CHAPTER IV

HYDRODEALKYLATION PROCESS

4.1 Process Description

The hydrodealkylation HDA of toluene process (alternative 1) by Douglas (1988) on conceptual design as in Fig. 4.1 contain nine basic unit operations: reactor, furnace, vapor-liquid separator, recycle compressor, two heat exchangers, and three distillation columns. Two raw materials, hydrogen, and toluene, are converted into the benzene product, with methane and diphenyl produced as by-products. The two vapor-phase reactions are

$$\text{Toluene} + H_2 \longrightarrow \text{benzene} + CH_4$$

$$\text{2BenZene} \longleftrightarrow \text{diphenyl} + H_2$$

The kinetic rate expressions are functions of the partial pressure (in psia) of toluene p_T, hydrogen p_H, benzene p_B, and diphenyl p_D, with an Arrhenius temperature dependence. Zimmerman and York (1964) provide the following rate expression:

$$r_1 = 3.6858\times10^6 exp(-25616/T)p_T p_H^{1/2}$$

$$r_2 = 5.987\times10^4 exp(-25616/T)p_B^2 - 2.553\times10^5 exp(-25616/T)p_D p_H$$

Where r_1 and r_2 have units of lb×mol/(min×ft3) and T is the absolute temperature in Kelvin. The heats of reaction given by Douglas (1988) are -21500 Btu/lb×mol of toluene for r_1 and 0 Btu/lb×mol for r_2.

The effluent from the adiabatic reactor is quenched with liquid from the separator. This quenched stream is the hot-side feed to the process-to-process heat exchanger, where the cold stream is the reactor feed stream prior to the furnace. The reactor effluent is then cooled with cooling water and the vapor (hydrogen, methane) and liquid (benzene, toluene, diphenyl) are separated. The vapor stream from the separator is split and the remainder is sent to the compressor for recycle back to the reactor.

The liquid stream from the separator (after part is taken for the quench) is fed to the stabilizer column, which has a partial condenser component. The bottoms stream from the stabilizer is fed to the product column, where the distillate is the benzene product from the process and the bottoms is toluene and diphenyl fed to the recycle column. The distillate from the recycle column is toluene that is recycled back to the reactor and the bottom is the diphenyl byproduct.

Makeup toluene liquid and hydrogen gas are added to both the gas and toluene recycle streams. This combined stream is the cold-side feed to the process-to-process heat exchanger. The cold-side exit stream is then heated further up to the required reactor inlet temperature in the furnace, where heat is supplied via combustion of fuel.

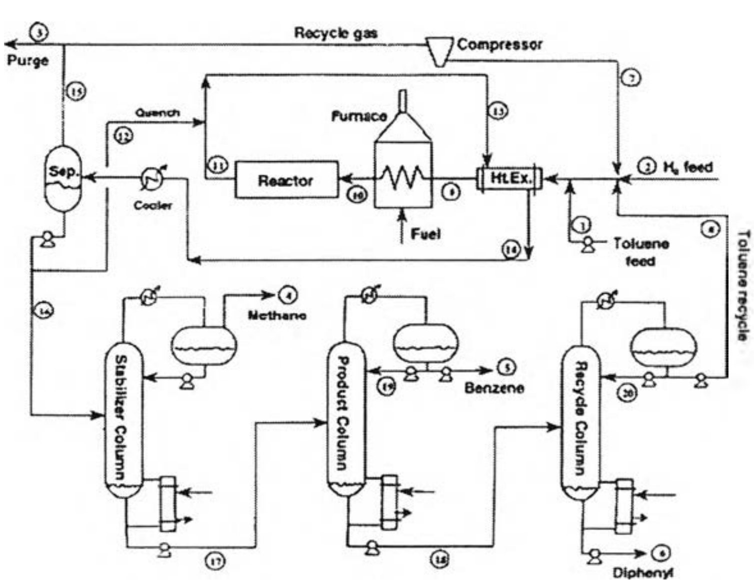


Figure 4.1: Hydrodealkylation HDA of toluene process (alternative 1).

Component physical property data for the HDA process were obtain from William L. Luyben, Bjorn D. Tyreus, Michael L. Luyben (1999)

4.2 Hydrodealkylation Process Alternatives

Terrill and Douglas (1987b) design six different energy-saving alternatives to the base case. The simplest of these designs (alternative 1) recovers an additional 29% of the base case heat consumption by making the reactor preheated larger and the furnace smaller.

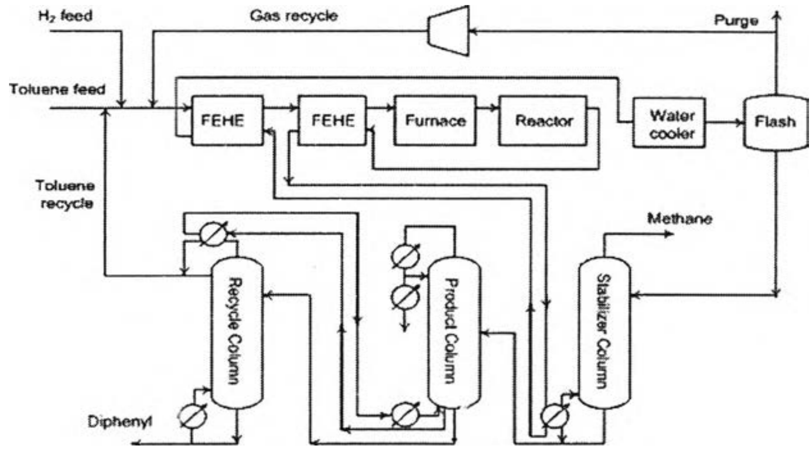


Figure 4.2: HDA process -alternative 1.

In alternative 2 (Figure 4.3) is the same as alternative 1, except that recycle column was pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler.

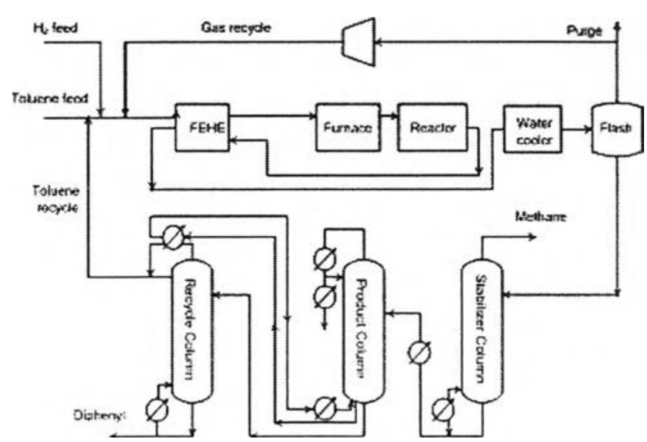


Figure 4.3: HDA process -alternative 2.

In alternative 3 part of the heat in the reactor effluent stream is used to drive the stabilizer reboiler, recycle column was pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler as in Figure 4.4.

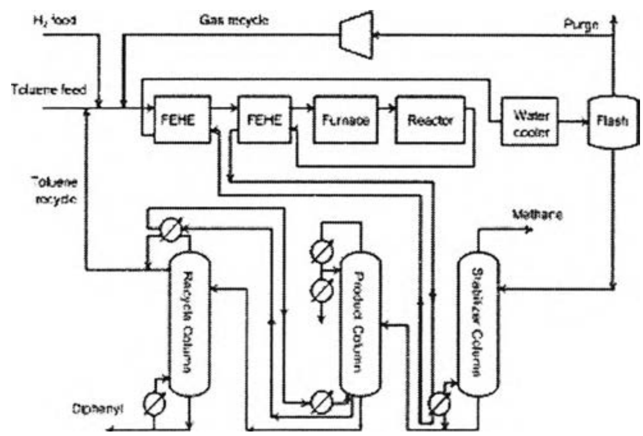


Figure 4.4: HDA process -alternative 3.

In alternative 4 the reactor effluent is used to drive the product column reboiler. recycle column was pressure shifted to be above the pinch temperature as in Figure 4.5.

