CHAPTER V

CONTROL STRUCTURES DESIGN

5.1 New Plantwide Control Strategies

Here, this will apply the above plantwide control structure are designed procedure to the monoisopropylamine process.

Step 1: Gather relevant plant information and control objective including constraints for control.

Before initiating work on the control structure design, is it necessary to obtain all information relevant to process control. The process objectives and constraints will determine the lower/upper bounds on the control variables as well as set points on quality variables.

Table 5.1 Constraints for control

	Constrains	Unit
Temperature inlet reactor	158.3	°C
Vapor fraction. inlet reactor	1	
MIPA product flow	≥45.36	kgmole/hr.
MIPA purity in product	0.999	
Top pressure column 1	300	psia
Top pressure column 2	30	psia
Top pressure column 3	5	psia
Pressure in flash tank	335	psia

Step 2: List manipulated variables (control degree of freedom, CDOF).

The CDOF can be obtained using the guideline given in table 5.2 and the guideline for pairing the controlled variables with the manipulated as shown in table 5.3.

Table.5.2 The control degree of freedom for monoisopropylamine production process.

Unit	Manipulated Variable	Quantity	DOF
Independent stream	Flow rate	2	2
Vaporizer	Level	1	1
Flash tank	Level Pressure	1	2
Compressor	Power	2	2
Distillation column	Distillation flow Bottom flow Reflux flow Reboiler heat removal Condenser heat removal	2	10
Distillation column	Distillation flow Bottom flow Vent flow Reflux flow Reboiler heat removal Condenser heat removal	1	6
Total of degree for freedom			23

Table5.3 Guideline pairing of manipulated and controlled variables.

Controlled Variables (CVs)	Available Manipulated Variables (MVs)
Inlet temperature reactor	Heater
Vaporizer level	Duty of vaporizer
Flash tank level	Bottom flow
Flash tank pressure	Top flow
Ammonia column pressure	Condenser heat removal Reboiler heat input Column feed flow rate
Ammonia column Tray temperature	Reboiler heat input Reflux flow rate
Ammonia column Reflux flow rate	Reflux ratio Reflux-to-feed Boil up ratio Distillate flow rate
Reflux drum Level (ammonia column)	Distillate flow rate Reflux flow rate Vapor boil up flow rate Condenser heat removal Column feed flow rate
Ammonia column Bottom Level	Bottom flow rate Vapor boil up flow rate Column feed flow rate
MIPA product column pressure	Condenser heat removal Reboiler heat input Column feed flow rate
MIPA product column Tray temperature	Reboiler heat input Reflux flow rate

Table5.3 (Continues) Guideline pairing of manipulated and controlled variables.

Controlled Variables (CVs)	Available Manipulated Variables (MVs)
MIPA product column Reflux flow rate	Reflux ratio Reflux-to-feed Boil up ratio Distillate flow rate
Reflux drum Level (MIPA product column)	Reflux ratio Reflux-to-feed Boil up ratio Distillate flow rate
MIPA product column Bottom Level	Bottom flow rate Vapor boil up flow rate Column feed flow rate
DIPA recycle column pressure	Condenser heat removal Reboiler heat input Column feed flow rate
DIPA recycle column Tray temperature	Reboiler heat input Reflux flow rate
DIPA recycle column Reflux flow rate	Reflux ratio Reflux-to-feed Boil up ratio Distillate flow rate
Reflux drum Level (DIPA recycle column)	Distillate flow rate Reflux flow rate Vapor boil up flow rate Condenser heat removal Column feed flow rate

Step 3: Establish fixture plant.

The material entered, and in-process stream must be maintained to ensure that the plant is smoothly operated.

Step 3.1: The flow rates of the major components entered the process are fixed. Since there is no isopropyl alcohol recycled, then the fresh feed of isopropanol is flow-controlled. This is the throughput handle. The total ammonia (fresh feed plus ammonia recycle D1from column C1) is flow controlled.

Step 3.2: Adjust the flow of exit material streams according to their accumulations. The flowrate of monoisopropylamine product is regulated by the level of the reflux drum of the second column. The water by-product flow rate is adjusted by the base level of the third column.

Step 3.3: Locate the quantifiers for the rest of the components. The quantifier of diisopropylamine is the reflux drum of the third column. The drum level indicates the amount of diisopropylamine resided in the process. The drum level loop is used to accommodate its amount.

There are 5 components; IPA, ammonia, monoisopropylamine, di-isopropylamine, and water in the process. The quantifiers of each component which has a potential to balance their inventories are provides the control loops below.

1. The rests of the component is di-isopropylamine: The reflux drum C3 is a quantifier of di-isopropylamine. The composition of di-isopropylamine is manipulated by the distillate flow of product C3.

2. The rests of the component is monoisopropylamine: The reflux drum C2 is a quantifier of monoisopropylamine. The component monoisopropylamine is manipulated by the distillate flow of product C2.

3. The rests of the component is ammonia: The reflux drum C1 is a quantifier of ammonia is manipulated by the distillate flow of product C1.

4. The rests of the component is water: The column base C3 is a quantifier of water. The component is water is manipulated by the bottom flow of product C3.

5. The rests of the component is IPA: The fresh feed is a quantifier of IPA.

Although there is no hydrogen feed, there is a small amount of hydrogen presented in the reactor to improve catalyst life and lower dew-point temperature, so hydrogen is circulated around in the reaction separation in two gas streams. Hydrogen is not consumed in the reaction. We must provide control to regulate the amount of hydrogen presented in this section via pressure control at the flash tank and composition control at the compressor stream 14.

The resulted control structure *Step 3* is given in Fig. 5.1.

Step 4: Handling the disturbances.

In this step, the disturbances are handled by configuring the control loops employing the principle of disturbances management:

(4.1) Heat Disturbances

This step is to design the control loops to deal both thermal and material disturbances, which the thermal disturbance is divided into 2 categories.

Step 4.1.1 HDC1: Heat Disturbance Category 1 (HDC1) is rejected at the reactor preheater HX1 and cooler HX2. In the heat exchangers, the cool stream out temperatures are controlled by adjust the by-pass stream of the heat exchanger. So the heat disturbances enter the heat exchangers are directed to the hot stream, as shown in Fig. 5.5. *Step 4.1.2* HDC2: Heat Disturbance Category 2 (HDC2) is the thermal disturbances that enter the distillation columns. The three distillation column feed temperatures are control by adjusting by heat duty reboiler, as shown in Fig. 5.4.

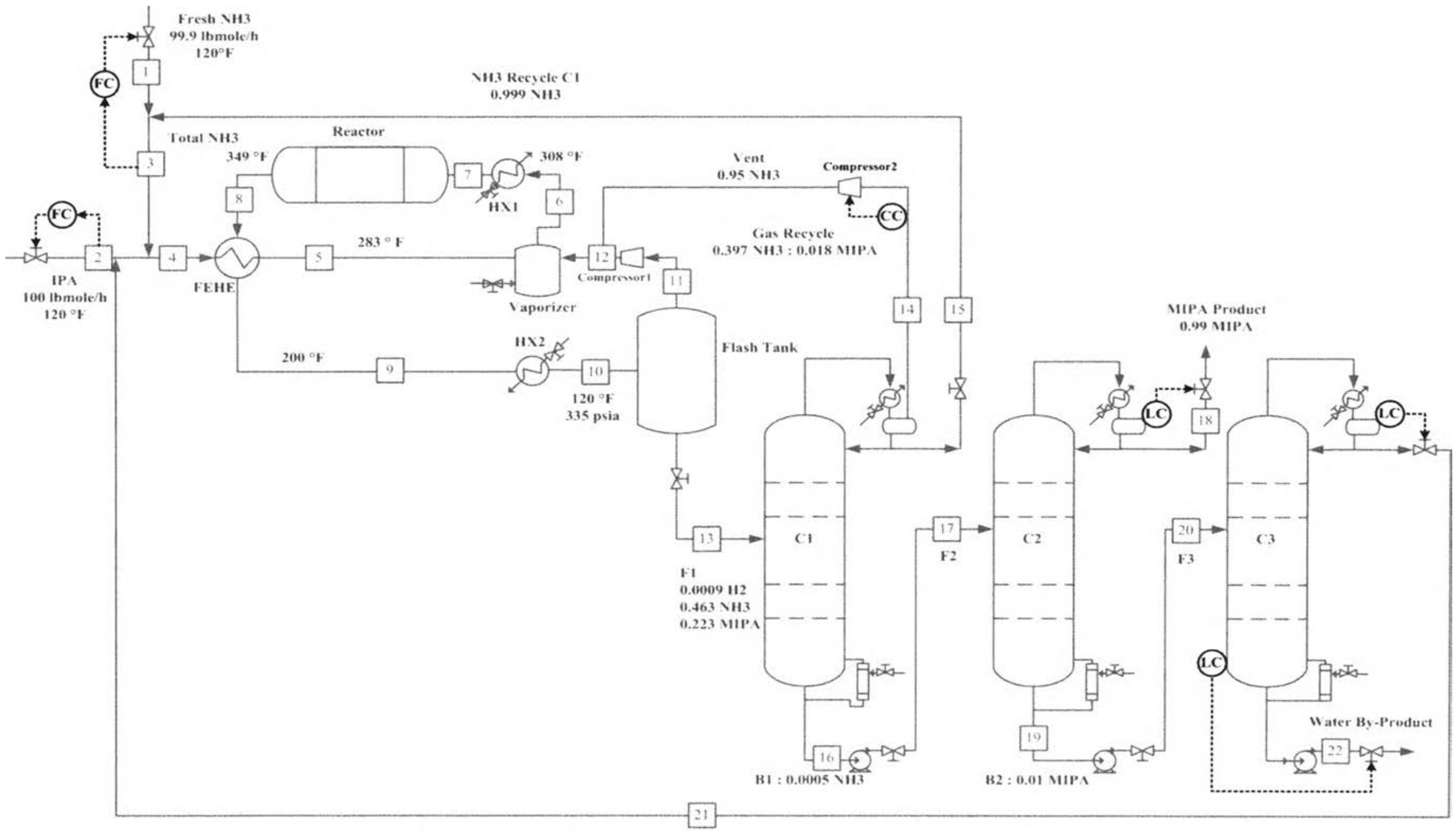


Figure 5.1: Control loops to balance in the Step 3 establishes fixture plant

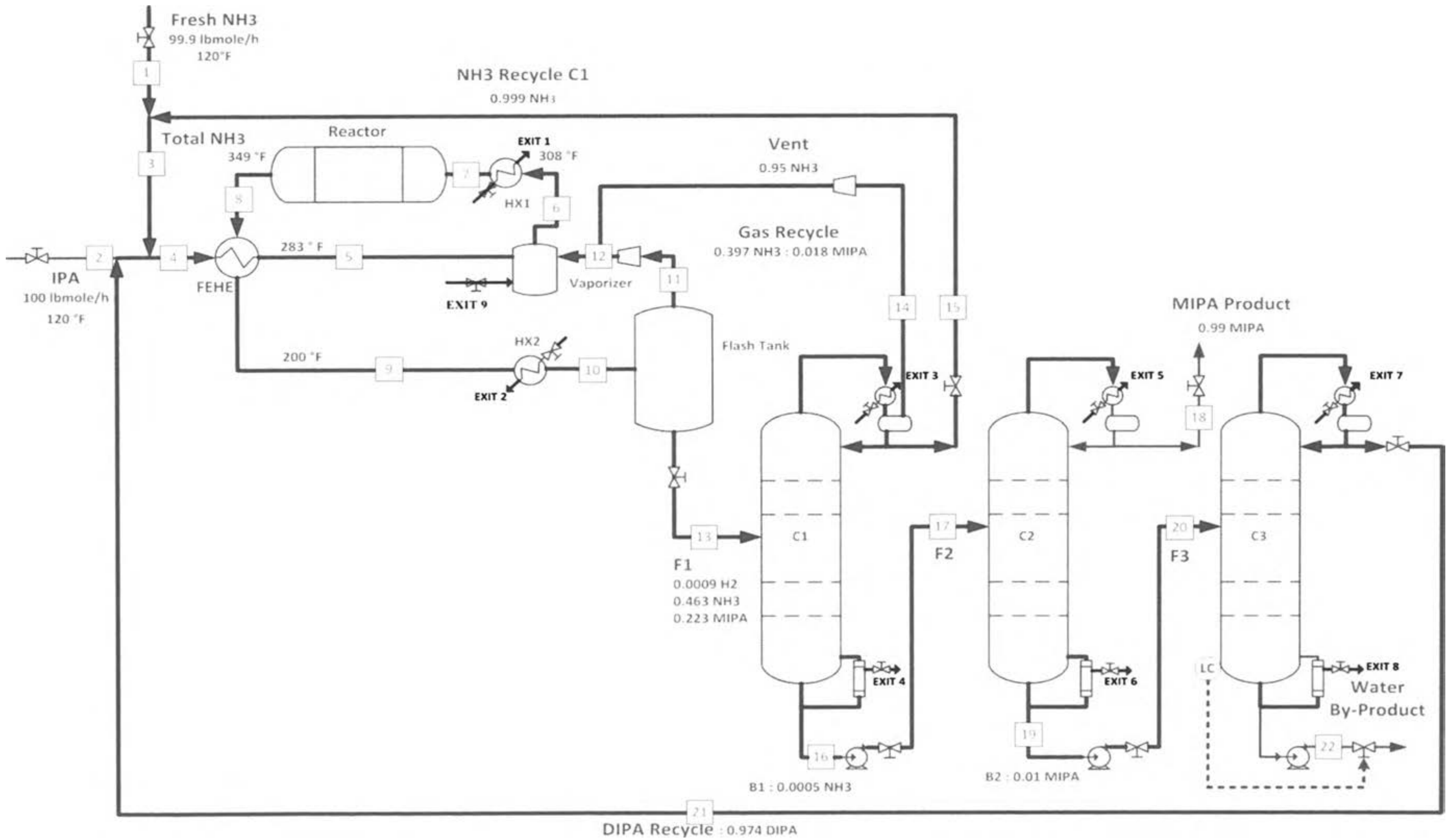


Figure 5.2: The heat disturbances pathways from ammonia feed flow.

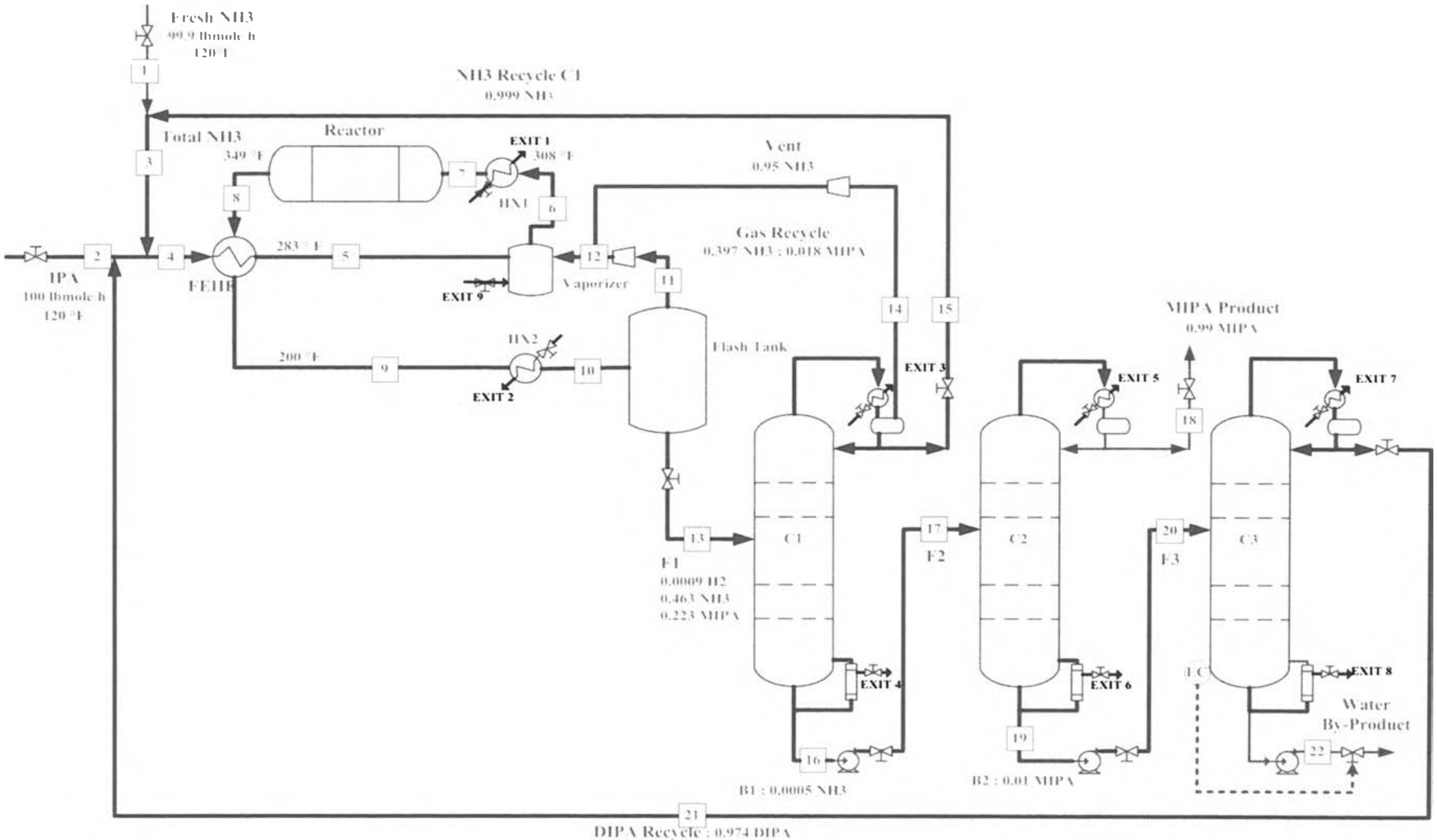


Figure 5.3: The heat disturbances pathways from IPA feed flow.

