EVALUATION of CONTROL STRATEGIES of PRESSURE of GAS PLANT USING MATLAB Xpc-TARGET

By

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FINAL PROJECT REPORT

Submitted to the Electrical & Electronics Engineering Programme
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(Electrical & Electronics Engineering)

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CERTIFICATION OF APPROVAL

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Approved:

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UNIVERSITI TEKNOLOGI PETRONAS
TRONOH, PERAK

June 2007

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.

Mohd Syahrul Ridhwan B Zailani

ABSTRACT

Instrumentation and control engineers are often responsible for the operation of process control such as chemical and gas processes. As these processes and controllers are becoming more complex, the role of process automation becomes more and more important. For the monitoring and controlling of an industrial plant process, a proper controller needs to be design to achieve a good strategy.

The main objective of this project is to study a plant behavior and to optimize its performance by simulating it using several types of control strategies.

The first task is to model the pressure plant using an empirical modeling method. The purpose is to produce the plant transfer function that can be used for pressure controller analysis. By having the model, there is an alternative way to obtain forecasted data and result for use in this study.

The second task is to reconstruct a MATLAB / Simulink model. The Simulink model will be interfaced with the plant via an Xpc Target card to enable real-time analysis of the control strategies.

The third task is to analyze the model pressure controller performance using different ways of tuning method. Different tuning techniques give similar but not identical result. Five tuning method in open loop and three closed loop tuning method are being compared, namely open loop and closed loop Ziegler – Nicholas, Cohen-Coon, Fertik, Giancone and Lopez and Tyreus-Luyben and Ziegler-Nichols Bode plot methods are compared. The tuning methods were analyzed to yield the best result for the control valve. Finally the data of the PI controller from each of the tuning method is fed into closed loop Matlab and also to the Simulink Xpc-Target for the real-time analysis and comparisons of the control strategies performance.

ACKNOWLEDGEMENTS

In the name of ALLAH, the Most Graceful and Most Merciful,

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Last but not least, my appreciation goes out the individuals or group that have helped me in any possible way to complete this project.

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ABBREVIATIONS AND NOMENCLATURES

PID	Proportional, Integral and Derivatives
GUI	Graphical User Interface
PT	Pressure Transmitter
PIC	Pressure Indicator Controller
FC	Flow Transmitter
P	Pump
VE	Vessel
UTP	Universiti Teknologi PETRONOAS
SP	Set Point
PV	Process Variable
MV	Manipulated Variable
CV	Control Variable

INTRODUCTION

CHAPTER 1
INTRODUCTION

1.1 Background of Study

In UTP, there are several pilot plants specifically built to study process control The pilot plant closely resembles the actual plant with all the transmitter and control valves but in laboratory-scale. The gaseous pressure plant in the process control laboratory in the Electrical and Electronics (EE) Department of UTP will be used to study and apply real control via MATLAB / Simulink. The main component of this plant is pressure vessel and pressure transmitter.

In general, this project aims at modeling and simulation of pressure pilot plant. There are various methods that can be used to model the plant for example neural network, system identification and empirical method. In this project, the modeling part is done based on the pressure plant input and outputs. The simulation will involve with validation and testing the functionality of the model.

In process control, the basic objective is to regulate the value of some quantity. To regulate means to maintain that quantity at some desired value regardless of external influence. The desired value is called the reference value or set-point. After designing stage has done, the system need to be simulate to check the functionality and the convenience of the system. MATLAB / Simulink is a tool that will be used in this project. In order to perform real-time rapid prototyping, the Simulink need to be connected to xPC Target. This will provide a high-performance, host-target prototyping environment that enables the connecting Simulink and State-flow models to physical systems and execute them in real time on PC-compatible hardware.

By using the model developed, the pressure controller performance will be analyzed. Existing PID controller will be tested and compared to new PID with different of tuning method such as Ziegler-Nichols, Cohen-Coon and Lopez. The outcome from this analysis can be used to optimize the pressure plant performance.

1.2 Problem Statement

Currently, there are no models in MATLAB / Simulink of the pressure pilot plant that can be used for student to analyze its overall performance in real time. Modeling here refers to the process of analysis and synthesis at arriving to a suitable mathematical description of the plant parameter.

This project will be focusing on development of the model block and its controller that can be used for further analysis particularly for optimizing the pressure process. The purpose of this model is to evaluate the system and predict its output in real-time.

In the process control lab, the plant that is to be used in this study is used mostly for pressure control experiment using existing PID controller. There are several occasions where the pressure controller performance exhibit poor result. Therefore, the PID controller analysis and feasibility study of new controller implementation in real-time using MATLAB / Simulink is required. Perhaps the control strategy can be improved and smooth operation can be ensured.

1.3 Objective and Scope of Study

1.3.1 Objective

- To model and simulate the pressure pilot plant.
- To test the existing PID controller performance.
- To improve performance by using others tuning method.

1.3.2 Scope of Study

The modeling and simulation will be done on UTP's pressure pilot plant. The study will be based on input and output of the pressure together with its controller action. The modeling part is mostly involving empirical modeling approach.

The accuracy of the model will be observed based on its output reaction to input reaction. To validate the model, MATLAB Simulink is used to simulate its behavior based on certain input variation. Comparison is made on collected real-time data from plant experiment. The controller design will be implemented using Simulink specifically the Xpc-Target toolbox. Once the model of the controller is completed, the performance will be tested and compared. The criteria to measure such as the performance of reaching steady state, the rise and settling time, overshoot percentage and peak amplitude

LITERATURE REVIEW

CHAPTER 2
LITERATURE REVIEW

2.1 Process Flow of Gaseous Pressure Pilot Plant

This study will be on a gas plant and the focus in on monitoring and simulating control of a pressure vessel. This one loop pressure plant consists of gas vessel, pressure controller, fail open diaphragm valve and pressure sensor. Appendix 1 shows the plant that involve on this project.

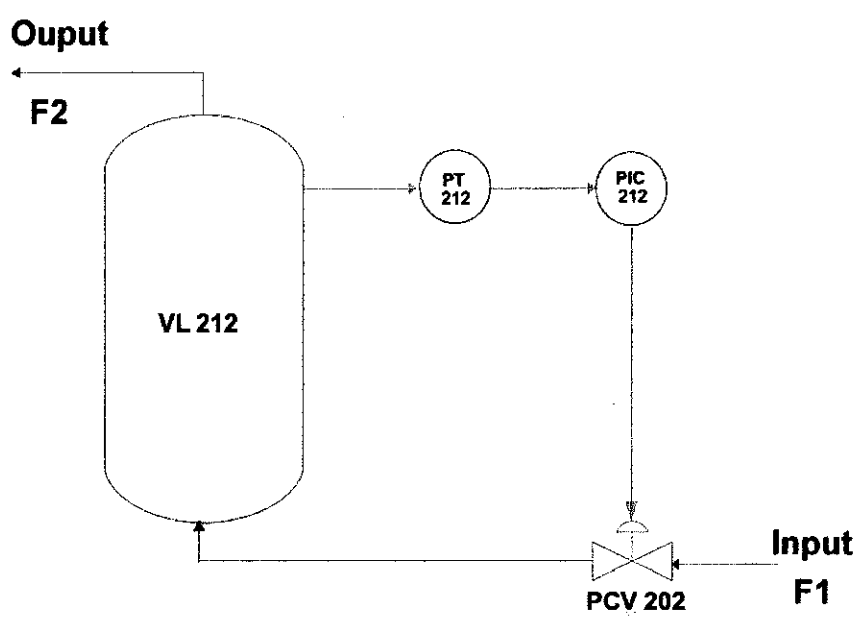

Figure 2.1: Simplified Pressure Flow Diagram for Gaseous Pressure Plant

Figure 2.1 shows the output regulates the flow rate F2 where F2 disturbance stream since the regulation of F2 is determined by another system. In this case F1 is the manipulated variable that is to be adjusted to manipulate a desired pressure.

The pressure transmitter (PT 212) sends the measured pressure with the desired pressure to the pressure controller (PIC 212). The PIC 212 compares the measured pressure with the desired pressure set-point and sends a signal to the valve (PCV202). This valve will step up or down for any change of the flow rate which will affect the pressure in the vessel.

In order to analyze the dynamic effect of feedback control loop, a control blocks diagram need to be developed. All dynamic elements in control loop are combined, using their laplace transfer function representation. Figure 2.2 shown control block diagram of gas pressure control. The closed-loop transfer function for figure 2.2 is

$$g_{cl}(s) = \frac{g_p(s)g_v(s)g_c(s)}{1 + g_p(s)g_v(s)g_c(s)g_m(s)}$$

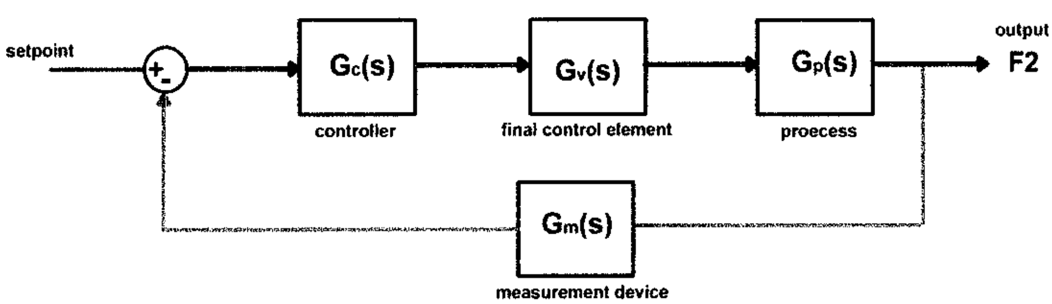

Figure2.2: Block diagram of the gaseous pressure plant

2.2 Empirical Model

The gaseous pressure empirical model is developed based on process reaction curve obtained from the experiment. Gaseous pressure transfer function can be expected to be a first order plus dead time model which represented by transfer function relationship

$$y(s) = \frac{k_p e^{-\theta s}}{\tau_p s + 1} u(s)$$

has the following output response to a step input change,

$$t \geq \theta, y(t) = k_p \Delta u\{1 - \exp(-(t-\theta)/\tau_p)\}$$

where the measured output is in deviation variable form. The three process parameters can be estimated by performing a single test on the process input. The gin is found as simply the long-term change in process output divided by the change in process input. Also the time delay is the amount of time, after the input change, before a significant output response is observed.

2.3 Process Reaction Curve

The purpose of using process reaction curve is to identify the dynamic model which will be used on the first-order with dead time model. The process reaction curve method involves the following four actions:

1. Allow the process to reach steady state
2. Introduce a single step change in the input variable
3. Collect input and output response data until the process again reaches steady state.
4. Perform the graphical process reaction curve calculations

Based on Ziegler and Nichols (1942) method, there are two different graphical technique can be used[5].

2.3.1 Method 1

The intermediate values determined from graph are

- The magnitude of the input change, δ;
- The magnitude of the steady state change in the output, Δ
- The maximum slope of the output-versus-time plot, S
- Intercept of maximum slope with initial values, θ

The values from the plot can be related to the model parameter according to the following relationship for first-order-with-dead-time model. The general method for a step in the input with $t \geq \theta$ is

$$Y'(t) = Kp\delta[1 - e^{-(t-\theta)/t}]$$

The slope for this response at any time $t \geq \theta$ is

$$\frac{dY'(T)}{dt} = \frac{d}{dt}\{K_p\delta[1 - e^{-(t-\theta)/\tau}]\} = \frac{\Delta}{\tau}e^{-(t-\theta)/\tau}$$

The maximum slope for this slope occurs at $t = \theta$, so $s = \Delta/\tau$. Thus the model parameters can be calculated as

$$Kp = \Delta/\delta$$
$$\tau = \Delta/s$$
$$\theta = \text{intercept of maximum slope with initial value}$$

2.3.2 Method II

Method II involve on times at which the output reaches 28 and 63 percent of its final value. Any two values of time can be selected to determine the unknown parameters θ and τ. The typical times are selected where the transient response is changing rapidly so that the model parameters can be accurately determined in spite of measurement noise. Thus the values of time at which the output reaches 28 and 63 percent of its final value are used to calculate the model parameter.

$$t_{28\%} = \theta + \frac{\tau}{3} \qquad t_{63\%} = \theta + \tau$$

$$\tau = 1.5(t_{63\%} - t_{28\%}) \qquad \theta = t_{63\%} - \tau$$

Others intermediate value determined form the graph are:

1. Magnitude of the input changes, δ
2. Magnitude of the steady-state change in the output, Δ

2.4 Open Loop Test

In open loop control or manual operation, the measured values of process variable are displayed to the operator, who has ability to manipulate the final control element (valve) by making an adjustment in the control room to a signal that is transmitted to a valve. There will be no comparisons and adjustment to regulate the process. The purpose of open loop is to evaluate the process reaction curve for identifying dynamic models. The process reaction curve method involves the following four actions:

- Allow the process to reach steady state
- Introduce a single step change in the input variable

- Collect input and output response data until the process again reaches steady state
- Perform the graphical process reaction curve calculation.

