# STUDY OF THE PARAMETER INTERACTION IN A CLOSED_LOOP CONTROL OF A PROCESS INTENSIFICATION SYSTEM

**NOR HAIRIE BIN RAMLI**

**UINVERSITI SAINS MALAYSIA**

**2022**

# STUDY OF THE PARAMETER INTERACTION IN A CLOSED_LOOP CONTROL OF A PROCESS INTENSIFICATION SYSTEM

**by**

**NOR HAIRIE BIN RAMLI**

**Project submitted in partial fulfilment of the requirement for the degree of**

**Bachelor of Chemical Engineering**

**July 2022**

## ACKNOWLEDGEMENTS

This final year project is for the completion of degree of Bachelor of Chemical Engineering. Support and commitments from several individuals have contributed to the success in completion of my final year project. Therefore, I would like to show my highest gratitude and appreciation to all the individuals involved for their contribution throughout the project.

First and foremost, I would like to express my sincere appreciation to my supervisor, Professor Dr. Mohd Roslee Othman for his invaluable guidance, insightful supervision and endless encouragement throughout this project. Without the guidance and advice from him, this final year project will not be done completely on time.

To begin with, I would want to thank Allah S.W.T. for allowing me the strength to complete this report for my thesis. Then, I would also want to express my thanks to my supervisor, Dr. Syamsul Rizal Abd Shukor, whose expertise, and experience assisted me in completing my report. I would also want to thank my parents and friends for their moral support throughout the preparation of this report.

Once again, I would like to thank all the people, including those who I might have missed out mentioning their names but have helped me directly or indirectly in accomplishment of this project. Thank you very much.

Nor Hairie bin Ramli

July 2022

## TABLE OF CONTENTS

# LIST OF TABLES

# LIST OF FIGURES

## LIST OF ABBREVIATION

| Symbol | Description |
|---|---|
| $\tau_p$ | Time constant of Process |
| $\tau_v$ | Time constant of Valve |
| $\tau_m$ | Time constant of Sensor |
| $\theta$ | Time delay |
| $G_c$ | Transfer Function of Controller |
| $G_v$ | Transfer Function of Valve |
| $G_p$ | Transfer Function of Process |
| $G_m$ | Transfer Function of Sensor |
| PID | Proportional Integral Derivative Controller |
| PIDD | Proportional Integral Derivative Derivative Controller |
| PIDDD | Proportional Integral Derivative Derivative Derivative Controller |

## ABSTRAK

Dalam Intensifikasi Proses, peralatan inovatif dan penyepaduan dua atau lebih unit ke dalam satu peralatan digunakan. Ini mempunyai pengaruh yang signifikan terhadap teknologi hijau, loji kimia yang lebih selamat, lebih murah dan unggul. Unit pengecilan dan hibrid termasuk kawasan intensifikasi proses. Pengecilan memerlukan pengecilan saiz, yang mempercepatkan masa tindak balas proses. Reaksi pantas menyebabkan mekanisme kawalan biasa tidak mampu mengawalnya. Untuk memenuhi keperluan ini, teknik kawalan baharu bersama penggerak atau penderia baharu mesti direka. Dalam kerja ini, pendekatan Sintesis Langsung telah digunakan untuk membangunkan pengawal untuk sistem yang dipergiatkan, dengan mengambil kira semua elemen gelung kawalan serta beberapa anggaran kelewatan masa. Anggaran pade kelewatan masa menghasilkan struktur pengawal PIDDD, manakala anggaran pengembangan siri Taylor bagi kelewatan masa menghasilkan struktur pengawal PIDD. Variasi pemalar masa injap dan penderia untuk setiap pemalar masa proses mendedahkan interaksi antara keadaan proses dan unit proses untuk kedua-dua pengawal, yang diterokai. Untuk membolehkan perbandingan prestasi pengawal dan interaksi parameter dengan borang pengawal PID, susunan pengurangan pengawal juga dibekalkan. Prestasi diukur menggunakan Ralat Mutlak Integral (IAE). Untuk pemalar masa proses pantas $\tau_p$=0.01s, julat pemalar masa injap yang dibenarkan, pemalar masa penderia, dan juga kelewatan masa adalah 10 kali lebih tinggi dan 10 kali lebih rendah daripada nilainya, seperti yang ditentukan oleh ketiga-tiga pengawal. Pemalar masa proses pertengahan dalam penyelidikan ini, $\tau_p$=1S, mungkin mengambil nilai yang sama atau separuh nilainya. Pemalar masa proses perlahan $\tau_p$=10s menghasilkan tindak balas yang tidak stabil dan tidak memuaskan merentasi julat pemalar masa yang sama.

**STUDY OF THE PARAMETER INTERACTION IN A CLOSED_LOOP CONTROL OF A PROCESS INTENSIFICATION SYSTEM**


## ABSTRACT

In Process Intensification, innovative equipment and the integration of two or more units into a single piece of equipment are used. This has a significant influence on green technology, safer, cheaper, and superior chemical plants. Miniaturization and hybrid unit include the area of process intensification. Miniaturization entails the diminution of size, which expedites the process's reaction time. The rapid reaction renders the usual control mechanism incapable of controlling it. To meet this need, new control techniques together with a new actuator or sensor must be devised. In this work, the Direct Synthesis approach was utilised to develop the controller for an intensified system, taking into consideration all of the control loop's elements as well as several time delay approximations. Pade approximation of time delay yields PIDDD controller structure, while Taylor series expansion approximation of time delay yields PIDD controller structure. Variation of the time constant of the valve and sensor for each process time constant reveals the interaction between process state and process units for both controllers, which is explored. To enable a comparison of controller performance and parameter interaction with the PID controller form, the reduction order of the controller was also supplied. Performance is measured using Integral Absolute Error (IAE). For a fast process time constant $\tau_p$=0.01s, the allowable range of valve time constant, sensor time constant, and even time delay is 10 times higher and 10 times lower than its values, as determined by all three controllers. The intermediate process time constant in this research, $\tau_p$=1S, may take on the same value or one-half its value. Slow process time constant $\tau_p$=10s yields an unstable and unsatisfactory response across the same time constant range.