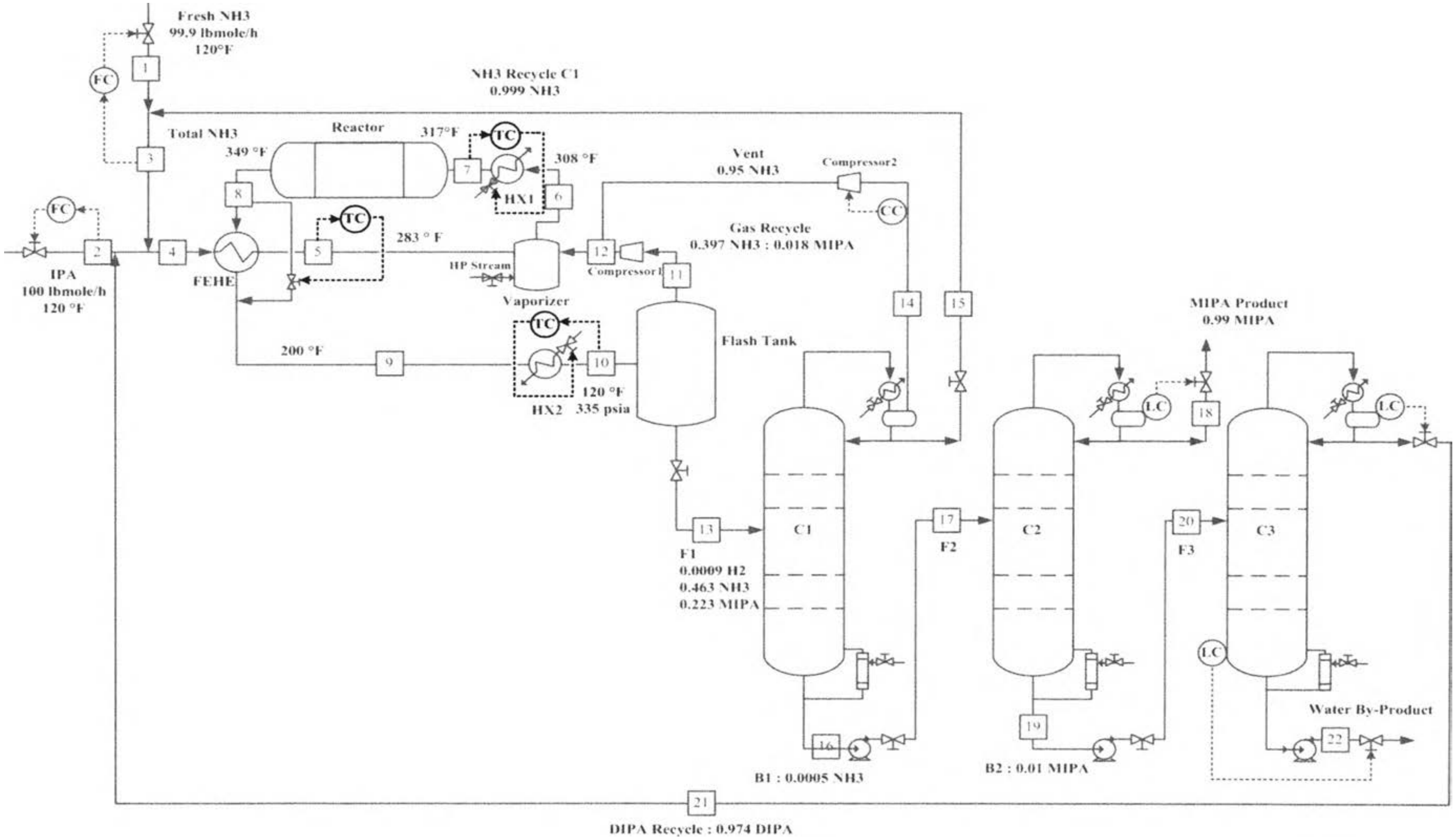


Figure 5.4: Flowsheet of control method of heat disturbance that does not directly effect on product qualities (By passing hot streams)

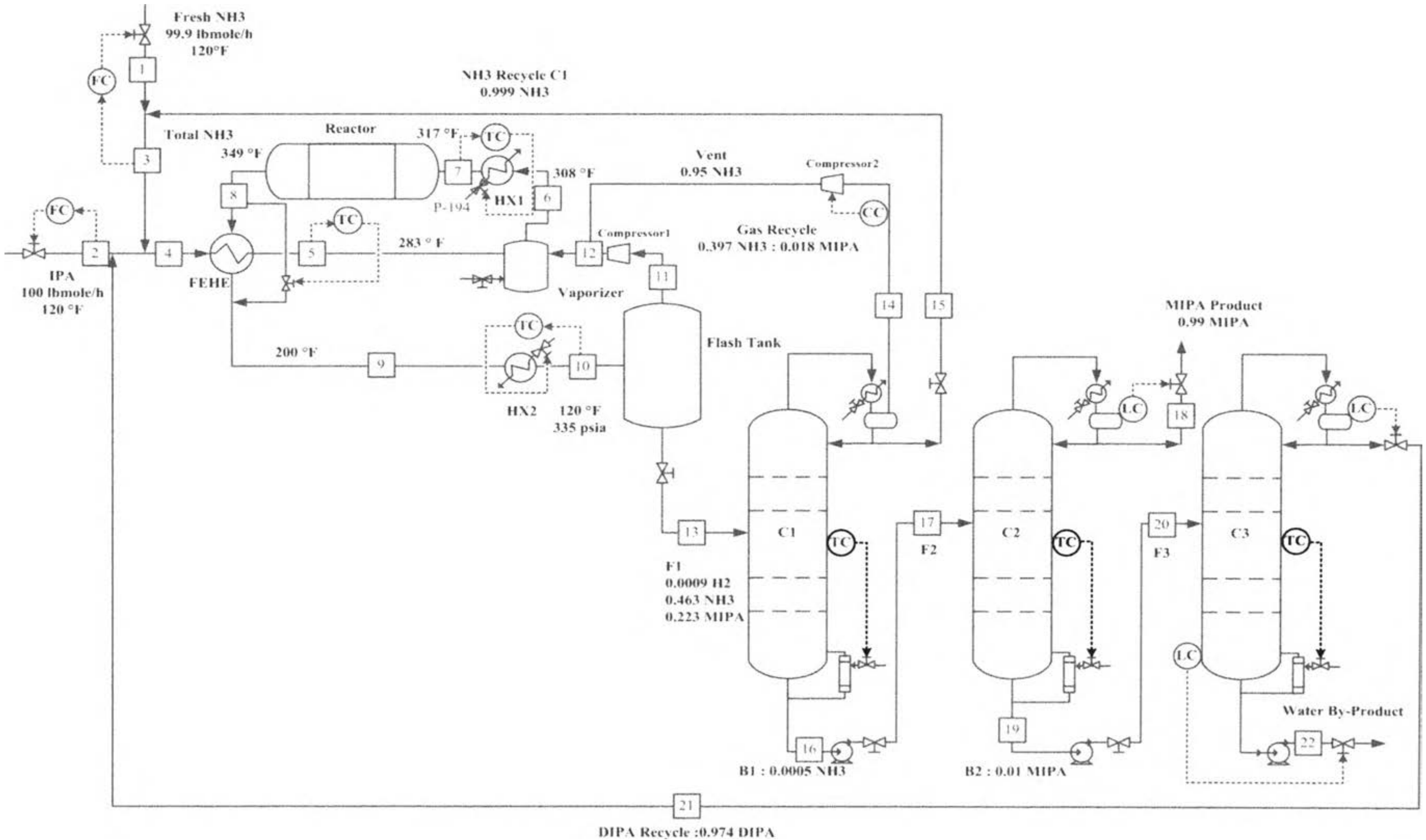


Figure 5.5 Flowsheet of control method of heat disturbance that directly effect on product qualities (By passing hot streams)

Step 4.2: Material disturbances

There are path ways of any material showed in figure 5.2-7. The material pathways are indicated the control structures to manage each material according to their pathways. Isopropyl alcohol is controlled to react in reactor R1. Ammonia is controlled to react in reactor R1. Unconverted ammonia is recycled back to the process combining the fresh feed ammonia by distillated column C1 and vent at column C1. The MIPA is generated at reactor is distillated at column C2. The DIPA is generated at reactor and distillated at column C3 to be recycled back. The water by-product is generated at reactor and carried out as a product from the bottom of column C3. All control structures established from step 3 are shown as following figure 5.6 - 5.10.

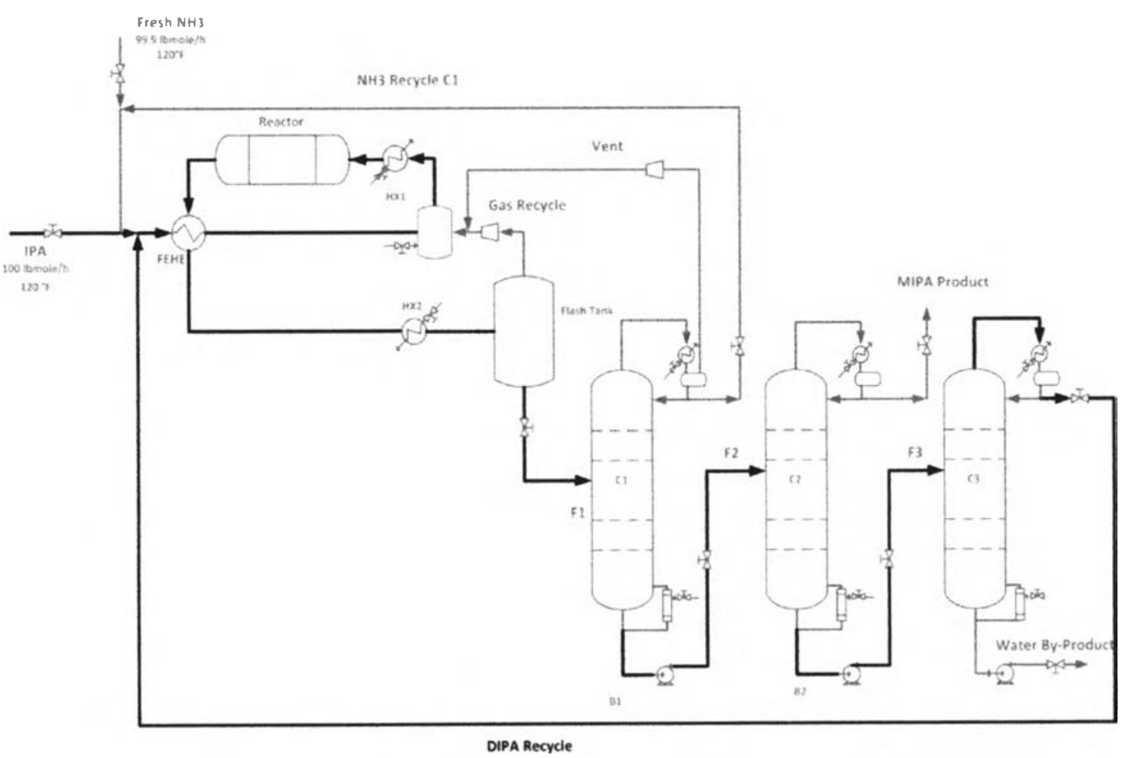


Figure 5.6: The material pathway of IPA

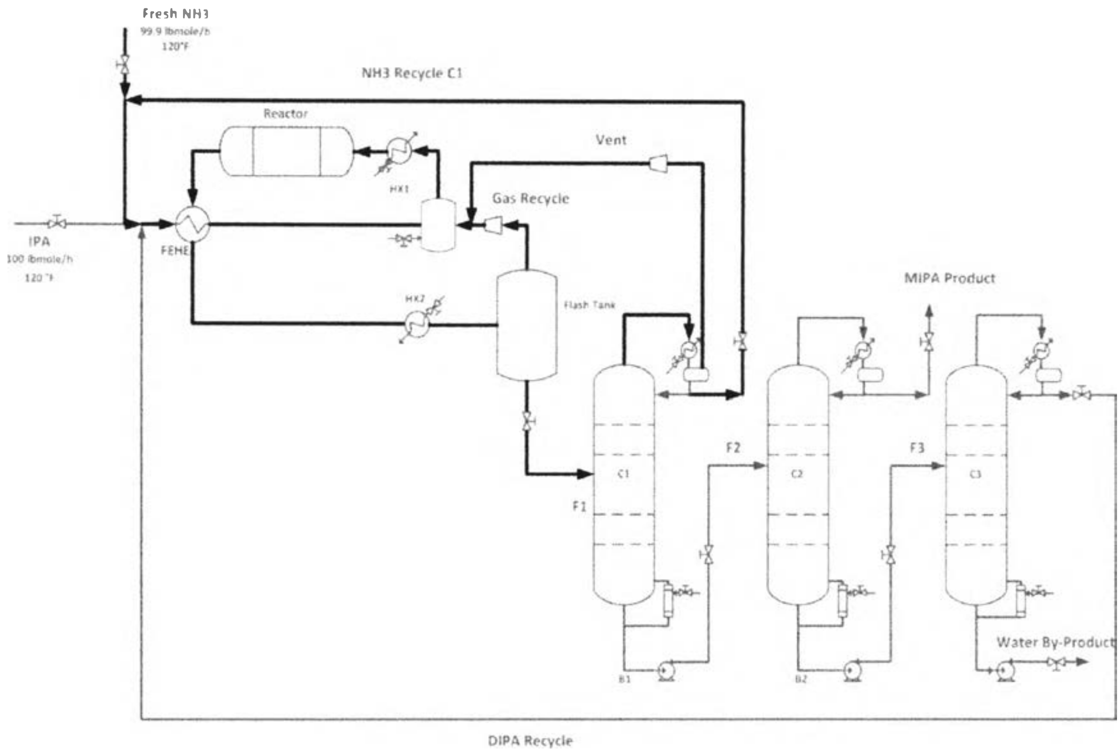


Figure 5.7: The material pathway of NH_3

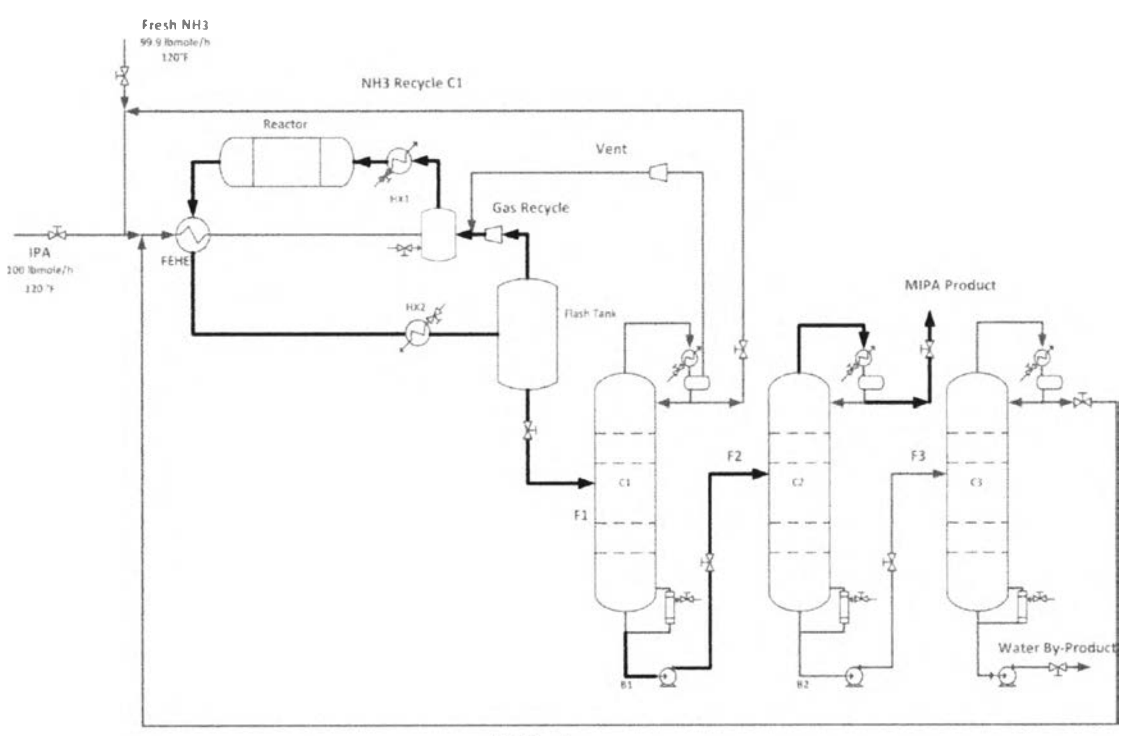


Figure 5.8: The material pathway of MIPA.

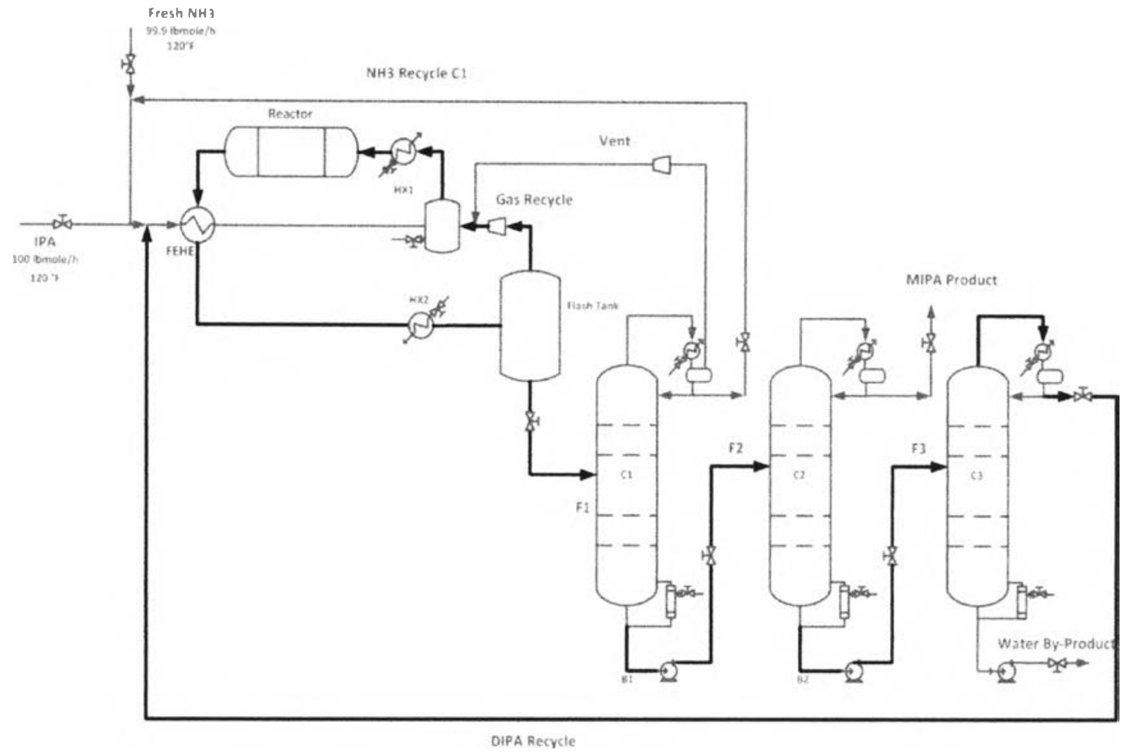


Figure 5.9: The material pathway of DIPA.

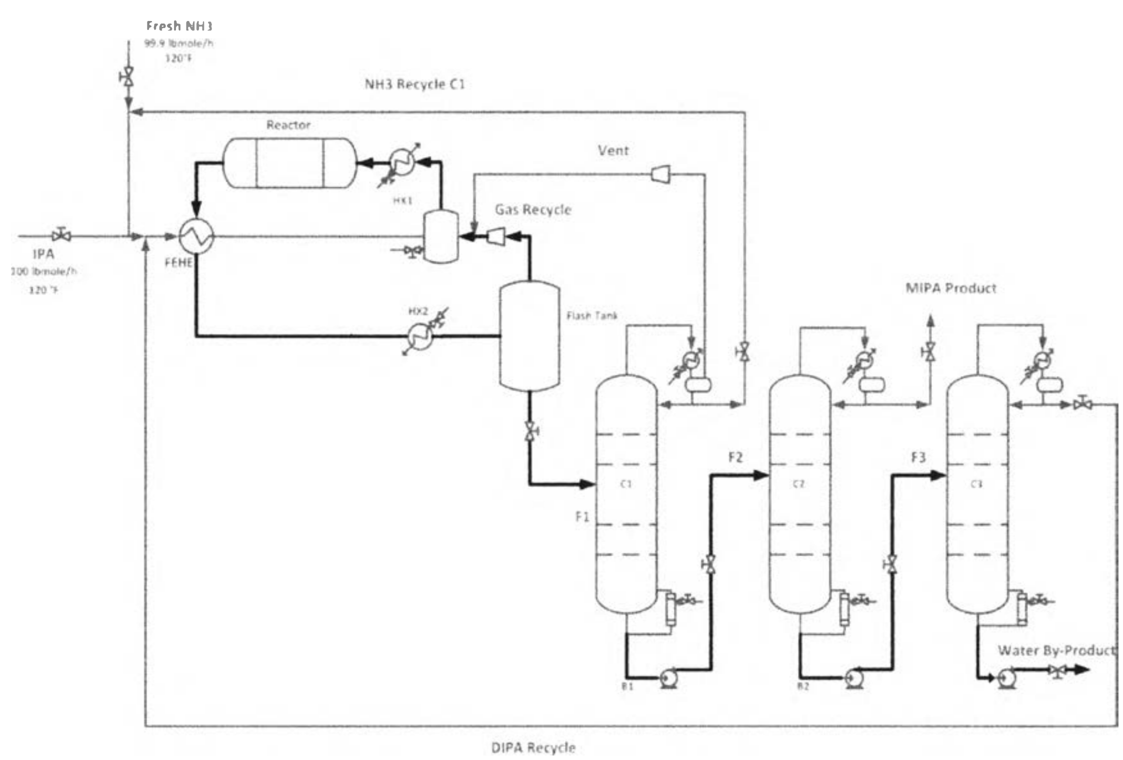


Figure 5.10: The material pathway of water.