The graphical calculations determine the parameters for a first-order-with dead-time model: the process reaction curve is restricted to this model. The form of the model is as follows with X(s) denoting the input and Y(s) denoting the output which expressed in deviation variables:

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

2.5 Closed Loop Test

The closed loop control system provides a form of feedback to the process under control. A process is measured, compared to a set point and a final control element is adjusted accordingly. In process environment that has an established set point and control algorithm in controller, a change in the process load will determine if corrective action is needed. The controller which provides the excitation for the plant was referred to PID to control the overall system behavior

2.6 PID Controller

A proportional-integral-derivative controller (PID controller) is a common feedback loop component in an industrial system. The controller compares a measured value from a process with a reference set point value. The difference or "error" signal is used to calculate a new value for a manipulatable input to the process that brings the process' measured value back to its desired set point. The PID controller can adjust process outputs based on the history and rate of change of the error signal, which gives more accurate and stable control. Table 1.1 below shows the effect of the each controller.

- Kp = Proportional gain
- KI = Integral gain
- Kd = Derivative gain

Proportional – A proportional controller (Kp) will have the effect of reducing the rise time and will reduce, but never eliminate, the steady-state error.

Integral - An integral control (Ki) will have the effect of eliminating the steady state error, but it may take the transient response worse

Derivative - A derivative control (Kd) will have the effect of increasing the stability of the system, reducing the overshoot, and improving the transient response.

Table1.1: Effect of each controller

Controller Respond	**Rise Time**	**Overshoot**	**Settling Time**	**SS Error**
Kp	Decrease	Increase	Small Change	Decrease
Ki	Decrease	Increase	Increase	Eliminate
Kd	Small Change	Decrease	Decrease	Small Change

METHODOLOGY

CHAPTER 3
METHODOLOGY

3.1 Procedure Identification

There are several procedures that have been identified for the execution of this project. First is overall flow chart that describes the implementation of the project. Next is the methodology on designing the plant control with involves empirical modeling design.

3.1.1 Overall Project Flow chart

On early stage, the literature review on the gaseous pilot plant and PID controller has been done. It is to get better understanding of the pressure control at the gaseous pilot plant. The research covers on familiarize of the component that involve on the simulation, understand the characteristic of the PID, function and effect of each controller elements.

In order to evaluate the PID performance a simple model of closed and open loop has been develop using the MATLAB / Simulink. Basically, the research and literature review stage has been finished before the middle of the first semester.

Figure 3.1 shows the implementation steps for modeling and simulation of gaseous pressure project. It consists both modeling and simulation exercise.

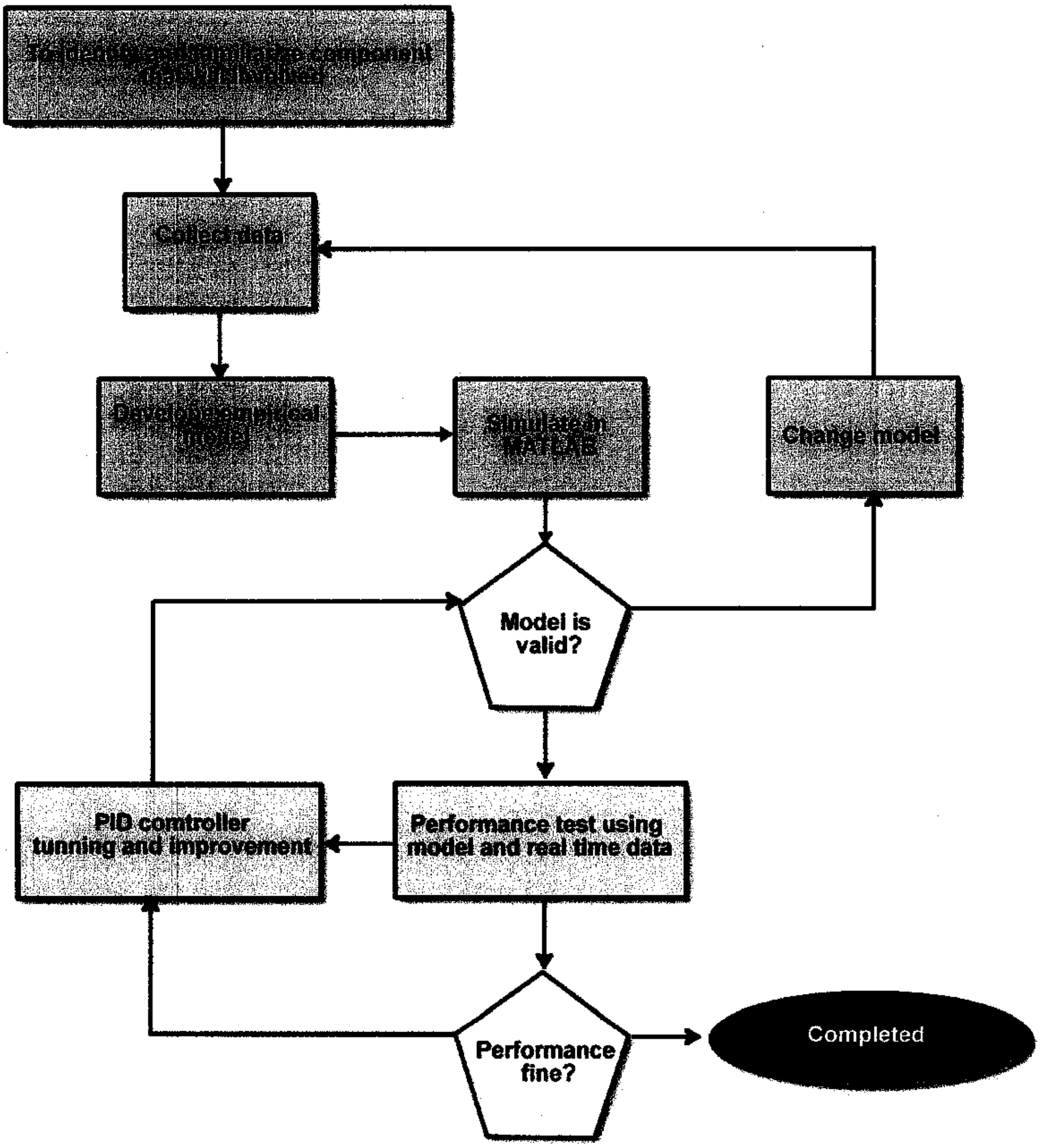

Figure 3.1: Overall project flow chart

3.1.2 Development of Control Strategy

The development of control strategy consists of formulating or identifying the control problem. Figure 3.2 shows the step in formulating a control problem.

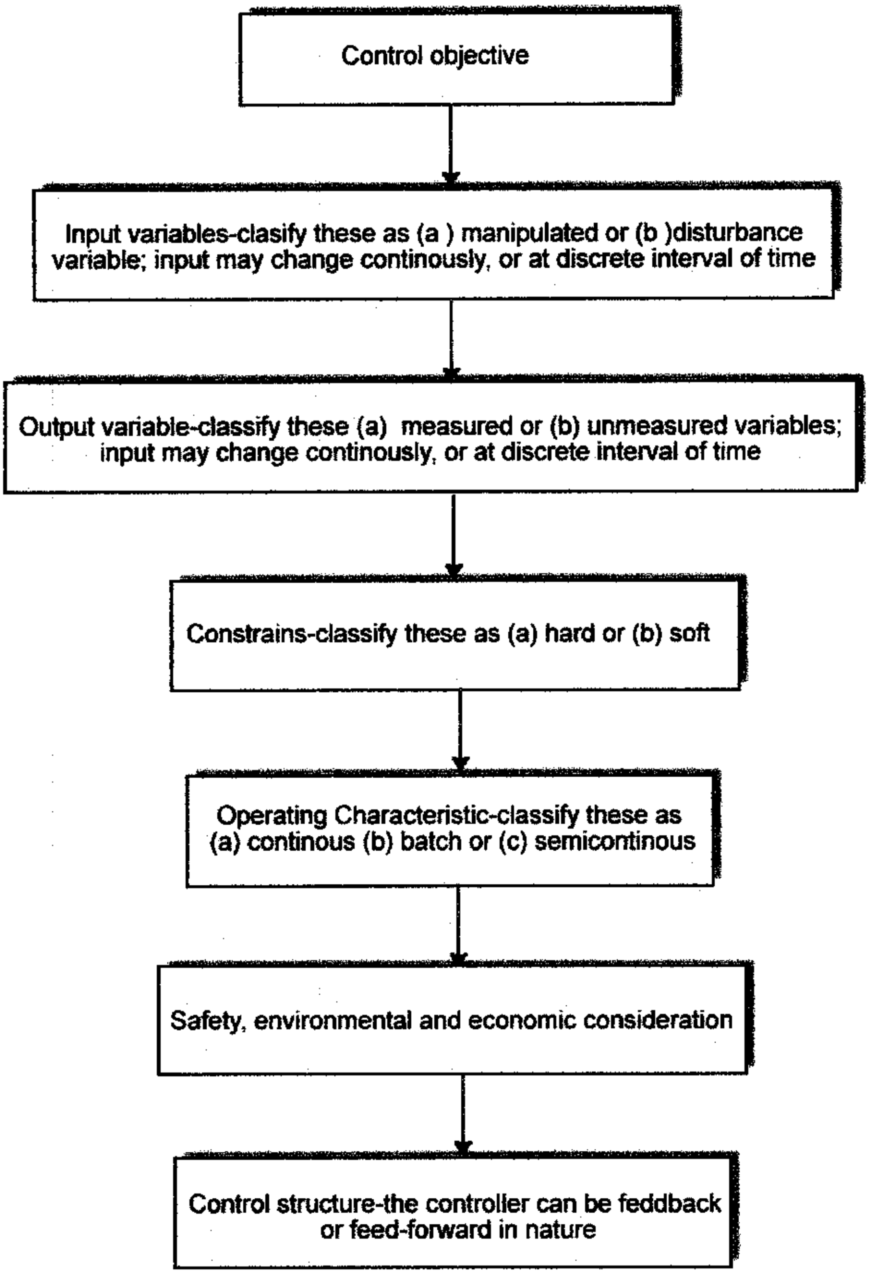

Figure3.2: Steps in formulating a control problem

Vessel tanks are often used as intermediate storage for fluid streams to be transferred between several process units. Consider the process flow diagram show in Appendix 2 where a pressure stream from process 1 is fed to the vessel tank and the effluent from the surge tanks is send to process 2.

There are obvious constrains on the height in the vessel. If the vessel is overpressure, it may create safety and environment hazards, which may also have ergonomic significant. Step-by-step procedure on this analysis will be discussing bellow.

Control Objective: The control objective is to maintain the pressure within certain bounds. If it is too high, it will overflow and if it is too low there will be problem with the flow to process 2. Usually a specific desired pressure will be selected. This desired pressure is known as the set point.

Input Variable: The input variable is the flow from process 1 and the flow to process 2. Notice that an outlet flow rate is considered input to this problem.

Output variable: The most important output variable is the pressure level which is measured.

Constrains: There are a number of constrains in this problem. There is a maximum pressure level, if this exceeded, the vessel will be overpressure, There are maximum flow rate through the inlet and outlet valves

Operating Characteristic: Assume that this is a continuous process which will be a continuous flow in and out. It would be semiconscious process for example there was an inlet with no outlet flow (if the tank was simply being filled)

Safety, Environment and Economic Considerations: These aspects depend somewhat on the gas characteristics. If is is a hazardous chemical, then the tremendous incentive from safety and environment consideration is not to allow the tank to overflow. Indeed this is also an economic consideration , since injuries to employment or environment cleanup costs money

Control Structure: There are numerous possibilities of controlling of this system. The gaseous pressure empirical model is developed based on process reactions curve obtained from the pilot plant. Based on the process reaction curve, there are two graphical methods that can be used to calculate the transfer function parameters.

3.1.3 Empirical Modeling

There are six steps for developing empirical model of a system. The steps are shown in Figure 3.3

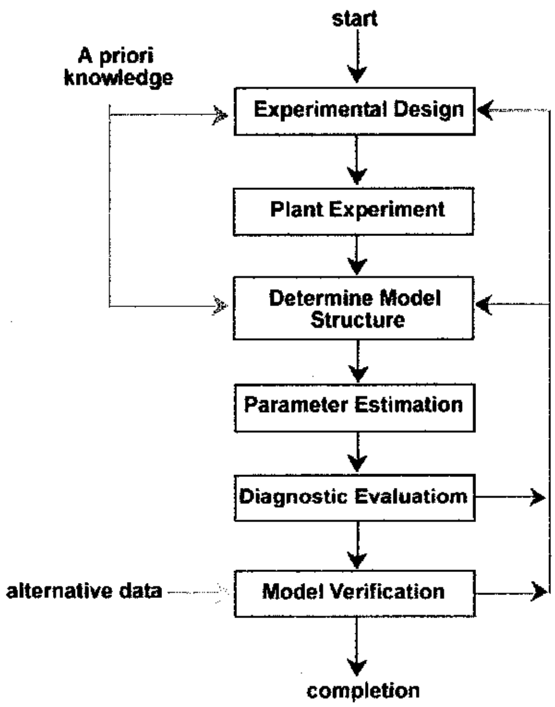

Figure 3.3 Steps in formulating a control problem

For a start proper experimental design is required so that its shape, duration and base operating condition can be determined. In plant experiment, it should be executed as close to the experiment design plant as possible. To determine model structure, many methods are available but initial structure is selected based on prior knowledge.

