INDUSTRIAL PROCESS MODELLING AND SIMULATION

By

ABID AKBAR SOOMRO

FINAL PROJECT REPORT

Submitted to the Electrical & Electronics Engineering Programme
in Partial Fulfillment of the Requirements
for the Degree
Bachelor of Engineering (Hons)
(Electrical & Electronics Engineering)

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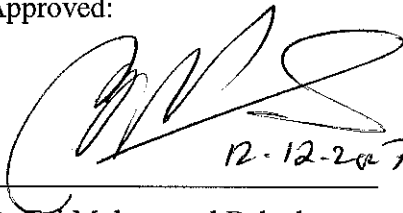
CERTIFICATION OF APPROVAL

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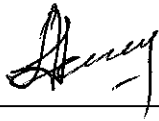(Electrical & Electronics Engineering)

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December 2007

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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.

Abid Akbar Soomro

ABSTRACT

The process systems are modeled to know the dynamic behavior of the system. The models are based on the fundamental theories or laws of the conservation of mass, energy, and momentum. The objective of this project is to develop model that need to address the process that is difficult (complex) to model. In Industrial Process control, the process variables such as flow, level, pressure and temperature are widely controlled process variables. The modeling technique that is used in this project is empirical modeling. The experiment was conducted on the existing Gas Processing Pilot Plant located in the Process Control & Instrumentation Laboratory at University Technology PETRONAS. The empirical model is developed for pressure and it is single input single output (SISO) system using first order with dead-time model. The process reaction curve was obtained from the empirical model and the parameters of the system were obtained from process reaction curve. The parameters obtained from the empirical model are used to form the transfer function which is used for pressure controller analysis.

After finding the transfer function of the model, the task was to reconstruct the model in MATLAB/SIMULINK. Different tuning methods are used to analyze the controller performance. The tuning methods include both the open loop and closed loop and then the results with both tuning techniques are compared to yield the best result. The tuning parameters were then fed into the closed loop MATLAB and SIMULINK for the real time analysis.

ACKNOWLEDGEMENTS

I would like to take this opportunity to express my deepest gratitude to Allah the Almighty for the courage to put my plucky efforts in the Final Year Project. I would like to thank to following individuals or groups for giving me moral support through out the semester.

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I would like to thank my friends who have morally supported me and my family who encouraged me through out my academic studies.

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LIST OF ABBREVIATIONS

ABV	Actuator Ball/Block Valve
ASD	Alarm and Shutdown Equipment
ANSI	American National Standards Institute
API	American Petroleum Institute
barg	Bar gauge (pressure)
CCR	Central Control Room
DOSH	Department of Occupational Safety and Health. Formerly known as Factory and Machinery Department (FMD)
ESD	Emergency Shutdown with Venting
FCV	Flow Control Valve
FI	Flow Indicator
FLC	Fuzzy Logic Controller
FR	Flow Recorder
FRC	Flow Recorder Controller
FT	Flow Transmitter
HP	High Pressure
LP	Low Pressure
K	Compressor
LC	Level Controller
LCV	Level Control Valve
LCP	Local Control Panel
LI	Level Indicator
LIC	Level Indicating Controller
LR	Level Recorder

LT	Level Transmitter
M	Motor
MV	Manipulated Variable
NC	Normally Closed
NO	Normally Open
P	Pumps for Equipment, Hydrocarbon for Piping
PC	Pressure Controller
PC	Personal Computer
PCV	Pressure Control Valve
PV	Process Variable
PR	Pressure Recorder
psig	Pound per square inch gauge (pressure)
RV	Relief Valve
SP	Set Point
V	Vessel
UTP	University Technology PETRONAS

CHAPTER 1

INTRODUCTION

1.1 Background of Study

Plants consist of numerous loops. Each of these control loops is designed to control some important process variable such as pressure, flow, level, temperature etc. within a required operating range to ensure the quality of the end product. Flow is controlled from the tank which affects the level of the other tank.[2].

Process control system identification is achieved through number of modeling techniques and in this project the empirical modeling is done which is easy approach compared to mathematical modeling due to the complexities in mathematical models. The identification job is carried to know the order of the system dynamics whether they are slow or fast, stability properties, operation range, and time delay.

Empirical methods are adequate for many process control designs and implementations. In empirical model building two identification methods will be discussed. The first is termed as process reaction curve which employs simple, graphical procedures for model fitting. The second and more general method employs statistical principles for determining the parameters. In empirical model building, models are built by making the small changes in the input variables about the required operating conditions.

Another important stage in the design of the control system is the design of the controller for the system. In order to develop the controller first step is to analyze the system to be controlled and the information is given by the empirical model of the system.

The purpose of modeling the process is to emulate the relationship between the system's input and output data. For modeling the certain process designs an experiment is performed on the existing plant to collect the identification data which is also called the process data, select the model structure and calculate the model parameters using the process reaction curve. Then evaluate the performance of the model and recommend the remodeling if the model parameters differ by more than ± 20 percent.

Process modelling plays a vital role in the process control system industry. Process control engineers are heading forward to build the accurate model (the model with minimum errors) and that represents the dynamics of that system. This modelling facilitates us to implement the different control strategies where performing the experiment on the running plant becomes impossible due to the operation condition and the quality of the end product.

1.2   Problem Statement

The present project explains the modeling practice and solves the following problems, which can arise in industrial applications.

- Building the empirical model using the data from the plant.
- To test the existing PI/PID controller for the plant. The controller is tuned and optimized to maintain the controlled variable close to the desired set point.
- To simulate the model in MATLAB/Simulink using the data acquired from the plant.

1.3   Objectives

- To gather the process data from the experiment conducted on the UTP Pilot Plant (Gaseous Pilot Plant).
- To develop the empirical model from the process data using MATLAB and Simulink.
- To design the controller to improve performance of the plant by maintaining the selected variables close to their desired optimum values.

1.4   Scope of Study

Fundamental and empirical models are based on such principles as material and energy conservation and can provide great insight as well as predictive power. Fundamental systems can be very complex for many systems and simplified empirical models based on the data taken from the experiment from the plant are sufficient for many process control tasks.

The model can be constructed using the data from the existing UTP plant (Pilot Plant) and then it is simulated using Simulink (model validation). The modeling involves empirical modeling approach.

The model accuracy is verified using the MATLAB/SIMULINK. Comparison is made with the real time data. The controller performance is tested and compared. The PI controller was used which very common in industry.

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CHAPTER 2

LITERATURE REVIEW

2.1 Basic Elements in the Process Control Loop

The selected process control systems consist of a control loop having four main components:

- A Flow Transmitter measures the flow of the liquid.

- A controller calculating an action based on measured value against a preset or desired value (set point).

- An output signal resulting from the controller calculation, which is used to manipulate the process action through some form of actuator.

- The final action or corrective action is by the Flow Control valve (FCV) [2].
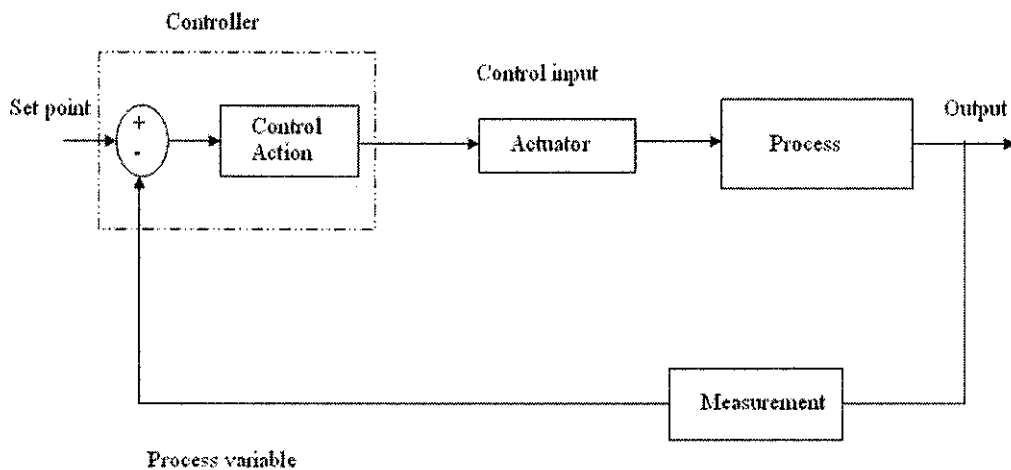
Figure 2.1 Block Diagram showing the elements of a process control loop

2.2 Pressure Control Loop

The study is carried out on the gas plant which is located in University Technology PETRONAS Process laboratory. The pressure control loop is shown below:
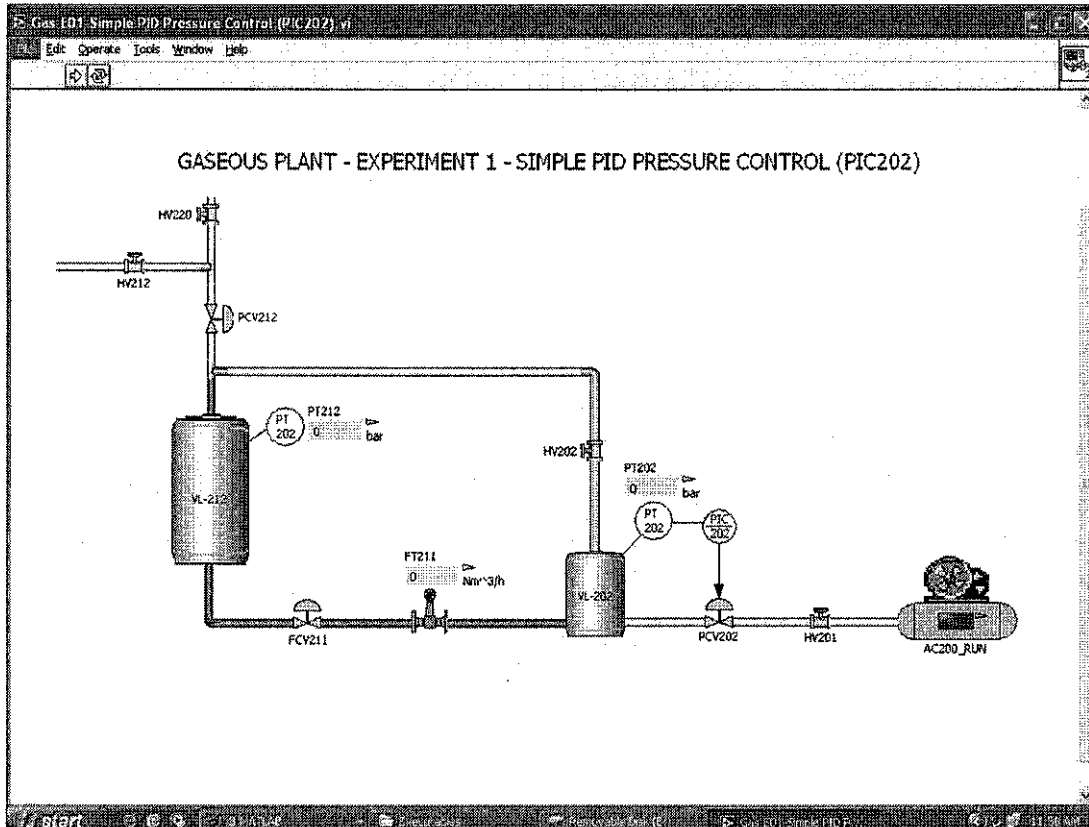
Figure 2.2: Simplified Pressure Diagram for Gaseous Pressure Plant

This pressure plant consists of gas vessel VL212, Pressure controller PIC 212, fail open diaphragm valve and pressure sensor.

The task is to regulate the flow from vessel VL212. The manipulated variable is the flow to the vessel VL212 that is to be adjusted to manipulate the desired pressure.

The pressure transmitter (PT212) sends the measured pressure signals to the pressure controller (PIC 212) which compares the measured pressure with the desired pressure

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set-point and then sends the signal to the valve (PCV 202). The PCV 202 will step up or down to any change of the flow rate that affects the pressure in the vessel.

