

**SLOVAK UNIVERSITY OF TECHNOLOGY  
IN BRATISLAVA**

**FACULTY OF CHEMICAL AND FOOD TECHNOLOGY**

Reg. No.: FCHPT-

# **Modelling and Control of a Distillation Column**

**BACHELOR THESIS**

**2024**

**Bc. Kadir Ercan Özdemir**



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**BACHELOR THESIS**

Study programme: Process Control  
Study field: Cybernetics  
Training workspace: Institute of Information Engineering, Automation, and Mathematics  
Thesis supervisor: prof. Ing. Miroslav Fikar, DrSc.

**2024**

**Bc. Kadir Ercan Özdemir**



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# Abstract

This thesis focuses on modelling the distillation column of methanol-water mixture in MATLAB environment and implementing PID controllers using Simulink. Controlled variables are distillate molar fraction and bottom molar fraction. The best MV and CV couples are selected using Relative Gain Array. It turns out that manipulated variables should be the distillate flow rate for distillate composition and bottom flow rate for bottom composition. According to this, open-loop schemes are constructed for tuning purposes separately for distillate molar fraction and bottom molar fraction. It is notable to say that the system does not exhibit FOPDT response. Four different tuning methods, of which MATLAB tuner and other tuning method are chosen to be the best ones for both distillate molar fraction and bottom molar fraction according to a specific purpose, are compared. Lastly, the model with two PID controllers interacting simultaneously is simulated. It is important to note that they do not cause too much deviations when simulated.

Keywords: Process control, Distillation column, Methanol-water mixture, PID tuning



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# Chapter 1

## Introduction

The process of distillation involves applying or removing heat in order to separate a liquid or vapour mixture of two or more substances into its component fractions of desired purity using different volatility of substances in the mixture and liquid-vapour contact.

The operation of distillation column first involves feeding the mixture into the distillation column and then heating it in the reboiler located at the bottom of the distillation column. After that, the vapour formed by reboiler ascends through column and goes to condenser located at the top of the column where the vapour is liquified. Products are the bottom product from reboiler and distillate from condenser.

Drawbacks of distillation are that energy is consumed in vast amounts and that operating cost is too much. One way to eliminate the drawbacks aforementioned is to implement control strategies according to plant's needs.

However, implementing control strategies are often faced with nonlinearities in the system. Control configurations such as LV, LB, DV and LD which combines input and output variables are used for effective operation. The control loops for composition may be located at the trays, reboilers, condensers and feed part. Predictive control has been so popular lately but there is still reliance on conventional PID control loops because it is used more widely than predictive control all around the world [10].

The aim of this work is to model the distillation column in MATLAB and control

it with a PID control. Controlled variables are distillate molar fraction and bottom molar fraction. Several PID controllers are presented and compared.

The work is divided into two sections namely, theoretical and practical. The theoretical section begins by outlining the steps involved in distillation. Furthermore, the description of PID controllers as well as binary mixtures are covered in this section, with particular attention paid to the methanol and water mixture that was the subject. The second part is practical which is dedicated to calculations and results for PID tuning methods and comparing them.

# Chapter 2

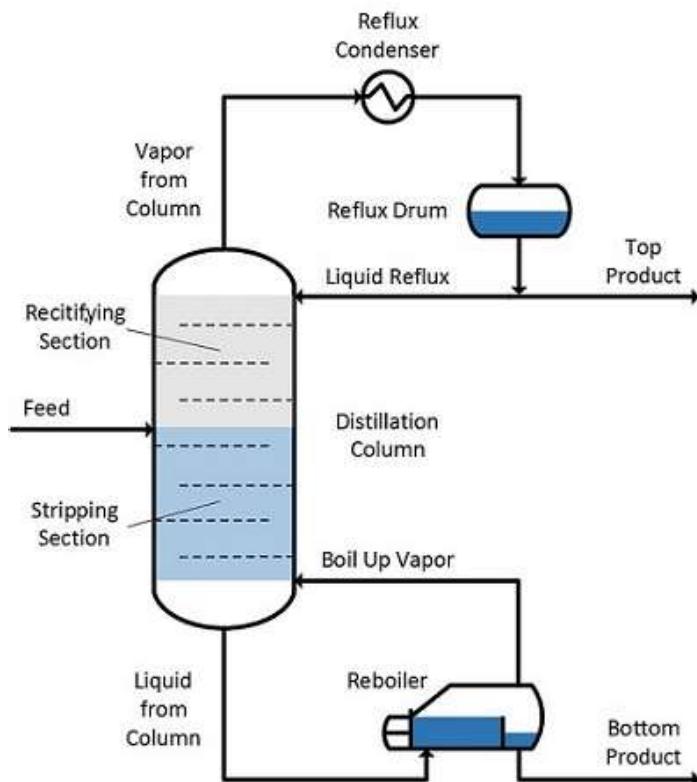
## Theoretical Part

In this part, distillation column basics, its modelling and PID in theory are introduced.

### 2.1 Distillation Column Basics

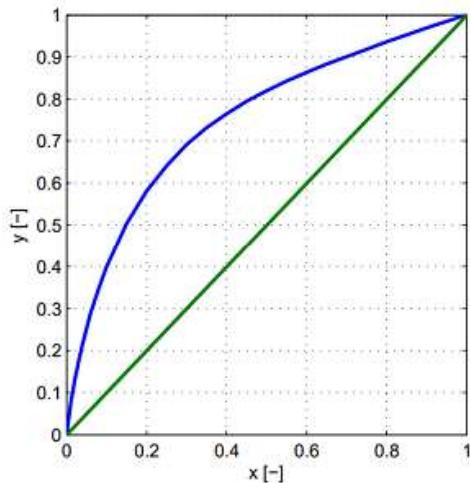
The separation of two or more mixtures according to their different volatilities is the main idea behind the distillation column. The substance with a lower boiling point is thought to be the more volatile component. A characteristic of distillation is the phase transition from liquid to vapor. A more volatile component is added to the vapour phase once the liquid phase has evaporated in the boiler located at the bottom. This results in the liquid and vapor phases having distinct compositions. Condensed vapour from top of the column often recycles back to the distillation column to enhance the contact between liquid and vapour [14, 10, 12].

Typically, a distillation column consists of a tower or column tray. Compared to other types, tray types are more efficient. Trays above the feed tray are called enriching section and trays below the feed tray are called stripping section. The most volatile component is separated from enriching section (condenser) and the least volatile component is separated from stripping section. Simple distillation column is depicted in Figure 2.1.