For parameter estimation, two methods can be used; a graphical techniques or statistical principle. The diagnostic level of evaluation determines how well the model fits the date used for parameter estimation. Lastly, the final check on the method is to verify by comparison with additional data not used in parameter estimation.

Method 1 concerns with the value of maximum slope of the process reaction curve. Because of difficulty in getting the best maximum slope, Method 1 usually will produce inconsistent and imprecise result hence it has large errors in parameter estimation. Method II concerns with the time at which the output reaches 28% and 63% of its final value. Thus it is more reliable and accurate. Therefore, Method II is used to determine all process parameter (e.g dead time, time constant and process gain.)

3.2 Tools and Software

3.2.1 MATLAB-Simulink

MATLAB offers array operations that allow one quick manipulated sets of data in wide variety of ways. MATLAB also offers programming features similar to those other computer programming language. In addition, MATLAB offers graphical user interface (GUI) tools to allow one to use it as an application development tool. Therefore this project will utilize most of MATLAB programming application and its development feature.

Simulink is an extension to MATLAB that allow engineers to rapidly and accurately build computer model of dynamic systems, using block diagram notation. With Simulink, it is easy to model complex nonlinear system.

Additionally, a Simulink model can be produce graphical animation that shows the progress of a simulation visually, significantly enhancing understanding of system behavior.

3.2.2 Gaseous Pressure Pilot Plant with xPC Target.

The pilot plant is am important equipment in this project. The element that are required for analysis are pressure transmitter, control valve and the gas vessel it self. xPC Target will perform data acquisition that is vital for modeling and analysis of data.

RESULTS AND DISCUSSION

CHAPTER 4

RESULT

4.1 Empirical Model

The pressure model is developed based on process reaction where the pressure function is estimated to be of first order plus dead time. Figure 4.1 shows the process reaction curve obtained from experiment. The step change is 10% and the pressure on the vessel increment is about 1.916bar. Both plots are used for pressure and flow transfer function calculation.

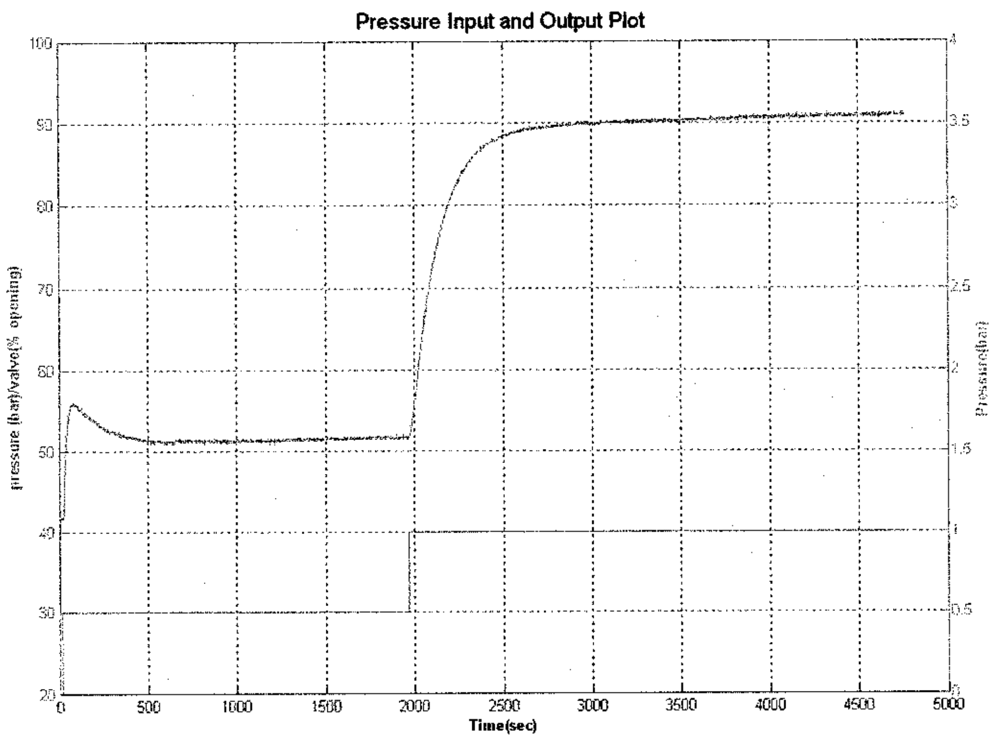

Figure 4.1: Pressure Input and Output Plot

There are two methods that can be used to calculate the transfer function parameters. Method 1 concerns with the value of the output-versus-time plot. Because of difficulty in evaluating the slope, Method 1 typically has larger error in the parameter estimates; thus, Method II is preferred. Method II concerns with the time at which the output reaches 28% and 63% of the final value. Appendix 3 and Appendix 4 shows the detail calculation on how to get transfer function from the process reaction curve. The simplified results of model parameters calculated are:

	Method I	Method II
Process Gain K_p	0.194bar/%opening	0.194bar/%opening
Time Constant τ	271 Seconds	160.5seconds
Time Delay θ	18 Seconds	15.5seconds

Hence the general first order plus dead time model transfer function for Method II is

$$\frac{Y(s)}{X(s)} = \frac{K_p e^{-\theta s}}{\tau s + 1} \rightarrow \frac{Y(s)}{X(s)} = \frac{0.194 e^{-15.5s}}{160.5s + 1}$$

4.1.1 Open Loop Test

Figure 4.2 shows the block diagram for the empirical modeling of pressure process model. The simulation is a step input that resembles valve opening

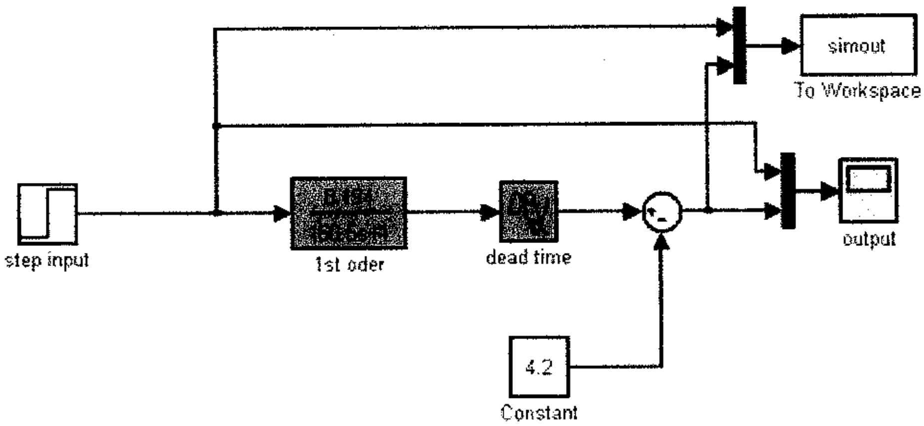

Figure 4.2: Simulink block diagram of pressure process model

The output from the model is as shown in Figure 4.3. From that curve, it can be observed that the pressure increment is about 1.91bar. The change is quite similar to reaction curve pressure change as shown in Figure 4.1. Therefore, this model is valid to resemble the actual pressure plant at pilot plant.

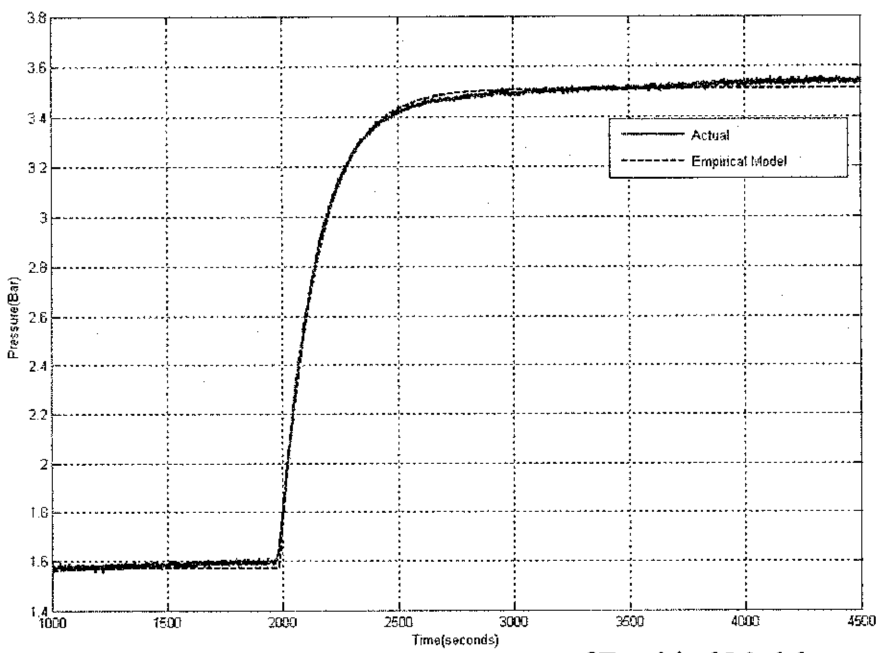

Figure 4.3: Open Loop Response of Empirical Model

4.2 Tuning Rules for first-Order + Dead Time Process

Appendix 5 shows the empirical model with PID controller. Using parameters obtained from the process reaction curve, the tuning coefficient for PID controller can be calculated by referring to Table 4.1 and Table 4.2. The formula shown is based on Ziegler-Nichols open loop method and Cohen – Coon method. Both method were empirically developed to yield a closed loop response which should give roughly quarter-wave damping.

Table 4.1: Ziegler-Nichols open loop tuning based on reaction curve

Controller	K_c	T_i	T_d
PI	$(\frac{1}{K_p})/(\frac{\tau}{\theta})$	3.3θ	–
PID	$(\frac{0.9}{K_p})/(\frac{\tau}{\theta})$	2.2θ	0.5θ

Table 4.2: Cohen-Coon open loop tuning based on reaction curve

Controller	K_c	T_i	T_d
PI	$\frac{\tau}{K_p\theta}\left[0.9+\frac{\theta}{12\tau}\right]$	$\theta\left[\frac{30+\frac{3\theta}{\tau}}{9+\frac{20\theta}{\tau}}\right]$	–
PID	$\frac{\tau}{K_p\theta}\left[\frac{4}{3}+\frac{\theta}{4\tau}\right]$	$\theta\left[\frac{32+\frac{6\theta}{\tau}}{13+\frac{8\theta}{\tau}}\right]$	$\frac{4\theta}{11+\frac{2\theta}{\tau}}$

In addition, there is another tuning parameter suggested by Fertik (1974) which emphasizes on minimum ITAE (Integral Absolute Error) which limit the overshoot. The Formula to calculate the PI tuning coefficient

$$K_c = 0.859(\theta / \tau)^{-0.977}$$

$$T_i = (\tau / 0.674)(\frac{\theta}{\tau})^{0.680}$$

By referring to Marlin[5], there are another two tuning method parameter calculations suggested by Giancone and Lopez. The detail calculation for this method is shown in Appendix 6. The tuning parameters are listed in Table 4.3.

Table 4.3: PI tuning parameters for Empirical Model

PI controller	Ziegler-Nichols	Cohen-Coon	Fertik	Giancone	Lopez
K_c	0.5	48.5	8.43	10.3	28.4
T_i	51.15	43	48.6	128.5	88

Figure 4.4 shows the PV response of all controllers after a step change in SP from 0bar to 5bar. In general, all of them settle at new set point mark but with various transient response behaviors. Cohen-Coon parameters are far more aggressive than Ziegler-Nichols since it has bigger controller gain value. However its MV behavior is too aggressive and not suitable for normal control valve. From the graph it shows also that the response of Giancone, Fertik and Lopez becomes sluggish. This is because this project involves fast response and the dead time was too small which will result bigger reset time and small controller gain. Potential problem for a system with low time-delay –constant ratio, since this causes the proportional gain to become very small. Similarly, the integral time tends to be high causing overdamped with large amount energy absorption.