Table 2.1: Conversion Table

Current (mA)	Pressure (psig)	Percentage (%)
4	3	0
8	6	25
12	9	50
16	12	75
20	15	100

The control block diagram is developed for the feedback control loop. The elements in control loop are combined and the closed transfer function is obtained using the figure 2.3 below
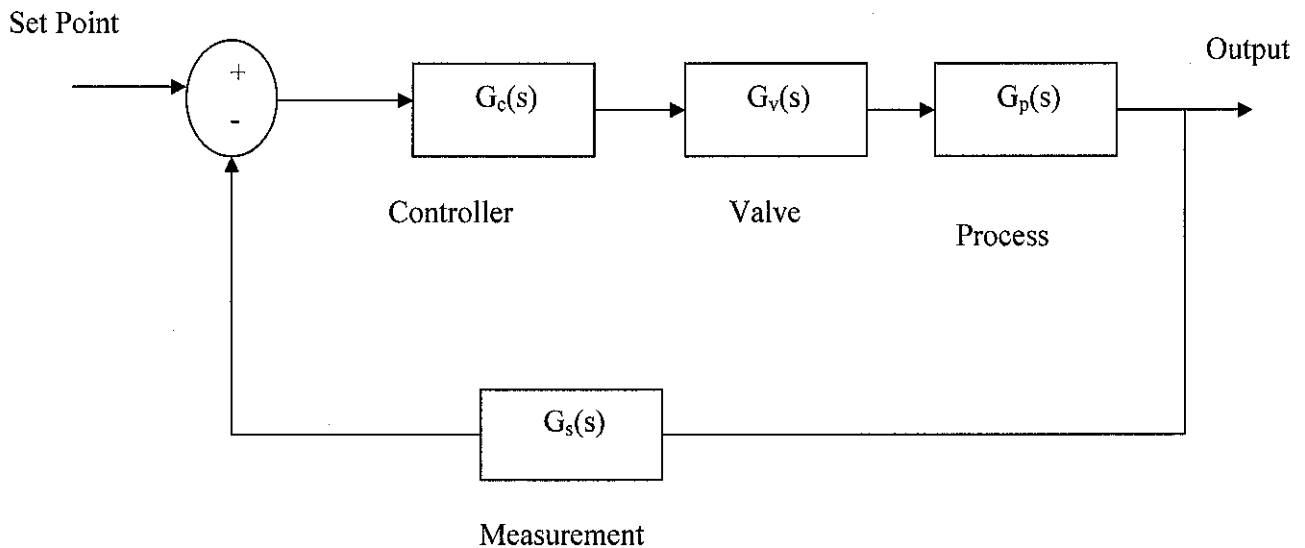
Figure 2.3: Block Diagram of the pressure plant

2.3 Empirical Modelling

To succeed in process control the designer must first establish a good understanding of the process to be controlled. Since it is not advisable to become too deeply involved in chemical or process engineering, we need to find a way of simplifying the representation of the process we wish to control. This is done by adopting a technique modeling and the idea is to use empirical modelling.

Empirical modelling is the alternative to the mathematical modelling for the process control system. Empirical modelling provides the dynamics of the system (i.e. dynamic behavior of the system between the selected inputs and outputs).

In empirical model building two common methods used to design the experiments. The perturbation to the system is the step response in this project to generate the dynamic response. [3].

2.4 Process Reaction Curve

The process reaction curve method involves the following four actions:

- Allow the process to reach the steady state.
- Introduce the perturbation which is the perfect step input.
- Collect the process data (input and output response data) until the response reaches the steady state.
- Perform the calculations on the process reaction curve.
-

The graphical calculations determine the parameters for the first order with dead time. The process reaction curve is restricted to the first order and the transfer function from the process data is easily achieved. [3].

$$Y(t) = K_P\,(t)\,\delta[1\text{-}e^{-(t\text{-}\theta)\,/\,t}]$$

The differential equation then transformed into transfer function given below.

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

Based on the Ziegler and Nichols) method, there are two graphical techniques that can be used [3].

2.4.1 Method I for Process Reaction Curve

The values determined from the graph are

- The magnitude of change in input,δ

- The magnitude of change in output,Δ

- The maximum slope of output versus time, S

- The intercept of maximum slope with initial values, also called Dead timeθ.

The values from the plot are used to get the transfer function for the plant.

The maximum slope occurs at $t=\theta$ and $s= \Delta/\tau$. The model parameters can be calculated as follows: [3].

$$K_p = \Delta / \delta$$

$$\tau \quad = \Delta / s$$

$$\theta \quad = \text{intercept of maximum slope with initial value}$$

2.4.2 Method II for Process Reaction Curve

Method II involves the calculation of model parameters when the output reaches the 28 and 63 percent of its final value. Any two values are picked where the transient response is changing rapidly so that the model parameters can be calculated accurately.

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The process gain is the same as Method I but the other parameters calculated as follows:

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

2.5 Open Loop Test

The open loop test is also called the manual test or manual control. The measured value of process variable is displayed to the operator and he has the ability to manipulate the control valve while sitting in the control room and sending signal to control valve. The purpose of open loop is to evaluate the process reaction curve for the model. For the process reaction it involves the following actions:

- First allow the reaction to get steady state

- Then introduce the single step change to the input

- Let the process again reach the steady state

- Collect the input and output values

- Perform the graphical calculations on the process reaction curve.

-

2.6 Closed Loop test

In the closed loop test the control is transferred to the controller rather than the operator. The closed loop control system provides a form of feedback to the process. The process is measured and then it is compared with the provided set point and final control element (control valve) is used to adjust the process. The process load will determine whether corrective action in the controller algorithm is needed. The PI controller was used for plant to control the overall system behavior.

2.7 Controllers

A controller is a device which monitors and affects the operational conditions of a given dynamic system. The controller takes a measured value from process and

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compares it with a reference set point value. The difference (error) is then used to adjust some input to the process in order to bring the process' measured value to its desired set point.

2.7.1 Proportional Integral Derivative (PID) Controller

A Proportional Integral Derivative (PID) Controller consists of three modes namely Proportional, Integral and Derivative. [7].
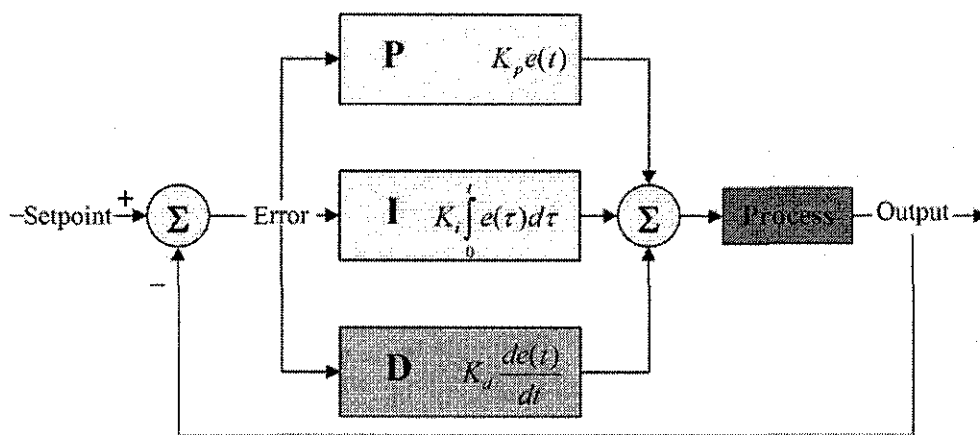
Figure 2.4: Schematic of a PID Control Loop

Proportional mode is used to correct or handle the error but it operates with non zero offset. To ensure that the controller must achieve zero offset an integral mode is added to it. In Integral mode the steady state error is eliminated. An integral controller reacts to the error by accumulating a value that is added to the output value (Plant output). While this will force the controller to approach the set point quicker than a proportional controller alone and eliminate steady state error, and it also guarantees that the process will overshoot the set point since the overshoot value will continue to be added to the output value. The derivative mode or derivative controller is added to reduce the magnitude of the overshoot produced by the integral component, but the controller will be a bit slower to reach the set point. [7].

A PID controller is called a PI, PD, or P controller in the absence of respective control actions. In designing the PID controller the following steps should be considered.

- Obtain an open-loop response and determine what needs to be improved.
- Add a proportional control to improve the rise time.
- Add a derivative control to improve the overshoot.
- Add an integral control to eliminate the steady state error.
- Adjust the gains of each controller (Proportional, Integral and derivative) until a desired output response is obtained.

It is not necessary to implement all three controllers (Proportional, Integral and derivative) into a single system. If a PI controller gives a good enough response then no need to implement the derivative controller to the system. The controller is kept as simple as possible. [7].

CHAPTER 3

METHODOLOGY

3.1 Experimental Design

An important aspect of empirical modelling is the need for proper experimental design. The whole experiment is divided into few stages. The initial stage is to understand the control loop for gaseous pressure plant. After getting the priori knowledge about the system the open loop test of the loop is carried. From the open loop test the transfer function is obtained. Then the open loop tuning procedure is performed. In this the suitable controller tuning and control strategy in order to achieve better process dynamics.

3.2 Basic Operating Procedure

Step by step procedure to operate the pilot plant has been given in each experiment manual. As a general guideline, some common start up and shut down procedure is including here:

3.2.1 Start up
- Switch on Power to the local control panel and Server Panel.
- Run the XPc target Program in Simulink.
- Turn the Selector switch to XPc to control the process via computer in the Laboratory assigned to do research.
- Switch the main air supply before running any experiment.

3.2.2 Shut Down
- Shutdown the DCS first
- Switch of the Main Air Supply.

3.2.3 *Hand Valves*

- Open and close the various hand valves based on the experiment instructions.
- All hand valves need to be fully open or fully close.

3.2.4 *Pumps*

- All pumps may be run directly from the DCS
- Open the valve partially before starting the related pump.
- Pump must be started or stopped at the Local Control Panel (LCP) if local control mode is selected.
- Do not run the pump until its suction is completely empty. This will damage the pump.
- Stop the pump when the pressure level in the tank is approaching the outlet of the tank.

3.2.5 Dependency of one experiment on another

- Make sure the preceding experiment have been completed before running the new one.

3.3   Project Flow Chart

At beginning stage, the literature review was studied on the gaseous pilot plant and about the PID controller. The study was also done on the equipment involved in the loop such as control valve, hand valve, pump, and pressure transmitter and about the control panel. The understanding about the simulation and the characteristic of PID controller and its effects was also done.

It is the combination of both modeling and simulation. In order to evaluate the PID performance a simple model of closed and open loop is developed using the MATLAB/Simulink. The figure 3.1 shows the implementation steps of modeling and simulating the gaseous pressure plant.

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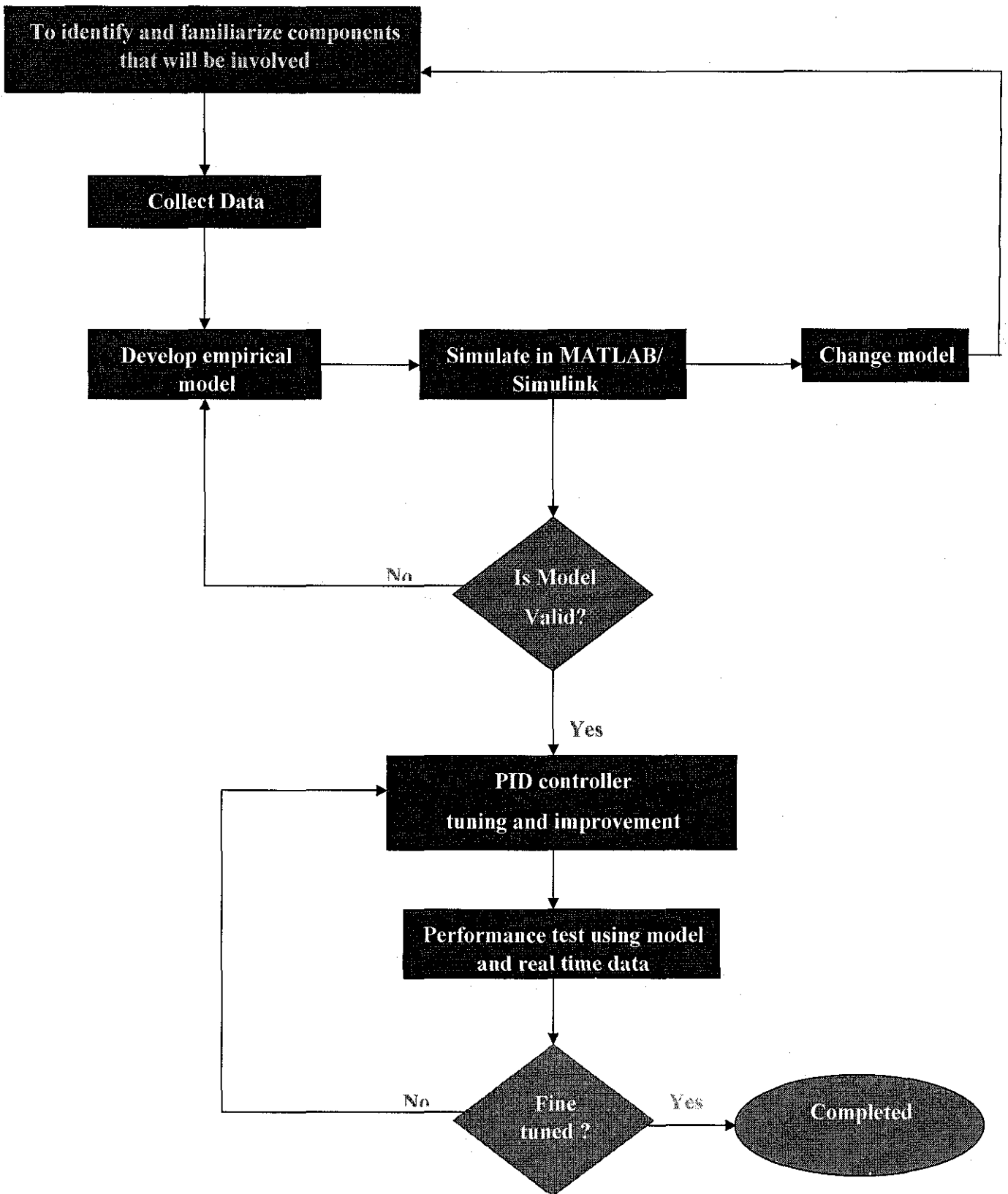
Figure 3.1: Overall Project Flow Chart

