controller for temperature process control of a gas medium in a Gaseous Pilot Plant.

ii. To integrate two PID controllers into one single functioning control design for temperature control of a gas medium in a Gaseous Pilot Plant through cascade method.

The objectives above are relevant in investigating the viability of implementing cascade PID control on an industrial process. Tentatively, the objectives above are to be achieved within two semesters. The modelling and testing of the PID controller using MATLAB/Simulink is to be completed within the first semester of the FYP year, while the real life implementation is to be carried out during the second half of the year.

CHAPTER 2

LITERATURE REVIEW

2.1 THEORY

2.1.1 Proportional-Integral-Derivative controller

A proportional–integral–derivative controller (PID controller) is a generic control loop feedback mechanism widely used in the field of control systems. A PID controller attempts to correct the error between a measured process variable and a predetermined setpoint by calculating the corrective value based on calculations done within the PID.

The PID controller algorithm involves three separate parameters; the Proportional, the Integral and Derivative values. The Proportional value determines the reaction to the current error, the Integral determines the reaction based on the sum of recent errors and the Derivative determines the reaction to the rate at which the error has been changing. This can also be described through a mathematical representation:

$$Kc = [E(t) + \frac{1}{T_1}\int_0^1 E(t')dt' + Td\,\frac{dE(t)}{dt}] + 1$$

$$Kc = [E(t) + \frac{1}{T_1}\int_0^1 E(t')dt' + Td\,\frac{dCV(t)}{dt}] + 1$$

(Recommended by Thomas E. Marlin)

There are two forms of expressions when the derivative mode is expressed. The first of which is the Instrument Society of America (ISA) standard, and the second is the form recommended by Thomas E. Marlin, 1995. The second form is recommended as it prevents set point changes from causing excessive response. The derivative mode amplifies sudden changes in the controller inputs signal, and can potentially cause a large variation in the controller output. This may be unwanted primarily for two reasons, first of which is that step changes to the set point can lead to step changes in the error. The derivative of a step change goes to infinity, or in a more practical scenario, to a completely open or closed control valve. This could lead to a severe process upset and could even be a health risk. The recommended form will reduce the extreme variation in the manipulated variable[4].

By "tuning" the three constants in the PID controller algorithm, the controller manipulative control action designed for specific process requirements. The response of the controller can be described in terms of the responsiveness of the controller to an error, the degree to which the controller overshoots the set point and the degree of settling time.

2.1.2 Cascade Design Architecture

The work involves the integration of two individual PID controllers to cooperate and gain control of one common control variable. In the case of this project, it goal is to control the temperature (control variable) of the gas medium in the pipe along the gaseous plant via the flow and temperature transmitter (process variable).

There are a few design approaches; each boasts its own improvements in performance and adaptability. In this project, the cascade design architecture is selected due to a few primary reasons. The dynamics of the temperature behavior is differs than that of the typical flow behavior dynamics. Temperature is a much slower process and its reaction time is slow compared to that of flow. This is one of the main criteria needed to be fulfilled when designing via cascade architecture, one variable must respond well compared to its cascaded partner variable[3].

The cascade architecture is comprised of two ordinary controllers from the PID group, and is specifically designed for improved disturbance rejection.

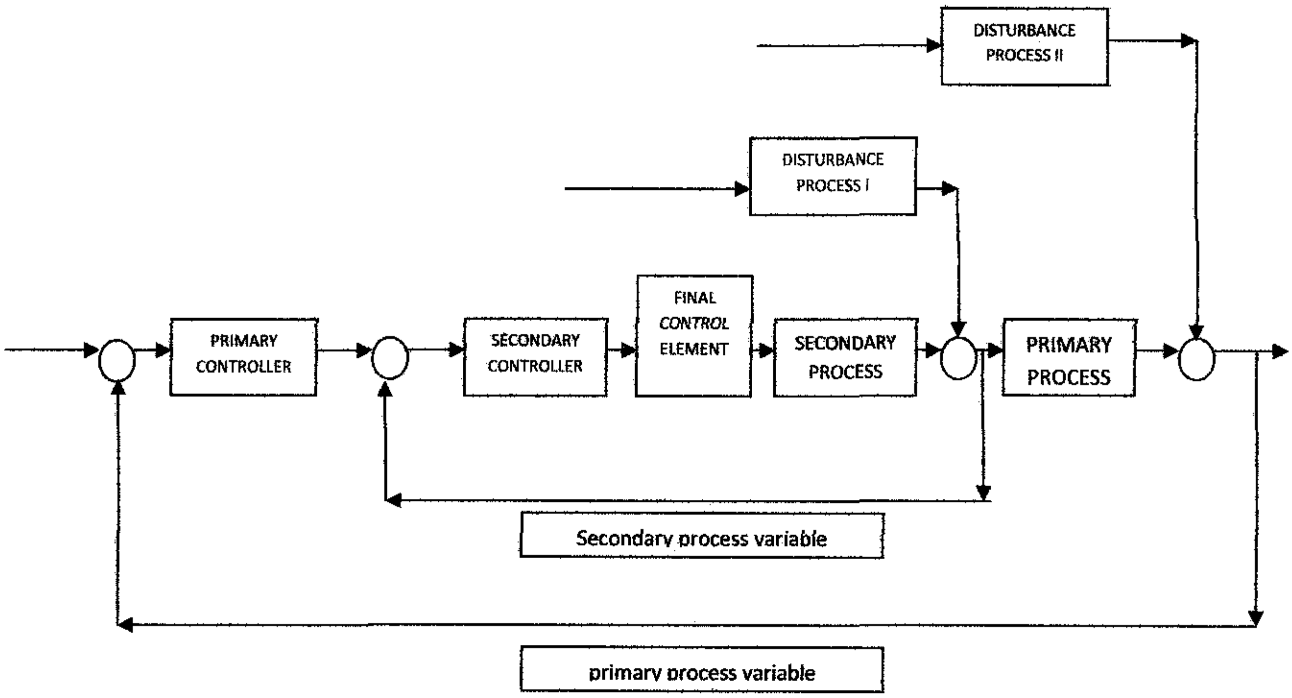

Figure 2: Block diagram of cascade architecture

Figure 2 shows the block diagram of the cascade architecture. It is noted that the secondary loop uses the normal feedback control loop, and is nested within the primary loop. The success of this cascade implementation requires that the settling time of the secondary loop is significantly faster than the settling time of the primary loop[3]. The cascade architecture caters to two process variables to control one control variable. To implement this architecture, the variables must meet a certain criteria:

- The variables must be measurable with a sensor.
- Both primary and secondary variable must be able to be manipulated by one common valve.
- The secondary variable must respond well before the primary variable to disturbances and final control element manipulations.
- Both primary and secondary variable are disrupted by the same disturbance.

A cascade will require two individual sensors and two controllers, but only one final control element. This is because the output of the primary controller will be the set point of the secondary controller.

CHAPTER 3

METHODOLOGY

3.1 PROCEDURE IDENTIFICATION

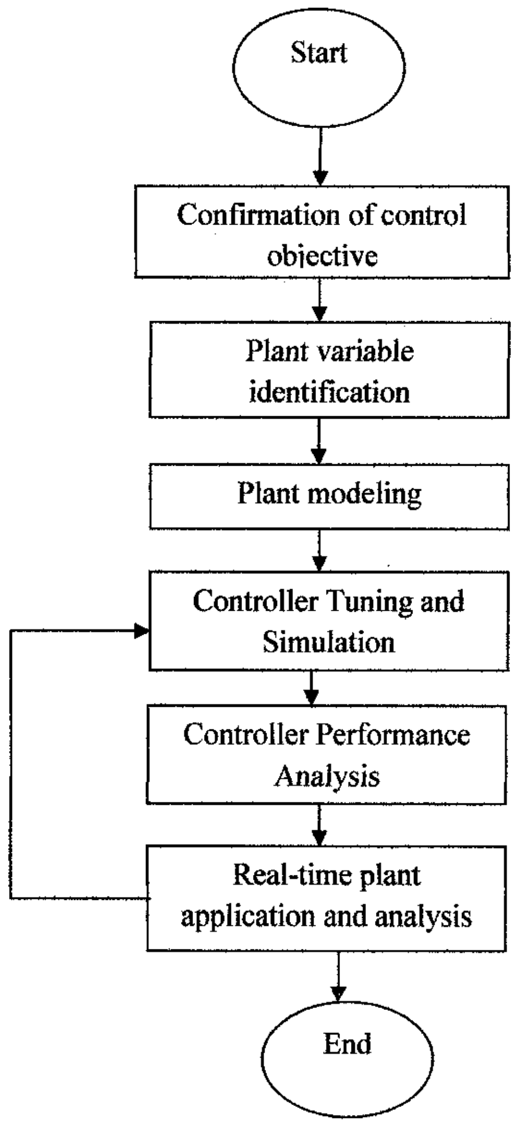

Figure 3 : Overall workflow chart

3.2 LITERATURE REVIEW

Several papers [3] has been referred to develop understanding on controller design and the variables involved in the project. A revision on the process control [4] provides an improved understanding of the temperature and flow control in the gaseous pilot plant. Research on the PID controller also has been conducted to better understand it function and definition. The research covers on the characteristic of the PID, function and effect of each controller elements.

Another aspect of the controller design will involve the parameters identification and controller tuning. This involves modelling and simulation, tuning and implementation. The process model is derived via Empirical Modelling.

3.3 PARAMETERS INDENTIFICATION

Empirical modelling is used to yield the plant transfer function. The empirical

of this reaction curve is to identify the dynamic model, which in turn will be used on the

flow variables. The process reaction curve is obtained through the following steps:

2. Introduce a single step change in the input variable.

4. Perform the graphical process reaction curve calculations

Figure 4

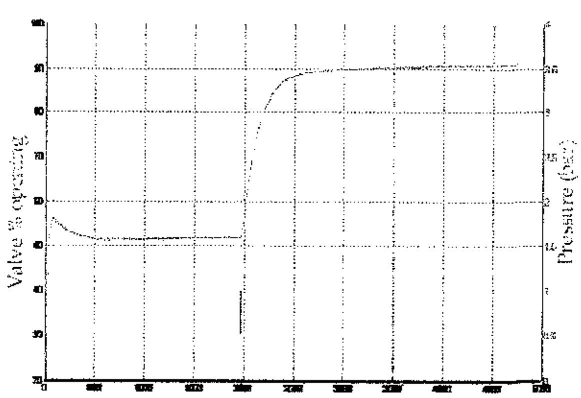

Figure 4: Sample process reaction curve.

calculate the function parameters and the general first order transfer function.

The general first order plus dead time model can be obtained through the

$$\frac{Y(s)}{X(s)} = \frac{Kp * e^{-\theta s}}{\tau s + 1}$$

Where

$$Kp = \frac{\Delta}{\delta}$$

$$\tau = \frac{\Delta}{s}$$

$$\theta = intercept\ of \max slope\ with\ initial\ value$$

$$s = slope\ of\ graph$$

Alternatively, Method II can also be used, where the equations are expressed as

$$Kp = \frac{\Delta}{\delta}$$

$$\tau = 1.5(t63\% - t28\%)$$

$$\theta = t63\% - \tau$$

$$t63\% = time\ taken\ to\ reach\ 63\%\ of\ final\ value$$

$$t28\% = time\ taken\ to\ reach\ 28\%\ of\ final\ value$$

Once the transfer function is obtained, we can then find the tuning coefficient for the PI controller using either the Ziegler-Nicholas open loop or the Cohen-Coon method. This will involve computer simulation involving MATLAB/Simulink as well a LabVIEW. Each simulation responses will be compared to obtain the best parameter and used in the cascade PID controllers.

3.4 TOOLS & CONFIGURATION

This project will utilize a combination of hardware and software setup. The setup is configured so as to allow signal from the workstation to be transferred to the Gaseous Pilot Plant. The list of hardware and software are listed as following:

3.4.1 MATLAB/Simulink software

Used for modelling, simulating and analyzing the dynamical systems. This will be the main software used to design, tune and test the PID controller. The control block diagram will be constructed using the Simulink application within MATLAB.