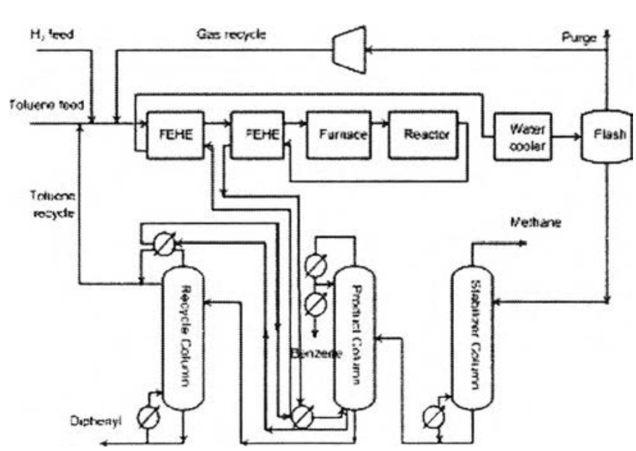


Figure 4.5: HDA process -alternative 4.

For alternative 5, both the stabilizer reboiler and the product column reboiler are driven consecutively by the reactor effluent stream, recycle column was pressure shifted to be above the pinch temperature as in Figure 4.6.

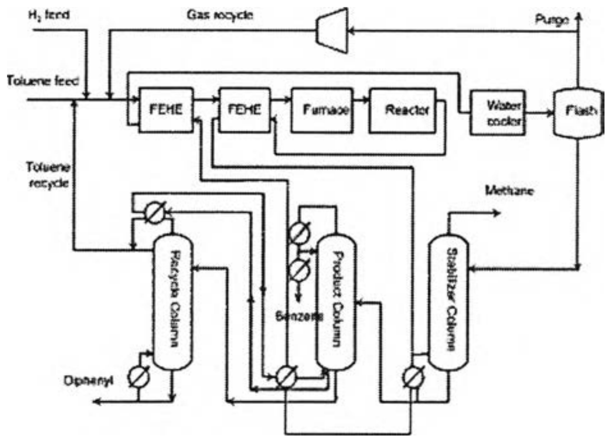


Figure 4.6: HDA process -alternative 5.

The benefit obtained from energy integration with the base-case flow rate for the six alternatives is given in table 4.1. The energy saving from the energy integration fall between 29 and 43 %, but the cost saving are in the range from -1 to 5 %. The cost saving are not as dramatic the raw-material costs dominate the process economics.

Table 4 1: Energy integration for HDA process

	Base case	Alternatives					
		1	2	3	4	5	6
1. TAC ($\$10^6$/yr) base-case flows	6.38	6.40	6.45	6.38	6.11	6.04	6.03
2. Utilities usage (MW), base-case flows	12.70	9.06	7.68	7.34	7.30	7.30	7.30
3. Energy saving %		29	40	42	43	43	43
4. Cost saving %		-0.3	-1	0	4	5	5

4.3 Steady-State Modeling

First, a steady-state model is built in HYSYS.PLANT, using the flowsheet and equipment design information, mainly taken from Douglas (1988); Luyben et al. (1998). Table A.2 presents the data and specifications for the equipment employed other than the three columns. For our simulation, Peng-Robinson model is selected for physical property calculations because of its reliability in predicting the properties of most hydrocarbon-based fluids over a wide range of operating conditions. The reaction kinetics of both reactions are modeled with standard Arrhenius kinetic expressions available in HYSYS.PLANT, and the kinetic data are taken from Luyben et al. (1998). Since there are four material recycles, four RECYCLE operations are inserted in the streams, Hot-In, Gas-Recycle, Quench, and Stabilizer-Feed . Proper initial values should be chosen for these streams, otherwise the iterative calculations might converge to another steady-state due to the non-linearity and unstable characteristics of the process.

When columns are modeled in steady-state, besides the specification of inlet streams, pressure profiles, numbers of trays and feed tray, two specifications need to be given for columns with both reboiler and condenser. These could be the duties, reflux rate, draw stream rates, composition fractions, etc. We chose reflux ratio and overhead benzene mole fraction for the stabilizer column. For the remaining two columns, bottom and overhead composition mole fractions are specified to meet the required purity of products given in Douglas (1988). The detailed design data and specifications for the columns are summarized in Table A.3.This table also includes details of trays, which are required for dynamic modeling. The tray sections of the columns are calculated using the tray sizing utility in HYSYS, which calculates tray diameters based on Glitsch design parameters for valve trays. Though the tray diameter and spacing, and weir length and height are not required in steady-state modeling, they are required for dynamic simulation.

4.3.1 Steady State Simulation of HDA Process Alternative 1

Figure 4.8 shows the HYSYS flowsheet of HDA process alternative 1. The steady state simulation results are summarized in table A1.1 . For the comparison, the steady state simulation results given by Luyben et al. (1999) are also listed in those tables. The data and specification for the different equipments are given in Appendix B.

Since there are four materials recycle streams in HDA process alternative 1, four recycle modules are inserted in the streams: hot stream to FEHE, gas recycle, quench, and toluene recycle stream. Proper initial values should be chosen for these streams; otherwise the iterative calculations might converge to another steady state due to the non-linearity and unstable characteristics of the process.

All of the three columns are simulated using the "distillation column" module. When columns are modeled in steady state, besides the specification of inlet streams, pressure profiles, number of trays and feed tray, two additional variables should be additionally specified for columns with condenser or reboiler. These could be the duties, reflux rate, draw stream rates, composition fraction, etc. We chose to specify a priori overhead and bottom component mole fraction for all columns. These mole fractions are specified to meet the required purity of product given in Douglas (1988). The tray sections of the columns are calculated using the tray sizing utility in HYSYS, which calculates tray diameters based on sieve trays. The column specifications of HDA process alternative 1 are given in Appendix B . Although the tray diameter and spacing, weir length and height are not required for steady state modeling, they are required for dynamic simulation.

4.3.2 Steady State Simulation of HDA Process Alternative 2 and 5

The steady state simulation results of HDA process alternative 1 have been compared with the earlier study by Luyben et al. (1999), and the results are found consistent with those in the earlier study. Then, considering the consistency of the simulation results of the HDA process alternative 1 with respect to the previous work, the other alternatives considered in this work, i.e. alternative 2 and 5 are also developed in the HYSYS software environment. Figures 4.9 and 4.10 show the HYSYS flowsheets of the HDA process with energy integration schemes for alternative 2 and 5, respectively. The data for the selected streams for these alternatives are listed in Appendix A. The data and specifications for the equipments ate summarized in Appendix B.