Step 5: **Design the control loops for the remaining control variables and/or adding enhanced controls, i.e. cascade, feed forward controls.**

The remaining control loops are the rest of inventory control loops. They level loops are of vaporizer, and flash tank, bottoms of columns C1 and C2, and condenser level of column C1. The pressure loops are of flash tank, columns C1, C2, and C3. The resulted control loops are shown Fig 5.12. The function of the vaporizer is to deliver the reactor vapor feed; then its level is regulated by the vaporizer's heat input. The level of flash tank and bottoms of columns C1 and C2 are regulated by bottom flowrates. The condenser level of C1 is regulated by ammonia recycle flowrate. The pressure of flash tank is regulated by compressor work. The pressures in all columns are control by condenser duties. The other control structure (CS0 and CS2) is no controlling of temperature heat exchanger control as for studied the effects of controlling the heat exchange outlet temperature.

Step 6: **Energy management via heat exchanger networks.**

If potential heat exchanger networks or alternative heat integrated processes (HIPs) exist, list additional control variables and manipulated variables.

In the process is no alternative heat integrated available in this monoisopropylamine process. Therefore, adding more heat exchanger networks results as increasing operating costs to power the run of the refrigeration system.

Table 5.4 Thermal data of an auto-refrigerated mono-isopropylamine process for pinch analysis

Stream Name	Start temperature T_s(°F)	Target temperature T_T(°F)	W (KW/°F)	Duty (MW)
H1: Outlet stream of reactor	349	200	2.732	1.3149
C1: Total feed stream (IPA+NH_3)	120	283	3.920	1.729

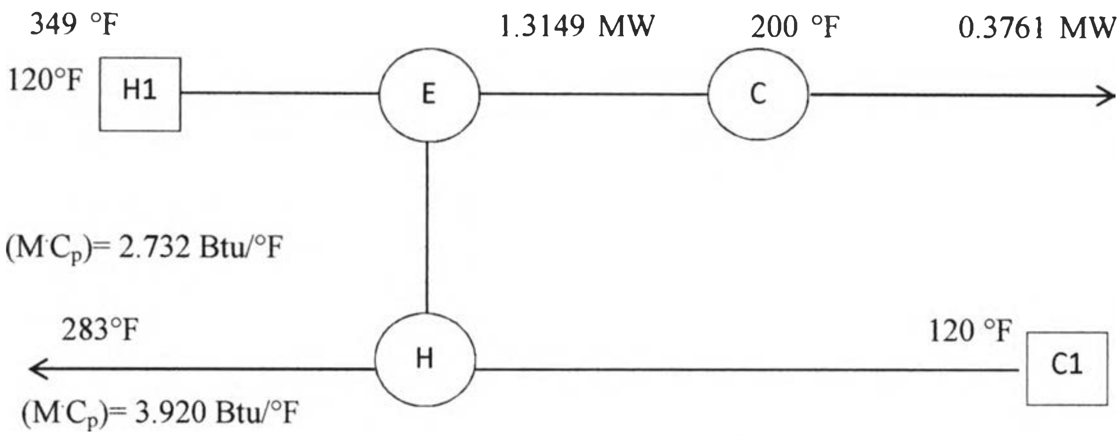


Figure 5.11 shown the thermal data of an auto-refrigerated mono-isopropylamine process for pinch analysi

Step 7: **Optimize economics and/or improve control performance. For example, the controls scheme/structure of the reactor (e.g. temperature/composition sensor location), the control scheme of the distillation column (e.g. reflux to feed ratio control), the optimal operating temperatures of the reactors, the recycle flow rates, the sequence of separation, etc. If the opportunity of optimization exists, we might backtrack to the previous step as dictated.**

The flow disturbances entered column C1, C2, and C3 are fed to reflux flow controllers to adjust by reflux rates. Alternatively, the reflux ratio is used to adjust the reflux of column C3 since it reflux ratio is greater than 4.

The temperature disturbance is minimal resulted from our thermal disturbance handling in Step 4.1, the alternative of rationing the reboiler heat input to the feed flow rate with the ratio reset by the temperature controller as discussed in Luyben (2009) will not be relevant here.

That the PID controllers are used control temperature and concentration controls of the three columns. All control structures are established in figure 5.13-5.16, control structure presented by Luyben (Base case), control structure 1-3 (CS1-CS3).

5.2 Design of Plantwide control structures

The new control structures (CS1, CS2, and CS3) design using Wongsri design procedure and tests fresh IPA feed flow, feed temperature and IPA composition changed. The responses of the entire process are shown as a graph plotted versus operates time (15 hours).

In the procedure, three plantwide control structures are designed CS1-CS3. Figure 5.13-5.16 show the control structure of base case (CS0) and designed control structures. The list of all control loops of each control structure are listed in Table 5.5-5.8 below. The difference between CS1 and CS2, the MIPA distillation column C2, concentration controller is designed at distillate which sends the signal to be the set point of the temperature controller by adjusting heat duty reboiler and the column C3 are different are used reflux flow rate manipulating the tray 5 temperature adjust by reflux flow, as concentration controller in tray the column (C3) control by heat duty reboiler. For the difference between CS1 and CS2, at the heat exchanger in which the pathway cool stream outlet controlled temperature. The difference between CS3 and CS3, at heat exchanger is controlled outlet temperature cool stream by hot stream flow pathway.

At the ammonia distillation column (CS1), an average temperature is controlled in column by manipulating reboiler heat input. This all structures are used because of the large temperature change over the column. The temperature on stage 7, 8, and 9 are measured and averaged.

The temperature profile of ammonia distillation column C1 is shown as table A.3b in the Appendix A. The stages in the column, there are large temperature changes from tray to tray are focused. Table A.3b in the Appendix A, the slope of the temperature profile is steepest from trays 7-9. Hence trays 7-9 are selected to be used as the temperature control by manipulating the reboiler heat input in order to maintain the distillate composition. Figure 5.12a, changing the fresh feed of the ammonia distillation column (C1) effect the temperature profile significantly changes around steep slope showing the strictly control must locate around here. The changing temperature profile when isopropyl alcohol (IPA) feed composition change, figure 5.12b, same as the changing of the fresh feed. Figure 5.12c, feed temperature change highly effect the temperature profile effect the temperature profile significantly changes around steep slope showing the strictly control must locate around here. Increase in feed temperature causes the temperature profile increase.

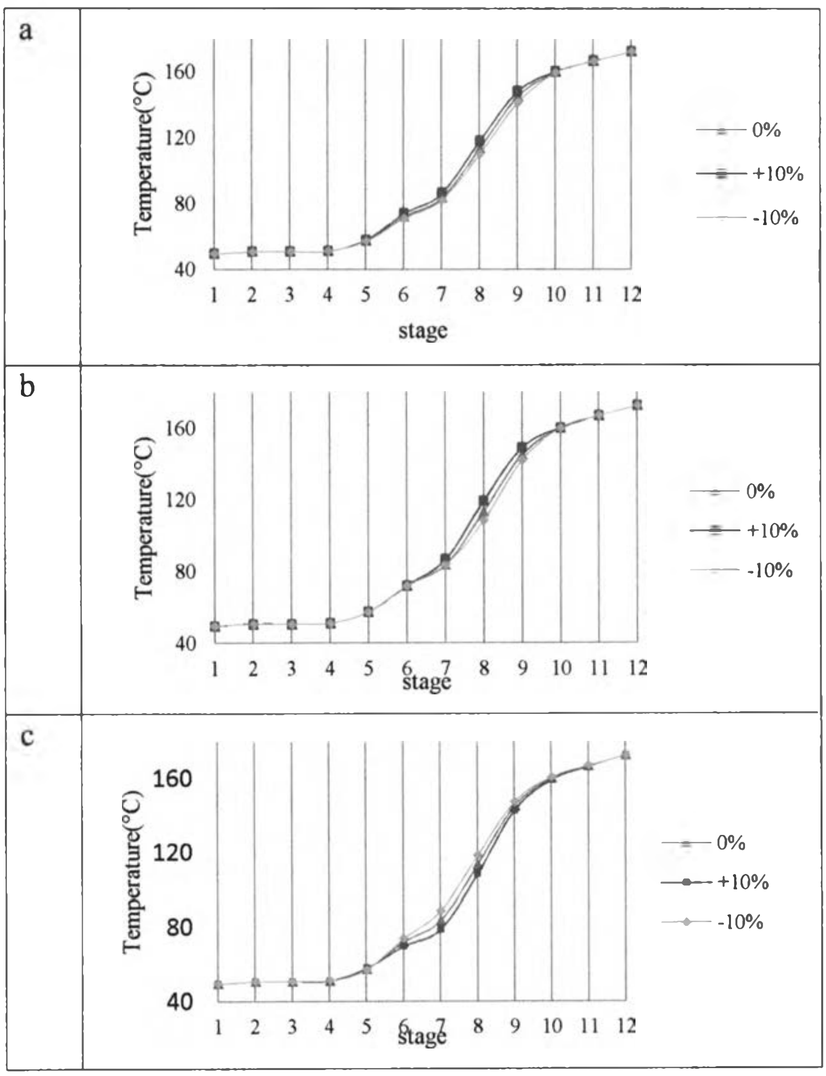


Figure 5.12 temperature profile of column C1 when a.) Total feed, b.) Feed composition and c.) Temperature change.

The temperature profile of product distillation column C2 is shown as table A.3d in the Appendix A. The slope of the temperature profile is steepest from tray 9-10. Tray 9 are selected to control by adjust the reboiler duty for keeping the composition of bottom product constant.

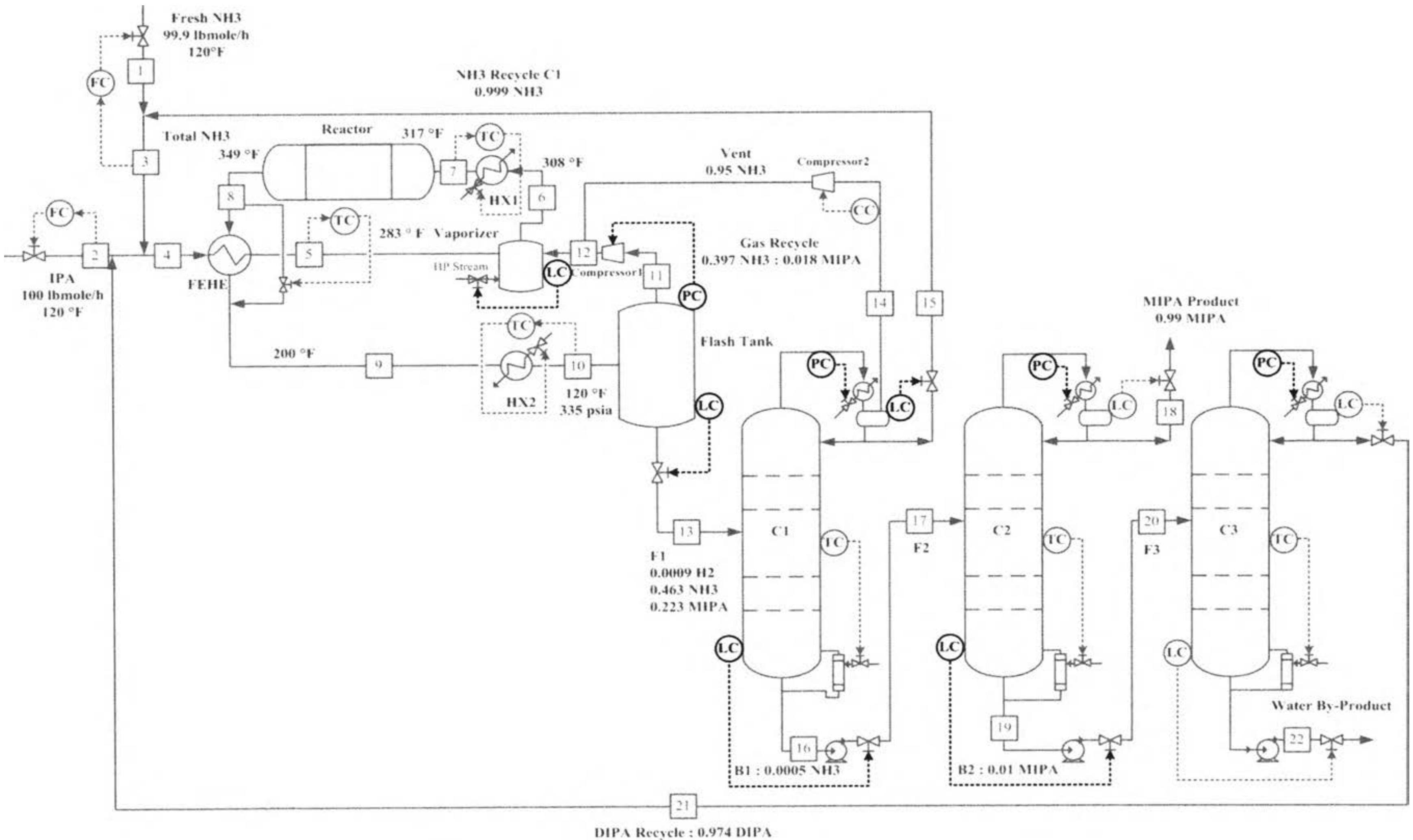


Figure 5.13 Shown the control loops for the remaining control variables

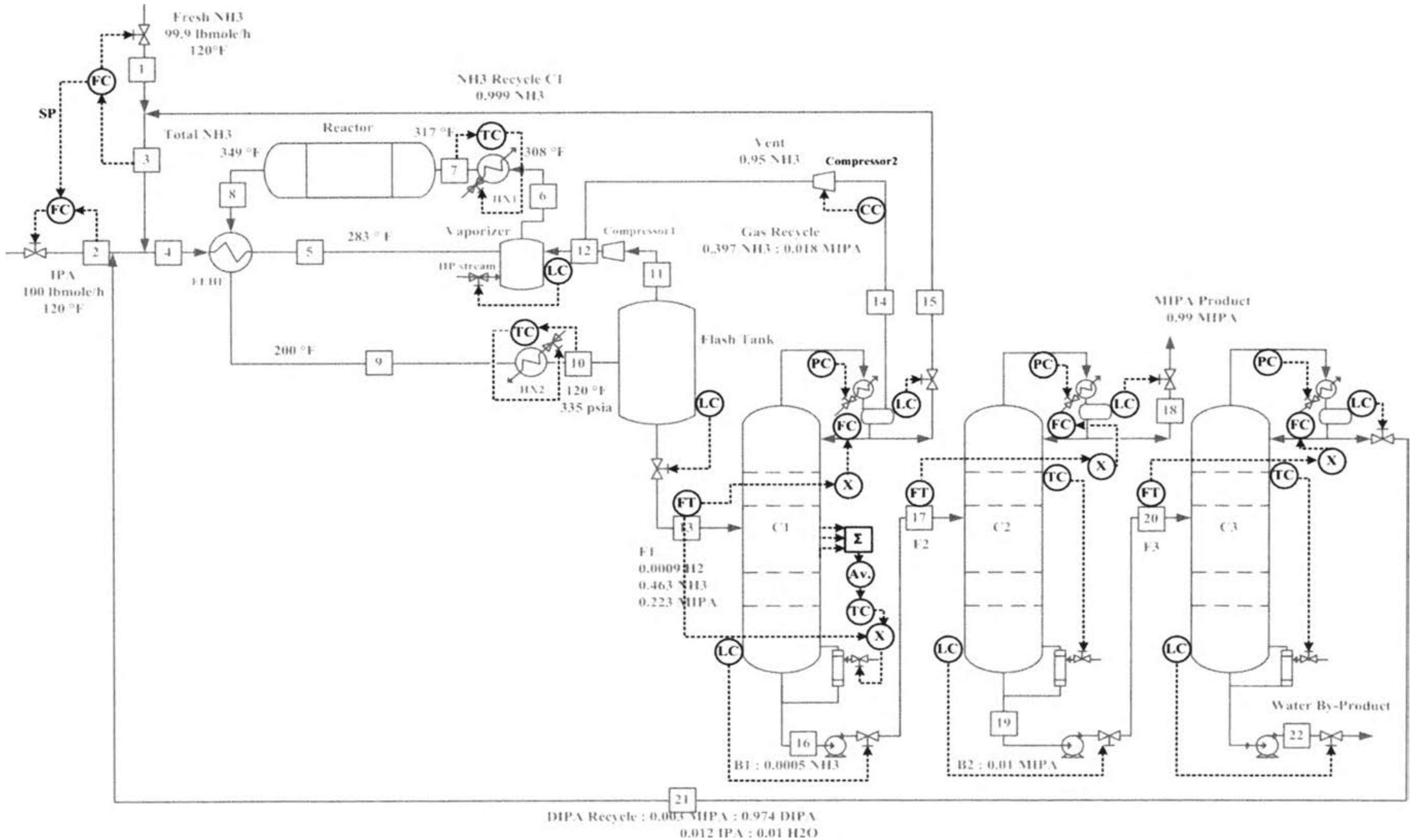


Figure 5.14 shown the control structure base case

Table 5.5 Control structure lists of base case

Equipment	Control	CV	MV	Action
Total Ammonia feed	FC	Ammonia/IPA ratio Flow rate	Feed flow rate	Reverse
IPA fresh feed	FC	Flow rate	Feed flow rate	Reverse
Vaporizer	LC	Level	Duty flow rate	Direct
Heater (HX1)	TC	Temperature outlet	Heater duty	Reverse
Cooler (HX2)	TC	Temperature outlet	Cooler duty	Reverse
Flash tank	PC	Pressure	Compressor work	Reverse
	LC	Level	Duty flow rate	Direct
Column (C1)	TC	Temperature tray 7,8,9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C2)	TC	Temperature tray 9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C3)	TC	Temperature tray 5	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse