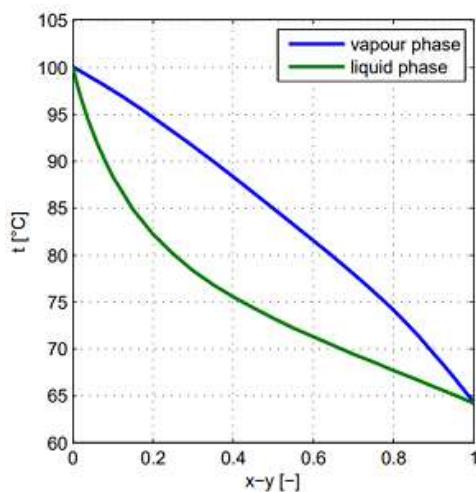


**Figure 2.1:** Distillation column [8].

In this work, methanol-water mixture is considered for separation in distillation column since it is convenient to use them in laboratory conditions. Methanol has a lower boiling point than water ( $64.7^{\circ}\text{C}$ ), making it more volatile component than water. Methanol and water are examples of two components that can be separated using a binary distillation column as they have no azeotrope point. Since vapour and liquid phases are present in the distillation column, there is a plot relating how many fractions of vapour or liquid phase of one of the component. Figure 2.2 shows x-y plot for methanol-water mixture at 1 atm pressure for  $x$  being methanol mole fraction in liquid phase and  $y$  being methanol mole fraction in vapour phase. Also, there is a T-x-y graph of methanol-water mixture shown in Figure 2.3. When the temperature is set, methanol mole fractions in the mixture can be found. With T-x-y plot, it is possible to find temperature of each tray with the help of liquid fraction [2, 20].



**Figure 2.2:** Plot between vapour phase and liquid phase for methanol [20].



**Figure 2.3:** Plot between vapour phase and liquid phase for methanol in different temperatures [20].

## 2.2 Distillation Column Modelling

In order to model the distillation column, some assumptions must be made [16, 20]. But first, variables and their abbreviations are introduced. These parameters are going to be used in MATLAB (See Section 3.1).

- $k$  is the feed tray number.
- $N_i$  is holdup in  $i$ -th tray.
- $N_b$  is reboiler holdup.
- $N_t$  is the tray and condenser holdup.
- $eff_T$  is the efficiency of the trays.
- $eff_B$  is the efficiency of the reboiler.
- $x_F$  is the molar fraction of volatile component in feed.
- $F$  is the molar flow of feed.
- $D$  is the molar flow of distillate.
- $V$  is the molar flow of vapour.
- $B$  is the molar flow of bottom.
- $x_i$  is the molar fraction of methanol in  $i$ -th tray in liquid phase.
- $(x_i)^*$  is the ideal molar fraction of methanol in  $i$ -th tray in liquid phase.
- $y_i$  is the molar fraction of methanol in  $i$ -th tray in vapour phase.
- $(y_i)^*$  is the ideal molar fraction of methanol in  $i$ -th tray in vapour phase.

Assumptions are:

- Stages are mixed ideally.
- Each plate which contains liquid and vapour leaving is in equilibrium.
- There is no vapour holdup on trays.
- There is constant liquid holdup on trays, reboiler and condenser.

- The column is operated at constant 1 atm pressure.
- The actual liquid phase molar fraction is equal to ideal liquid phase molar fraction  $x = x^*$ .
- Efficiency and vapour molar fractions are correlated with Equation (2.1).
- Theoretical vapour and liquid molar fractions are correlated with Equation (2.2). This polynominal is created by equilibrium data [3].

$$eff_T = \frac{y_i - y_{i+1}}{(y_i)^* - y_{i+1}} \quad (2.1)$$

$$y^* = -23.805x^6 + 82.142x^5 - 112.57x^4 + 78.529x^3 - 29.821x^2 + 6.5244x + 0.003 \quad (2.2)$$

From the assumptions above, there are some equations for molar fractions of each tray, condenser and reboiler. In this work, condenser and reboiler are treated as a tray and numbering of trays starts from at the top of the column with 1 (condenser) and ends at the bottom of the column with 10 (reboiler). Between 1 and 10 are real trays. Since stripping section, feed section, enriching section, condenser and reboiler differ in their motion within the column, there are different equations for each of them. For example in 7th tray (stripping part), state-space model to be used in MATLAB can be written in Equation (2.3)[4]. Also, 5th tray which is feed tray, reboiler, condenser and 3rd tray (enriching section) state-space equations are given in Equation (2.4) (2.5) (2.6) (2.7) respectively. 3rd tray and 7th tray is chosen arbitrarily. It must be noted that 1st and 10th indices are condenser and reboiler respectively. Other tray equations can be easily found by changing the indices.

In MATLAB environment, in order to conduct a PID scheme, s-function is used. S-function is a function that represents the entire column for each tray with given parameters. It is a description of a Simulink model. Without any specific transfer functions, s-functions enables the user to create control schemes in Simulink. These equations below as well as other trays' equations are used to build the s-function.

$$\frac{dx_8}{dt} = -\frac{(L+F)}{N_8}x_8(t) + \frac{(L+F)}{N_8}x_7(t) + \frac{V}{N_8}y_9(t) - \frac{V}{N_8}y_8(t) \quad (2.3)$$

$$\frac{dx_6}{dt} = -\frac{(L+F)}{N_6}x_6(t) + \frac{L}{N_6}x_5(t) + \frac{F}{N_6}x_F(t) + \frac{V}{N_6}y_7(t) - \frac{V}{N_6}y_6(t) \quad (2.4)$$

$$\frac{dx_{10}}{dt} = -\frac{B}{N_{10}}x_{10}(t) + \frac{L+F}{N_{10}}x_9(t) - \frac{V}{N_{10}}y_{10}(t) \quad (2.5)$$

$$\frac{dx_1}{dt} = -\frac{(L+D)}{N_1}x_1(t) + \frac{V}{N_1}y_2(t) \quad (2.6)$$

$$\frac{dx_4}{dt} = -\frac{L}{N_4}x_4(t) + \frac{L}{N_4}x_3(t) + \frac{V}{N_4}y_5(t) - \frac{V}{N_4}y_4(t) \quad (2.7)$$

After the equations are set out, the initial steady state values for each index can be found via `fsolve` built-in function in MATLAB [6]. `fsolve` tries to solve a system of equations by minimizing the sum of squares of its components. If the sum of squares equals zero, the system of equations is solved. It is known that vapour fraction is correlated with liquid fraction and flow rates of distillate, bottom and feed are constant. First, arbitrary initial values are assigned to each tray from 0 to 1 and then, by iteration from `fsolve`, the real initial steady state values are found. The initial steady state values are given below in MATLAB snippet. `x0` denotes initial steady state value array.

```
1 x0 = [0.9526 0.8926 0.8406 0.7916 0.7398 0.6703 0.6478
      0.5953 0.4857 0.2023];
```

## 2.3 Control Configurations and PID Controllers

Pairing the manipulated variables and controlled variables may be a problem when the system is complex. Relative Gain Array calculation is used to better the pairing of controlled and manipulated variables [13]. Several control configurations such as LV, LB, DV and LD are proposed in the literature. These two letters show the manipulated variable. The LV setup suggests adjusting the reflux flow rate and the heating steam

flow rate. Controlling both the bottom and reflux flow rates is recommended by the LB setup. The distillate flow and the heating steam flow rate are the two primary control loops in the DV arrangement. The reflux and distillate flow rates are managed by the LD setup. LV configuration can be seen in Figure 2.4 [13, 10]. Here, L is liquid flow rate through the column,  $M_B$  is liquid hold-up at the bottom,  $M_D$  is liquid hold-up at the top,  $z_F$  is feed molar fraction for volatile component.