# CHAPTER 1

# INTRODUCTION

## 1.1 Research Background

The term "process intensification" refers to the practice of decreasing the size of the equipment or plant used in a process while keeping the same output level. Apart from the idea of intensification is the improvement of a process technique, as well as the consolidation of many pieces of equipment into a single, multipurpose piece of machinery. The major goal in introducing the idea was to reduce the size of the equipment, which ultimately led to the construction of micro-reactors and other unique pieces of machinery.

These can function at higher efficiencies under optimal conditions while maintaining a high level of product quality. A drastic decrease in size by a factor of 100 or more opens the door to the possibility of developing environmentally friendly and sustainable technology, such as one that uses less energy, generates less waste, pollutes less, and is intrinsically safe. The reduced size of the equipment will result in the chemical plant being smaller in size, which will save space, reduce the company's effect on the environment, and improve control and automation.

However, process control of the intensified system, whether it be a single piece of intensified equipment or the plant as a whole, becomes exceedingly difficult. It would not be possible to make full use of the advantageous aspects of the Pl idea if these intensified systems were not thoroughly controlled and automated. When focusing on novel equipment as a representation of process intensified equipment, the control system is confronted with the fact that the reduction in size of this equipment leads to a fast dynamic behavior that cannot be controlled by using conventional control systems directly.

This behaviour is impossible to control. Despite this, the traditional, restricted interaction that exists between process design and process control seems to be inadequate when attempting to cope with process intensification. In order to realize the objective of having a precise control system of P, the control system has to initiate itself at the very beginning of the control loop's most fundamental stage. The aspect of the control loop's dynamic behaviour as well as how it influences the performance of the control loop is something that has to be thoroughly deliberated on and investigated.

## 1.2 Problem Statement

When the spatial domain of an equipment or system that has been intensified is reduced, this results in a quicker response from the system, which in turn means that the process time constant is shortened. As a result, there is a requirement for a control system that is able to precisely control the quick dynamic of the system. In order to accomplish this objective, the investigation into the control system must start with a dynamic analysis from the very beginning, incorporating controller design into the process. In addition, the interaction between the process state, which is the dynamic of the process itself, and the process unit, which is the element that made up the control loop and may include an actuator, a valve, or a sensor, becomes significant in equipment that has been intensified.

This interaction is important because it determines the significance of the control loop. Therefore, while building a controller or control system, it is vital to take into consideration all of these elements in addition to the dynamic behaviour of each process unit. The quantity of time delay is still another challenge for the process control. Due to the quicker responsiveness of the process in comparison to these process units, the measurement delay and action delay that are caused by the final control element are very important in process intensifying equipment. This problem is solved when using traditional equipment or a conventional system since the reaction time of the process is slower. Because of the rapid response of the process, the capabilities of the current sensor and actuator are restricted in conventional systems, and it is not possible to use them in the control systems of intensified equipment or systems because of this restriction. It is necessary for the sensor or actuator to have a reaction time that is comparable to or even quicker than that of the process. The development of a modified version of the new kind of sensor and actuator offers a great deal of cause for optimism.

## 1.3 Objectives

The objective of this research:

i. To investigate the relationship between the process state and the process unit.

ii. To investigate the dynamic relationship between the many components that comprise process intensification.

iii. To investigate the influence of time delay approximation on controller design.

iv. To investigate the influence that the reduction order controller has on the dynamics of the process.

# CHAPTER 2

# LITERATURE REVIEW

## 2.1 Process Intensification

Process Intensification (PT) is a technique that may be described as the act of making a significant decrease in the size of a chemical plant in order to achieve a certain level of output (Ramshaw, 1995). Either decreasing the size of individual pieces of equipment or decreasing the total number of pieces of equipment needed may accomplish the decrease. The degree of contraction or decrease is more than one hundred times the component. The term "process intensification" may also refer to any innovation in chemical engineering that results in a technology that is noticeably more compact, environmentally friendly, and efficient in terms of energy use (Stankiewicz and Moulijn, 2000). There are a few other ways to define what is meant by the term "process intensification." One of these ways refers to the utilization of technologies that, in place of large, high-priced, or energy-intensive equipment or processes, smaller, more affordable, and more effective alternatives are used. These alternatives may also combine multiple processes into a smaller number of devices (Tsouris and Procelli, 2003).

As a result of these benefits and advantages, the field of process intensification within chemical engineering evolves into something unique and intriguing. The primary goal of the Pl system is to maximize output while minimizing both energy consumption and the amount of waste and by-product created. Colin Ramshaw made a number of important technological advancements throughout the 1970s and early 1980s at Imperial Chemical Industries (ICI). These advancements led to a breakthrough in the field of process intensification, which is now undergoing fast growth. Since 1994, Andrzej Stankiewicz, who holds the position of Professor of Process Intensification at the Delft University of Technology, has been carrying on research into the Pl system.

Process intensification may be motivated in part by the desire of chemical and allied businesses, as well as process engineering, to solve or get rid of difficulties. One of the technologies that helps fulfil the evolution of chemical and related industries and process engineering is process intensification. This leads to the idea of latest technology or unit operations that are multipurpose and more economical, such as catalytic distillation. Process intensification is one of the innovations that helps fulfil the progression of chemical and related industries and process engineering (Charpentier and Trambouze, 1998).

### 2.1.1 Monoliths reactor

In-line-monolithic reactor (ILMR) is a novel gas liquid catalytic reactor. According to this concept, a horizontal catalytic reactor is an integral part of the pipeline. Conventional monolith reactors have many drawbacks, including mass and heat transfer limitations, low catalyst surface area, and high pressure drop. The reactor concept solves these problems. Based on experimental testing, this reactor's size reduction and reaction time is seconds, compared to 20 minutes for a conventional industrial packed-bed reactor. This reduces ILMR volume and plant size (Andrzej, 2001).