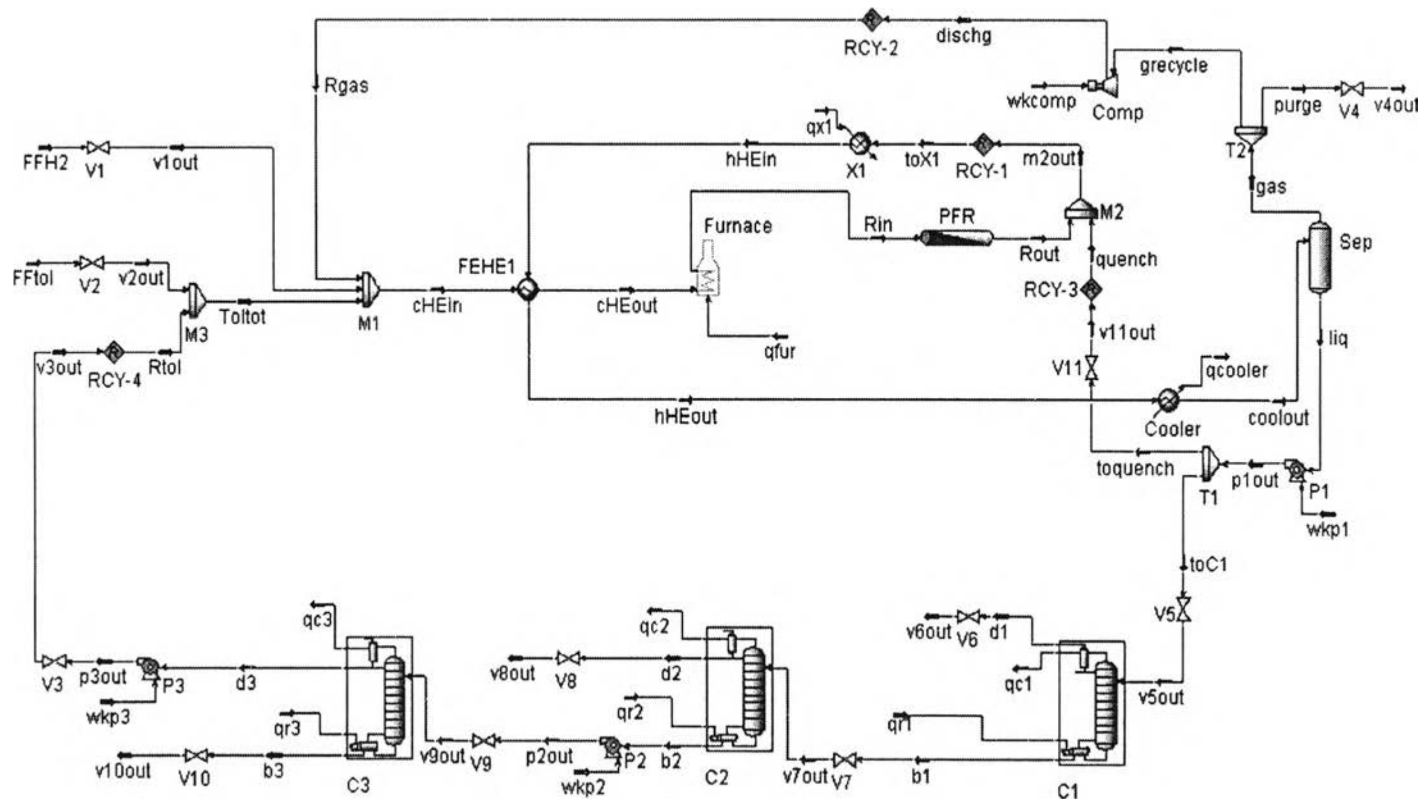


Figure 4.7: The simulated HDA process alternative 1 at steady-state by HYSYS

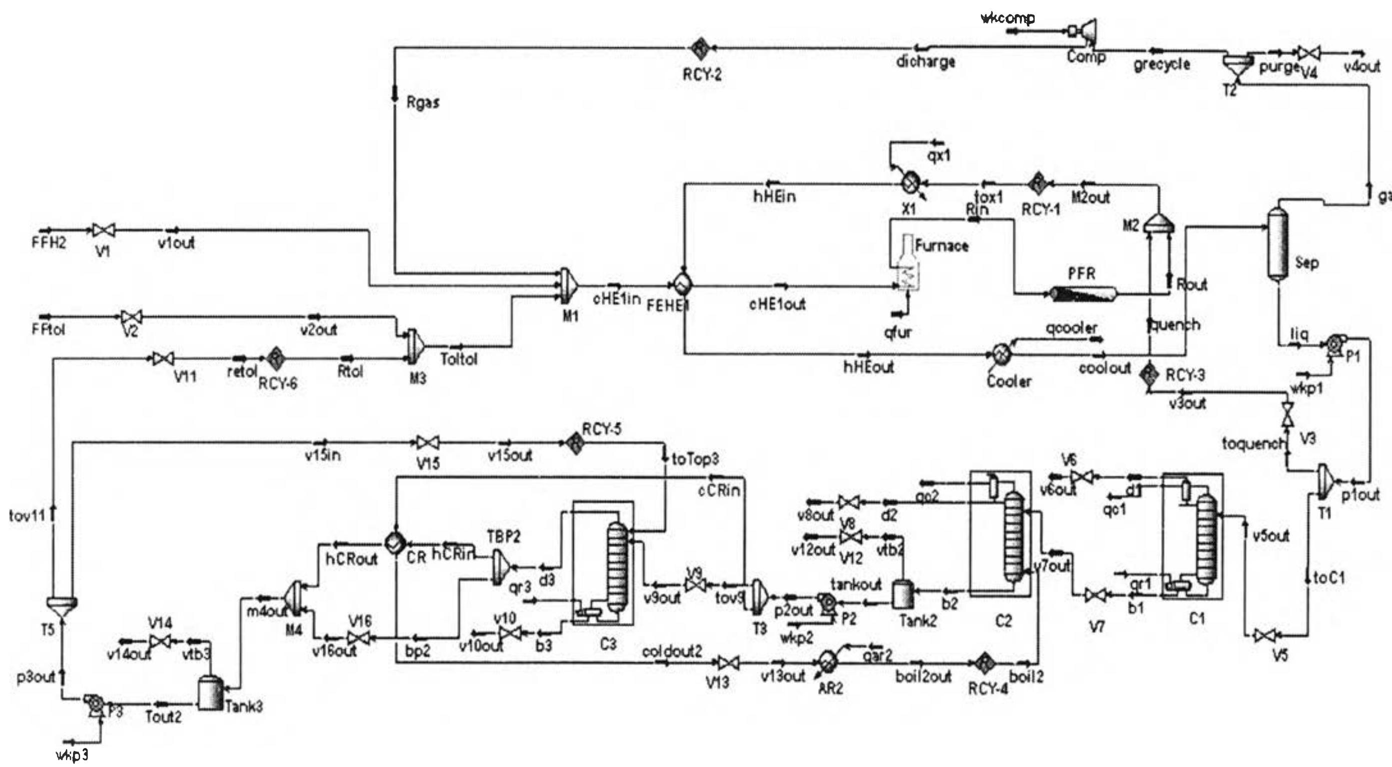


Figure 4.8 The simulated HDA process alternative 2 at steady-state by HYSYS

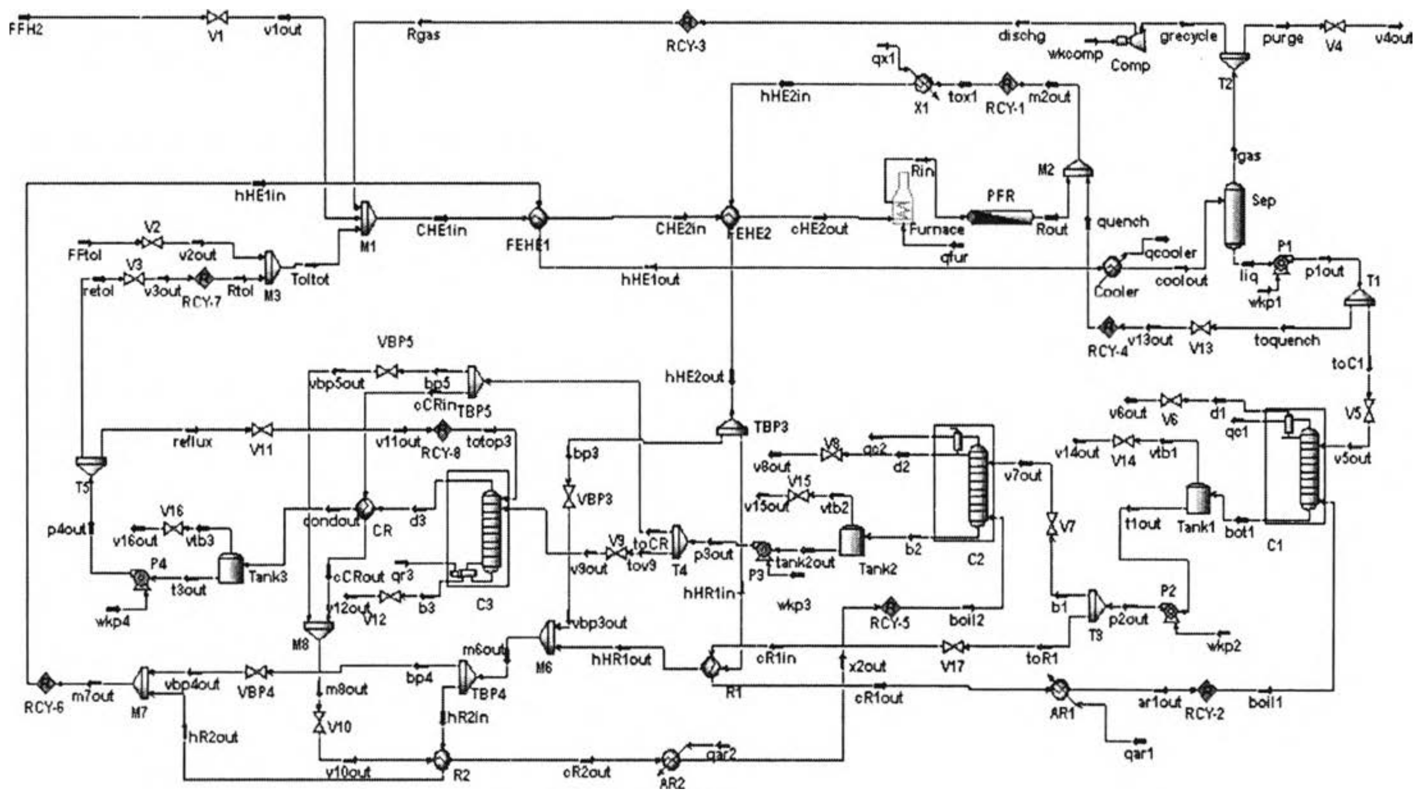


Figure 4.9 The simulated HDA process alternative 5 at steady-state by HYSYS

4.4 Plantwide control design procedure

Step 1. Establish Control Objectives.

For this process, the essential is to produce pure benzene while minimizing yield losses of hydrogen and diphenyl. The reactor effluent gas must be quenched to 621.11°C. The design a control structures for process associate with energy integration can be operated well.

Step 2. Determine Control Degree of Freedom.

There are 23 control degrees of freedom. They include; two fresh feed valves for hydrogen and toluene, purge valve, separator base and overhead valves, cooler cooling water valve, liquid quench valve, furnace fuel valve, stabilizer column steam, bottoms, reflux, cooling water, and vapor product valves; product column steam, bottoms, reflux, distillate, and cooling water valves; and recycle column steam, bottoms, reflux, distillate, and cooling water valves.

Step 3. Establish Energy management system.

The reactor operates adiabatically, so for a given reactor design the exit temperature depends upon the heat capacities of the reactor gases, reactor inlet temperature, and reactor conversion. Heat from the adiabatic reactor is carried in the effluent stream and is not removed from the process until it is dissipated to utility in the separator cooler.

Energy management of reaction section is handled by controlling the inlet and exit streams temperature of the reactor. Reactor inlet temperature must be controlled by adjusting fuel to the furnace and reactor exit temperature must be controlled by quench to prevent the benzene yield decreases from the side reaction. In the reference control structure, the effluent from the adiabatic reactor is quenched with liquid from the separator. This quenched stream is the hot-side feed to the process-to-process heat exchanger, where the cold stream is the reactor feed stream prior to the furnace. The reactor effluent is then cooled with cooling water. But in alternative 3 part of the heat in the reactor effluent stream is used to drive the stabilizer reboiler before go to cooling water. And recycle column is pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler for saving cost from the utility. However, this method gives up degree of freedom for temperature control. The solutions to restore one degree of freedom fairly easily have two ways. It is possible to oversize the P/P exchanger and provides a controlled bypass around it. And it is possible to combine the P/P exchanger with a utility exchanger.

Step 4. Set Production Rate.

Many control structures, there are not constrained to set production either by supply or demand. Considering of the kinetics equation is found that the three variables alter the reaction rate; pressure, temperature and toluene concentration (limiting agent).

- Pressure is not a variable choice for production rate control because of the compressor has to operate at maximum capacity for yield purposes.
- Reactor inlet temperature is controlled by specify the reactant fresh feed rate and reactant composition into the reactor constant. The reactor temperature is constrained below 704.44 C for preventing the cracking reaction that produce undesired byproduct.

- Toluene inventory can be controlled in two ways. Liquid level at the top of recycle column is measured to change recycle toluene flow and total toluene feed flow in the system is measured for control amount of fresh toluene feed flow.

For on demand control structure the production rate is set; distillate of product column is flow control instead of level control so condenser level is controlled by manipulating the total flow rate of the toluene. This on-demand structure might be used when the downstream customer desires immediate responses in the availability of the product stream from this unit.

Step 5. Control Product Quality and Handle Safety, Operational, and Environmental Constraints.