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3.4 Empirical Modeling

Empirical model is build after obtaining the plant data from the experiment. Empirical model building should be undertaken using the six step procedure shown below in flow chart. This procedure ensures that proper data is generated through careful experimental design and execution.[3].
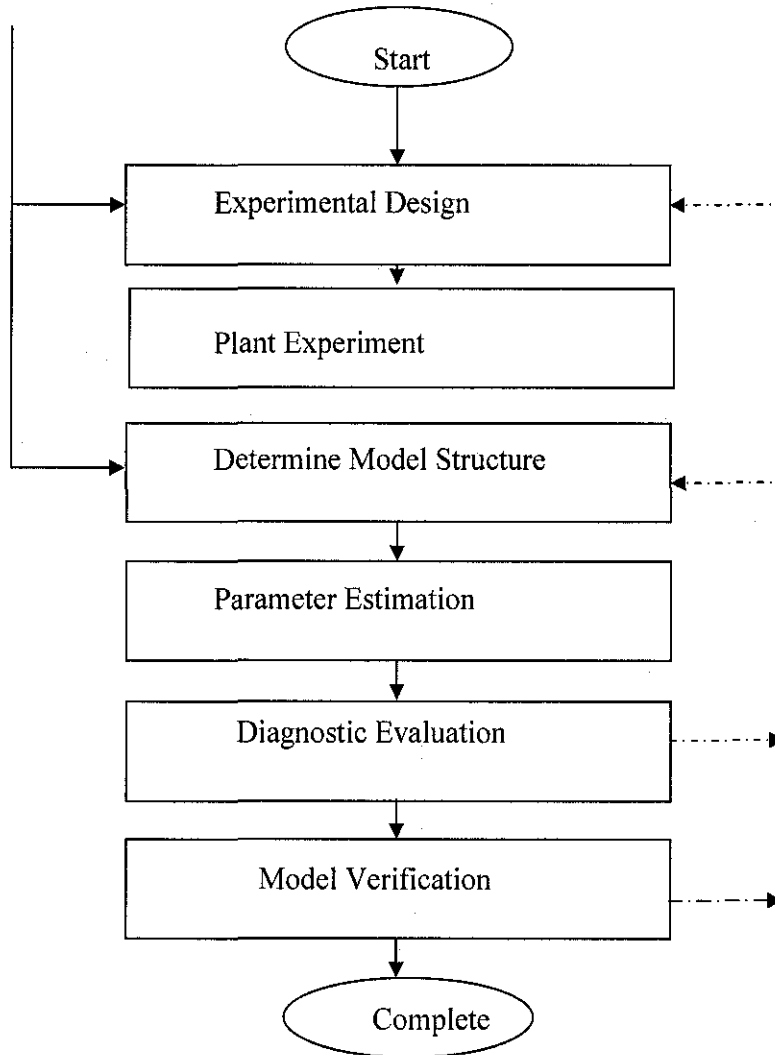
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   ────▶│        Experimental Design          │◀ ─ ─ ─ ─ ┐
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   ────▶│      Determine Model Structure       │◀ ─ ─ ─ ─ ┤
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Figure 3.2: Empirical Modeling Flow Steps

3.4.1 Experimental Design

An important aspect of the empirical modeling is the need for proper experimental design. Determine the input perturbation and its shape and duration, base operating conditions for the process. Get the priori knowledge before the experiment and its dynamic responses.

3.4.2 Plant Experiment

Plant operation monitored during the experiment because variation in plant operation is inevitable and large disturbances during the experiment can invalidate the results. The experiment is designed to establish the relationship between one input and one output (Single Input Single Output), changes in other inputs during the experiment can make the data unusable for identifying a dynamic model.

3.4.3 Determining Model Structures

First step in determining the model is to know the order of the system (e.g. first order, second order), based on the data. The initial structure is selected based on the prior knowledge of the unit operation, perhaps based on the structure of the fundamental model, and based on the patterns in the experimental data. The goal is to develop the model that matches the experimental data and to know the input output behavior of the system.

3.4.4 Parameter Estimation

The process reaction curve (Plant dynamic response) is obtained during the experiment is used to approximate the process parameters. The process parameters are:

- Dead Time (time delay), θ
- Time constant, τ
- Steady state Gain (Process gain), K_p

3.4.5 Diagnostic Evaluation

Evaluation is required before the model is used for the control. The diagnostic evaluation determines how well the model fits the data used for parameter estimation. Usually the comparison of the model prediction with the measured data is done in diagnostic evaluation.

3.4.6 Verification

In verification comparing the model to data collected at another time to make sure that typical variation in the plant operation does not significantly degrade model accuracy.

3.5 Tools and Software

3.5.1 MATLAB and Simulink

MATLAB is interactive software whose basic data element is an array that does not require dimensioning. This allows us to solve many technical computing problems, especially those with matrix and vector formulations, in a fraction of the time

MATLAB is offering array operations and since the data received is used in MATLAB to get the Process Reaction Curve (PRC). MATLAB also offers some programming features similar to other languages. The Graphical User Interface (GUI) in MATLAB allows use the Process Reaction Curve for our calculations of the transfer function. MATLAB features were really utilized in this project.

Simulink is an extension to MATLAB. It is a software package for modeling, simulating, and analyzing the dynamic systems. It allows to accurately build computer model of dynamic systems using block diagram notation. Simulink model can produce the simulation and is as easy approach to know the dynamic behavior of the system.

CHAPTER 4

RESULTS AND DISCUSSIONS

4.1 Empirical Model Results

The experiment was conducted at UTP in the Plant Process laboratory. The pressure and flow model is developed based on process reaction curve. Figure 4.1 shows the process reaction curve obtained from the experiment. The step change is 10% which is from 30% to 40% valve opening and the pressure on the vessel increment is about 2.787 bar. The parameters were estimated from the process reaction curve. The transfer function calculation using the parameters.
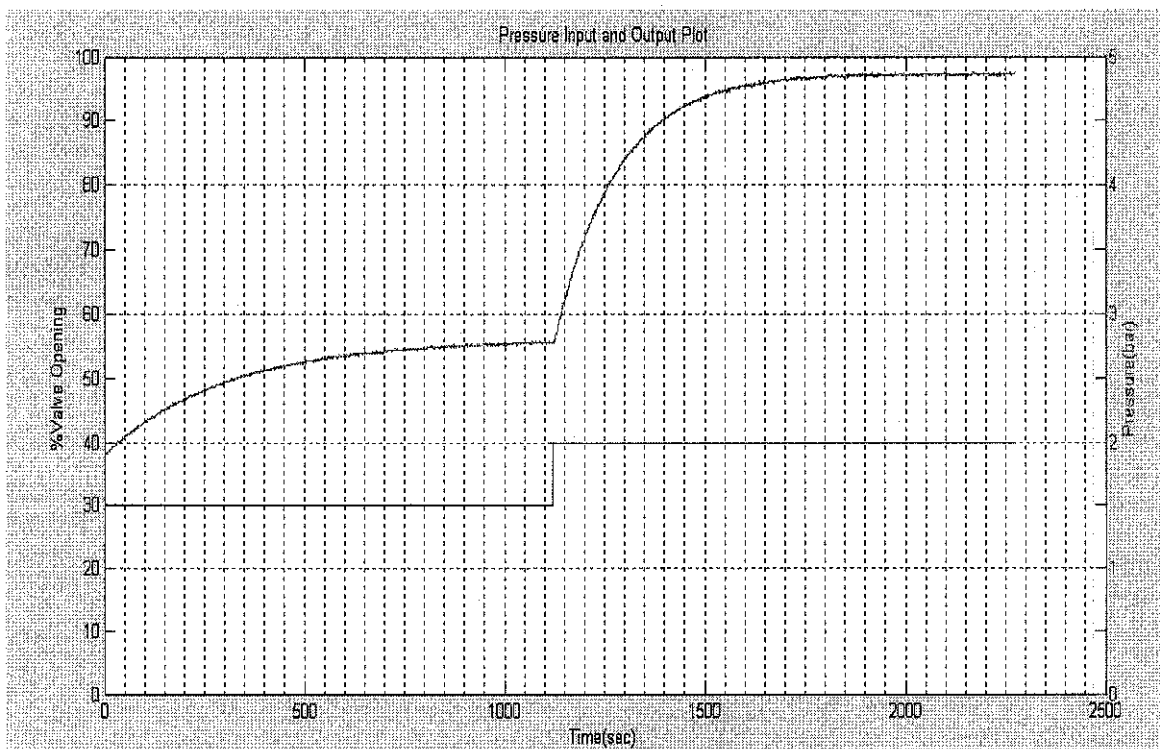
Figure 4.1: Pressure Input and Output Plot