3.4.2 LabVIEW Application

LabVIEW Application is one of the software used for real time monitoring. This application will be used to monitor the process during the experiment. The process variable that have to be monitored can be specified and represent in graphic form in LabVIEW.

3.4.3 Gaseous Pilot Plant

The Gaseous Pilot Plant is the process plant that will be used in the case study. The pilot plant is located in the Plant Process Laboratory at Block 23 of the Universiti Teknologi PETRONAS academic complex. The plant consists of real functioning equipments and components which is similar to any industrial process plant. These include valves, transmitters, controller and so forth. It should be noted however, that the plant is at laboratory scale. This plant is able to cater to simulation of a plant that

CHAPTER 4

RESULTS AND DISCUSSION

4.1 PLANT DESCRIPTION

The part of the plant that is concentrated on is the pipe along which the gaseous medium is transferred as indicated with the red line along the diagram below.

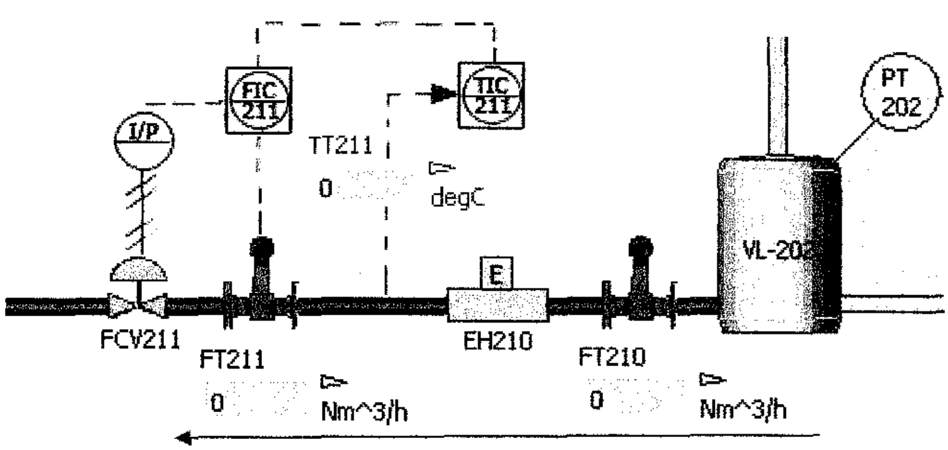

Figure 5: Plant block diagram (area of interest) with proposed cascade architecture

The gaseous medium will travel along the pipe passing first through a temperature transmitter (TT 211) and then through a flow transmitter (FT 211). The aim is to control the temperature of the gas medium along the pipe using both the flow and temperature control loop. The cascade design approach is most suited for this task as there are two variables available for control, both with one common manipulated variable. Since the dynamics of temperature control is slower than the dynamics of flow, it is more suited that the primary variable is temperature, and the secondary variable is flow. This can best be described graphically:

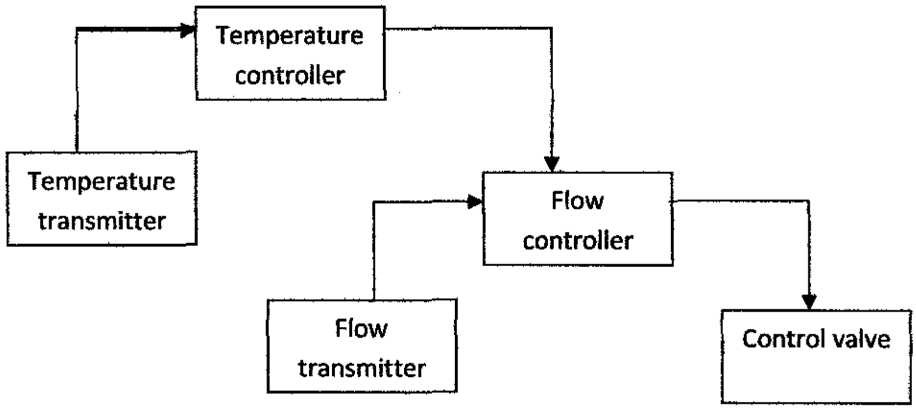

Figure 6: Flow diagram of suggested cascade architecture

Note that the output of the temperature controller (primary) will become the set point of the flow controller (secondary), and that in turn will manipulate the control valve.

4.2 RESULTS

The first step toward designing the cascade configuration is the secondary loop (flow loop) tuning. An initial plant experiment was conducted to obtain the Process Reaction Curve. The flow behaviour of the gas medium was observed and the data collected for analysis. The control valve selected to control the flow of the gas medium is FCV211. The valve input change in the valve opening is set at 20%. The flow rate of the gas medium was initially at 30 kg/hr Celsius, and the change in flow after the input change is observed until change is no longer observable or the flow rate has reached a relatively constant value. The flow rate data trend can best be viewed through the graphs.

4.2.1 Secondary Loop Open loop analysis

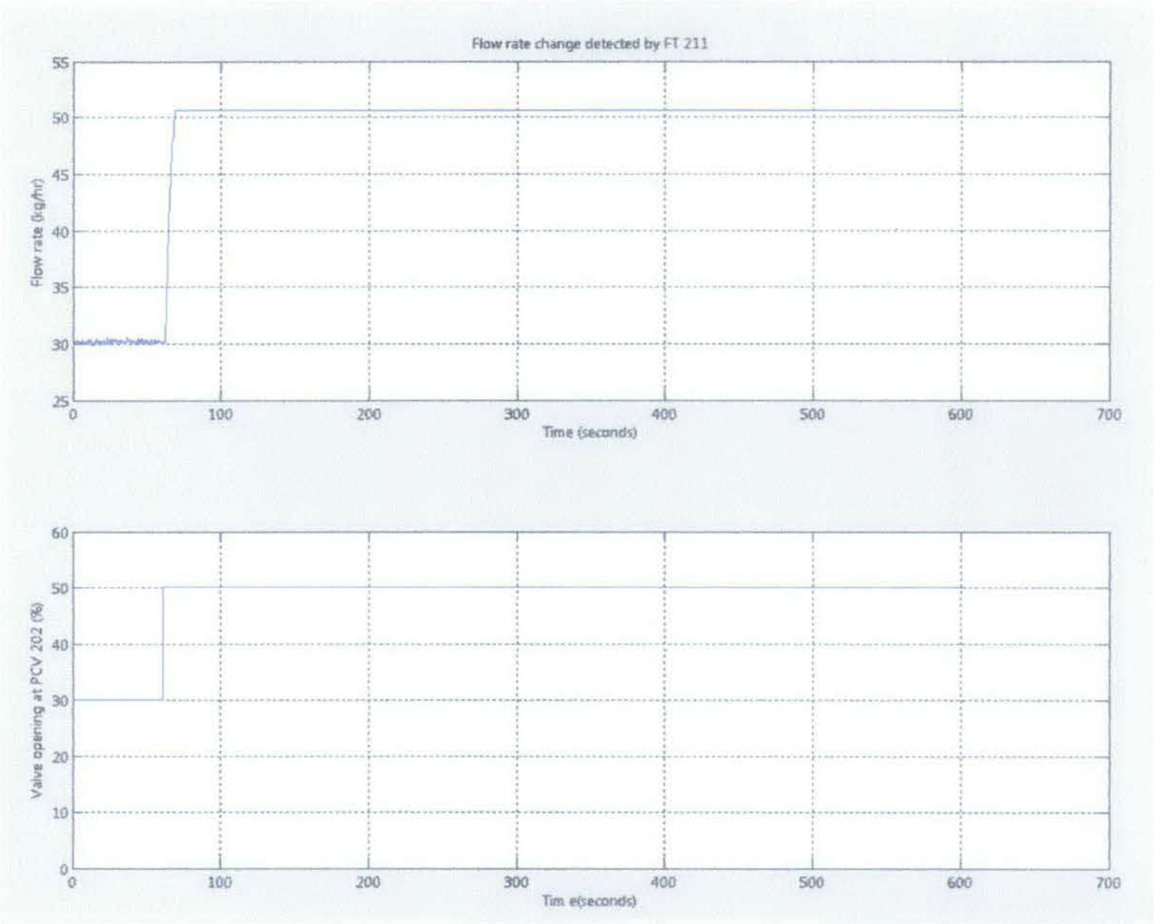

Figure 7: Graph of flow rate versus elapsed time (top) and input change versus elapsed time (bottom).

The graph obtained from the experiment was analyzed to produce the first order transfer function. In this case, Method II was used to evaluate the parameters. Method II was chosen because the graph obtained contains a lot of noise, and thus is difficult to evaluate based using Method I [4]. Method II also is generally preferred as Method I typically has larger errors in the parameter evaluation [4]. From the process reaction curve, again Method II was used to calculate the parameters for the first order transfer function.

Table 1: Process Reaction Curve Analysis for flow using method I

Parameters	Value	Unit
Change in manipulate variable	20	%
Change in ultimate value,dBu	20.5	m^3/hr
Slope, S	2	(m^3/hr)/seconds
Apparent time constant, τ	10.25	seconds
Apparent dead time, θ	10	seconds
Steady state Process Gain, Kp	1.025	$(m^3/hr)/\%$

The first order plus dead time model obtained using Method I are as follows

$$\frac{Y(s)}{X(s)} = \frac{1.025e^{-10s}}{10.25s + 1}$$

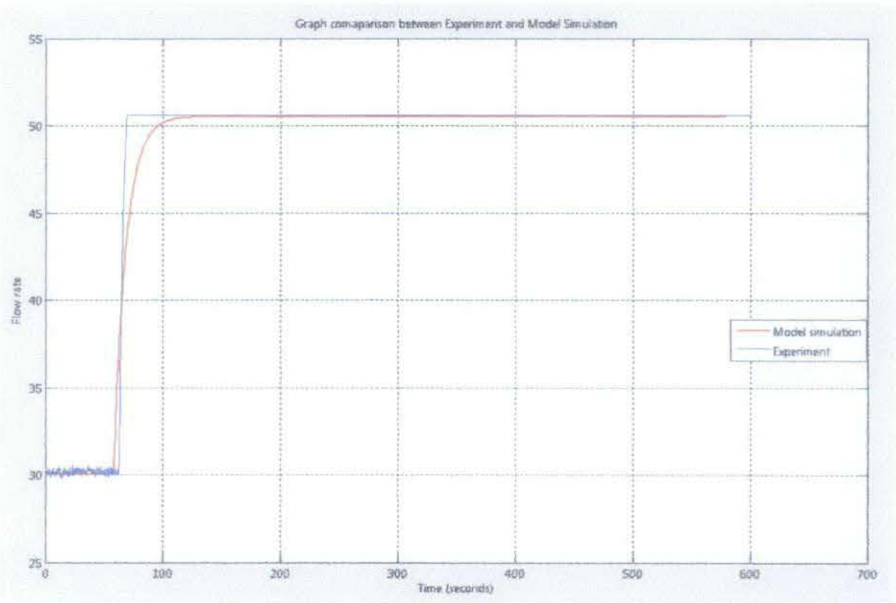

Figure 8: Comparison between experimental (blue) and simulated result (red) for flow rate using Method I

Figure 8 illustrates the comparison between the dynamic response of the simulated and experimental process reaction curve for flow rate, when using Method I. The simulated curve shows deviation to the experimental response during the transient stage. The experimental curve shows a more step like response compared to the model simulated curve. This could be due to the limitation of the flow transmitter in detecting rapid change in values. When the disturbance is applied to FCV 211, flow rate changes drastically over a short period of time, as illustrated in Figure 10. During the experiment, data was sampled at 0.1 seconds intervals. This interval could be the cause for the error between the two graphs during the transient stage. The simulated model has an error of -0.97%.