Benzene quality can be affected primarily by two components, methane and toluene. Any methane that leaves in the bottoms of the stabilizer column contaminates the benzene product. The separation in the stabilizer column is used to prevent this problem by using a temperature to set column stream rate (boilup). Toluene in the overhead of the product column also affects benzene quality. Benzene purity can be controlled by manipulating the column steam rate (boilup) to maintain temperature in the column.

Step 6. Control Inventories and Fix a Flow in Every Recycle Loop.

In most processes a flow control should be present in all recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows, while the process is perturbed by small disturbance. We call this high sensitivity of the recycle flowrates to small disturbances the "snowball effect".

Four pressures and seven liquid levels must be controlled in this process. For the pressures, there are in the gas loop and in the three distillation columns. In the gas loop, the separator overhead valve is opened and run the compressor at maximum gas recycle rate to improve yield so the gas loop control is related to the purge stream and fresh hydrogen feed flow. In the stabilizer column, vapor product flow is used to control pressure. In the product column, pressure control can be achieved by manipulating cooling water flow, and in the product column pressure control can be set by bypass valve of P/P heat exchanger to regulate overhead condensation rate.

For liquid control loops, there are a separator and two receivers in each column (base and overhead). The most direct way to control separator level is with the liquid flow to the stabilizer column. The stabilizer column overhead level is controlled with cooling water flow and base level is controlled with bottom flow. In several cases of this research; the product column, distillate flow controls overhead receiver level but on demand control structure condenser level is controlled by cascade the total flow rate of the toluene and bottom flow controls base level. In the recycle column manipulate the total toluene flow to control level. The base level of recycle column in the reference is controlled by manipulating the column steam flow because it has much larger effect than bottoms flow. But the column steam flow does not obtain a good controllability, so base level is controlled with bottom flow.

Step 7. Check Component Balances.

Component balances control loops consists of:

- Methane is purged from the gas recycle loop to prevent it from accumulating and its composition can be controlled with the purge flow.

- Diphenyl is removed in the bottom stream from the recycle column, where bottom stream controls base level. And control temperature (or concentration) with the reboiler steam.

- The inventory of benzene is accounted for via temperature and overhead receiver level control in the product column. But on demand structure the inventory of benzene is accounted for via temperature and distillate flow control in the product column.

- Toluene inventory is accounted for via level control in the recycle column overhead receiver.

- Gas loop pressure control accounts for hydrogen inventory.

Step 8. Control Individual Unit Operations.

The rest degrees of freedom are assigned for control loops within individual units. These include:

- Cooling water flow to the cooler controls process temperature to the separator.

- Refluxs to the stabilizer, product, and recycle columns are flow controlled.

Table 4.2: Component Material Balance

	Input	+Generation	-Output	-Consumption	Accumulation
Component					Inventory Controlled by
H_2	Fresh Feed	$0.5V_Rr_2$	Purge stream	V_Rr_1	Pressure control of recycle gas loop
CH_4	0	V_Rr_1	Purge stream	0	Composition control of recycle gas loop
C_6H_6	0	V_Rr_1	Product stream	$2V_Rr_2$	Temperature control in product column
C_7H_8	Fresh Feed	0	0	V_Rr_1	Level control in recycle column reflux drum
$C_{12}H_{10}$	0	$0.5V_Rr_2$	Purge stream	0	Temperature control in recycle column

Where V_R= reactor volume

r_1 = first reaction rate

r_2 = second reaction rate

Step 9. Optimize Economics or Improve Dynamic Controllability.