### 2.1.2 Reactive Separation

Multifunctional reactors combine at least one function (typically a unit operation) that would normally be done in a separate piece of equipment. Most multifunctional reactors combine reaction and separation. Integration reduced plant size and cost (the height of the plant with respect to the conventional technology decreased almost 2.5 times). Reactive distillation, adsorption, membrane reactor, and other reactive separation are examples. The process intensification decreased reactive distillation equipment from 28 to 3. By separating equilibrium educts and products, reactive absorption increases conversion and yield. Membrane reactor reaction time is less than conventional (Andrzej, 2003). The chosen combination must have a big operating window and be flexible and controllable. To achieve process intensification, technological barriers to greater use of reactive separations must be overcome (Andrzej, 2003).

**2.2 Difficulties of Process Intensification**

Process intensification delivers safety, environmental, and economical advantages. Due to hurdles, process-intensified systems are sluggish to become industrial. Intensive equipment hasn't been demonstrated on a wide scale; thus, industrial implementation is challenging. Most companies won't replace traditional reactors with microreactors. Their biggest fear is that more equipment may malfunction, reducing productivity. Full Process Intensification advantages can only be realized by intensifying the entire plant. They also need proof of the system's success. Due to its proven safe operation and dependability, conventional equipment with comprehensive design and code gains increasing industry attention (Ramshaw et al., 2008).

Other restrictions must be eliminated or reduced. One is the present tendency of industrial R&D, which focuses on new product production rather than process intensification. Chemical manufacturers are less interested in unique or integrated equipment. Insufficient activity. Due to chemical engineers' unfamiliarity with the idea, process intensification has grown slowly. Process Intensification is well-studied and developed in the lab, but there is no instrument to translate it to industrial scale (Andrzej and Moulijn, 2002).

Lack of simulation, model, and assessment tool for intensified equipment or procedure caused industries choose traditional, more mature equipment. The function of universities as research centers must be highlighted by introducing this notion in chemical engineering curriculum and courses. Chemical engineering curriculum need updating (Andrzej and Moulijn, 2002).

## 2.3 Control Strategies on Process Intensification System

### 2.3.1 Miniaturization

System volume reduction while maintaining product throughput speeds up process responses. The time constant decreases. The smaller equipment is more susceptible to disturbances, which complicates process management and operation. To reject unwanted disturbances quickly, controller, actuator, and sensor dynamic should be equivalent or less than reaction time. When building a control structure for an intensified system, the dynamics of all control loop parts should be considered to properly integrate the control and operation of the system. Effective control of increased processes sometimes needs a new or modified sensor. Sensors must meet quick response and sophisticated control needs, yet present instrumentation is sluggish, imprecise, and off-line (Nikaevi et al., 2011).

### 2.3.2 Anisotropic Membrane

The hybrid control system has fewer degrees of freedom. This decrease is due to integrating many units into one that shares process variables. The challenge limits design and control of these machines. Integration of the system makes it harder to smooth out operational disturbances. In addition, the integration system becomes more complicated owing to increasing interaction between process variables, resulting in non-linear behavior that is difficult to regulate (Nikaevi et al., 2011).

Table 2.1 The opportunities and challenges of process control for PI

| **Type** | **Opportunities** | **Challenges** |
|---|---|---|
| Process control for PI | - Molecular efficiency. Micro-level process actuation and control<br><br>- Force and surface area. New driving forces offer meso and macro actuation. | - Higher driving pressures, particular equipment, and shorter process time constants need quicker control.<br><br>- "Field" instrumentation, particularly sensors, must reduce. Combining functions minimises DOF (actuation surfaces lead to compact controllability).<br><br>- Partially synergetic.<br><br>- More interactivity, smaller window. |

## 2.4 Advanced Control Strategies of Process Intensification System

### 2.4.1 Model Based

Shukor and Tham (2003) researched the dynamic of control loop component for process intensification systems. Analyze process status and unit interaction. The controller for an intensified system was designed using direct synthesis to account for process unit dynamics. The controller innovation attributes to a PIDD controller in series with a first-order filter. When the sensor's dynamic behavior is insignificant, this controller reduces to a PID controller. When the transmitter's dynamics are considerable, anticipatory intervention is needed, leading to the second-order derivative term. A low-pass filter is needed to reduce high-order derivative activity (Shukor and Tham, 2003).

Integral absolute error measures system performance (IAE). 3-D graphic summarizes performance assessment. Studies found that only process delay and transmitter affect control performance. Synthesis Equation controllers cancel process and actuator dynamics. Since forward route dynamics are absent, the finding applies to all closed-loop systems with the same controller architecture. This dispels early concerns about current actuators' capacity to control enhanced systems (Shukor and Tham, 2003).

### 2.4.2 Internal Model Control

Another contribution compares IMC and PID controllers. IMC technique forms PID-$D^2$ controller This controller is similar to classical PID, PID-$D^2$, and industrial PID. IMC may be used to establish optimal parameters for traditional PID and industrial PID algorithms for enhanced processes. Cutting controller order terms is ad hoc. Simulated set-point tracking performances of these controllers, including the original PID+$D^2$, were comparable. Industrial PID has superior disturbance rejection, particularly with a short time-delay (Jones and Tham, 2006).

## 2.5 Difficulties in Controlling Process

Process intensification control systems might be hybrid or miniaturized or low volume. Hybrid systems combine equipment, such as catalytic distillation. Strong interplay of process variables and loss of degree of freedom make process control for this equipment more challenging. Slower control than low-volume equipment. Low-volume equipment includes micro-reactors and spinning disc reactors. As more persistent control issues occur, such as instrumentation problems, instrumentation error to system, controller tuning relation, difficulty utilizing digital control, and high time delay problems, the equipment's control system becomes more important (Barzin et al., 2007).

The enhanced unit lacks standard instrumentation like valves and sensors. Conventional instrumentation has a slower reaction time and can't keep up with the increased unit's speed. For example, a sluggish control valve leads the control system to fail. In traditional control systems, the process time constant is dominant because the time constants of other units may be ignored. The measuring sensor also limits the enhanced system. Current measurement sensors can't measure fast-reacting systems in intensified systems (Barzin et al., 2007).