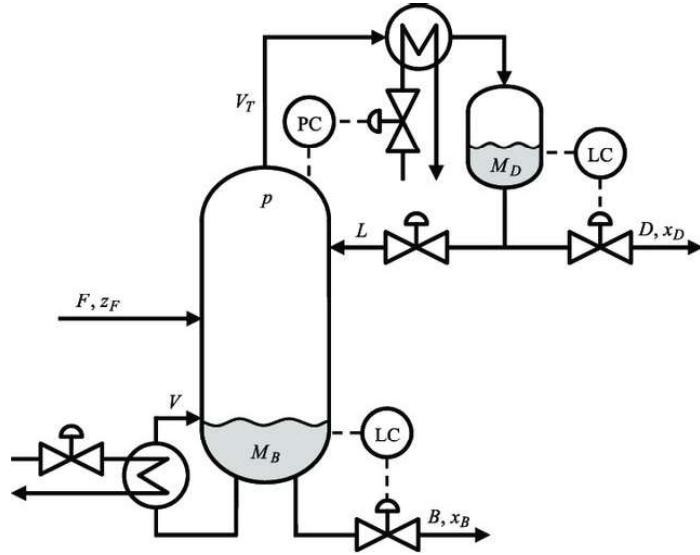


Figure 2.4: LV configuration [11].

PID controllers are predominantly used in the industry since the beginning of 1930s. It showed its robustness among every industry. PID helps understand process dynamics and makes a connection between controlled variable and manipulated variable using three distinct control loop strategies. One part of implementing a PID controller requires trial and error for their distinctive coefficients. This trial and error can be called tuning. Other part of implementing PID is selecting the structure of PID. Some academics developed tuning rules, for example Ziegler-Nichols [7]. Process dynamics usually have a form of first-order plus delay model which is described by dead time and time constant. These type of controllers are still being used by several academics and industries due to its robustness. It is the primary control strategy used in every industry [9, 10].

The difference between set point and measured value is called error. The objective of all control modes, namely Proportional, Derivative and Integral is to eliminate

the difference between set point and measured value given in Equation (2.8).  $y_s$  is the set point and  $y_m$  is measured value at any instant  $t$ .  $K_c$  value which is called controller gain and its value is changed according to the system specifications. The controller output of proportional controllers is given in Equation (2.9). where  $p(t)$  is controller output and  $p^*$  is steady state value. The biggest disadvantage of proportional controllers is that there is always an offset present, meaning the measured value will never reach the set point. Offset could be diminished by increasing the controller gain but this in turn increases oscillation which causes unstable process [9, 10].

$$e(t) = y_s - y_m(t) \quad (2.8)$$

$$p(t) = p^* + K_c e(t) \quad (2.9)$$

The biggest advantage of integral controllers are to eliminate the offset. Integral controller output depends on error over time until error is zero. Oscillatory behaviours of the integral controllers are eliminated by choosing correct integral time constant. They are usually used with P controllers since proportional control takes immediate action when an error is present. Its formula is given in Equation (2.10), where  $\tau_I$  is reset time or integral time. PI controllers are sufficient and it is easier to do tuning by trial and error method [9, 10].

$$p(t) = p^* + K_c(e(t) + \frac{1}{\tau_I} \left[ \int_0^t e(t^*) dt^* \right]) \quad (2.10)$$

The objective of derivative controllers is to predict the behaviour of the output by rate of change of error, meaning that if it detects a change in the error, it takes action rapidly. It is important to say that P-controllers behaves according to the deviation only and integral controllers are not effective for abrupt deviation in temperature. If the error signal does not change, derivative controllers alone could be useless. Because of that, derivative controllers are used with proportional or integral controllers. Proportional and derivative controllers in transfer function form is given in Equation (2.11) where  $P(s)$  is difference between controller output signal and nominal (steady-state) controller output,  $E(s)$  is error signal and  $\tau_D$  is derivative time. Derivative controllers make the process stable. That is to say it is utilized to make up for destabilizing tendency

created by integral mode. Derivative control enhances the controlled variable response by settling time which is the time it takes reducing to reach steady state. If the process is noisy, derivative action is useless. Temperature controllers with deadtime in heat exchangers or distillation column trays utilizes the derivative action. When process deadtime is big derivative controllers are used [9, 10].

$$\frac{P(s)}{E(s)} = K_c(1 + \tau_D s) \quad (2.11)$$

Apart from the modes mentioned above, PID mode exists. This mode is predominantly used in the industry [17]. Parallel form of this mode is given in Equation (2.12).

$$\frac{P(s)}{E(s)} = K_c \left( 1 + \frac{1}{\tau_I s} + \tau_D s \right) \quad (2.12)$$

PID controllers have a disadvantage in that a sudden change in set point (or error,  $e$ ) can cause the derivative term to become very large, resulting in a derivative kick to the final control element. To avoid a quick spike the derivative action should be based on the measurement  $y_m$  rather than the error signal  $e$ .

The parallel or textbook formula is frequently referred to as a decoupled PID form. Changing the value of D term merely affects the size of the derivative action, which is independent of the proportional and integral terms. The PID controller's parallel construction leads to the decoupling of the three terms [9].



# Chapter 3

## Practical Part

This chapter is about MATLAB parameters, PID tuning methods and PID tuning calculation for both distillate and bottom molar fractions. In MATLAB parameters section, important MATLAB parameters that define the s-function and their importance are given, PID tuning methods exclude some of the methods and lastly, PID tuning calculation reveals the best P,I and D coefficients.

### 3.1 MATLAB Parameters

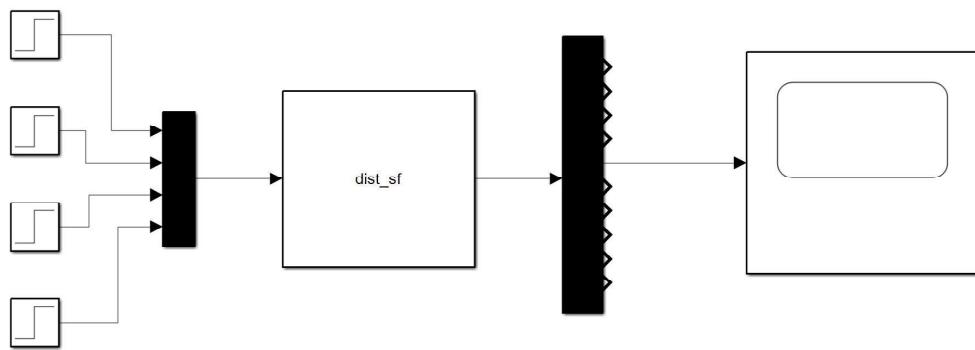
Number of trays are 8 because the laboratory distillation column consists of 8 trays. Number of inputs to the Simulink model is determined as 4 namely, feed concentration, feed flow, distillate and bottom flow. These input are very commonly used [18]. In this case, number of states are equal to number of outputs and they are the molar fractions of each tray in liquid phase of volatile component. Number of output is 10 because reboiler and condenser are treated as a tray. The reboiler holdup is 1 kmol and the tray and condenser holdup is 0.2 respectively. The efficiency of tray and reboiler is 0.8 and 1. Feed concentration is 0.65, feed flow is 0.300 kmol/h, distillate flow is 0.179 kmol/h and vapour flow is 0.328 kmol/h. These data can be used as a source for methanol-water distillation considering the size of the laboratory distillation column [1, 15]. Below is the MATLAB snippet. Inside this s-function are state-space equations of 10 states. Also, the simplest Simulink scheme with 4 inputs and 10 outputs without any controller is given in Figure 3.1.