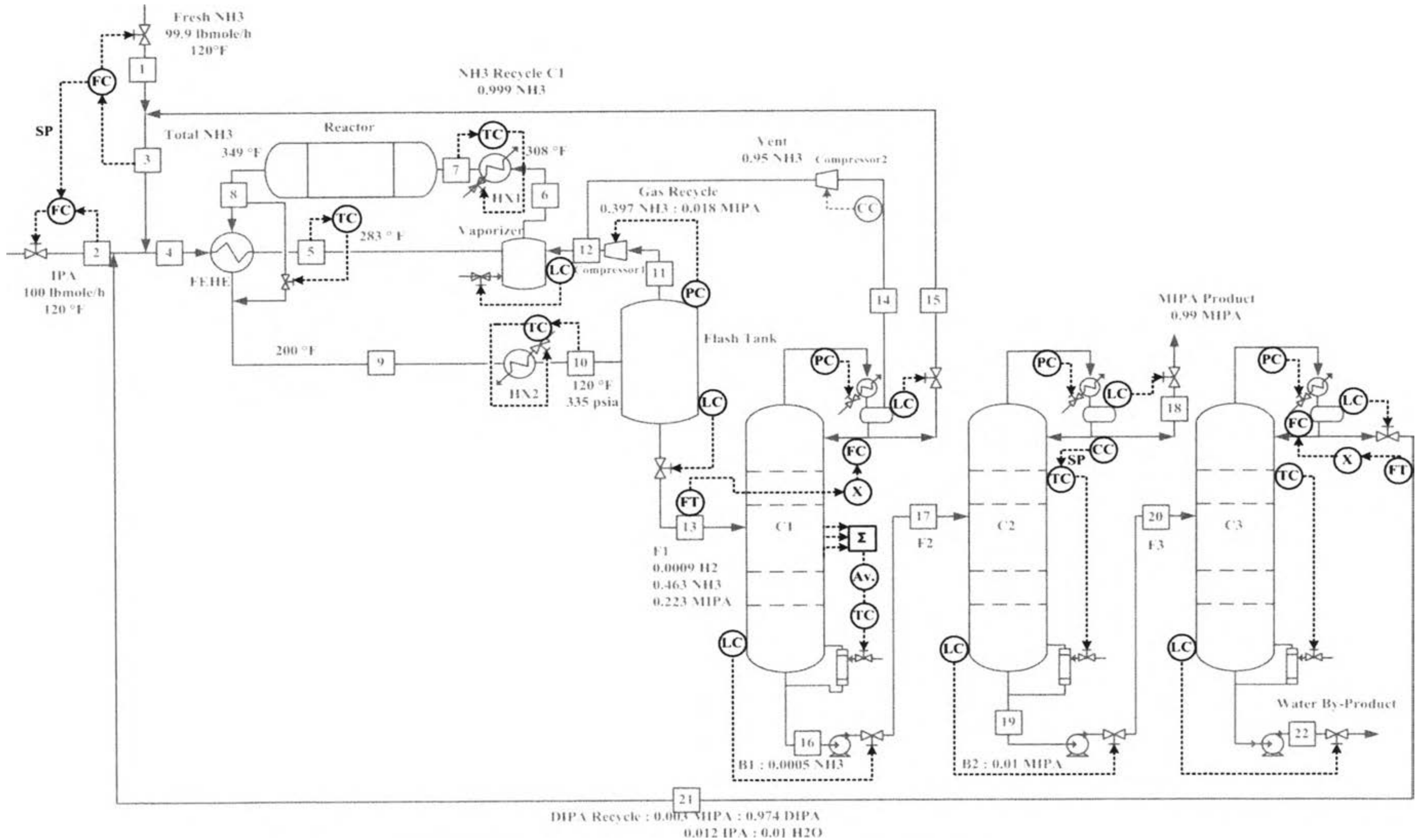


Figure 5.15 shown the control structure CS1 case

Table 5.6 Control structure lists of CS1

Equipment	Control	CV	MV	Action
Total NH_3 feed	FC	Ammonia/IPA ratio Flow rate	Feed flow rate	Reverse
IPA fresh feed	FC	Flow rate	Feed flow rate	Reverse
Vaporizer	LC	Level	Duty flow rate	Direct
Heater (HX1)	TC	Temperature outlet	Heater duty	Reverse
Cooler (HX2)	TC	Temperature outlet	Cooler duty	Reverse
Heat exchanger By-pass	FC	Temperature outlet	By-pass flow rate	Reverse
Flash tank	PC	Pressure	Compressor work	Reverse
	LC	Level	Duty flow rate	Direct
Column (C1)	TC	Temperature tray 7,8,9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C2)	TC	Temperature tray 9 set point CC(MIPA)	Reboiler duty	Reverse
	CC	Composition MIPA	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C3)	TC	Temperature tray 5	Reboiler duty	Reverse
	FC	Reflux ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse

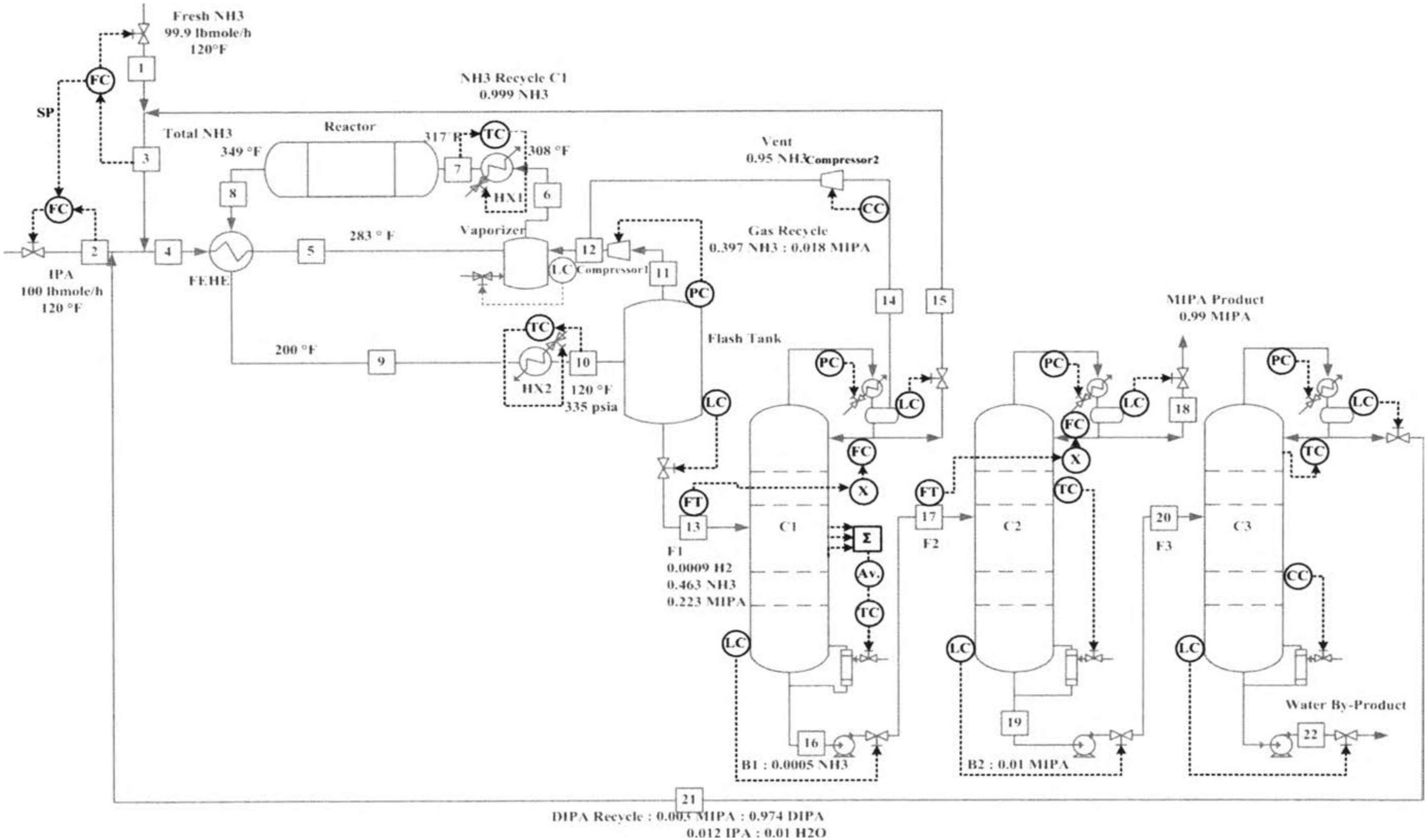


Figure 5.16 shown the control structure CS2 case

Table 5.7 Control structure lists of CS2

Equipment	Control	CV	MV	Action
Total NH_3 feed	FC	Ammonia/IPA ratio Flow rate	Feed flow rate	Reverse
IPA fresh feed	FC	Flow rate	Feed flow rate	Reverse
Vaporizer	LC	Level	Duty flow rate	Direct
Heater (HX1)	TC	Temperature outlet	Heater duty	Reverse
Cooler (HX2)	TC	Temperature outlet	Cooler duty	Reverse
Flash tank	PC	Pressure	Compressor work	Reverse
	LC	Level	Duty flow rate	Direct
Column (C1)	TC	Temperature tray 7,8,9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C2)	TC	Temperature tray 9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C3)	TC	Temperature tray 5	Reflux flow rate	Reverse
	CC	Composition DIPA	Reboiler duty	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse

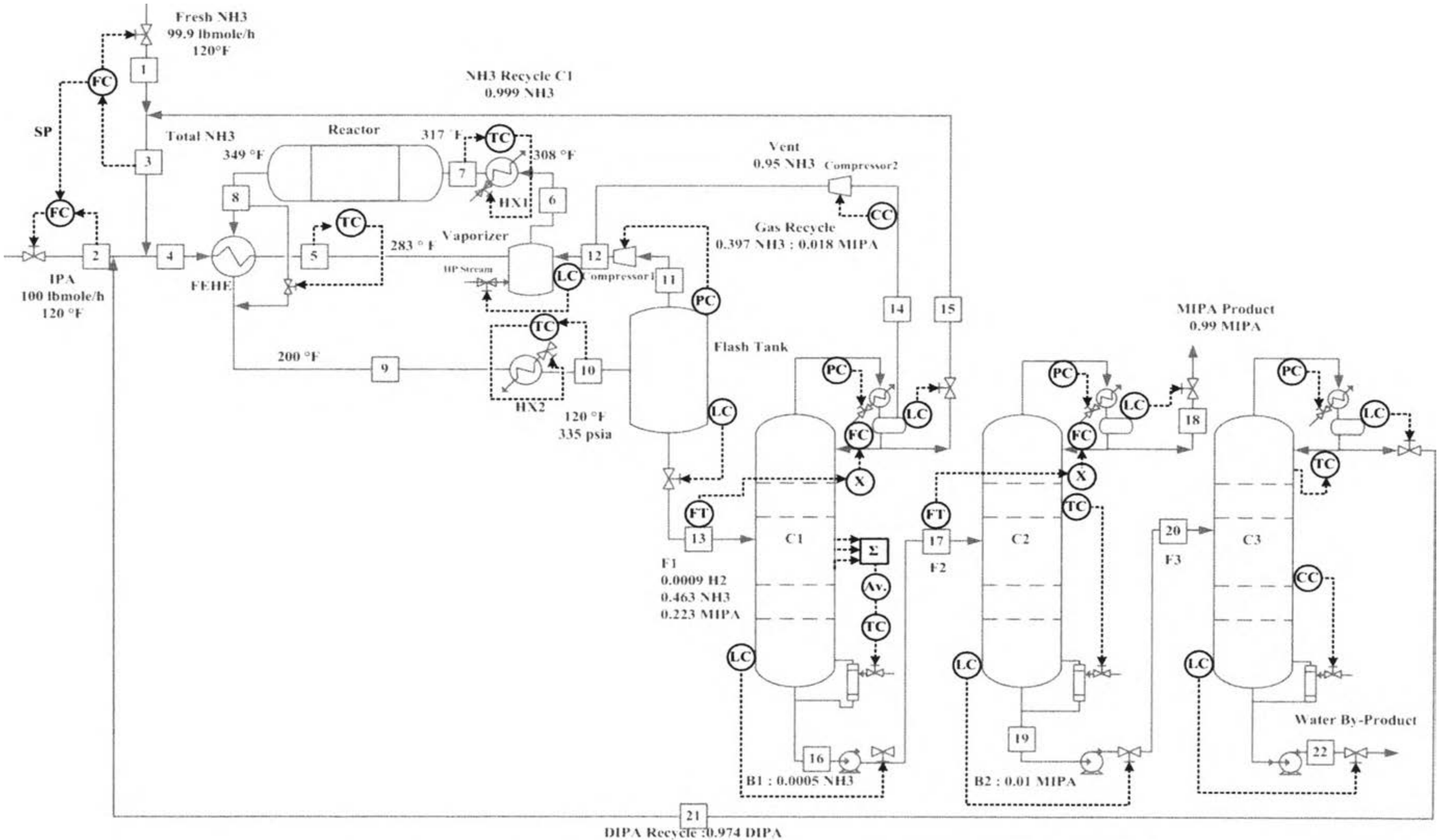


Figure 5.17 shown the control structure CS3 case

Table 5.8 Control structure lists of CS3

Equipment	Control	CV	MV	Action
Total Ammonia feed	FC	Ammonia/IPA ratio Flow rate	Feed flow rate	Reverse
IPA fresh feed	FC	Flow rate	Feed flow rate	Reverse
Vaporizer	LC	Level	Duty flow rate	Direct
Heater (HX1)	TC	Temperature outlet	Heater duty	Reverse
Heat exchanger By-pass	FC	Temperature outlet	By-pass flow rate	Reverse
Cooler (HX2)	TC	Temperature outlet	Cooler duty	Reverse
Flash tank	PC	Pressure	Compressor work	Reverse
	LC	Level	Duty flow rate	Direct
Column (C1)	TC	Temperature tray 7,8,9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C2)	TC	Temperature tray 9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C3)	TC	Temperature tray 5	Reflux flow rate	Reverse
	CC	Composition DIPA	Reboiler duty	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse

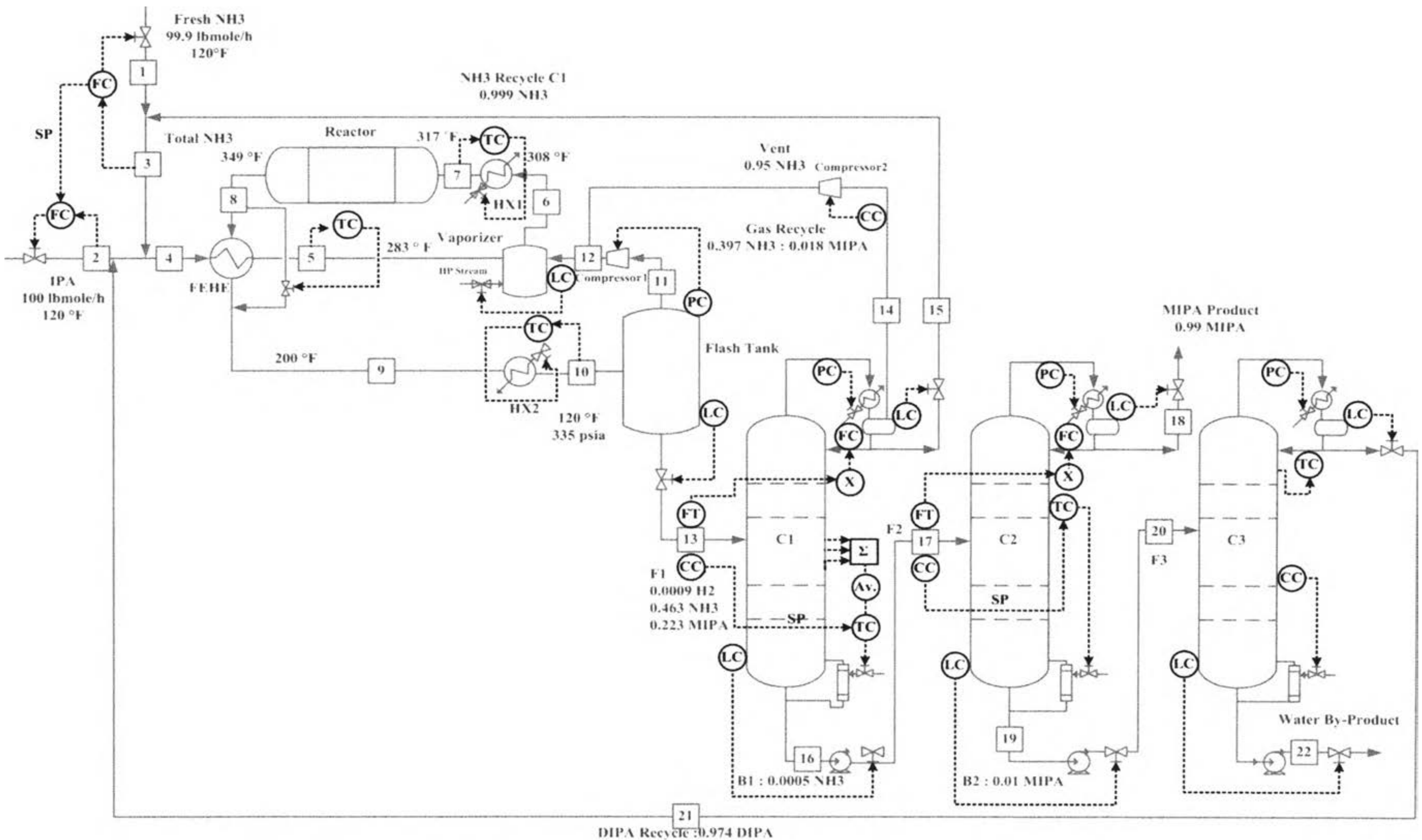


Figure 5.17 shown the control structure CS4 case

Table 5.9 Control structure lists of CS4

Equipment	Control	CV	MV	Action
Total Ammonia feed	FC	Ammonia/IPA ratio Flow rate	Feed flow rate	Reverse
IPA fresh feed	FC	Flow rate	Feed flow rate	Reverse
Vaporizer	LC	Level	Duty flow rate	Direct
Heater (HX1)	TC	Temperature outlet	Heater duty	Reverse
Heat exchanger By-pass	FC	Temperature outlet	By-pass flow rate	Reverse
Cooler (HX2)	TC	Temperature outlet	Cooler duty	Reverse
Flash tank	PC	Pressure	Compressor work	Reverse
	LC	Level	Duty flow rate	Direct
Column (C1)	TC	Temperature tray 7,8,9	Reboiler duty	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	CC	Composition NH_3	Cascasd TC @ C1	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C2)	TC	Temperature tray 9	Reboiler duty	Reverse
	CC	Composition MIPA	Cascasd TC @ C2	Reverse
	FC	Reflux/feed ratio	Reflux flow rate	Reverse
	LC	Reboiler level	Bottom flow rate	Direct
	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse
Column (C3)	TC	Temperature tray 5	Reflux flow rate	Reverse
	CC	Composition DIPA	Reboiler duty	Reverse
	LC	Reboiler level	Bottom flow rate	Direct

Table 5.9 (continue) Control structure lists of CS4

Equipment	Control	CV	MV	Action
Column (C3)	LC	Reflux drum level	Distillate flow rate	Direct
	PC	Top column	Condenser duty	Reverse

Step 8: **Validate the designed control structures by rigorous dynamic simulation. The measures can be costs, raw material and energy consumptions, control performances of the total plant or some selected loops, etc.**

The commercial process simulator is used for obtaining the responses of the variables when the disturbances occur, isopropyl alcohol fresh feed flow, feed temperature and composition change. The responses are collected and plotted as graph versus time.

Figure 5.14-5.17 gives the basecase plantwide control structure are developed. Inlet temperature of reactor is controlled by heat duty of heater (HX1). The new control structure (CS1 and CS3) are discussed in heat disturbance, designed temperature control loop at vaporizer inlet temperature by manipulating the by-pass flow rate around the FEHE (heat exchange) as shown Fig 5.14 and 5.16. In CS2 and CS3, the MIPA distillation column C2, concentration controller is designed at distillate which sends the signal to be the set point of the temperature controller by adjusting heat duty reboiler. In the recycle distillation column (C3), the temperature on tray 5 is controlled by manipulating by reflux flowrate and concentration DIPA on tray 31 is controlled by reboiler duty. We could try to directly control the DIPA concentration in the bottoms, but this is the high-purity distillation column. The desired bottoms impurity is 0.1 mole% DIPA. The response of the bottom composition would be very nonlinear and present difficult control problems.

5.3 Dynamic simulation results.

An analysis the dynamic of base on control structure by Luyben (2009) and all designed control structures are using new design procedure of Wongsri (2012). The material disturbance (flow rate and composition) and thermal disturbance are tested the response of the system at dynamic mode, the performance of each structure. Temperature and composition controllers are used PID control by relay-feedback test. The temperature loops have 1 min dead time and the composition loop have 3 min dead times. Flow and pressure controllers are PI control. Level loops are proportional-only control. There parameters are heuristics values. At the nominal operating condition, the control valve is half-opened. The dynamic results are explained in this part.

5.3.1 Changes in material disturbances of the isopropyl alcohol (IPA) flow rate for all control structures (base case (CS0), designed control structure (CS1-CS4)

Table 5.10 shows dynamic response of the monoisopropylamine process by step change molar flow rate ± 20% in the isopropyl alcohol increase from 45.36 to 45.45 kgmol/hr. at 15 hr., decrease from 45.36 to 45.27 kgmol/hr. at 15 hr. The graph of dynamic responses of base case control structure (CS0), designed control structures (CS1-CS3) in each section.