In decreasing instrumentation error to system, dynamic process unit behaviour must also be addressed. In intensified systems, the controller's design must consider the ultimate control element's dynamics. Good controllers for amplified systems take each element's dynamic behaviour into consideration. The dynamic behaviour of the constituents may affect the enhanced system's short residency duration (Shukor and Tham, 2003).

Conventional controller tuning is extensively developed, but not adequate for intense systems. Conventional controller tuning neglects certain measuring elements. Due to additional considerations, it's not ideal for intensive systems. New controller tuning approaches are needed (Barzin et al.,2007). The enhanced system may have excessive delay. The time delay exceeds the process time constant, making control impossible.

## CHAPTER 3

## METHODOLOGY

### 3.1 Overview of Research Methodology

**Direct Synthesis Method of controller designed (Seborg et al., 2004)**

-Pade Approximation of time delay PIDDD controller structure

-Taylor Series Expansion of time delay-PIDD controller structure

**Interaction between process state and process unit**

-tau p=0.01, 1s, 10s

-tau v, tau m,theta = 0.1 tau p-10tau p -JAE

**Reduction order by truncation method**

-PIDD controller to PID controller

-Interaction between process state and process unit comparison of PIDD and PID controllers

## 3.2 Controller Design Method

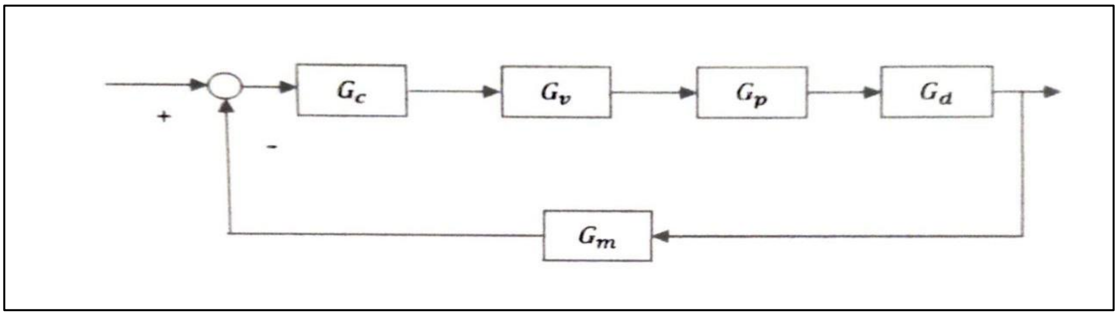


Figure 3.1 Feedback control loop controllers

Feedback control loop controllers are designed via direct synthesis (Seborg et al., 2004). The design considers the dynamic behavior of feedback loop components including valves, sensors, and time delays. First-order transfer function represents the component's dynamic behavior. First order with time delay represents process dynamics.

$$G_v = \frac{1}{\tau_v s + 1}, G_p = \frac{e^{-\theta s}}{\tau_p s + 1}, G_m = \frac{1}{\tau_m s + 1}, G_d = e^{-\theta s}$$

Where G, represent the valve, G, sensor, G, process and Gd is the time delay. The close loop transfer function for set point changes with process delay was derive as

$$\frac{Y}{Ysp} = \frac{G_c G_v G_p G_d}{1 + G_c G_v G_p G_m G_d}$$

The desired close loop transfer function and the time delay as

$$\frac{Y}{Ysp} = \frac{e^{-\theta s}}{\tau s + 1}$$

$$G_d = e^{-\theta s}$$

To examine their impact on the controller structure, several time delay approximation methods are applied. Below is a detailed explanation of how the controller was derived using a time delay approximation.

### 3.2.1 First order Pade approximation for time delay

In the beginning, the Pade Approximation of the Time Delay was chosen since it contains more terms to approximate the value of or in comparison to the Taylor Series Expansion (Seborg et al., 2004). Following the derivation, one arrives at the Proportional-Integral-Derivative-Derivative-Derivative (PIDDD) controller structure, which includes a first order low pass filter denoted by $G_f$. the configuration can be referred to figure 3.1.

$$e^{-\theta s} = \frac{1 - \frac{\theta}{2}s}{1 + \frac{\theta}{2}s}$$

$$G_c = K_c G_f\left(\tau_{d_3}s^3 + \tau_{d_2}s^2 + \tau_{d_1}s + 1 + \frac{1}{\tau_i s}\right)$$

$$K_c = \left(\tau_v + \tau_p + \tau_m + \frac{\theta}{2}\right)$$

$$G_f = \frac{1}{\bar{\tau}\bar{\tau}_m\bar{s} + (\bar{\tau} + \bar{\tau}_m + \frac{\theta}{2})}$$

$$\tau_i = \left(\tau_v + \tau_p + \tau_m + \frac{\theta}{2}\right)$$

$$\tau_{d_1} = \frac{(\tau_v\tau_p + \tau_p\tau_m + \tau_m\tau_v) - 0.5\theta(\tau_v + \tau_p + \tau_m)}{\left(\tau_v + \tau_p + \tau_m + \frac{\theta}{2}\right)}$$

$$\tau_{d_2} = \frac{\tau_v\tau_p\tau_m + 0.5\theta(\tau_v\tau_p + \tau_p\tau_m + \tau_m\tau_v)}{\left(\tau_v + \tau_p + \tau_m + \frac{\theta}{2}\right)}$$

$$\tau_{d_3} = \frac{0.5\theta\tau_v\tau_p\tau_m}{\left(\tau_v + \tau_p + \tau_m + \frac{\theta}{2}\right)}$$