---

```

1 N = 8; % number of trays
2 nu = 4; % inputs = [xF nF nD nV]
3 nx = N+2; % number of states
4 ny = N+2; % number of outputs
5
6 % model parameters
7
8 k = 5; % feed tray number
9 nB = 1; % reboiler hold up [kmol]
10 nT = 0.2; % tray and condenser hold up [kmol]
11 effT = 0.80; % efficacy - tray
12 effB = 1.00; % efficacy - reboiler
13
14 % inputs
15 xF = 0.65; % molar ratio of light component in feed
16 F = 0.300; % molar flow of feed
17 D = 0.179; % molar flow of distillate
18 V = 0.328; % molar flow of vapour

```

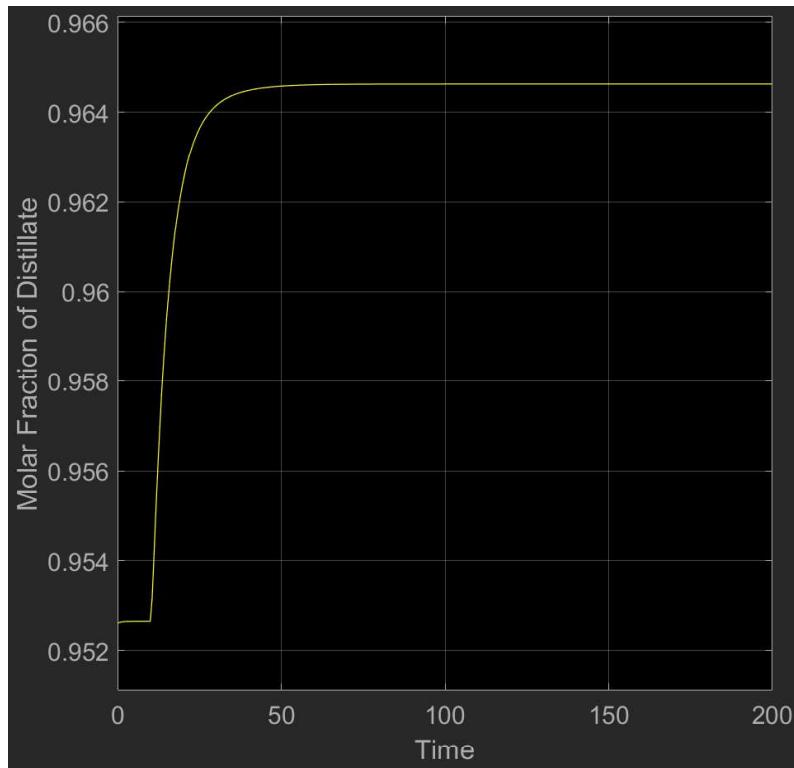


**Figure 3.1:** The simplest Simulink scheme.

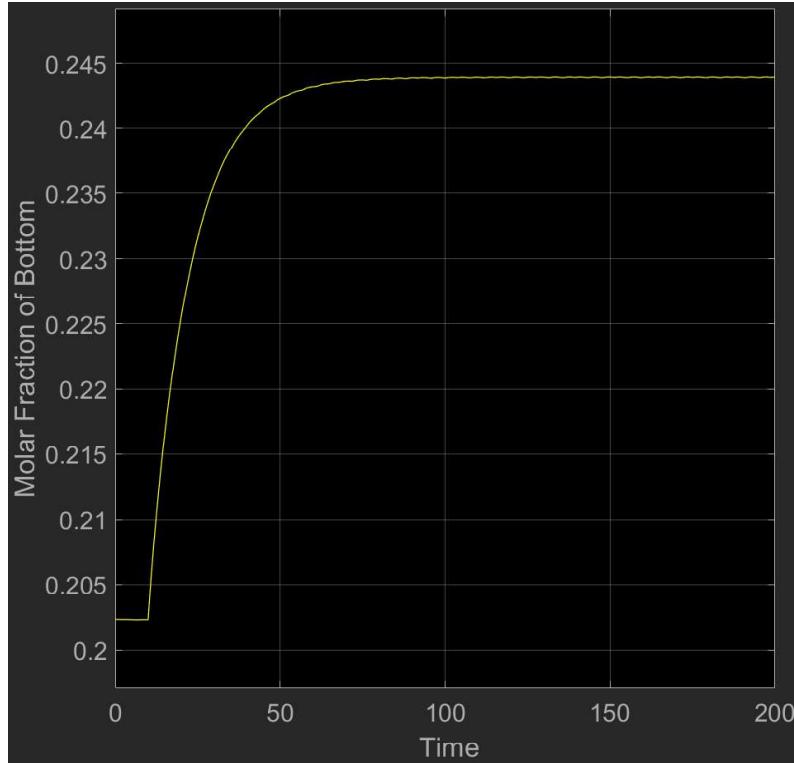
### 3.2 PID Tuning Methods

In the literature, there are lots of PID tuning methods available such as Ziegler-Nichols, Cohen Coon which involve deadtime [7]. They require to make an open-loop scheme

and calculate gain, deadtime and time constant to determine P, I and D constants. Apart from open-loop scheme, oscillatory behaviour can be used for tuning. However, in MATLAB outputs, there is no deadtime in the response for both distillate molar fraction and bottom molar fraction in Figure 3.2 and 3.3 respectively. Distillate flow decreases from 0.179 to 0.160 and bottom flow decreases from 0.121 to 0.100 as a step change in 10th second which in turn increases the compositions of volatile component from 0.9526 and 0.2023 to 0.9646 and 0.2439. Also, oscillatory behaviour is not observed when increasing P constant as the system is non-linear. In this case, other PID tuning methods are used. In one of the methods which uses PI controller, the gain and time constant is calculated and P constant is determined by  $1/(K_p)$ ,  $2/(K_p)$  and  $5/(K_p)$  in order and time constant  $T$  determines the integral part. Integral coefficients are  $1/(T)$ ,  $2/(T)$  and  $5/(T)$  respectively for each block [5]. Time constant of distillate molar fraction is 6.0665 and for bottom molar fraction, it is 9.5. Another tuning method is the MATLAB itself.



**Figure 3.2:** Distillate open loop response.



**Figure 3.3:** Bottom open loop response.

### 3.3 PID Tuning Calculation

Before doing the calculation for tuning, Relative Gain Array is constructed to determine the best manipulated variables and controlled variables [19]. In order to construct a relative gain array, step changes are carried out. Our candidate manipulated variable for distillate molar fraction is distillate flow and for bottom molar fraction is bottom flow. Remember that 0.179 kmol/h is the initial steady state value for distillate flow and 0.121 kmol/h is the initial steady state value for bottom flow.  $x_D$  is distillate molar fraction and  $x_B$  is bottom molar fraction. Table 3.1 shows the manipulated variable step changes and the responses in fractions.

