

**AN INVESTIGATION ON IMC BASED DUAL-PHASE PID
CONTROLLERS**

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ABBREVIATIONS

CC	: Cohen Coon
DCS	: Distributed Control System
DM	: Dual Mode
DP	: Dual Phase
FOPDT	: First Order Plus Dead Time
GM	: Gain Margin
GMC	: Generic Model Control
GPIB	: General Purpose Instrumentation Bus
GPIO	: General Purpose Input/Output
IAE	: Integral of the Absolute of the Error
IMC	: Internal Model Control
ISE	: Integral of the Squared Error
ITAE	: Integral of the Time Weighed Absolute Error
MPC	: Model Predictive Control
P	: Proportional
PI	: Proportional Integral
PID	: Proportional Integral Derivative
PM	: Phase Margin
OLTF	: Open Loop Transfer Function
SISO	: Single Input Single Output
VEE	: Visual Engineering Environment
VME	: Virtual Machine Environment
VXI	: VME Extensions for Instrumentations
ZN	: Ziegler Nichols

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NOMENCLATURE

AR	: Amplitude Ratio
C	: Controlled Variable
G	: Process
$\check{G}$	: Process model
K_p	: Process gain
K_c	: Gain, proportional tuning parameter
M	: Manipulated variable controller
q	: IMC controller
R	: Set point
T	: Temperature
t	: Time
τ_D	: Derivative time constant
τ_F, λ	: Filter time constants
τ_I	: Integral (reset) time constant
τ_P	: Process time constant
θ	: Delay time

AN INVESTIGATION ON IMC BASED DUAL-PHASE PID CONTROLLER

SUMMARY

The chemical industry is known to have very dynamic and changing market conditions with a wide range of conditions. Since it is not feasible to re-design the processes for economical concerns, effective control schemes are essential.

In the scope of that project, having enough knowledge on Dual-Mode Controllers by literature, a Dual-Phase Controller is designed that adapts two different controller parameters for two different phases in the re-boiler of distillation columns having a wide range of usage in separation processes of chemical engineering. The designed controller system is simulated via a computer and the results are examined according to different well known performance criteria.

The temperature control of liquid and vapor phases in the reboiler of distillation columns, having commonly used in most of the separation processes of chemical engineering, has a vital role since the temperature of the head of the distillation column must be kept at a certain critical level.

In this study, the modeling and control of the reacting distillation process are performed using the data obtained from a previous study of the reacting distillation process producing acetate esters from the fusel alcohols. The control of the process is very difficult due to its fusel oil content. The major components of fusel oil (i-amyl, i-butyl, n-propyl, ethyl alcohol and water) indicate an azeotropic property (closer boiling points) that results in difficulty in the control of a distillation column. Having complex characteristics, the control of a reboiler of the fusel oil distillation is designed for both liquid and vapor phases. It is observed that, the system response obtained by using three-term controllers based on internal model control (DPIMCPID_L and DPIMCPID_V) gives better results. Those controllers use the model transfer function of the liquid phase when heating-up in progress, and the model transfer function of the vapor phase when cooling-down the system.

Model transfer functions are obtained by using the process reaction curve. It is achieved that by addition of the process integrator to the first order plus time delay transfer function obtained by process reaction curve, the system response becomes faster compared to the previous studies results.

An IMC based PID controller is designed by using model transfer functions. A pole placement method used during the internal model control design removed the oscillation effects of the system responses. Thus, IMC based PID controllers ($DPIMCPID_L$ and $DPIMCPID_V$) are designed for both phases. Then, simulation studies are performed.

Process simulation is done by using Matlab 6.5 and VEE Pro 7.0. In simulation studies, beside the $DPIMCPID$ controller, the classical controller Ziegler Nichols tuning method is also used. After obtaining simulation results, it is obviously seen that using the model transfer functions of the liquid and vapor phases is beneficial to reach the desired set point values.

All the simulation results of the controllers used in this thesis are analyzed according to several performance criteria by using MathCad and Matlab 6.5. It is seen that, by examining the numerical values of the controller performance indices, $DPIMCPID$ controller gives better results compared to the classical controllers. Thus, more precised temperature control of the reacting distillation process is aimed in this study by using the $DPIMCPID$ controller.

DAHİLİ MODEL KONTROL BAZLI İKİ FAZLI PID KONTROL EDİCİLER ÜZERİNE BİR ARAŞTIRMA

ÖZET

Kimya endüstrisinin geniş bir yelpazede, son derece dinamik ve değişken pazar koşullarına sahip olduğu bilinmektedir. Endüstri açısından proseslerin tekrar dizaynı ekonomik olarak uygun olmadığından, etkili kontrol mekanizmaları bu pazarlarda öncelik kazanmaktadır.

Bu proje kapsamında, İkili Modda Çalışan (Dual Mode) Kontrol Ediciler hakkında elimizde bulunan mevcut literatür bilgilerine dayanarak, kimya mühendisliği ayırma işlemlerinde oldukça yaygın olarak kullanılan distilasyon kolonlarının reboyer bölümündeki sıvı ve buhar fazlarının ısınma ve soğuma aşamalarındaki ısı transferinin farklı durumları için dönüşümlü olarak çalışacak iki fazlı (dual-phase) bir kontrol edici tasarlanmıştır. Oluşturulan sistemin bilgisayar yardımı ile simülasyonu yapılmış ve tasarlanan bu kontrol edici çeşitli performans kriterlerine göre incelenmiştir.

Kimya mühendisliği uygulaması olarak çok sık rastlanan ayırma işlemlerinde kullanılan distilasyon kolonlarının reboyer sıvı ve buhar faz sıcaklıklarının kontrolleri, kolonunun tepe noktasındaki sıcaklığının belirli bir kritik değerde tutulması gerektiğinden büyük önem arz etmektedir.

Bu çalışmada, daha önce yapılmış olan fuzel alkollerinden asetat esterleri üreten reaksiyonlu distilasyon prosesinin verileri kullanılarak sistem modellenmesi ve kontrolü gerçekleştirilmiştir. Proses, fuzel yağı içerdiğinden kontrolü oldukça zordur. Fuzel yağının bileşimindeki ana komponentler (i-amil, i-bütil, n-propil, etil alkol ve su) azeotropik özellik (kaynama noktalarının yakınlığı) gösterdiğinden, distilasyon kolonunun kontrolünü zorlaştırmaktadır. Karmaşık özelliğe sahip olan fuzel yağının distilasyonunun reboyer kontrolü hem sıvı, hem de buhar fazı için tasarlanmıştır. Dahili model kontrole dayanan üç terimli kontrol edicilerin ($DPIMCPID_L$ ve $DPIMCPID_V$) kullanılması ile elde edilen sistem cevaplarının çok daha hızlı olduğu görülmüştür. Söz konusu kontrol ediciler sistem ısınırken sıvı fazın model denklemini, soğurken de buhar fazın denklemini kullanmaktadır.

Model denklemleri proses reaksiyon eğrisinden faydalanarak oluşturulmuştur. Proses reaksiyon eğrisinden elde edilen zaman gecikmeli birinci mertebeden transfer denklemine proses integratörü eklenmesiyle, sistem simülasyon cevaplarının önceki çalışmalara göre daha hızlı olması sağlanmıştır.

Elde edilen proses model denklemleriyle dahili model kontrole dayanan PID kontrol edici tasarlanmıştır. Dahili model kontrolü yaparken kutup yerleştirme yöntemi

kullanılması sistem cevaplarının osilasyon etkisini kaldırmıştır. Sonuçta, her iki faz için de PID kontrol edicileri (DPIMCPID_L ve DPIMCPID_V) elde edilmiştir. Daha sonra simülasyon işlemlerine geçilmiştir.

Prosesin simülasyonu Matlab 6.5 ve VEE Pro 7.0 yardımıyla yapılmıştır. Simülasyon çalışmalarında, DPIMCPID kontrol ediciye ek olarak klasik kontrolör ayar yöntemi olan Ziegler Nichols'a da yer verilmiştir. Simülasyonlar neticesinde, sistem ısınırken sıvı faz denkleminin, soğurken de buhar faz denkleminin kullanımasının avantajlı olduğu görülmüştür.

Tez içerisinde elde edilen bütün kontrol edicilerin simülasyon sonuçları, çeşitli performans kriterlerine göre MathCad ve Matlab 6.5 yardımıyla incelenmiştir. Kontrol edicilerin sayısal performans değerleri incelenerek, DPIMCPID kontrol edicisinin diğer klasik kontrol edicilere kıyasla daha iyi sonuçlar verdiği görülmüştür. Böylece, DPIMCPID kontrol edicisinin kullanılmasıyla reaksiyonlu distilasyon prosesinin daha kesin sıcaklık kontrolü sağlanması hedeflenmiştir.

1. INTRODUCTION

1.1 History of Process Control

Most chemical processing plants were run essentially manually prior to the 1940s. Only the most elementary types of controllers were used. Many operators were needed to keep watch on the many variables in the plant. Large tanks were employed to act as buffers or surge capacities between various units in the plant. These tanks, although sometimes quite expensive, served the function of filtering out some of the dynamic disturbances by isolating one part of the process from upsets occurring in another part [1].

With increasing labor and equipment costs and with the development of more severe, higher-capacity, higher performance equipment and processes in the 1940s and early 1950s, it became uneconomical and often impossible to run plants without automatic control devices. At this stage feedback controllers were added to the plants with little real consideration for the dynamics of the process itself. Rule-of-thumb guides and experience were the only design techniques [1].

In the 1960s chemical engineers began to apply dynamic analysis and control theory to chemical engineering processes. Most of the techniques were adapted from the work in the aerospace and electrical engineering fields. In addition to designing better control systems, processes and plants were developed or modified so that they were easier to control [1].

The rapid rise in energy prices in the 1970s provided additional needs for effective control systems. The design and redesign of many plants to reduce energy consumption resulted in more complex, integrated plants that were much more interacting. So the challenges to the process control engineer have continued to grow over the years. This makes the study of dynamics and control even more vital in the chemical engineering curriculum than it was 30 years ago [1].

1.2 Objective of the Study

In the scope of this study, a Dual-Phase PID Controller is designed, that adapts two different controller parameters for two different phases in a reboiler of the reacting distillation column having a wide range of usage in separation processes of chemical engineering. The designed controller system is simulated via a computer and the results are examined according to the different well known performance criteria and compared with other classical type of controllers.

1.3 Structure of the Thesis

The first part of this thesis from Chapters 2 to 3 concentrates on the basic elements of the controllers and the literature review. The second part deals with the controller algorithm derived for the Reacting Distillation Process, and its simulation results.

The structure and the layout of this thesis are as follows:

Chapter 2 provides a brief information of the advances in dual mode controllers and Internal Model Control studies.

Chapter 3 describes the controller mechanisms used in this study and the process identification techniques.

In Chapter 4, the modeling and control of the Reacting Distillation Process are explained and DPIMCPID controller parameters are obtained for liquid and vapor phases.

Chapter 5 shows the simulation studies and results obtained by Matlab 6.5 and VEE Pro 7.0.

Chapter 6 indicates the comparison of DPIMCPID results by using controller performance criteria.

Chapter 7 summarises the results of the study and provides further suggestions for the direction of future work.

2. LITERATURE SURVEY

Aziz, Hussain and Mujtaba examined the performance of three different types of controllers in tracking the optimal reactor temperature profiles in batch reactor. Dual Mode (DM) control with proportional-integral (PI) and proportional-integral-derivative (PID) and generic model control (GMC) algorithms are used to design the controllers to track the optimal temperature profiles (dynamic set points). Neural network technique is used as the on-line estimator the amount of heat released by the chemical reaction within the GMC algorithm [2].

Allgöwer and Ogunnaike studied the dual-mode adaptive control of nonlinear processes. In the first mode the controller is designed to achieve stability and performance in a neighborhood of the operating point where the system is operated under normal conditions. The second mode is based on a very robust adaptive nonlinear high-gain feedback law. They proved that global asymptotic stability was achieved robustly with that scheme [3].

Rubio and Aracil examined the performance of a dual-mode controller followed by deadbeat regulator. The goal of the system is to follow a given target in as short a time as possible, and with a minimum overshoot. The study shows the results of controller performance both in theoretical and in real processes [4].

Tan, Marquez and Chen introduced a modified IMC structure proposed for unstable processes with time delays. The structure extends the standard IMC structure for stable processes to unstable processes and controllers do not have to be converted to conventional ones for implementation. An advantage of the structure is that setpoint tracking and disturbance rejection can be designed separately [5].

Lee, Lee and Park proposed a new method for PID controller tuning based on process models for integrating and unstable processes with time delay. The tuning rule is based on the process model and the desired closed-loop response. The results show that the proposed tuning method is superior to the existing methods [6].

Skogestad studied Simple analytic rules for model reduction and PID controller tuning. The starting point has been the IMC-PID tuning rules that have achieved widespread industrial acceptance. The rule for the integral term has been modified to improve disturbance rejection for integrating processes. Furthermore, rather than deriving separate rules for each transfer function model, there is just a single tuning rule for a first-order or second-order time delay model [7].

Wang, Hang and Yang examined a new internal model control (IMC)-based single-loop controller design. The model reduction technique is employed to find the best single-loop controller approximation to the IMC controller. Compared with the existing IMC-based methods, the proposed design is applicable to a wider range of processes, and yields a control system closer to the IMC counterpart [8].

Kaya studied on relay autotuning of a plant to find parameters for its control using a Smith predictor. A Smith predictor configuration is represented as its equivalent internal model controller (IMC) which provides the parameters of the proportional-integral (PI) or proportional-integral-derivative (PID) controller to be defined in terms of the desired closed-loop time constant, which can be adjusted by the operator, and the parameters of the process model [9].

3. THEORY OF CONTROLLERS AND PROCESS IDENTIFICATION

3.1. Conventional PID Controller

3.1.1 History

Despite rapid evolution in control hardware over past 50 years, the PID controller remains the workhorse in process industries. The proportional action (P mode) adjusts controller output according to the size of the error. The integral action (I mode) can eliminate the steady-state offset and the future trend is anticipated via the derivative action (D mode) [10]. These usefull functions are sufficient for a large number of process applications and the transparency of the features leads to wide acceptance by the users. On the other hand, it can be shown that the internal model control (IMC) framework leads to PID controllers for virtually all models common in industrial practice [11]. Note that this includes systems with inverse responses and integrating (unstable) processes.

PID controllers have survived many changes in technology. It begins with pneumatic control, through directional control to the distributed control system (DCS). Nowadays the PID controller is far different from that of 50 years ago. Typically, logic, function block, selector and sequence are combined with the PID controller. Many sophisticated regulatory control strategies, override control, start-up and shut-down strategies can be designed around the classical PID control. This provides the basic means for good regulatory, smooth transient, safe operation and fast start-up and shut-down. Moreover, even with the Model Predictive Control (MPC), the PID controllers still served as the fundamental building block at the regulatory level. The computing power of microprocessors provides additional features such as automatic tuning, gain scheduling and model switching to the PID controller. Eventually, all PID controllers will have the above mentioned intelligent features [10].

In process industries, more than 90% of the control loops are of the PID type [12]. Most loops are actually under PI control (as a result of the large number of flow loops). Fifty years after the publication of the Ziegler-Nichols tuning rule (1942) and with the numerous papers published on the tuning methods since, one might think the use of PID controllers has already met our expectations. Unfortunately, this is not the case. Surveys of Bialkowski (1993), Ender (1993), McMillan (1995) and Hersh and Johnson (1997) show that :

1. Pulp and paper industry over 2000 loops [13];

- 30% gave poor performance due to poor controller tuning
- 30% gave poor performance due to control valve problems (e.g., control valve stiction-slip, dead band, backlash)
- 20% gave poor performance due to process and control system design problems

2. Process industries [14];

- 30% of loops operated on manual mode
- 20% of controllers used factory tuning
- 30% gave poor performance due to sensor and control valve problems

3. Chemical process industry [15];

- Half of the control valves needed to be fixed (results of the Fisher diagnostic valve package).

4. Manufacturing and process industries [16];

- Engineers and managers cited PID controller tuning as difficult problem.

Surveys indicate that the process control performance is, indeed, "not as good as you think" [14]. The reality leads us to reconsider the priorities in process control research. First, an improved process and control configuration redesign (e.g., selection and pairing of input and output variables) can improve control performance. This research direction has received a great deal of attention in recent years under the title "interaction between design and control"[10].

Second, control valves contribute significantly to the poor control performance. It is difficult, if not impossible, to replace or to restore all the control valves to the expected performance. In other words, in many cases, this is a fact we have to face (e.g., dead band, stic-slip etc.). Third and probably the easiest way to improve control performance is to find appropriate tuning constants for PID controllers [10].

Fifty years after Ziegler and Nichols published their famous tuning rule numerous tuning methods were proposed in the literature. It is expected that engineers have gained proficiency in the design of simple PID controller. The reality indicates that this is simply not the case. Moreover, the structure of current leaner corporations does not offer much opportunity to improve the situation. Another factor is the time required for the tuning of many slow loops (e.g., temperature loops in high purity distillation columns). In many occasions, engineers simply do not have the luxury and patience to tune a loop over a long period of time (not being able to complete the task in a shift). It then becomes obvious that the PID controller with an automatic tuning feature is an attractive alternative for better control. That is, instead of continuous adaptation, the controller should be able to find the tuning parameters by itself: it is an autotuner [10].

3.1.2 PID Structure

Consider the simple SISO control loop shown in Figure 3.1.

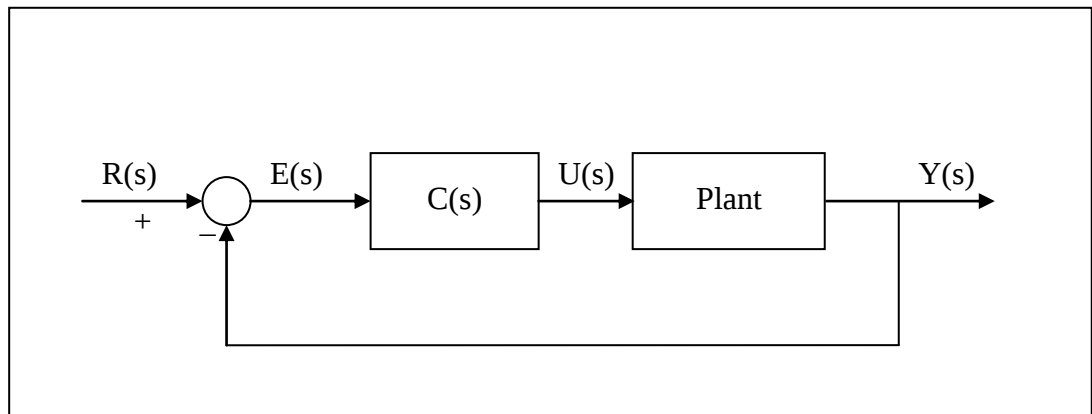


Figure 3.1 : Basic feedback control loop

The traditional expressions for PI and PID controllers can be described by their transfer functions, relating error $E(s) = R(s) - Y(s)$ and controller output $U(s)$ as follows:

$$C_P(s) = K_C \quad (3.1)$$

$$C_{PI}(s) = K_C \left(1 + \frac{1}{\tau_I s} \right) \quad (3.2)$$

$$C_{PD}(s) = K_C \left(1 + \frac{\tau_D s}{T_d s + 1} \right) \quad (3.3)$$

$$C_{PID}(s) = K_C \left(1 + \frac{1}{\tau_I s} + \frac{\tau_D s}{T_d s + 1} \right) \quad (3.4)$$

$$C_{PID}(s) = K_C \left(1 + \frac{1}{\tau_I s} + \tau_D s \right) \quad (3.5)$$

where τ_I and τ_D are known as the reset time (integral time) and derivative time, respectively [17].

The form given in (3.5) is the ideal PID used in most of the processes. As seen from (3.1) to (3.4), the members of this family include, in different combinations, three control modes or actions: proportional (P), integral (I), and derivative (D) [17].

Proportional action provides a contribution which depends on the instantaneous value of the control error. A proportional controller can control any stable plant, but it provides limited performance and nonzero steady-state errors. This latter limitation is due to the fact that its frequency response is bounded for all frequencies [17].

It has also been traditional to use the expression proportional band (PB) to describe the proportional action. The equivalence is

$$PB[\%] = \frac{100[\%]}{K_C} \quad (3.6)$$

The proportional band is defined as the error required (as a percentage of full scale) to yield a 100% change in the controller output [17].

Integral action, on the other hand, gives a controller output that is proportional to the accumulated error, which implies that it is a slow reaction control mode. This characteristic is also evident in its low-pass frequency response. The integral mode plays a fundamental role in achieving perfect plant inversion at $\omega = 0$. This forces the steady-state error to zero in the presence of a step reference and disturbance. The integral mode has two major shortcomings: its pole at the origin is detrimental to loop stability and it also gives rise to the undesirable effect (in the presence of actuator saturation) known as wind-up [17].

Derivative action acts on the rate of change of the control error. Consequently, it is a fast mode which ultimately disappears in the presence of constant errors. It is sometimes referred to as a predictive mode, because of its dependence on the error trend. The main limitation of the derivative mode, is its tendency to yield large control signals in response to high-frequency control errors, such as errors included by set-point changes or measurement noise. Its implementation requires properness of the transfer functions, so a pole is typically added to the derivative as is evident in equations (3.3) and (3.4). In the absence of other constraints, the additional time constant T_d is normally chosen such that $0.1\tau_D \leq T_d \leq 0.2$. This constant is called the derivative time constant; the smaller it is, the larger the frequency range over which the filtered derivative approximates the exact derivative [17].

3.1.3 Methods for Tuning PID Controllers

There are several methods for tuning a controller. In this study a PID controller is modeled by the function:

$$p(t) = \bar{p} + K_c \left[e(t) + \frac{1}{\tau_I} \int_0^t e(t) dt + \tau_D \frac{de(t)}{dt} \right] \quad (3.7)$$

which has the transfer function,

$$G(s) = \frac{K e^{-\theta s}}{\tau s + 1} \quad (3.8)$$

Trial and error

Assume the controller algorithm is as follows

1. Eliminate the integral and derivative action by setting τ_D to 0 and τ_I to as large a value as possible
2. Set K_c at a low value and put the controller on automatic
3. Increase the controller gain K_c by small increments until continuous cycling occurs after a small set point or load change. The term “continuous cycling” refers to a sustained oscillation with constant amplitude.
4. Reduce K_c by a factor of 2.
5. Decrease τ_I in small increments (this increases integral control) until continuous cycling occurs again. Set τ_I to 3 times this value
6. Increase τ_D until continuous cycling occurs. Set τ_D equal to one third of this value [18].

Disadvantages

1. It is quite time consuming if a large number of trial are required or if the process dynamics are slow. Testing can be expensive because of lost productivity or poor product quality
2. Continuous cycling may be objectionable because the process is pushed to the stability limit. Consequently, if external disturbances or a change in the process occurs during controller tuning, unstable operation or a hazardous situation could result.
3. The tuning process is not applicable to processes that are open loop unstable because such processes typically are unstable at high and low values of K_c but are stable at an intermediate range of values.
4. Some simple processes do not have an ultimate gain (e.g. first and second order OLTF (open loop transfer function) without time delays) [18].

Continuous cycling method (Ziegler Nichols Method)

1. Determine the ultimate controller gain K_{cu} as described in the first three steps, above (or from the root locus plot of the OLTf)
2. The period of oscillation at this value is the ultimate period, P_u . The Ziegler Nichols settings are calculated from these values to provide a $1/4$ decay ratio shown in Table 3.1 [18].

Table 3.1 : Ziegler Nichols Controller Settings Based on the Continuous Cycling Method

Controller	K_c	τ_I	τ_D
P	$0.5 K_{cu}$	--	--
PI	$0.45 K_{cu}$	$P_u / 1.2$	--
PID	$0.6 K_{cu}$	$P_u / 2$	$P_u / 8$

Process Reaction Curve Methods (Cohen-Coon method)

Assume the controller algorithm is as follows

$$p(t) = \bar{p} + K_c \left[e(t) + \frac{1}{\tau_I} \int_0^t e(t) dt + \tau_D \frac{de(t)}{dt} \right] \quad (3.9)$$

For self-regulating responses (Cohen-Coon method), also assume that the process model can be described as a first order lag with dead time. In other words, the OLTf can be approximated by the following transfer function [18].

$$G(s) = \frac{K e^{-\theta s}}{\tau s + 1} \quad (3.10)$$

With the controller in the manual mode, the controller output is suddenly changed. The process output curve follows one of the curves shown in Figure 3.2 below and the times are noted. In the second curve, the straight line to determine τ and the slope S is drawn tangent to the system response at the inflection point. In the bottom curve, the slope S is taken from a straight portion of the ascent [19].

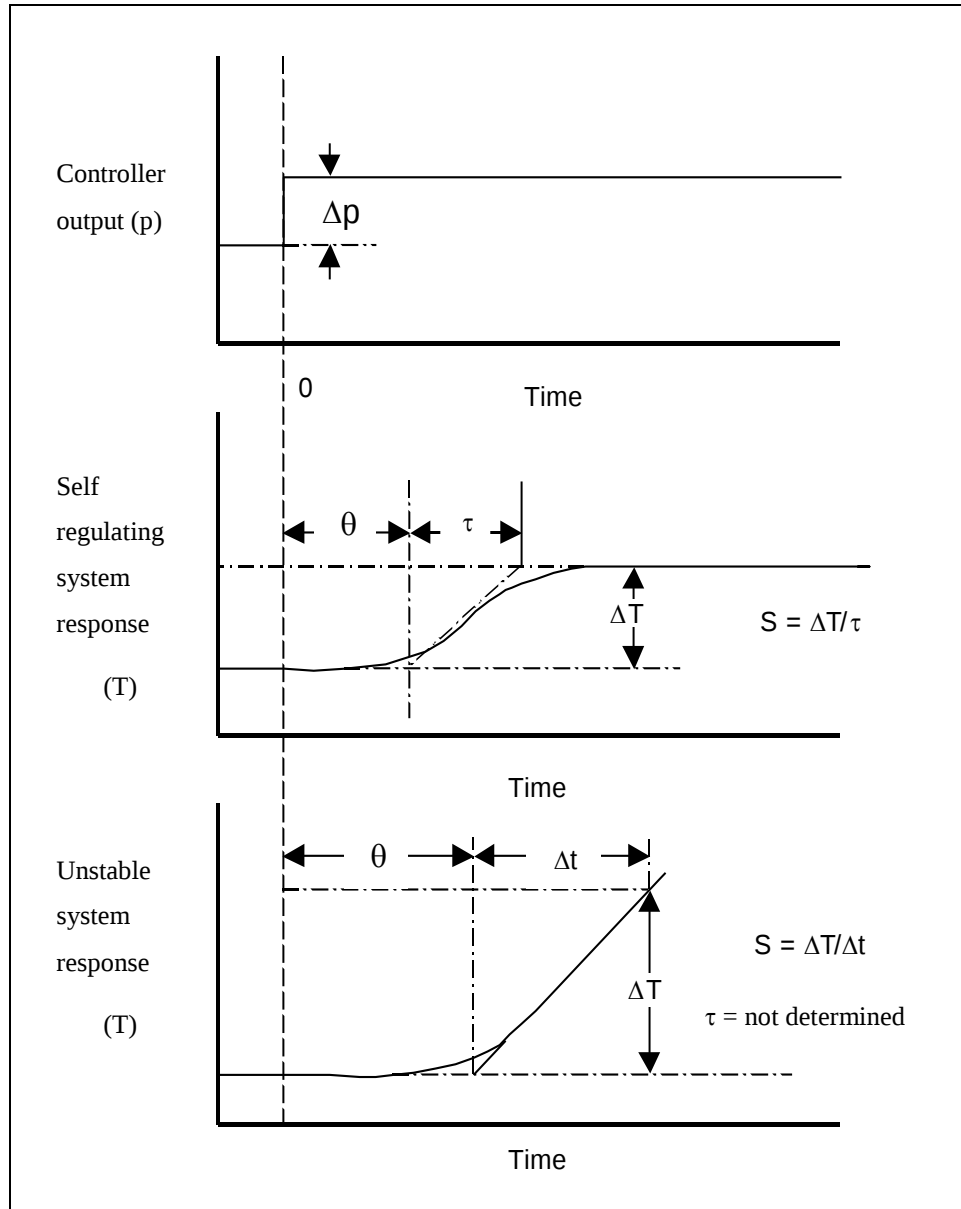


Figure 3.2 : Process output curve (S-curve)

1. Ziegler-Nichols method (both types of responses)
 Compute the following:
 $S = \Delta T/\tau$ or $S = \Delta T/\Delta t$ (the average slope of the rise)
 $S^* = S/\Delta p$ (the normalized slope)
2. Compute the controller settings from Table 3.2.