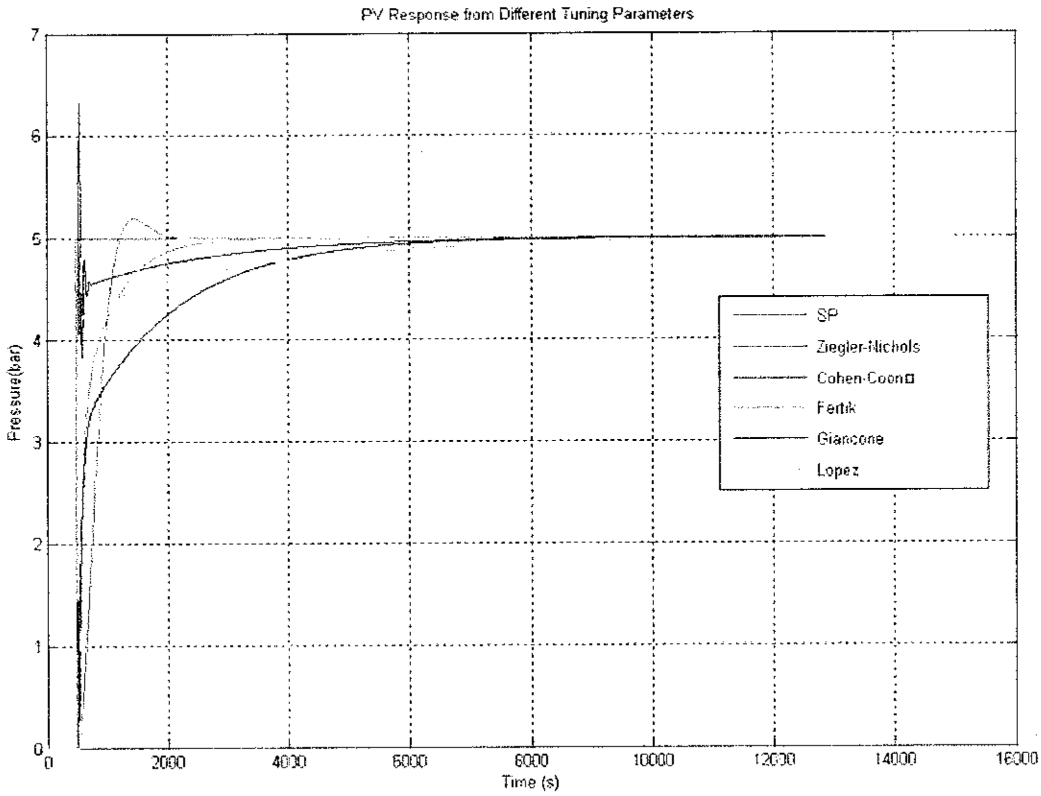

Figure 4.4: PV Response from Different Tuning Parameters

4.3 Closed-Loop Based Tuning

It is based on a simple closed loop experiment, using proportional control only. The proportional gain is increased until a sustained oscillation of the output occurs (which neither grows nor decays significantly with time).

With P=only closed-loop control-, the magnitude of the proportional gain will be increase until the closed-loop is in a continuous oscillation. The values of the proportional gain that causes the continuous oscillation is called the critical (or ultimate) gain k_{cu} .The peak-to-peak period (time between successive peaks in the continuously oscillating process output) is called the critical period P_u .Depending on the controller chosen, P, PI, and PID, use the values in Table 4.4 for the tuning parameter, based on the critical gain and period.

Table 4.4 Ziegler-Nichols Closed-Loop Oscillation Tuning Parameters

Controller	K_c	T_i	T_d
PI	$0.45k_{cu}$	$P_u/1.2$	–
PID	$0.6k_{cu}$	$P_u/2$	$Pu/8$

From the experiment, continues oscillation caused when the value of controller gain was k_{cu}= 40. From the response shown in Appendix 7, it shows that the ultimate period (period oscillation) is p_u = 0.644minutes.

Tyreus and Luyben modified the suggested Ziegler-Nichols parameters for increased robustness. Tyreus and Luyben have suggested tuning parameter rules in less oscillatory response and that are less sensitive to changes in the process condition. Their rules are shown in Table 4.5.

Table 4.5: Tyreus-Luyben Suggested Tuning Parameter Based on the Ziegler-Nichols Closed Loop Oscillation Tuning Method

Controller	K_c	T_i	T_d
PI	$k_{cu}/3.2$	$2.2P_u$	–
PID	$k_{cu}/2.2$	$2.2P_u$	$Pu/6.3$

Other tuning methods that can be used to determined the initial tuning for PI is [5] Ziegler-Nichols closed loop-Bode Plot method. This tuning method has two advantages:

1. It can be applied to process that are not well modified by first-order with dead-time models
2. It provides considerable insight into the effect of all loop elements (process, instrumentation and control algorithm) on stability and proper tuning constant values.

The system will be closed-loop stable if and only if the amplitude ratio is less that 1 at the crossover frequency.

$$AR_{co} = |g_c(\omega)g_p(\omega)| < 1$$

where ω_{co} is known as the crossover frequency which is defined as the frequency where the phase angle is -180^{o}. Ultimate gain k_u is the value of the proportional gain that brings the system to the boundary of stability at the critical frequency and the Ultimate period, P_u is the period of oscillatory of the system at the margin of stability.

$$k_u = \frac{1}{|G_{oL}(j\omega_c)|} = \frac{1}{AR_c}$$

$$P_u = \frac{2\pi}{\omega_c}$$

The Ziegler-Nichols closed-loop tuning correlation given in Table 4.6 need to be used to calculate the controller tuning constant values,. Figure 4.5 shows the Bode plot of the transfer function. Appendix 8 shows the detail calculation the tuning constant of PI controller from the Bode plot. From the Bode plot, the ultimate gain and period can be determined to be P_u = 59.8 and k_u = 49.8.

Table 4.6: Ziegler-Nichols closed-loop tuning correlations

Controller	K_c	T_i	T_d
PI	$k_u/2.2$	$P_u/1.2$	–
PID	$k_u/1.7$	$P_u/2.0$	$Pu/8$

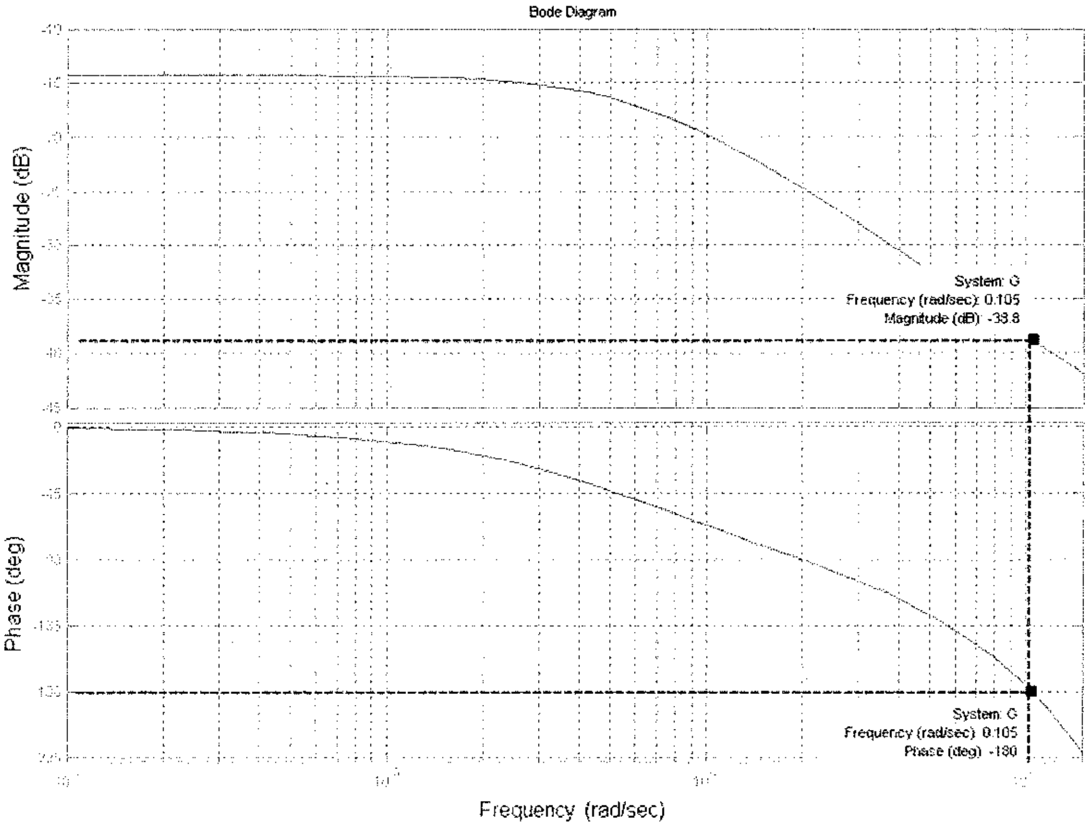

Figure 4.5: Bode plot of $G_{oL}(j\omega)$ for $K_p = 0.194, \tau = 160.5, \theta = 15.5$

Finally, all the tuning parameters are listed in Table 4.7.

Table 4.7: PI Tuning Parameters for Closed Loop Oscillation Based Tuning

Method		
Ziegler-Nichols	**Proportional**	18
	Integral	1.82
Tyreus-Luyben	**Proportional**	1.41
	Integral	12.5
Ziegler-Nichols closed loop-Bode Plot method	**Proportional**	39.6
	Integral	49.8

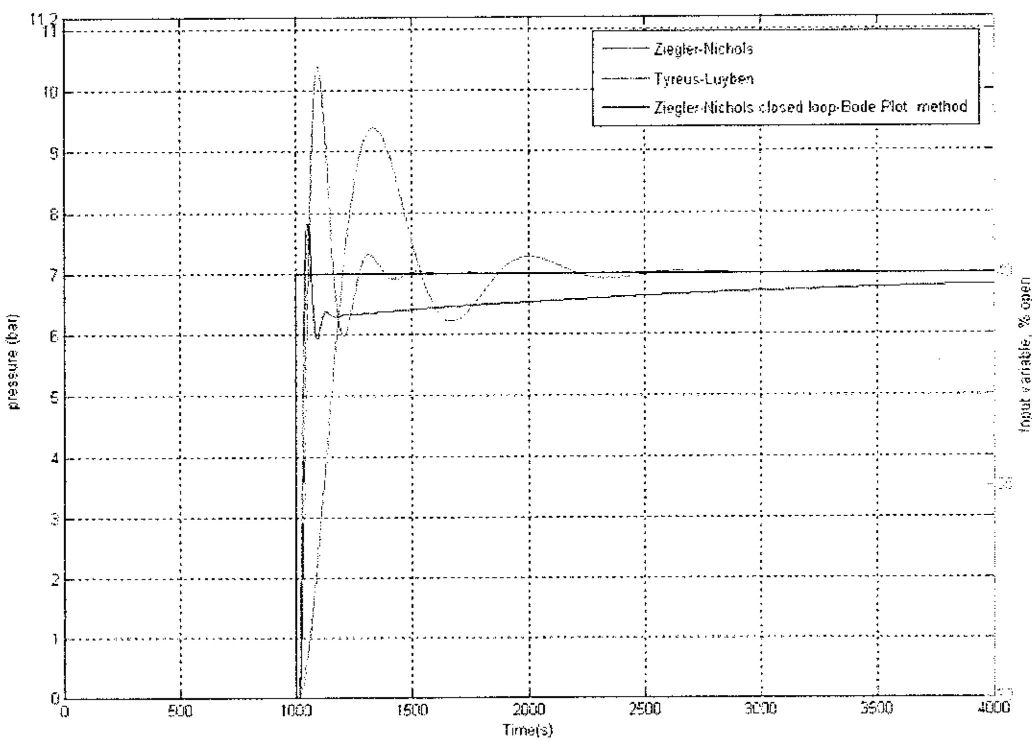

Figure 4.6: Response to unit Step set-point change. Comparison of Ziegler-Nichols, Tyreus-Luyben and Ziegler-Nichols closed loop-bode plot tuning method

The closed loop responses for Ziegler-Nichols, Tyreus-Luyben and Ziegler-Nichols closed loop-bode plot tuning method are shown in Figure 4.6. Notice that the Ziegler-Nichols closed-loop tuning method more aggressive with high overshoot but in a short settling time. It typically results in more oscillatory behavior that would be allowable in typical process plant.

The Tyreus-Luyben parameter results in less overshoot and gain which will be less sensitive to uncertainty. The Ziegler-Nichols closed loop-bode plot response was sluggish. This is because the proportional gain was become very large. Similarly, the integral time tends to be large causing oscillatory behavior on early response and it become overdamped.

4.4 Construction of MATLAB Simulink Xpc Target.

MATLAB / Simulink is a powerful software to for modeling, simulating and analyzing dynamical systems. In this project, it is use to design, tune, test and simulates the PID controller. The control block diagram is constructed in the Simulink. To create a Simulink model, it involves the selection of the necessary block such as step input, scope, xPC target scope block or output block based on the application needed. The suitable parameters are entered for the scope block to view the result. The Simulink output block is added to log the result for analysis. The xPC target is added to visualize signal while running the target application. In this project, the external simulation mode is chosen because the model is connected to the Gaseous Pilot Plant. Figure 4.7 shows the block diagram of MATLAB Simulink which connected to the xPC Target.