Table 2: Process Reaction Curve Analysis for flow rate using method II

Parameters	**Value**	**Unit**
initial value	30	m^3/hr
final value	50	m^3/hr
Δ (initial value-final value)	20	m^3/hr
28.3% of final value	35.66	m^3/hr
63.2% of final value	42.64	m^3/hr
Time to reach 28% of final value, t28	63.7	seconds
Time to reach 63% of final value, t63	65.9	seconds
Apparent time constant, τ	3.3	seconds
Apparent dead time, θ	62.6	seconds
Change in manipulated variable	20	%
Gain, Kp	1.025	$(m^3/hr)/\%$

The first order plus dead time model obtained using Method II are as follows:

$$\frac{Y(s)}{X(s)} = \frac{1.025e^{-62.6}}{3.3s+1}$$

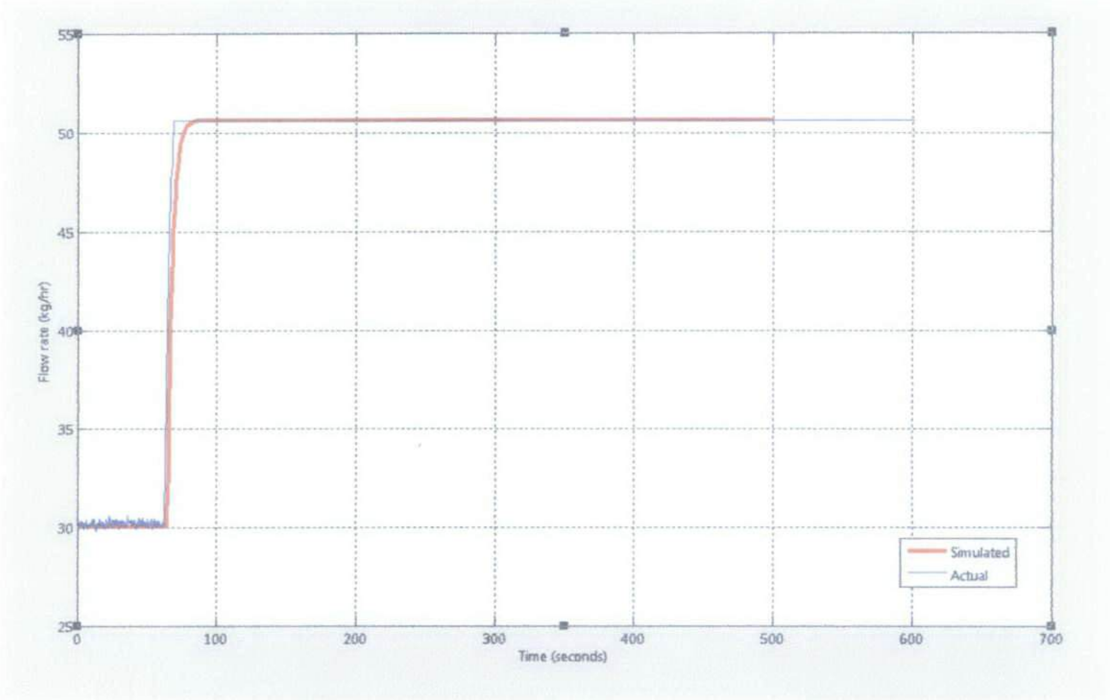

Figure 9: Comparison between experimental (blue) and simulated result (red) for flow rate using method II

Figure 9 is the comparison between the dynamic response of the simulated and experimental process reaction curve for flow rate, when using Method II. Similarly, the simulated curve shows deviation to the experimental response during the transient stage. The experimental curve shows a more step like response compared to the model simulated curve. Again, as mentioned above could be due to the limitation of the flow transmitter in detecting rapid change in values. However, when using Method II, the error between the model simulated curve and the experimental curve is only -0.14%. This clearly indicates that using Method II yields a more accurate first order plant transfer function.

4.2.2 Validation for secondary loop

The first order plant transfer function and its error in relative to the experimental model are tabulated as follows.

Table 3: Error analysis between Method I and Method II

Variable	|Error| using Method I (%)	|Error| using Method II (%)
Flow	0.97	0.14

From the table, it is clear that for both cases, using Method II yields a more accurate plant transfer function. Thus, the plant transfer functions selected to represent the flow response of the plant is:

$$\frac{Y(s)}{X(s)} = \frac{1.025e^{-62.6}}{3.3s+1} \quad \text{..................(4.2.2.1)}$$

First order transfer function selected to represent the process reaction curve for flow.

4.2.3 Ziegler Nichols Closed loop tuning (secondary loop)

To determine the tuning parameters, a couple of methods can be used. The first of which is using the Zieger –Nichols closed loop-Bode plot method. This tuning method provides two advantages [1]:

- Can be applicable to processes that are not well modified by first-order with dead time models.
- Provides considerable insight into the effect of all loop elements (process, instrumentation and control algorithm) on stability and proper tuning constant values

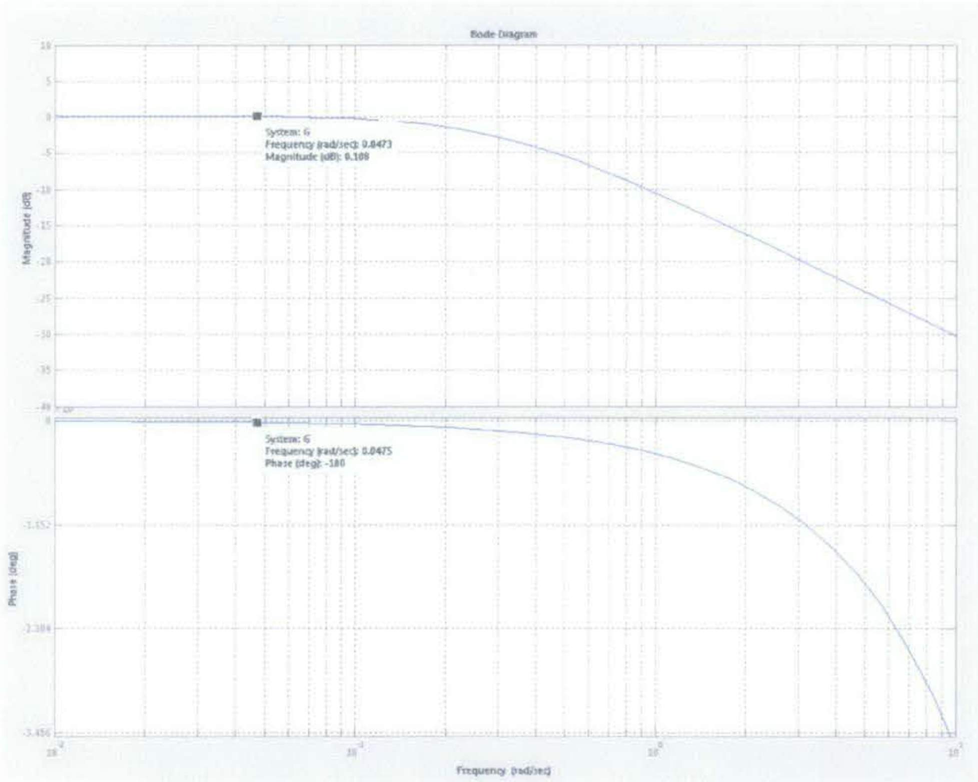

Figure 10: Bode plot for Kp = 1.025, $\tau = 3.3$, $\theta = -62.6$

From the bode plot, we are able to see that at the critical frequency of -180 degrees, the corresponding frequency is 0.0475 rad/sec and the magnitude is 0.108dB. Thus, this means that the ultimate gain Ku is 1.102 and the Pu is 132.27 (refer to Appendix). Using these two values, the tuning parameters are calculated based on the formula (refer to Appendix) and tabulated as follows:

Table 4: Zieger-Nichols closed loop Tuning Parameters for flow

Controller Parameter	P Controller	PI Controller	PID controller
Kc	0.506	0.46	0.595
Ti	n/a	110.23	66.14
Td	n/a	n/a	16.53

Once we have obtained the tuning parameters, we can then proceed to simulate the response of the system. From the simulation, we can then deduce whether the system is stable or not stable, and decide whether the tuning is useful for improving the system's performance.

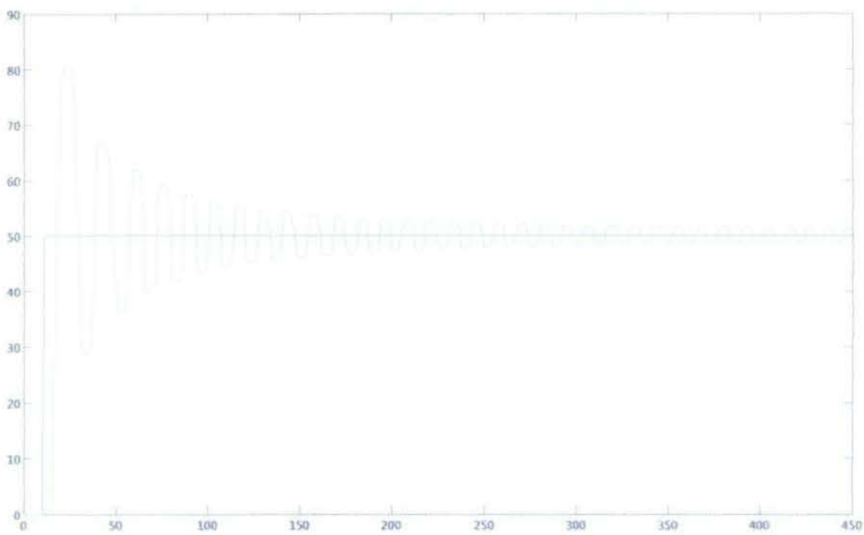

Figure 11: The response for P controller of Z-N Closed loop Method flow

From Figure 11, we can see that the system response is not stable. The manipulated variable keeps oscillating when a set point change is introduced. The oscillating manipulated variable reflects that the control variable will not settle at the predetermined setpoint.

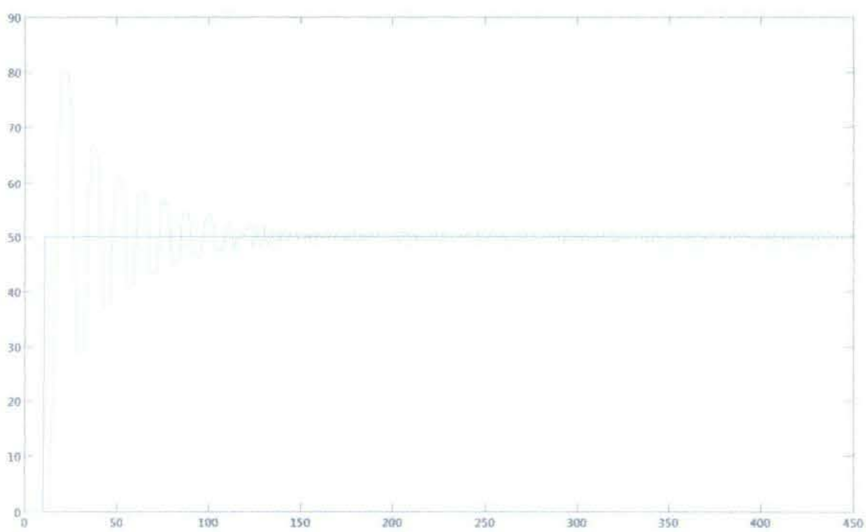

Figure 12: The response for PI controller of Z-N Closed loop Method for flow (bottom)

The second performance evaluation is the PI controller the Ziegler-Nichols Closed Loop method. From the result shown as in Figure 12, the response for the flow system is better compare to the P only controller performance. The oscillatory response of the controller decreases towards the end of the simulation, however fails to settle about the set point.

From observation of the Ziegler Nichols Closed Loop Method performance evaluation, the PI controller shows the most desirable performance. Despite having a large setting time, the controlled variable settles about the set point with less than 5% error.

4.2.3 Cohen Coon Tuning Method (secondary loop)

Table 5: Cohen coon Tuning Parameters for flow

Controller Parameter	P Controller	PI Controller	PID controller
Kc	0.38	0.13	0.31
Ti	n/a	14.04	55.55
Td	n/a	n/a	5.12

Table 5 shows the tuning parameters obtained using the Cohen Coon Tuning method (refer to Appendix). The Cohen Coon tuning method is based on the values obtained from the process reaction curve; gain, dead time, and time constant. The parameters are then used in the controller modes, and the performance of each controller is analyzed as previously done with the Ziegler Nichols Closed loop tuning method done in the previous section.