As can be seen, in first 3 runs, only distillate flow is changed by step change and last 2 belong to only bottom flow step change. Let's denote the distillate flow  $u_1$ , the bottom flow  $u_2$ , the distillate fraction  $y_1$  and the bottom fraction  $y_2$ . All 5 runs' gain

**Table 3.1:** Different Runs for Distillate and Bottom Flow.

| Run | D (kmol/h) | B (kmol/h) | $x_B$  | $x_B$  |
|-----|------------|------------|--------|--------|
| 1   | 0.160      | 0.121      | 0.9646 | 0.3360 |
| 2   | 0.135      | 0.121      | 0.9777 | 0.5208 |
| 3   | 0.110      | 0.121      | 0.9875 | 0.7138 |
| 4   | 0.135      | 0.101      | 0.9780 | 0.6235 |
| 5   | 0.135      | 0.141      | 0.9774 | 0.4472 |

is denoted as  $g_{ij}$  with i being controlled variables and j being manipulated variables. For example,  $g_{21}$  is the gain on bottom fraction change when distillate flow changes. Any run couple can be selected to calculate the gain  $g$ . For gains  $g_{11}$  and  $g_{21}$ , the first two run are chosen and for gains  $g_{12}$  and  $g_{22}$ , the last 2 are chosen. Gain calculations are given in Equation (3.1), (3.2), (3.3) and (3.4). Minus comes from the fact that if the flow rate increases, fraction decreases and vice versa.

$$g_{11} = \frac{0.9777 - 0.9646}{0.135 - 0.160} = -0.524 \quad (3.1)$$

$$g_{21} = \frac{0.5208 - 0.3360}{0.135 - 0.160} = -7.392 \quad (3.2)$$

$$g_{12} = \frac{0.9774 - 0.9780}{0.141 - 0.101} = -0.015 \quad (3.3)$$

$$g_{22} = \frac{0.4472 - 0.6235}{0.141 - 0.101} = -4.4075 \quad (3.4)$$

After finding the gains, Relative Gain Array must be created. Relative Gain Array matrix is given in Equation (3.5) where  $\lambda$  is ratio of open loop gain to closed loop gain given in Equation (3.6). According to Relative Gain Array method, the couples whose cells are equal to one or near one must be selected for best coupling. In this case,  $u_1$  must be coupled with  $y_1$  and  $u_2$  must be coupled with  $y_2$ . Calculations can be seen in Equation (3.7) and (3.8).

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix} = \begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} \quad (3.5)$$

$$\lambda_{11} = \frac{1}{1 - \frac{g_{12}g_{21}}{g_{11}g_{22}}} \quad (3.6)$$

$$\lambda_{11} = \frac{1}{1 - \frac{(-0.015)(-7.392)}{(-0.524)(-4.4075)}} = 1.05 \quad (3.7)$$

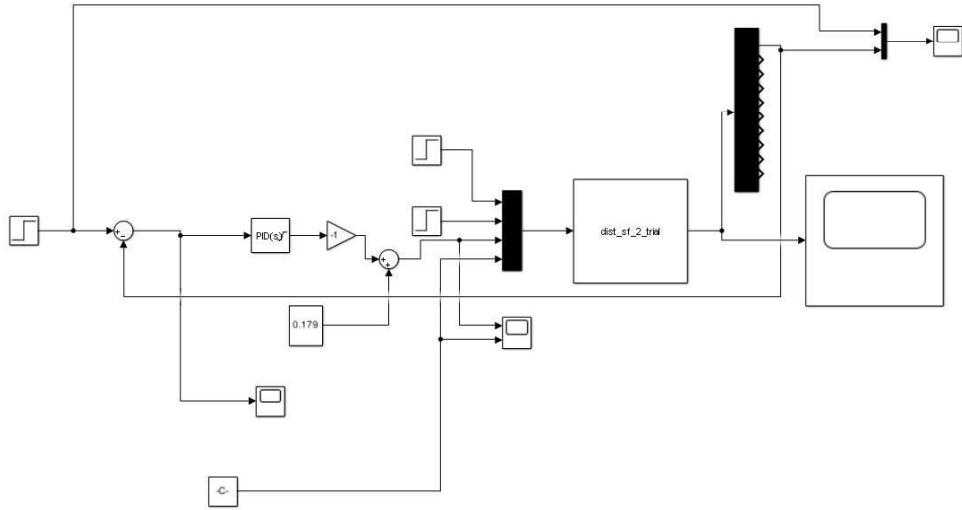
$$\Lambda = \begin{bmatrix} 1.05 & 0.05 \\ 0.05 & 1.05 \end{bmatrix} \quad (3.8)$$

As for PID formula, in Sections 3.3.1 and 3.3.2, two compensators (PID formula) have the same structure which is ideal. Its equation is given in Equation (3.9). Here  $n$  is filter coefficient for derivative part.

$$P \left( 1 + I \frac{1}{s} + D \frac{n}{1 + n \frac{1}{s}} \right) \quad (3.9)$$

### 3.3.1 Distillate Calculation

Now that the couples are determined, PID tuning can be carried out. Distillate fraction is tuned separately from bottom fraction and vice versa. Simulink scheme for distillate tuning is shown in Figure 3.4. For the distillate flow, step change of flow rate from 0.179 kmol/hr to 0.160 kmol/hr is done. This results in concentration increase in distillate liquid fraction from 0.9526 to 0.9646 shown in Figure 3.2. Gain  $K$  from this step change is 0.632. Applying tuning methods yields following shown in Equation (3.10) (3.11) (3.12). I parts are calculated by the method in Section 3.2. I parts are 0.1648, 0.3296 and 0.824. In order to compare and enhance the tuning, MATLAB tuner is introduced. As for MATLAB parameters, it is suggested that P



**Figure 3.4:** Distillate fraction control scheme.

constant be 2.1995, I constant be 0.30388, D part be 0.07259 and n part be 35.537. The comparison between different PID tuning methods is given in Table 3.2.

$$\frac{1}{K} = 1.582 \quad (3.10)$$

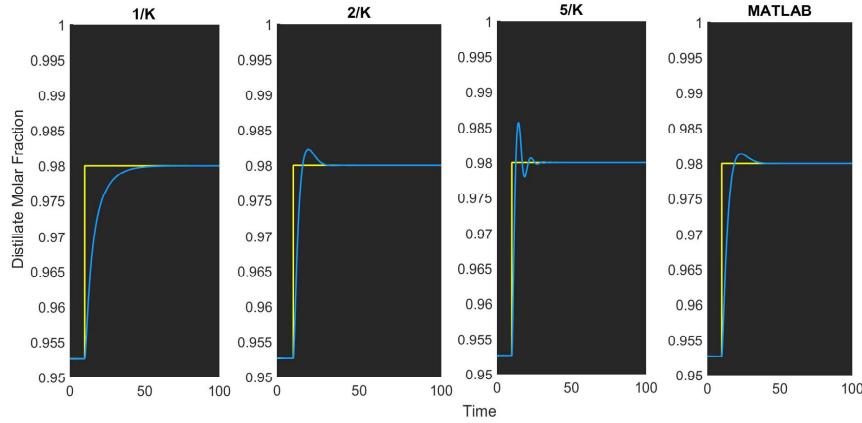
$$\frac{2}{K} = 3.164 \quad (3.11)$$

$$\frac{5}{K} = 7.910 \quad (3.12)$$

In order to visually compare PID methods, The setpoint is given as 0.98 so it will rise from the initial steady-state value which is 0.9526 to 0.98. Respective PID controllers's performance is shown in Figure 3.5 and Table 3.2. As a result, MATLAB tuner's coefficients are chosen because it has lower rise time and settling time than 1/K. Other are eliminated because they have so large overshoot.