As mentioned earlier that two methods that are used to calculate the transfer function parameters. Method I is about the value of output versus time plot. Due to difficulty in getting the values from the slope Method has a large error and therefore Method II is preferred over Method I. Method II concerns with the time at the process variable (output) reaches 28% and 63% of the final value. The detailed calculations on how to get the transfer function by calculating the parameters from the process reaction curve are given below.
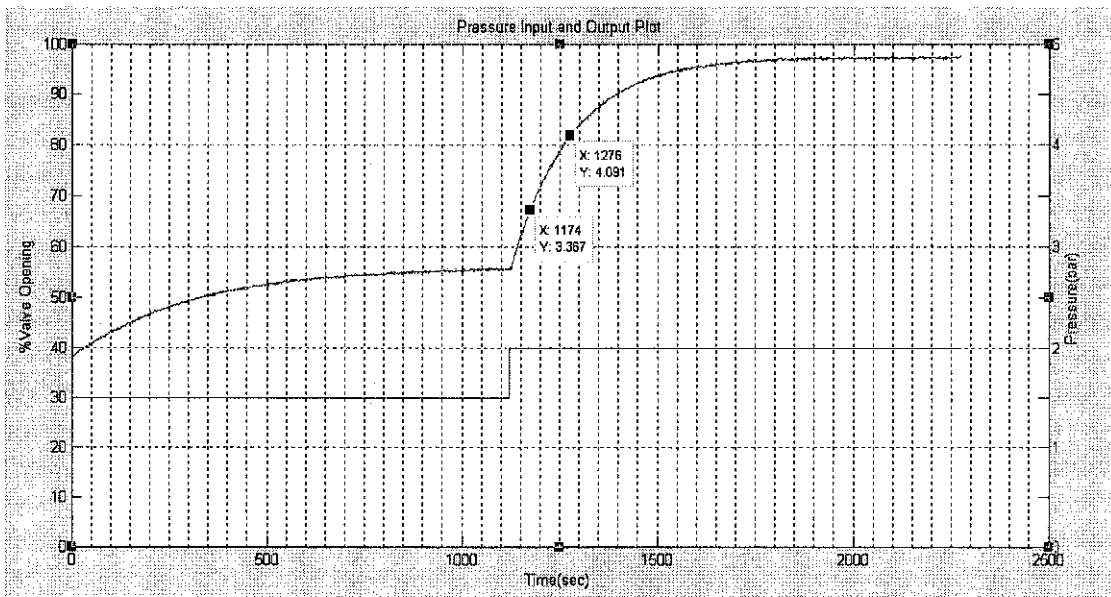
4.1.1 Method I
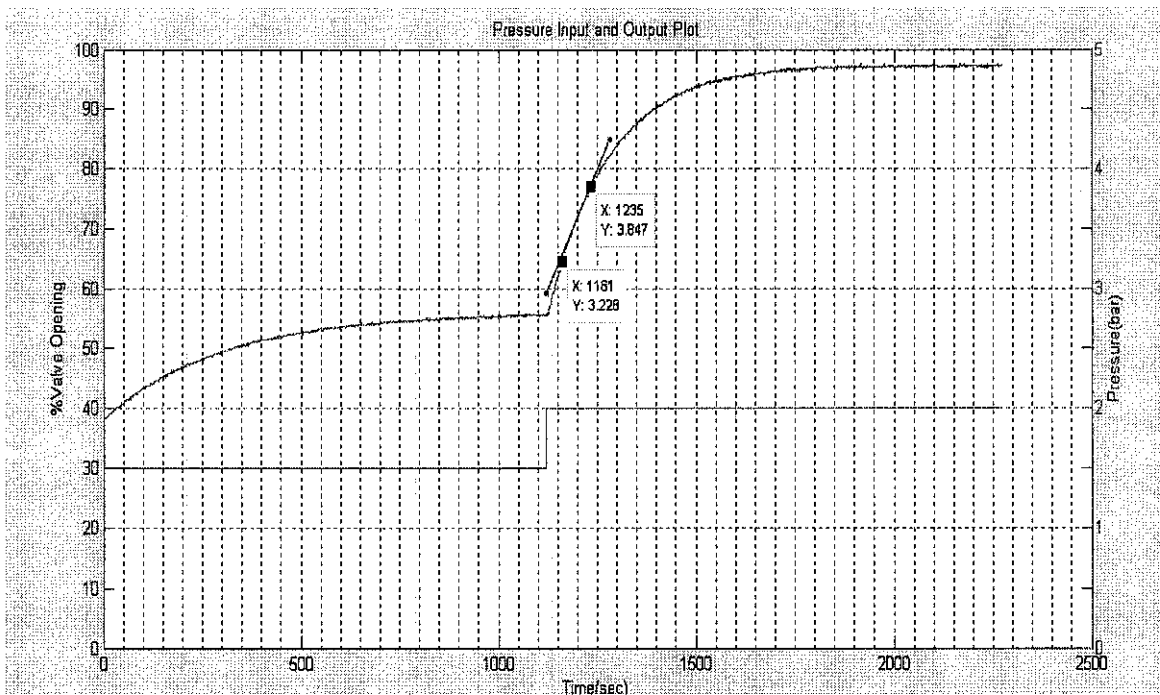
Figure 4.2: Method I Process Reaction Curve from Open Loop

Calculation:

Change in input $= \delta = 40\% - 30\% = 10\%$ Open

Change in output $= \Delta = 4.8650$ bar $- 2.787$ bar

$= 2.078$ bar

K_p $=$ Δ / δ

$=$ $2.078 / 10\%$

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| | = | 0.2078 bar / % |
| Slope | = | $(3.847 - 3.228)/(1235 - 1161)$ |

| | | |
| Slope | = | 0.008365 |

| Time Constant = | τ | = | 2.078/0.008365 |

| | | |
| | = | 248 seconds |

| Dead Time | $= \theta$ | = | 10 seconds |

The model transfer function:

$$G(s) = \frac{0.2078\ e^{-10s}}{248s+1}$$

4.1.2 Method II

Figure 4.3: Method II Process Reaction Curve from Open Loop

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Calculation:

Change in input = δ = 40 % - 30% = 10 % Open

Change in output = Δ = 4.8650 bar – 2.787 bar

$= 2.078$ bar

K_p = Δ / δ

= 2.078/ 10 %

= 0.2078 bar / %

0.63Δ = 1.30914+2.787

= 4.09614

0.28Δ = 0.58184+2.787

0.28Δ = 3.36884

$T_{63\%}$ = 156 seconds

$T_{28\%}$ = 54 seconds

τ = 1.5 (156 -54)

= 153 seconds

θ = $t_{63\%} - \tau$

= 156 -153 = 3 seconds

The model transfer function:

$$G(s) = \frac{0.2078\,e^{-3s}}{153s+1}$$

The results of both the methods are tabulated below:

Table 4.1: The results of Method I and Method II

Parameter	Method I	Method II
Process Gain K_p	0.2078 bar / %opening	0.2078 bar / %opening
Time Constant τ	248 seconds	153 seconds
Time Delay θ	10 seconds	3 seconds

The method II is preferred over method I and hence the transfer function for Method II is

$$G(s) = \frac{0.2078\,e^{-3s}}{153s+1}$$

4.1.3 Open Loop Test

Figure 4.2 shows the block diagram for the empirical modeling of pressure process model. The valve opening is given from 30% to 40% which means the step change of 10%. This step change is fed into the transfer function of method II and the output from the transfer function is displayed by the scope.
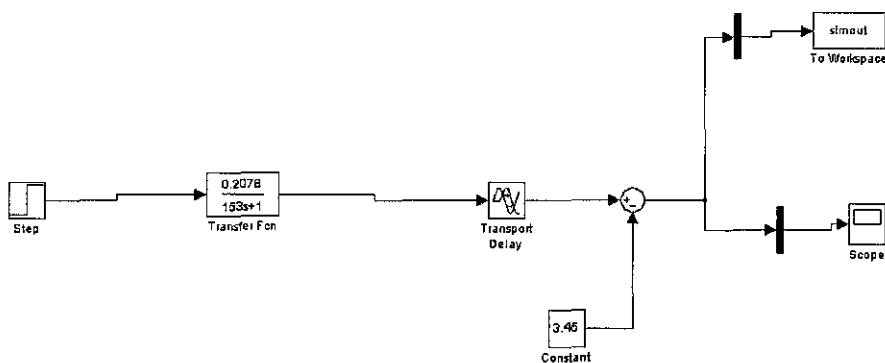

Figure 4.4: Simulink Block Diagram of Pressure Process Model

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The output of the model in Simulink is shown in Figure 4.3 shown below. The pressure increment is observed as 2.787 bar from the process reaction curve.
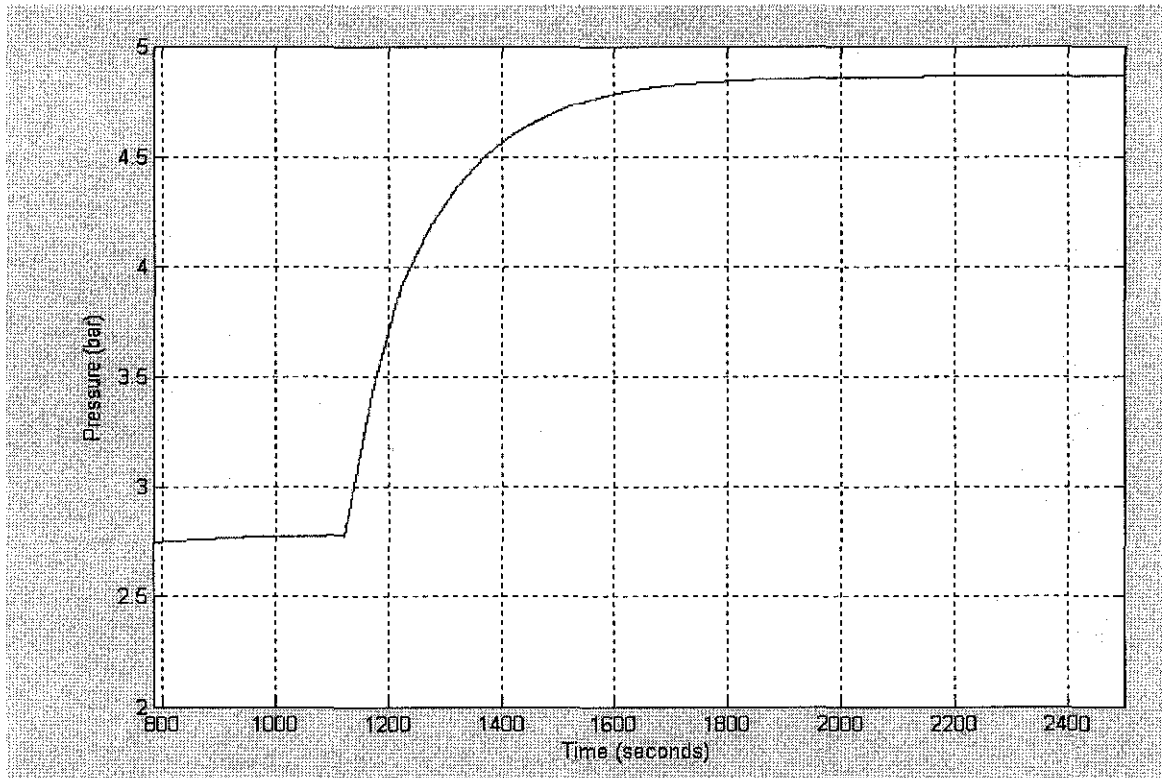

Figure 4.5: Empirical model using Simulink/MATLAB

The comparison is done with both the models and the change is not quite large and both look similar to each other. The model is valid by comparing the models from the pressure pilot plant.
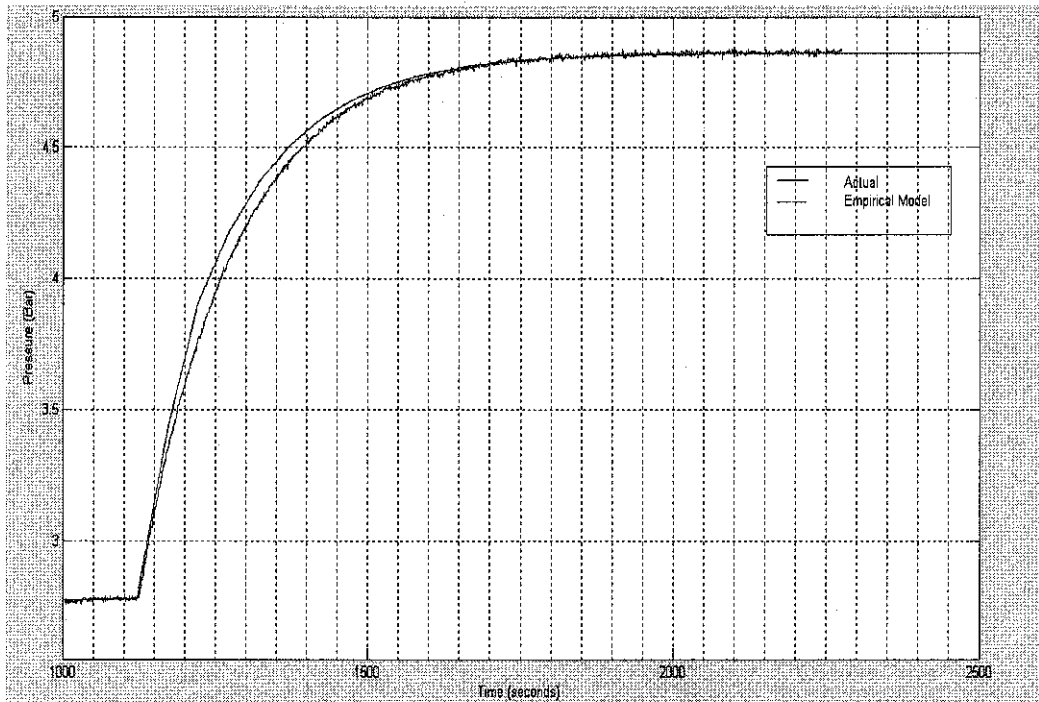
Figure 4.6: Simulink and Plant Open Loop Response of Empirical Model

4.2 Tuning Techniques

The PID controller Tuning methods are classified into two main categories namely Open loop tuning methods and closed loop tuning methods. Closed Loop tuning techniques refer to methods that tune the controller during automatic state in which the pilot plant is operating in closed loop. The open loop techniques refer to methods that tune the controller when it is in manual state and the pilot plant is operating in open loop.