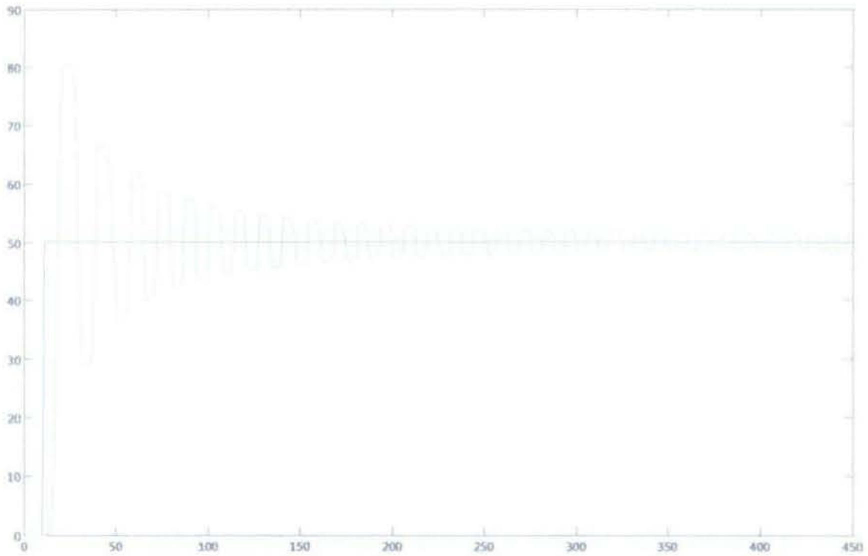

Figure 13: The response for P controller of Cohen coon Method for flow

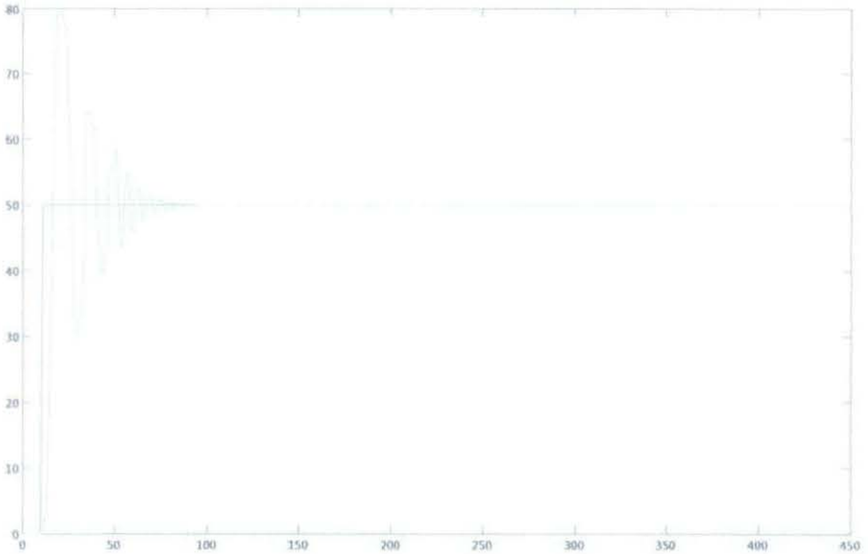

Figure 14: The response for PI controller of Cohen coon Method for flow

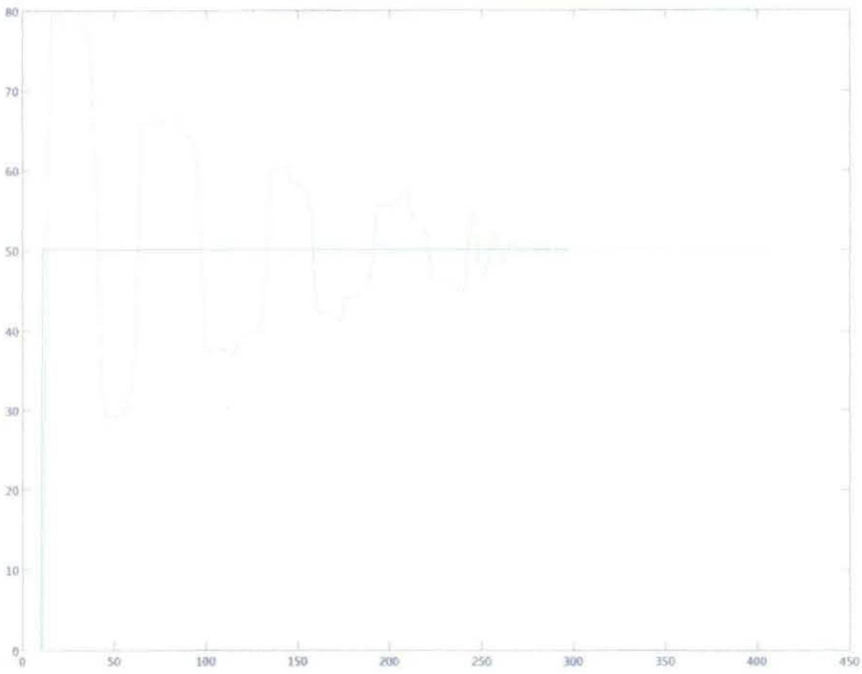

Figure 15: The response for PID controller of Cohen coon Method flow

The controlled variable for the flow loop does settle about the setpoint when in the PI and PID controller mode, as shows in Figure 14 and Figure 15. For both modes, the system settles about the set point as required. However the settling time for the PID controller is notably larger than when the system is using the PI controller. This indicates that for the case of flow controller, the PI controller shows a better performance.

As a conclusion, the Cohen Coon tuning method is applicable for flow control. The system shows a stable response when using the PI and PID controller modes. The oscillatory response settles at the setpoint, with the PI controller showing a more desirable performance.

4.2.4 Ziegler Nichols Open Loop Tuning Method (secondary loop)

The Ziegler-Nichols Open Loop tuning method is a tuning method in the open loop analysis. This method provides correlations that are applicable to process models developed from open loop process reaction [4] such as was done to obtain the process reaction curve. Table 6 shows the tuning parameters calculated using the Ziegler Open loop method formula (refer to Appendix).

Table 6: Zieger-Nichols open loop Tuning Parameters for flow

Controller Parameter	P Controller	PI Controller	PID controller
Kc	0.05	0.05	0.06
Ti	n/a	206.58	125.2
Td	n/a	n/a	31.3

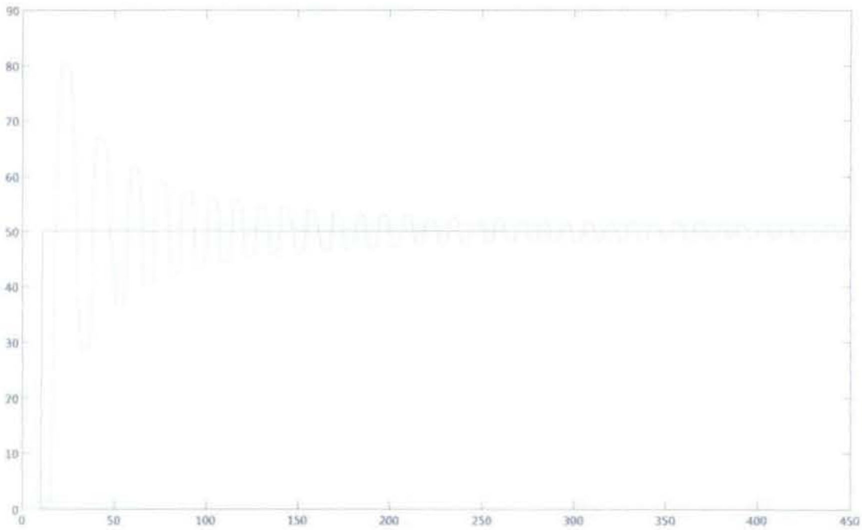

Figure 16: The response for P controller of Ziegler Nichols Open Loop tuning Method for flow

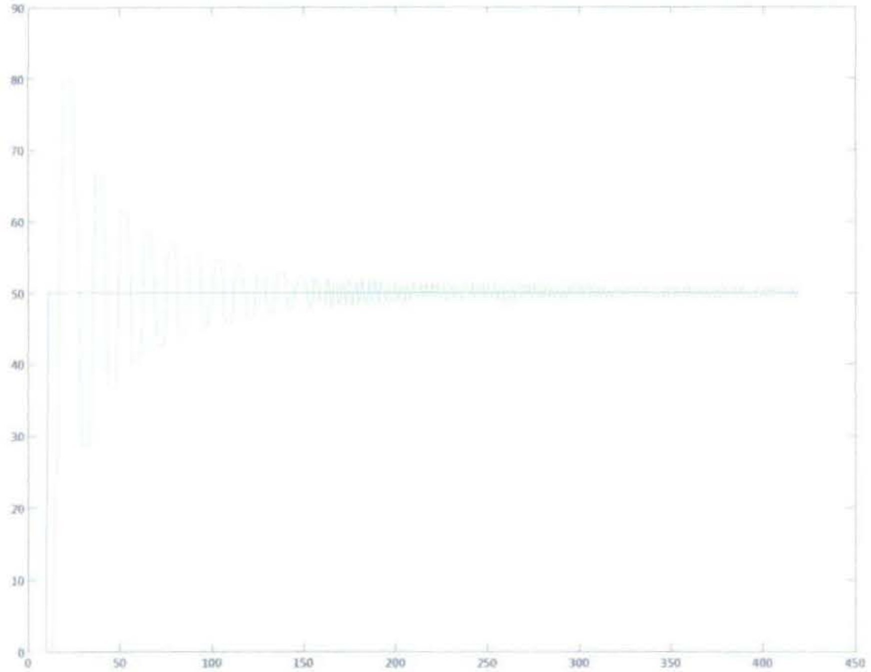

Figure 17: The response for PI controller of Ziegler Nichols Open Loop tuning Method for flow

The controlled variable continues to oscillate when about the setpoint when a change is done. This is true for all cases of controller modes.. As can be seen in Figures 16 and 17, the response of the system shows that it is unstable; concluding that the Ziegler Nichols Open loop method is unsuitable for this particular application.

4.2.5 Primary Loop Open loop analysis

Once we have tuned the secondary loop, we can now proceed with tuning the primary loop. The primary loop is tuned with the secondary loop in auto (closed loop) mode. The primary loop is now modelled by perturbing the secondary variable set point, which in turn will cause the primary variable to respond. The process is allowed to reach an initial steady state before a step disturbance is applied. The method used to obtain this process reaction curve is similar to the method used in the secondary loop open loop analysis.

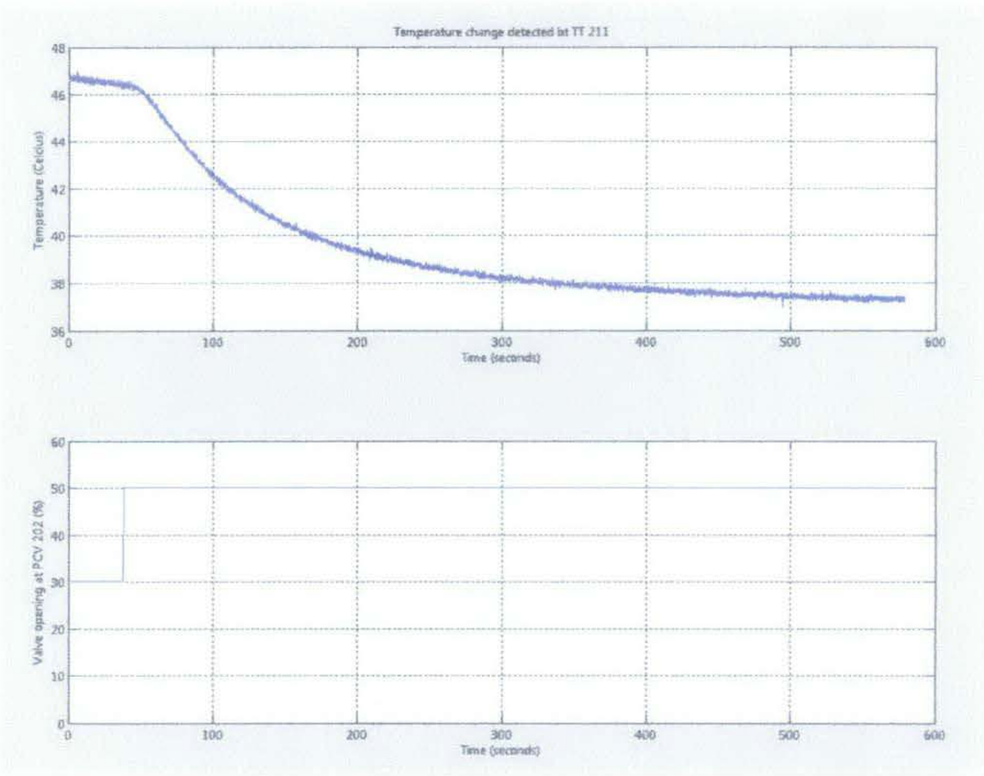

Figure 18: Graph of temperature versus elapsed time (top) and input change versus elapsed time (bottom).