**Table 3.2:** Comparison of Different PID Controllers for Distillate.

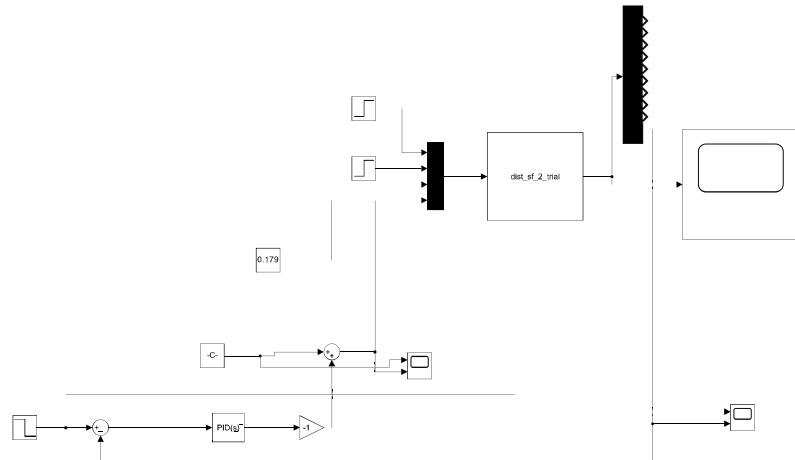
| Parameters    | $1/K$     | $2/K$     | $5/K$     | MATLAB    |
|---------------|-----------|-----------|-----------|-----------|
| Rise Time     | 11.8 sec. | 3.11 sec. | 1.07 sec. | 4.72 sec. |
| Settling Time | 20.8 sec. | 13.9 sec. | 9.74 sec. | 18.8 sec. |
| Overshoot     | 0.43%     | 9.68%     | 38.30%    | 7.05%     |
| Peak          | 1         | 1.1       | 1.38      | 1.07      |

**Figure 3.5:** Distillate PID comparison.

### 3.3.2 Bottom Calculation

Simulink scheme for bottom tuning is shown in Figure 3.6. For the bottom flow, step change of flow rate from 0.121 kmol/hr to 0.100 kmol/hr is done. This results in concentration increase in distillate liquid fraction from 0.2023 to 0.2439 shown in Figure 3.3. Gain  $K$  from this step change is 1.98. Applying tuning methods yields following shown in Equation (3.13) (3.14) (3.15). I parts are calculated by the method in Section 3.2. I parts are 0.1053, 0.2106 and 0.5265. In order to compare and enhance the tuning, MATLAB tuner is introduced. As for MATLAB parameters, it is suggested that P constant be 0.736, I constant be 0.224, D part be 0 and n part be 100. The comparison between different PID tuning methods is given in Table 3.3.

$$\frac{1}{K} = 0.505 \quad (3.13)$$



**Figure 3.6:** Bottom fraction control scheme.

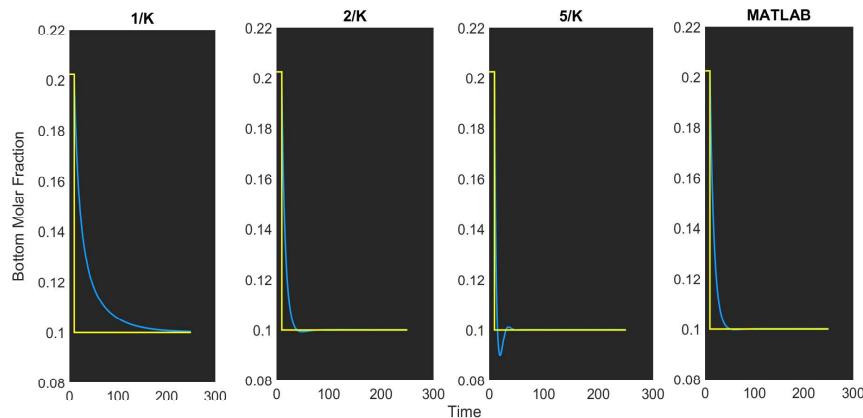
$$\frac{2}{K} = 1.01 \quad (3.14)$$

$$\frac{5}{K} = 2.525 \quad (3.15)$$

In order to visually compare PID methods, The setpoint is given as 0.1 so it will decrease from the initial steady-state value which is 0.2023 to 0.1. Respective PID controllers's performance is shown in Figure 3.7 and Table 3.3. As a result, 2/K tuner's coefficients are chosen because it has lower rise time, settling time, overshoot and peak than MATLAB tuner. 5/K has so much overshoot and 1/K has large rise time and settling time.

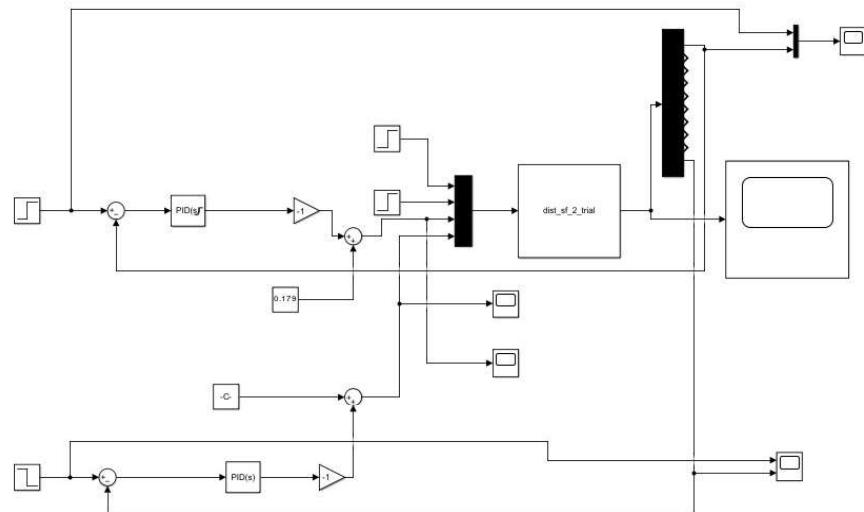
**Table 3.3:** Comparison of Different PID Controllers for Bottom.