The basic regulatory strategy has now been established. Some freedom is used to select several controller setpoints to optimize economics and plant performance. Such as, the setpoint for the methane composition controller in the gas recycle loop must balance the trade-off between yield loss and reactor performance. Reflux flows to the stabilizer, product, and recycle columns must be determined based upon column energy requirement and potential yield losses of benzene (in the overhead of the stabilizer and recycle columns) and toluene (in the base of the recycle column).

4.5 Design of plantwide control structure

In this current work three control structures were designed and compared, the first control structure we modify of Luyben et al. (2002) control system, namely control structure 1 (CS1) to the HDA process as show in Figure 4.10 - 4.12. The second control structure we apply with control structure 1 with Kietawarin (2002) control system, namely control structure 2 (CS2) to the HDA process as show in Figure 4.13 - 4.15. The third control structures CS3 as show in Figure 4.16 - 4.18, this control structure a ratio control was induced to the second control scheme.

4.5.1 Design of control structure 1 (CS1).

The plant wide control structures in the HDA plant alternatives 1, 2 and 5 are designed based on the heuristic design procedure given by Luyben et al. (2002). The major loops are the same as those used in Luyben et al. (1999), but we have used valve position control concept Luyben (1990) which can reduce energy-cost of utility. In this control structure both valve bypart of column heat exchanger and column auxiliary heater is used to control tray temperature of column. When valve bypart decrease to 5% open but temperature cannot achieve to its setpoint the auxiliary will operate to control temperature as show in figure4.10 - 4.12. The size of disturbance in this study is about 5 to 10% according to Luyben's recommendations.

4.5.2 Design of control structure 2 (CS2).

The second control structure we apply control structure 1 with Kietawarin (2002) control system by adding a cooling unit to control the outlet temperature from reactor, instead of using internal process flow (from bottom of vapor-liqiud seperator) to reduce material and seperation ratio fluctuations within the process flow.

4.5.3 Design of new plantwide control structure (CS3)

This control structure a ratio control of fresh feed toluene and fresh feed toluene hydrogen was induced to the second control scheme for controlling the ratio of hydrogen and toluene within the process.