Table 5.10 shows responses of variables disturbed by fresh feed is changing in base case CS0, CS1, CS2, CS3, and CS4. There are gradually increased rapidly changes because of the higher flow control time constant.

The inlet temperature of reactor can handled the disturbances after operating times of 3 hours for all control structures. The monoisopropylamine product composition and flowrate at the product distillation column C2 are controlled after 5 hours operating times. For the ammonia distillation column (C1), the impurity of ammonia is maintained at the set point for all control structures since the tray temperature is well controlled at set point by the reboiler duty. The ammonia flowrate changes directly when the fresh feed flow rate is changed. At the product distillation column (C2), the composition of IPA is also kept at the set point as and the monoisopropylamine product changes in the same way of the fresh feed flowrate change. The DIPA distillation column (C3), the distillate of column is affected to fresh feed change and the water by-product change in the same way of the fresh feed flowrate change.

5.3.2 Changes in material disturbances of the composition isopropyl alcohol (IPA) and ammonia (NH_3) for all control structures (base case (CS0), designed control structure (CS1-CS4)

5.3.2.1 Decreased IPA in fresh IPA feed.

Composition of isopropyl alcohol (IPA) feed is decreased by 5%mole IPA. The responses of the variables are controlled for considering the effect of composition change, showed in table CS0, CS1, CS2, CS3, and CS4. The test disturbances are concentrating conversation value in process because the conversion of reaction increase effecting the recycle streams. The recycle streams are decreasing makes the total feed control adjust the total flow to achieve set point. The temperature conversion is increased allow the reactor to obtain the limit value.

5.3.2.2 Increased IPA in fresh IPA feed.

Because, that changes are occur affected reaction in reactor and recycle streams. The recycle decreasing causes the fresh feed control adjusts the total flow to reach the set point. The inlet temperature is controlled at the reactor by adjusting the duty HX1. Vaporizer has affected from the reactor and sends effect to purity of product. The flash tank is controlled inlet temperature by heat duty of cooler (HX2). Therefore, there is the composition control of MIPA at condenser in column 2 (CS3) and control temperature in all column adjust by heat duty of reboiler (CS0, CS1, CS2, CS3, and CS4). The graph of dynamic responses of base case control structure (CS0), designed control structures (CS1-CS4) in each section as show in table 5.9.

5.3.3 Changes in material disturbances of the temperature isopropyl alcohol (IPA) feed flow for all control structures (base case (CS0), designed control structure (CS1-CS4)

Testing condition start at 0.5 h by change fresh feed temperature at normal value 60 °C to 65 °C and decrease to 55 °C. The dynamic responses of thermal disturbance are shown in Table 5.8

When the fresh feed temperature change, the reactor and heat exchangers can cope with the disturbance as shown in Table 5.8, and at the both column, the C1 and C2 is less effect to the disturbance so that the purity of the product is maintained at the set point.

The temperature of heat exchanger (FEHE) outlet is deviated from the steady value because there is no control in temperature outlet of base case, CS1, and CS2. The CS4 are can controlled the heat disturbances giving the temperature out of a set point by flow by-path way. By-pass control temperature is cooling flow before fed to vaporizer.

Table 5.10 Dynamic responses with fresh IPA feed changed.

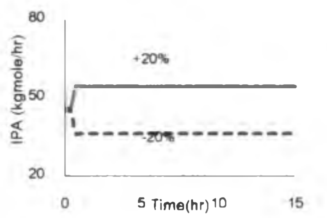
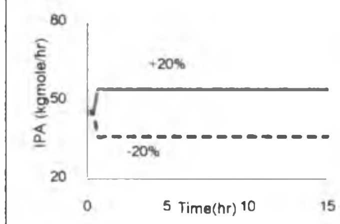
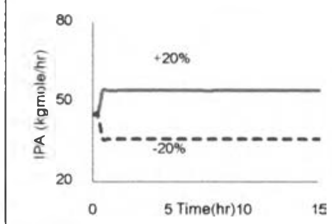
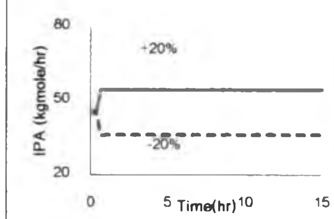
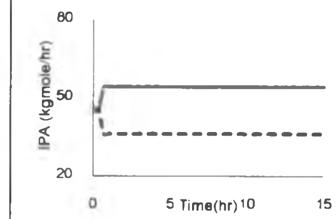
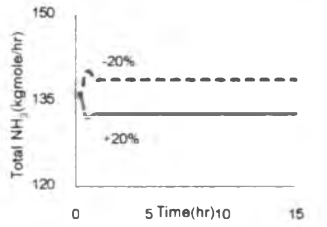
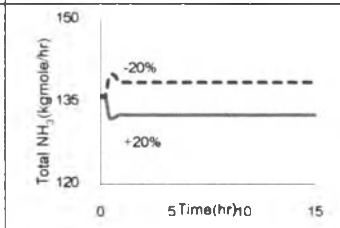
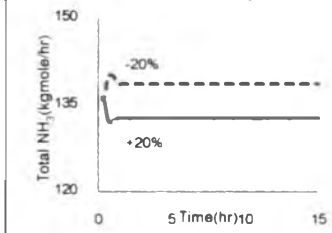
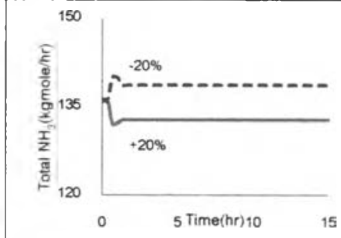
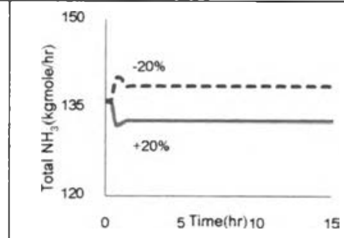
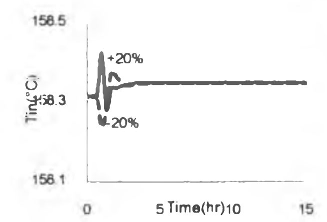
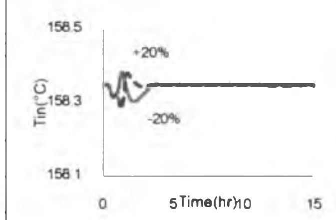
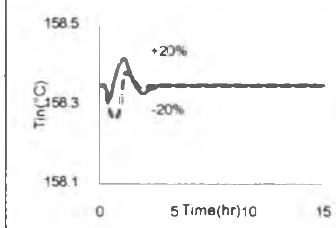
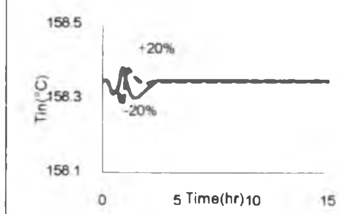
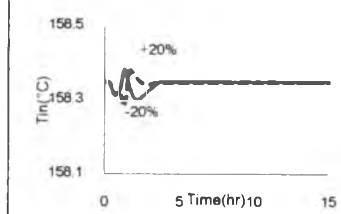
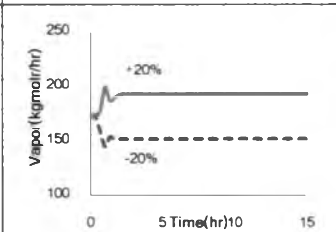
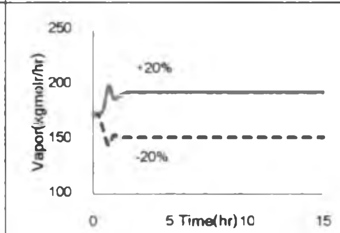
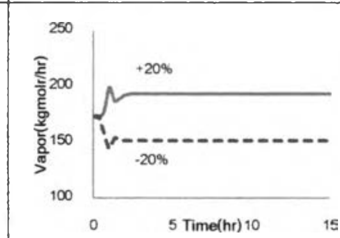
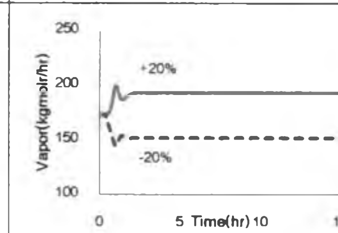
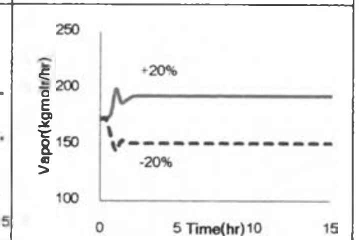
	CS0	CS1	CS2	CS3	CS4
A					
B					
C					
D					

Table 5.10 (Continue) Dynamic responses with fresh IPA feed changed.

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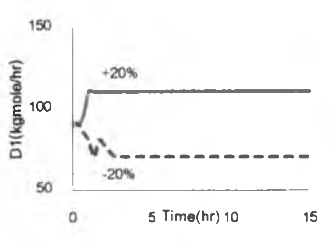
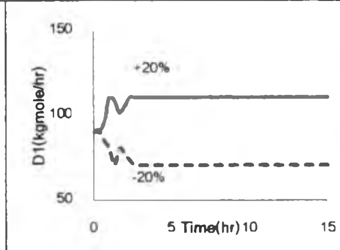
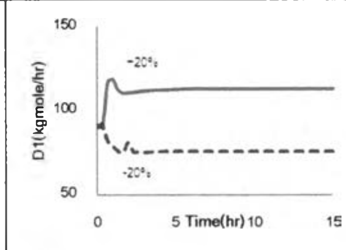
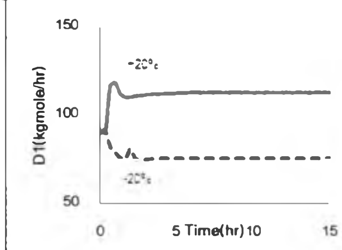
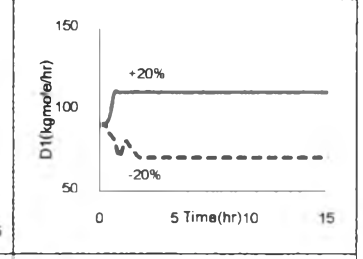
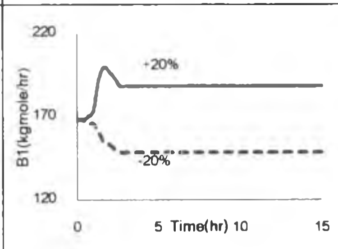
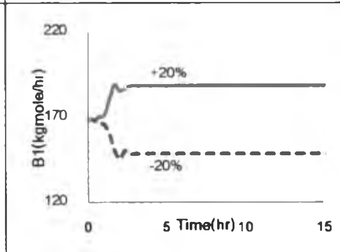
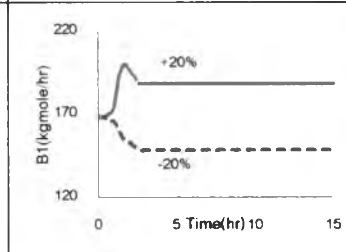
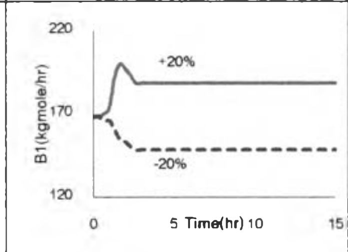
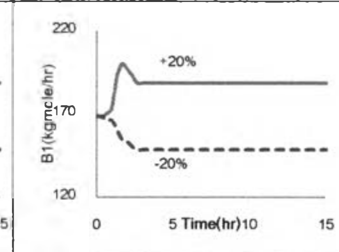
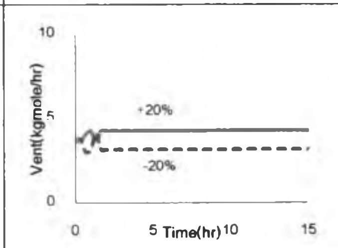
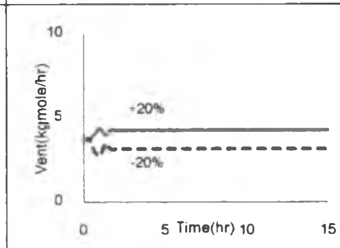
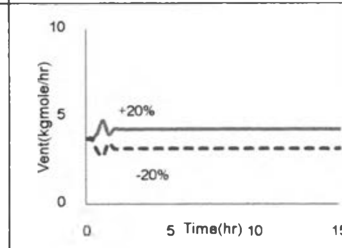
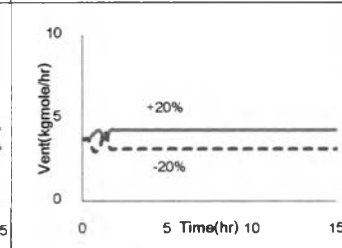
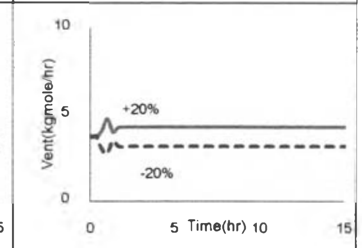
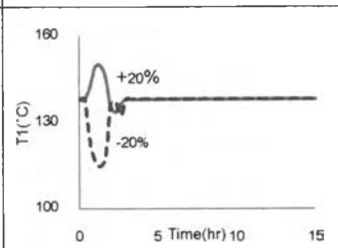
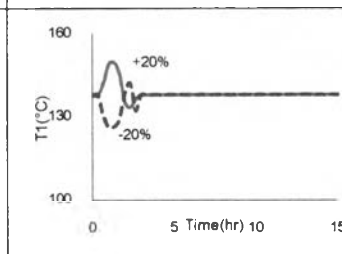
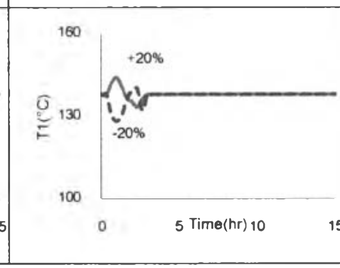
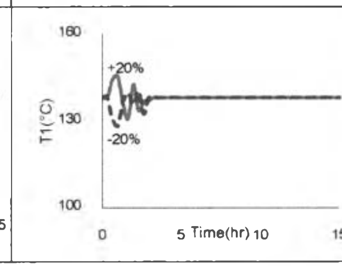
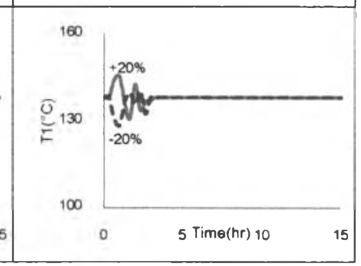
E					
F					
G					
H					

Table 5.10 (continue) Dynamic response with fresh IPA feed changed

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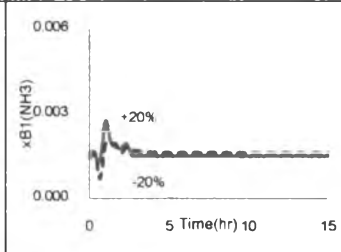
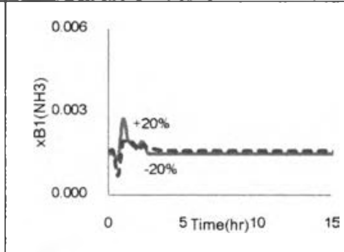
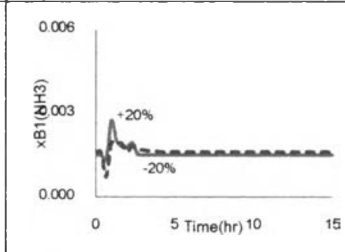
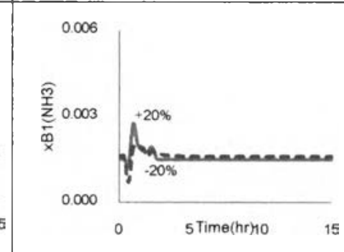
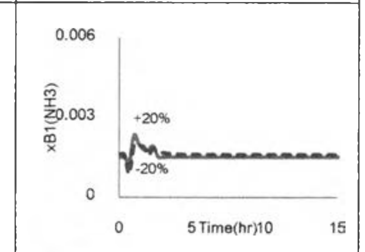
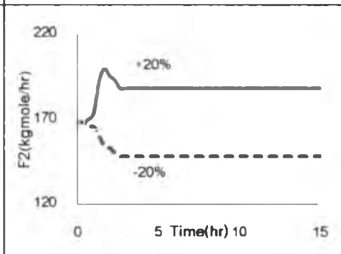
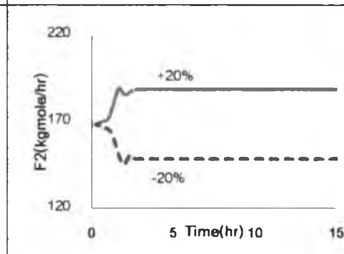
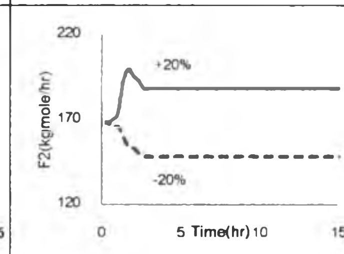
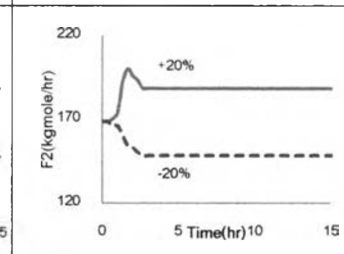
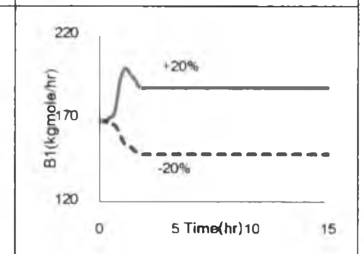
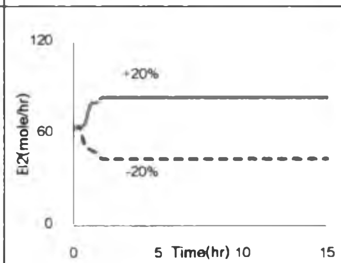
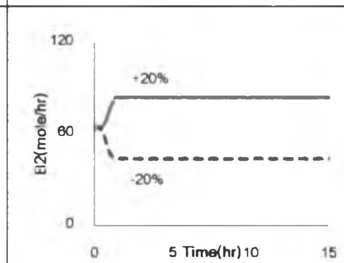
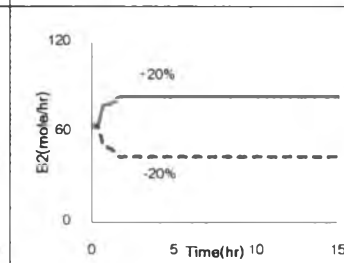
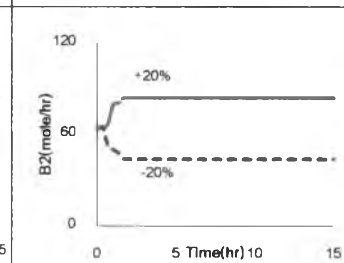
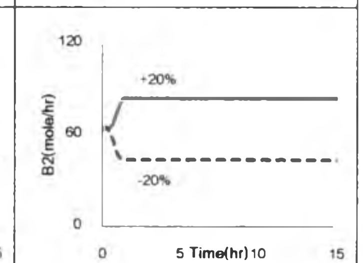
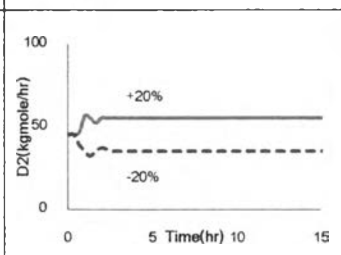
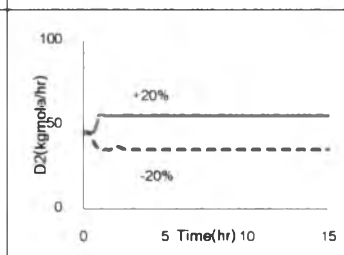
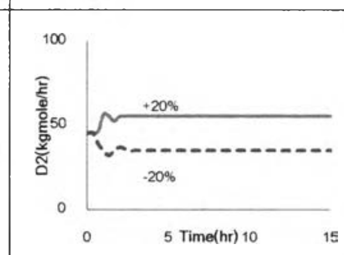
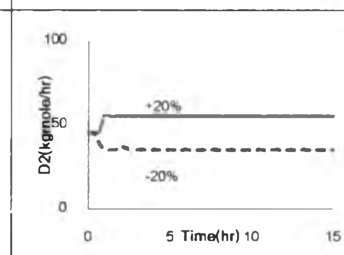
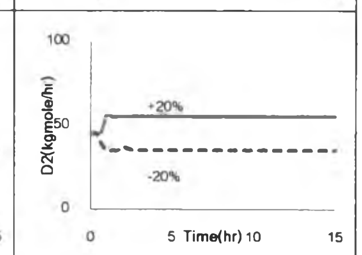
I					
J					
K					
L					