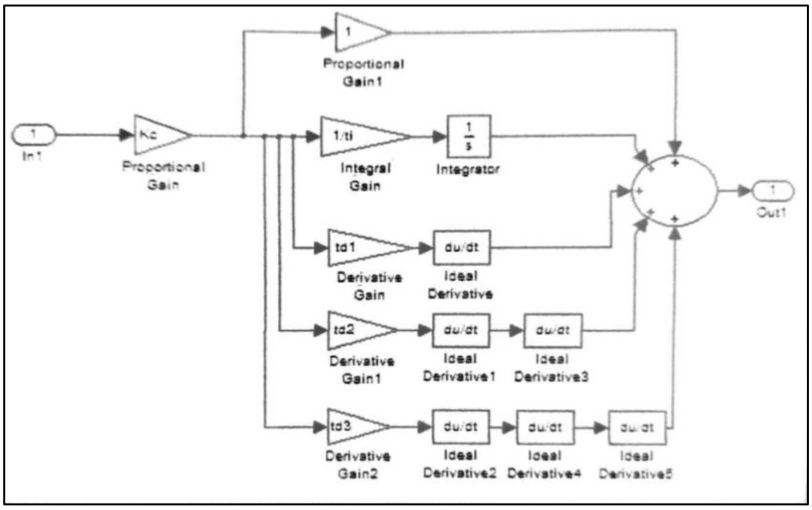


Figure 3.2 PIDDD Controller in Simulink

### 3.2.2 Expansion of Taylor Series for time delay

After then, an estimate of the time delay based on the Taylor Series Expansion is utilized so that additional investigation may be done. This straightforward approximation of time delay techniques led to the development of a controller structure known as proportional-

$$e^{-\theta s} = 1 - \theta s$$

$$G_c = K_c G_f(\tau_{d_2} s^2 + \tau_{d_1} s + 1 + \frac{1}{\tau_i s})$$

$$K_c = (\tau_v + \tau_p + \tau_m)$$

$$G_f = \frac{1}{\tau \tau_m s + (\tau + \tau_m + \theta)}$$

$$\tau_i = (\tau_v + \tau_p + \tau_m)$$

$$\tau_{d_1} = \frac{(\tau_v \tau_p + \tau_p \tau_m + \tau_m \tau_v)}{(\tau_v + \tau_p + \tau_m)} \qquad \tau_{d_2} = \frac{\tau_v \tau_p \tau_m}{(\tau_v + \tau_p + \tau_m)}$$

integral-derivative-derivative, or PIDD for short (Shukor and Tham, 2003). The controller's configuration for Simulink can be referred in figure 3.2.

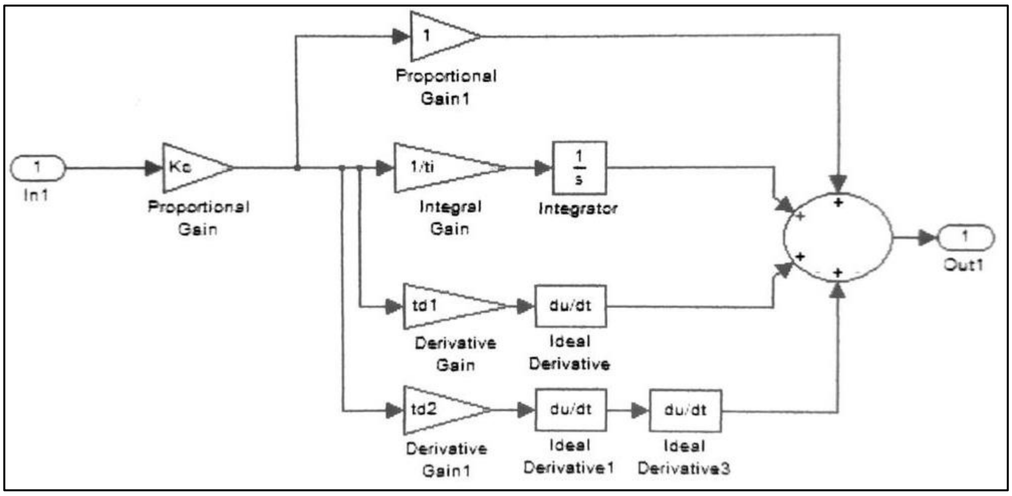


Figure 3.3 PIDD Controller in SIMULINK

### 3.3 Study the Interaction between Process State and Process Unit

#### 3.3.1 Parameter interaction

Process state is represented by the order of the transfer function. Process unit consist of valve, controller, and sensor. The interaction between process state and process unit is study by focusing on the time constant at each unit. The time constant will be varying according to table below.

Table 3.1: Configuration of the time constant

| $\tau_p$ | $\tau_v$ | $\tau_m$ | $\theta$ |
|---|---|---|---|
| 0.01s | vary | constant | constant |
| 1s | constant | vary | constant |
| 10s | constant | constant | vary |

Table 3.2: The ranges of time constant used

| $\tau_p$ | $\tau_v$ | $\tau_m$ | $\theta$ |
|---|---|---|---|
| 0.01s | $0.1\tau_p < \tau_v$ or $\tau_m$ or $\theta < 10\tau_p$ | | |
| 1s | $0.1\tau_p < \tau_v$ or $\tau_m$ or $\theta < 10\tau_p$ | | |
| 10s | $0.1\tau_p < \tau_v$ or $\tau_m$ or $\theta < 10\tau_p$ | | |

This simulation uses three process time constants: $\tau_p$=0.01s, $\tau_p$=1s, and $\tau_p$=10s. $\tau_p$=1s and $\tau_p$=10s describe an extremely quick reaction and a nominal process, respectively (EnTech, 1998). Other component time constant is $0.1\tau_p$ to $10\tau_p$. Within ranges, component time constants are varied to determine their influence on process time constant. Smaller and bigger $\tau_v$, $\tau_m$ or $\theta$ values should provide different results. The simulation is done in SIMULINK software. The simulation uses a fixed-step integration interval of 0.001s to prevent numerical errors.