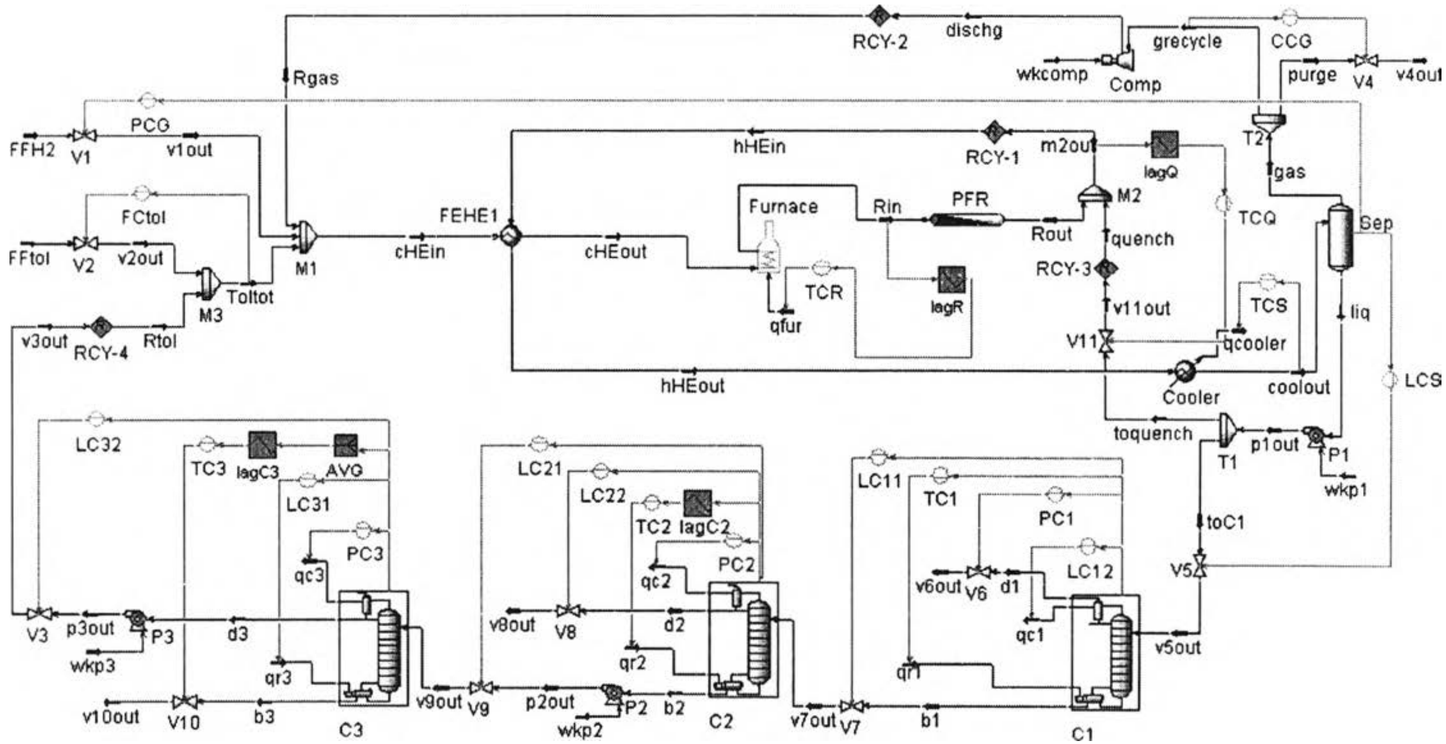


Figure 4.10 Application of control structure 1 to HDA plant alternative 1

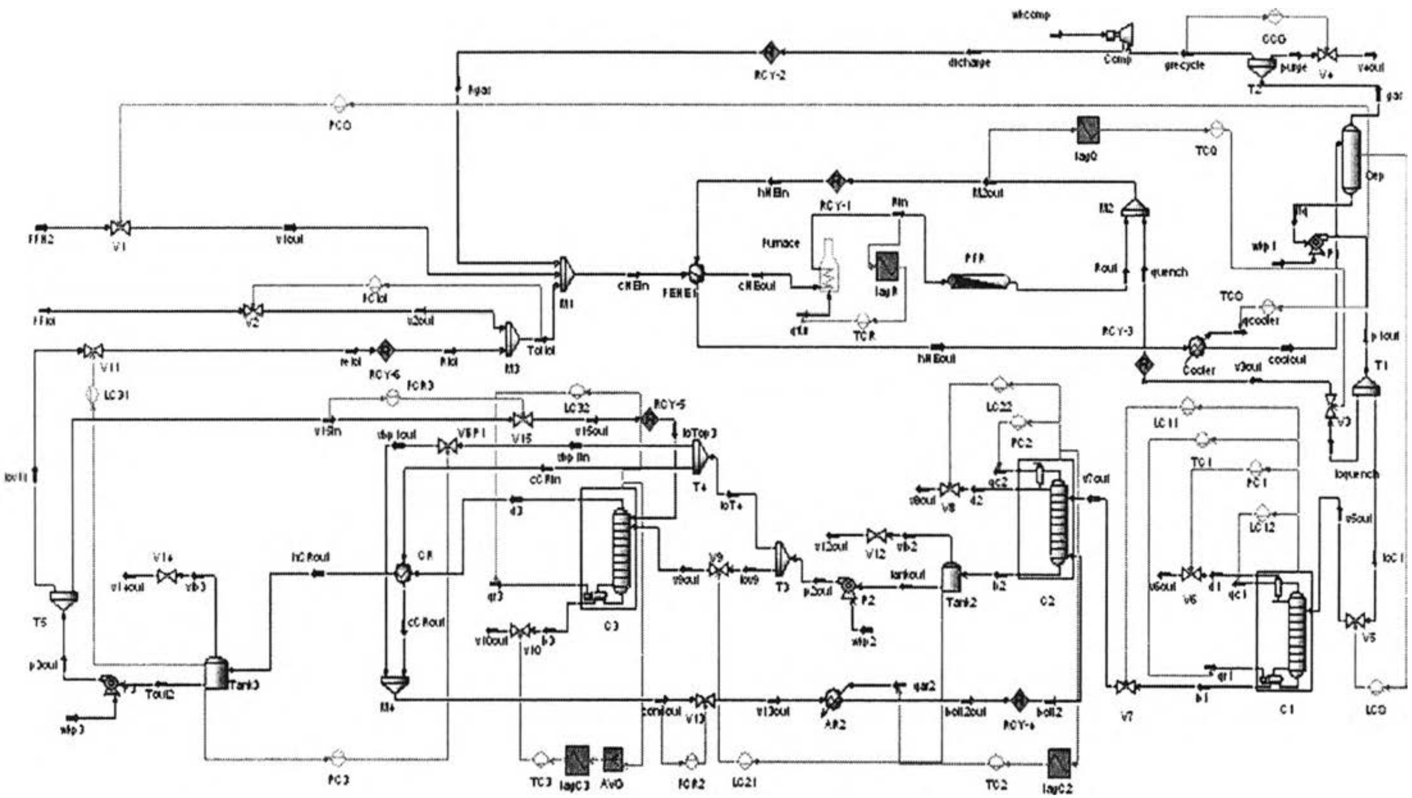


Figure 4.11 Application of control structure 1 to HDA plant alternative 2

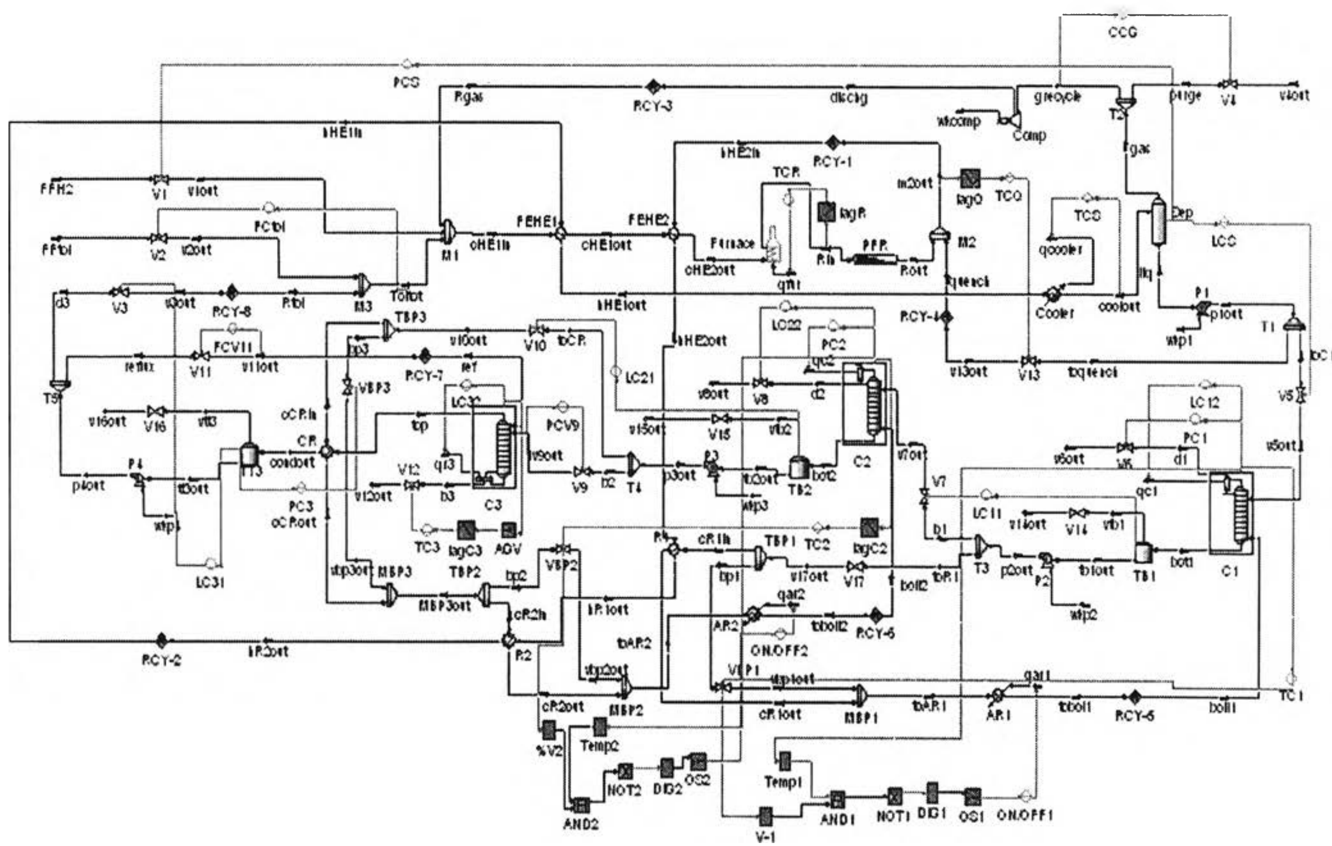