Table 5.10 (continue) Dynamic response with fresh IPA feed changed

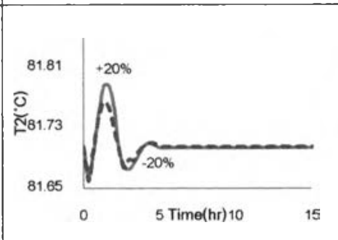
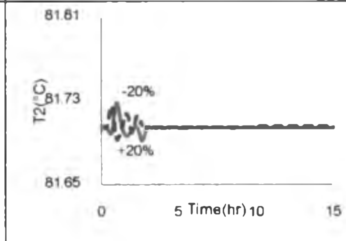
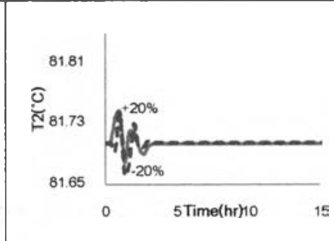
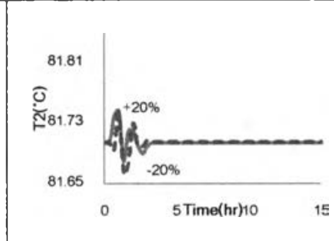
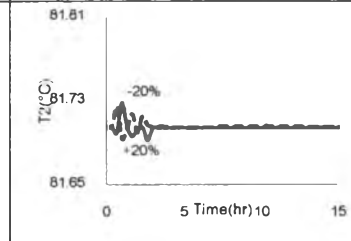
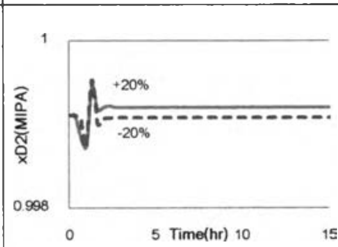
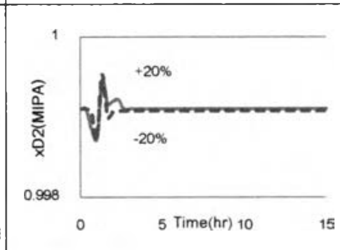
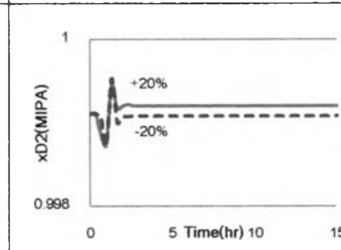
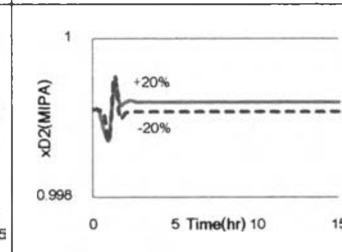
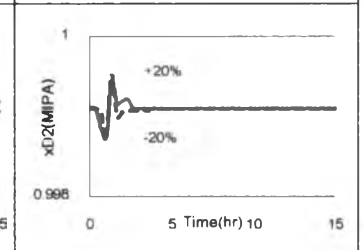
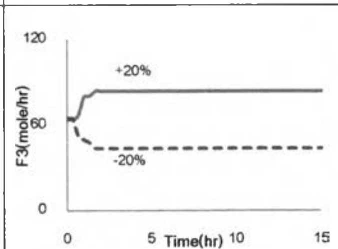
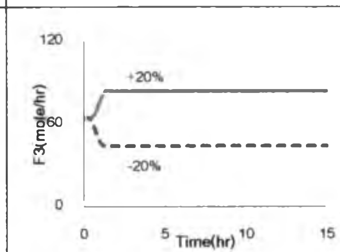
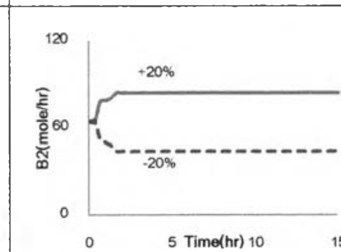
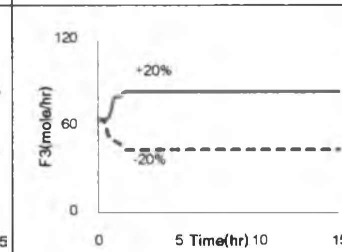
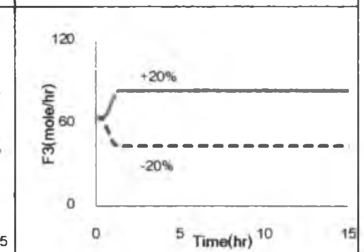
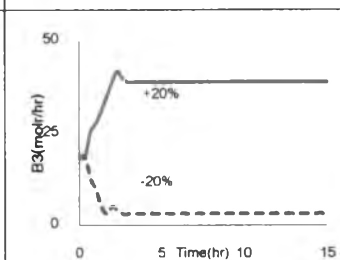
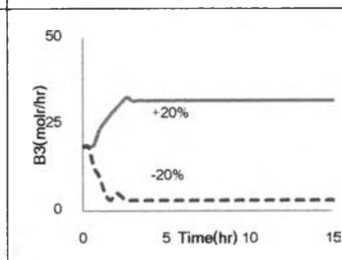
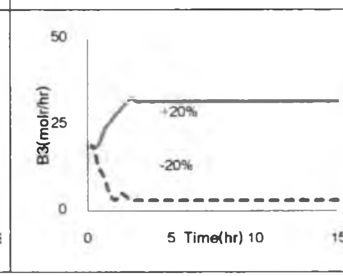
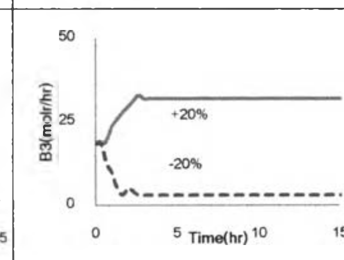
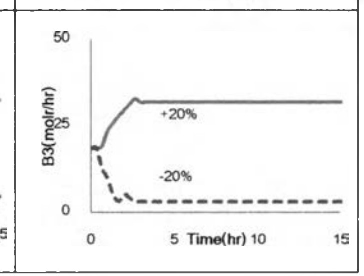
	CS0	CS1	CS2	CS3	CS4
M					
N					
O					
P					

Table 5.11 Dynamic response with fresh IPA feed temperature changed

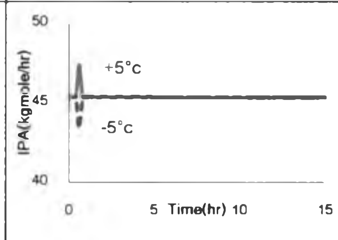
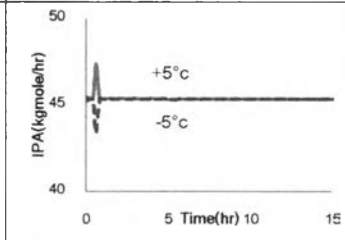
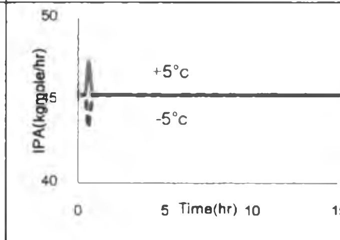
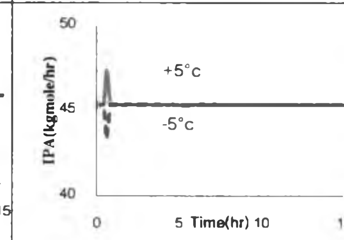
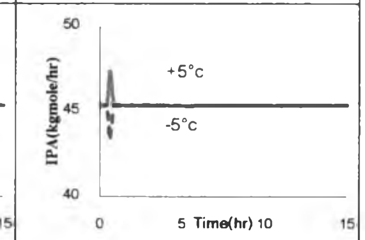
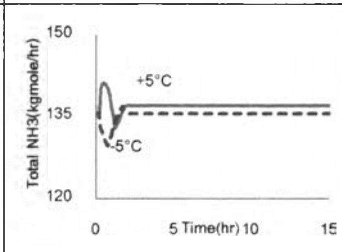
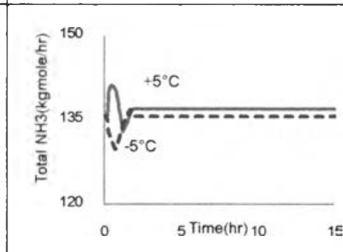
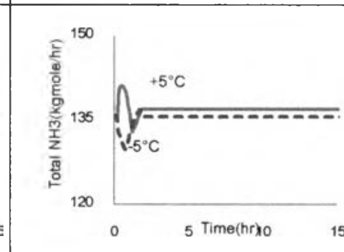
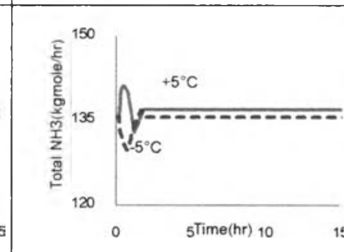
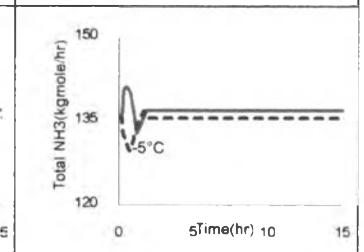
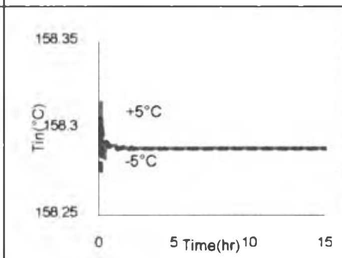
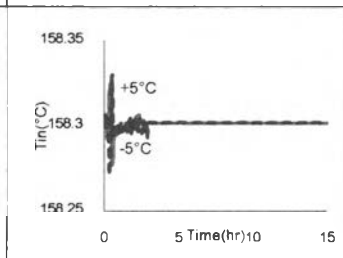
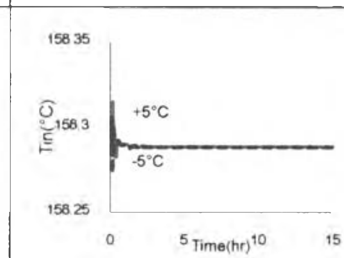
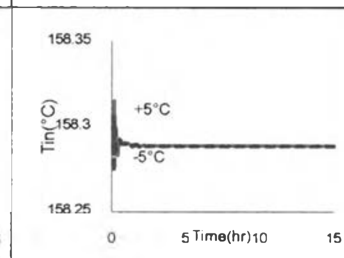
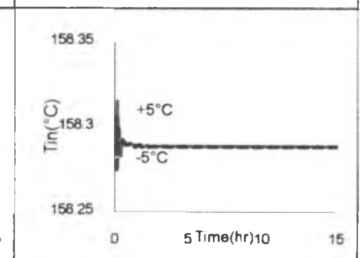
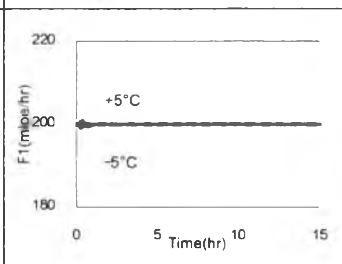
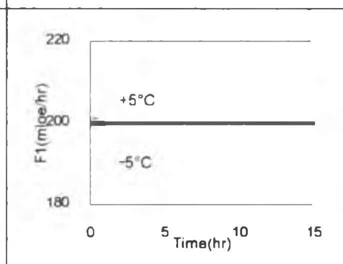
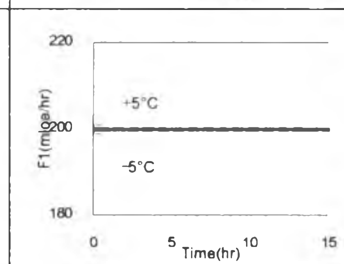
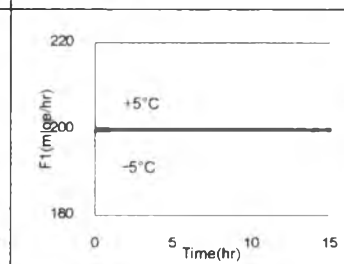
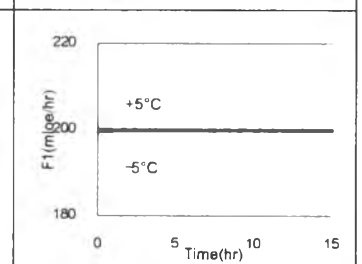
	CS0	CS1	CS2	CS3	CS4
A					
B					
C					
D					

Table 5.11 (continue) Dynamic response with fresh IPA feed temperature changed

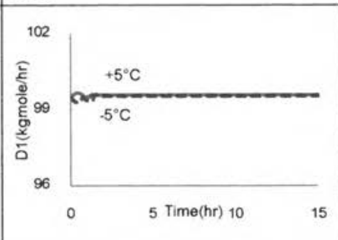
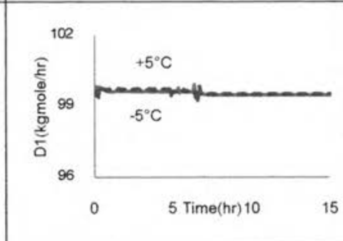
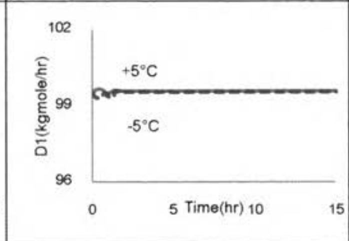
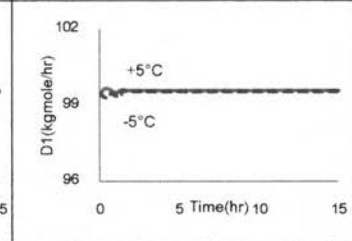
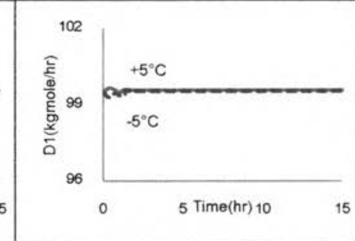
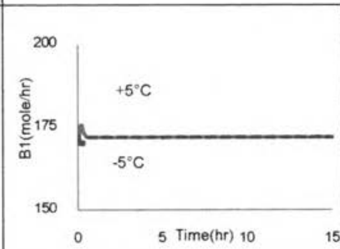
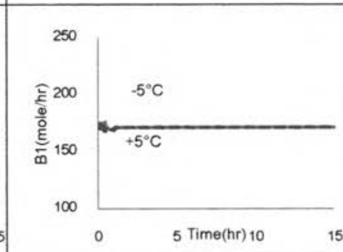
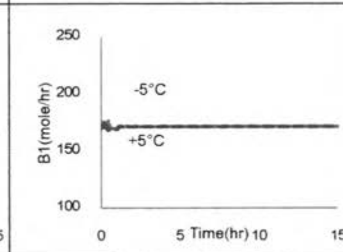
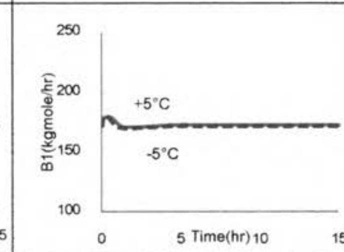
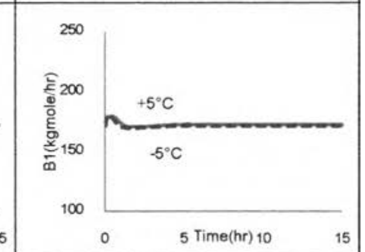
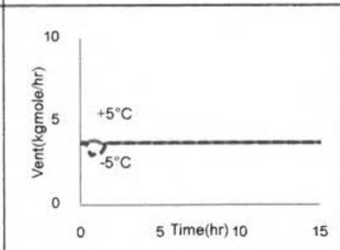
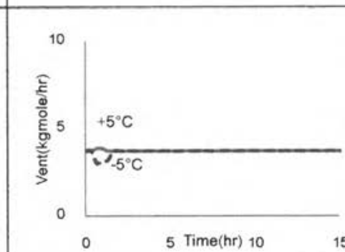
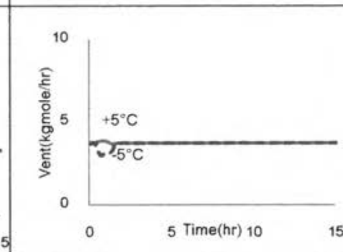
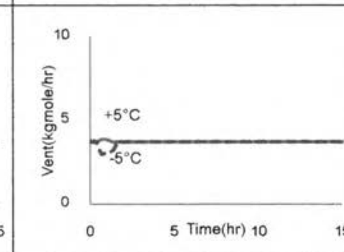
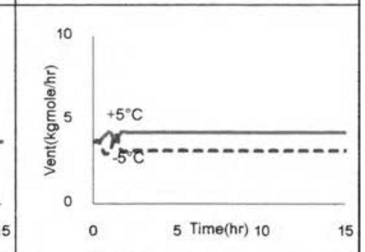
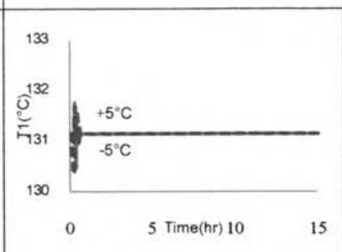
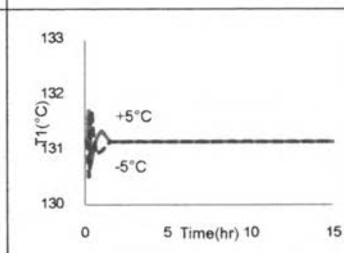
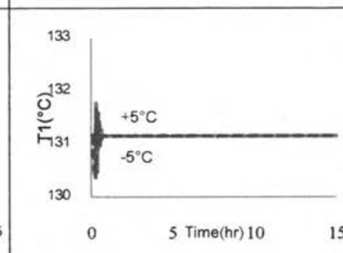
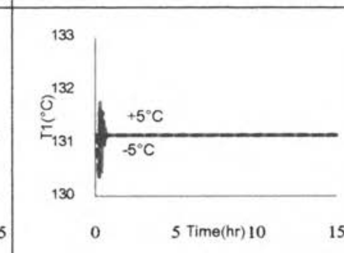
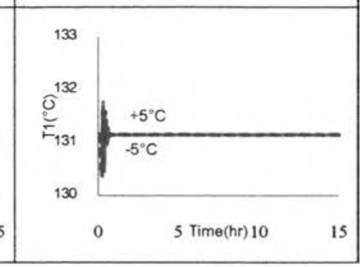
	CS0	CS1	CS2	CS3	CS4
E					
F					
G					
H					