4.2.1 Open Loop Tuning Methods

Tuning Constants are obtained using the parameters calculated from the process reaction curve. The tuning coefficients for PI and PID controller can be calculated by referring to Table 4.2 and Table 4.3.

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The formulae shown in below are based on Ziegler Nicholas open loop method and Cohen – Coon open loop method. The task is to get the closed loop response for quarter wave damping using both the methods which are empirically developed.

Table 4.2: Cohen – Coon open loop tuning method based on process reaction curve

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

Table 4.3: Ziegler-Nichols open loop tuning based on process reaction curve

Controller	K_c	T_i	T_d
PI	$(1/K_p)/(\tau/\theta)$	$3.3\,\theta$	-
PID	$(0.9/K_p)/(\tau/\theta)$	$2.2\,\theta$	0.5θ

Fertik suggested tuning method which emphasizes on minimum Integral Absolute error which limits the overshoot of the system. The formulae to calculate the tuning coefficients are given below

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

$$T_I = (\tau\,/\,0.674)\,(\theta/\,\tau)^{0.680}$$

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Calculations of Different Tuning Methods

1. Ziegler Nichols Open Loop Tuning

PI controller

K_c = (1 /0.2078) (153 / 3) = 245.428

T_I = 3.3 (3) = 9.9 seconds

PID Controller

K_c = (0.9 /0.2078) (153 / 3)= 220.88

T_I = 2.2 (3) = 6.6 seconds

T_d = 0.5 (3) = 1.5 seconds

2. Cohen- Coon Open Loop Tuning

PI controller

K_c = 245.4283 (0.901634) = 221.29

T_I = 3 (30.0588 /9.392) = 9.6 seconds

PID Controller

K_c = 245.4283 (1.33490) = 327.623

T_I = 3 (32.11765 / 13.15686) = 7.63234 seconds

T_d = 12 / 11.0392 = 1.087 seconds

3. Fertik Open Loop Tuning

PI controller

K_c = $0.859 (3 / 153)^{-0.977}$ = 40.021

T_I = $(153 / 0.674)(3 / 153)^{0.680}$ = 15.66332 seconds

4. Ciancone Correlations

The model transfer function:

$$G(s) = \frac{0.2078 \, e^{-3s}}{153s + 1}$$

Maximum Execution Period

$\Delta t < 0.05 (3 + 153)$

$\Delta t = 7.8$ seconds

New Dead time, $\theta' = \theta + \Delta t / 2 = 6.9$ seconds
Fraction Dead Time

θ' = $\theta / (\theta' + \tau)$ = 0.04315

$K_c K_p$ = 1

K_c =   1 / 0.2078   =   4.812

$T_I / (\theta + \tau)$ =   0.25

T_I =   0.25 (156)   =   39 seconds

5.   Lopez Tuning Technique

$K_c K_p$ =   5.0

K_c =   24.06

$T_I / (\theta + \tau)$ =   0.4

T_I =   0.4 (156)   =   62.4 seconds

Ciancone and Lopez also presented tuning correlations. The goal is to simply IAE (minimize the Integral Absolute Error) without concern for potential variation in feedback dynamics or effect on manipulated variable's transient behavior.
The comparison of tuning constants for PI controller is tabulated below.

Table 4.4: PI tuning parameters for Empirical model

PI Controller	Ziegler Nichols	Ciancone	Lopez	Fertik	Cohen-Coon
K_c	245.428	4.812	24.06	40.021	221.29
T_i	9.9	39	62.4	15.66332	9.6

The comparison of tuning constants for PID controller is tabulated below

Table 4.5: PID tuning parameters for empirical model

PI Controller	Ziegler Nichols	Cohen-Coon	Ciancone
K_c	220.88	327.623	4.812
T_i	6.6	7.6334	39
Td	1.5	1.087	-

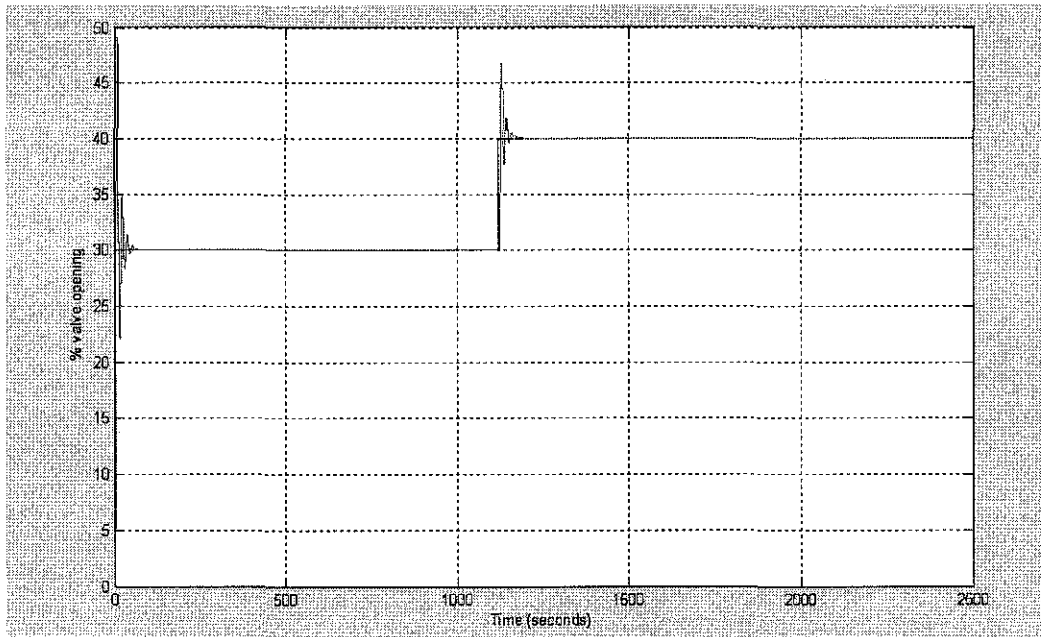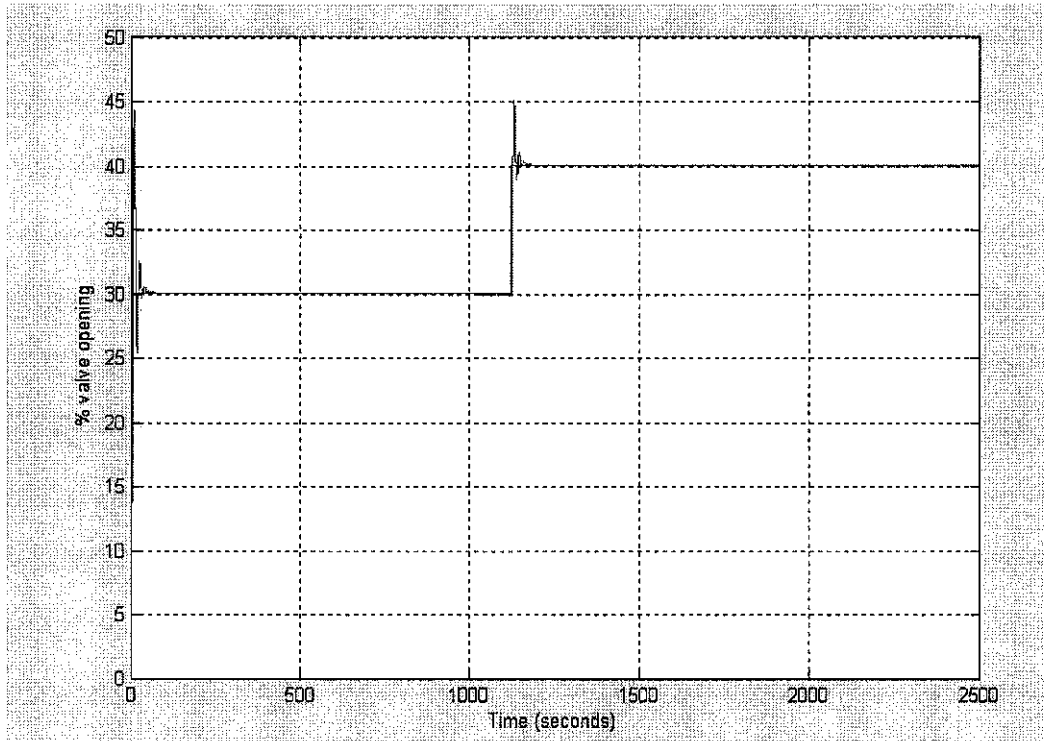The PV (Process Variable) response of the best controller is compared when a step change is provided.

Figure 4.7: The response of PI controller with Ziegler Nichols open loop tuning constants

Table 4.6: Performance values of PI controller for Ziegler Nichols open loop tuning

Settling Time	14 seconds
Rise Time	20 seconds
Overshoot	60%
Decay Ratio	16.67%

Figure 4.8: The response of PID controller with Ziegler Nichols open loop
tuning constants

Table 4.7: Performance values of PID controller for Ziegler Nichols open loop tuning

Settling Time	9 seconds
Rise Time	24 seconds
Overshoot	50 %
Decay Ratio	20%

The PID controller has high settling time and less overshoot than PI controller but it
has less rise time and high decay ratio compared to PI controller.

4.2.2 Closed Loop Tuning Methods

The closed loop experiment is conducted on the pilot Plant using the proportional control only. The proportional gain is increased until the process variable gets sustained oscillation which means the output neither grown nor decays significantly large in magnitude with time.
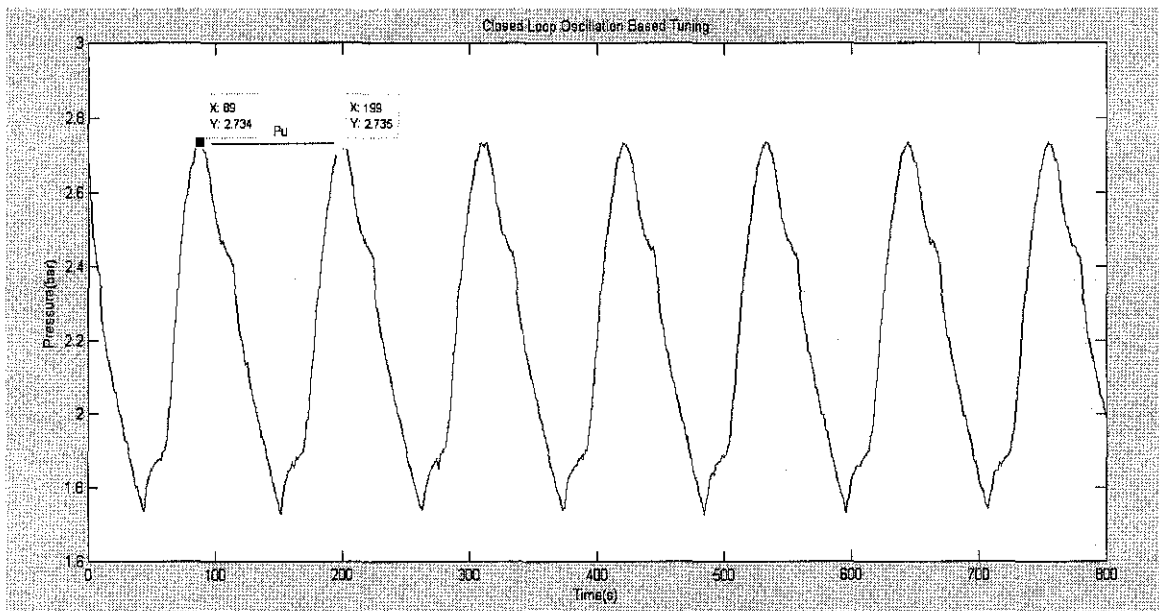

Figure 4.9: Closed Loop Oscillation Based Tuning

The value of the proportional control gain is noted which makes the output to get oscillatory response. The value of the proportional gain that causes the continuous oscillation is called the ultimate gain and is denoted as K_u. Ultimate period or Critical period is obtained from peak to peak (time between the successive peaks for the output). The gain that causes the oscillatory response of the output is noted as 0.5 and the critical period is 110 seconds ($K_u = 0.5$ and $P_u = 110$ seconds). The value of the critical period is received of the Oscillation based curve.