### 3.3.2 Performance Behavior

The performance behavior of the interaction is analyzing by integral error criteria which is integral of the absolute value of the error (IAE). (Seborg et al., 2004)

$$IAE = \int_0^\infty |e(t)|dt$$

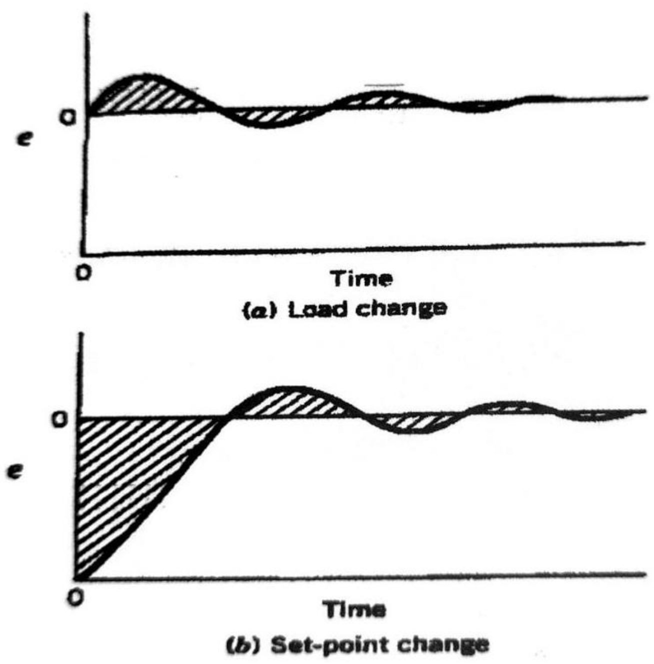


Figure 3.4 Graphical Interpretation of IAE

This penalizes all errors equally in simulation, regardless of direction, and the error can be measured directly from controller input. The error is represented by a shaded area between the response curve and the line of the set point's value, or it can be defined as the sum of the areas above and below the set point. The IAE value of each $\tau_v$, $\tau_m$ or $\theta$ value in this simulation is listed with the accompanying $\tau_p$ values.

### 3.4 Truncation Method of Controller Reduction

When the upper order of the controller equation has to be removed, one approach to do so is through truncation (Jones and Tham, 2006). Based on the value of the IAE produced, one of the controllers will be chosen to truncate into standard Proportional Integral-Derivative (PID) controller in this research. The PIDDD controller structure only requires the removal of one derivative term, while the PIDD controller structure requires the removal of two derivative terms. Utilizing all the aforementioned values and methodologies, the PID controller is studied, and the results are compared.

# CHAPTER 4

# RESULTS AND DISCUSSIONS

## 4.1 Process State and Process Unit Interaction

Three sets of time constant of process, $\tau_p$, with values of 0.01s, 1s, and 10s are chosen to represent quick, medium, and slow processes, respectively. The interaction between process state and process units for each $\tau_p$, value is determined by adjusting the time constant of the valve, $\tau_v$ , the sensor, $\tau_m$ , and the time delay $\theta$ from $0.1\tau_p$ to 10 $\tau_p$. The data are tabulated as follows.

Table 4.1 The values and constants for simulation.

| | Variation of $\tau_v$ | | | Variation of $\tau_m$ | | | Variation of θ | | |
|---|---|---|---|---|---|---|---|---|---|
| $\tau_p$ | $\tau_m$ | $\tau_v$ | $\theta$ | $\tau_m$ | $\tau_v$ | $\theta$ | $\tau_m$ | $\tau_v$ | $\theta$ |
| | | 0.001 | | 0.001 | | | | | 0.001 |
| | | 0.0051 | | 0.0051 | | | | | 0.0051 |
| 0.01 | 0.01 | 0.011 | 0.01 | 0.011 | 0.01 | 0.01 | 0.01 | 0.01 | 0.011 |
| | | 0.051 | | 0.051 | | | | | 0.051 |
| | | 0.011 | | 0.011 | | | | | 0.011 |
| | | 0.11 | | 0.11 | | | | | 0.11 |
| | | 0.51 | | 0.51 | | | | | 0.51 |
| 1 | 1 | 1.1 | 1 | 1.1 | 1 | 1 | 1 | 1 | 1.1 |
| | | 5.1 | | 5.1 | | | | | 5.1 |
| | | 11.0 | | 11.0 | | | | | 11.0 |
| | | 1.1 | | 1.1 | | | | | 1.1 |
| | | 5.1 | | 5.1 | | | | | 5.1 |
| 10 | 10 | 11.0 | 10 | 11.0 | 10 | 10 | 10 | 10 | 11.0 |
| | | 51.0 | | 51.0 | | | | | 51.0 |
| | | 110.0 | | 110.0 | | | | | 110.0 |

By altering $\tau_v$, the interaction between the process time constant, $\tau_p$, and the valve time constant, $\tau_v$, may be used to find the appropriate valve for a given process and the dynamic of the valve that affects the process. The sensor and the time delay both used the same mechanism. This interaction was developed to include several types of controllers, which will be covered further below.

### 4.2 Time Delay Approximation Effect

Using Pade approximation for the time delay and the first order plus time delay transfer function of the valve, sensor, and process, the controller produced a PIDDD structure. PIDD controller is the result of Taylor series expansion estimate of time delay. As described above, both resulting controller architectures are investigated by altering the time constant of each component in the feedback control loop. At time 30s, the change in set point from 0 to 1 is implemented. The graphs of output response and controller output response are then displayed, and the integral absolute error (IAE) for all three controllers is tabulated.