Table 3.2 : Ziegler-Nichols Tuning Relations (Process Reaction Curve Method)

Controller Type	K_c	τ_I	τ_D
P	$\frac{1}{\theta S^*}$	--	--
PI	$\frac{0.9}{\theta S^*}$	3.33 θ	--
PID	$\frac{1.2}{\theta S^*}$	2 θ	0.5 θ

Cohen-Coon Method (self regulating response only)

Alternatively, the constants K , τ and θ can be used in the process model when the system exhibits self-regulating response (i.e. the response curve follows the second curve so τ can be determined). Controller settings can be calculated using the Cohen-Coon relations (designed for $\frac{1}{4}$ decay ratio) indicated in Table 3.3 [18].

Table 3.3 : Cohen Coon Controller Design Relations

Controller	Settings	Cohen-Coon
P	K_c	$\frac{1}{K} \frac{\tau}{\theta} \left[1 + \frac{\theta}{3\tau} \right]$
PI	K_c	$\frac{1}{K} \frac{\tau}{\theta} \left[0.9 + \frac{\theta}{12\tau} \right]$
	τ_I	$\frac{\theta[30 + 3(\theta/\tau)]}{9 + 20(\theta/\tau)}$
PID	K_c	$\frac{1}{K} \frac{\tau}{\theta} \left[\frac{16\tau + 30}{12\tau} \right]$
	τ_I	$\frac{\theta[32 + 6(\theta/\tau)]}{13 + 8(\theta/\tau)}$
	τ_D	$\frac{4\theta}{11 + 2(\theta/\tau)}$

Advantages

1. Only a single experiment is necessary
2. Does not require operating at stability limit
3. Does not require trial and error
4. Controller settings are easily calculated [19].

Disadvantages

1. The experiment is performed under open loop conditions (controller on manual). Thus, if a significant load change occurs during the test, no corrective action is taken and the test results may be easily distorted.

2. It may be difficult to determine the slope at the inflection point accurately, especially if the measurement is noisy and a small recorder chart is used
3. The methods tends to be sensitive to controller calibration errors.
4. The Cohen and Coon recommended settings tend to result in oscillatory responses since they were developed to provide a $\frac{1}{4}$ decay ratio
5. The method is not recommended for processes that have oscillatory open-loop responses since the process model will be quite inaccurate [19].

Additional Comments on the Process Model

The process model used for this tuning method (first order lag plus dead time) is actually much applicable than might seem on first consideration. The explanation for this fact lies in the Padé approximation for the Laplace transform of dead time:

$$e^{-\theta s} \cong \frac{1 - \frac{\theta}{2}s}{1 + \frac{\theta}{2}s} \quad (1^{\text{st}} \text{ order Padé}) \quad (3.11)$$

Substituting this approximation into the OLTf yields

$$G(s) \cong \frac{1 - \frac{\theta}{2}s}{\left(1 + \frac{\theta}{2}s\right)(\tau s + 1)} \quad (3.12)$$

From this approximation, we can see that the system model approximates a system of two first order lags and a first order lead. Many systems can be fit to this approximate model [19].

3.2 Process Identification

In practice many of the industrial processes to be controlled are too complex to be described by the application of fundamental principles. Either the task requires too much time and effort or the fundamentals of the process are not understood. By means of experimental tests, one can identify the dynamic nature of such processes and from the results obtain a process model, which is at least satisfactory for use in designing control systems. The experimental determination of the dynamic behaviour of a process is called process identification.

Process models are needed in developing feed-forward control algorithms, self-tuning algorithms, internal model control algorithms and in the use of tuning methods. Process identification provides several forms that are useful in process control; some of these forms are [20]:

- Process reaction curve (obtained by step input)
- Frequency response diagram (obtained by sinusoidal input)
- Pulse response (obtained by pulse input)

In the case of the Z-N method, the procedure obtained one point on the openloop frequency response diagram when the ultimate gain was found. (This point corresponds to a phase angle of -180° and a process gain of $1/K_{cu}$ at the crossover frequency ω_{co}). In the case of the C-C method, the process identification took the form of the process reaction curve [20].

3.2.1 Step Testing

A step change in the input to a process produces a response, which is called the process reaction curve. For many processes in the chemical industry, the process reaction curve is an S-shaped curve as shown in Figure 3.3.

It is important that no disturbances other than the test step enter the system during the test, otherwise the transient will be corrupted by these uncontrolled disturbances and will be unsuitable for use in deriving a process model. For systems that produce an S-shaped process reaction curve, a general model that can be fitted to the transient is the following second-order with transport lag model [20].

$$G_p(s) = \frac{Y(s)}{X(s)} = \frac{K_p \cdot e^{-\ell s}}{(\tau_1 s + 1)(\tau_2 s + 1)} \quad (3.13)$$

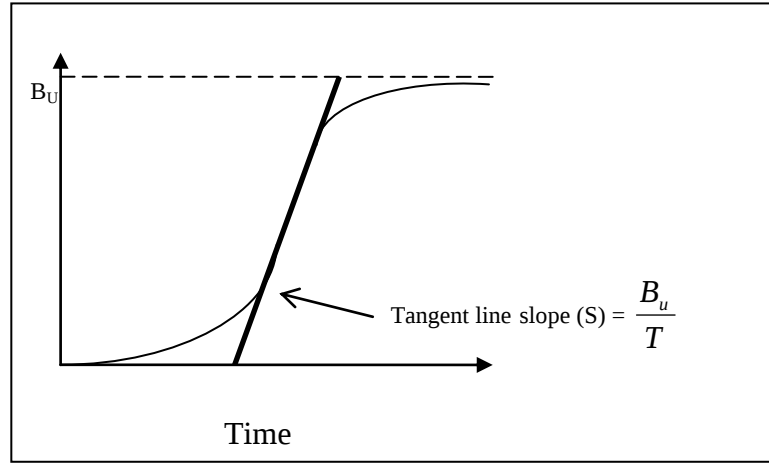


Figure 3.3 : Typical process reaction curve showing graphical construction to determine first-order with transport lag model

3.2.2 Frequency Testing

Process having a transfer function $G(s)$ can be represented by a frequency response diagram (or Bode plot) by taking the magnitude and phase angle of $(j\omega)$. This can be reversed to obtain $G(s)$ from an experimentally determined frequency response diagram. The procedure requires that a device be available to produce a sinusoidal signal over a range of frequencies. We describe such a device as a sine wave generator. In frequency testing of an industrial process, a sinusoidal variation in pressure is applied to the top of the control valve so that manipulated variable can be varied sinusoidal over a range of frequencies. The block diagram that applies during frequency testing is the same as the one in Figure 3.4 with the step input (M/s) replaced by a sinusoidal signal. The sine wave generator used to test electronic devices operates at frequencies that are too high for many slow moving chemical processes. For frequency testing of chemical processes, special low-frequency generators must be built that can produce sinusoidal variation in pressure to a control valve. To preserve the sinusoidal signal in the flow of manipulated variable through the valve, the valve must be linear [20].

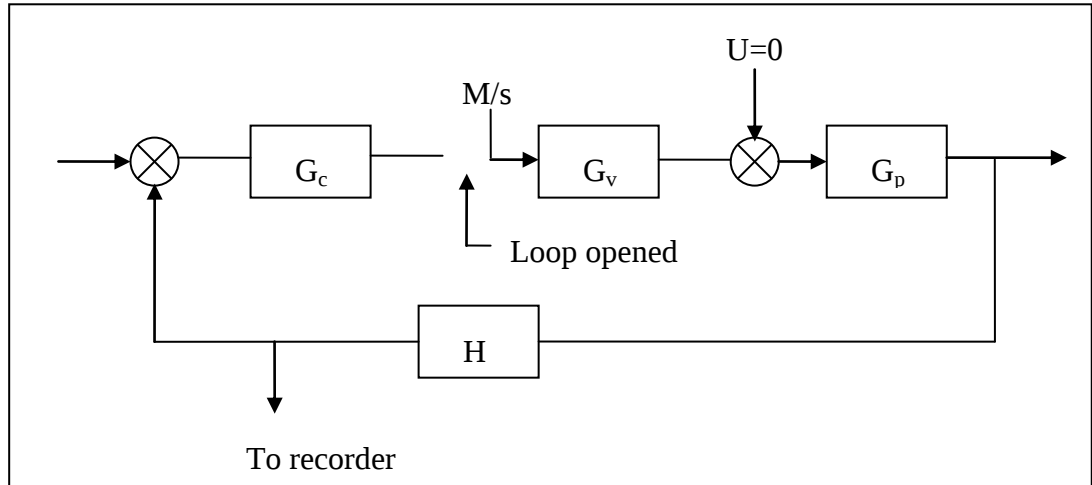


Figure 3.4 : Block diagram of a control loop for measurement of the process reaction curve

3.2.3 Pulse Testing

Pulse testing is similar to step testing; the only difference in the experimental procedure is that a pulse disturbance is used in place of a step disturbance. The pulse is introduced as a variation in valve top pressure as was done for step and frequency testing (see Figure 3.4). In applying the pulse, the open-loop system is allowed to reach steady state, after which the valve top pressure is displaced from its steady state for a short time and then returned to its original value. The response is recorded at the output of the measuring element. An arbitrary pulse and a typical response are shown in Figure 3.5. Usually the pulse shape is rectangular in experimental work, but other well-defined shapes are also used. The input-output data obtained in a pulse test are converted to a frequency response diagram, which can be used to tune a controller. The transfer function of the valve, process, and measuring element (referred to as the process transfer function, for convenience) is given by [20]:

$$G_p(s) = Y(s)/X(s) \quad (3.14)$$

where $Y(s)$ = Laplace transform of the function representing the recorded output response

$X(s)$ = Laplace transform of the function representing the pulse input

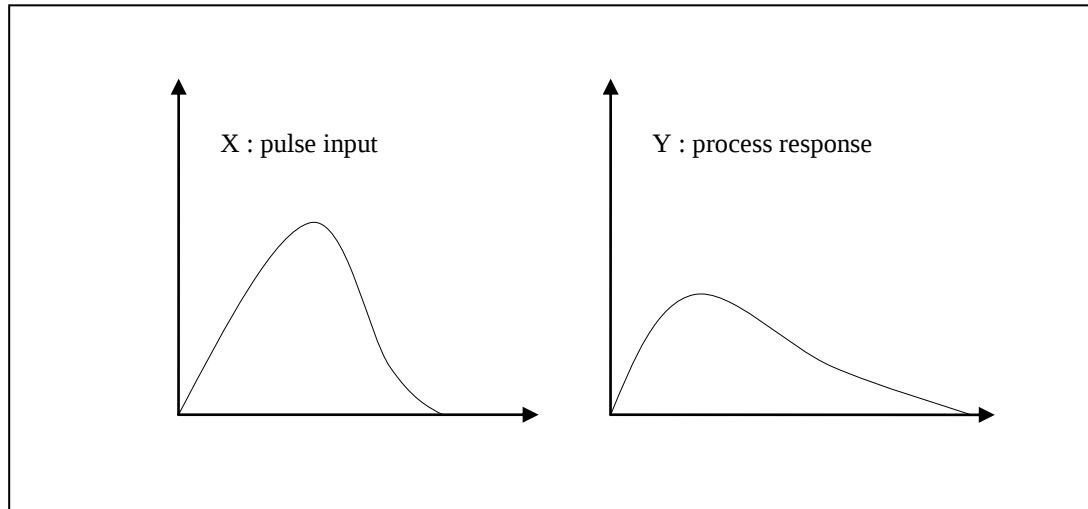


Figure 3.5 : Typical process response to a pulse input

3.2.4 The System Identification Procedure

The construction of a model from data involves three basic entities:

- 1.The data.
- 2.A set of candidate models.
- 3.A rule by which candidate models can be assessed using the data

1. The data record. The input-output data are sometimes recorded during specifically designed identification experiment, where one may determine which signals to measure and when to measure them and may also choose the input signals. The object with experiment design is thus to make these choices so that the data become maximally informative subject to constraints at hand. In other cases the user may not have the possibility to affect the experiment, but must use data from the normal operation of the system [20].

2. The set of models. A set of candidate models is obtained by specifying within which collection of models we are going to look for a suitable one. This is no doubt the most important and, at the same time, the most difficult choice of the system identification procedure. It is here that a priori knowledge and engineering intuition and insight have to be combined with formal properties of models. Sometimes the model set is obtained after careful modeling.

Then a model with some unknown physical parameters is constructed from basic physical laws and other well established relationships. In other cases standard linear models may be employed, without reference to the physical background. Such a model set whose, parameters are basically viewed as vehicles for adjusting the fit to the data and do not reflect physical considerations in the system, is called a black box. Model sets with adjustable parameters with physical interpretation may, accordingly, be called gray boxes [20].

3. Determining the "best" model in the set, guided by the data. This is the identification method. The assessment of model quality is typically based on how the models perform when they attempt to reproduce the measured data [20].

3.3 Internal Model Control (IMC)

3.3.1 The IMC Structure

The IMC structure is shown in Figure 3.6. The distinguishing characteristic of this structure is the process model, which is in parallel with the actual process (plant). Figure 3.7 illustrates that both the controller and model exist as computer computations; it is convenient to treat them separately for design and analysis. A list of transfer function variables shown in the IMC block diagram are given below [21].

$d(s)$ = disturbance

$\tilde{d}(s)$ = estimated disturbance

$g_p(s)$ = process

$\tilde{g}_p(s)$ = process model

$q(s)$ = internal model control

$r(s)$ = setpoint

$\tilde{r}(s)$ = modified setpoint (corrects for model error and disturbances)

$u(s)$ = manipulated input (controller output)

$y(s)$ = measured process output

$\tilde{y}(s)$ = model output

Notice that the feedback signal is

$$\tilde{d}(s) = (g_p(s) - \tilde{g}_p(s))u(s) + d(s) \quad (3.15)$$

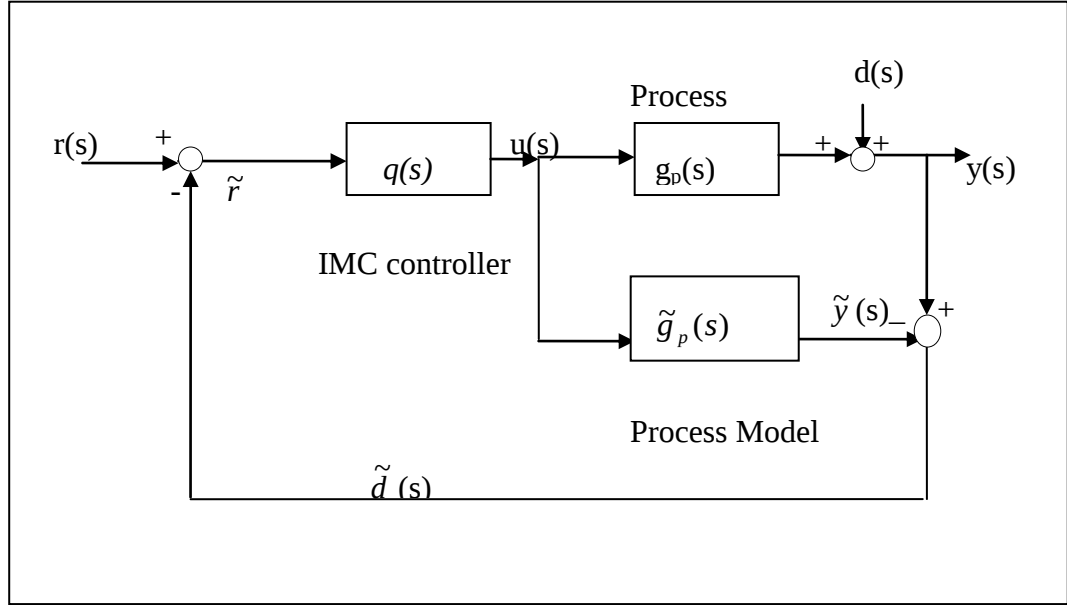


Figure 3.6 : The Internal Model Control Structure.

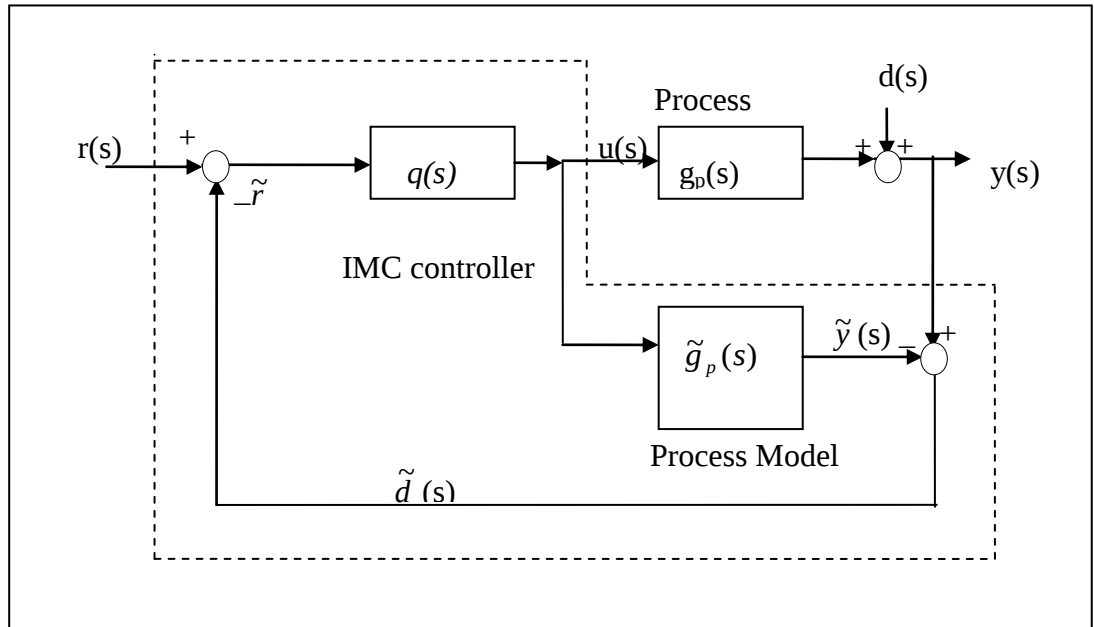


Figure 3.7 : The IMC strategy. The dotted line indicates the calculations performed by the model-based controller.

The signal to the controller is

$$\tilde{r}(s) = r(s) - \tilde{d}(s) = r(s) - (g_p(s) - \tilde{g}_p(s))u(s) - d(s) \quad (3.16)$$

Consider now same limiting cases.

Perfect Model, No Disturbances

If the model is perfect ($\tilde{g}_p(s) = g_p(s)$) and there are no disturbances ($d(s) = 0$), then the feedback signal is zero. The relationship between $r(s)$ and $y(s)$ is then

$$y(s) = g_p(s)q(s)r(s) \quad (3.17)$$

Notice that this is the same relationship that is obtained for an openloop control system design.

Recall that a standard feedback controller could actually destabilize if the tuning parameters are not correctly chosen. An analysis of the poles of the closed-loop transfer function must be performed to determine the stability of standard feedback controllers [21].

Perfect Model, Disturbance Effect

If the model is perfect ($\tilde{g}_p(s) = g_p(s)$) and there is a disturbance, then the feedback signal is

$$\tilde{d}(s) = d(s) \quad (3.18)$$

This illustrates that feedback is needed because of unmeasured disturbances entering a process [21].

Model Uncertainty, No Disturbances

If there are no disturbances [$d(s) = 0$] but there is model uncertainty ($\tilde{g}_p(s) \neq g_p(s)$) which is always the case in the real world, then the feedback signal is;

$$\tilde{d}(s) = (g_p(s) - \tilde{g}_p(s))u(s) \quad (3.19)$$

This illustrates that feedback is needed because of model uncertainty.

The closed-loop relationship is

$$y(s) = \frac{g_p(s)q(s)}{1 + q(s)(g_p(s) - \tilde{g}_p(s))} r(s) + \left[\frac{1 - \tilde{g}_p(s)q(s)}{1 + q(s)(g_p(s) - \tilde{g}_p(s))} \right] d(s) \quad (3.20)$$

Recapitulating, the reasons for feedback control include the following:

- Unmeasured disturbances
- Model uncertainty
- Faster response than the open-loop system (with a static controller)
- Closed-loop stability of open-loop unstable system

The primary disadvantage of IMC is that it does not guarantee stability of open-loop unstable systems [21].

3.2.2 The IMC Design Procedure

The IMC design procedure for SISO systems is identical to the design procedure that we developed for open-loop controller design earlier. The assumption we are making is that the model is perfect, so the relationship between the output, y , and the setpoint, r , is given by Equation (3.17). Model uncertainty is handled by adjusting the "filter factor" for robustness (tolerance of model uncertainty) and speed of response. The IMC design procedure consists of the following four steps [21].

1. Factor the process model into invertible and noninvertible (bad time delays and RHP zeros) elements.

$$\tilde{g}_p(s) = \tilde{g}_{p+}(s) \tilde{g}_{p-}(s) \quad (3.21)$$

This factorization is performed so that the resulting controller will be stable.

2. Form the idealized IMC controller. The ideal internal model controller is the inverse of the invertible portion of the process model.

$$\tilde{q}(s) = \tilde{g}_{p-}^{-1}(s) \quad (3.22)$$

3. Add a filter to make the controller proper. A transfer function is proper if the order of the denominator polynomial is at least as high as the numerator polynomial.

$$q(s) = \tilde{q}(s)f(s) = \tilde{g}_{p-}^{-1}(s)f(s) \quad (3.23)$$

If it is most desirable to track step setpoint changes, the filter transfer function usually has the form

$$f(s) = \frac{1}{(\lambda s + 1)^n} \quad (3.24)$$

and n is chosen to make the controller proper (or semiproper). If it is most desirable to track ramp setpoint changes (often used for batch reactors or transition control problems), then

$$f(s) = \frac{n\lambda s + 1}{(\lambda s + 1)^n} \quad (3.25)$$

4. Adjust the filter-tuning parameter to vary the speed of response of the closed-loop system. If the λ is "small," the closed loop system is "fast," if λ is "large," the closed-loop system is more robust (insensitive to model error) [21].

If the process model is perfect, then we can easily calculate what the output response to a setpoint change will be. Substituting Equation (3.9) into Equation (3.4), found as

$$y(s) = g_p(s)q(s)r(s) = g_p(s)\tilde{q}(s)f(s)r(s) = g_p(s)\tilde{g}_{p-}^{-1}(s)f(s)r(s) \quad (3.26)$$

If the model is perfect, then

$$g_p(s) = \tilde{g}_p(s) = \tilde{g}_{p+}(s)\tilde{g}_{p-}(s) \quad (3.27)$$

which yields

$$y(s) = \tilde{g}_{p+}(s)f(s)r(s) \quad (3.28)$$

Equation (3.15) indicates that the bad stuff must appear in the output response. That is, if the open-loop process has a RHP zero (inverse response), then the closed-loop system must exhibit inverse response. Also, if the process has dead time, then dead time must appear in the closed-loop response [21].

The most common process model is a first-order plus time-delay transfer function. The design procedure for this system is given below.

Example: First Order Plus Dead Time Process

Consider a first order plus time delay model:

$$g_p(s) = \frac{K_p e^{-\ell s}}{\tau_p s + 1} \quad (3.29)$$

By using the four-step design procedure, first factor out the noninvertible elements,

$$\tilde{g}_p(s) = \tilde{g}_{p+}(s) \tilde{g}_{p-}(s) = e^{-\ell s} \cdot \frac{K_p}{\tau_p s + 1} \quad (3.30)$$

Then form the idealized IMC controller,

$$\tilde{q}(s) = \tilde{g}_p^{-1}(s) = \frac{\tau_p s + 1}{K_p} \quad (3.31)$$

and add a filter to make the controller proper

$$q(s) = \tilde{q}(s) f(s) = \tilde{g}_p^{-1}(s) f(s) = \frac{\tau_p s + 1}{K_p} \cdot \frac{1}{\lambda s + 1} = \frac{1}{K_p} \cdot \frac{\tau_p s + 1}{\lambda s + 1} \quad (3.32)$$

Once again, the controller is of lead-lag form. Finally, adjust λ for response speed and robustness. The closed-loop response (assuming a perfect model) to a setpoint change is [21];

$$y(s) = \tilde{g}_{p+}(s) f(s) r(s) = \frac{e^{-\ell s}}{\lambda s + 1} r(s) \quad (3.33)$$

For a step setpoint change of magnitude R,

$$y(t) = 0, \quad 0 \leq t < \theta$$

$$y(t) = R(1 - e^{-(t-\theta)/\lambda}), \quad t \geq \theta$$

3.4 The Dual Mode Concept

The most severe demand would certainly be to follow a step change in set point perfectly. This could be demanded of the controller but not of the process because it requires infinite process gain. The speed at which a variable can change is limited by the maximum rate at which energy can be delivered to the process. A valve may only open fully, not infinitely. Therefore it can only be asked that the controller not interfere with the maximum speed of the process. To duplicate the remainder of the step input, the control loop must be stable to the point that no overshoot or oscillation is observable. Nor should there be any offset. Finally, the controller ought to be insensitive to input noise, which is usually present in some form. To summarize, the ultimate controller should be capable of achieving the following loop-response characteristics [22]:

1. Maximum speed
2. Critical damping
3. No offset
4. Insensitivity to noise

Any control system that can satisfy the above demands will also satisfy any minimum-integral-error criterion, regardless of what function of the error may be used and regardless also of the nature of the input signals. The character of the process determines the complexity of the controller which is to accomplish the goals listed above. If the process is a pure single capacity, an on-off controller will provide maximum speed, critical damping, and no offset. An on-off controller is sensitive to noise, however. Significantly, this simplest control device is capable of achieving the ideal closed-loop response on the simplest process. As the process complexity increases, on-off control is no longer optimum, and combinations of less severe linear or nonlinear elements must be used to provide stability [22].

With regard to difficult processes, control functions which approach the demands of the four points of performance listed above need to be set forth:

1. Maximum speed implies that the controller be saturated for any measurable deviation. Such a demand limits the selection to an on-off controller. But if the tolerance may be widened somewhat, the controller need only saturate in response to a large signal. (How large the signal must be will vary with the difficulty of the process.)

2. Critical damping can be achieved by both low gain and derivative action, but the latter amplifies noise. Critical damping implies an asymptotic approach to the set point. To accommodate maximum speed, the zone of critical damping must be restricted to a narrow band about the set point. Therefore this criterion and its solution apply specifically to small-signal response.

3. Zero offset requires a controller with infinite gain in the steady state. An integrator is sufficient to satisfy this criterion.

4. Low noise response can be obtained through low-gain or low-pass filtering, but low-pass filtering degrades the speed of response of the loop. The only condition, then, which tends to reconcile this requirement with the others, is the application of low gain to small signals [22].

Of particular significance is the combination of high gain to large signals and low gain to small signals. The exact combination of parameters that will be most effective for a specific application may not be obtainable in a single controller. It may then be necessary to use two controllers, intelligently programmed to take the best advantage of their individual features. The combination of two controllers operating sequentially in the same loop has been called a dual-mode system.

The most valuable application of a dual-mode systems in the process industries is in batch-reactor control. Reactor temperature is usually controlled by setting the temperature controller for the heat-transfer medium in cascade.

The dual-mode system needs seven adjustments, which fall into two independent groups. Settings of proportional, integral, and derivative only pertain to the steady state, while the program settings are in effect elsewhere [22].

3.5 Controller Performance Criteria

Specifications for Closed Loop Response

There are a number of time-domain specifications. A few of the most frequently used dynamic specifications are listed below:

1. Closed Loop damping coefficient
2. Overshoot: The magnitude by which the controlled variable swings past the setpoint.
3. Rise Time (speed of response): The time it takes the process to come up to the new setpoint.
4. Decay Ratio: The ratio of the maximum amplitudes of successive oscillations.
5. Settling Time: The time it takes the amplitude of the oscillations to decay to some fraction (5 %) of the change in set point [1].

The design relations represented in Table 2 were developed to provide a closedloop response with a $\frac{1}{4}$ decay ratio. This performance criterion has several disadvantages:

- a) Responses with $\frac{1}{4}$ decay ratios are often judged to be too oscillatory by plant operating personnel.
- b) The criterion considers only two points of the closed loop response $c(t)$, namely the first two peaks [1].

An alternative approach is to develop controller design relations based on a performance index that considers the entire closed loop response. Three popular performance indices are given in Table 3.4.

Table 3.4 : Design Relations Based on Integral Error Criteria

Name of the error criterion	Formula
Integral of the absolute value of the error IAE	$IAE = \int_0^{\infty} e(t) dt$
Integral of the squared error ISE	$ISE = \int_0^{\infty} e(t)^2 dt$
Integral of the time weighted absolute error ITAE	$ITAE = \int_0^{\infty} t e(t) dt$

Design relations for PID controllers have been developed that minimize these integral error criteria for simple process models and a particular type of load or set-point change. In order to establish values of the settings of a tuner, those criterias must be satisfied [18].