After getting the ultimate gain and ultimate period the task is to calculate the closed loop tuning constants. By using the K_u (ultimate gain) and P_u (ultimate period) the tuning constants for PI and PID controller can be calculated using two methods named as Ziegler Nichols and Tyreus-Luyben.

Tyreus Luyben suggested tuning parameters which gives less oscillatory response and is less sensitive to changes in process oscillation. The formulae for these methods are tabulated in Table 4.8 and Table 4.9.

Table 4.8: Ziegler Nichols Closed Loop Oscillation Based Tuning

Controller	K_c	T_i	T_d
PI	$0.45K_u$	$P_u/1.2$	-
PID	$0.6K_u$	$P_u/1.8$	$P_u/8$

Table 4.9: Tyreus-Luyben Closed Loop Oscillation Based Tuning

Controller	K_c	T_i	T_d
PI	$K_u / 3.2$	$2.2\,P_u$	-
PID	$K_u / 2.2$	$2.2\,P_u$	$P_u / 6.3$

The other closed loop tuning method is Bode Plot method which is also called Ziegler Nichols closed loop - Bode Plot method. This method can be applied to the process which is not the first order with dead time. It provides good insight into the loop elements stability and suitable tuning coefficient values.

The real importance of Bode Plot is in highlighting the contributions of various components to the stability of feedback system. Most process models have amplitudes that decrease with increasing frequency; a lower critical frequency yields higher amplitude for G_{OL} (jw). Since lower amplitude ratio is desired to maintain the amplitude ratio below 1.0 for stability, elements with the larger phase angle tend to destabilize a feedback control system. Simply the Bode plot stability analysis provides a way to determine whether a process and feedback controller, with all elements completely specified is stable.

The system will be closed-loop stable if and only if the amplitude ratio is less than one at the crossover frequency

$$AR_{CO} = \left| g_c(\omega) \, g_p(\omega) \right| < 1$$

Where ω is the crossover frequency which is defined as the frequency where the phase angle is -180°. Ultimate gain K_u is the proportional gain that brings the system to the margin of stability at critical frequency. Ultimate period P_u is the period of oscillation of the system at the margin of stability.

$$K_u = \frac{1}{\left| G_{OL}(j\omega) \right|} = \frac{1}{AR_{CO}}$$

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

From the Bode plot the ultimate gain and ultimate period can be determined and $P_u = 19.63$ and $K_u = 223.87$. The Ziegler Nichols closed loop tuning correlations are to be calculated using the formulae given in table 4.10. Bode Plot is shown in Figure 4.10 below:

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Table 4.10: Ziegler Nichols Closed Loop Tuning Correlation

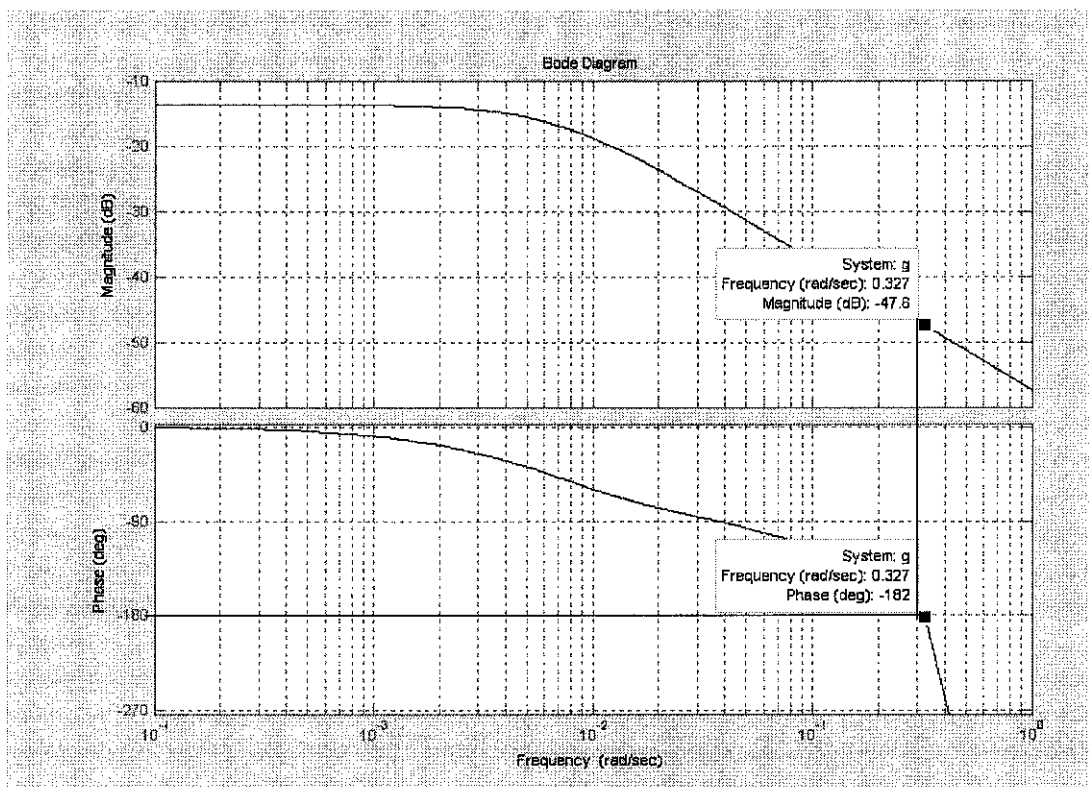| Controller	K_c	T_i	T_d
PI	$K_u / 2.2$	$P_u/1.2$	-
PID	$K_u / 1.7$	$P_u/2.0$	$P_u/8$

Figure 4.10: Bode Diagram for Closed Loop

Calculations of Different Tuning Methods

1.   Ziegler Nichols Closed Loop Oscillation Based Tuning

PI controller

K_c =   0.45 (0.5)   =   0.225

T_I =   110 / 1.2   =   91.66 seconds

PID Controller

K_c =   0.6 (0.5)   =   0.3

T_I =   110 / 1.8   =   61.11 seconds

T_d =   110 / 8   =   13.75 seconds

2.   Tyreus- Luyben

PI controller

K_c =   0.5 / 3.2   =   0.15625

T_I =   2.2 (110)   =   242 seconds

PID Controller

K_c =   0.5 / 2.2   =   0.227

T_I =   2.2 (110)   =   242 seconds

T_d =   110 / 6.3   =   17.46 seconds

3. Bode Plot Tuning Calculation

$$P_u \quad = \quad 2\pi / \omega c \quad = \quad 19.63$$

$$K_u \quad = \quad 1 /Arc \quad = \quad 223.87$$

PI Controller

$$Kc \quad = \quad K_u / 2.2 \quad = \quad 101.76$$

$$T_i \quad = \quad P_u / 1.2 \quad = \quad 16.36 \text{ seconds}$$

PID Controller

$$K_c \quad = \quad k_u /1.7 \quad = \quad 131.69$$

$$T_i \quad = \quad P_u / 2 \quad = \quad 9.82 \text{ seconds}$$

$$T_d \quad = \quad P_u / 8 \quad = \quad 2.45 \text{ seconds}$$

The tuning constants for all three closed loop methods for PI controller are tabulated below.

Table 4.11: PI Tuning Parameters for Closed Loop

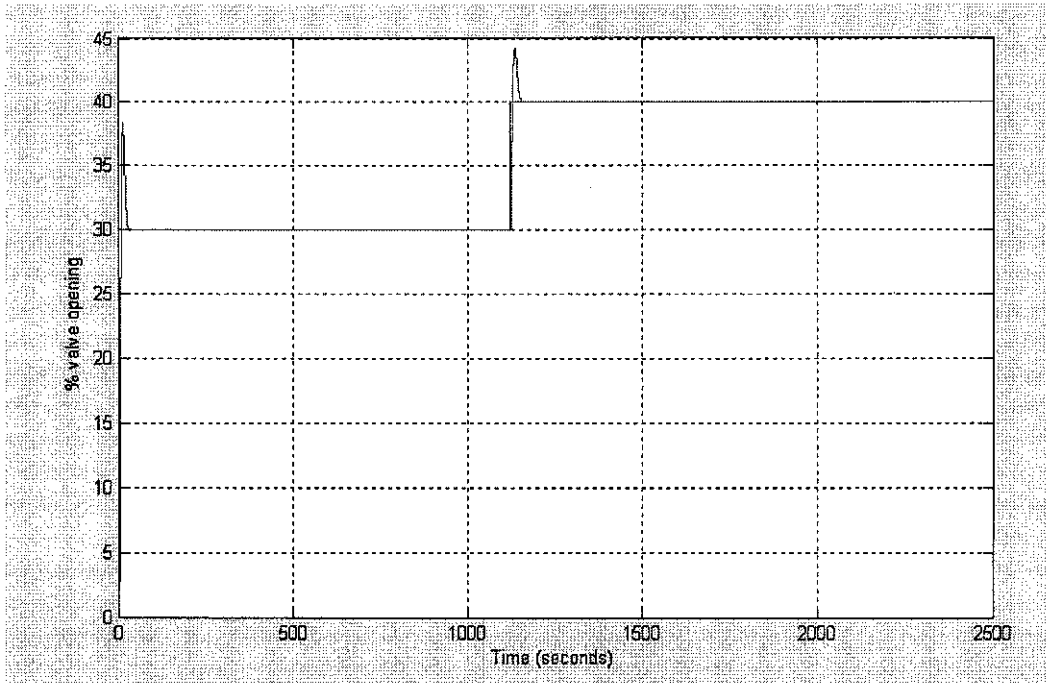| PI Controller	Ziegler Nichols	Tyreus- Luyben	Bode Plot
K_c	0.225	0.15625	101.76
T_i	91.66	242	16.36
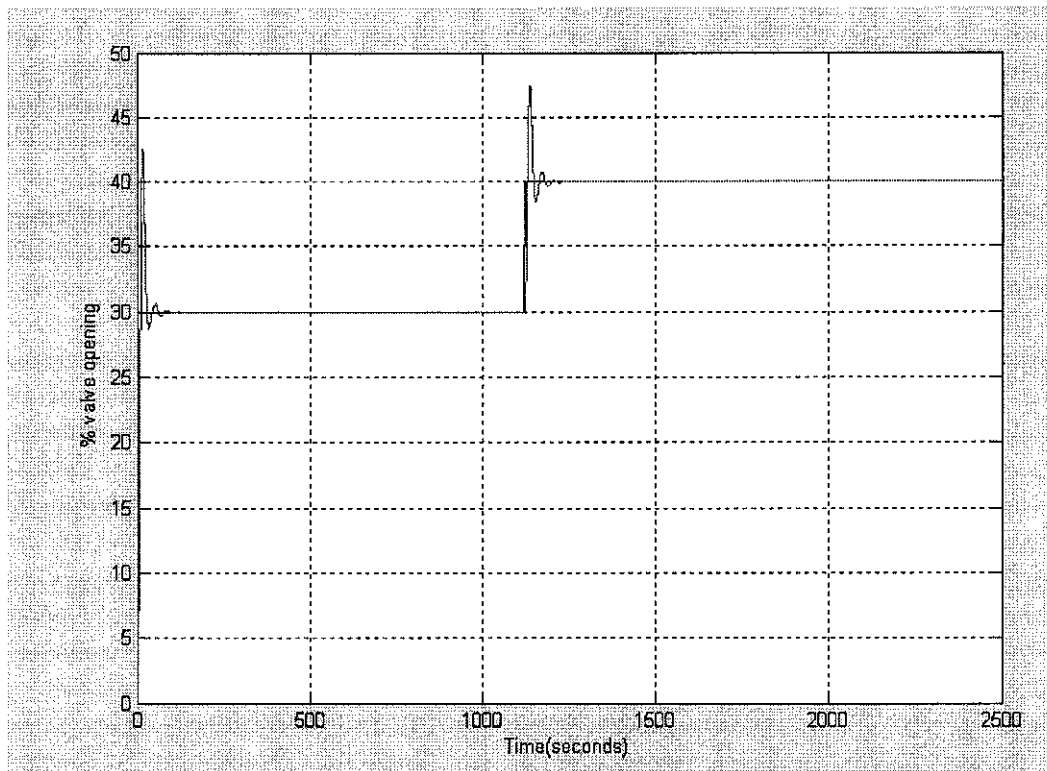
Figure 4.11: Response to unit step set point change using Ziegler Nichols closed loop
bode plot tuning method for PI controller

Table 4.12: Performance values of PI controller for Ziegler Nichols closed loop
tuning

Settling Time	39 seconds
Rise Time	9 seconds
Overshoot	70 %
Decay Ratio	14.28%

Figure 4.12: Response to unit step set point change using Ziegler Nichols closed loop
bode plot tuning method for PID controller

Table 4.13: Performance values of PID controller for Ziegler Nichols closed loop
tuning

Settling Time	29 seconds
Rise Time	8 seconds
Overshoot	40 %
Decay Ratio	0.5 %

The PID controller with Ziegler Nichols bode plot closed loop tuning method
provides the optimum performance and satisfies all the criteria. It has less rise, short
settling time, less overshoot and also satisfies the quarter decay ratio. That's why it is
the best tuned controller.