### 4.2.1 PIDDD Controller

### 4.2.1.a Variation of $\tau_v$

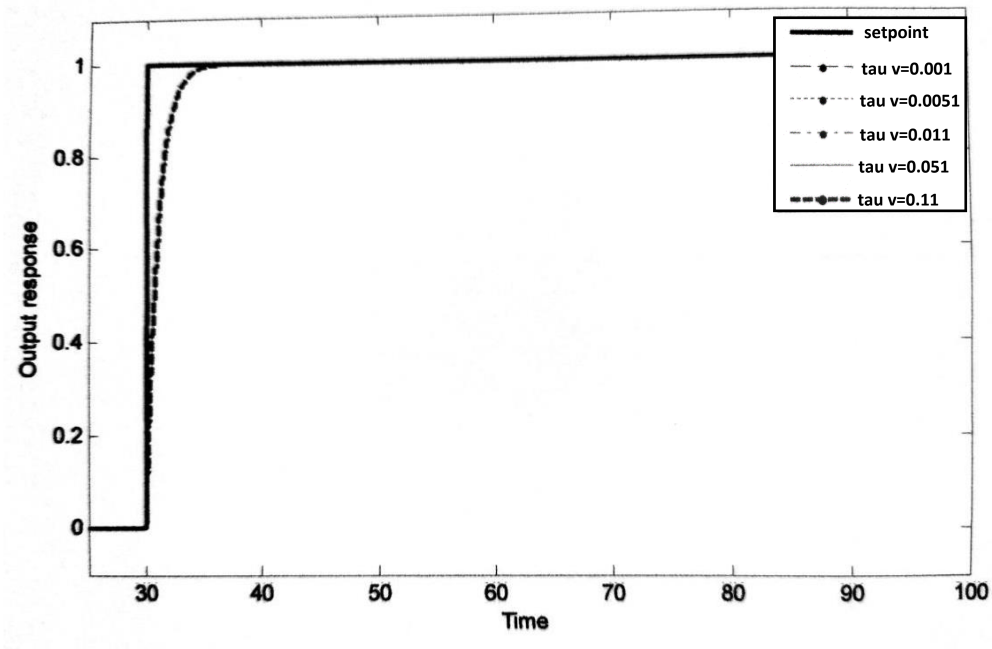


Figure 4.1 Output response of $\tau_p$=0.01s at different $\tau_v$

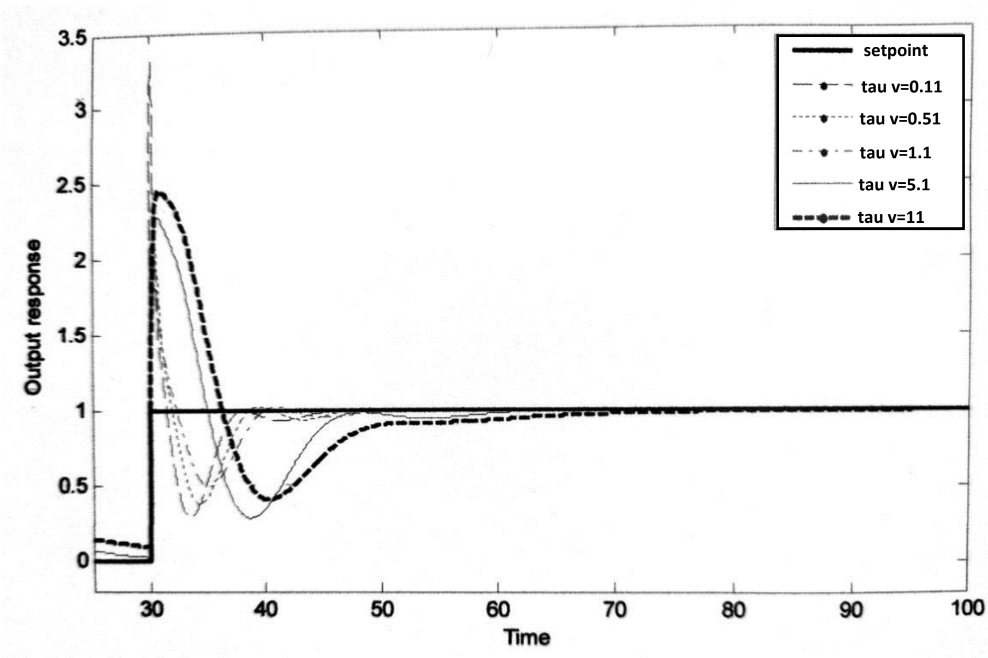


Figure 4.2 Output response of $\tau_p$=1s at different $\tau_v$

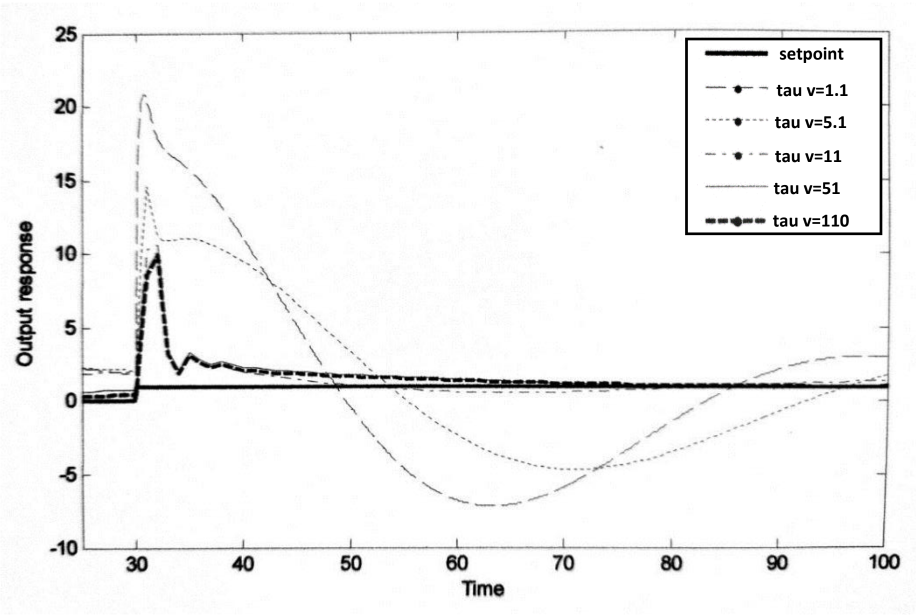


Figure 4.3 Output response of $\tau_p$=10s at different $\tau_v$

In figure 4.1, the output response of the rapid system $\tau_p$=0.01s exhibits the overlapping of the lines. Changes in $\tau_v$, values have a negligible impact on the output response variation for all $\tau_p$.