Figure 4.12 Application of control structure 1 to HDA plant alternative 5

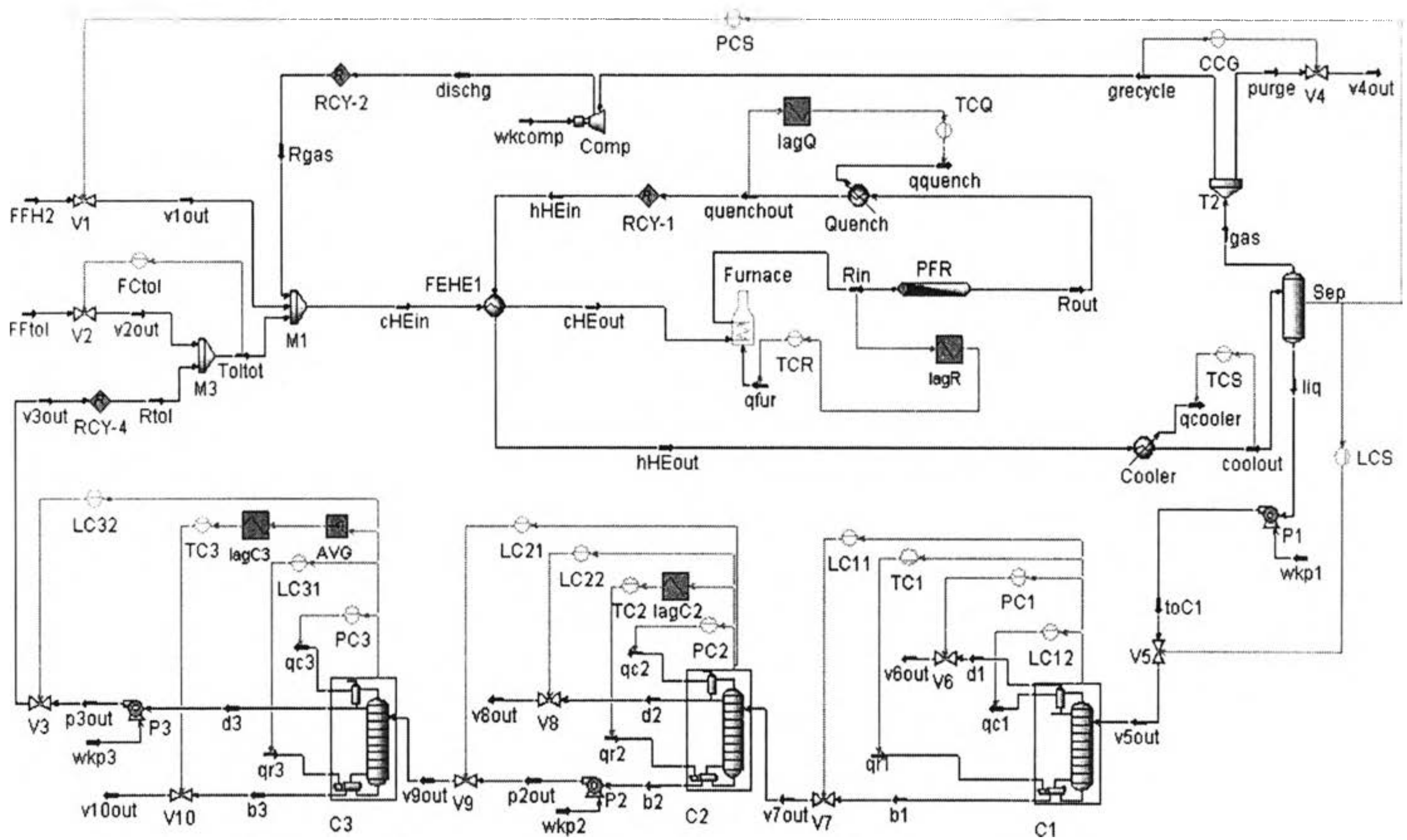


Figure 4.13. Application of control structure 2 to HDA plant alternative 1

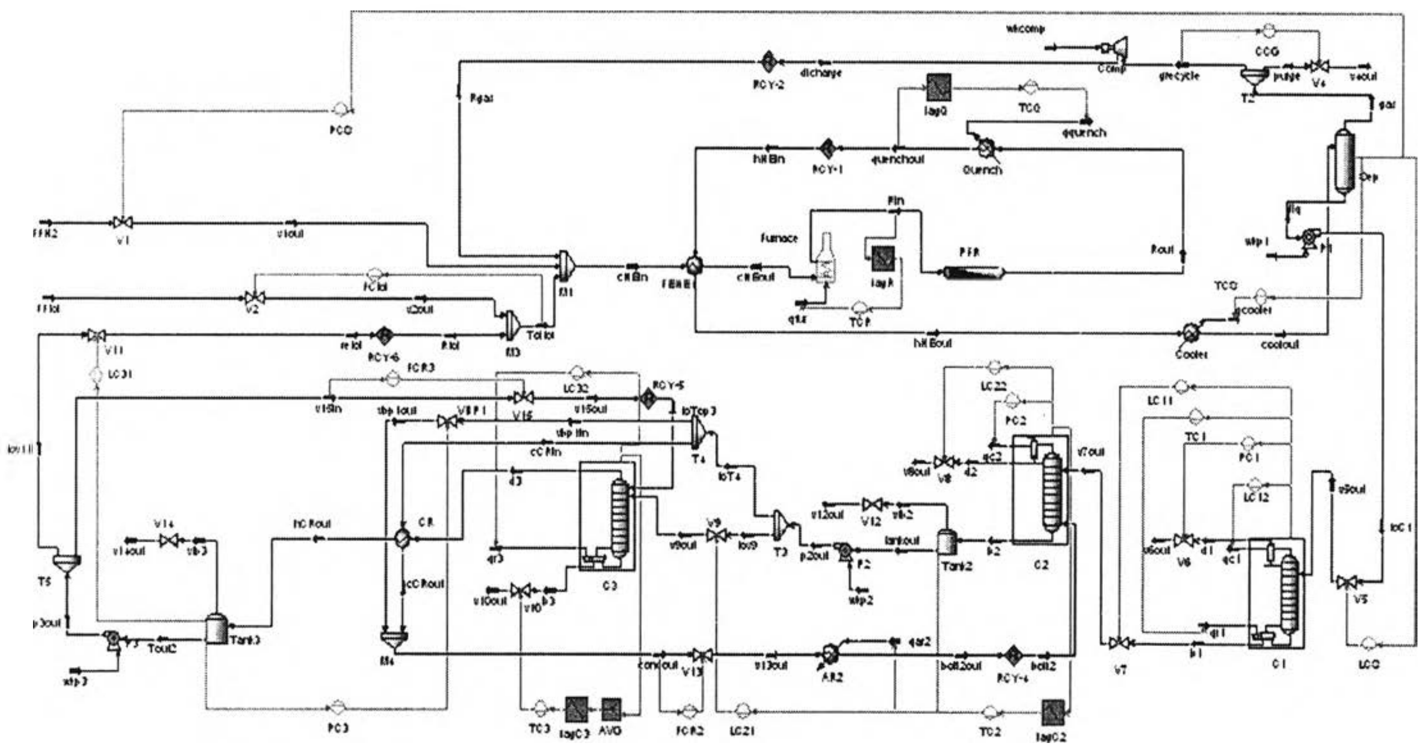


Figure 4.14. Application of control structure 2 to HDA plant alternative 2

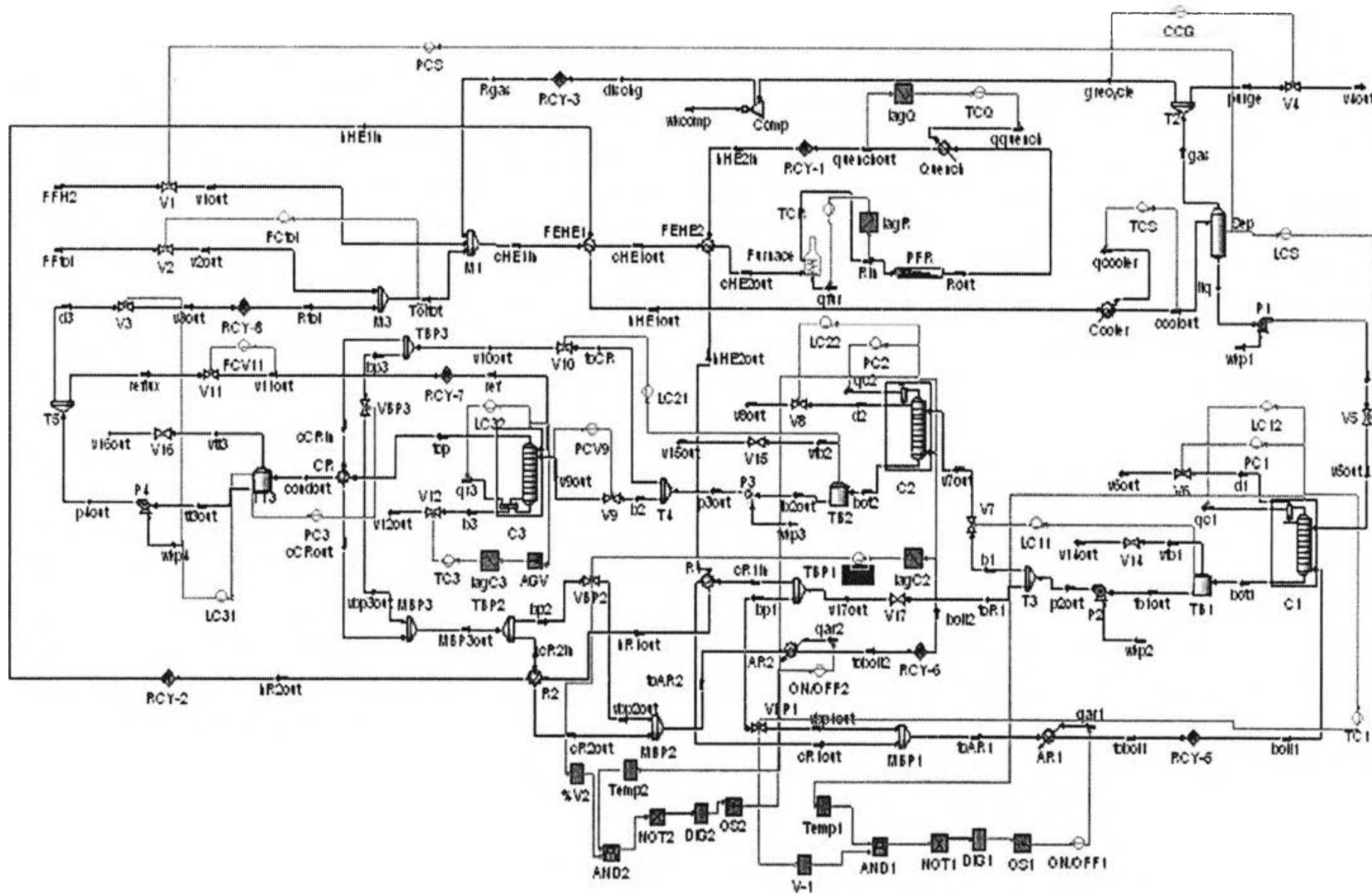