Figure 18 shows the process reaction curve obtained. From the process reaction curve, again Method II was used to calculate the parameters for the first order transfer function.

Table 7: Process Reaction Curve Analysis for temperature Using Method I

Parameters	Value	Unit
Change in manipulate variable	20	%
Change in ultimate value,	9.2245	celcius
Slope, S	0.04	celcius/seconds
Apparent time constant, τ	230.6125	seconds
Apparent dead time, θ	30	seconds
Gain, Kp	0.461225	celcius/%

The first order plus dead time model obtained using Method I are as follows:

$$\frac{Y(s)}{X(s)} = \frac{0.461225e^{-30}}{307.48s + 1}$$

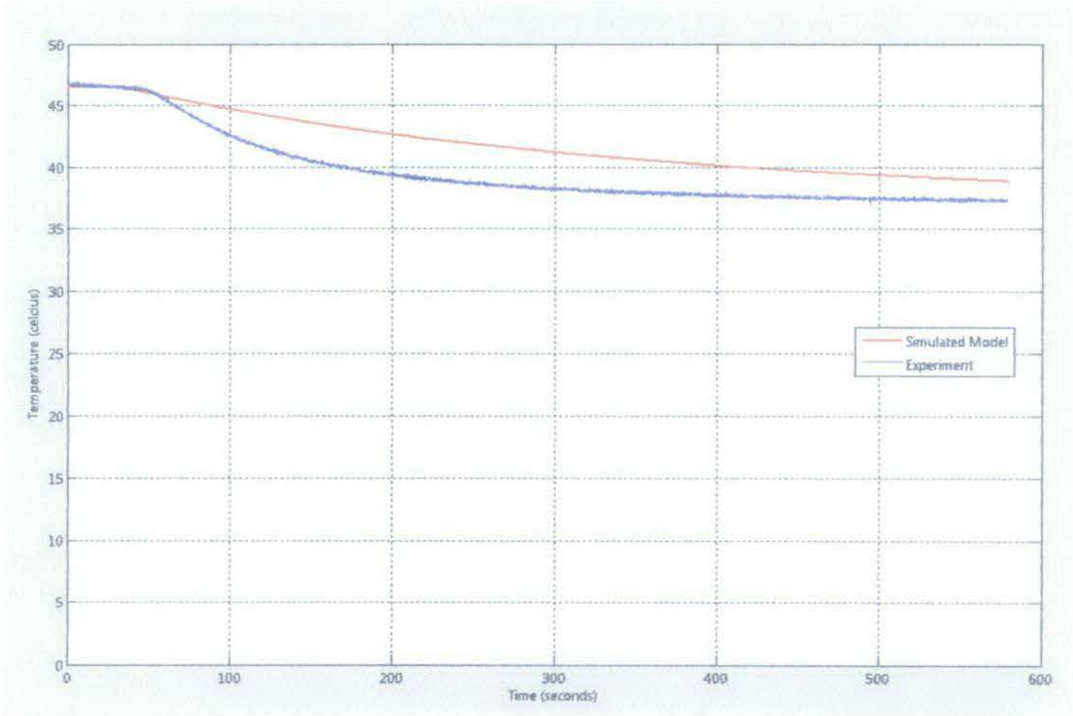

Figure 19: Comparison between experimental (blue) and simulated result (red) for Temperature using method I.

Figure 19 illustrates the dynamic response of the simulated and experimental process reaction curve. The simulated model curve shows a sizable deviation to the experimental response. This could be because the experimental process reaction curve has a lot of noise, which causes difficulty in evaluating the slope as required when using Method I. In signals with high frequency noise, Method I typically has larger errors in the parameter estimates [4]. The model simulated curve has a positive error of 4.65%.

Table 8: Process Reaction Curve Analysis for temperature Using Method II

Parameters	***Value***	***Unit***
initial value	46.5833	celcius
final value	37.3588	celcius
Δ (initial value-final value)	9.2245	celcius
28.3% of final value	43.9727665	celcius
63.2% of final value	40.753416	celsius
Time to reach 28% of final value, t28	78.3	seconds
Time to reach 63% of final value, t63	137.1	seconds
Apparent time constant, τ	88.2	seconds
Apparent dead time, θ	47.9	seconds
Change in manipulated variable	20	%
Gain, Kp	-0.461225	celcius/%

The first order plus dead time model obtained using Method II are as follows:

$$\frac{Y(s)}{X(s)} = \frac{-0.461225e^{-47.9}}{88.2s + 1}$$

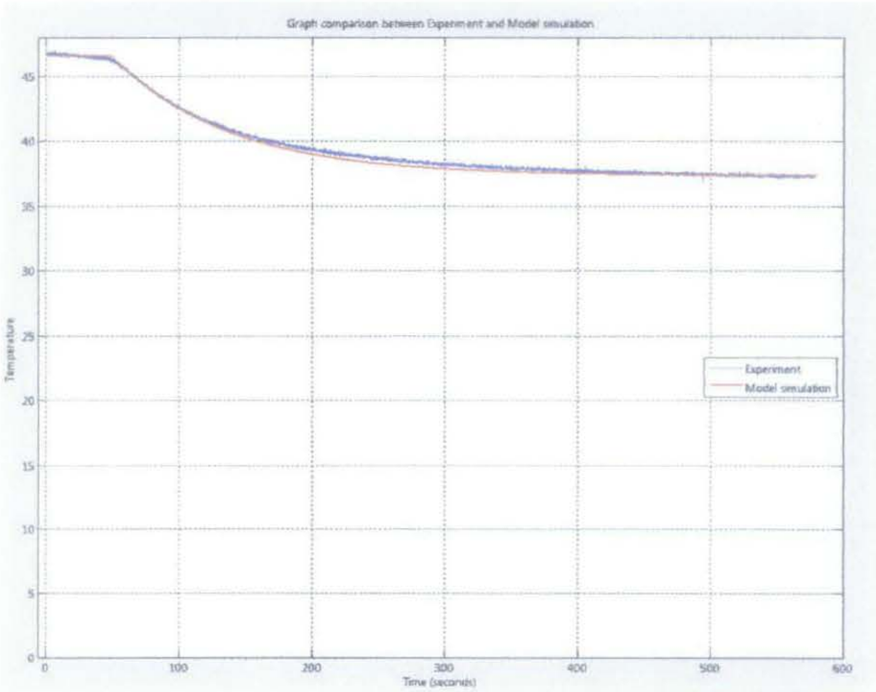

Figure 20: Comparison between experimental (blue) and simulated result (red) for Temperature using method II.

Figure 20 illustrates the dynamic response of the simulated and experimental process reaction curve when using Method II. The simulated curve shows little deviation to the experimental response as compared to when using Method I. This could be because the experimental process reaction curve has a lot of noise, thus calculations done to obtain the general first order plus dead time model transfer function using Method II is inaccurate. The model simulated curve has an error of 0.02% which indicates that using Method II yields a more accurate first order model.

4.2.6 Validation for primary loop

The first order plant transfer function and its error in relative to the experimental model are tabulated as follows.

Table 9: Error analysis between Method I and Method II

Variable	|Error| using Method I (%)	|Error| using Method II (%)
Temperature	4.65	0.02

From the table, it is clear that for both cases, using Method II yields a more accurate plant transfer function. Thus, the plant transfer function selected to represent the temperature response of the plant is:

$$\frac{Y(s)}{X(s)} = \frac{-0.461225e^{-47.9}}{88.2s+1}$$

4.2.7 Ziegler Nichols Closed loop tuning (Primary)

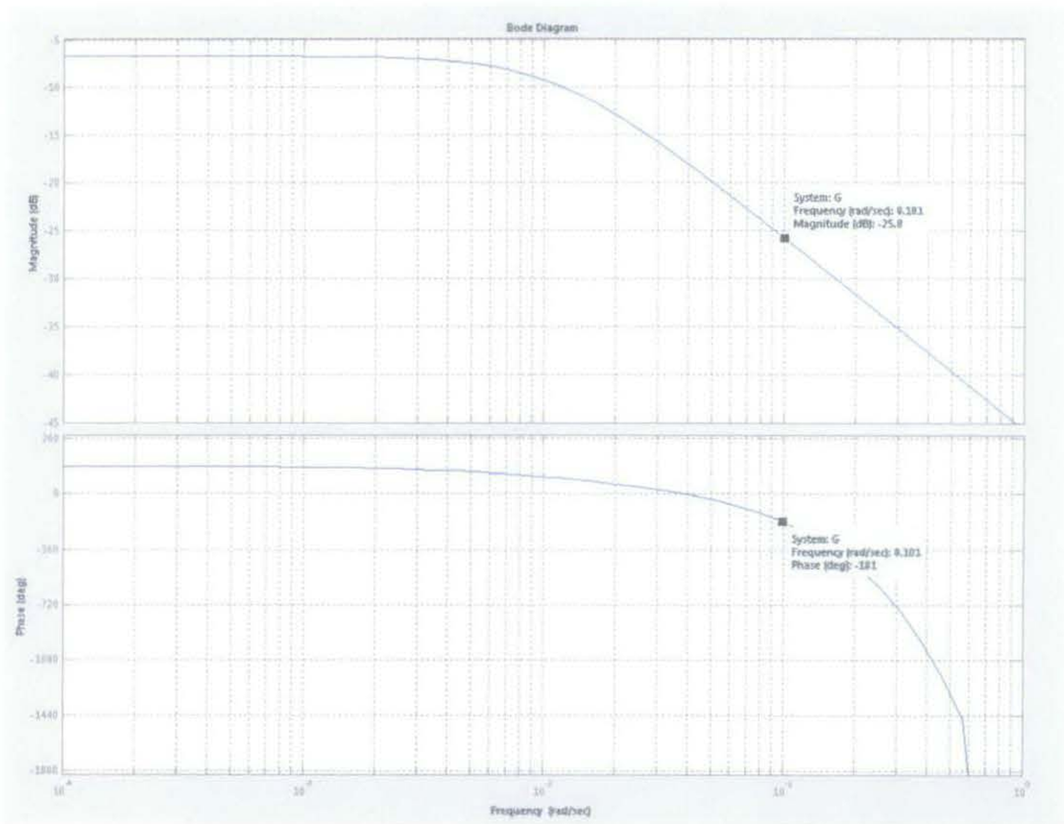

Figure 21: Bode plot for Kp = -0.46, τ = 88.2, θ = 47.9

From the bode plot, at -180 degrees, the frequency is 0.101 rad/sec at a magnitude of -25.8dB. This means that the ultimate gain, Ku, is equal to 19.5 and the ultimate period, Pu, is 1.03 (refer to Appendix). Simillarly, these two values are used to calculate the tuning parameters for the Ziegler Nichols Closed loop method.