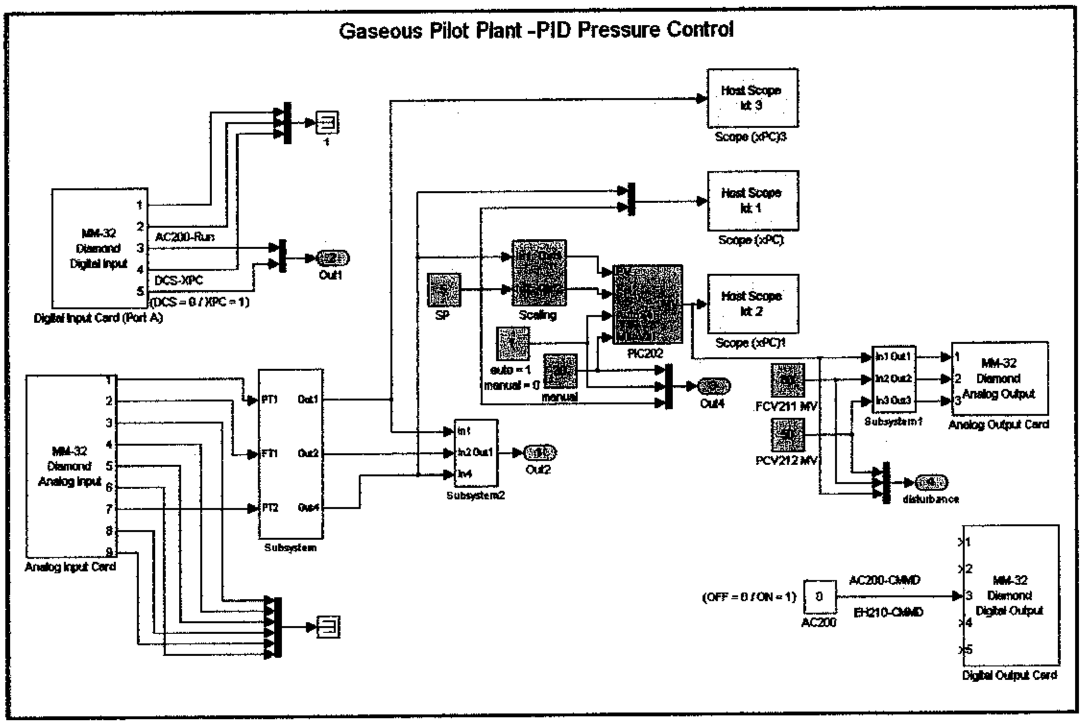

Figure 4.7: Block diagram of MATLAB / Simulink xPC Target

The xPC target industrial PC acts as server and interface system connecting the Simulink model to Gaseous Pilot Plant. The signal from Simulink model is written to the server via xPC target scope block. The xPC target kernel automatically creates the scope on the target when the target application is downloaded to the target PC. The software and hardware configuration of the xPC Target can be referred to Appendix 9.

In this project the controller running in real-time connected to a real plant. The xPC Target will validate the design in real-time without the need for custom target hardware. Figure 4.8 shows the block diagram of xPC Target box. The Diamond-MM-32-AT Frame Analog Input block executes the model in which it occurs each time it converts a new frame of data.

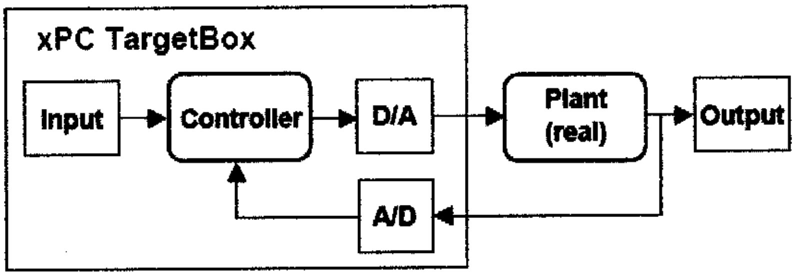

Figure 4.8: Block diagram of xPC Target box

4.4.1 Reconstruct the existing Simulink program.

The existing Simulink program needs to be rebuilt since some of it have an errors especially the connection of the component For example, the diamond Analog Input should connect to PT201 main vessel rather than PT202 the buffer vessel. In this project, the main objective is to control the main vessel in certain set point with in the pressure of 0-5 bar. Appendix 10 shows the new MATLAB / Simulink program.

On other hand, there was an error with the existing Simulink program on the PID itself. Supposedly in control system

Error = Setpoint - Process value

In this case the existing Simunlik program was using

Error = Process value - Setpoint

Figure 4.9 shows the existing PID and Figure 4.10 shows the new PID.

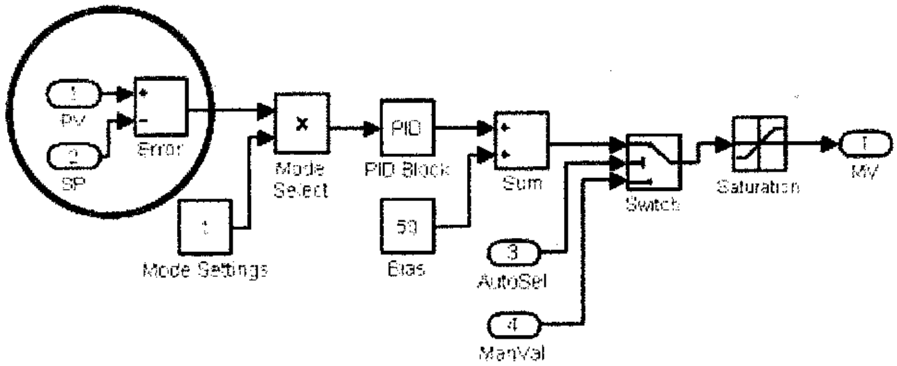

Figure 4.9: Default ideal PID using difference Error approach

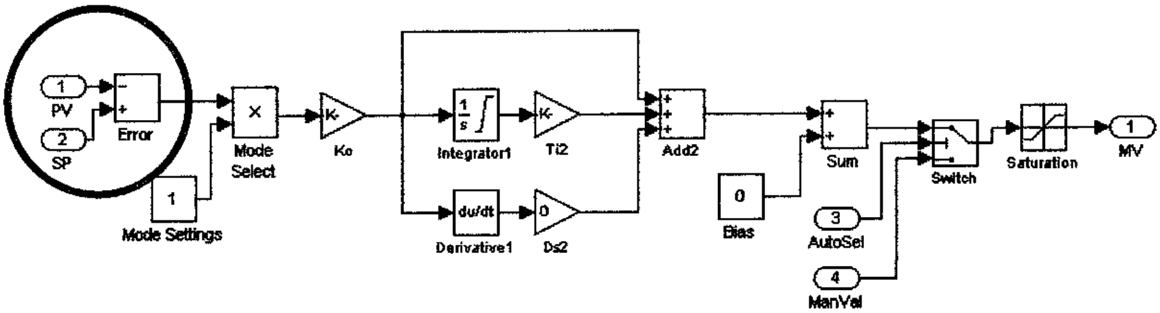

Figure 4.10: New Error approach

4.4.1.1 Developing Alternative Controller Icons for PID

The default PID controller icon was used in existing Simulink program was difference from ordinary PID. The default PID controller icon is shown in Figure 4.11. The structure of the default PID has the form of

$$g_c(s) = P + \frac{I}{s} + D_s$$

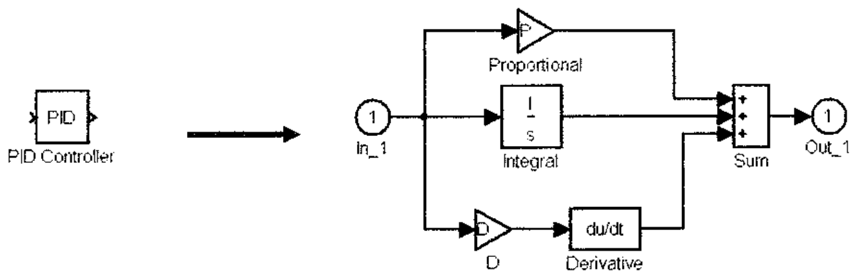

Figure 4.11: Default ideal PID controller

In other hand normally the preferred PID structure was

$$g_c(s) = k_c[1 + \frac{1}{\tau_I s} + \tau_D s]$$

The two algorithm are related by

$$P = k_c \qquad I = \frac{k_c}{\tau_I} \qquad D = k_c \tau_d$$

A new ideal analog PID is shown in Figure 4.8. Notice that there are two inputs to the controller, the set-points(r) and the measured output (y), rather than just the error signal that is to the default Simulink PID controller.

4.5 Performance Test Using On Line Simulation

The performance check is perform to compare the controller performance based on the calculated values. The controller performance check using online simulation to have the feedback control loop to maintains a small deviation between the controlled variable and the set point by adjusting the manipulated variable.

From a real time experiment, the real process response can be observed and analyzed. In the MATLAB block diagram, the PID controller parameters that have gained from the tuning parameters were assigned into the respective block. The controller mode is set to 1 that is for automatic controller which means that the PID will do the analysis and controlling the manipulated valve. From the tuning method, the PI value from the Ziegler-Nichols closed loop tuning method has been chosen. This is because the Ziegler-Nichols closed loop tuning method has been done using the plant scale it self which is difference with others tuning method.

Figure 4.12 shows the single set-point change from 5% opening to 5.5% opening. The response of this step change has shown in Figure 4.13. As can be seen, the control performance is approaching to the set point in short settling time. The control performance has shown bellow:

%overshoot	22.6%
Decay ratio	0.33
Rise time	15 seconds
Settling time	81 seconds

In order to maximize the performance check, the step change has been change from 3.5% opening to 6% opening with increment of 0.5. Figure 4.14 shows the set-point change is 0.5 bar for each 300seconds. Results shows that the process value approaching the set-point. The integral absolute error (IAE) is 0.1152. Notice that the process value was in good performance when it is approaching the set point at 4bar to 5.5bar. This is due to the set-point that is approaching to the maximum pressure on the vessel set at 6bar. Figure 4.15 shows the pressure response due to the set point change.