In contrast, for $\tau_p$=1s in figure 4.2, the output response exceeds the set point value for $\tau_v$, equal to, $\tau_v$/10 by more than 300 percent. As $\tau_v$ increases, the overshoot decreases to 200 percent; if $\tau_v$ equals or exceeds $\tau_p$, the overshoot is 250 percent. There is a brief period of oscillation before the oscillation settles to its ultimate value. The rising time for each $\tau_v$, value is same, however the settling time varies for each $\tau_v$ value. In terms of time, the $\tau_v$ value with the shortest settling time is less than 10 seconds, while the $\tau_v$ value with the longest settling time is around 30 seconds.

As seen in Figure 4.3, the output response becomes unstable when $\tau_p$=10s. For $\tau_v$ higher than $\tau_p$, the output response is overshoot and the line is not smooth, but the settling time is reduced. The output response overshoots by more than 2000 percent for the tiny $\tau_v$=$\tau_p$/10 and decreases to 1500 percent for the $\tau_v$=5s. For $\tau_v$ equal to or larger than $\tau_p$, the overshoot is around 1000 percent. The output response of smaller televisions oscillates rather than settling at the setting specified by the user. However, the remaining three responses are established between t=50s and t=60s.

The controller output response is investigated based on the output response. Due to the inadequate output responses, the controller output response for slow processes $\tau_p$=10s is omitted.

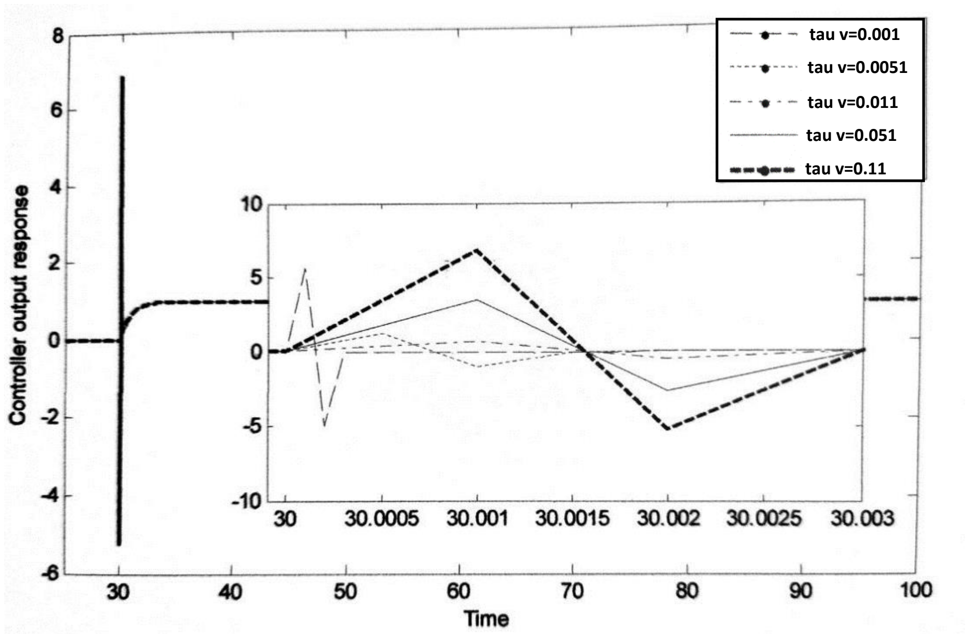


Figure 4.4 Controller output response of $\tau_p$=0.01s at different $\tau_v$

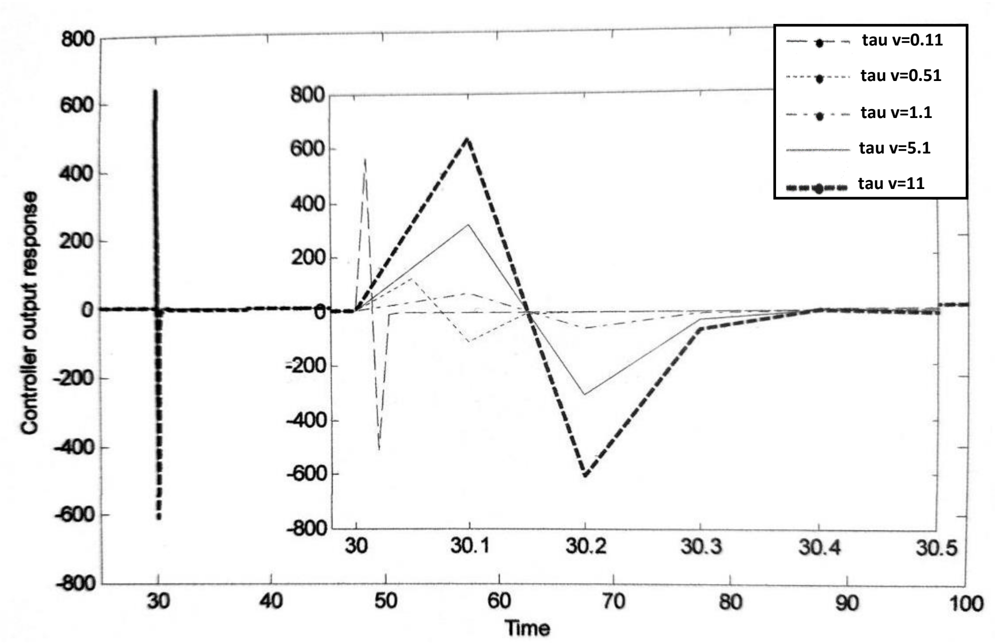


Figure 4.5 Controller output response of $\tau_p$=1s at different $\tau_v$

The controller output response may represent the signal sent from the controller to the ultimate control device, a valve or actuator. The greater the peak, the greater the controller's output signal, which means the controller must use more effort to adjust the process output so that it reaches the set point. The largest output response peak in figures 4.4 and 4.5 corresponds to the maximum $\tau_v$, which is 10 times bigger than $\tau_p$.

In terms of time, the controller takes action to stabilise the $\tau_v$ value. The controller responds within 0.0005s when tiny $\tau_v$=0.001s for $\tau_v$=0.1s, the controller requires around 0.003s. The controller takes about 10 times longer for $\tau_p$=1s than with $\tau_p$=0.01s.