Gain and Phase Margins

If a poorly tuned control system operates with K_c near the stability limit K_{cu} (ultimate gain), the closed-loop system could approach unstable operation. As measures of relative stability, the terms gain margin (GM) and phase margin (PM) often are used. Figure 3.8 illustrates the concepts of gain and phase margin [18].

Let AR_c (critical amplitude ratio) be the value of the open-loop amplitude ratio at the critical (or phase crossover) frequency w_c . The gain margin is defined as

$$GM = \frac{1}{AR_c} \quad (3.34)$$

From the Bode stability criterion, AR_c must be less than one to have a stable closed-loop system. Thus $GM > 1$ is a stability requirement.

Define w_g as the frequency at which the open-loop gain is unity (the gain crossover frequency). Let ϕ_g denote the phase angle w_g . Phase margin PM is defined as

$$PM = 180 + \phi_g \quad (3.35)$$

Controller manufacturers recommend that a well-tuned controller have a gain margin between 1.7 and 2.0, while the phase margin should be between 30 and 45°. These ranges are approximate and it is often not possible to choose controller settings that fix both GM and PM at arbitrary values. The GM and PM concepts are not meaningful when the open-loop system has multiple values for w_c or w_g [18].

The recommended values of phase and gain margins provide a compromise between performance and safety. Large values of GM and PM cause sluggish closed-loop response, while smaller values result in a less sluggish, more oscillatory response. The choice of GM and PM should also depend upon the level of confidence in the process model and how much the process parameters can change. For example, if the dominant time constant or time delay for a process depends on the flow rate (throughput), the phase margin will change if the flow rate changes [18].

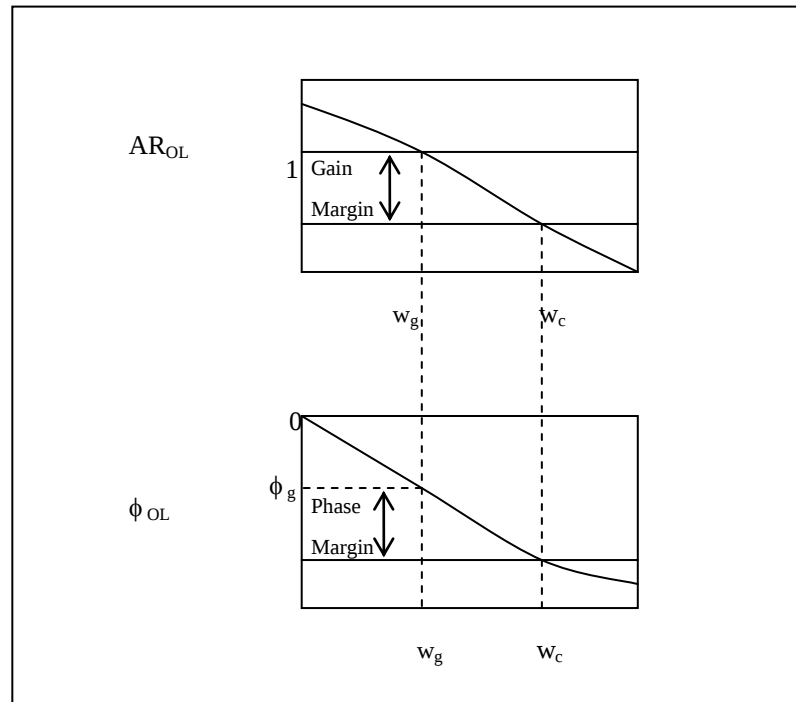


Figure 3.8 : Gain and phase margins on Bode plot

4. MODELING AND CONTROL OF THE PROCESS

In this part the modeling and control of the reacting distillation process are performed.

4.1 The Reacting Distillation Process

In this study, the temperature profile in a reboiler of reacting distillation process producing acetate esters from the fusel alcohols is analysed. The aim is to control the liquid and vapor phases in the reboiler of the reacting distillation column by using DPIMCPID. The experimental data is obtained from the study of Tanrıverdi (1996) and represented in Figure 4.1. In his study, the chosen by-product to be evaluated was fusel oil that has azeotropic properties [23]. To mention that, the separation of azeotrope mixtures by using the multicomponent reacting distillation column properly is difficult due to their closer boiling points which gives an excessive complexity to the system. Prior to describing the reacting distillation process, fusel oil properties and its usage is explained first.

Fusel oil is a by-product of the distillation of ethyl alcohol from the fermentation of molasses and contains mainly C3-C5 alcohols. Acetic acid and butyric acid esters of its major alcohol components have economic value as chemicals for flavor and fragrance manufacturing [24].

One of the main products of sugar manufacturing is molasses, which contains approximately 50% sucrose and 50% other components (water, various other organic components and inorganic salts). Because of its high sucrose content, a substantial portion of the molasses is used for the production of ethyl alcohol through fermentation. In recent decades, 12 million tons of sugar beet have been processed and approximately 550 thousand tons of molasses have been obtained per year, in Turkey [24].

Approximately, 30 million liters of ethyl alcohol per year is produced from these molasses. The by-products of the fermentation broths, more volatile than the alcohol, are mainly aldehydes with acetaldehyde being the principal component. The aldehyde is removed, as a distillation head product [24].

The other by-product of the distillation step, the bottom product, is fusel oil. It is composed of several alcohols, primarily C3, C4 and C5 aliphatic alcohols. Approximately 1 liter of acetaldehyde and 5 liter of fusel oil are obtained for every 1000 liter of ethyl alcohol from the distillation. Raw fusel oil is a relatively viscous liquid with a dark-reddish color and a very unpleasant odor. As a result of these properties, the direct utilization of fusel oil as a solvent has been very limited. In some countries it is burned to supply energy for processing plants. In Turkey, it is used mainly for denaturation of alcohol or for removing the foam from molasses during sugar manufacturing. A substantial portion of it, however, has generally been discarded. Recent studies have suggested that several alternatives uses for fusel oil are possible. For example, acetic acid and butyric acid esters, major alcohol components of fusel oil, have economic value as chemicals for flavor and fragrance manufacturing. Especially ethylbutyrate is in high demand as a component of pineapple-banana flavors in the food industry [24].

The main advantage of reacting distillation is the combination of complex process units by combining reaction and separation simultaneously. To obtain desired reaction conversions as well as product purities in one piece of equipment, it is very important to understand the interaction between reaction and separation.

4.2 Identification of the Reacting Distillation Process

Having explained the reacting distillation process, the identification of the process is explained. Firstly, the process reaction curve method is used. The process reaction curve method gives first order process transfer function. However the multi component distillation process could not have a first order transfer function. Then, process integrator is applied to the obtained transfer function which leads to better results.

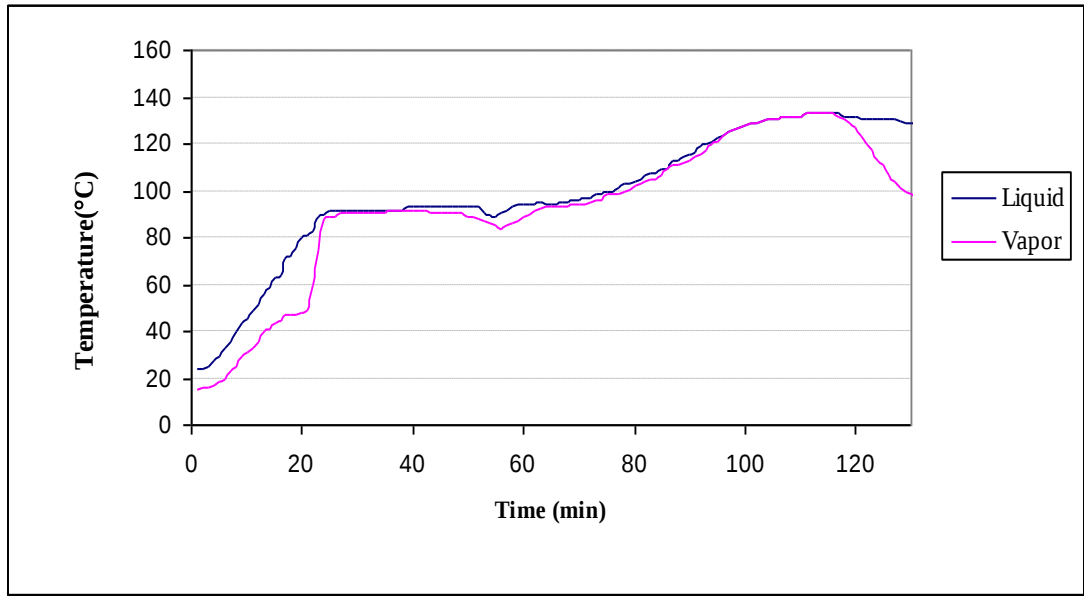


Figure 4.1 : Process reaction curve of a reboiler of the fusel alcohol reacting distillation column [23]

4.2.1 Determination of the Terms of Process Transfer Functions

Prior to explaining the determination of the process transfer functions, terms that describe the process characteristic are represented first.

Process Gain (K_p): It is defined as the ratio of the change in output, or responding variable, to the change in input, or forcing function.

Process Time Constant (τ_p): That is the amount of time counted from the moment the variable starts to respond that it takes the process variable to reach 63.2 % of its total change.

Process Dead Time (θ): It is a finite amount of time between the change in input variable and when the output variable starts to respond [25].

From Figure 4.1 the mentioned parameters were calculated as in the following.

$$\text{Liquid Phase } K_p = \frac{114.67^\circ\text{C}}{2500\text{W}} = 0.0458[^\circ\text{C}/\text{W}] = 0.182[^\circ\text{C}.\text{min}/\text{kcal}]$$

$$\text{Vapor Phase } K_p = \frac{117.33^\circ\text{C}}{(-)2500\text{W}} = 0.0469[^\circ\text{C}/\text{W}] = (-)0.187[^\circ\text{C}.\text{min}/\text{kcal}]$$

The vapor phase K_p has a minus sign that comes from the directional case. In calculations this sign is not shown.

Liquid Phase: $0.632 \cdot BK_p = 0.632 \cdot 114.67 = 72.24 \text{ } ^\circ\text{C}$ $\tau = 18 \text{ [min]}$

Vapor Phase: $0.632 \cdot BK_p = 0.632 \cdot 117.33 = 73.91 \text{ } ^\circ\text{C}$ $\tau = 30 \text{ [min]}$

Liquid Phase Delay Time θ : 2 [min]

Vapor Phase Delay Time θ : 0.3 [min]

4.2.2 Process Integration

According to the above calculations the process transfer function is found as a first order plus time delay (FOPDT) process. However, that does not give good results. The fusel oil distillation is a complex process due to its azeotrope components then that's why the integrator term is added into FOPDT, shown in (4.1)

$$G_p(s) = \frac{K_p \cdot e^{-\theta s}}{(\tau_p s + 1)} \quad \rightarrow \quad G_p(s) = \frac{K_p \cdot e^{-\theta s}}{(\tau_p s + 1)s} \quad (4.1)$$

4.3 Establishment of the Dual Phase IMC based PID (DPIMCPID)

4.3.1 Filter Selection

In the industrial processes the commonly used filters are analog and digital filters . In this study a low-pass filter (analog) with a steady state gain of one is used which has a function of $\frac{1}{(\tau_f s + 1)}$.

The analog and one type of digital filter forms are given in (4.3) and (4.4) respectively.

$$\frac{1}{(\tau_f s + 1)^n} \quad (4.3)$$

$$\frac{(\tau_f k s + 1)^n}{(\tau_f s + 1)^n} \quad (4.4)$$

The trial and error method is used to choose the best filter. Filter selection studies are shown in Appendix-A. Rivera (et al.) recommended that τ_f be selected for the process model FOPDT so that both $\tau_f / \theta > 0.8$ and $\tau_f > \tau_p / 10$ [26].

4.3.2 Pole Placement

Process transfer function has poles and zeros. The roots of the denominator are called the poles of the transfer function. Also the roots of the numerator are called the zeros of the transfer function. The roots of the characteristic equation, which are the poles of the transfer function, must be real or must occur as complex conjugate pairs. In addition to that, the real parts of all the poles must be negative for the system to be stable. A system is stable if all its poles lie in the left half of the s plane shown in Figure 4.2. The locations of the zeros of the transfer function have no effect on the stability of the system. They exactly affect the dynamic response, but they do not affect stability [1].

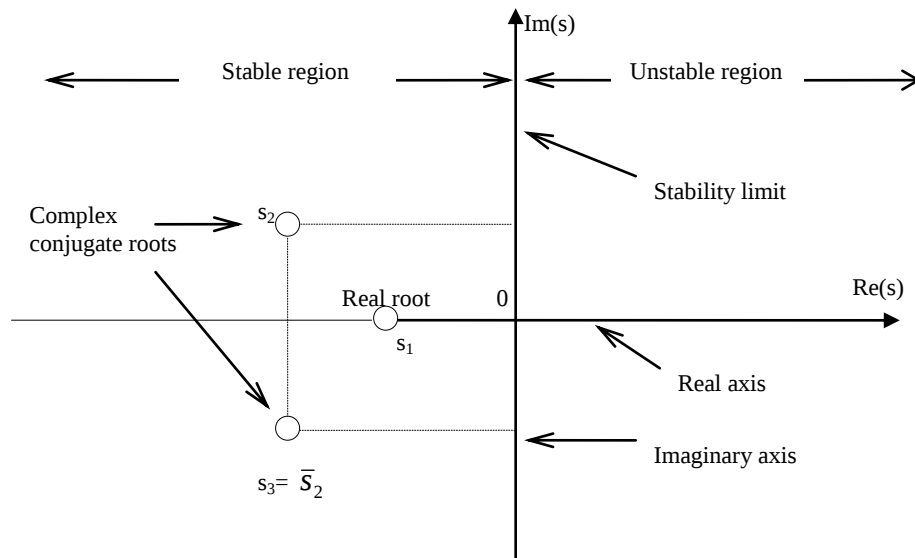


Figure 4.2 : S -plane of the roots of the characteristic equation

Processes having ‘process integration’ must be handled carefully as they have an extra root in the denominator which makes the system response faster but oscillatory. That integrator term acts as an accelerator. That is why; pole placement is used to get rid of this effect.

The pole placement can be used to the processes with integrator in the denominator of their transfer functions as in (4.1). Mathematically, when s is in the denominator, which means that pole is equal to zero. From the point of internal model control a transfer function is assumed to evaluate PID controller parameters. A perfect control is achieved, when the assumed transfer function is equal to the real process transfer function. The placement of a pole to the denominator has a good effect on achieving the perfect model. That means a small number that is closer to zero.

If 0.001 is chosen, the system indicates good results given in Figure 5.2 and its transfer function will be as in (4.5). However if too closer poles to zero are selected, the system will become too slow represented in Figure 4.3. Beside, when a pole is selected larger than 0.001(further from zero), not only will the system response make an overshoot, but also will be too slow as shown in Figure 4.4.

$$\tilde{G}_p(s) = \frac{K_p \cdot e^{-\theta s}}{(\tau_p s + 1)(s + 0.001)} \quad (4.5)$$

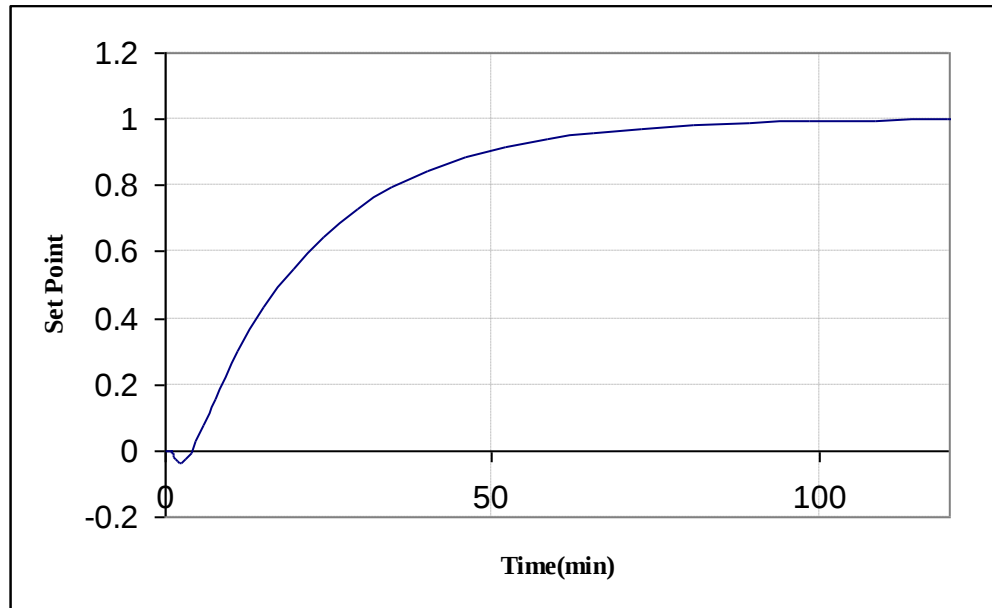


Figure 4.3 : The response of a process ($K_p=0.182$ $\tau_p=18$ $\theta = 2$ $\tau_f=1.9$) when the pole placement is ($s+0.0001$)

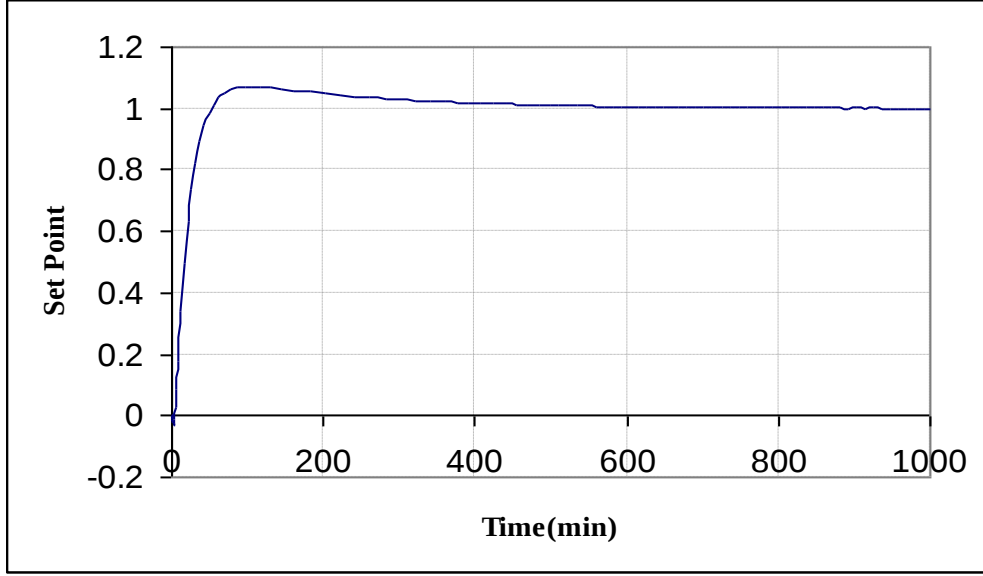


Figure 4.4 : The response of a process ($K_p=0.182$ $\tau_p=18$ $\theta = 2$ $\tau_f=1.9$) when the pole placement is $(s+0.005)$

On a different proses this pole placement is also examined and related calculations are given in Appendix-B. It is obviously seen that, pole placement gives better results in IMC design.

4.3.3 DPIMCPID Design

Let $\tilde{G}_p(s)$ as ;

$$\tilde{G}_p(s) = \frac{K_p \cdot e^{-\theta s}}{(\tau_p s + 1)(s + 0.001)} \quad (4.6)$$

To transform $e^{-\theta s}$, series approximation is used then the model transfer function becomes:

$$\tilde{G}_p(s) = \frac{K_p \cdot (1 - \theta \cdot s)}{(\tau_p s + 1)(s + 0.001)} \quad (4.7)$$

$$\tilde{G}_p(s) = \tilde{G}^+ \tilde{G}^- \quad \text{where} \quad \tilde{G}^+ = 1 - \theta \cdot s \quad \tilde{G}^- = \frac{K_p}{(s + 0.001)(\tau_p s + 1)} \quad (4.8)$$

The IMC filter is in the form of $f = \frac{1}{(\tau_F s + 1)}$ (4.9)

$$G_C^* = \frac{1}{\tilde{G}^-} f \quad (4.10)$$

$$G_C^* = \frac{(s+0.001)(\tau_P s+1)}{K(\tau_F s+1)} \quad G_{PID} = \frac{G_C^*}{1-G_C^* \tilde{G}_P} \quad (4.11)$$

$$G_{PID} = \frac{\frac{(s+0.001)(\tau_P s+1)}{K_P(\tau_F s+1)}}{1 - \frac{(s+0.001)(\tau_P s+1)}{K(\tau_F s+1)} \frac{K_P(1-\theta s)}{(s+0.001)(\tau_P s+1)}} \quad (4.12)$$

$$G_{PID} = \frac{\tau_P s^2 + s + 0.001\tau_P s + 0.001}{K_P(\tau_F s + \theta s)} \quad G_{PID} = \frac{\tau_P s^2 + (1+0.001\tau_P)s + 0.001}{K_P(\tau_F + \theta)s} \quad (4.13)$$

$$G_{PID} = \frac{\tau_P s}{K_P(\tau_F + \theta)} + \frac{(1+0.001\tau_P)}{K_P(\tau_F + \theta)} + \frac{0.001}{K_P(\tau_F + \theta)s} \quad (4.14)$$

$$G_{PID} = \frac{1}{K_P(\tau_F + \theta)} \left(\tau_P s + (1+0.001\tau_P) + \frac{0.001}{s} \right) \quad (4.15)$$

$$G_{PID} = \frac{1/(1+0.001\tau_P)}{K_P(\tau_F + \theta)} \left(\frac{\tau_P s}{(1+0.001\tau_P)} + 1 + \frac{0.001}{s(1+0.001\tau_P)} \right) \quad (4.16)$$

$$K_C = \frac{1}{K_P} \frac{1/(1+0.001\tau_P)}{\tau_F + \theta} \quad \tau_I = \frac{1+0.001\tau_P}{0.001} \quad \tau_D = \frac{\tau_P}{1+0.001\tau_P} \quad (4.17)$$

After obtaining those equations, IMC design is performed for both phases in Chapter 5.

5. SIMULATION STUDIES

In this study Matlab 6.5 and VEE Pro 7.0 programs are used to indicate the results.

5.1 Matlab 6.5 Applications

5.1.1 DPIMCPID Design for Both Phases

Liquid Temperature Control

$$K_p=0.182 \quad \tau_p=18 \quad \theta = 2 \quad \tau_f=1.9$$

$$K_c=1.384 \quad \tau_I=1018 \quad \tau_D=17.7$$

Transfer function of the feedforward path (G_{OL}) :

$$\frac{-9068s^3 + 4021s^2 + 255.9s + 0.2519}{18324s^3 + 1018s^2}$$

Simulink model of the feedforward path for step-up input is shown in Figure 5.1. The response curve is shown in Figure 5.2.

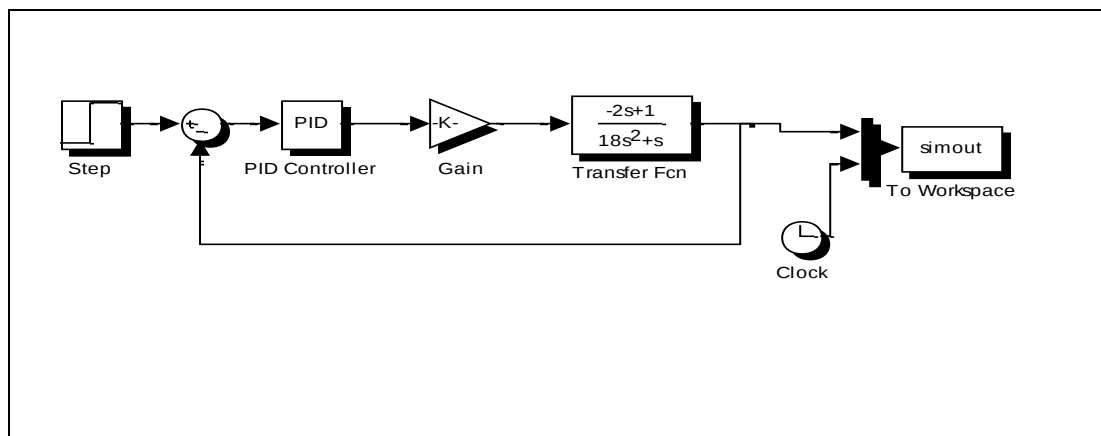


Figure 5.1 : SIMULINK model of the liquid phase controller

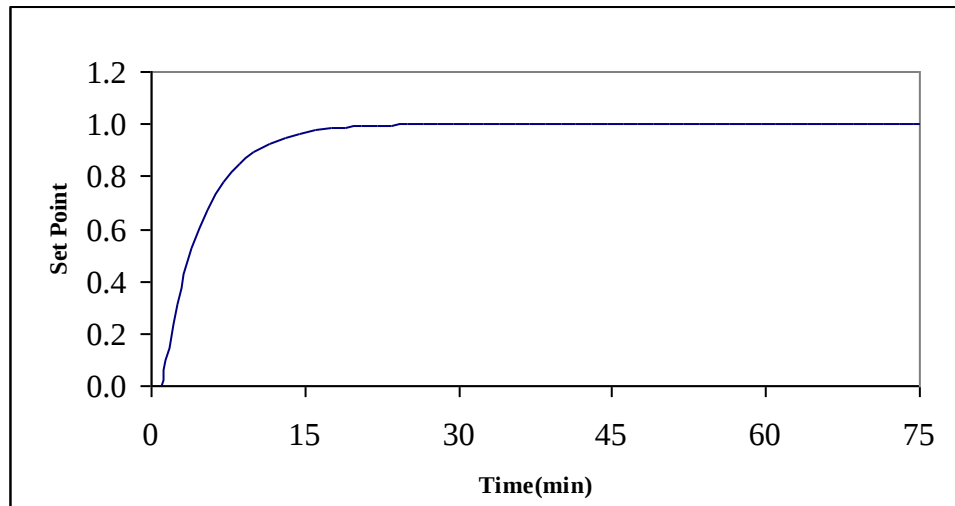


Figure 5.2 : The response of the liquid phase to the given step-up input

Simulink model of the feedforward path for step-down input is shown in Figure 5.3. The response curve is shown in Figure 5.4. Bode plot is also given in Figure 5.5.