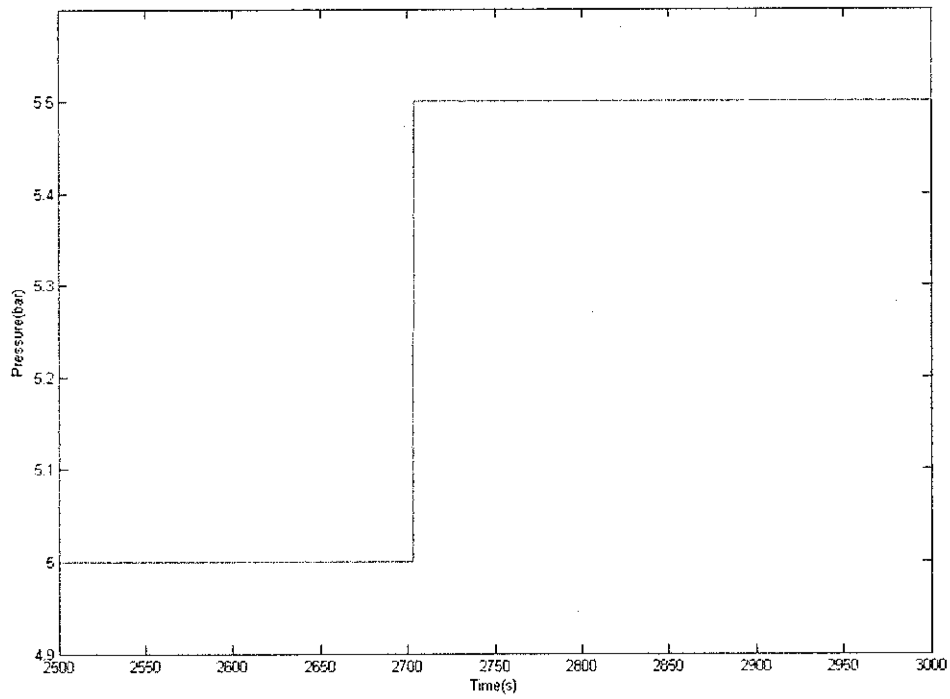

Figure 4.12: Single set-point change from 5% opening to 5.5% opening

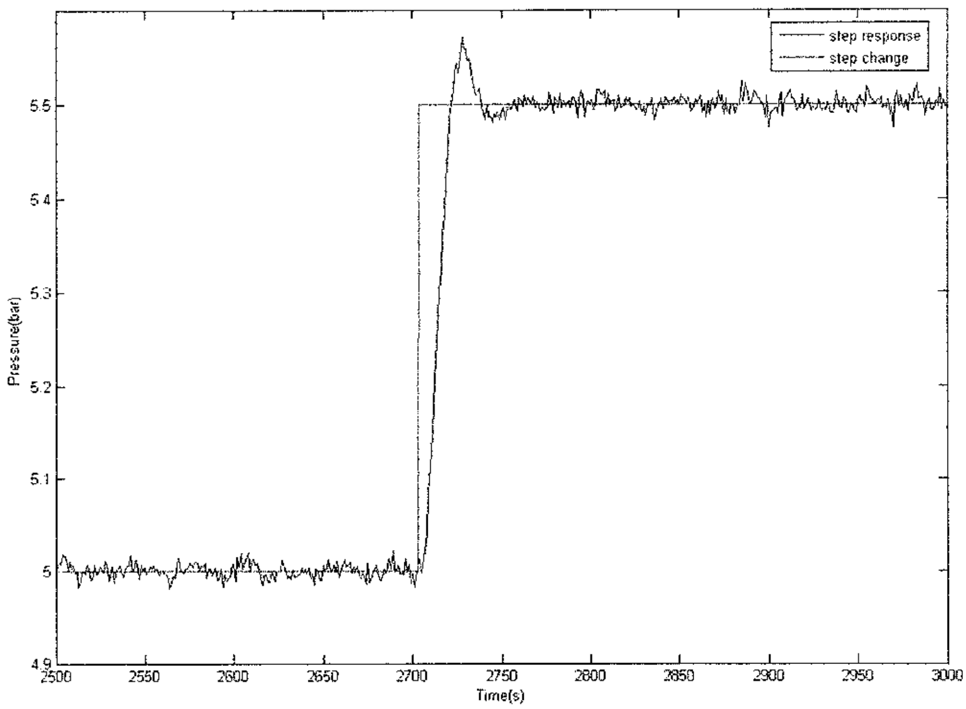

Figure 4.13: The response of this step change for single step change

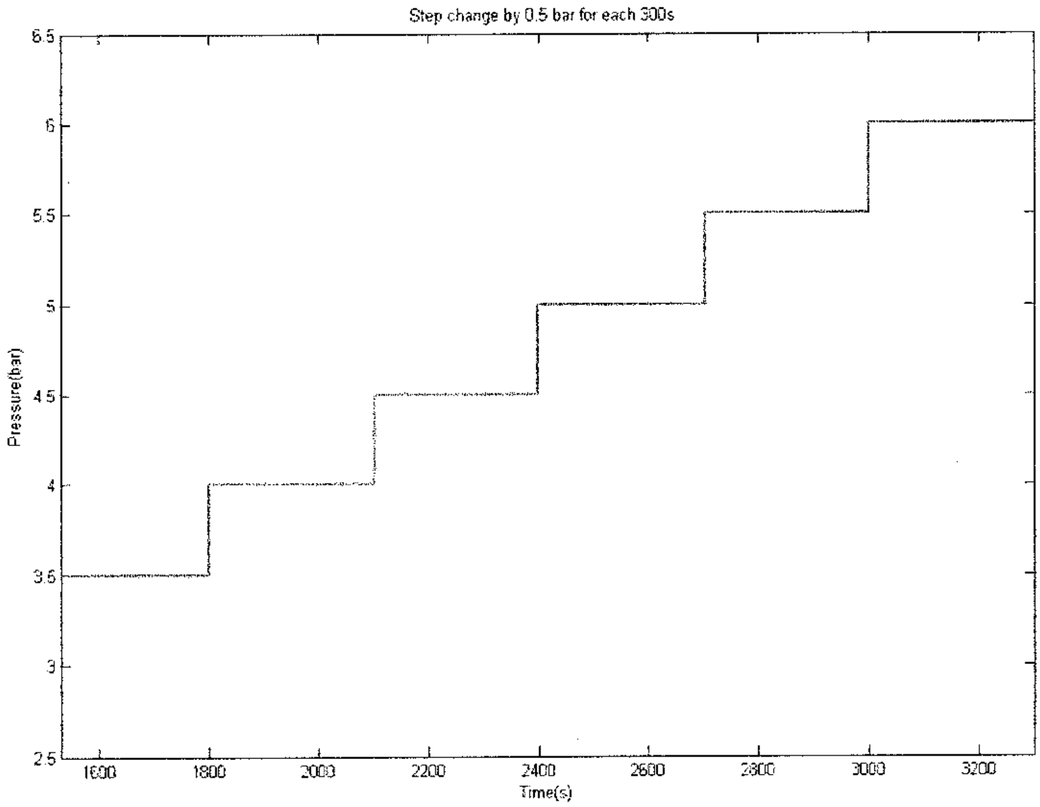

Figure 4.14: Step input for PID controlled

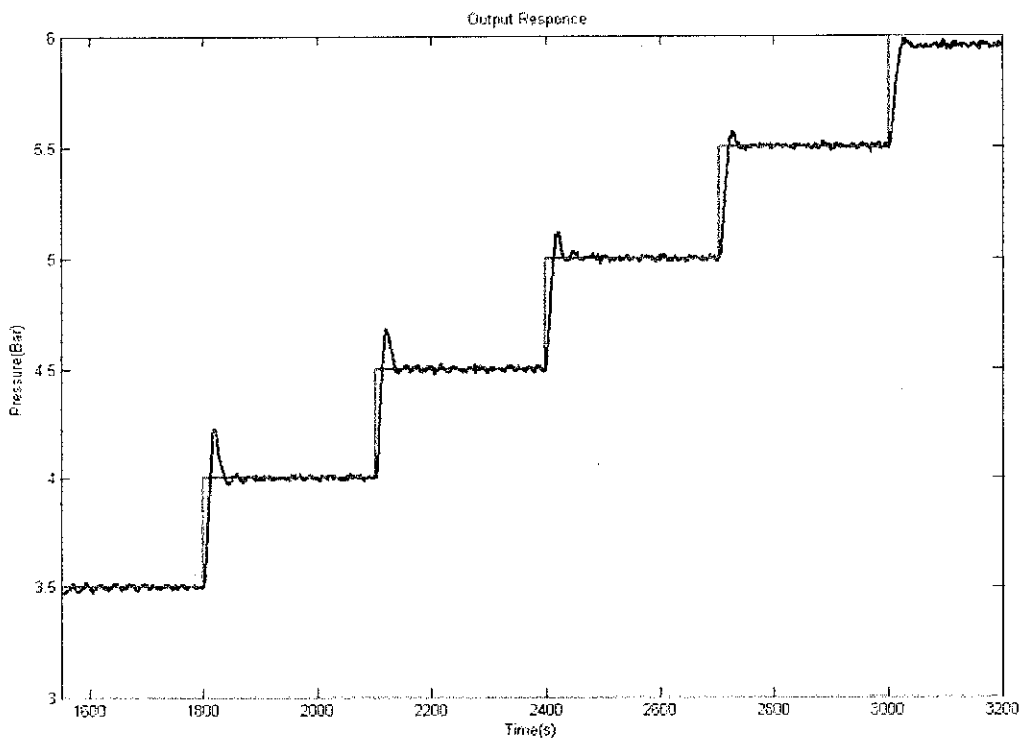

Figure 4.15: Pressure response due to the set point change

CONCLUSION & RECOMMENDATION

CHAPTER 5

CONCLUSION & RECOMMENDATION

5.1 Relevancies to Objective

The first task, which is to model and simulate the pressure process control, is finally completed. Empirical modeling approach is used to give a quite similar response after being compared with actual ones. Therefore, it is valid to use this model as an output predictor to the current pressure process control.

The second task which involves PID controller tuning and performance test is also accomplished. Several analyses on PID performance are conducted using Simulink model. Form the tuning exercise, the behavior of both process variable and manipulated variable are studied and the result obtained is useful for implemented in actual pilot plant.

The third task is to develop the MATLAB Simulink xPC Target use as an interface to do the real time experiment. The xPC target is added to visualize signal while running the target application to observe the output response due to the set point change. It also can show the response when adding the disturbance to the process. From the real time simulation using the Simulink xPC Target, the behavior of the process value is reasonably accurate. This is clearly shown in the accuracy of the values in term of rise time, overshoot, settling time and decay ratio and the behavior of the plant.

5.2 Next Project Recommendations

The system is a Single Input and Single Output (SISO) system. It only concerns with one input; shell inlet flow rate and one output; tube outlet pressure. Meaningful results can be obtained if the system can be extended to Multiple Inputs and Multiple Outputs (MIMO) system where the complete model of pressure process can be used to analyze the various performances and the effect of the various disturbances.

The reason why this project is recommended to involve with MIMO system is because with multiple inputs and output, the vast potential of Fuzzy Logic Controller (FLC) can be utilized in this system. The advantage of FLC is its ability to imitate (learn) another process behavior and it can be configured to handle more inputs than the PID controller. Therefore, it is possible that a single FLC can replace several PID controllers in controlling a MIMO system.

In the PID controller analysis, it is recommended that all the tuning parameters calculated and tested at actual plant. By doing this, the simulation result can be compared with the actual one more validations can be made. The problem faced in this project during the initial stage of the system used to control the pilot plant ware not synchronous in term of the tag number, the P&ID of the plant and also the PID design.

The Simulink block diagram developed is quite useful for analysis the pressure behavior. It is recommended that the diagram can be translated into a nice interface to have a clearer analysis presentation that can be viewed by user. For example by using Graphical User Interface (GUI) which is one of the MATLAB development environments to simplify the process of designing.

5.3 Conclusion

Modeling and Simulation of a pressure process control is a useful learning tools to understand process control technique. To start a modeling process, a good understanding of the process behavior is required to determine the important parameters and characteristics. The input and output correlations are also importance to be studied since it will affect the model performance and accuracy.

There are many modeling approaches that can be used to model the pressure process control and the selection depends on the user requirement. The most popular and simplest is empirical modeling method. The PI controller is chooses as the controller for the pressure control in this project. This is because it is a very common system used in the industrial control. The PI controller is designed based on the calculation of process model and parameter obtained from an experiment.

An investigation into the development of a controller for the pressure control in a gaseous plant has been presented. Real-time operation using the parameters gather in the on-line control is an important task to demonstrate the viability of this method.

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REFERENCES

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[3] Process Control Instrumentation Technology by Curtis D Johnson, Pearson International Edition 2006.