Table 10: Ziegler-Nichols closed loop Tuning Parameters for temperature

Controller Parameter	P Controller	PI Controller	PID Controller
Kc	9.75	8.86	11.47
Ti	n/a	0.858	0.515
Td	n/a	n/a	0.129

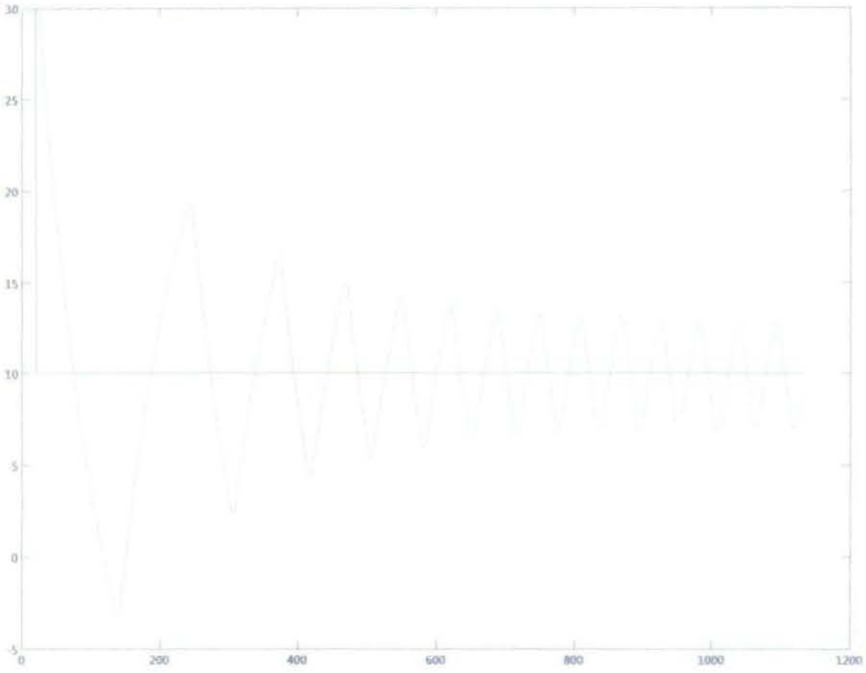

Figure 22: The response for P controller of Z-N Closed loop Method for temperature

From Figure 22, we can see that for temperature loop using the P only mode, the system response is not stable. The manipulated variable keeps oscillating when a set point change is introduced. The oscillating manipulated variable reflects that the control variable will not settle at the predetermined setpoint.

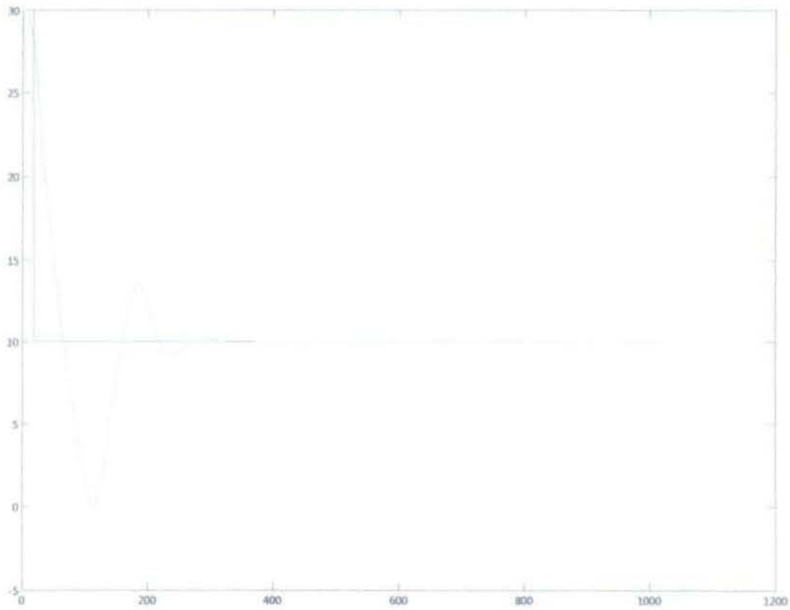

Figure 23: The response for PI controller of Z-N Closed loop Method for temperature.

The second performance evaluation is the PI controller the Ziegler-Nichols Closed Loop method. From the result shown as in Figure 16, it shows that the system reaches stability for temperature. The manipulated variable for has a fast settling time and stabilizes about the setpoint with no oscillation.

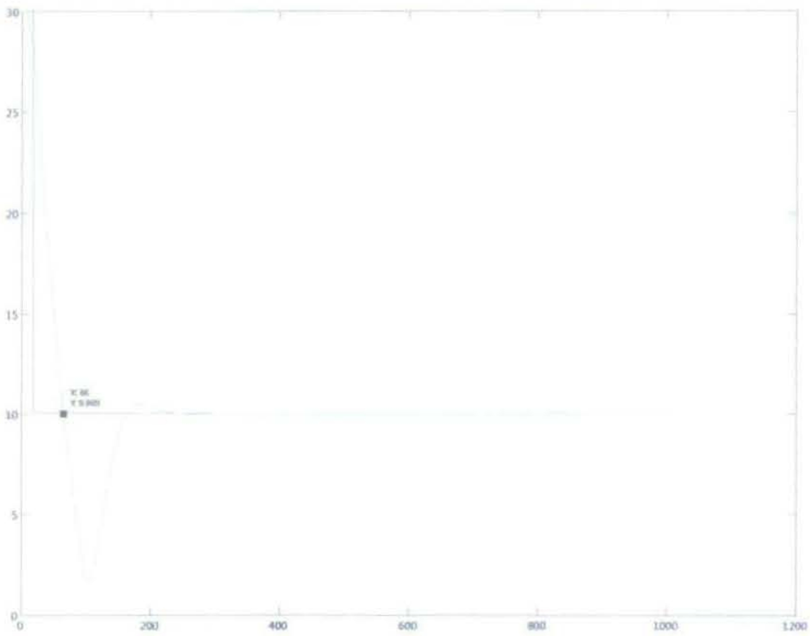

Figure 24: The response for PID controller of Z-N Closed loop Method for temperature

Figure 24 shows the performance evaluation of a PID controller for the Ziegler-Nicols Closed loop method. The system performance for the temperature system shows an improvement compared to the performance when using a PI controller. The PID controller results in a less overshoot of the manipulated variable, as well as a significantly faster settling time. PID control for flow is unsuited, as the derivative parameter obtained through the Ziegler-Nicols is large and results in a highly unstable response.

From observation of the Ziegler Nichols Closed Loop Method performance evaluation, the performance of the PI controller and PID controller results in a stable system response. When using the PI controller for the flow system, despite having a large setting time, the controlled variable settles about the set point with less than 5% error.

4.2.8 Cohen Coon loop tuning (Primary)

Table 11: Cohen coon Tuning Parameters for temperature

Controller Parameter	P Controller	PI Controller	PID controller
Kc	-4.73	-3.78	-5.88
Ti	n/a	76.28	97.37
Td	n/a	n/a	15.85

Table 11 shows the tuning parameters obtained using the Cohen Coon Tuning method based on the obtained plant model parameters. The performance of the controller using the above tuning parameters are analyzed as previously done with the Ziegler Nichols Closed loop tuning method done for the secondary (flow) loop.

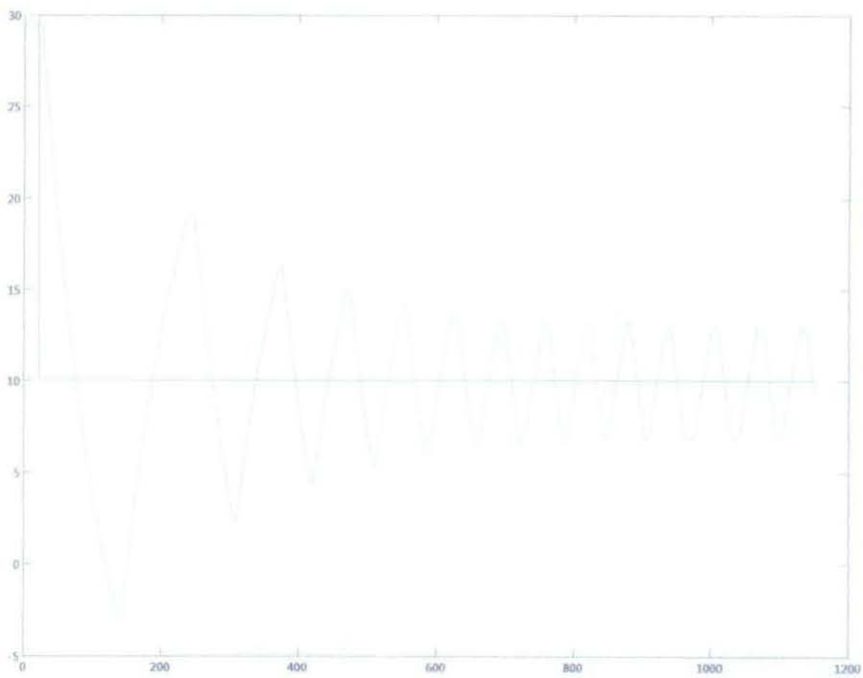

Figure 25: The response for P controller of Cohen coon Method for temperature

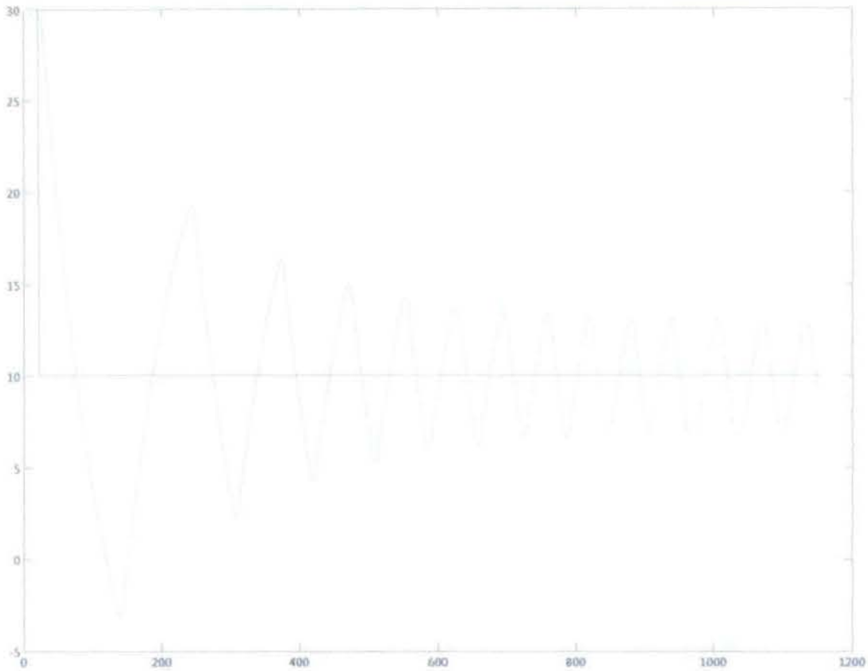

Figure 26: The response for PI controller of Cohen coon Method for temperature

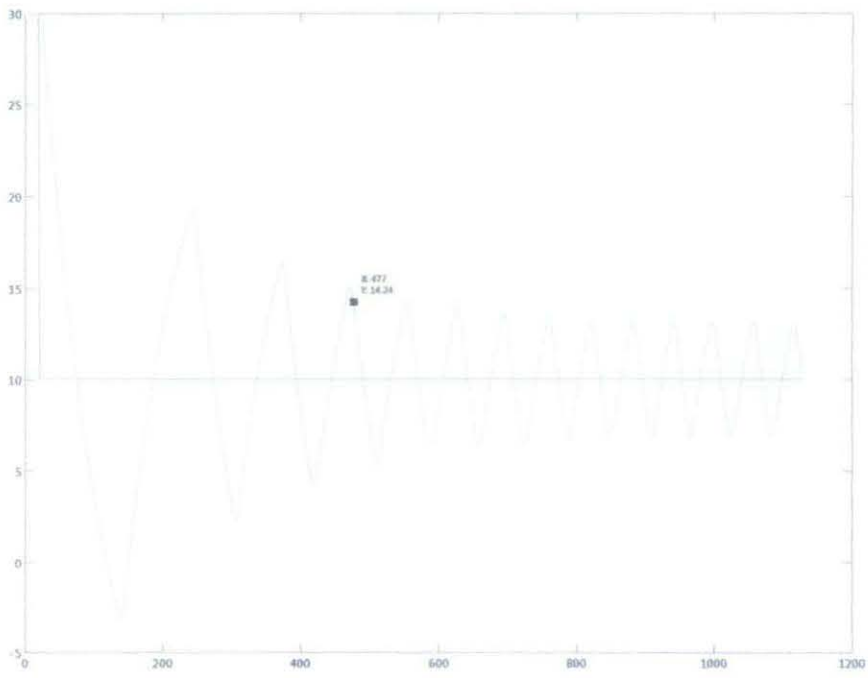

Figure 27: The response for PID controller of Cohen coon Method for temperature

The controlled variable for the temperature loop keeps oscillating for all modes of the controller. The system fails to stabilize at the set point, which indicates that the Cohen Coon tuning parameters are unsuitable to be used for temperature control.