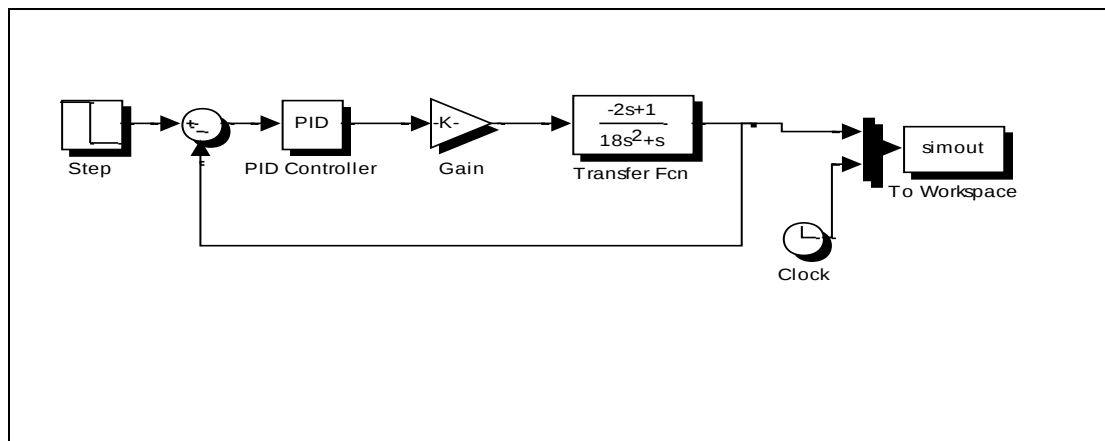


Figure 5.3 : SIMULINK model of the liquid phase controller

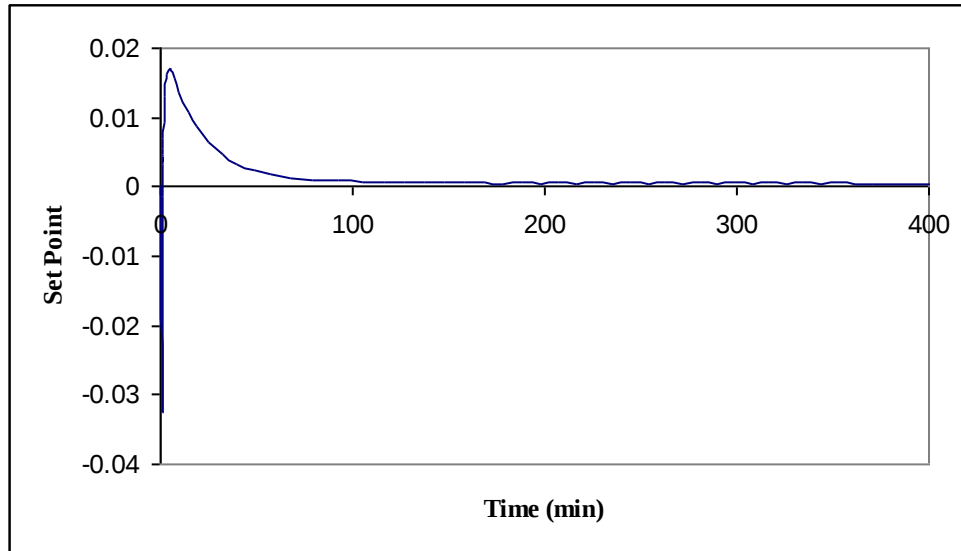


Figure 5.4 : The response of the liquid phase to the given step-down input

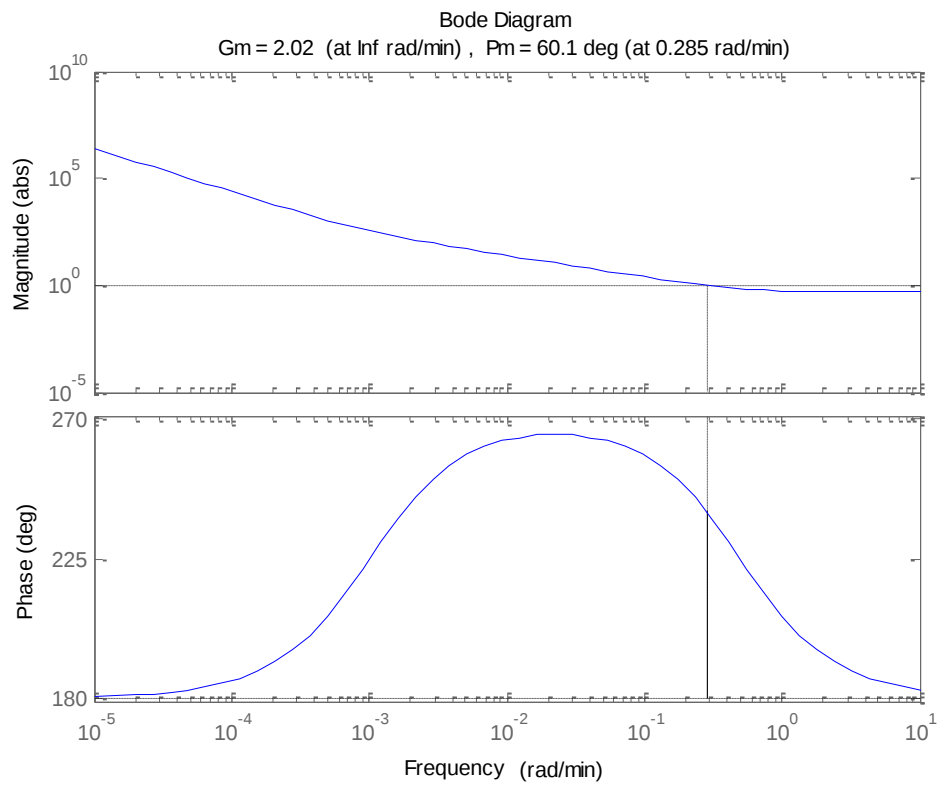


Figure 5.5 : The Bode plot of the DPIMCPID_L

Vapor Temperature Control

$$K_c = \frac{1}{K_p} \frac{1/(1+0.001\tau_p)}{\tau_F + \theta} \quad \tau_I = \frac{1+0.001\tau_p}{0.001} \quad \tau_D = \frac{\tau_p}{1+0.001\tau_p}$$

$$K_p = 0.187 \quad \tau_p = 30 \quad \theta = 0.3 \quad \tau_f = 3.1$$

$$K_c = 1.527 \quad \tau_I = 1030 \quad \tau_D = 29.13$$

Transfer function of the feedforward path:

$$\frac{-2570s^3 + 8478s^2 + 294s + 0.256}{30900s^3 + 1030s^2}$$

Simulink model of the feedforward path for step-up input is shown in Figure 5.6. The response curve is shown in Figure 5.7.

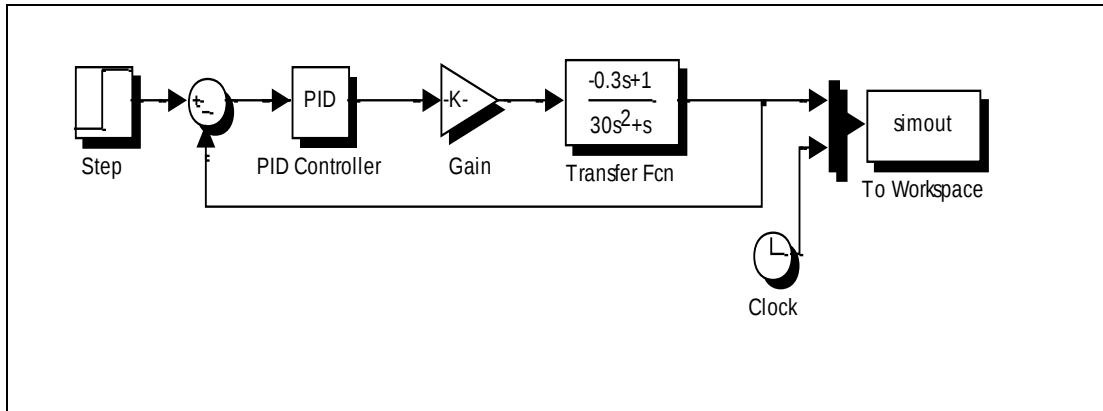


Figure 5.6 : SIMULINK model of the vapor phase controller

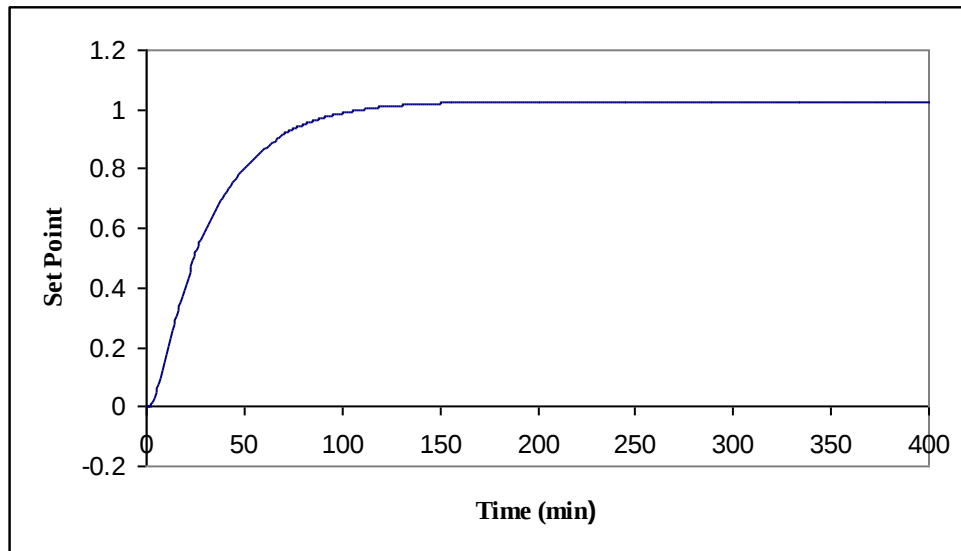


Figure 5.7 : The response of the vapor phase to the given step-up input

Simulink model of the feedforward path for step-down input is shown in Figure 5.8. The response curve is shown in Figure 5.9. Bode plot is also given in Figure 5.10.

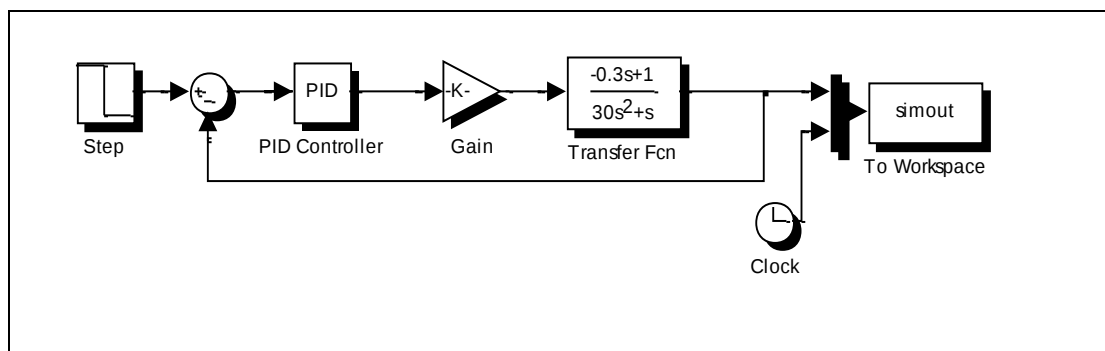


Figure 5.8 : SIMULINK model of the vapor phase controller

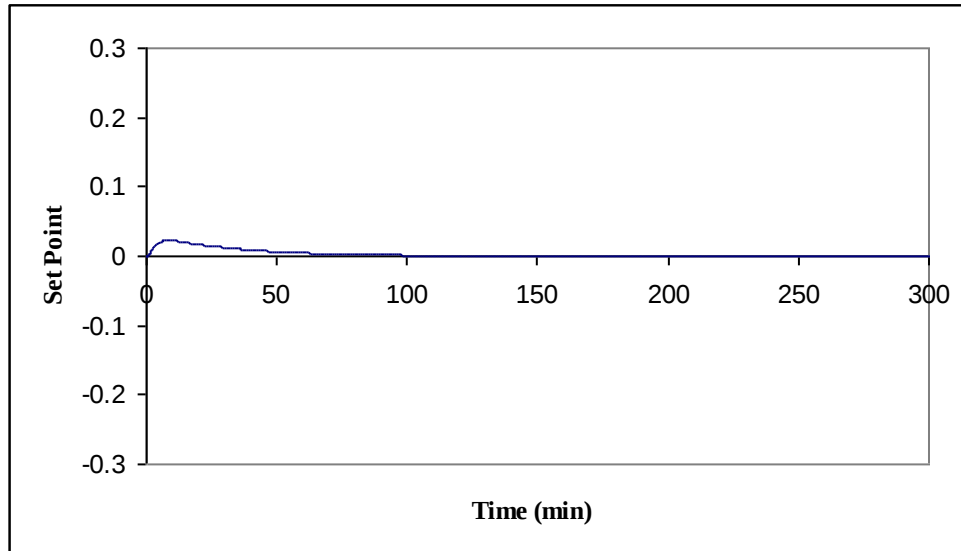


Figure 5.9 : The response of the vapor phase to the given step-down input

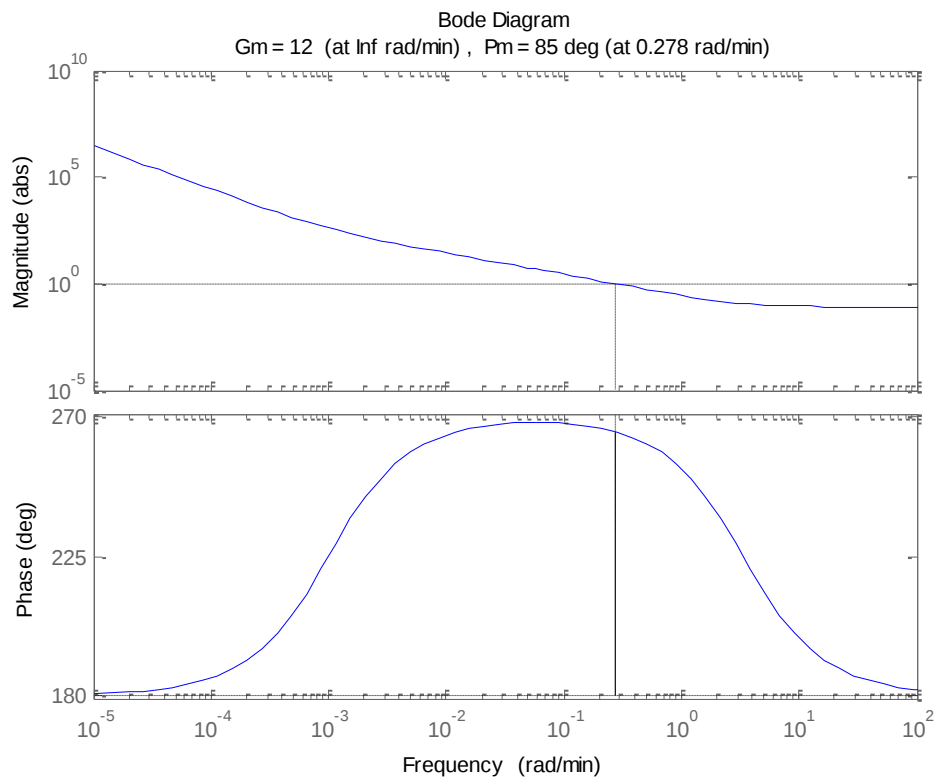


Figure 5.10 : The Bode plot of the DPIMCPID_v

5.2.1 Ziegler Nichols Controller Method

Liquid System

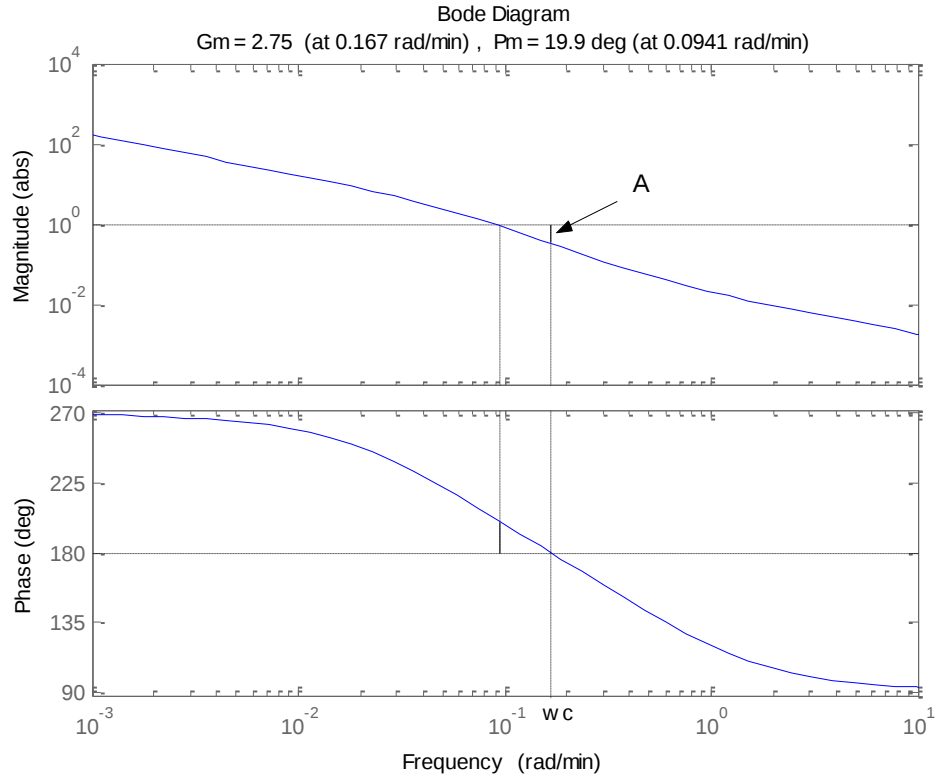


Figure 5.11 : The Bode plot of the liquid phase model transfer function

Firstly, the Bode plot is drawn to calculate ultimate gain and ultimate period and represented in Figure 5.11.

A (critical amplitude ratio) = 0.363

$$K_u = \frac{1}{A} = 2.75 \quad \text{at } w_c = 0.167 \text{ [rad/min]}$$

$$P_u = \frac{2\pi}{w_c} = 0.627 \text{ [min]}$$

➤ For a PID controller *original (1/4 decay ratio)* Z-N settings

$$K_c = 0.6 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{8}$$

$$K_c = 1.65 \quad \tau_I = 18.8 \quad \tau_D = 4.7$$

Transfer function:
$$\frac{-53.1s^3 + 15.25s^2 + 5.408s + 0.3003}{338.6s^3 + 18.81s^2}$$

The response curve is shown in Figure 5.12. Bode plot is also given in Figure 5.13.

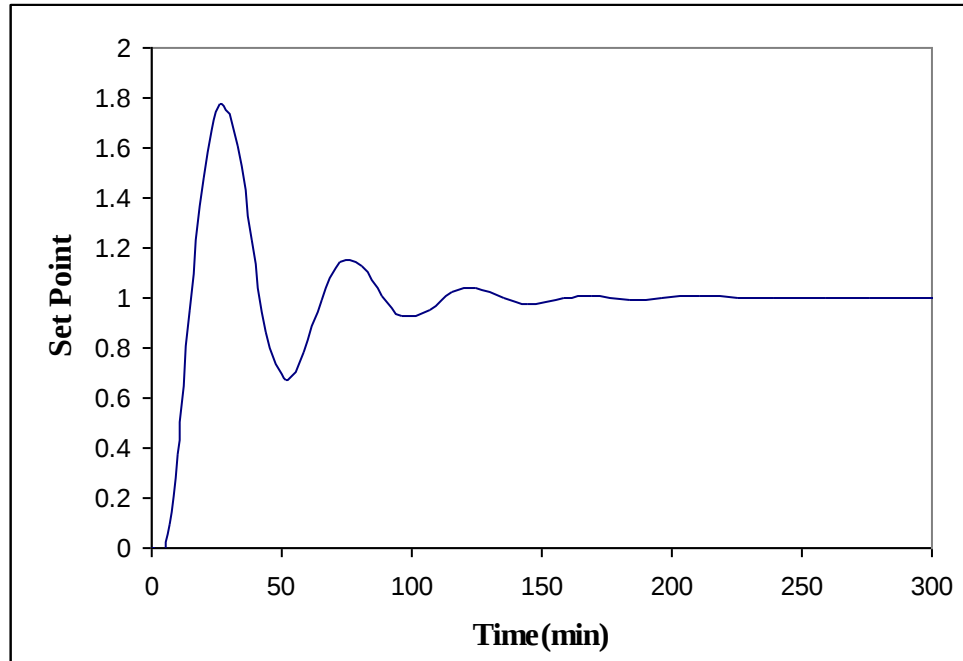


Figure 5.12 : The response of the liquid phase to the given step-up input by Ziegler Nichols Method (original)

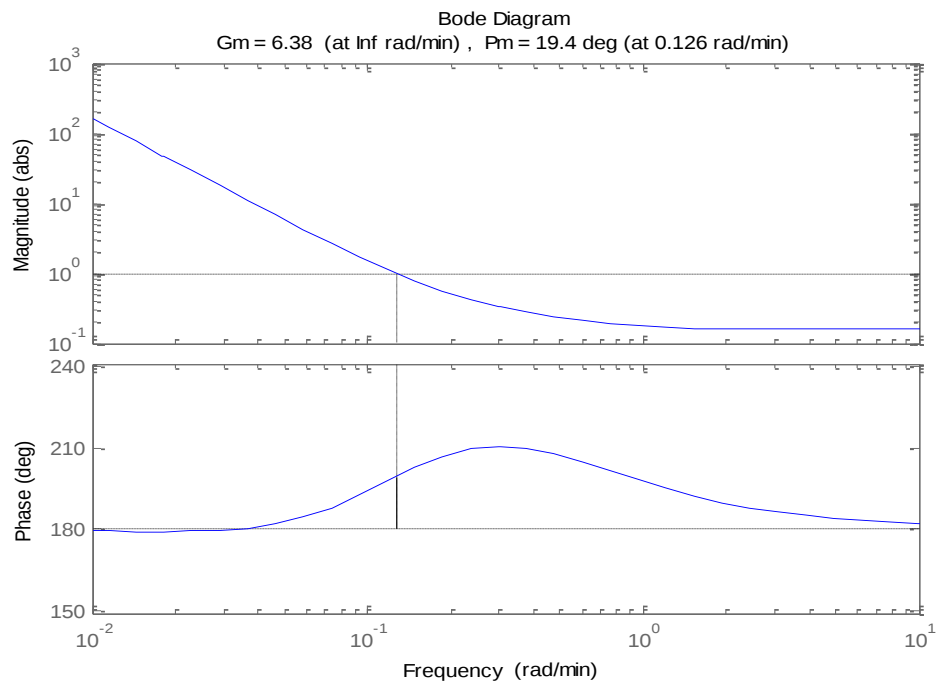


Figure 5.13 : The Bode plot of the liquid phase by Z-N method (original)

➤ For a PID controller Z-N settings with some overshoot

$$K_c = 0.33 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{3}$$

$$K_c = 0.907 \quad \tau_I = 18.81 \quad \tau_D = 12.54$$

Transfer function:

$$\frac{-77.87s^3 + 32.73s^2 + 2.775s + 0.1651}{338.6s^3 + 18.81s^2}$$

The response curve is shown in Figure 5.14. Bode plot is also given in Figure 5.15.

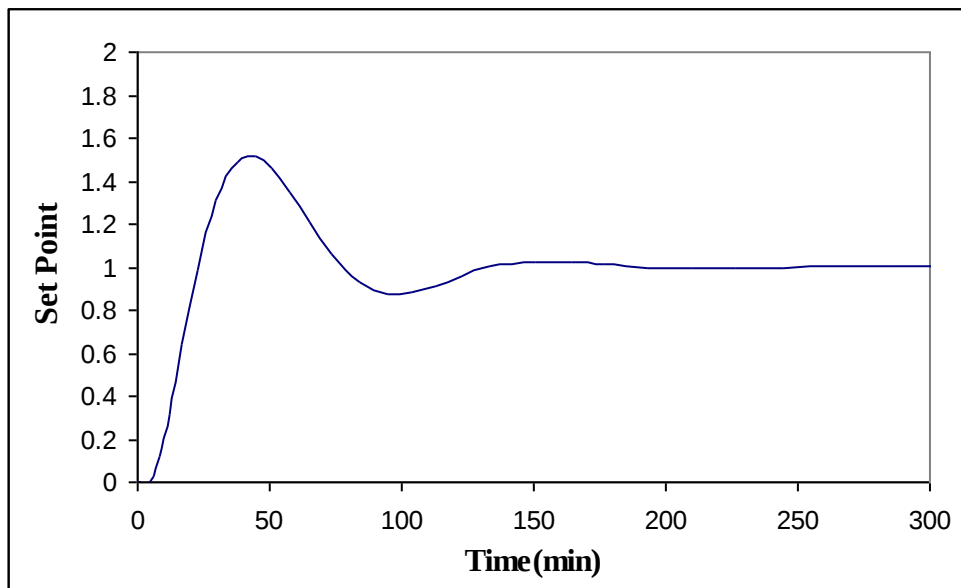


Figure 5.14 : The response of the liquid phase to the given step-up input by Ziegler Nichols Method (some overshoot)

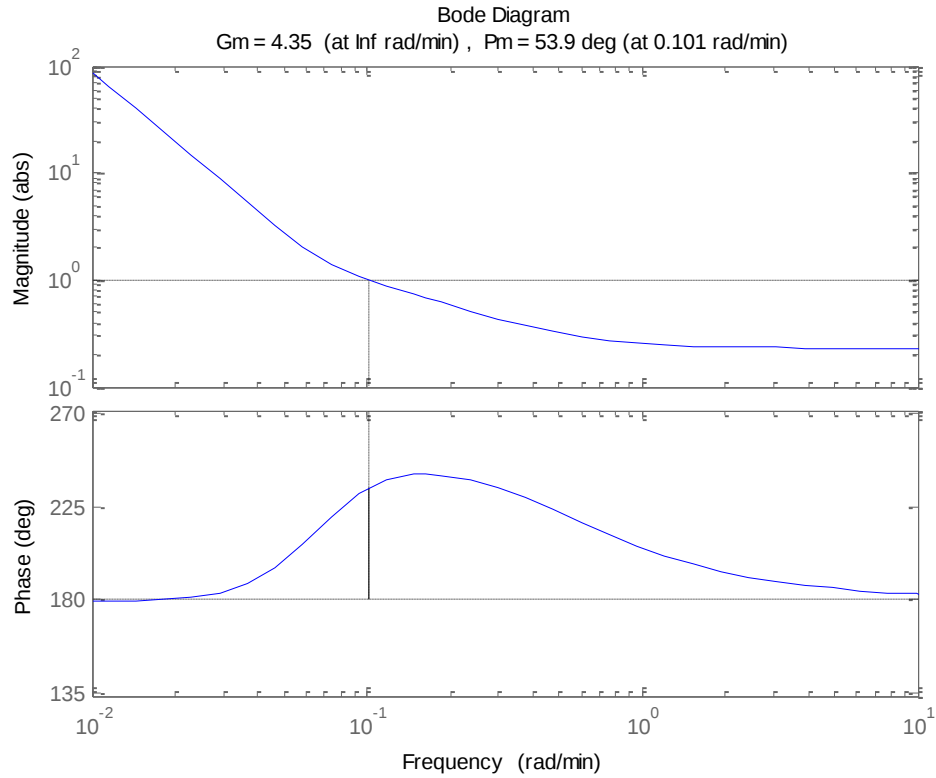


Figure 5.15 : The Bode plot of the liquid phase by Z-N method (some overshoot)

➤ For a PID controller Z-N settings with no overshoot

$$K_c = 0.2 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{3}$$

$$K_c = 0.55 \quad \tau_I = 18.81 \quad \tau_D = 12.54$$

Transfer function:

$$\frac{-47.22s^3 + 19.85s^2 + 1.683s + 0.1001}{338.6s^3 + 18.81s^2}$$

The response curve is shown in Figure 5.16. Bode plot is also given in Figure 5.17.