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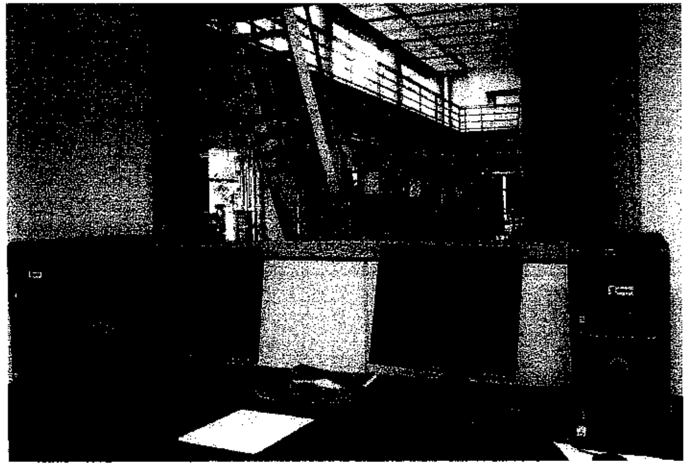
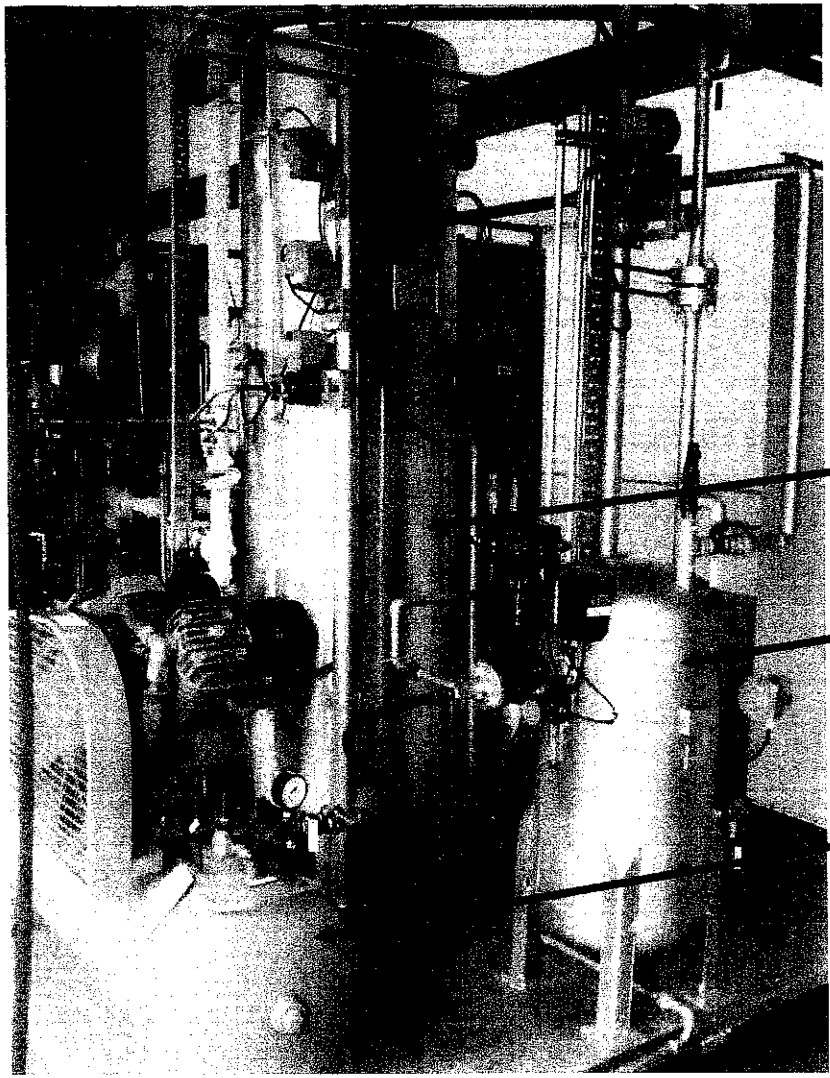
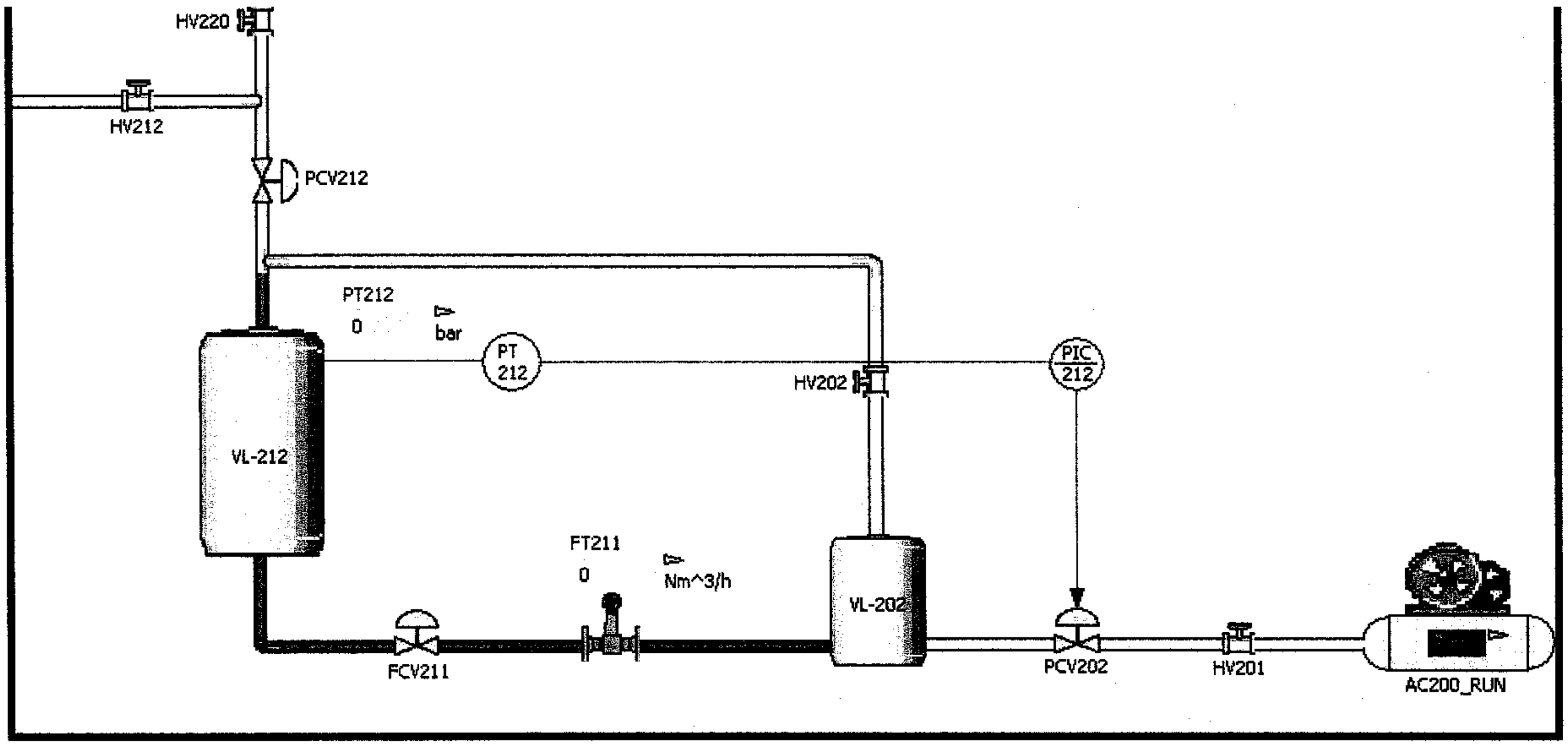
[8] http://en.wikipedia.org/wiki/PID_control

APPENDIXES

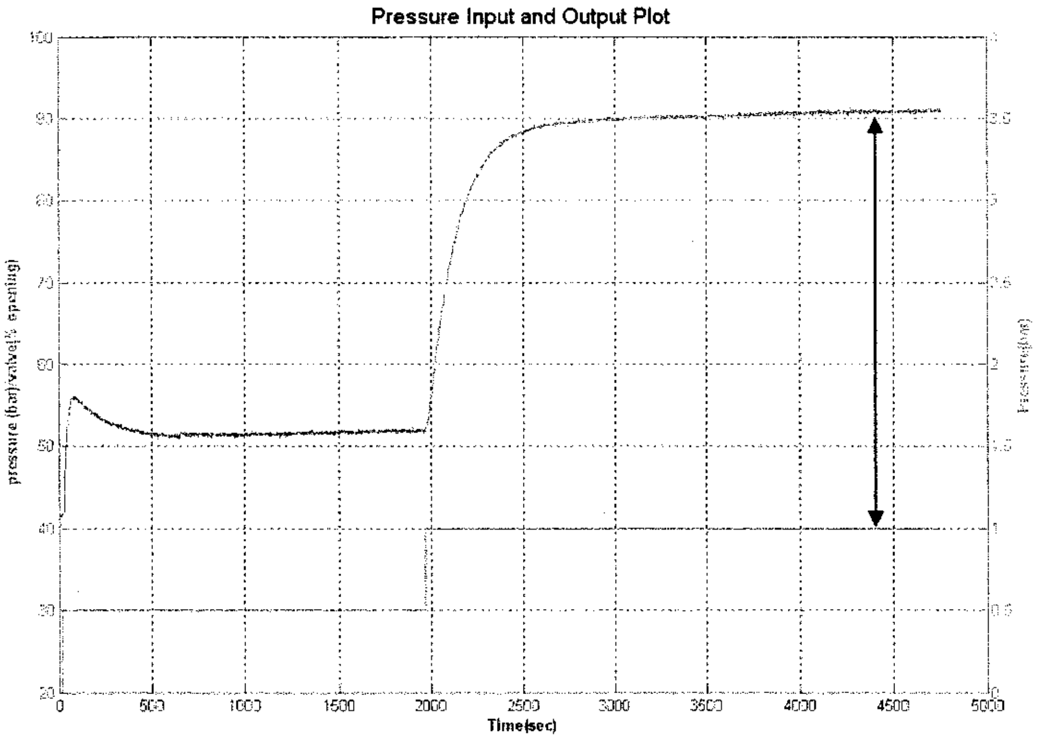




APPENDIX 1: Plant that involve on this project



APPENDIX 2: P&ID of Pressure Process Control



APPENDIX 3: Method II - Process Reaction Curve form Open Loop Test

δ = 40% – 30%

= <u>10% opening</u>

Δ = 3.53bar- 1.59bar

= <u>1.94bar</u>

K_p = Δ / δ

=1.94bar/ 10%

= <u>0.194 bar/ %</u>

0.63 Δ = 2.815bar

0.28 Δ = 2.13bar

$t_{63\%}$ = <u>176s</u>

$t_{28\%}$ = <u>69s</u>

τ = $1.5(t_{63\%} - t_{28\%})$

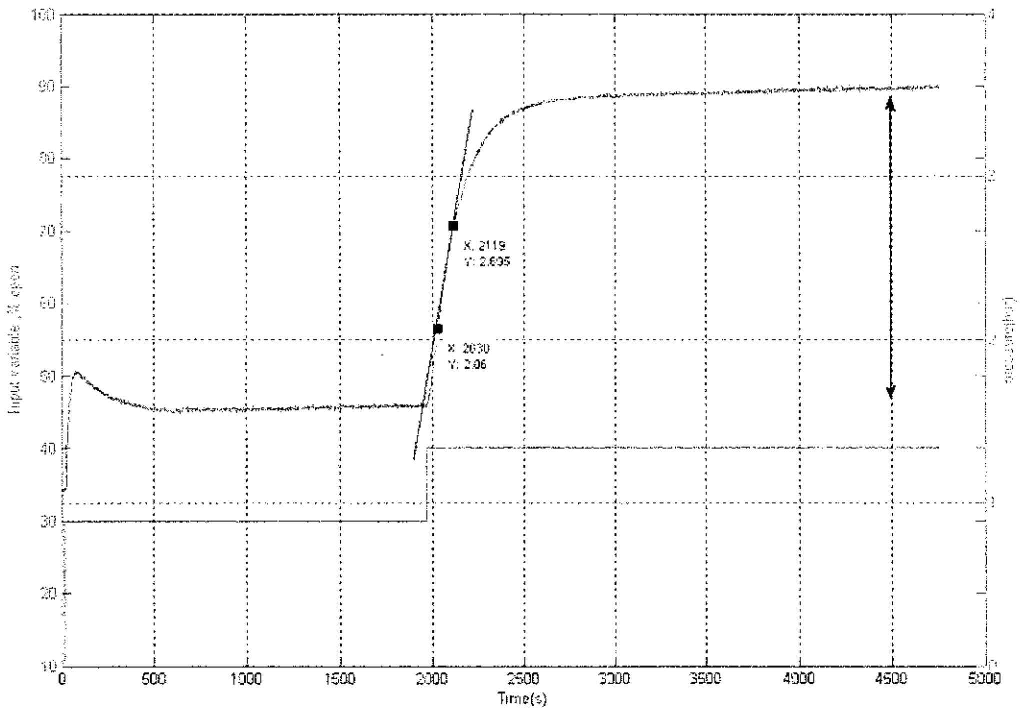
= 1.5(176-69)

= <u>160.5</u>

θ = $t_{63\%} - \tau$

= 176 – 160.5

= <u>15.5 s</u>



APPENDIX 4: Method I - Process Reaction Curve form Open Loop Test

$\delta = 40\% - 30\%$

$= \underline{10\%\ \text{opening}}$

$\Delta = 3.53\text{bar} - 1.59\text{bar}$

$= \underline{1.94\text{bar}}$

$K_p = \Delta / \delta$

$= 1.94\text{bar}/\ 10\%$

$= \underline{0.194\ \text{bar}/\ \%}$

$s = \dfrac{(2.695 - 2.06)}{(2119 - 2030)}$

$= 0.0071$

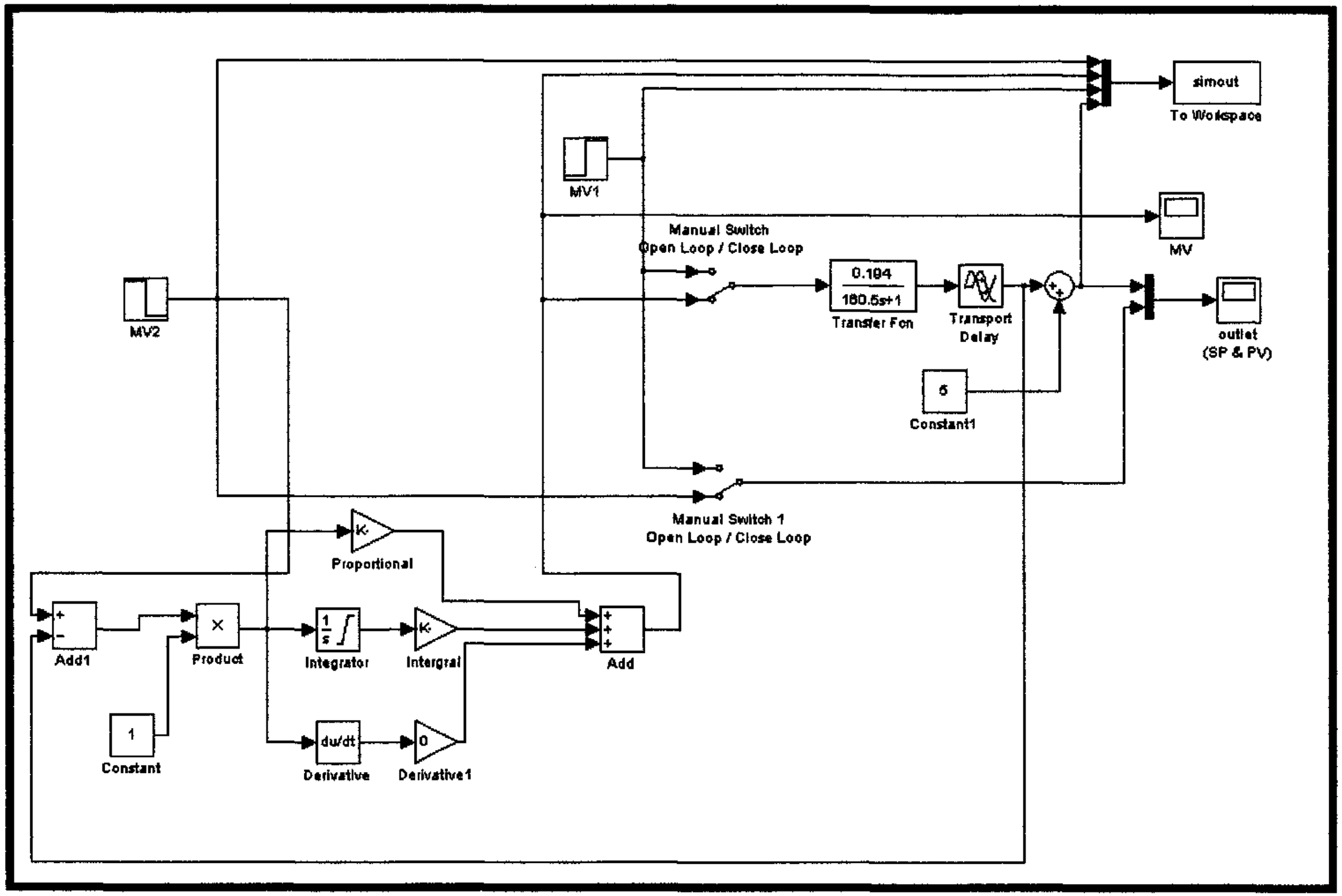
$\tau = \dfrac{\Delta}{s}$

$= \dfrac{1.94}{0.0071}$

$= \underline{271}$

$\theta = 1987 - 1969$

$= 18\text{s}$



APPENDIX 5:Simulink Block Diagram for Empirical Model with PID Controller

APPENDIX 6
PI Tuning Parameter Calculation for Giancone and Lopez

Given:
Process Gain K_p=0.194bar/% opening

Time Constant τ=160.5secon

Time Delay θ=15.5s

Fraction dead time, $\frac{\theta}{(\theta+\tau)}$=0.1

Giancone

From Figure 9.5 at Marlin[5] (pg 281)