As a conclusion, the Cohen Coon method is unable to control the temperature as it fails to maintain the controlled variable at the setpoint. The system is unstable for all controller modes.

4.2.7 Ziegler Nichols Open loop tuning (Primary)

Table 12: Zieger-Nichols open loop Tuning Parameters for temperature

Controller Parameter	P Controller	PI Controller	PID controller
Kc	-4.00	-3.60	-4.80
Ti	n/a	158.07	95.8
Td	n/a	n/a	23.95

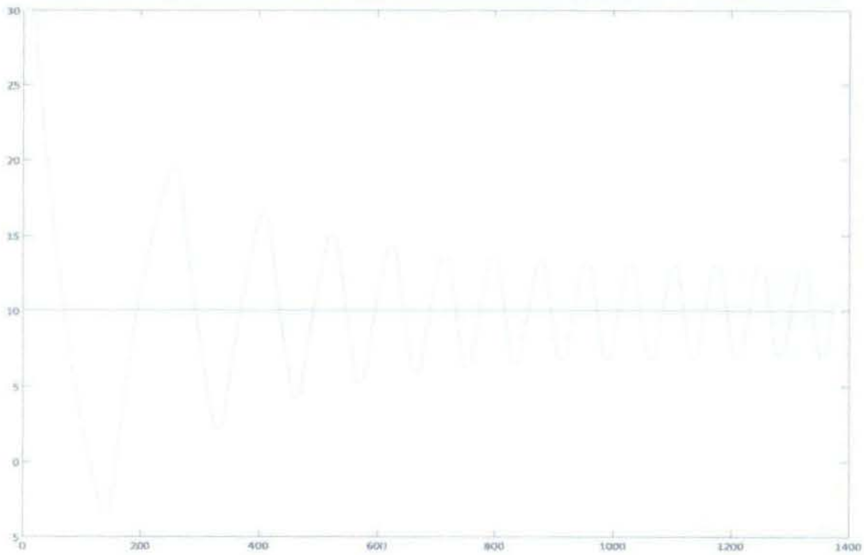

Figure 28: The response for P controller of Ziegler Nichols Open Loop tuning Method for temperature

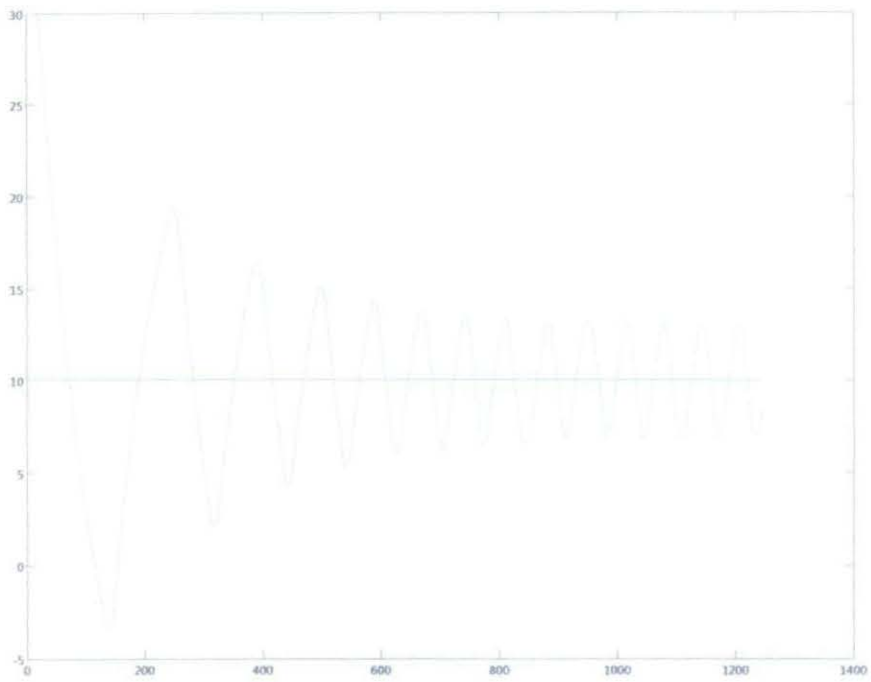

Figure 29: The response for PI controller of Ziegler Nichols Open Loop tuning Method for temperature (above) and flow (below)

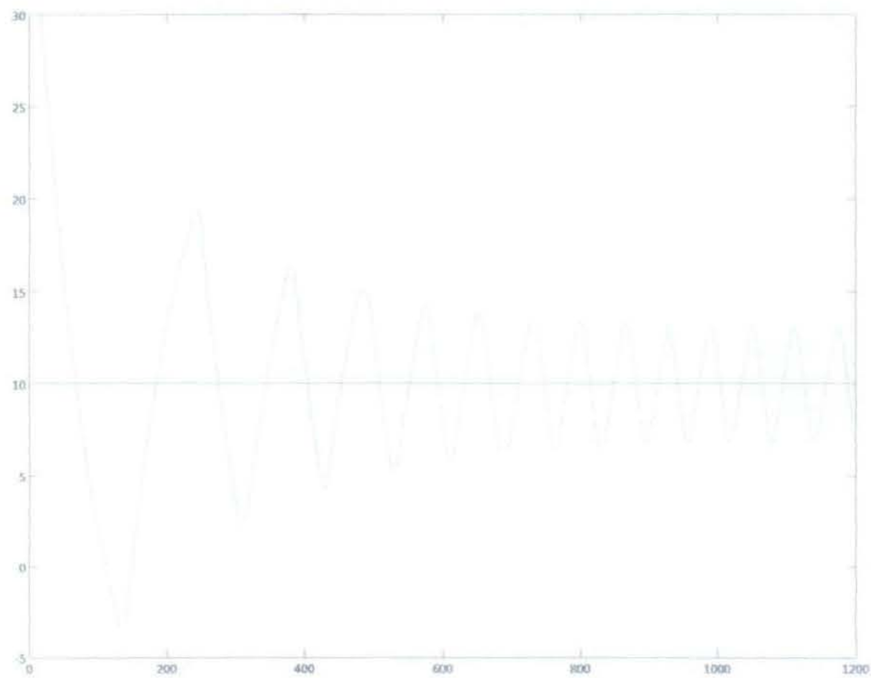

Figure 30: The response for PID controller of Ziegler Nichols Open Loop tuning Method for temperature.

The controlled variable continues to oscillate when about the setpoint when a change is done. This is true for all cases of controller modes, as can be seen in Figures 28, 29 and 30, the response of the system shows that it is unstable; concluding that the Ziegler Nichols Open loop method is unsuitable for this particular application.

4.2.8 Cascade Performance evaluation

The performances of each controller using different tuning methods are now evaluated so as to determine which tuning method yields the best performance. The system that shows a stable response is compared, and each control performance criteria is tabulated.

Table 13: Response Performance analysis for Cascade Control

Tuning Method	Controller Type	Control Performance			
		Rise time (s)	Settling Time (s)	Decay Ratio	CV overshoot (celcius)
Ziegler Nichols Closed loop	P only controller	88	N/A	0.7	14
	PI controller	30	285s	0.16	11
	PID controller	**56**	**136**	**0**	**8**
Cohen Coon	P only controller	75	475	0.679	13
	PI controller	76	476	0.68	12
	PID controller	75	477	0.68	13
Ziegler Nichols Open loop	P only controller	75	473	0.7	12
	PI controller	73	470	0.65	13
	PID controller	75	472	0.69	13

From Table 11, we can see that for the primary loop, a PID controller using the Ziegler Nichols closed loop tuning yields the best performance. The response has the fastest settling time and has a 0 decay ratio.

Based on the performance analysis, the tuning parameters that yields the best performance is used in the cascade architecture. The primary (temperature) loop will use the Ziegler Nichols Closed loop tuned parameters, while the secondary (flow) loop will use the Cohen Coon tuned parameters.

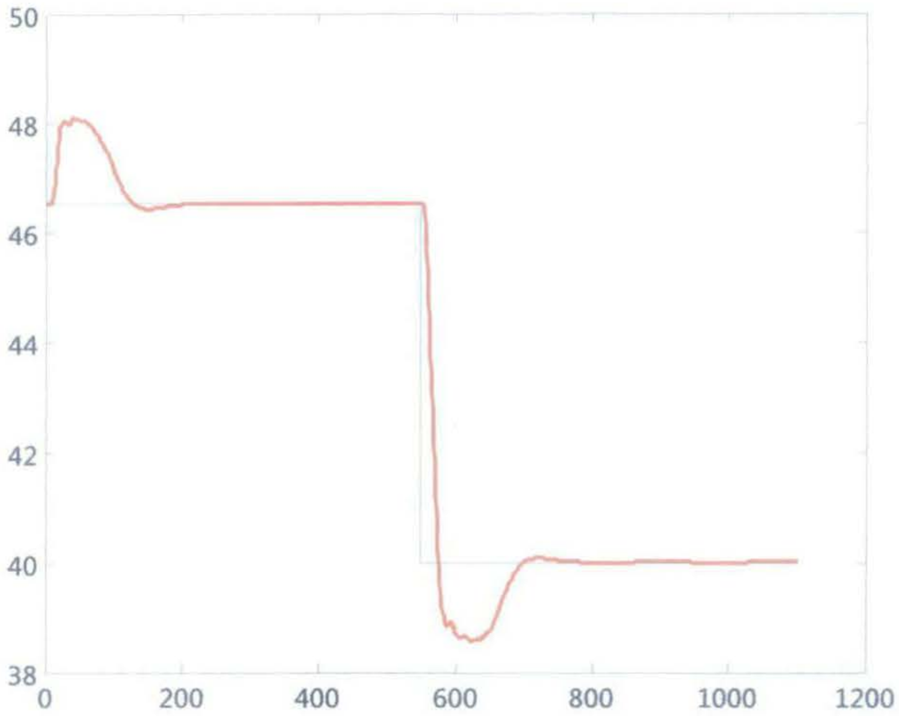

Figure 31: Cascade response of to a step change from 46.5 to 40

The cascade architecture is then tested with different input variation to see how it performs. This means varying on how the setpoint input change is applied to the cascade architecture.

Set point can be defined as the desired value for an operation variable. Set point is rarely changed when dealing with continuous production with the same condition. However, for batch operations, the set point may need to constantly change. So, we now observe the controller performance when it is subjected to varying setpoint values.

Figure 31: Cascade response to varying setpoint changes

Based on the result, the controlled variable is seen to respond to each setpoint variations. Another type of input is the linear ramp input. The input of the controller is increased by 2 celcius over a certain period both increasing and decreasing. A good controller should response fast to this type of input.