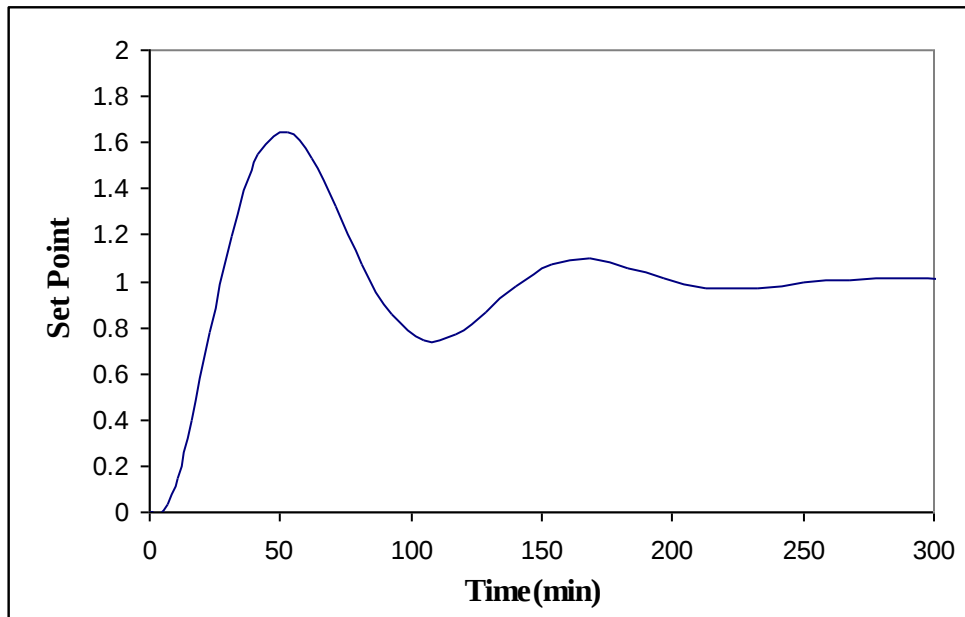


Figure 5.16 : The response of the liquid phase to the given step-up input by Ziegler Nichols Method (no overshoot)

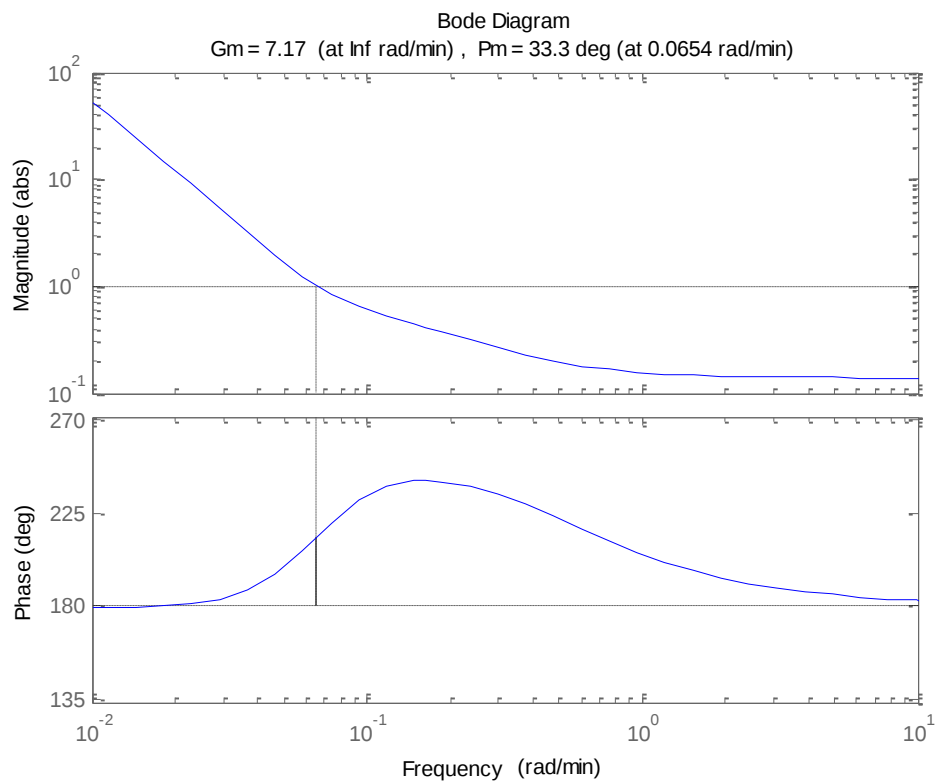


Figure 5.17 : The Bode plot of the liquid phase by Z-N method (no overshoot)

Vapor System

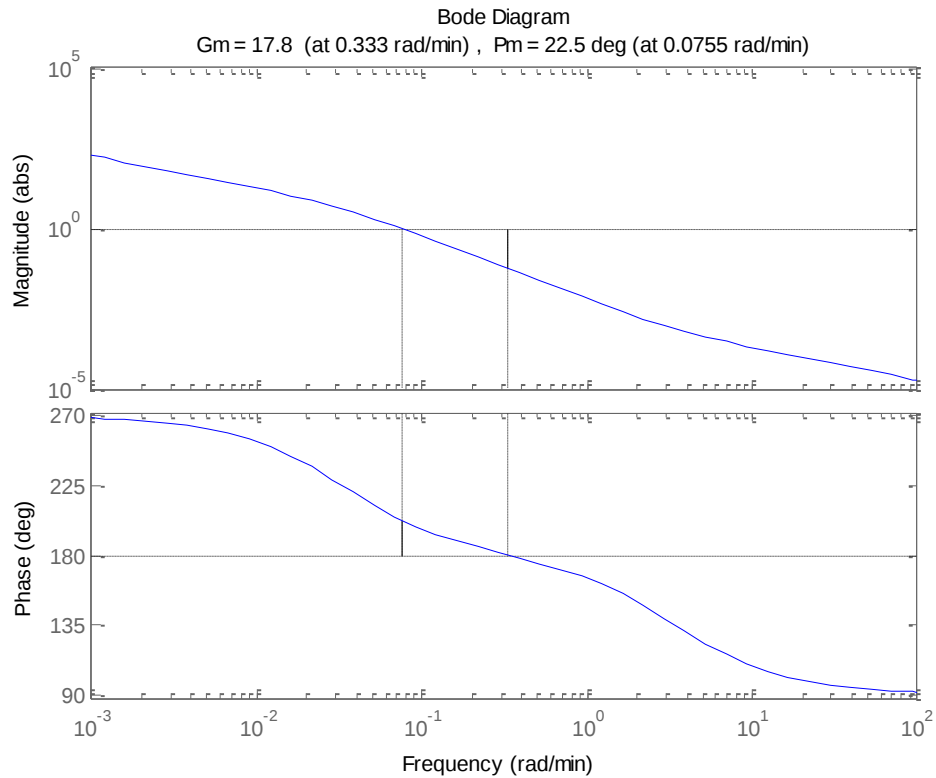


Figure 5.18 : Bode plot of vapor phase transfer function

Firstly, the Bode plot is drawn to calculate ultimate gain and ultimate period and represented in Figure 5.18.

$$A = 0.0561$$

$$K_u = \frac{1}{A} = 17.8 \quad \text{at } \omega_c = 0.333 \text{ [rad/min]}$$

$$P_u = \frac{2\pi}{\omega_c} = 18.87$$

➤ For a PID controller *original (1/4 decay ratio)* Z-N settings

$$K_c = 0.6 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{8}$$

$$K_c = 10.68 \quad \tau_I = 9.43 \quad \tau_D = 2.36$$

$$\text{Transfer function: } \frac{-13.33s^3 + 38.78s^2 + 18.23s + 1.997}{282.9s^3 + 9.43s^2}$$

Step-up and step-down inputs are given to the system and the response curves are shown in Figure 5.19 and Figure 5.20. Bode plot is also drawn to understand stability criterion in Figure 5.21.

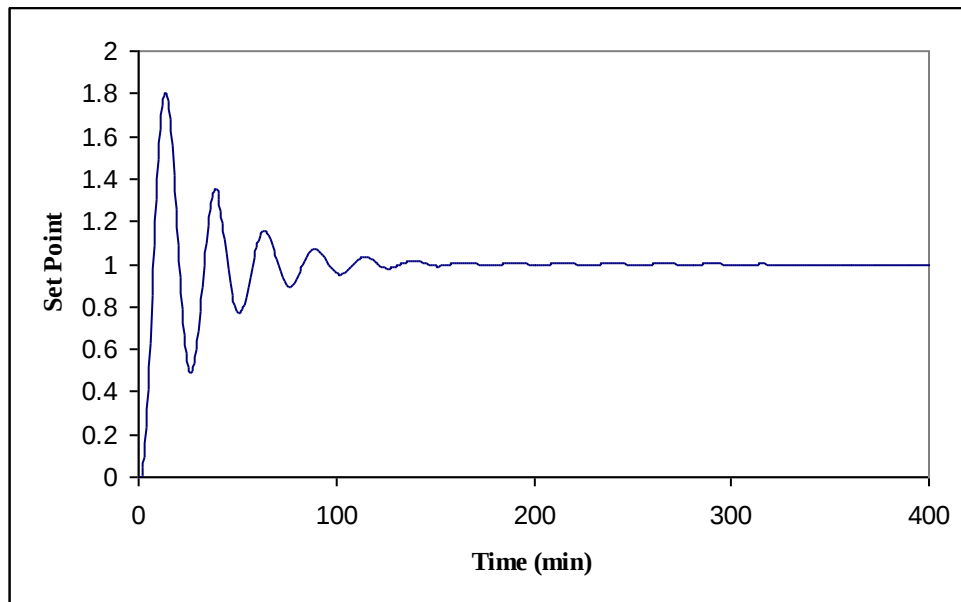


Figure 5.19 : The response of the vapor phase to the given step-up input by Ziegler Nichols Method (original)

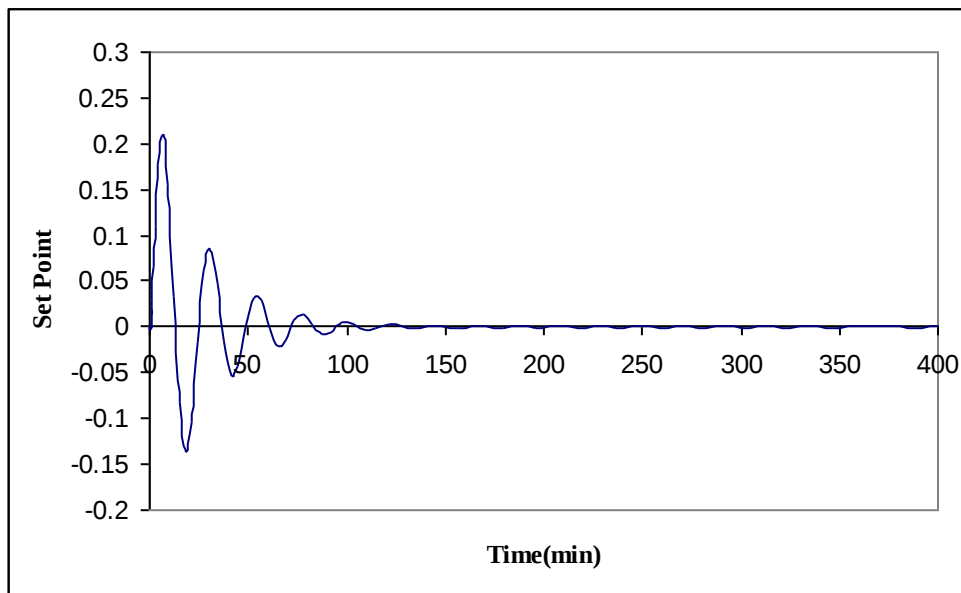


Figure 5.20 : The response of the vapor phase to the given step-down input by Ziegler Nichols Method (original)

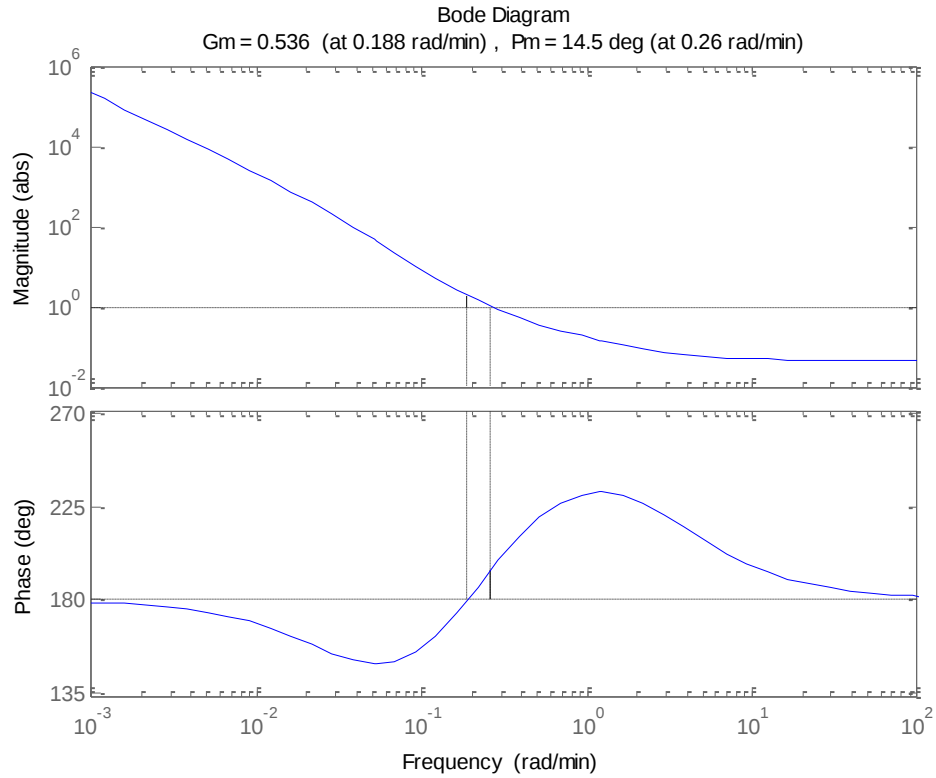


Figure 5.21 : The Bode plot of the vapor phase by Z-N method (original)

➤ For a PID controller Z-N settings with some overshoot

$$K_c = 0.33 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{3}$$

$$K_c = 5.9 \quad \tau_I = 9.43 \quad \tau_D = 6.29$$

Transfer function:

$$\frac{-19.63s^3 + 62.32s^2 + 10.07s + 1.103}{282.9s^3 + 9.43s^2}$$

Step-up and step-down inputs are given to the system and the response curves are shown in Figure 5.22 and Figure 5.23. Bode plot is also drawn to understand stability criterion in Figure 5.24.

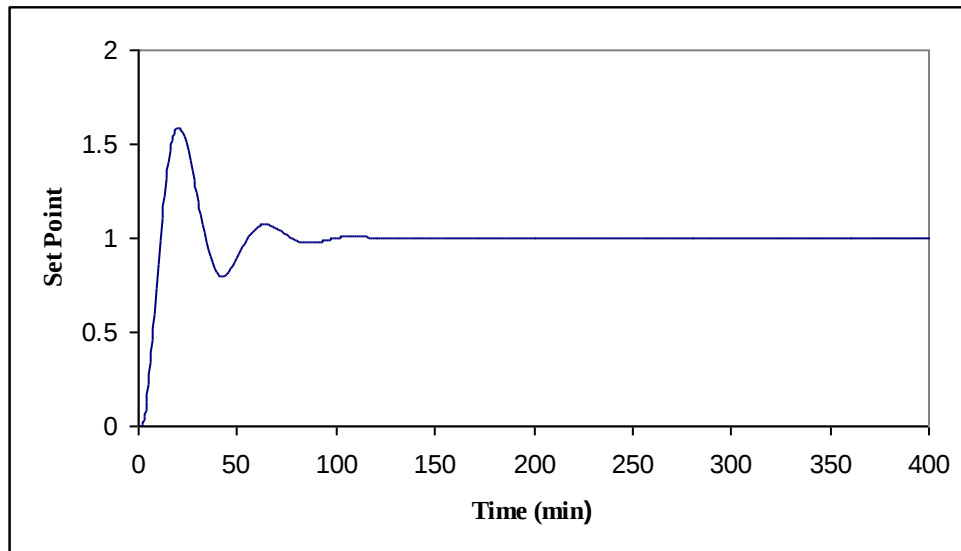


Figure 5.22 : The response of the vapor phase to the given step-up input by Ziegler Nichols Method (some overshoot)

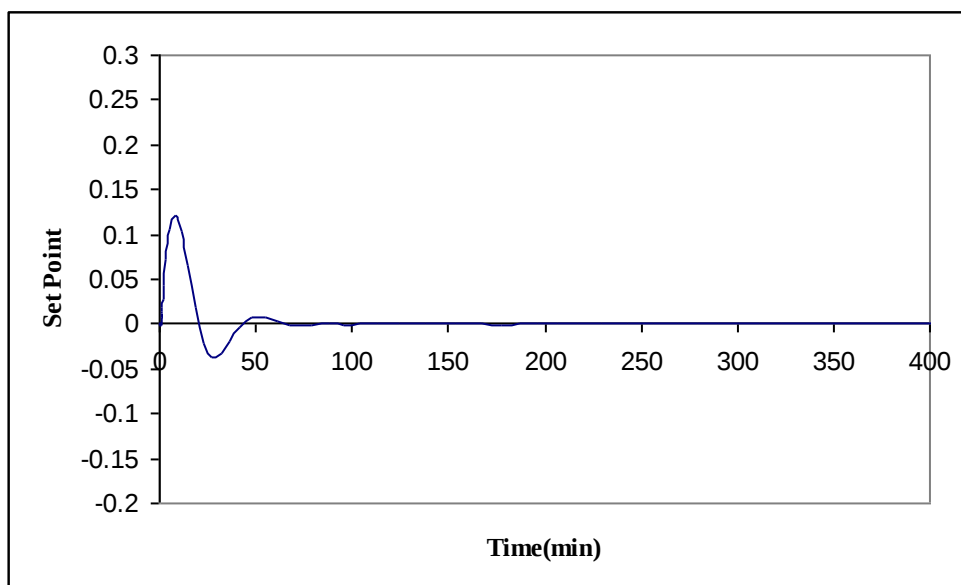


Figure 5.23 : The response of the vapor phase to the given step-down input by Ziegler Nichols Method (some overshoot)

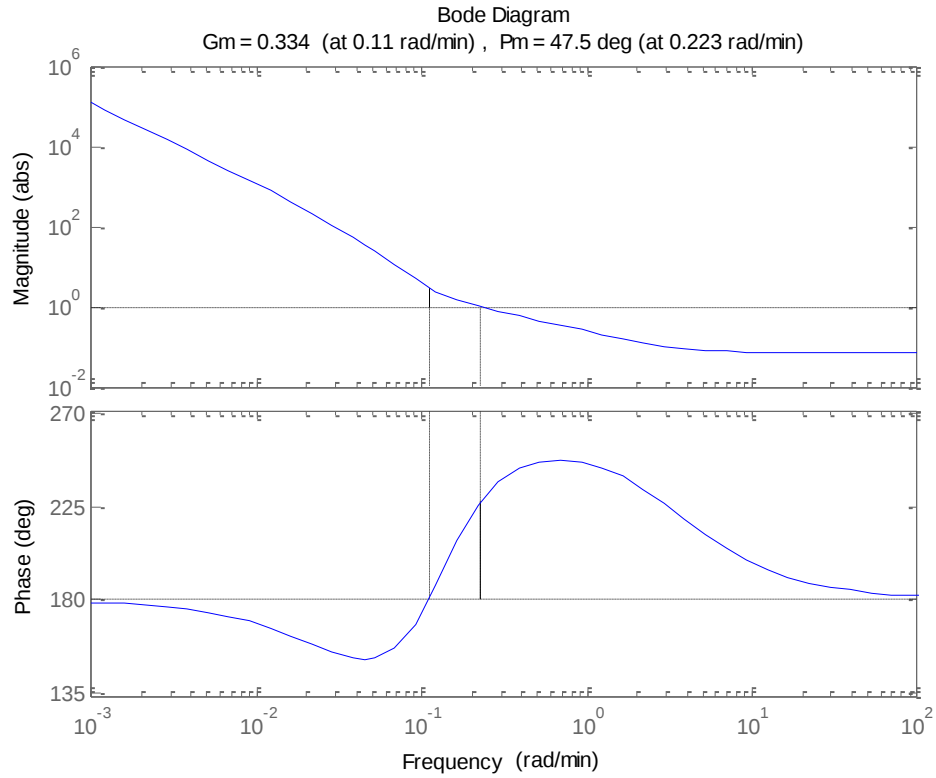


Figure 5.24 : The Bode plot of the vapor phase by Z-N method (some overshoot)

➤ For a PID controller Z-N settings with no overshoot

$$K_c = 0.2 * K_u \quad \tau_I = \frac{P_u}{2} \quad \tau_D = \frac{P_u}{3}$$

$$K_c = 3.56 \quad \tau_I = 9.43 \quad \tau_D = 6.29$$

Transfer function:

$$\frac{-11.85s^3 + 37.6s^2 + 6.708s + 0.6657}{282.9s^3 + 9.43s^2}$$

Step-up and step-down inputs are given to the system and the response curves are shown in Figure 5.25 and Figure 5.26. Bode plot is also drawn to understand stability criterion in Figure 5.27.

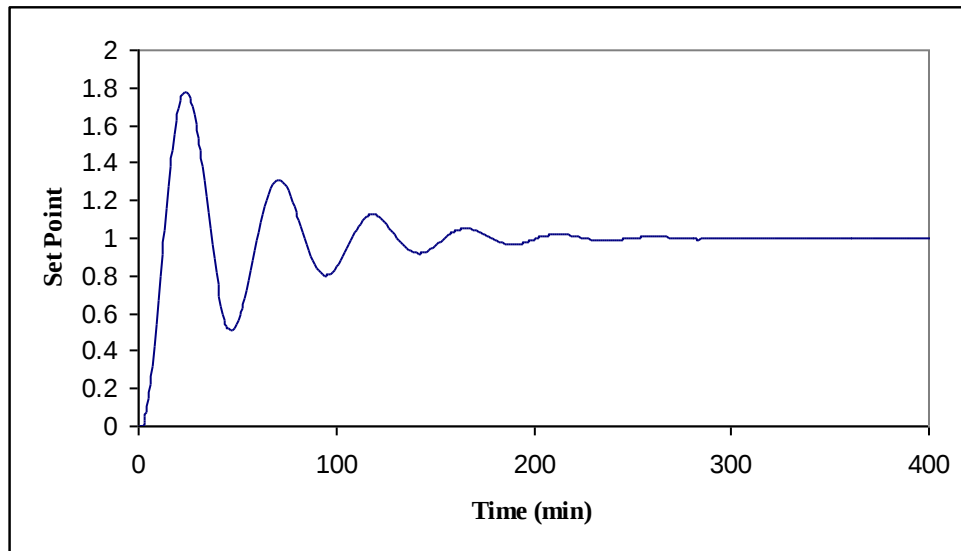


Figure 5.25 : The response of the vapor phase to the given step-up input by Ziegler Nichols Method (no overshoot)

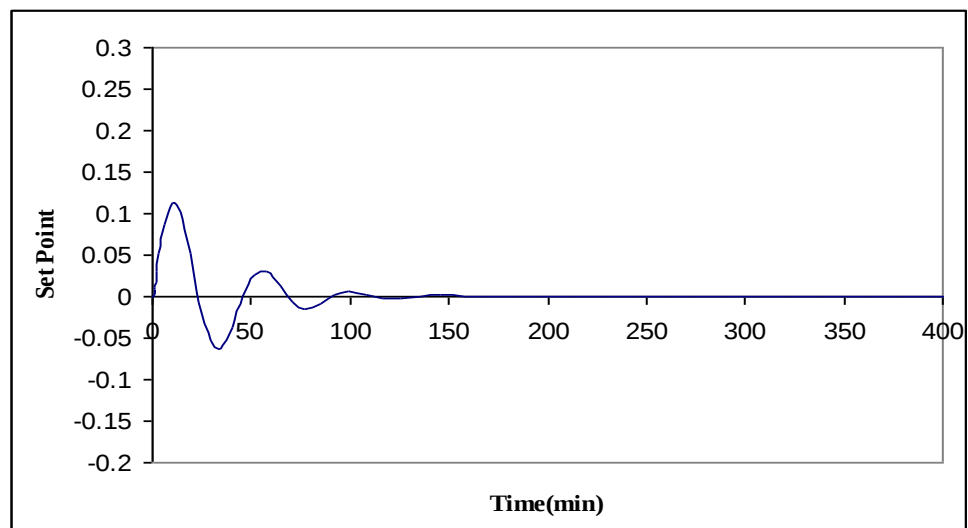


Figure 5.26 : The response of the vapor phase to the given step-down input by Ziegler Nichols Method (no overshoot)

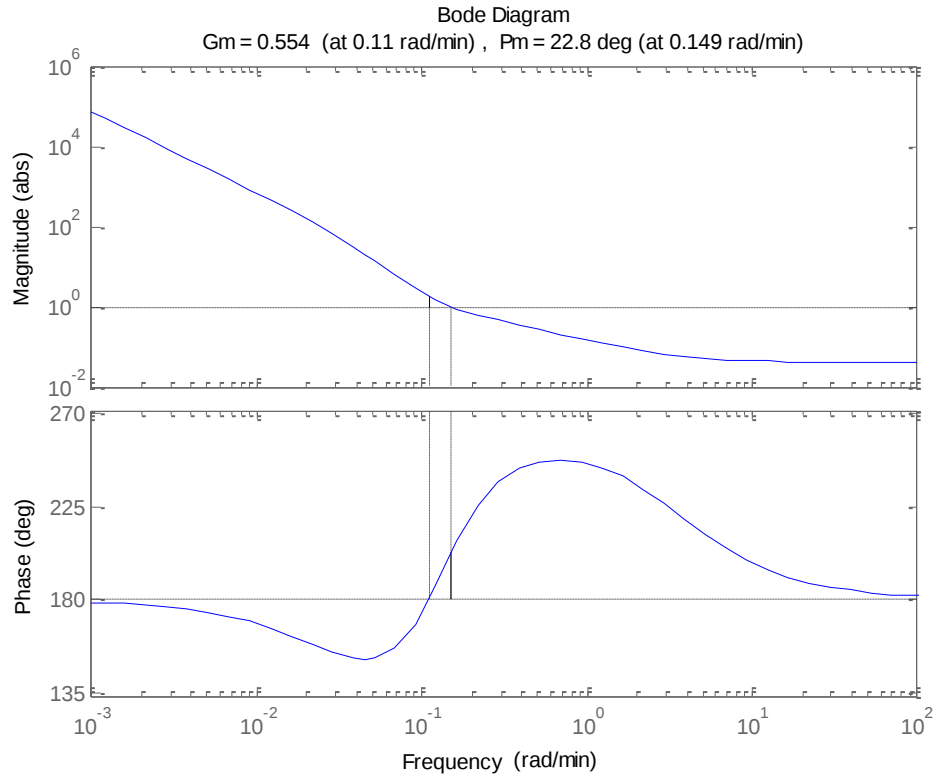


Figure 5.27 : The Bode plot of the vapor phase by Z-N method (no overshoot)

5.3.2 Disturbance Effects

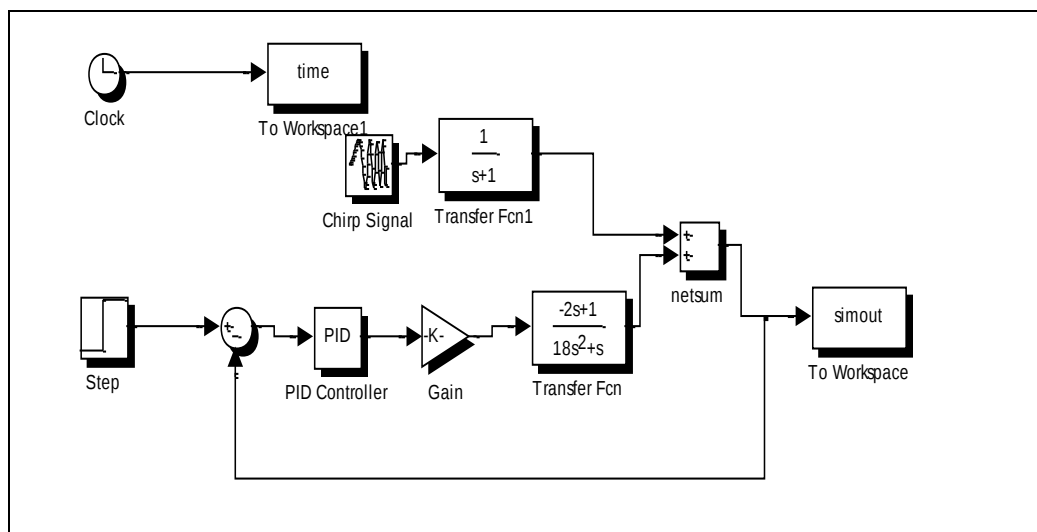


Figure 5.28 : SIMULINK Program for the disturbance on the DPIMCPID_L

By using the simulink program shown in Figure 5.28, the disturbance effects are analyzed and illustrated in Figure 5.29 and Figure 5.30.