4.3   MATLAB Simulink Xpc Target

The control block diagram is constructed in Simulink. The necessary blocks that are used to create model are step input, Xpc target scope, and other blocks are added in the application. The suitable parameters are entered for the scope block to view the result. The Simulink output block is added to log the output for analysis. The Xpc target is used to visualize signal while running the plant. The mode is selected as external because the model is connected to the Gaseous Pilot Plant. The block diagram which is connected to Xpc target is shown in figure 4.13 below.
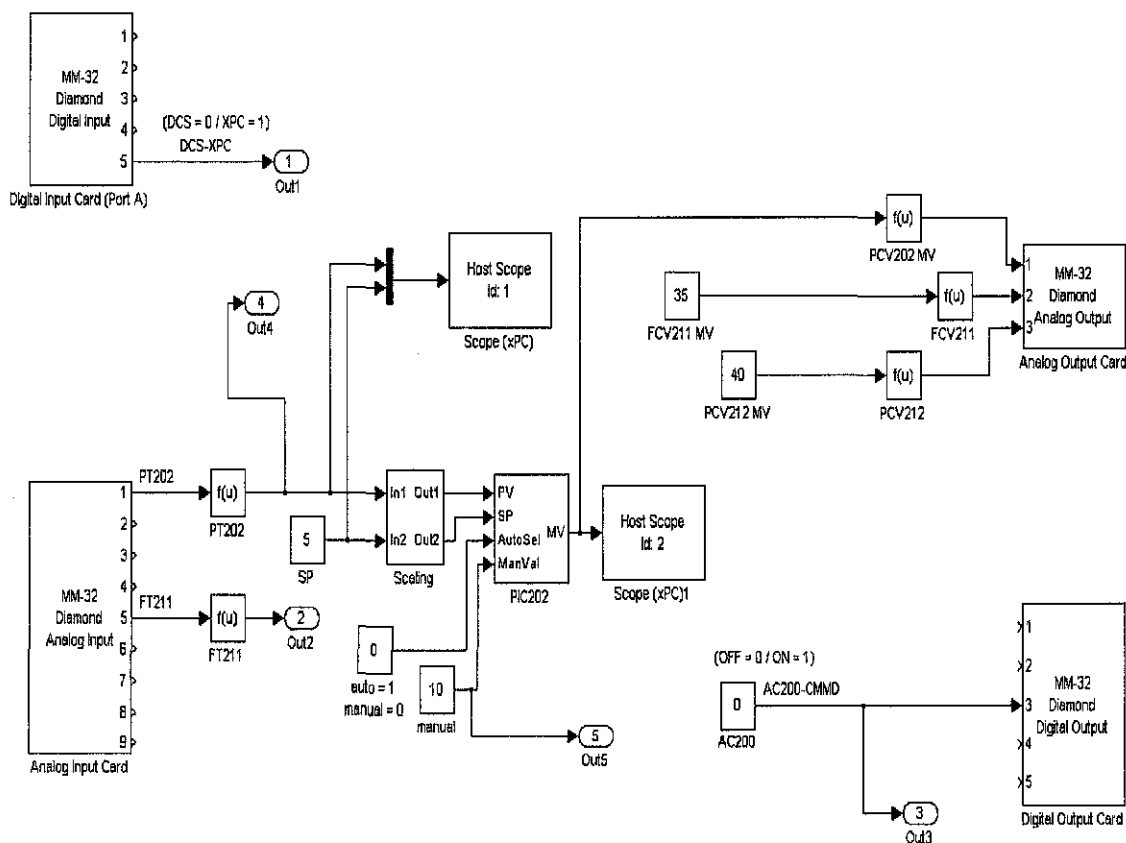

Figure 4.13: Block Diagram using Simulink Xpc Target

The Xpc Target industrial PC acts as server and is connected to the Gaseous Pilot Plant through the Simulink model. The signal from Simulink model is written to the server via Xpc target scope block. When the application is downloaded on the PC it creates the scope automatically.

The Xpc target validates the design in real time. In this project the controller is running in real time connected to the Pilot Plant. Figure 4.14 shows the block diagram of Xpc target box and Figure 4.15 shows the Default ideal PID using difference error approach.
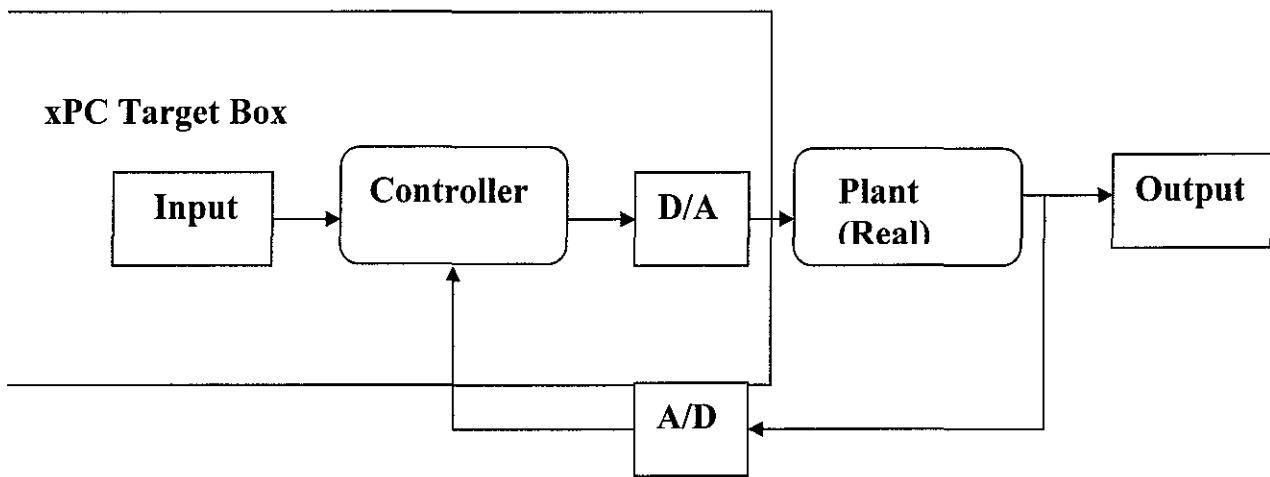

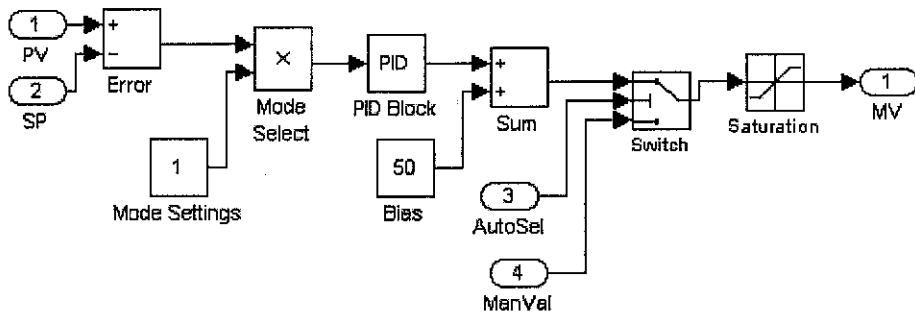Figure 4.14: Block diagram of xPC Target box

Figure 4.15: Default ideal PID using difference error approach.

4.4   Performance Test Using Online Simulation

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

From the real experiment the controlled variable response is observed which resembles the response that is obtained using Simulink block diagram with PID controller.

4.5   Discussion on the Results

The project objective is to model and simulate the gaseous plant for pressure control. Empirical modeling approach is used in this project. Empirical model is a good guide to start with the project. It is easy to implement and easy to evaluate if the model data is collected from the plant. Therefore, it is valid to use this model from the plant as an output predictor to the current pressure process control for the system.

After getting the model transfer function the second task which involves PID controller tuning and performance test was also done. Several analyses were done for the performance of the controller in Simulink model. The tuning constants were calculated for the system and implemented in actual pilot plant.

Finally the third task was to develop the real time MATLAB Simulink xPC Target an interface to the experiment. The xPC target was used in this project to visualize the signal while running the target application to observe the process variable (system response) due to set point change. The real time simulation from the plant using the Simulink xPC target, the process variable value was almost accurate comparing to the simulation with Simulink. The accuracy was also observed in the values in terms of rise time, overshoot, settling time and decay ratio of the plant.

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

To model and simulate the pressure process control a good understanding is required to determine the important parameters and characteristics. The selection of input and output variable was made as it is important to study 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 was made according to the requirement. The most popular and simplest technique used is empirical modeling method.

The PID controller is very common system used in the industry nowadays. The PID controller is chosen as the controller for the pressure control in this project. The PID controller is designed from the data obtained in the experiment on the pilot plant and the calculation of tuning constants on the basis of the real time data from the plant using different techniques.

The design of a controller for the gaseous pressure control was made in this project on real time operation. The dynamics and variability was observed on the on line control.

5.2   Project Recommendations

The system is a first order with dead time and is single input single output system. It only has the single input %valve opening (inlet flow rate) and single output which is the tube outlet pressure. The project can be extended with Multiple Inputs and Multiple Outputs (MIMO) system. It analyzes the various performances and the effect of disturbances to the system.

The recommendation to involve the MIMO system is because with the multiple inputs and multiple outputs is that the system becomes more complex and the introduction Fuzzy Logic Controller (FLC) in the system reduces the need for operator and the use of tuning constants for the system because FLC takes corrective action automatically.

The problem that occurred in doing the experiment many times at initial stage was because of doing the experiment on the pilot plant without restarting it. So it is recommended to do only one experiment and after it must be restarted for another experiment. In the PID controller analysis, it is recommended that all the tuning parameters calculated with different techniques are tested at actual point and by doing this more validations can be made easily.

The Simulink block diagram developed is quite useful for analysis of the pressure behavior and therefore it is recommended to transfer the diagram in nice and clearer format for understanding. It is also recommend to take the most of Graphical User Interface (GUI) of MATLAB for clearer graphs and getting the graphs in developed environments to simplify the process designing for understanding.

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APPENDICES

Appendix I   Acronyms

**Appendix II**   Figures

APPENDIX I

Acronyms

Accuracy: The difference between the reading of an instrument and the true value of what is being measured, expressed as percent of full instrument scale.

Automatic Control: The process of using the differences between the actual value and desired value of a variable to take the corrective action without human intervention.

Bias: It is the set value for control valve incases of failure of instrument air and is provided by the process operator for safety purpose.

Bumpless Transfer: Change from manual to automatic control or vice versa, without change in control system.

Control Valve: Control valves or proportional valves are power operated devices used to modify fluid flow or pressure rate in a process system.

Calibration: The procedure for determining, correcting or checking the absolute values to do the measurement on a measuring instrument.

Desired Value: The value of the controlled variable which is to be maintained.

Fluid: Any gas or liquid

Flapper Nozzle: A combination of elements in a pneumatic controller.

Instrumentation and Control System: Include pressure / temperature / flow / level switches, transmitter, controller, regulator, recorder, indicator, gauges.

Measured Variable: Analogous to controlled variable when it is used in connection with the control applications.

Measuring Element: The primary device for measuring the process variable.