| Parameters    | $1/K$     | $2/K$     | $5/K$    | MATLAB    |
|---------------|-----------|-----------|----------|-----------|
| Rise Time     | 25.2 sec. | 7.69 sec. | 2.4 sec. | 9.31 sec. |
| Settling Time | 41.6 sec. | 29 sec.   | 11 sec.  | 33.4 sec. |
| Overshoot     | 0.05%     | 8.41%     | 18.50%   | 9.01%     |
| Peak          | 1         | 1.08      | 1.19     | 1.09      |

**Figure 3.7:** Bottom PID comparison.

### 3.3.3 Interacting of Distillate and Bottom PID Controller

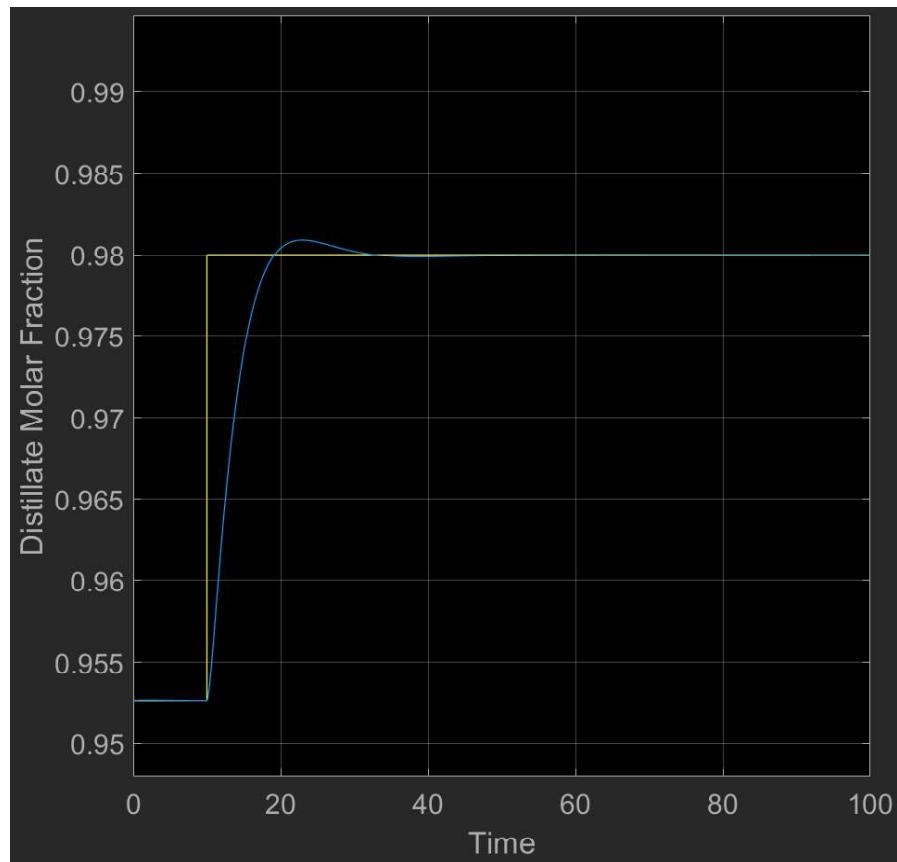
Previous PID performance results are obtained without interacting of distillate and bottom PID controller. It is desired to have the same results as separate controller in interacting mode. To test this, Simulink model shown in Figure 3.8 is created. Table 3.4 shows the difference between interacting and non-interacting for distillate fraction PID performance whereas Table 3.5 shows the difference between interacting and non-interacting for bottom fraction PID performance. Both controller performs better in interacting mode in terms of settling time, overshoot and peak. Respective graphs for distillate molar fraction and bottom molar fraction is given in Figure 3.9 and Figure 3.10. As it can be seen, for bottom molar fraction, the value initially rises and then goes down. Also, changes in both distillate flow rate and bottom flow rate are provided in Figure 3.11 and 3.12. Upper and lower limits are set in distillate flow rate for proper controller action. Upper limit is 0.179 kmol/h whereas lower limit is -0.121 kmol/h.

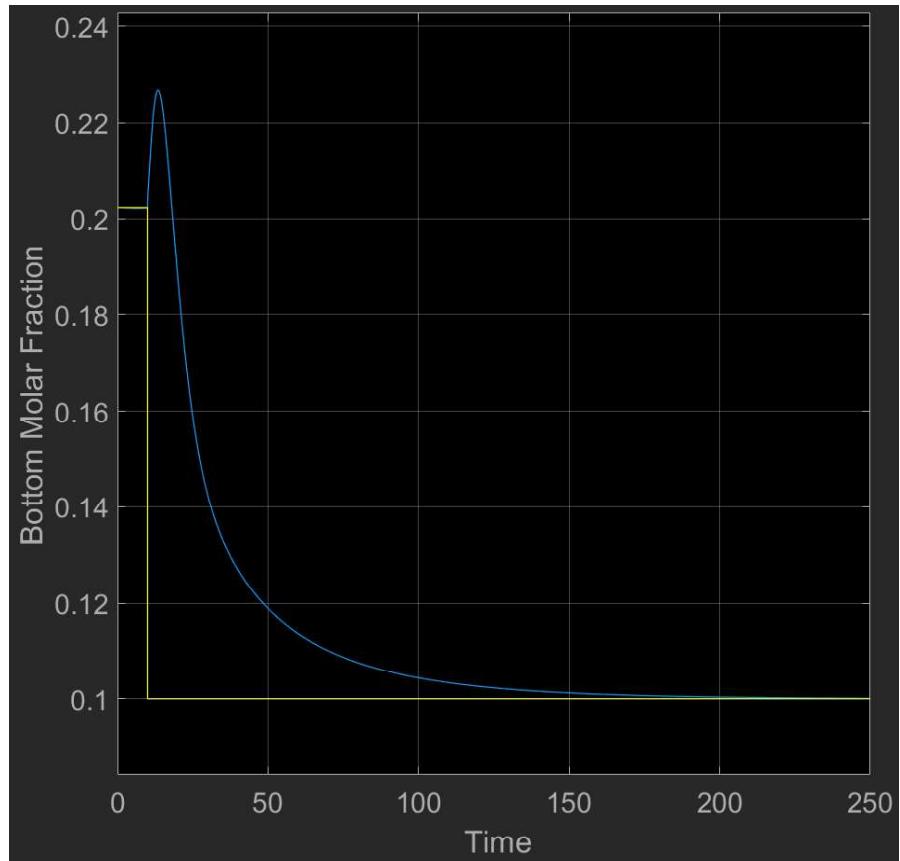
**Figure 3.8:** Interacting PID controllers scheme.**Table 3.4:** Comparison of Interacting and Non-Interacting Controllers for Distillate.

| Parameters    | Interacting | Non-Interacting |
|---------------|-------------|-----------------|
| Rise Time     | 4.75 sec.   | 4.72 sec.       |
| Settling Time | 15.6 sec.   | 18.8 sec.       |
| Overshoot     | 5.74%       | 7.05%           |
| Peak          | 1.06        | 1.07            |