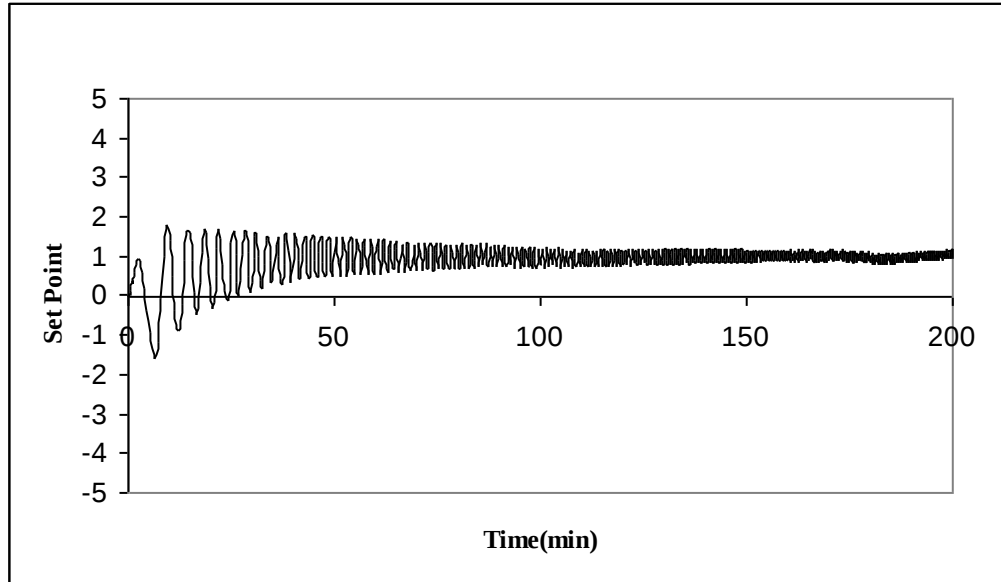


Figure 5.29 : The Effect of Disturbance (0.1 Hz. input) on the DPIMCPID_L

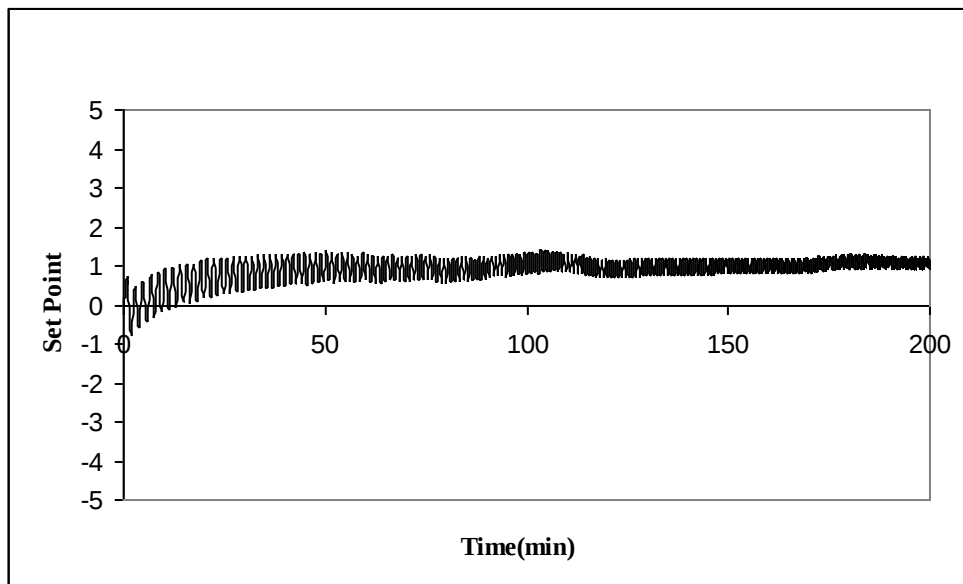


Figure 5.30 : The Effect of Disturbance (0.5 Hz. input) on the DPIMCPID_L

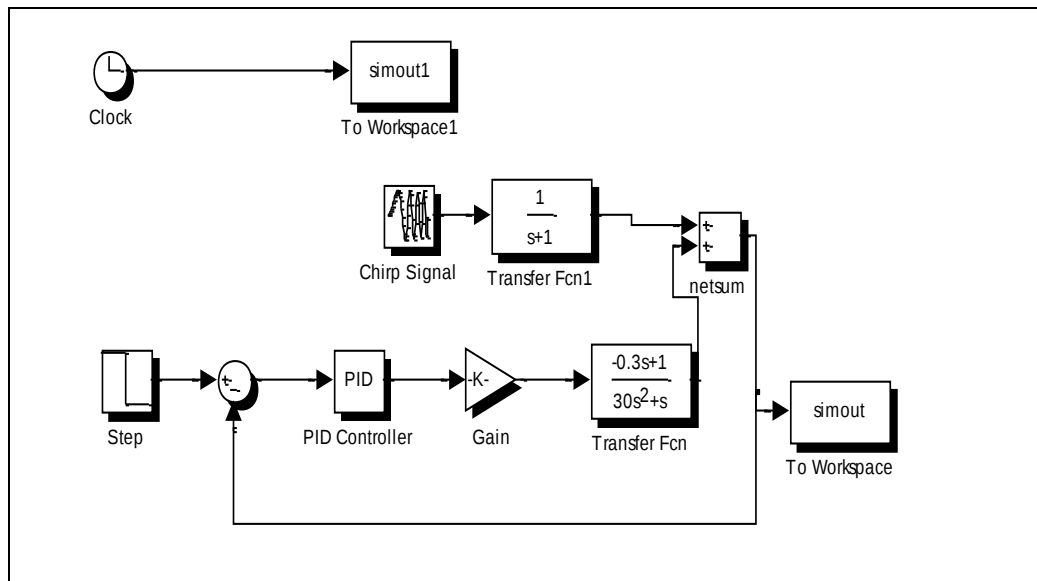


Figure 5.31 : SIMULINK Program for the disturbance on the DPIMCPID_v

By using the simulink program shown in Figure 5.31, the disturbance effects are analyzed and illustrated in Figure 5.32 and Figure 5.33.

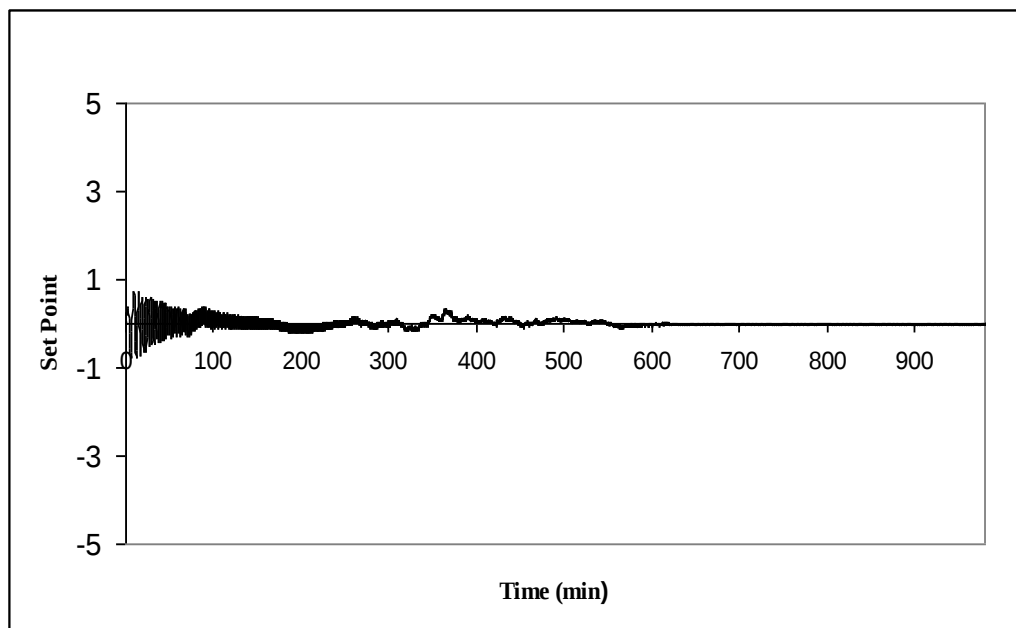


Figure 5.32 : The Effect of Disturbance (0.1 Hz. input) on the DPIMCPID_v

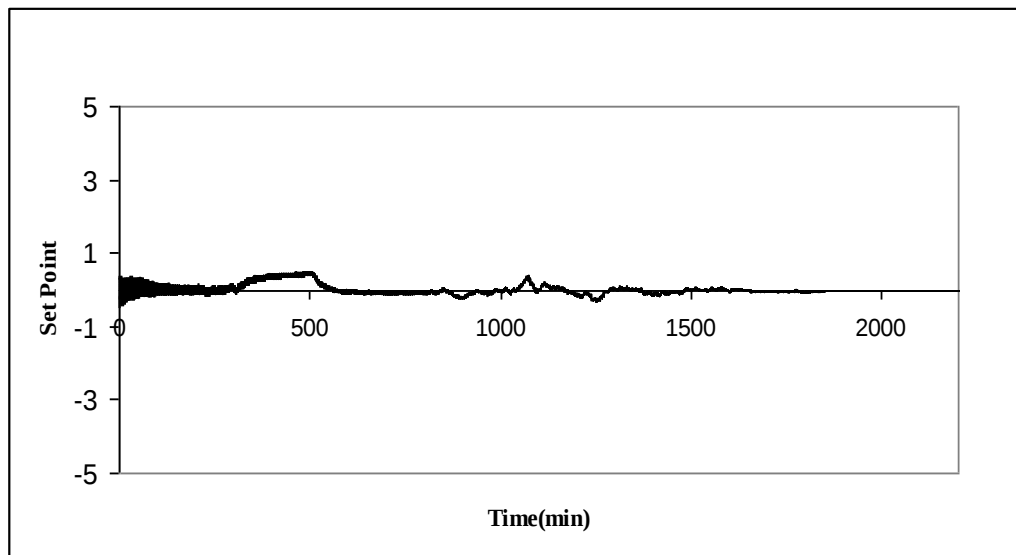


Figure 5.33 : The Effect of Disturbance (0.5 Hz. input) on the DPIMCPID_V

5.2 VEE Pro 7.0 Applications

5.2.1 Overview of VEE Pro 7.0

Agilent VEE is a graphical programming language optimized for building test and measurement applications, and programs with operator interfaces [27].

Advantages of Using Agilent VEE for Test Development

VEE offers many advantages in test development:

- Increase your productivity dramatically. Customers report reducing their program development time up to 80%.
- Use VEE in a wide range of applications including functional test, design verification, calibration, and data acquisition and control.
- Gain instrument I/O flexibility controlling GPIB, VXI, Serial, GPIO, PC Plug-in cards, and LAN instruments. Use “panel” drivers, *VXIplug&play* drivers, ODAS drivers, “direct I/O” over standard interfaces, or imported libraries from multiple vendors.

- Use ActiveX Automation and Controls on PCs to control other applications such as MS Word, Excel, and Access that assist with generating reports, displaying and analyzing data, or putting test results into a database for future use.
- Increase throughput, build larger programs with ease, and become more flexible in instrument management. VEE has a compiler; a professional development environment suited for large, complex programs; and advanced instrument management capabilities.
- Leverage your investment in textual languages such as C/C++, Visual Basic, Pascal, Fortran, and Rocky Mountain Basic [27].

5.2.2 VEE Simulation

In this study the written VEE program is shown in Figure 5.34 to Figure 5.38.

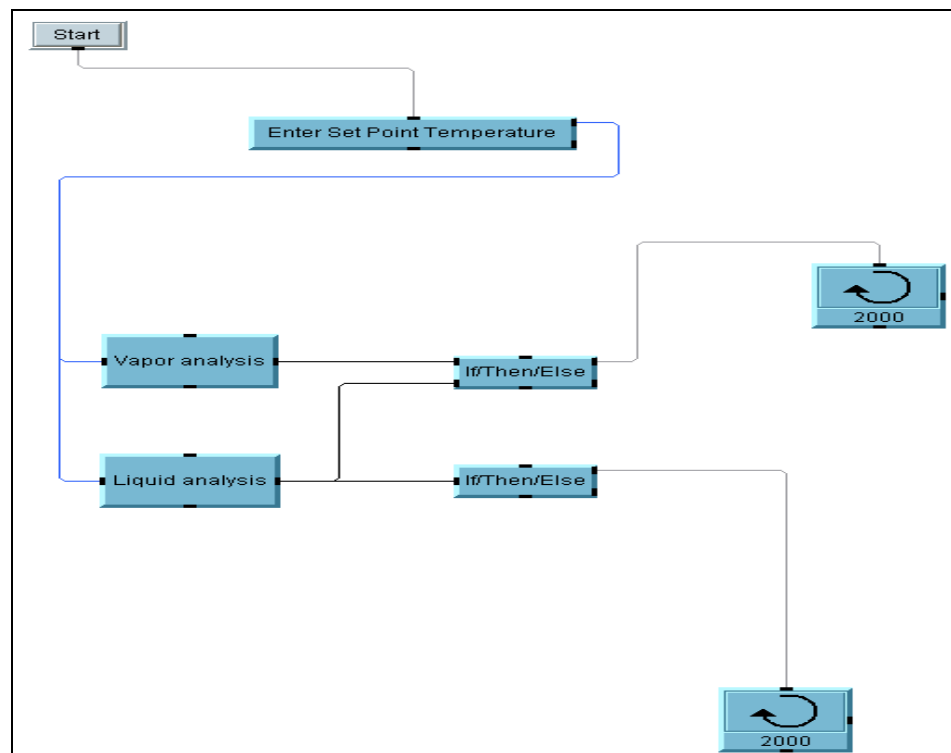


Figure 5.34 : First part of the VEE Program obtained for this study

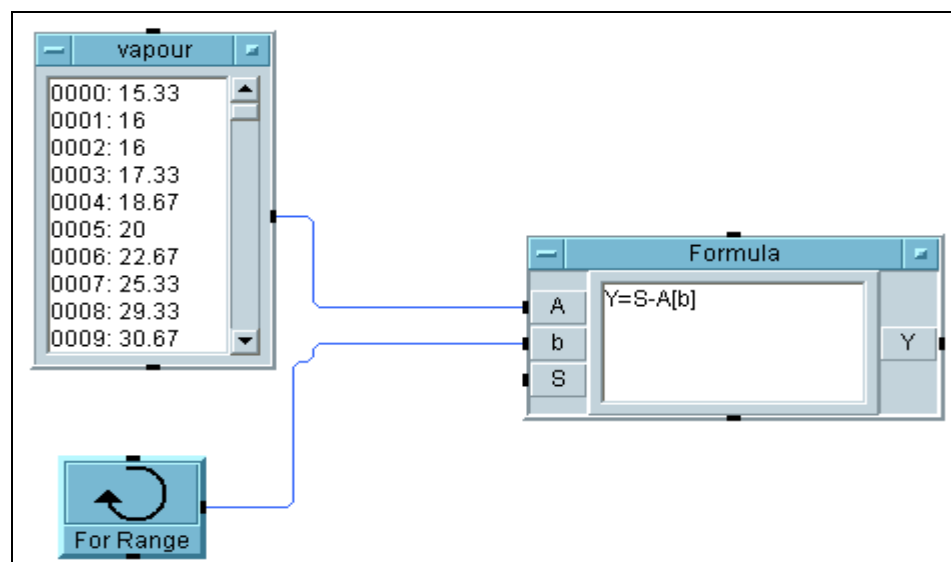


Figure 5.35 : Inside of the Vapor Analysis Block

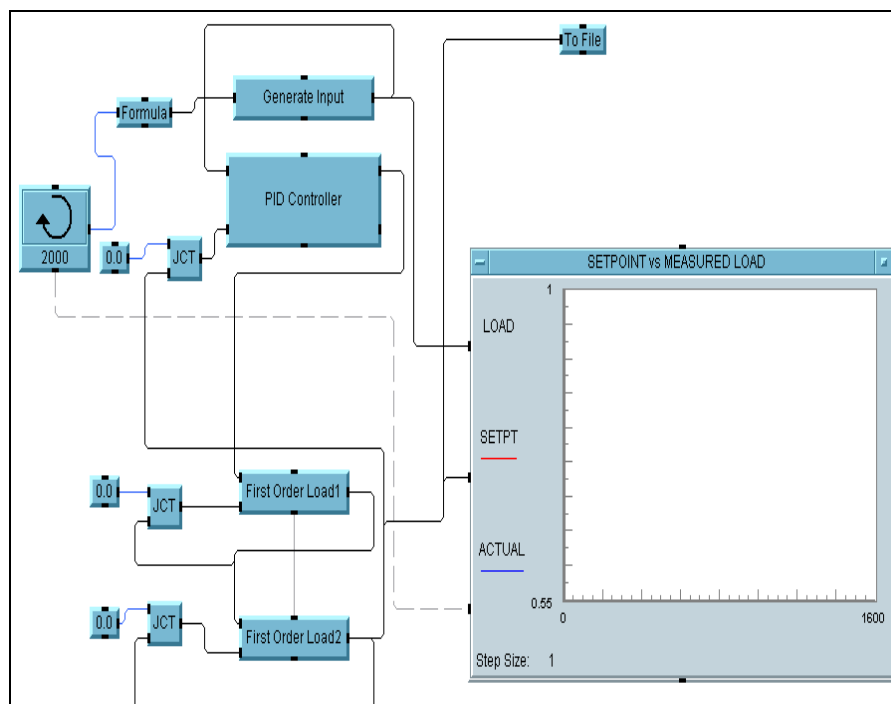


Figure 5.36 : Second part of the VEE Program obtained for this study

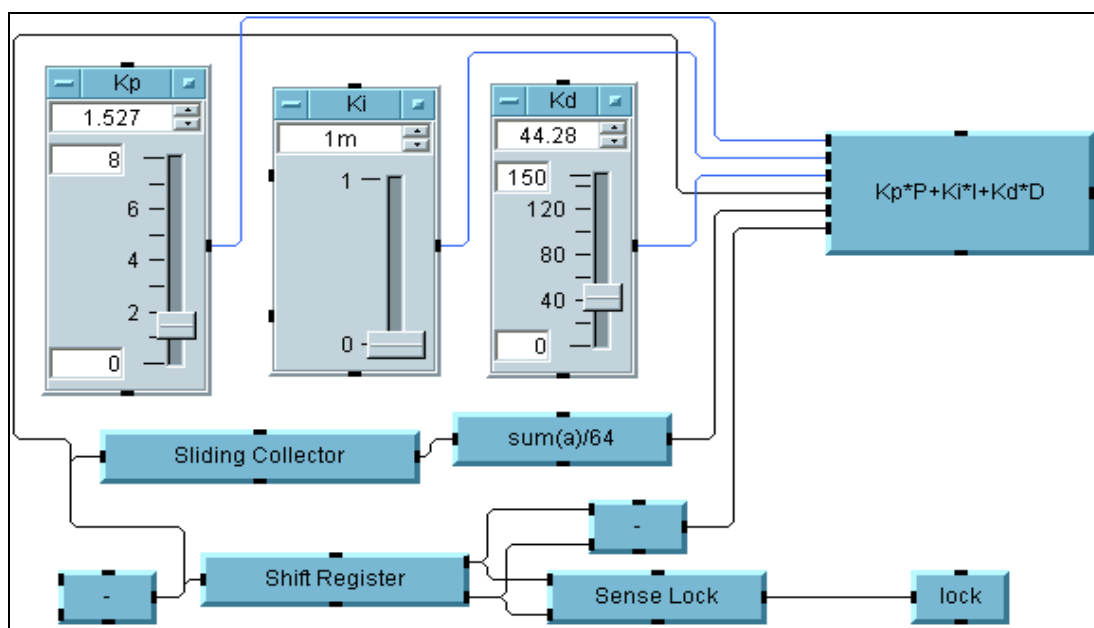


Figure 5.37 : Inside of the PID Controller Block

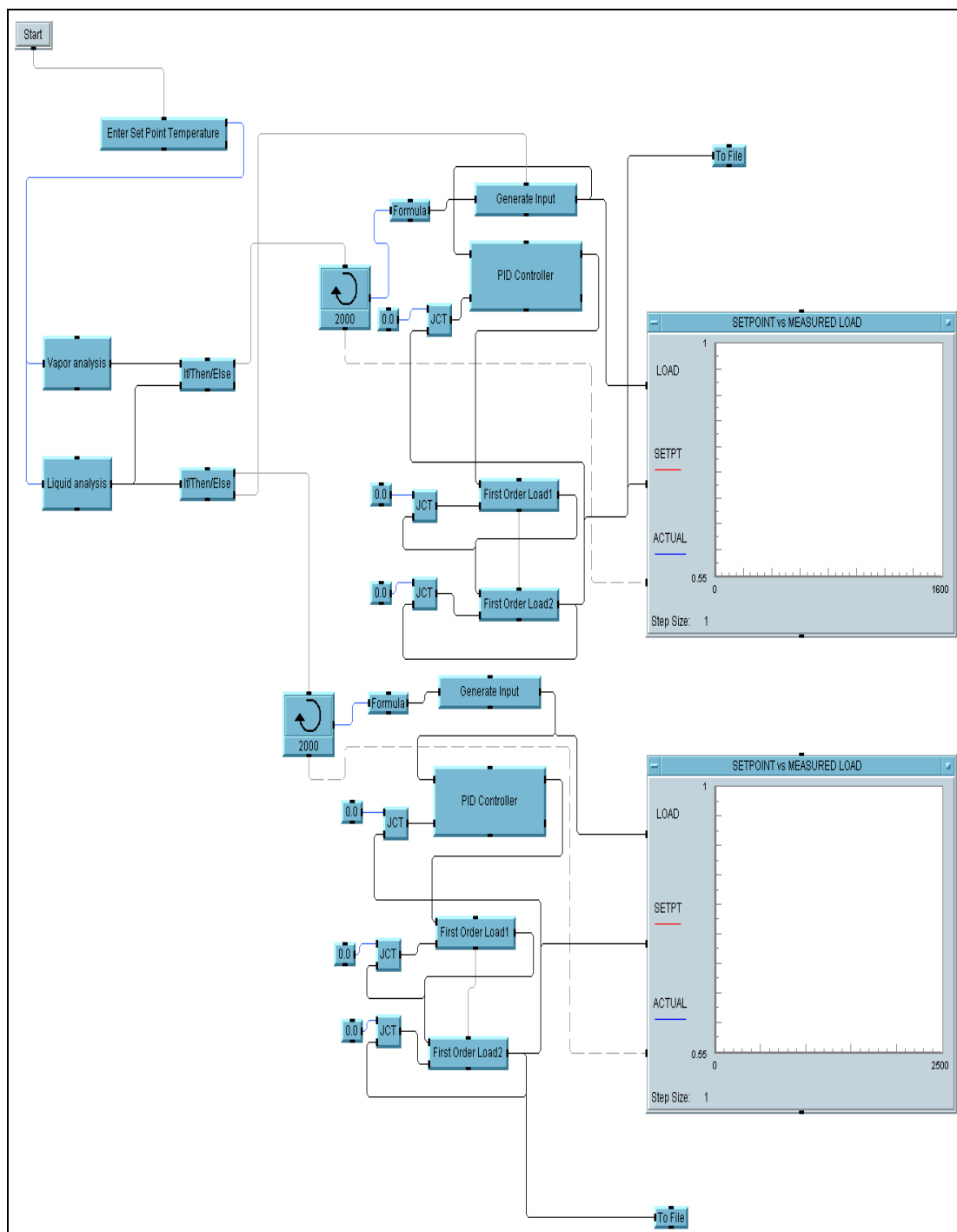


Figure 5.38 : Total view of the program written for this study

Vapor Phase Control Simulations

Step-up and step-down inputs are given to the process and shown in Figure 5.39 and Figure 5.40 respectively.

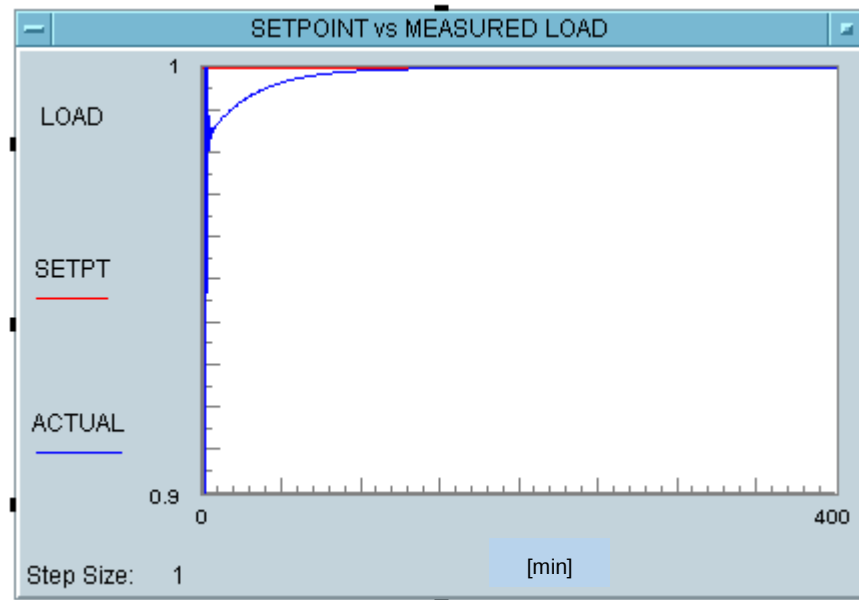


Figure 5.39 : Step-up Input Given to the Process controlled by DPIMCPID_v

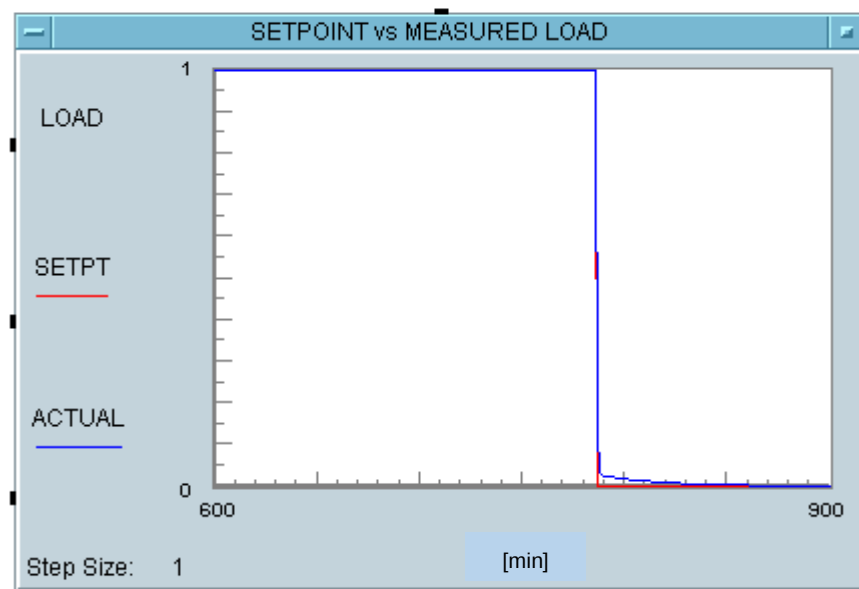


Figure 5.40 : Step-down Input Given to the Process controlled by DPIMCPID_v

Liquid Phase Control Simulations

Step-up and step-down inputs are given to the process and shown in Figure 5.41 and Figure 5.42 respectively.

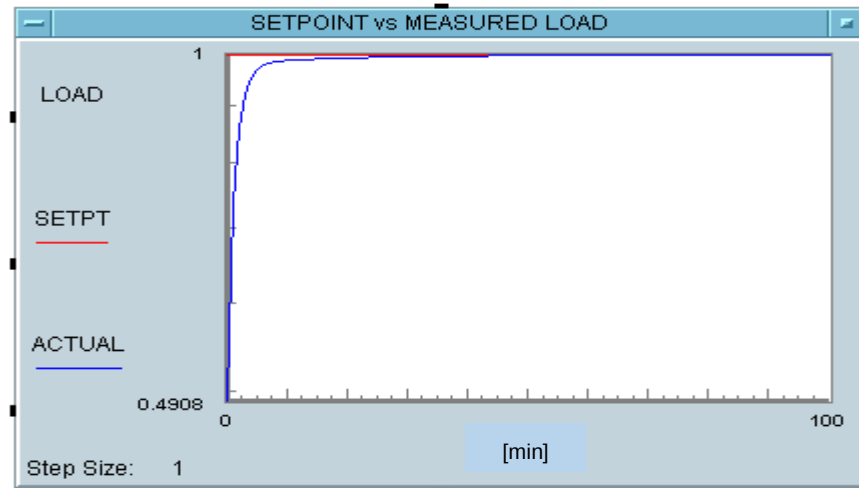


Figure 5.41 : Step-up Input Given to the Process controlled by DPIMCPID_L

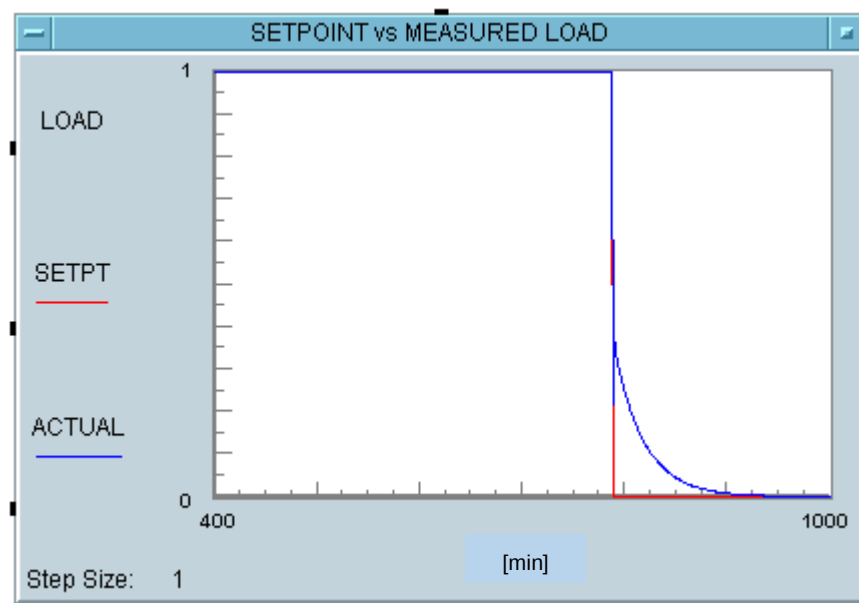


Figure 5.42 : Step-down Input Given to the Process controlled by DPIMCPID_L

The addition of function generator block which is cosine is a disturbance effect to the system shown in Figure 5.43 and Figure 5.44.