Offset: A sustained deviation of the controlled variable from the set point.

Open Loop: A system in which no comparison is made between actual the actual value and desired value of a process variable.

Orifice Plate: A thin, circular metal plate with an opening in it. It is used for measuring flow rate.

Overshoot: The amount by which a changing process variable exceeds the desired value as changes occur in a system.

Process Reaction Curve: A record of the reaction of a control system to a step change.

Set Point: The position at which the control point setting mechanism is set. This is the same as desired value of the controlled variable.

Span Check for Transmitter:	When there is signal present (it can be 8 mA, 12 mA, 16 mA and 20 mA) which is calculated as the output current and is the span check for transmitters..
Storage Tank:	Tanks or vessels including atmospheric and low pressure tanks / vessels used for storage and subsequent handling of petroleum or petroleum products.
Turbulent Flow:	Rapid flow of a fluid during which the Reynolds number is higher than 4000.
Zero Check for Transmitter	When there is no signals present the output current should be 4 mA as checked with the millimeter for transmitter.

APPENDIX II

FIGURES
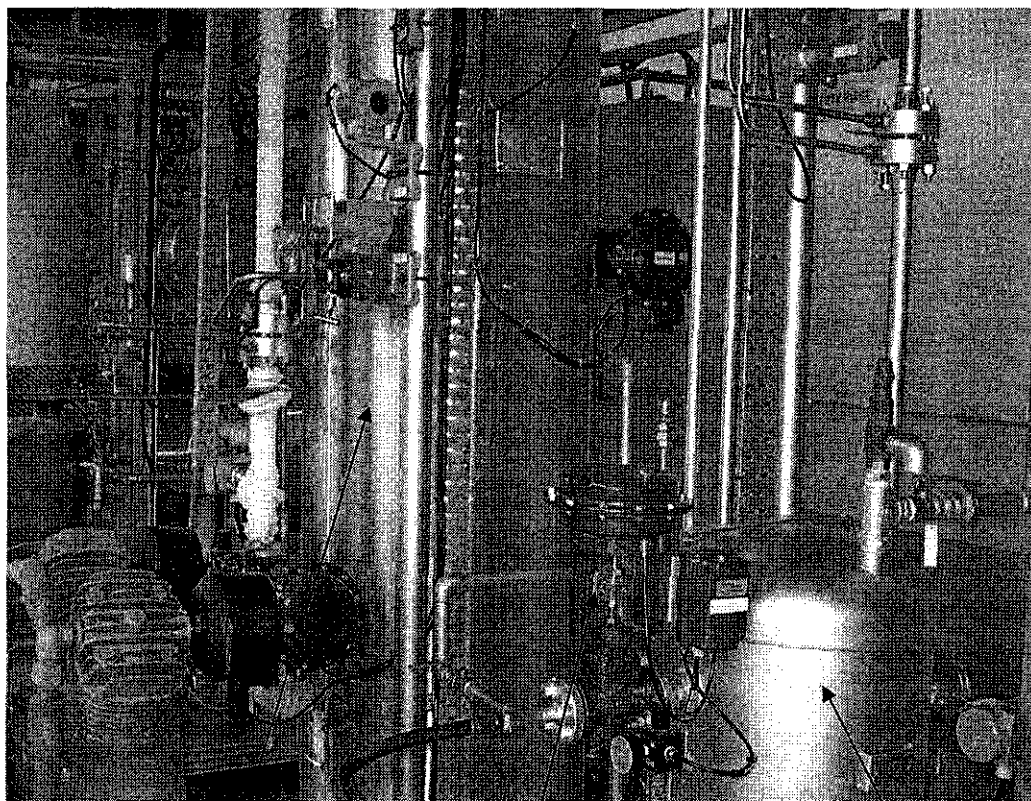
Control Valve

Buffer Vessel VL (202)

Main Vessel VL (212)

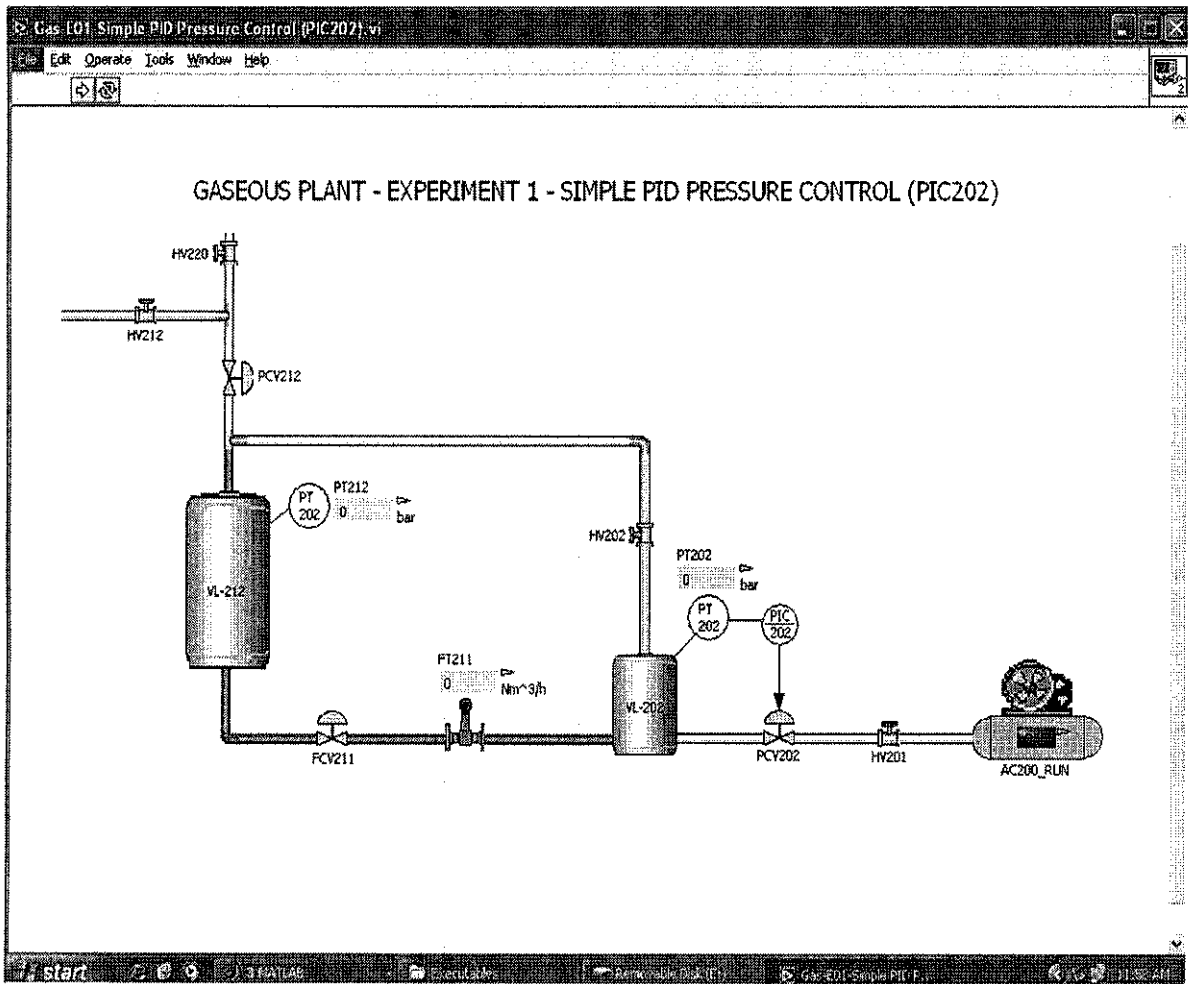Figure 1: Gaseous Pilot Plant

Figure 2: Simplified Pressure Diagram for Gaseous Pressure Plant
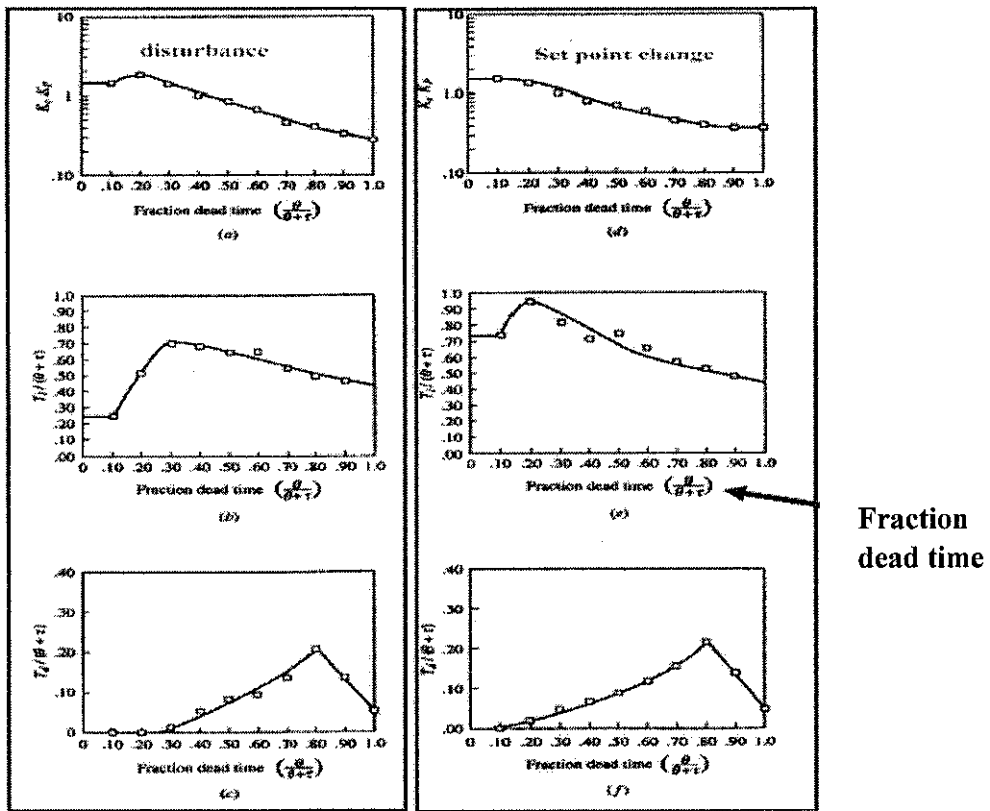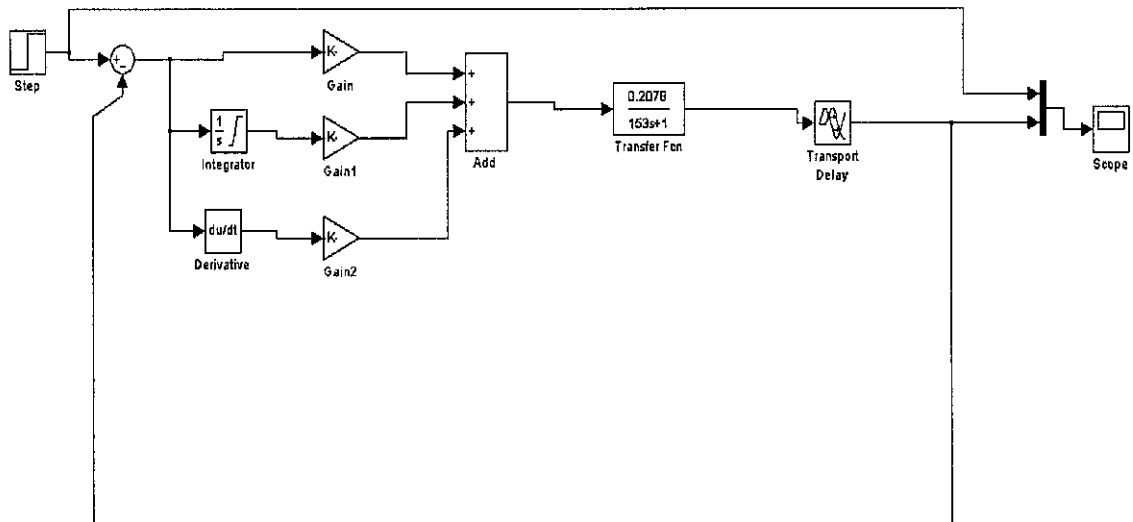
Figure 3: Ciancone Correlations for PID Controllers

Figure 4: Closed Loop Block Diagram with PID Controller

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