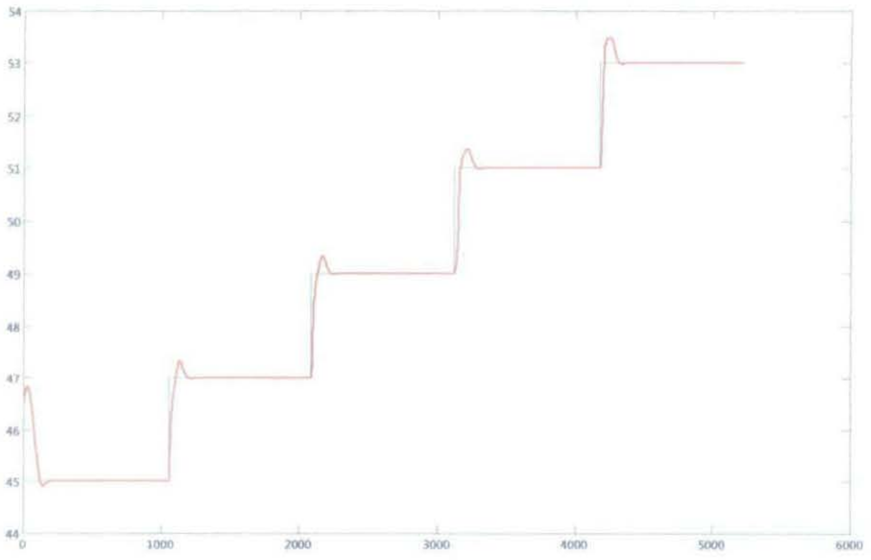

Figure 32: Cascade response to an increasing ramp input

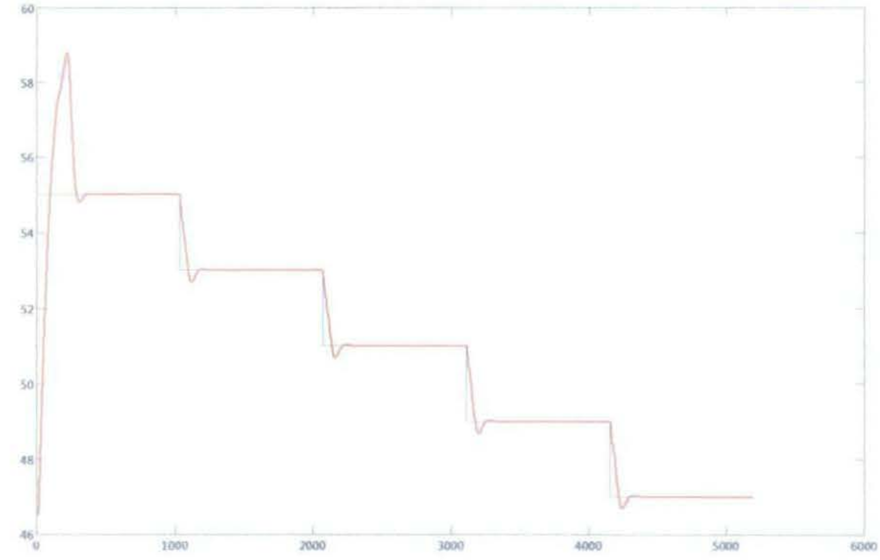

Figure 33: Cascade response to a decreasing ramp input

CHAPTER 5

CONCLUSION AND RECCOMENDATION

5.1 CONCLUSION

PID control is a very common approach used in industrial control. The PID controller is designed based on the calculation of the process model and plant parameter obtained from the experiment. The process model and plant parameters are also useful in calculating the tuning coefficients.

In calculating the first order plus dead time model of the plant, it is observed that Method II yields a more accurate result. This is especially true in cases where the process reaction curve obtained from plant experimentation contains a high frequency of noise. It should be noted that for cases where noise is apparent in the curve, Method II should be applied to obtain the plant model.

Computer simulation was conducted to obtain to enable observation and evaluation of controller performance using different tuning coefficients. The Cohen Coon tuning method resulted in the best tuning parameters for the flow loop. Similarly with the temperature loop, the Ziegler Nichols open loop method yields an unsatisfactory controller response; large settling time and rise time.

The Ziegler Nichols Closed loop tuning is proven to be the best method to calculate the PID parameters for the temperature loop. The Ziegler Nichols open loop tuning does not yield any satisfactory controller performance for the temperature loop.

The cascade PID control scheme indicates that it is able to perform well in controlling of temperature of the gas medium in the Gaseous Pilot Plant.

5.2 RECCOMENDATIONS

The approach presented in this project can be further improved as follows:

- Investigation of a detail study on optimizing the PID parameter of the primary and secondary loop.
- Implementation of intelligent system together with PID to form a hybrid controller such as Fuzzy-PI, and investigate if the implementation of such controller is worthwhile.
- Study the effects of different combination of tuning methods on cascade output response. Specify the best combination for temperature control in a gas plant.

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APPENDIX

Ziegler Nichols Closed loop Tuning Method based on values of ultimate gain,Ku, and ultimate period, Pu.

	Kc	Ti	Td
P only	Ku/2	-	-
PI controller	Ku/2.2	Pu/1.2	-
PID controller	Ku/1.7	Pu/2.0	Pu/8

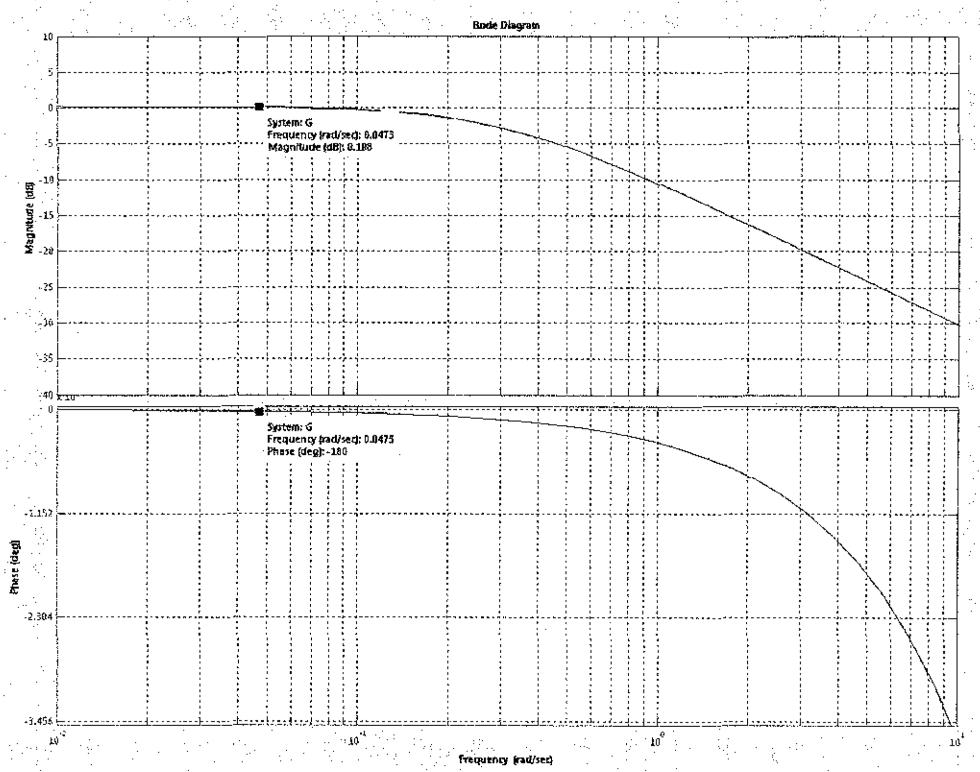
Ziegler Nichols Open loop Tuning Method formula

	Kc	Ti	Td
P only	$(1/K_p)/(\tau/\theta)$	-	-
PI controller	$(0.9/K_p)/(\tau/\theta)$	3.3θ	-
PID controller	$(1.2/K_p)/(\tau/\theta)$	2.0θ	0.5θ

Cohen Coon Tuning Method formula

	Kc	Ti	Td
P only	$(1/K_p)(\tau/\theta)(1+(\theta/3\tau))$	-	-
PI controller	$(1/K_p)(\tau/\theta)(0.9+(\theta/12\tau))$	$[\theta(30+3(\theta/\tau))]/(9+20(\theta/\tau)$	-
PID controller	$(1/K_p)(\tau/\theta)((3\theta+16\tau)/12\tau)$	$[\theta(32+6(\theta/\tau))]/(13+8(\theta/\tau)$	$(4\theta)/(11+2(\theta/\tau)$

Ziegler Nichols Closed loop bode plot calculations to obtain the ultimate gain, Ku, and ultimate period, Pu, for primary (temperature) loop. Bode plot obtained through MATLAB/simulink:



Given that

$$Ku = \frac{1}{10^{\frac{magnitude\,(db)}{20}}}$$

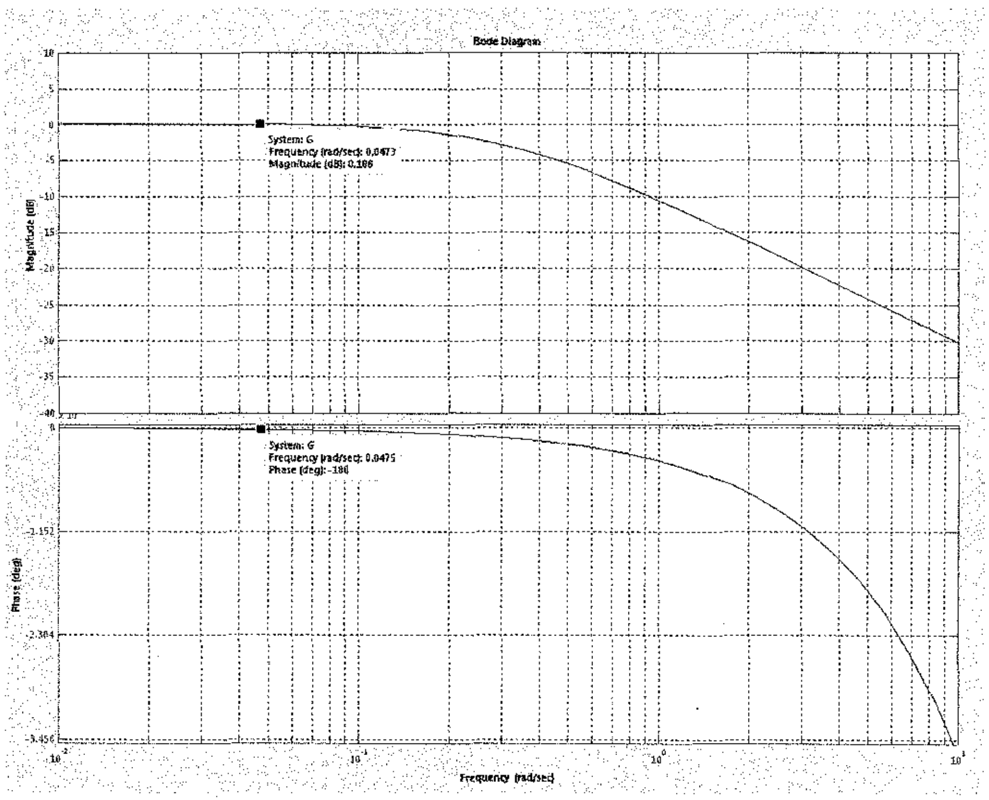
$$Pu = \frac{2\pi}{freq\ (rad/sec)}$$

Thus from the bode plot, the magnitude is -25.8dB at a frequency of 6.1 rad/sec. The calculated values for Ku, and Pu, are:

$$Ku = \frac{1}{10^{\frac{-25.8}{20}}} = 19.5$$

$$Pu = \frac{2\pi}{6.1} = 1.03$$

Ziegler Nichols Closed loop bode plot calculations to obtain the ultimate gain, Ku, and ultimate period, Pu, for secondary (flow) loop. Bode plot obtained through MATLAB/simulink:



Given that

$$Ku = \frac{1}{10^{\frac{magnitude\,(db)}{20}}}$$

$$Pu = \frac{2\pi}{freq\ (rad/sec)}$$

Thus from the bode plot, the magnitude is 0.107dB at a frequency of 0.0475 rad/sec. The calculated values for Ku, and Pu, are:

$$Ku = \frac{1}{10^{\frac{0.107}{20}}} = 1.012$$

$$Pu = \frac{2\pi}{0.0475} = 132.27$$