Table 5.11 (continue) Dynamic response with fresh IPA feed temperature changed

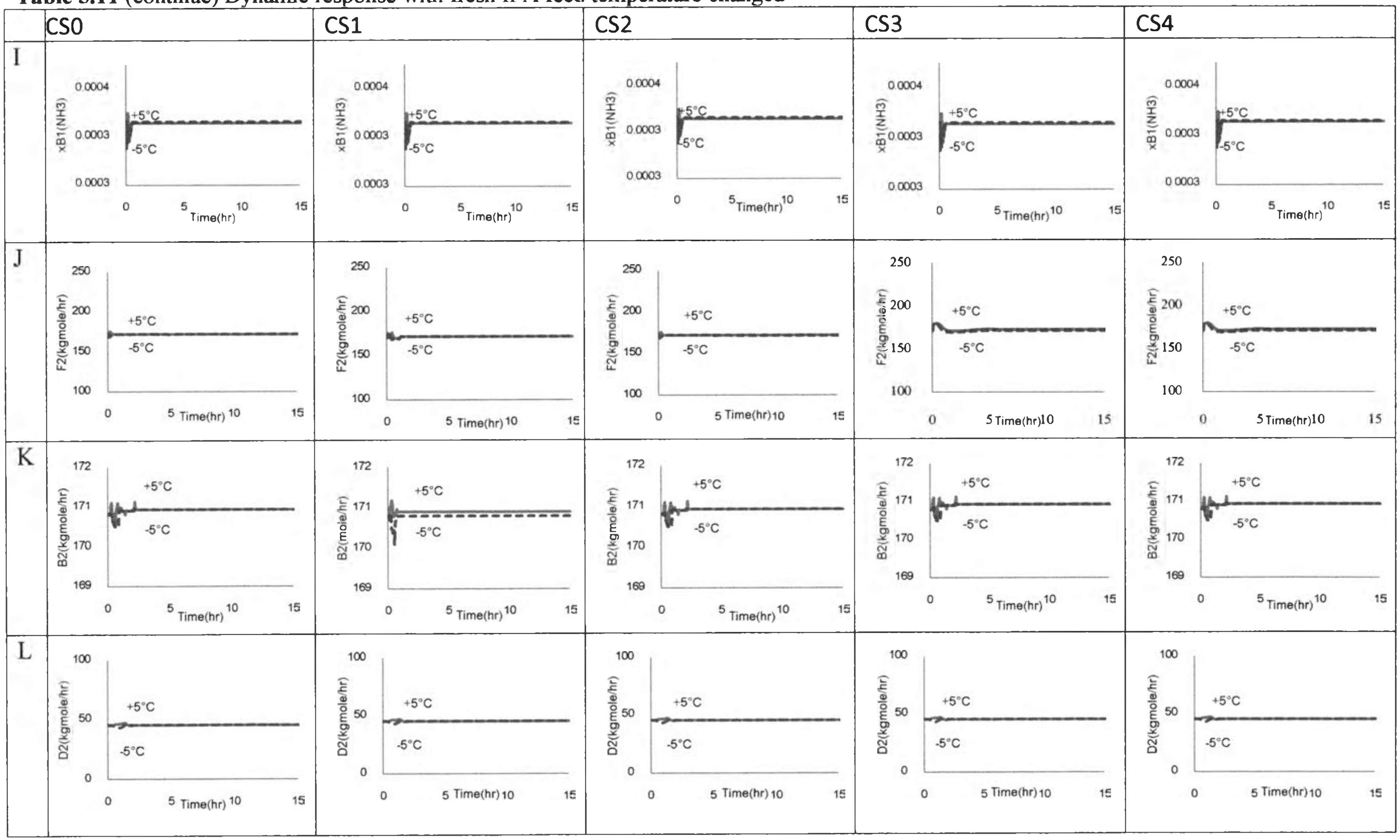


Table 5.11 (continue) Dynamic response with fresh IPA feed temperature changed

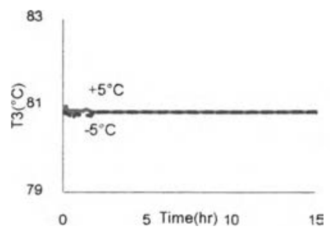
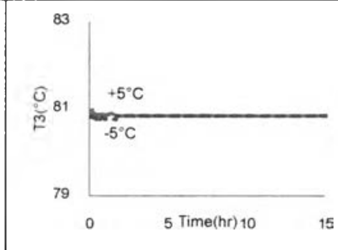
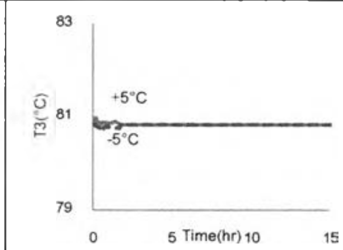
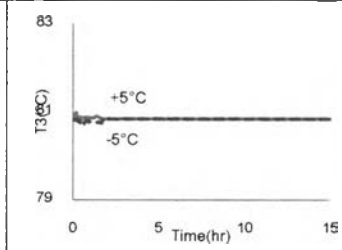
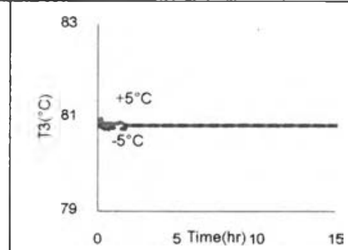
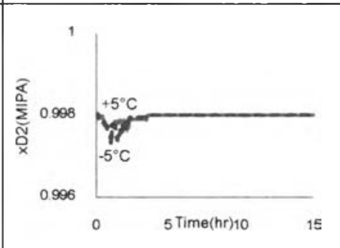
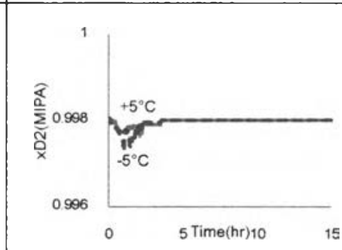
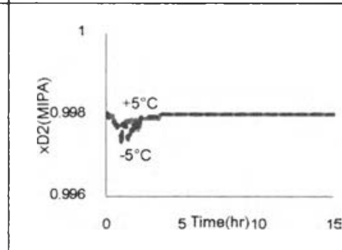
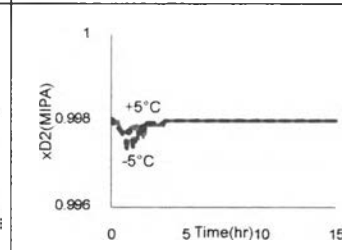
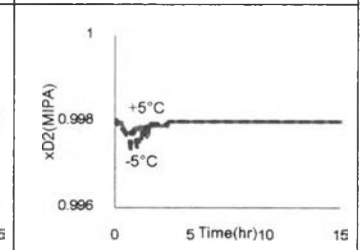
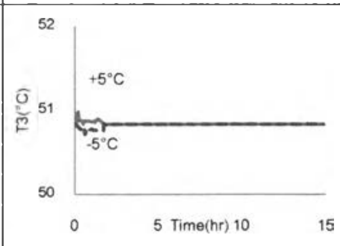
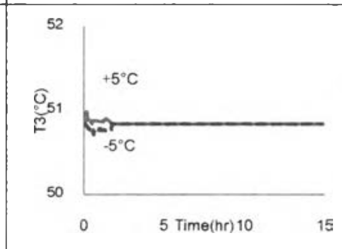
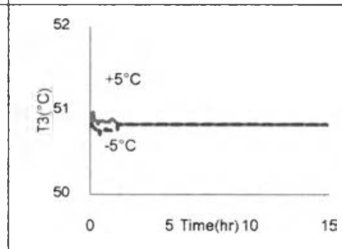
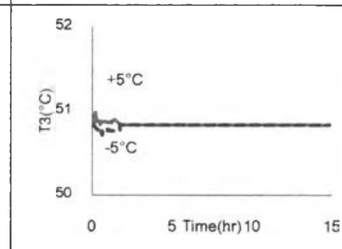
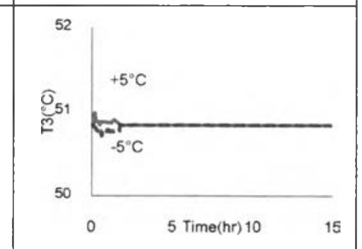
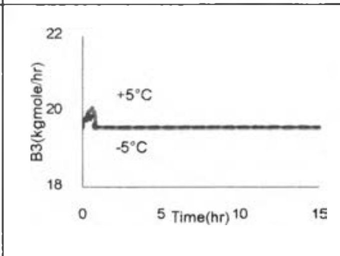
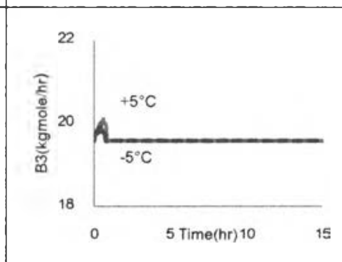
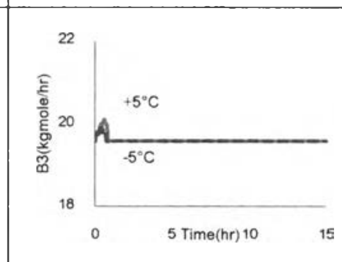
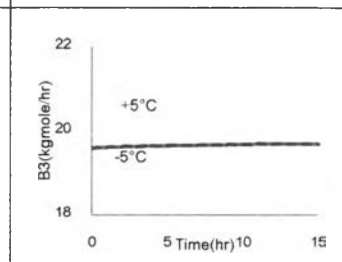
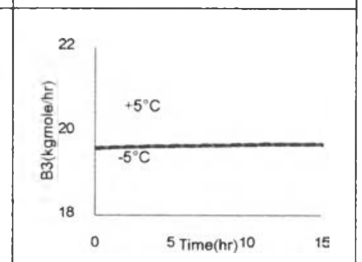
	CS0	CS1	CS2	CS3	CS4
M					
N					
O					
P					

Table 5.12 Dynamic response with total feed composition changed

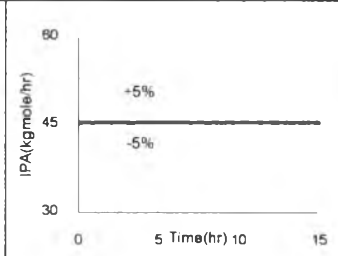
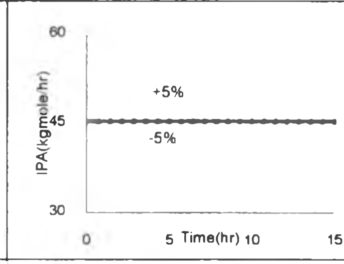
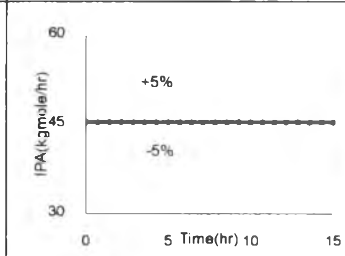
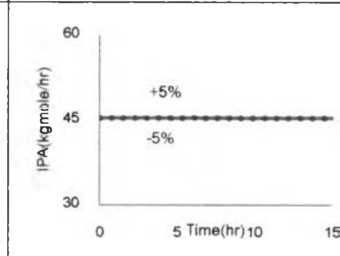
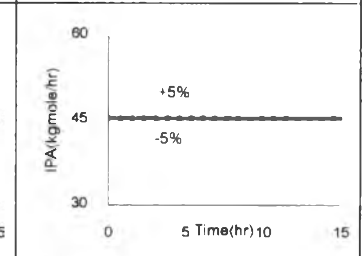
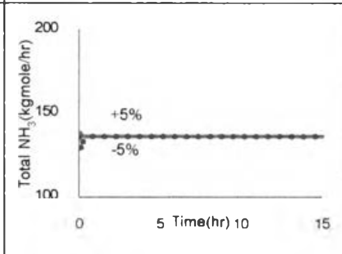
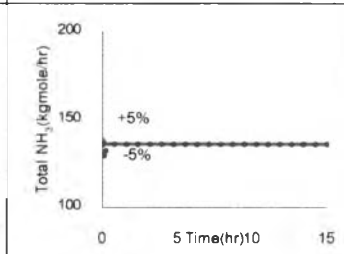
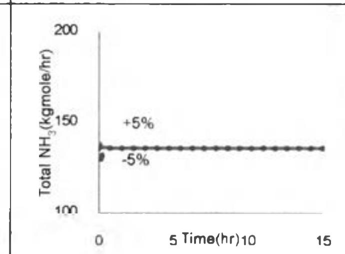
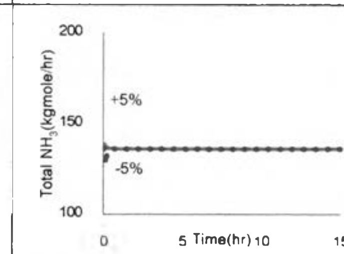
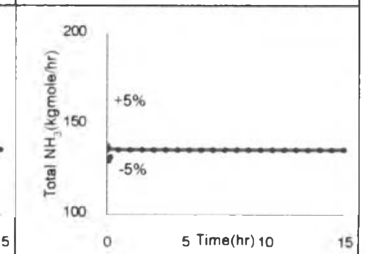
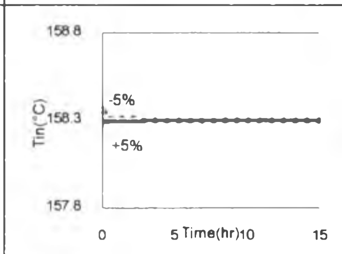
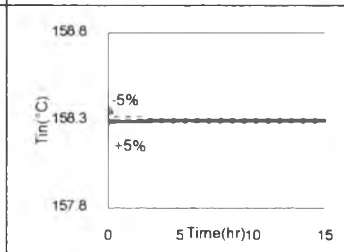
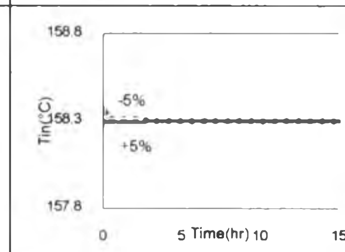
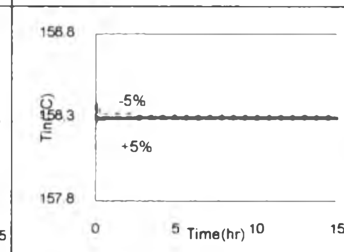
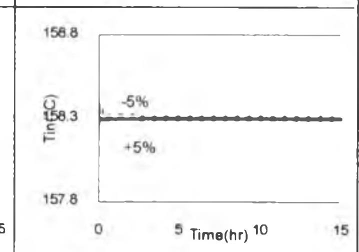
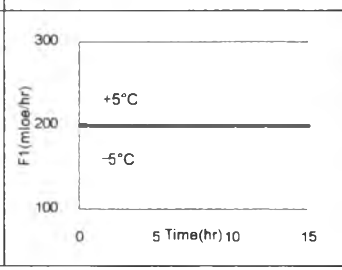
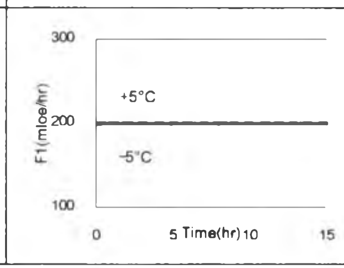
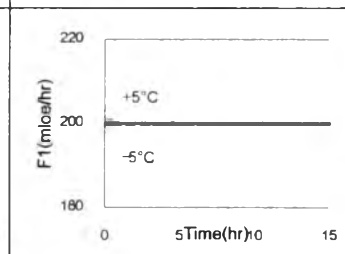
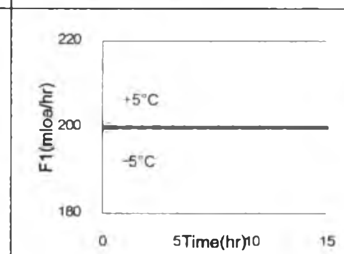
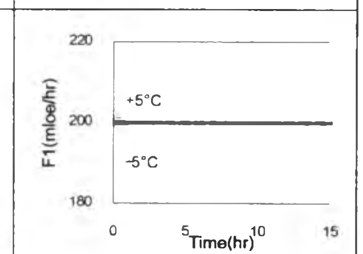
	CS0	CS1	CS2	CS3	CS4
A					
B					
C					
D					

Table 5.12 (continue) Dynamic response with total feed composition changed

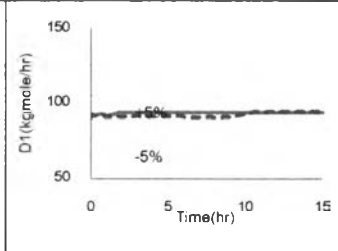
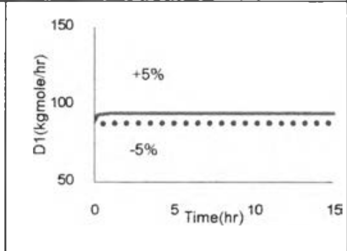
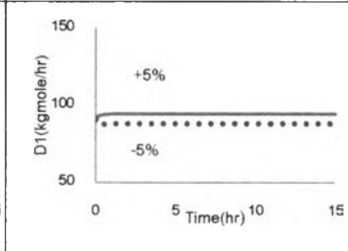
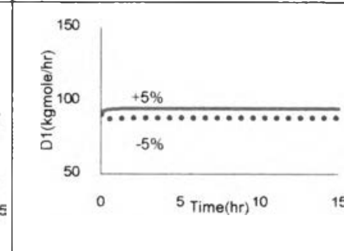
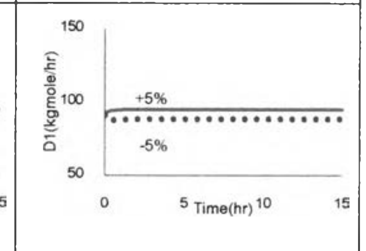
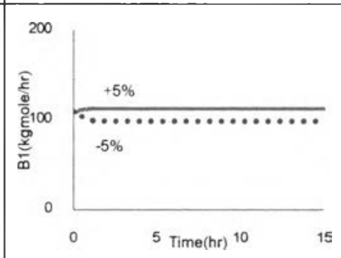
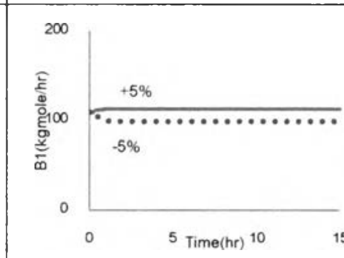
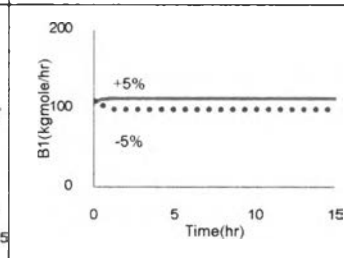
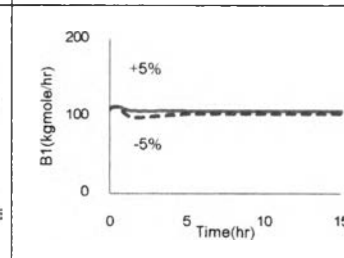
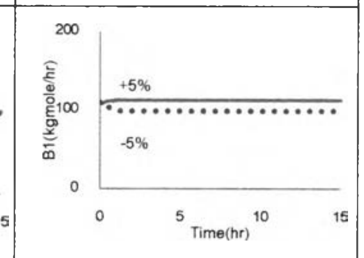
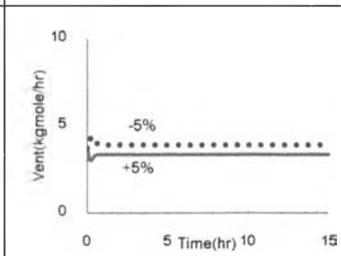
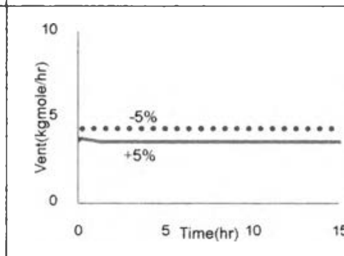
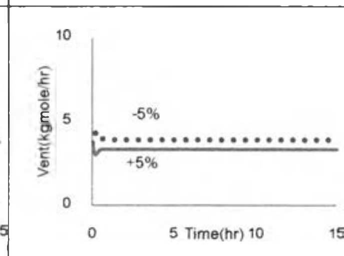
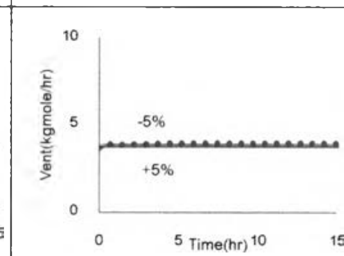
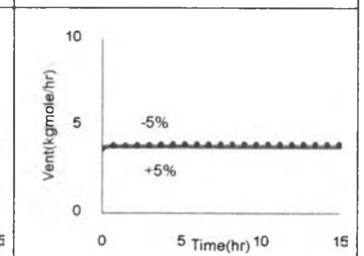
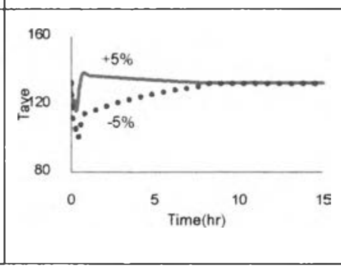
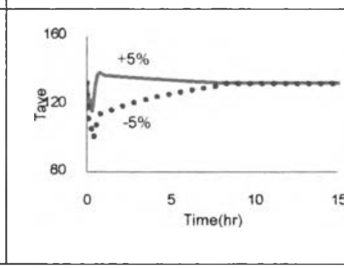
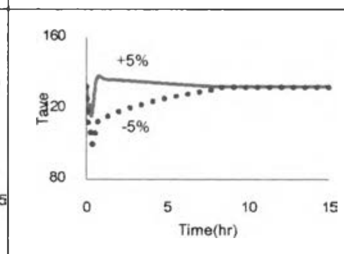
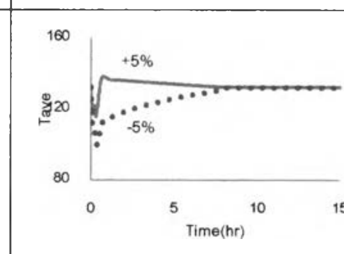
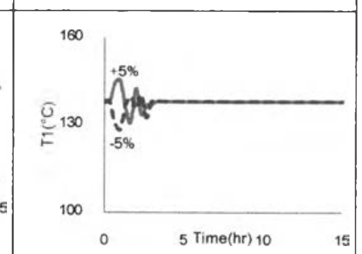
	CS0	CS1	CS2	CS3	CS4
E					
F					
G					
H					