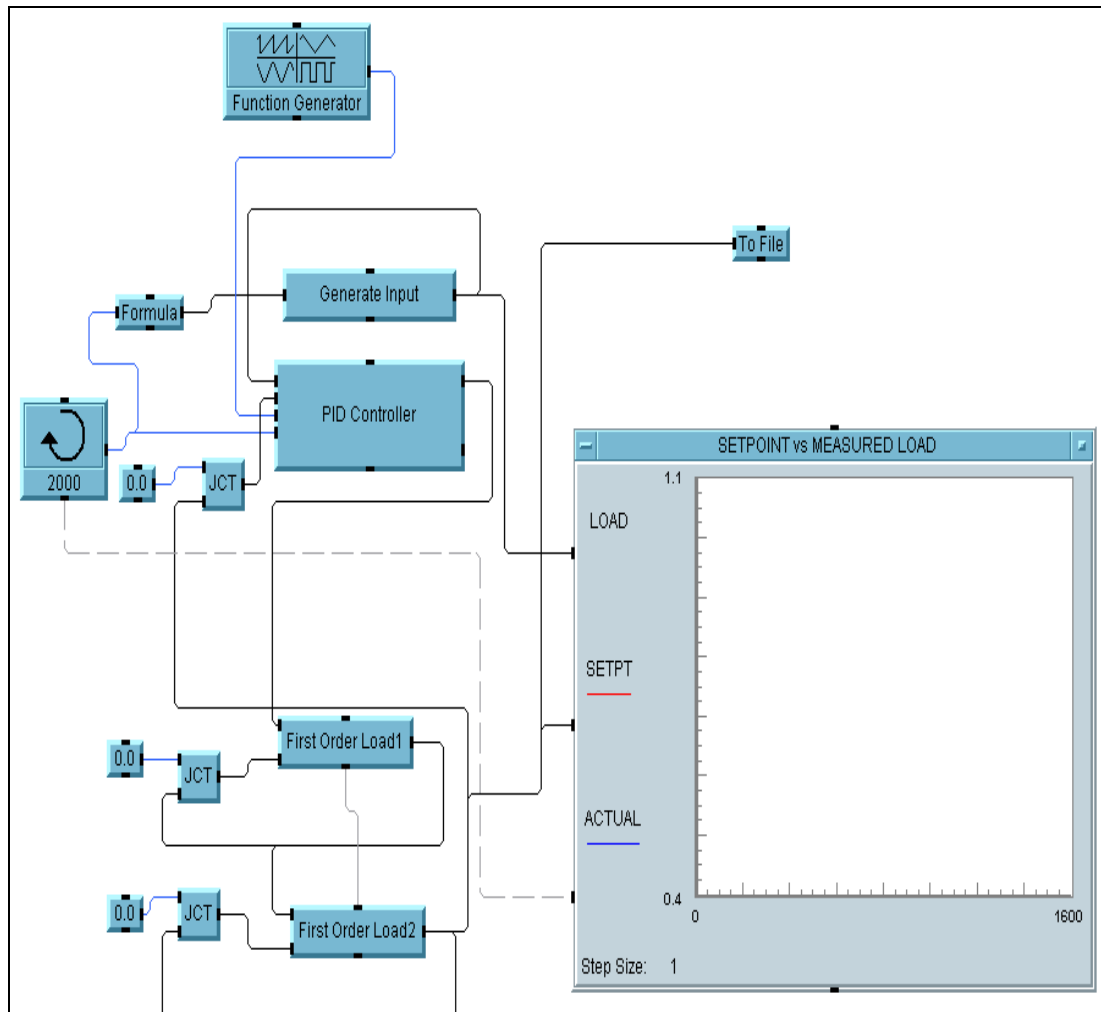


Figure 5.43 : After addition of disturbance effect (Function Generator) to the process

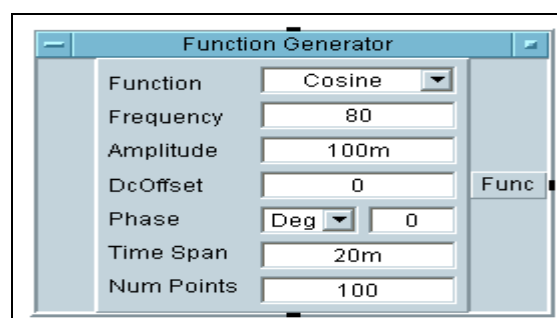


Figure 5.44 : Function Generator Block

Disturbance effects are shown in Figure 5.45 and Figure 5.46.

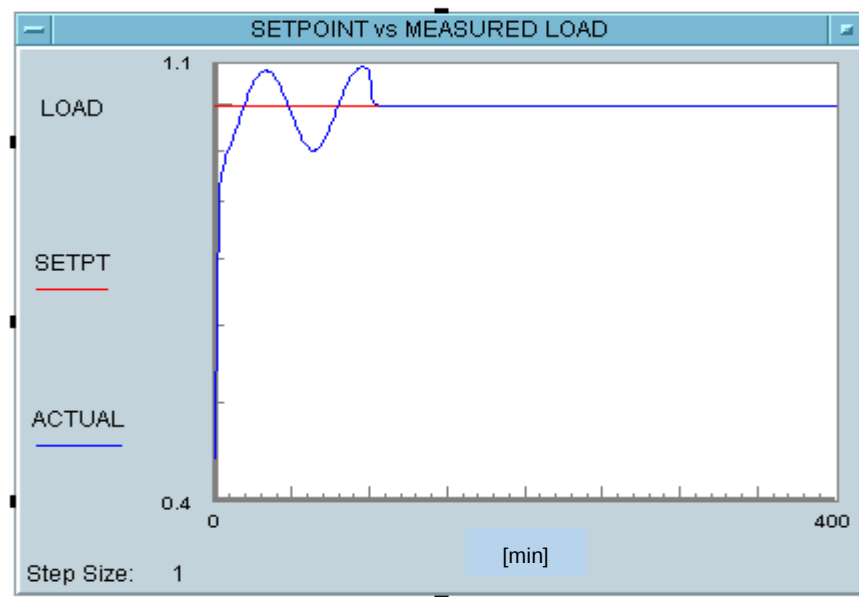


Figure 5.45 : Process Response to Disturbance by using DPIMCPID_L

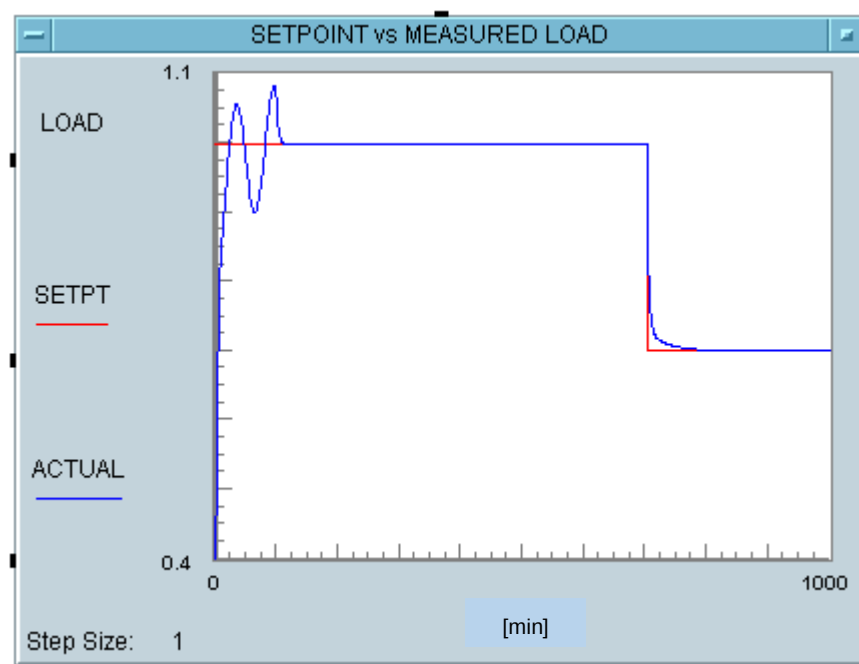


Figure 5.46 : Process Response to Disturbance by using DPIMCPID_V

6. COMPARISON OF DPIMCPID RESULTS

6.1 Performance Analysis of DPIMCPID

Time-integral performance criteria were analyzed for both DPIMCPID_V and DPIMCPID_L. To mention that, if it is desired to strongly suppress large errors, ISE is better than IAE because the errors are squared and thus contribute more to the value of the integral. For the suppression of small errors, IAE is better than ISE because when we square small numbers (smaller than one) they become even smaller. To suppress errors that persist for long times, the ITAE criterion will tune the controllers better because the presence of large t amplifies the effect of even small errors in the value of the integral [28]. The analysis of performance of the DPIMCPID is done by using Mathcad-2001 Program given in Appendix-C.

6.1.1 DPIMCPID_L Control Analysis

In this analysis IAE, ISE and ITAE that are time-integral performance criteria are performed for DPIMCPID_L control and shown in Figure 6.1, Figure 6.2 and in Figure 6.3 respectively.

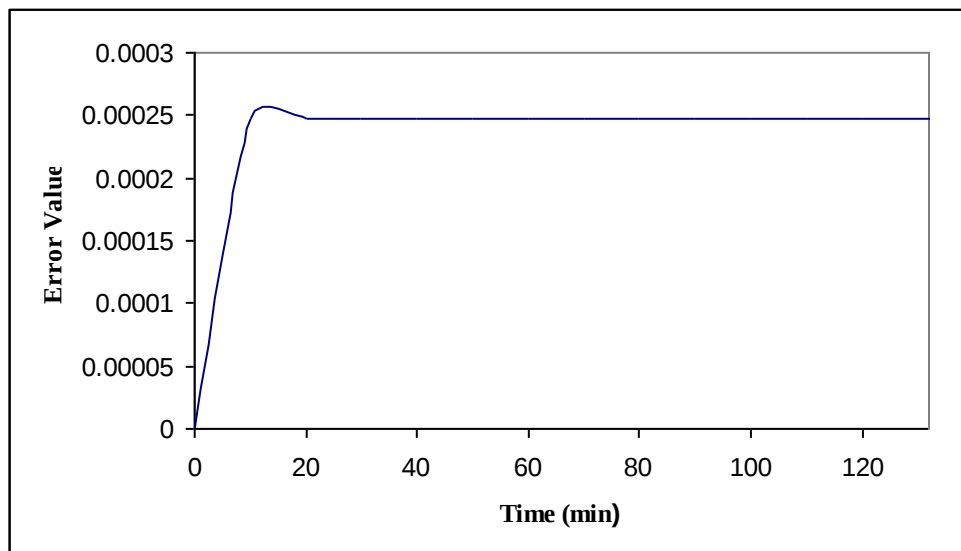


Figure 6.1 : IAE Analysis of the DPIMCPID_L

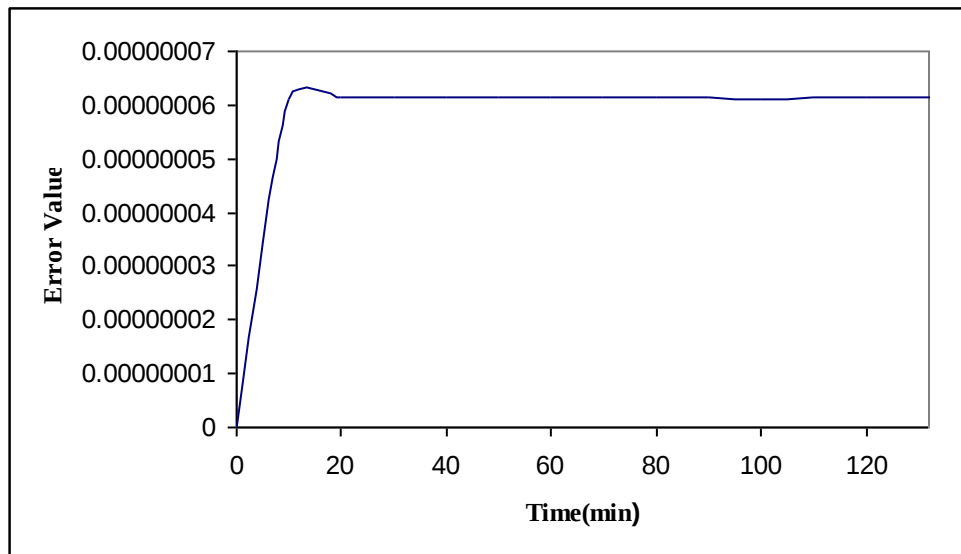


Figure 6.2 : ISE Analysis of the DPIMCPID_L

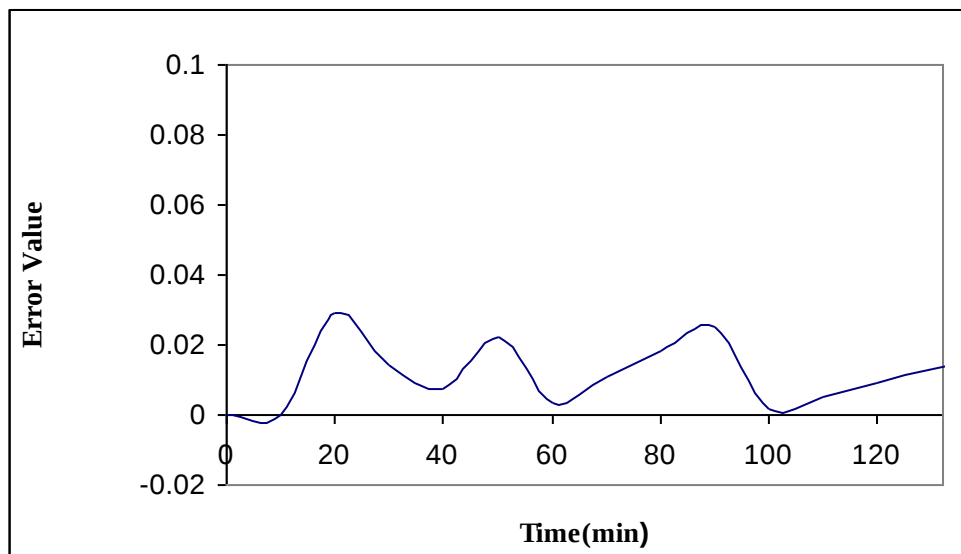


Figure 6.3 : ITAE Analysis of the DPIMCPID_L

6.1.2 DPIMCPID_v Control Analysis

In this analysis IAE, ISE and ITAE that are time-integral performance criteria are performed for DPIMCPID_v control and shown in Figure 6.4, Figure 6.5 and in Figure 6.6 respectively.

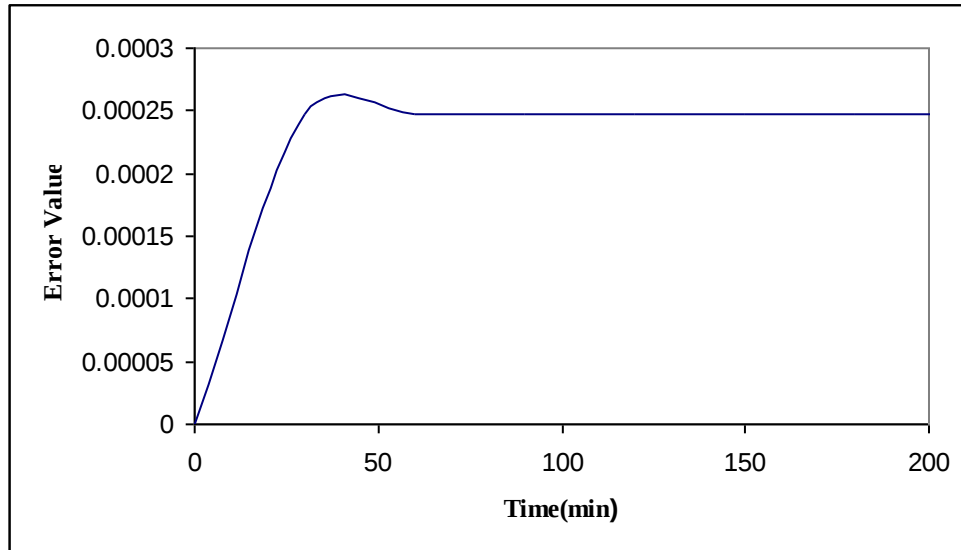


Figure 6.4 : IAE Analysis of the DPIMCPID_v

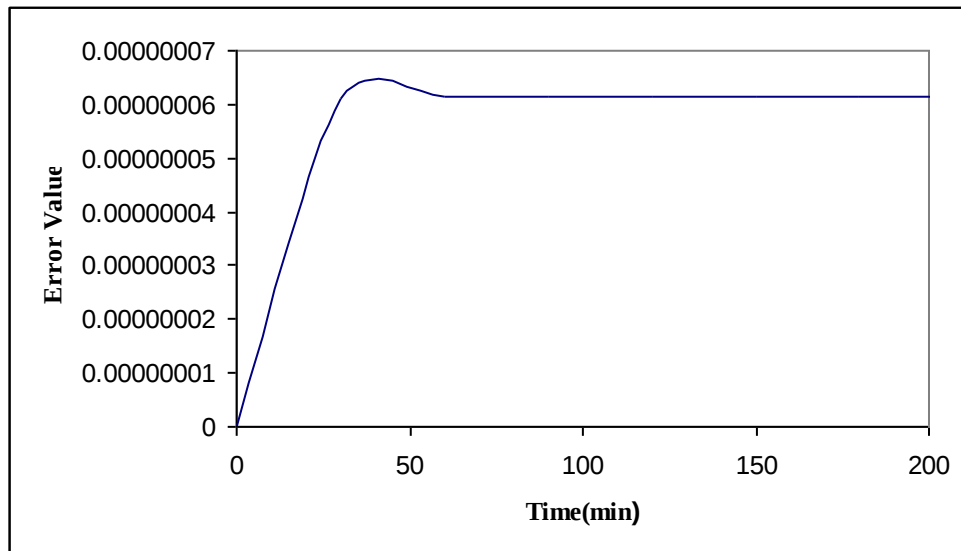


Figure 6.5 : ISE Analysis of the DPIMCPID_v

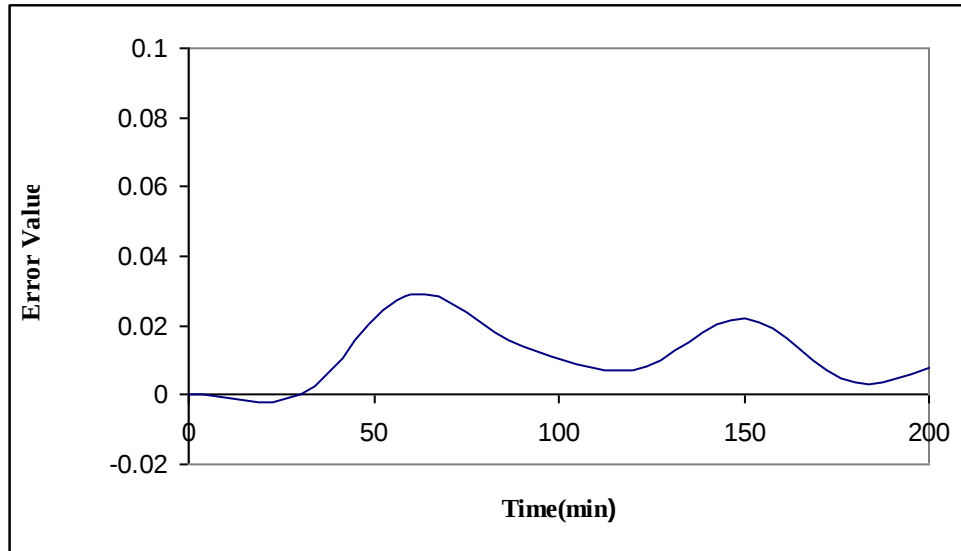


Figure 6.6 : ITAE Analysis of the DPIMCPID_v

6.2 Comparison of DPIMCPID with Classical Controllers

Results and the comparison of DPIMCPID are shown in the following tables. The calculated performance criteria are overshoot, reset time, settling time, gain margin phase margin, IAE, ISE and ITAE. To mention that overshoot values are evaluated only for underdamped systems.

Table 6.1 : Comparison of DPIMCPID by Classical Controllers (Step-up input given)

	Overshoot (%)	Rise time[min]	Settling time [min]	GM	PM
DPIMCPID _L	-	22.5	22.5	2.02	60.1
DPIMCPID _V	-	150	150	12	85
Z-N (original) Liquid Phase	78	12.5	225	6.38	19.4
Z-N (Some OS) Liquid Phase	53	22.5	250	4.35	53.9
Z-N (No OS) Liquid Phase	65	25	268	7.17	33.3
Z-N (original) Vapor Phase	80	6.7	317	0.54	14.5
Z-N(Some OS) Vapor Phase	57	10	117	0.33	47.5
Z-N(No OS) Vapor Phase	80	10	267	0.55	22.8

Table 5.2 : Comparison of DPIMCPID by Classical Controllers (Step-down input given)

	Rise time[min]	Settling time [min]	GM	PM
DPIMCPID _L	171	361	2.02	60.1
DPIMCPID _V	63	97	12	85
Z-N (original) Vapor Phase	13	400	0.54	14.5
Z-N(Some OS) Vapor Phase	20	190	0.33	47.5
Z-N(No OS) Vapor Phase	24	157	0.55	22.8

Table 6.3 : Comparison of DPIMCPID_L, DPIMCPID_V and Z-N (No OS) Controller

CONTROLLERS	Step-up input (Heating)		Step-down input (Cooling)		Time-integral Performance Analysis		
	Rise time[min]	Settling time [min]	Rise time[min]	Settling time [min]	IAE	ISE	ITAE
DPIMCPID _L	22.5	22.5	171	361	0.029	7.118×10^{-6}	1.67
DPIMCPID _V	150	150	63	97	0.029	7.120×10^{-6}	1.67
Z-N (No OS) Vapor Phase	10	267	24	157	6.648	0.402	432.5

7. CONCLUSION

In this study, the results of the reacting distillation process simulations done by using Matlab 6.5 and VEE Pro 7.0 are discussed according to well known performance criteria. Therefore, better temperature control of the reacting distillation process is achieved in this study by using the DPIMCPID controller.

7.1 Discussion and Conclusion

After performing several types of controllers, the simulation results given in Table 6.1 and Table 6.2 indicate the comparisons of DPIMCPID with classical controllers. In Table 6.1, overshoots are not calculated for DPIMCPID_L and DPIMCPID_V since the system responses of these controllers are overdamped. To mention that, the gain margin values of Z-N controllers are below than 1. When gain margin is smaller than 1 then the system is unstable which is not desired. From Table 6.3, it is obviously seen that DPIMCPID_L is better when heating-up in progress, since the settling time is much more lower than DPIMCPID_V and Z-N results. On the other hand, DPIMCPID_V is better when the system gets cooling-down because of its lower settling time. Although the rise time of Z-N (No OS) is smaller than others, the system can not be stable due to its high settling time.

In the previous studies, the model transfer function of the controller was designed for only one phase (liquid). By applying dual phase algorithm better results are achieved. For instance, Tanrıverdi (1996) designed the system for one phase and the controller's settling time was at about 38 minutes. In this study the settling time is found as 22.5 minutes which is better than previous study. Beside that, the transfer function is also different compared to Tanrıverdi (1996), in which he obtained a first order process transfer function. In this study, the transfer function is designed as first order plus time delay process with process integration.

According to the simulation results of the DPIMCPID controller, vapor phase of the reacting distillation process can be controlled more precisely, and the temperature of the head of the distillation column could be kept at a desired set point. This allows better separation of fusel oil having complexity due to its azeotropic property. Therefore, using DPIMCPID controller in a reacting distillation process gives better results than other PID controllers designed for only one phase.

7.2 Suggestions to Future Work

For the future work it is strongly recommended that feedforward long range predictive controller types of control mechanisms must be established. A good estimator such as summing neural net would be very useful. Iskender (1998) stated that it would be an advantage if recursive identification is used for estimating parameters of non-linear models [29].

In most of the chemical processes, the main parts having trouble for the control systems are the disturbances because of the chemical reactions and physical interactions such as heat and mass. These disturbances occur during the running process very spontaneously. It is not very easy to define them all and to have enough knowledge about them. That is why the modeling stage is the most important part, so perfect modeling is needed. Self tuning controllers with a good estimator can give us this opportunity.

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APPENDIX-A

$$K_p=0.182 \quad \tau_p=18 \quad \theta = 2 \quad \tau_f=1$$

$$K_c=1.799 \quad \tau_i=1018 \quad \tau_D=17.68$$

Transfer function:

$$\frac{-1.179 \cdot 10^4 s^3 + 5227 s^2 + 332.7 s + 0.3274}{18324 s^3 + 1018 s^2}$$

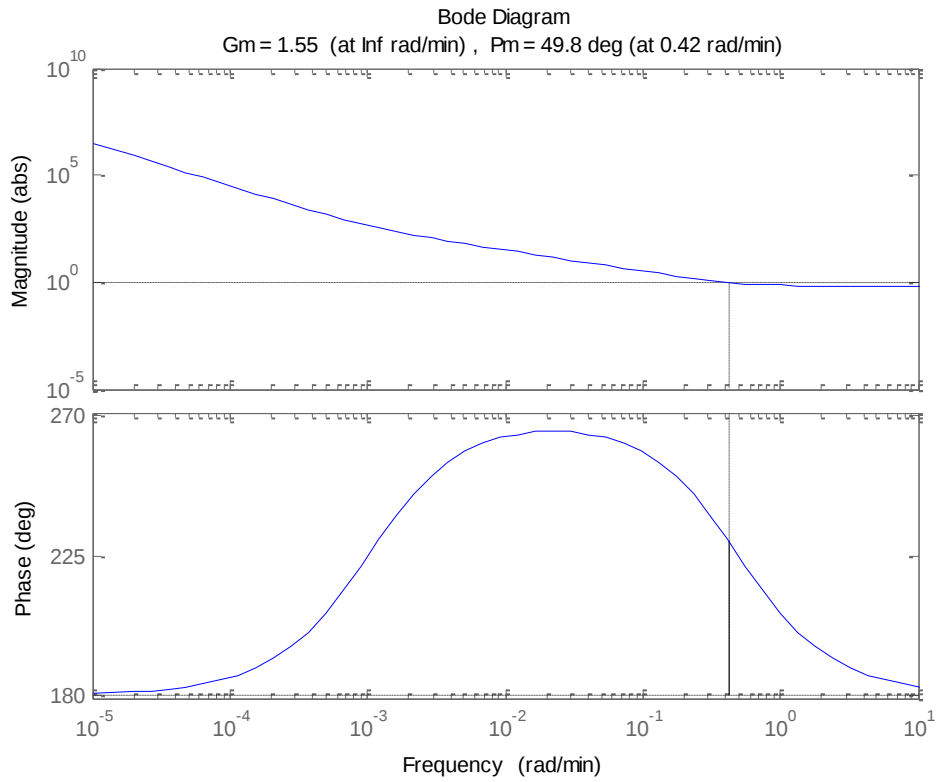


Figure A.1 : Bode Plot of DPIMCPID_L with $\tau_f=1$ filter time constant

$$K_p=0.182 \quad \tau_p=18 \quad \theta = 2 \quad \tau_f=0.85$$

$$K_c=1.894 \quad \tau_I = 1018 \quad \tau_D = 17.68$$

Transfer function:

$$\frac{-1.241 \cdot 10^4 s^3 + 5503 s^2 + 350.2 s + 0.3447}{18324 s^3 + 1018 s^2}$$

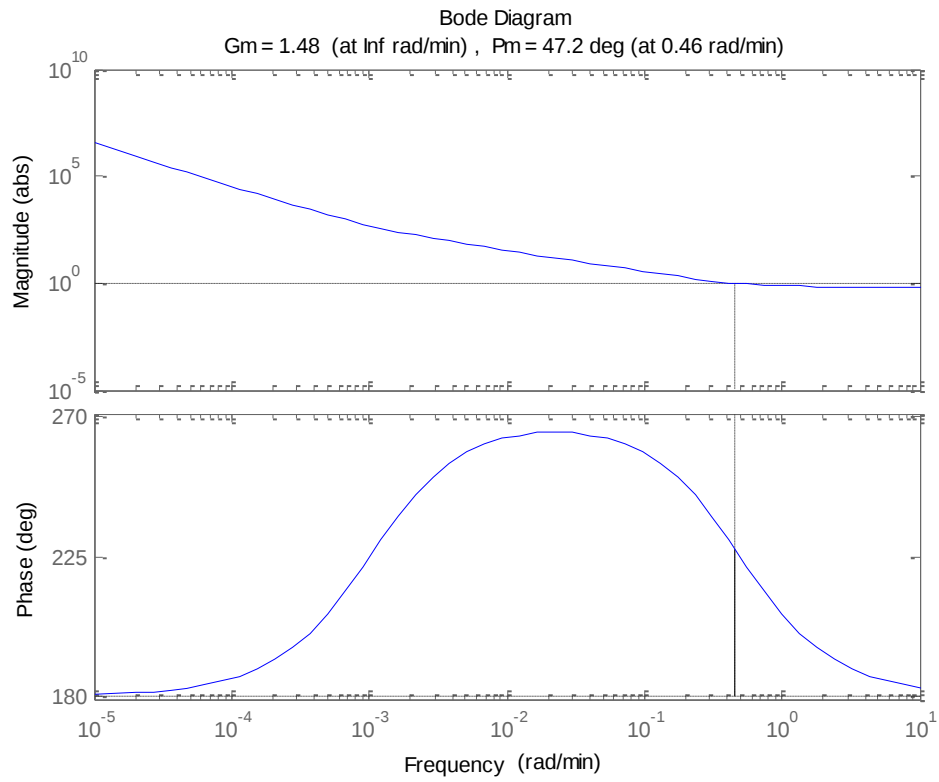


Figure A.2 : Bode Plot of DPIMCPID_L with $\tau_f = 0.85$ filter time constant

$$K_p=0.182 \quad \tau_p=18 \quad \theta = 2 \quad \tau_f=0.67$$

$$K_c=2.02 \quad \tau_I = 1018 \quad \tau_D = 17.68$$

Transfer function:

$$\frac{-1.324 \cdot 10^4 s^3 + 5869 s^2 + 373.5 s + 0.3676}{18324 s^3 + 1018 s^2}$$