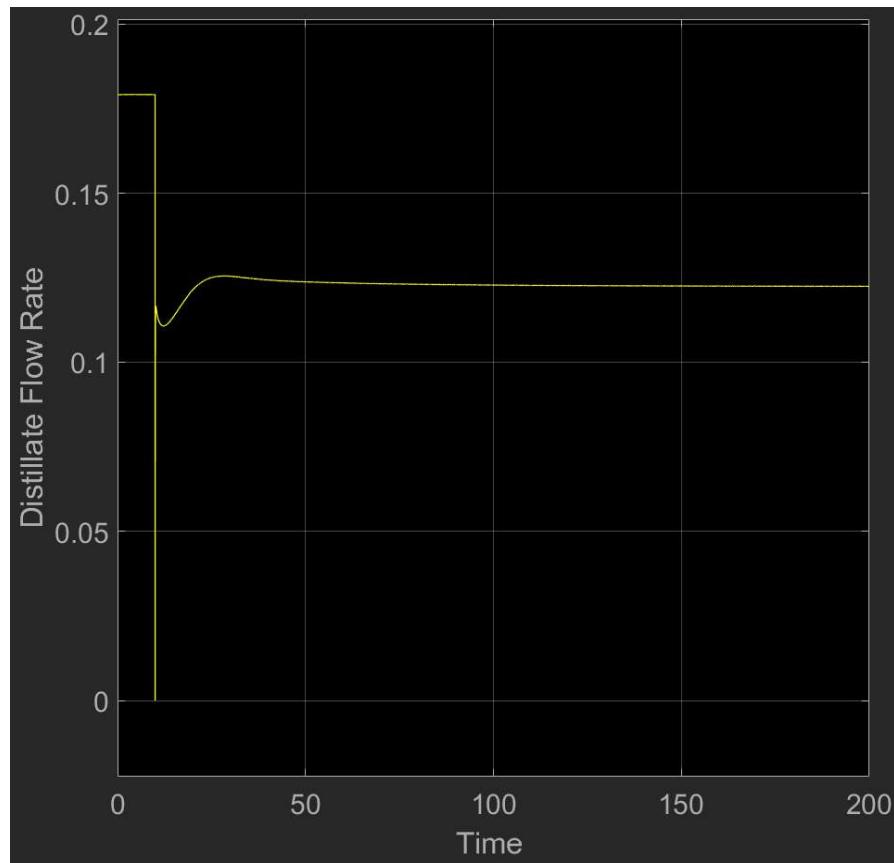
**Table 3.5:** Comparison of Interacting and Non-Interacting Controllers for Bottom.

| Parameters    | Interacting | Non-Interacting |
|---------------|-------------|-----------------|
| Rise Time     | 7.84 sec.   | 7.69 sec.       |
| Settling Time | 25.9 sec.   | 29 sec.         |
| Overshoot     | 5.57%       | 8.41%           |
| Peak          | 1.06        | 1.08            |

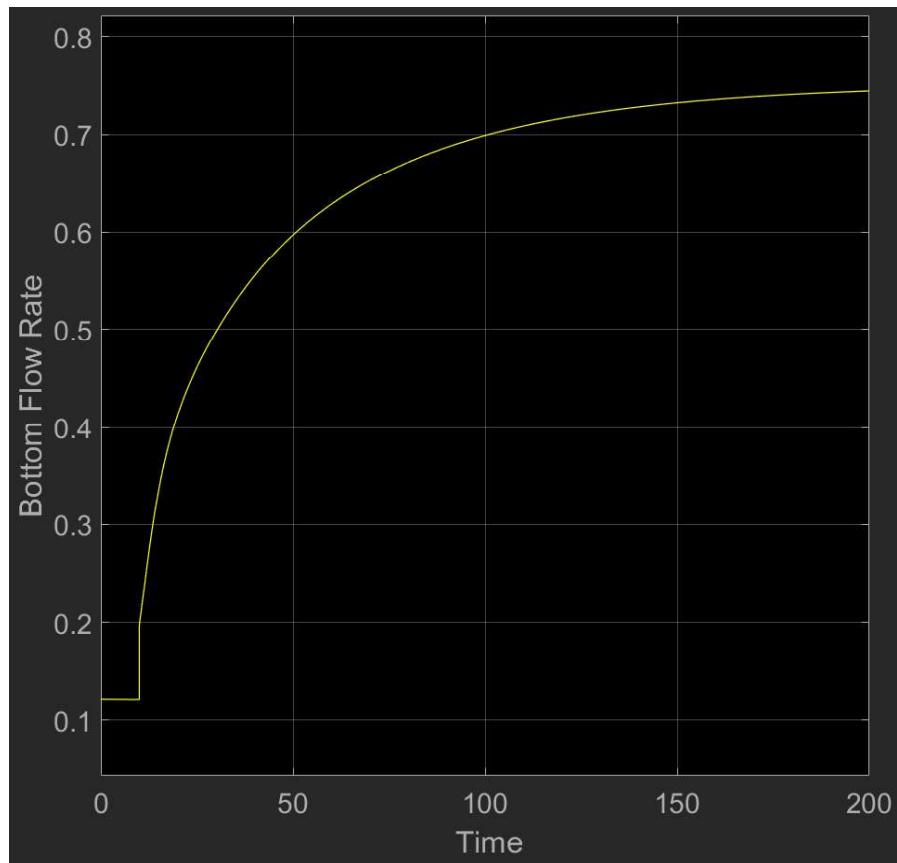
**Figure 3.9:** Distillate molar fraction in interacting mode.



**Figure 3.10:** Bottom molar fraction in interacting mode.



**Figure 3.11:** Distillate flow rate change in interacting mode.



**Figure 3.12:** Bottom flow rate change in interacting mode.



# Chapter 4

## Conclusions

It turns out that the altered variables should be the distillate flow rate and the bottom flow rate. It is worth noting that the system does not exhibit a FOPDT reaction. Four alternative tuning methods are tested, with MATLAB tuner and 2/K emerging as the best for distillate molar fraction and bottom molar fraction. For distillate molar fraction, 1/K has the largest rise time with 11.8 seconds and 5/K tuner has the largest overshoot and peak value with respective values of 38.30% and 1.38. 2/K is not chosen because it has large overshoot. Based on these values, MATLAB tuner is chosen for distillate molar fraction. For bottom molar fraction, 1/K has the largest rise time with 41.6 seconds and 5/K tuner has the largest overshoot and peak value with respective values of 18.50% and 1.19. MATLAB tuner has large overshoot. With these values, 2/K tuner is chosen for bottom molar fraction. Finally, the model with two PID controllers that interact simultaneously is simulated. It is vital to notice that they do not produce significant differences when simulated. For distillate molar fraction, rise time increases from 4.72 seconds to 4.75 seconds, settling time reduces from 18.8 seconds to 15.6 seconds, overshoot decreases from 7.05% to 5.74% and peak decreases from 1.07 to 1.06. For bottom molar fraction, rise time increases from 7.69 seconds to 7.84 seconds, settling time reduces from 29 seconds to 25.9 seconds, overshoot decreases from 8.41% to 5.57% and peak decreases from 1.08 to 1.06.



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