Table 5.12 (continue) Dynamic response with total feed composition changed

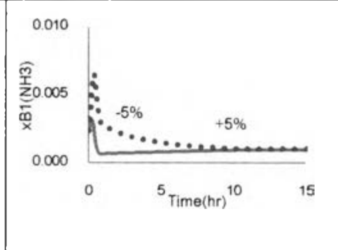
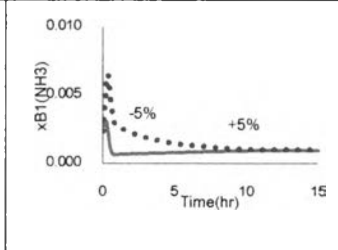
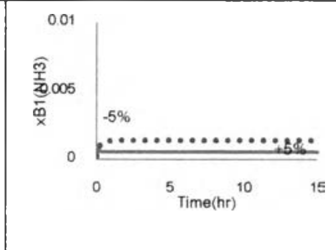
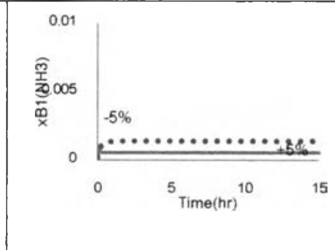
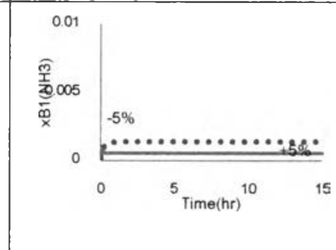
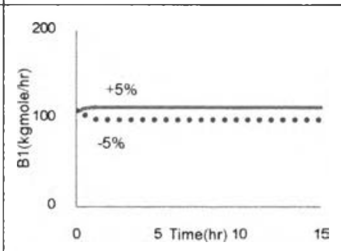
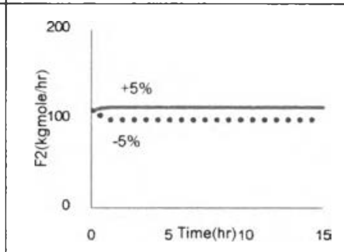
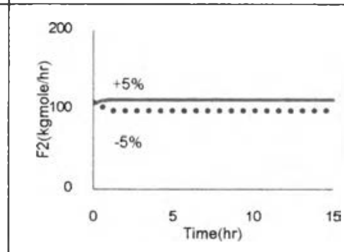
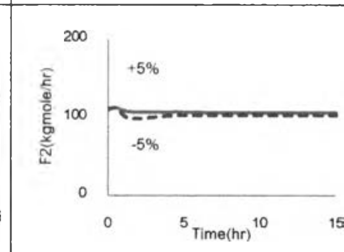
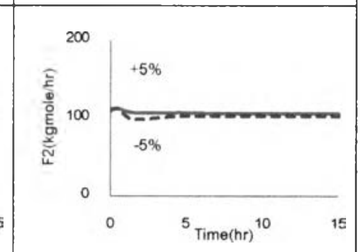
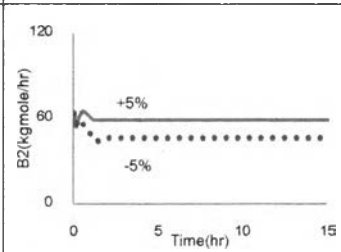
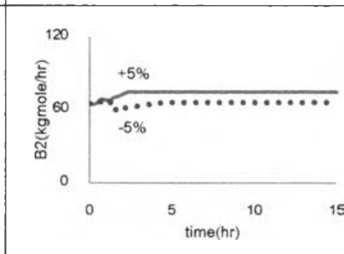
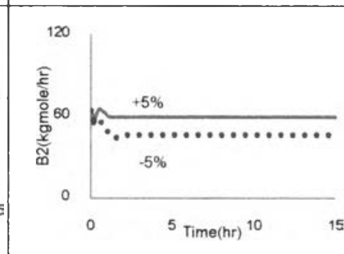
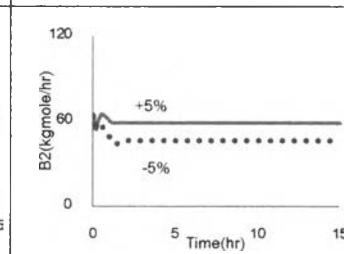
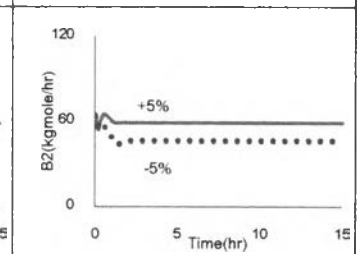
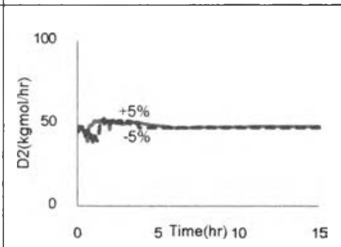
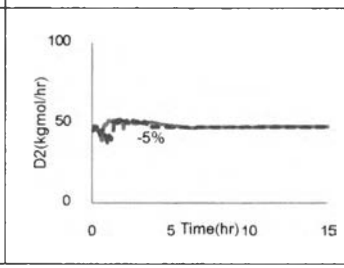
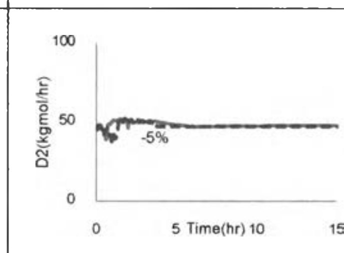
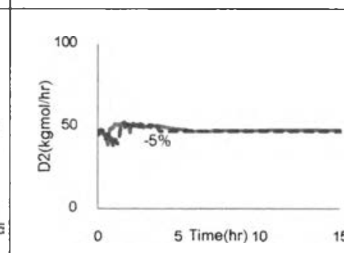
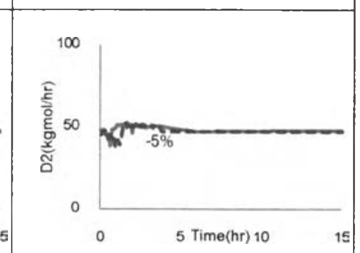
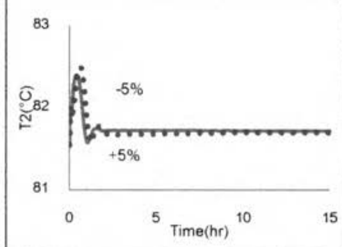
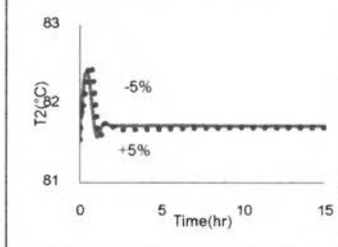
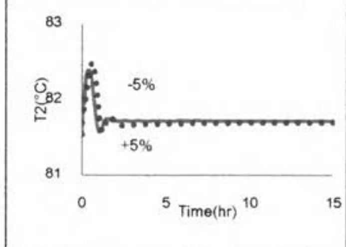
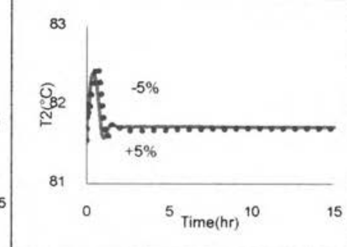
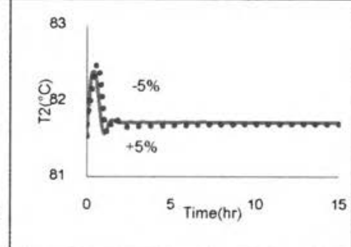
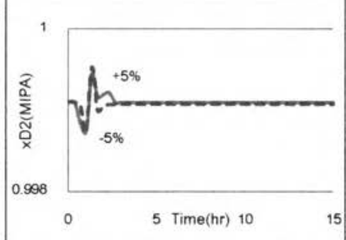
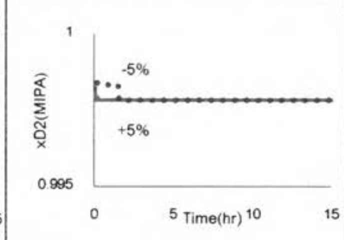
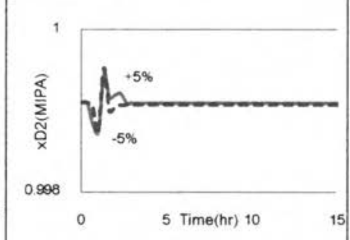
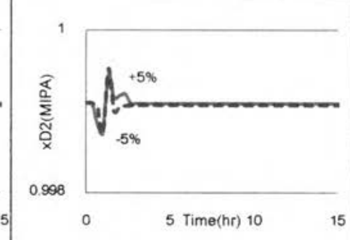
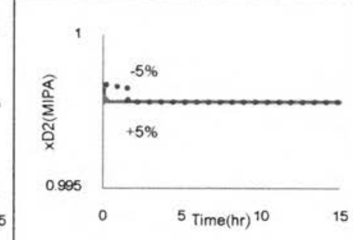
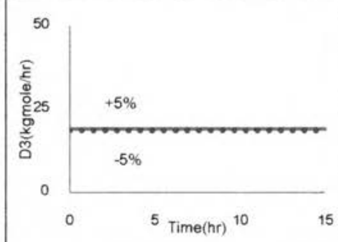
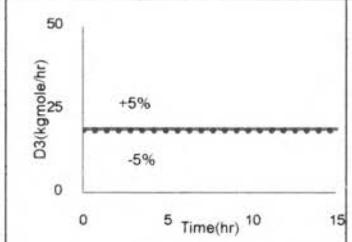
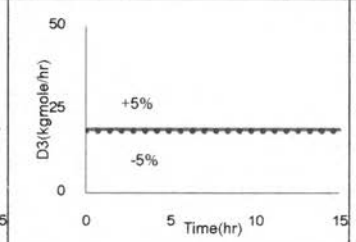
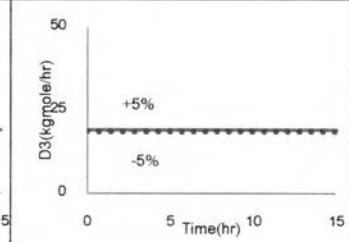
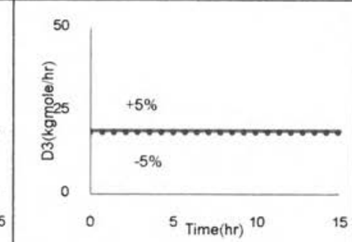
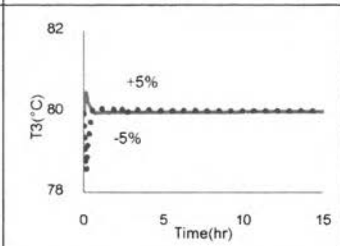
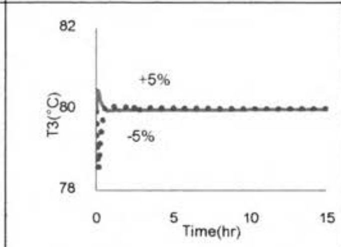
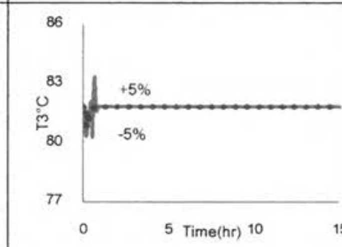
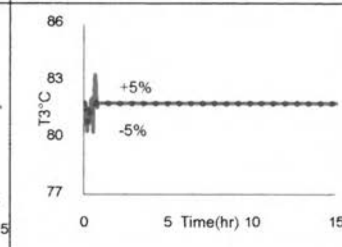
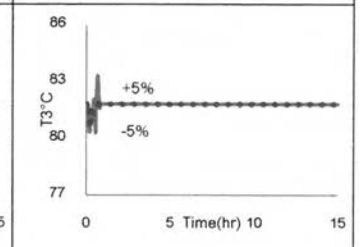
	CS0	CS1	CS2	CS3	CS4
I					
J					
K					
L					

Table 5.12 (continue) Dynamic response with total feed composition changed

	CS0	CS1	CS2	CS3	CS4
M					
N					
O					
P					

5.4 Evaluation of the Dynamic Performance

In this research, integral absolute error (IAE) is used to evaluate the dynamic performance of the control structure of base case and designed control structures. The formulation of a dynamic performance as shown below:

$$IAE = \int \left| \varepsilon (t) \right| dt$$

Note that $\varepsilon (t) = ysp(t)-y(t)$ is the deviation (error) of the response from the desired set point.

The IAE results are investigated form the safety loop control; temperature and pressure, and the composition of product control loop. The temperature and pressure control are taken into account because those control loop cause the operation smoothly and safely. The composition loops are also important because of requirement of the commercial product quality and maximizing the usage of raw material. The minimum IAE value indicates the best control structure handling the material and thermal disturbances which can operate smoothly and safely and also can keep the product quality at the specification value.

For changing in material disturbances test of fresh IPA fresh feed flow and temperature feed flow, and composition IPA feed flow and designed control structures (CS1-CS4). The IAE results for handle disturbances are shown in Table 5.13, 5.14, and Table 5.15, respectively.

Table 5.13 IAE of safety control loop with total feed change

Control structure	Temperature	Pressure	Composition	Sum IAE
Base case (CS0)	0.4649	0.2390	0.5523	1.5555
CS1	0.3333	0.7369	0.5165	1.5867
CS2	0.5649	0.2690	0.4543	1.2882
CS3	0.4139	0.2631	0.4435	1.1205
CS4	0.4253	0.2489	0.4631	1.1373

Table 5.14 IAE of safety control loop with temperature feed change

Control structure	Temperature	Pressure	Composition	Sum IAE
Base case (CS0)	0.7006	0.6692	0.4114	1.7812
CS1	0.6872	0.6035	0.6292	1.9199
CS2	0.6225	0.5012	0.5990	1.7227
CS3	0.3749	0.2875	0.4672	1.1296
CS4	0.3879	0.2789	0.3978	1.0646

Table 5.15 IAE of safety control loop with composition feed change

Control structure	Temperature	Pressure	Composition	Sum IAE
Base case (CS0)	0.2058	0.3867	0.6771	1.2697
CS1	0.7505	0.4207	0.6193	1.7905
CS2	0.3099	0.1365	0.5374	0.9839
CS3	0.2892	0.1256	0.4073	0.8223
CS4	0.3358	0.1342	0.3579	0.8279

For all disturbances testing, table 5.10, 5.11, and 5.13 show the results of Integral Absolute Error (IAE) of temperature controller, pressure controller, and composition controller, respectively for handle disturbances. See that the control structure CS4 can handle disturbances and the best. Control structure CS3 is the second. Control structure CS2 is the third and control structure CS1 is the four. From this result, if the designed control structure controls the all column with reflux to feed ratio controller and heat pathway for control temperature is fed to vaporizer it can handle all disturbances well.

5.3 Utilities cost

The process utilities include cooling water, heated water, compressed air, refrigerants and electricity. The prices of each utility are given as followed:

- Cooling water $0.354 per GJ
- Low pressure stream (LP) $6.08 per GJ
- High pressure stream (HP) $9.83 per GJ

Note that all cost are in unit of m$ per year. Table 5.16 to 5.18 shows the utilities cost of each control structure when facing various type of disturbances.

Table 5.16 Utilities cost with fresh feed changed

CASE	Cooling water		LP		HP		Utility cost
	GJ	m$ per year	GJ	m$ per year	GJ	m$ per year	
CS0	-41.388	0.01465	1120.280	6.811302	218.845	2.151246	8.964
CS1	-38.665	0.01368	1326.244	8.063786	159.210	1.565534	9.643
CS2	-33.600	0.01189	802.271	4.877810	377.828	3.714057	8.593
CS3	-48.597	0.01720	787.423	4.787534	366.838	3.606025	8.395
CS4	-46.458	0.01644	810.236	4.926234	367.038	3.60798	8.698

Table 5.17 Utilities cost with feed IPA composition changed

CASE	Cooling water		LP		HP		Utility cost
	GJ	m$ per year	GJ	m$ per year	GJ	m$ per year	
CS0	-41.535	0.01470	1264.755	7.689710	233.890	2.22991	9.9343
CS1	-39.835	0.01410	1357.460	8.253640	154.440	1.51814	9.7858
CS2	-35.088	0.01242	443.200	2.694656	541.705	5.32496	8.0232
CS3	-46.528	0.016.47	783.961	4.766482	338.779	3.33019	8.0966
CS4	-42.456	0.015029	756.263	4.598079	341.256	3.35454	7.9676

Table 5.18 Utilities cost with feed temperature changed

CASE	Cooling water		LP		HP		Utility cost
	GJ	m$ per year	GJ	m$ per year	GJ	m$ per year	
CS0	-42.670	0.01510	1081.823	6.577484	236.125	2.321109	8.9136
CS1	-39.169	0.01386	1573.838	9.568940	157.320	1.546456	11.129
CS2	-34.718	0.01229	1302.623	7.919950	276.474	2.717743	10.649
CS3	-45.639	0.01615	1281.040	7.788726	233.669	2.296970	10.101
CS4	-48.369	0.01712	1296.256	7.881236	228.622	2.247354	10.145

From the utilities cost table results that for the fresh feed, feed IPA composition, and feed temperature changed, the smallest utilities cost belongs to case CS3 followed by case CS4, CS1,CS2, and base case. As a result, base case has the highest utilities cost when facing all types of disturbances, though, case CS3 has the smallest utilities cost.