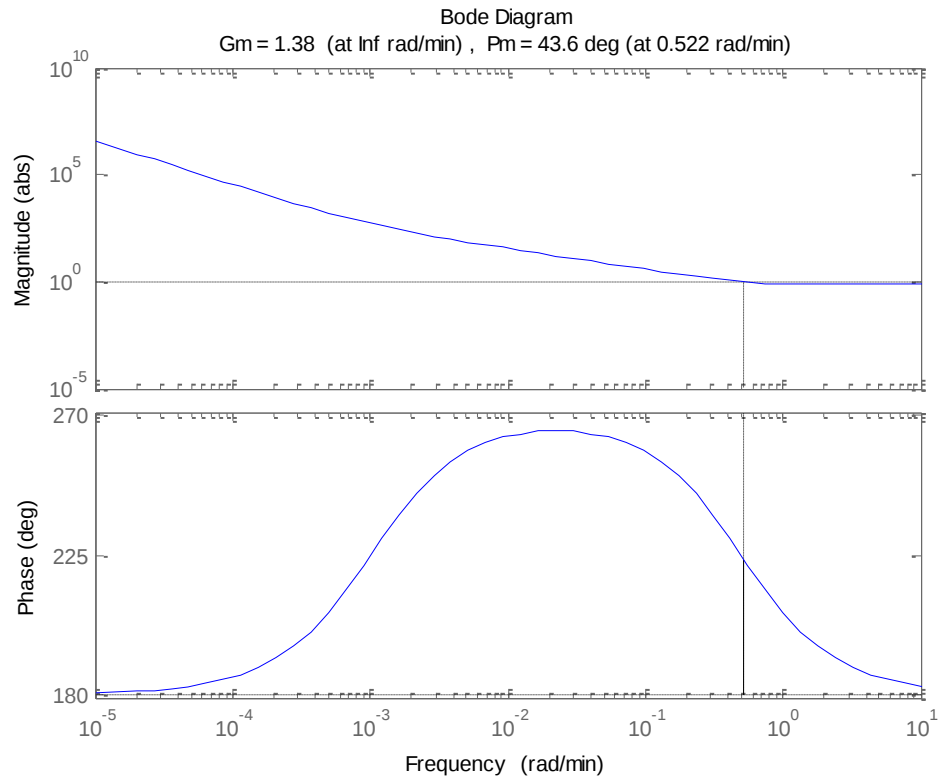


Figure A.3 : Bode Plot of DPIMCPID_L with $\tau_f=0.67$ filter time constant

$$K_p=0.182 \quad \tau_p=18 \quad \theta = 2 \quad \tau_f=0.9$$

$$K_c=1.86 \quad \tau_I=1018 \quad \tau_D=17.68$$

Transfer function:

$$\frac{-1.219 \cdot 10^4 s^3 + 5404 s^2 + 343.9 s + 0.3385}{18324 s^3 + 1018 s^2}$$

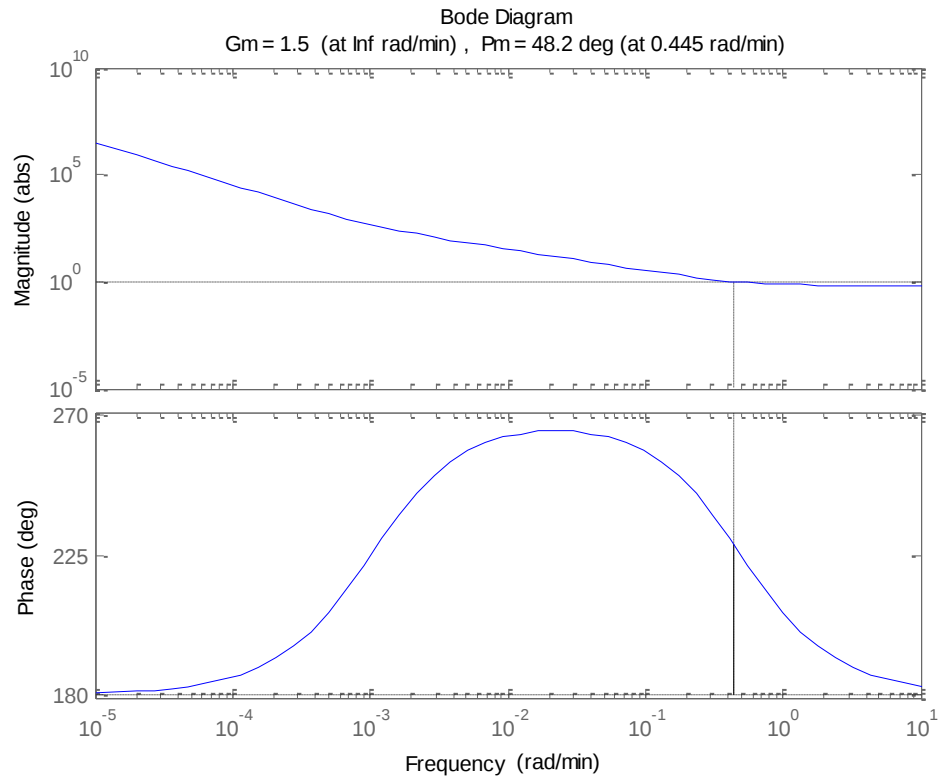


Figure A.4 : Bode Plot of DPIMCPID_L with $\tau_f=0.9$ filter time constant

$$K_p = 0.187 \quad \tau_p = 30 \quad \theta = 0.3 \quad \tau_f = 0.2$$

$$K_c = 10.383 \quad \tau_I = 1030 \quad \tau_D = 29.13$$

Transfer function:

$$\frac{-1.747 \cdot 10^4 s^3 + 57650 s^2 + 1999 s + 1.942}{30900 s^3 + 1030 s^2}$$

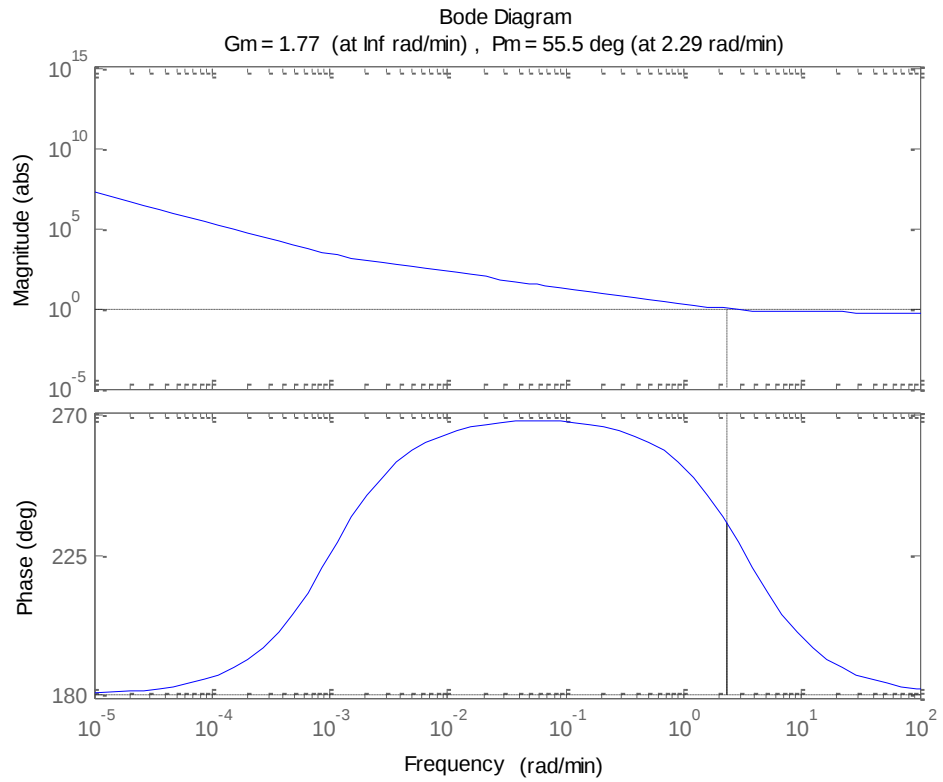


Figure A.5 : Bode Plot of DPIMCPID_V with $\tau_f = 0.2$ filter time constant

$$K_p = 0.187 \quad \tau_p = 30 \quad \theta = 0.3 \quad \tau_f = 0.5$$

$$K_c = 6.489 \quad \tau_I = 1030 \quad \tau_D = 29.13$$

Transfer function:

$$\frac{-1.092 \cdot 10^4 s^3 + 36030 s^2 + 1249 s + 1.213}{30900 s^3 + 1030 s^2}$$

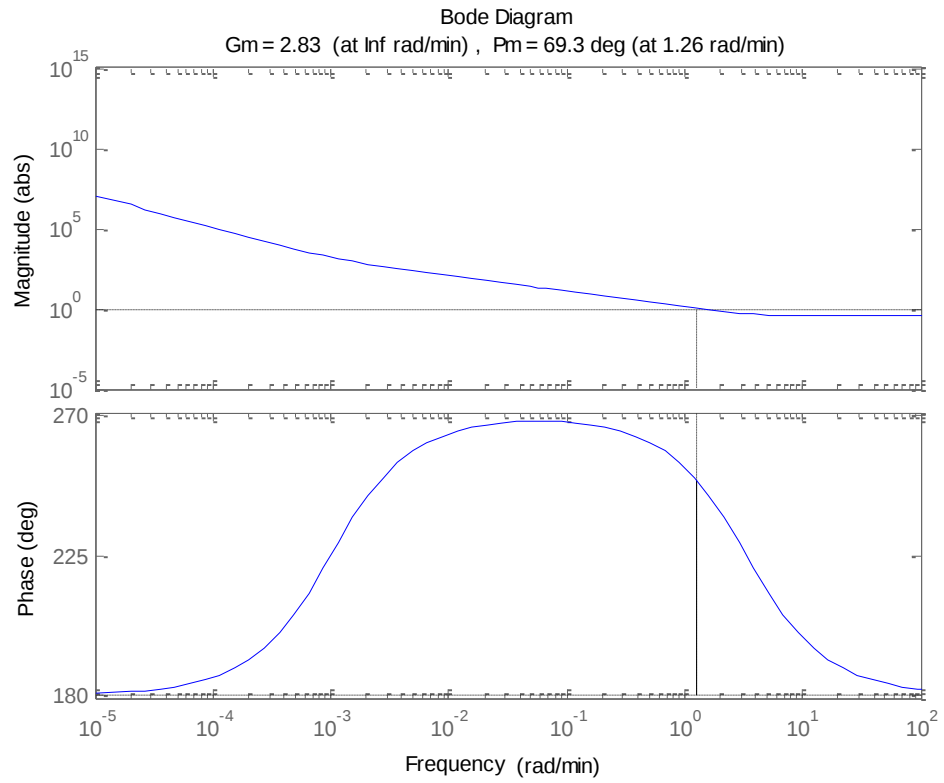


Figure A.6 : Bode Plot of DPIMCPID_V with $\tau_f = 0.5$ filter time constant

$$K_p = 0.187 \quad \tau_p = 30 \quad \theta = 0.3 \quad \tau_f = 0.67$$

$$K_c = 5.352 \quad \tau_i = 1030 \quad \tau_D = 29.13$$

Transfer function:

$$\frac{-9007s^3 + 2.971 \cdot 10^4 s^2 + 1031s + 1.001}{30900s^3 + 1030s^2}$$

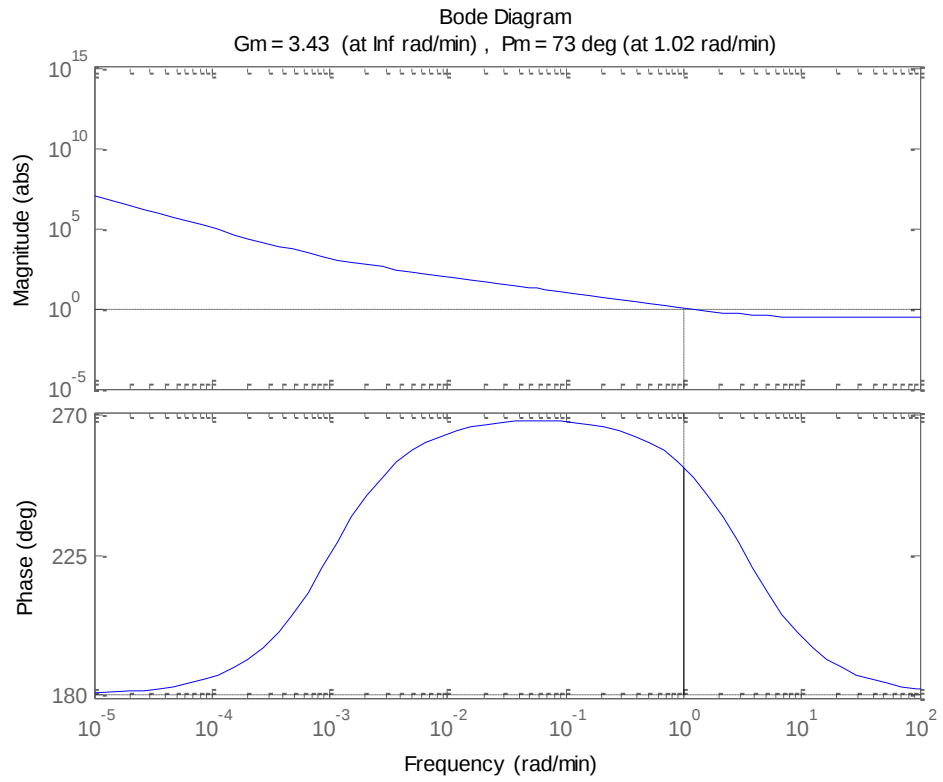


Figure A.7 : Bode Plot of DPIMCPID_V with $\tau_f = 0.67$ filter time constant

$$K_p = 0.187 \quad \tau_p = 30 \quad \theta = 0.3 \quad \tau_f = 0.9$$

$$K_c = 4.326 \quad \tau_i = 1030 \quad \tau_D = 29.13$$

Transfer function:

$$\frac{-7280s^3 + 2.402 \cdot 10^4 s^2 + 833s + 0.809}{30900s^3 + 1030s^2}$$

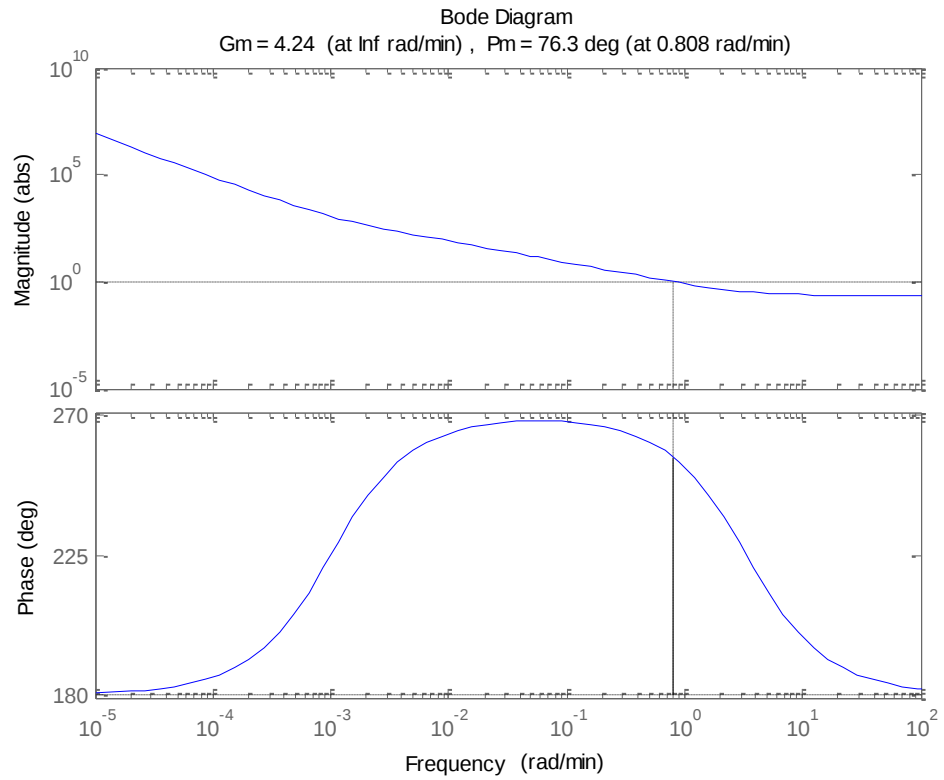


Figure A.8 : Bode Plot of DPIMCPID_V with $\tau_f = 0.9$ filter time constant

APPENDIX-B

An Application of IMC with Pole Placement on a Different Process

$K_p=0.25$ $\tau_p=0.4$ [sec] $\theta = 0.3$ [sec] $\tau_f=0.67$

$K_c= 4.122$ $\tau_I= 1000.4$ $\tau_D= 0.39$

Transfer function:

$$\frac{-123.4s^3 + 102.1s^2 + 1031s + 1.031}{400.2s^3 + 1000s^2}$$

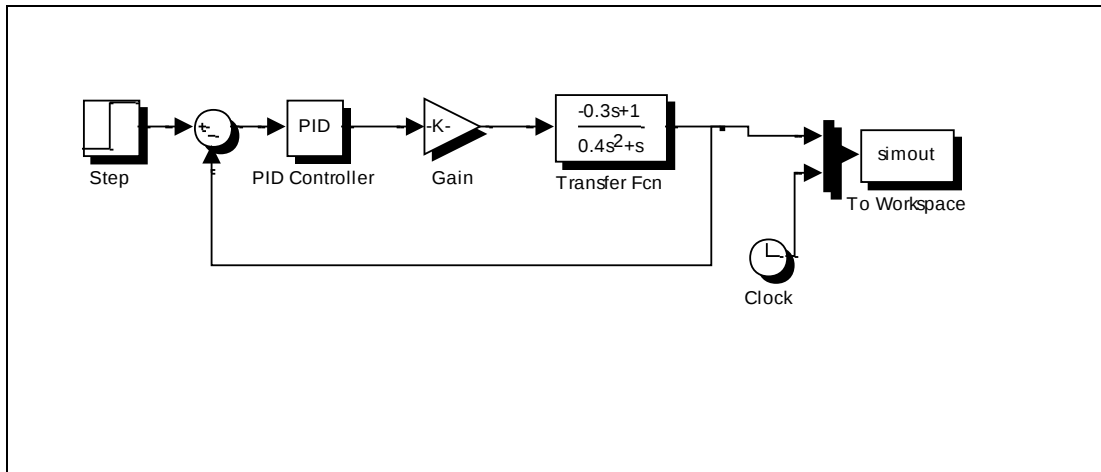


Figure B.1 : SIMULINK Model for the Process

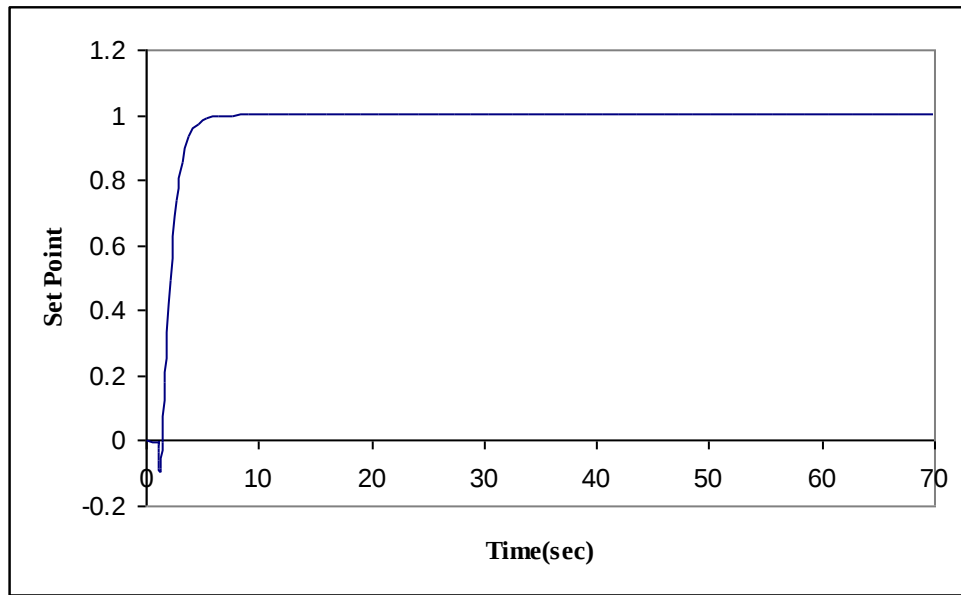


Figure B.2 : Step Response of the Process

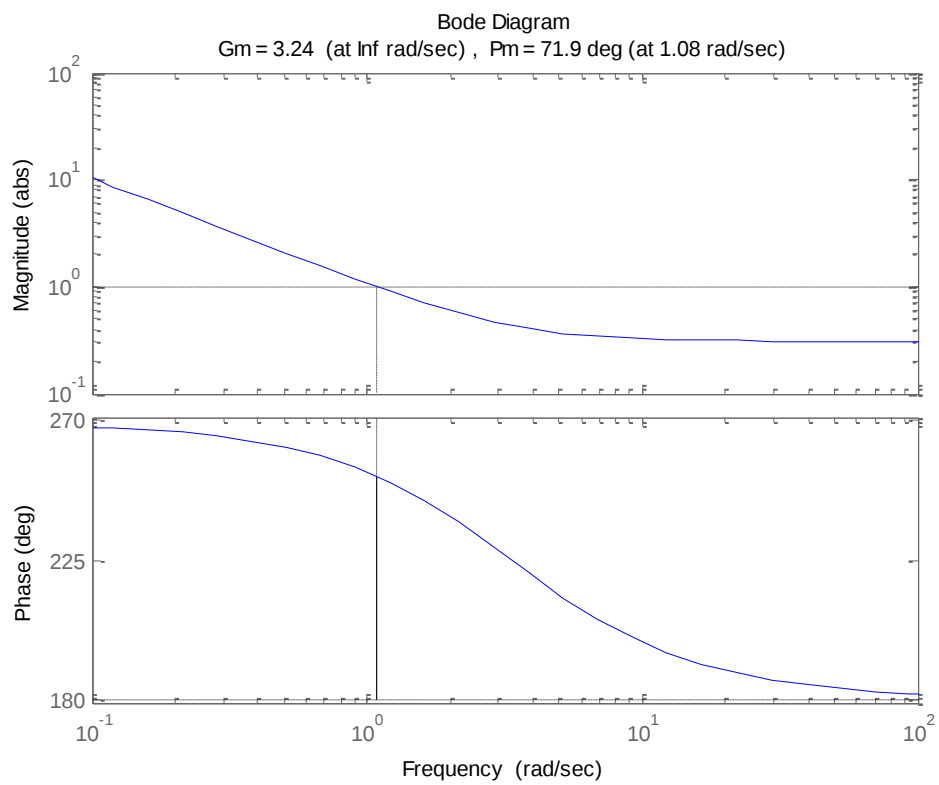


Figure B.3 : Bode Plot of the Process Response

APPENDIX-C

Liquid Phase Performance Analysis By Mathcad Program

$$\frac{(-9068t^3 + 4021t^2 + 255.9t + 0.2519)}{18324t^3 + 1018t^2}$$

Liquid phase control transfer function

$$-.495 \cdot \text{Dirac}(t) + 2.478 \cdot 10^{-4} \cdot t + .246 + 1.055 \cdot 10^{-5} \cdot \exp\left[-(5.56 \cdot 10^2) \cdot t\right]$$

$$b := 2 \quad p := 18 \quad H := 0.182$$

$$a := 0 \quad c := 116$$

$$E(t) := -0.495 + 2.478 \cdot 10^{-4} \cdot t + .246 + 1.055 \cdot 10^{-5} \cdot \exp\left[-(5.56 \cdot 10^2) \cdot t\right]$$

$$\int_a^c |E(t) - E(t-1)| dt = 0.029$$

IAE

$$F := \begin{cases} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \begin{cases} k \leftarrow k + 1 \\ v_k \leftarrow |E(t) - E(t-1)| \end{cases} dt \\ v \end{cases}$$

$$\text{ITAE} \int_a^c t \left| E(t) - E(t-1) \right| dt = 1.667$$

$$w := \left| \begin{array}{l} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \left| \begin{array}{l} k \leftarrow k + 1 \\ v_k \leftarrow t \left| E(t) - E(t-1) \right| \end{array} \right. dt \\ v \end{array} \right.$$

$$\text{ISE} \int_a^c (E(t) - E(t-1))^2 dt = 7.118 \times 10^{-6}$$

$$y := \left| \begin{array}{l} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \left| \begin{array}{l} k \leftarrow k + 1 \\ v_k \leftarrow (E(t) - E(t-1))^2 \end{array} \right. dt \end{array} \right.$$

Vapor Phase Performance Analysis By Mathcad Program

$$\frac{(-2570s^3 + 8478s^2 + 294s + 0.25)}{3090s^3 + 103s^2} \quad | \quad \text{Vapor phase control transfer function}$$

$$-(7.44 \cdot 10^{-2}) \cdot \text{Dirac}(t) + 2.479 \cdot 10^{-4} \cdot t + .247 + 1.55 \cdot 10^{-5} \cdot \exp[-(3.33 \cdot 10^{-2}) \cdot t]$$

$$b := 0.3 \quad p := 30 \quad l := 0.187$$

$$a := 0 \quad c := 116$$

$$H(t) := -(7.44 \cdot 10^{-2}) + 2.479 \cdot 10^{-4} \cdot t + .247 + 1.55 \cdot 10^{-5} \cdot \exp[-(3.33 \cdot 10^{-2}) \cdot t]$$

$$\int_a^c |H(t) - H(t-1)| dt = 0.029$$

IAE

$$F := \left| \begin{array}{l} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \left| \begin{array}{l} k \leftarrow k + 1 \\ v_k \leftarrow |H(t) - H(t-1)| \end{array} \right. dt \\ v \end{array} \right|$$

ITAE

$$\int_a^c t \left| H(t) - H(t-1) \right| dt = 1.667$$

$$w := \left| \begin{array}{l} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \left| \begin{array}{l} k \leftarrow k + 1 \\ v_k \leftarrow t \left| H(t) - H(t-1) \right| \end{array} \right. dt \\ v \end{array} \right.$$

ISE

$$\int_a^c (H(t) - H(t-1))^2 dt = 7.121 \times 10^{-6}$$

$$y := \left| \begin{array}{l} v_0 \leftarrow 0 \\ k \leftarrow 0 \\ \int_a^c \left| \begin{array}{l} k \leftarrow k + 1 \\ v_k \leftarrow (H(t) - H(t-1))^2 \end{array} \right. dt \\ v \end{array} \right.$$

Ziegler Nichols Method Controller Performance Analysis By Mathcad Program

$$\frac{-11.85^3 + 37.6^2 + 6.708 + 0.6657}{282.9^3 + 9.43^2}$$

$$-(4.189 \cdot 10^{-2}) \cdot \text{Dirac}(t) + 7.059 \cdot 10^{-2} \cdot t - 1.43 + 1.547 \cdot \exp[-(3.33 \cdot 10^{-2}) \cdot t]$$

$$b := 0.3 \quad p := 30$$

$$a := 0 \quad c := 116$$

$$H(t) := -(4.189 \cdot 10^{-2}) + 7.059 \cdot 10^{-2} \cdot t - 1.43 + 1.547 \cdot \exp[-(3.33 \cdot 10^{-2}) \cdot t]$$

$$\int_a^c |H(t) - H(t-1)| \, dt = 6.648 \quad \text{---■ IAE}$$

$$\int_a^c t |H(t) - H(t-1)| \, dt = 432.517 \quad \text{---■ ITAE}$$

$$\int_a^c (H(t) - H(t-1))^2 \, dt = 0.402 \quad \text{----■ ISE}$$

CURRICULUM VITAE

Ebru CEBECİ was born in Kayseri in 8 June 1981. She was graduated from Ahmet Paşa Primary School in 1992 and Nuh Mehmet Baldöktü Anatolian High School in 1999. In the same year she passed the university exam to be enrolled as a student to the department of Chemical Engineering at Middle East Technical University; following her graduation in June 2003, she entered to the Postgraduate Programme in Science and Technology Institute of Istanbul Technical University, in the department of Chemical Engineering.