$$k_c k_p = 2$$

$$\therefore k_c = 10.3$$

$$T_i/(\theta+\tau) = 0.73$$

$$\therefore T_i = 128.5$$

Lopez

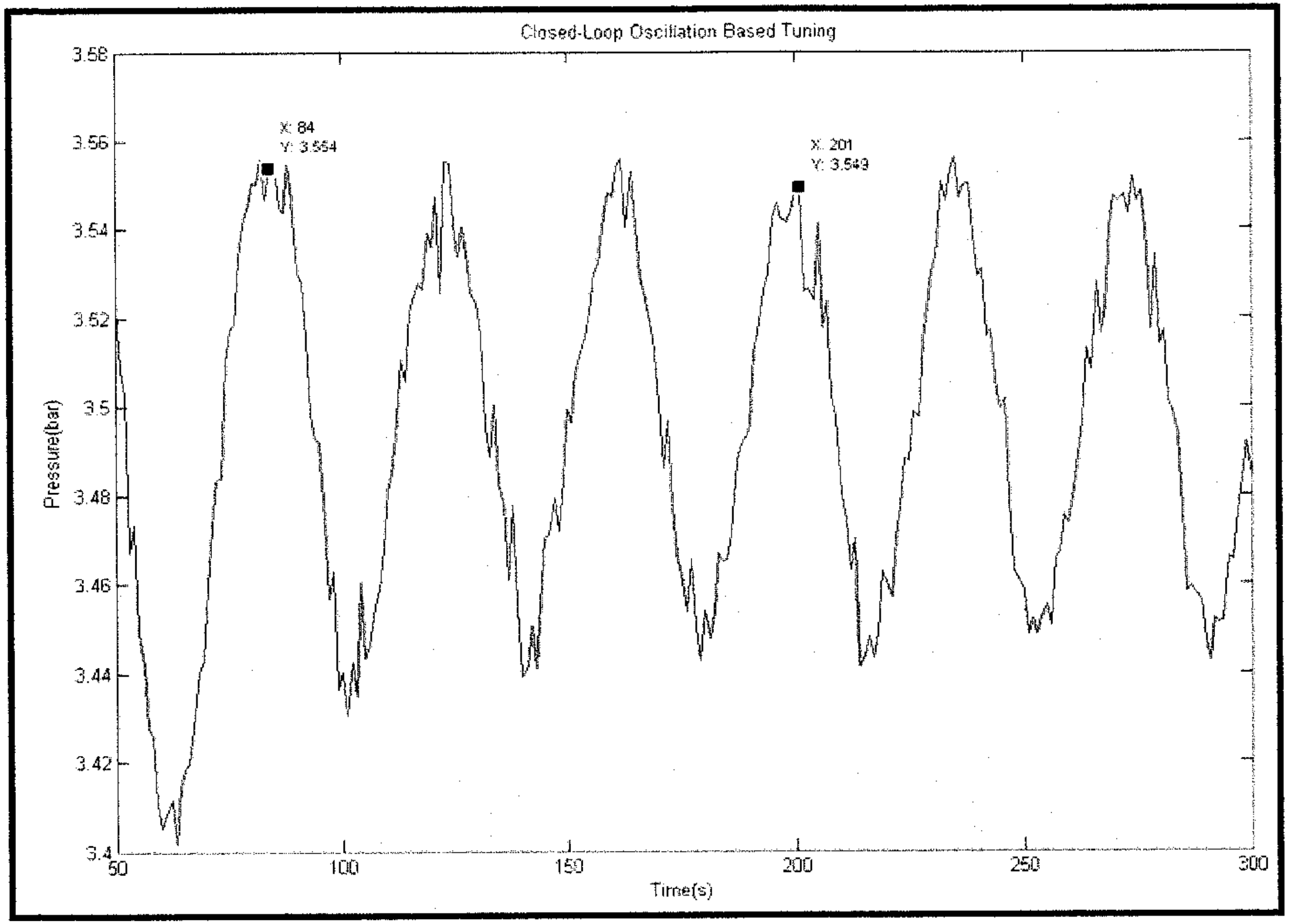
From Figure 9.10 at Marlin[5] (pg 287)

$$k_c k_p = 5.5$$

$$\therefore k_c = 28.4$$

$$T_i/(\theta+\tau) = 0.5$$

$$\therefore T_i = 88$$



APPENDIX 7: ClOscillation Based Tuning

APPENDIX 8
Bode plot tuning calculation

From the Bode plot,

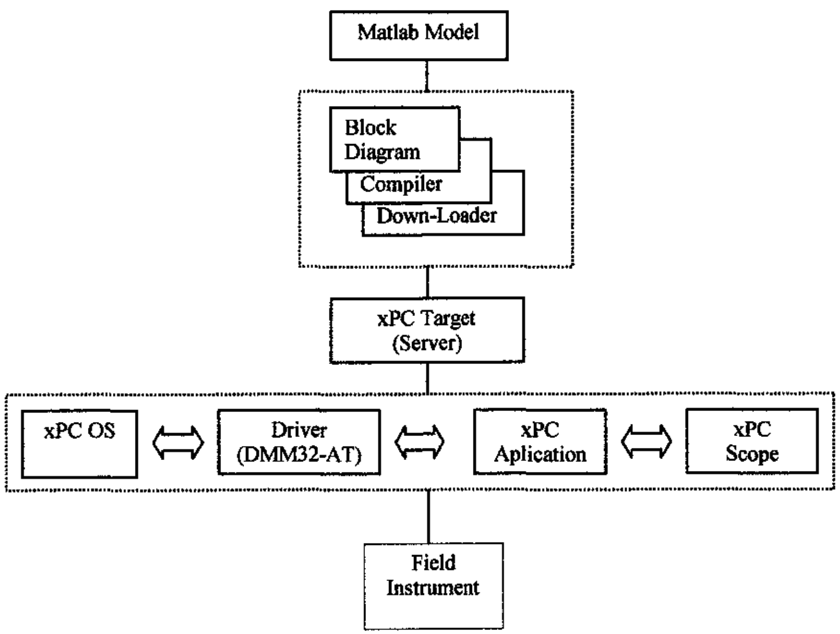
Frequency (rad/sec) = 0.105

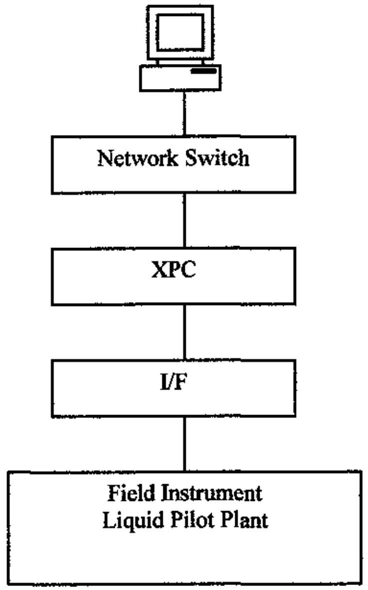
Magnitude (dB) = -38.8

$$K_u = \frac{1}{AR_c} = \frac{1}{10^{(-38.8/20)}} = 87.1$$

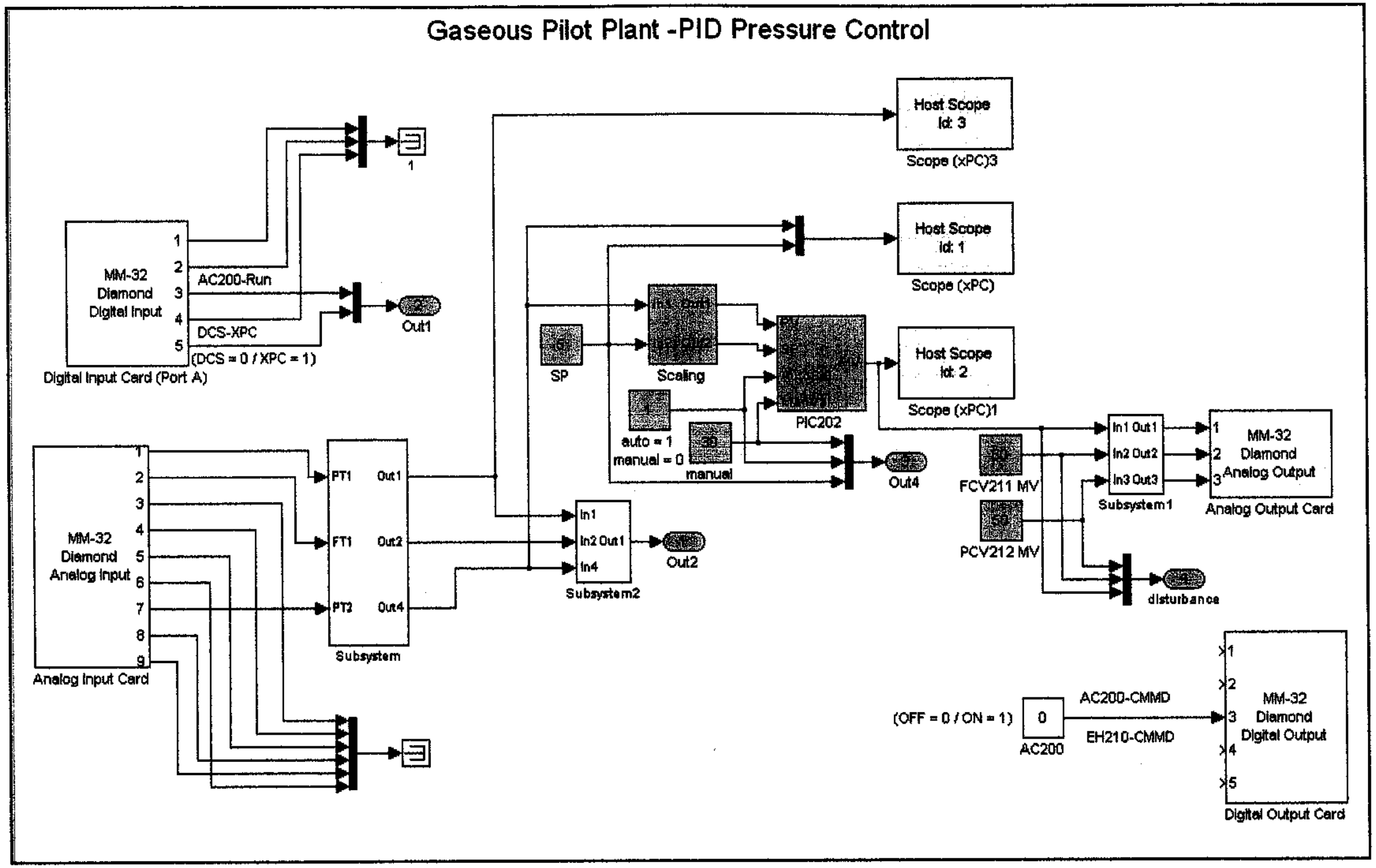
$$P_U = \frac{2\pi}{0.105} = 59.8$$

Controller	K_c	T_i	T_d
PI	39.6	49.8	–
PID	51.2	29.9	7.475





APPENDIX 9: Software and Hardware Configuration of the xPC Target



APPENDIX 10: Block diagram of MATLAB Simulink Xpc Target