Figure 4.15. Application of control structure 2 to HDA plant alternative 5

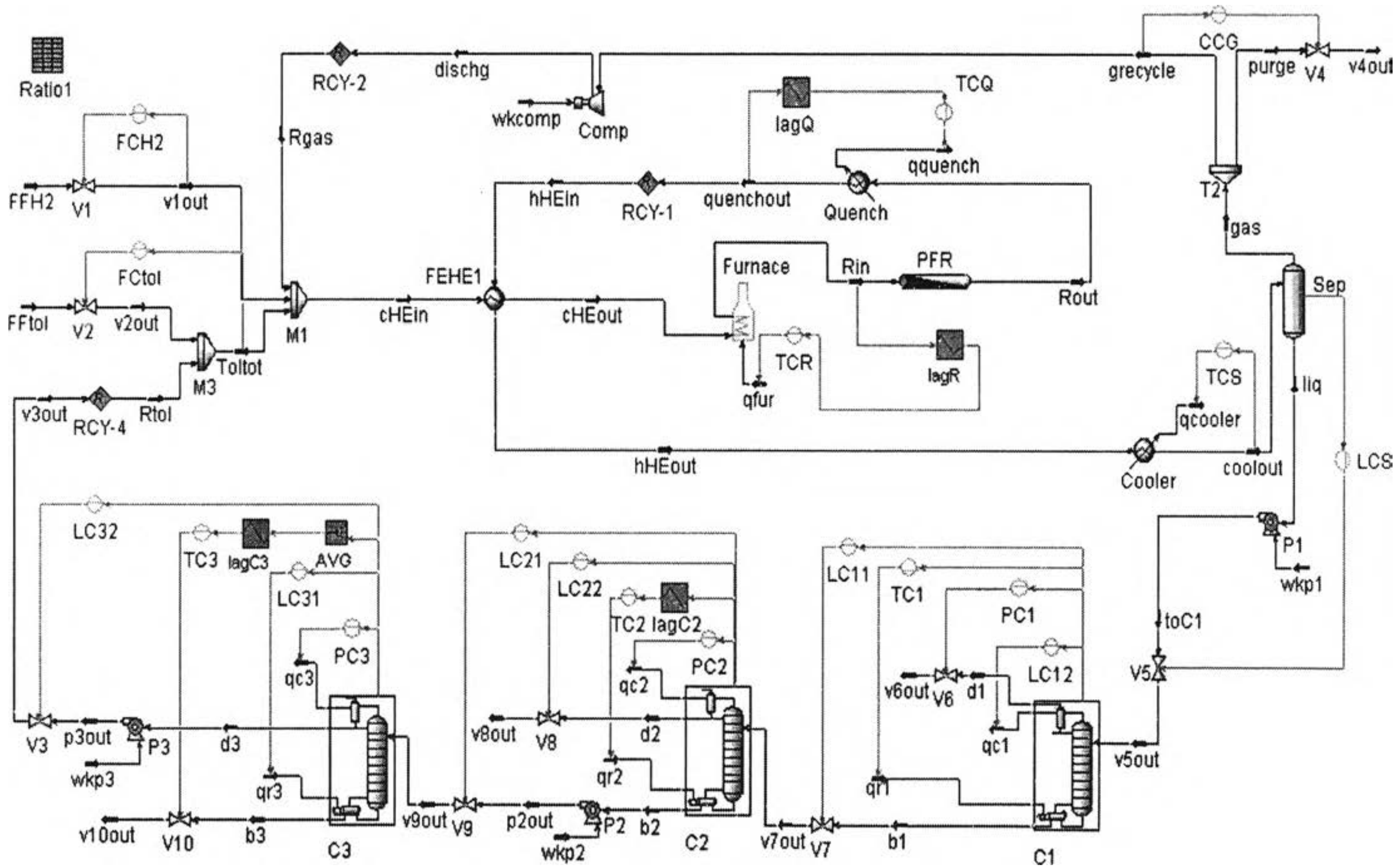


Figure 4.16 Application of control structure 3 to HDA plant alternative 1

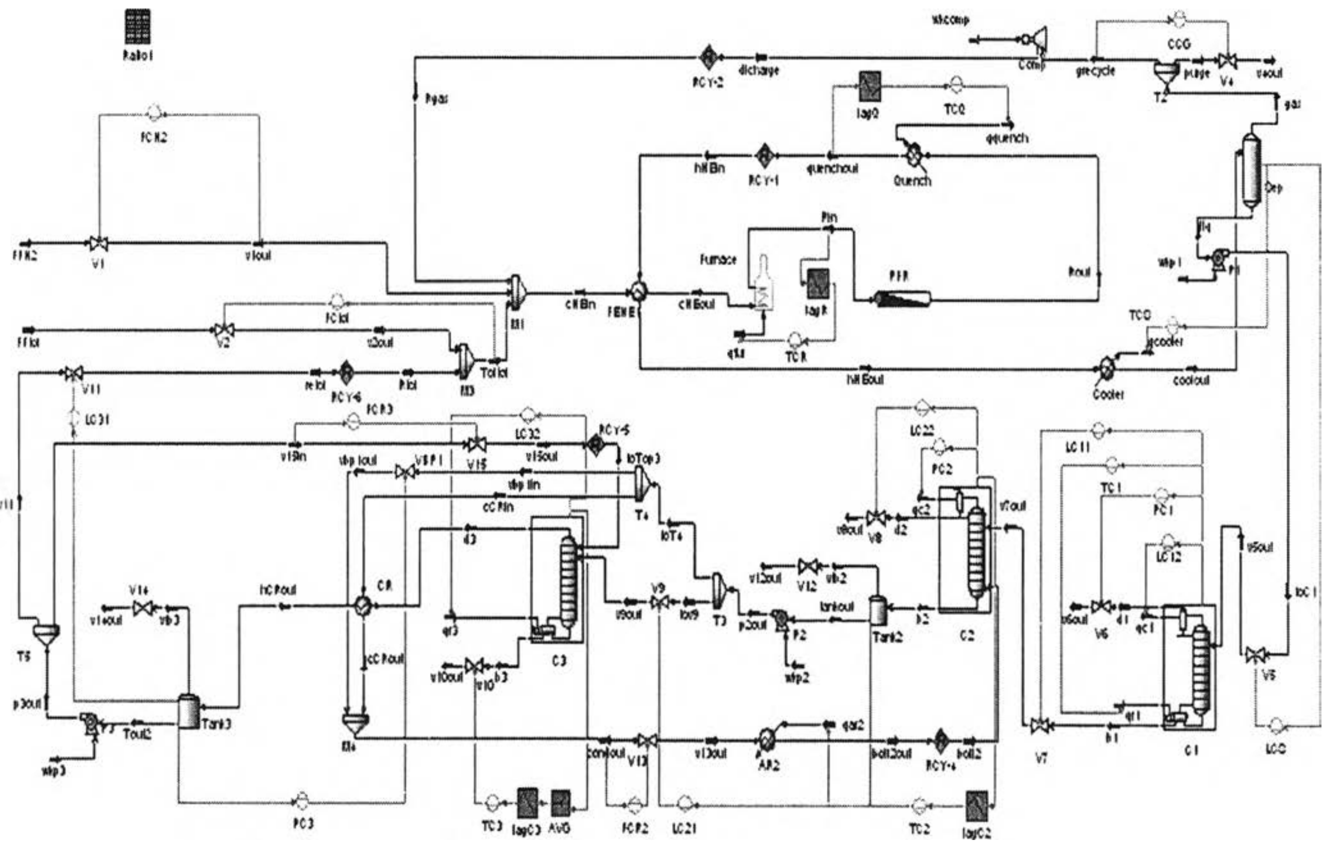


Figure 4.17 Application of control structure 3 to HDA plant alternative 2

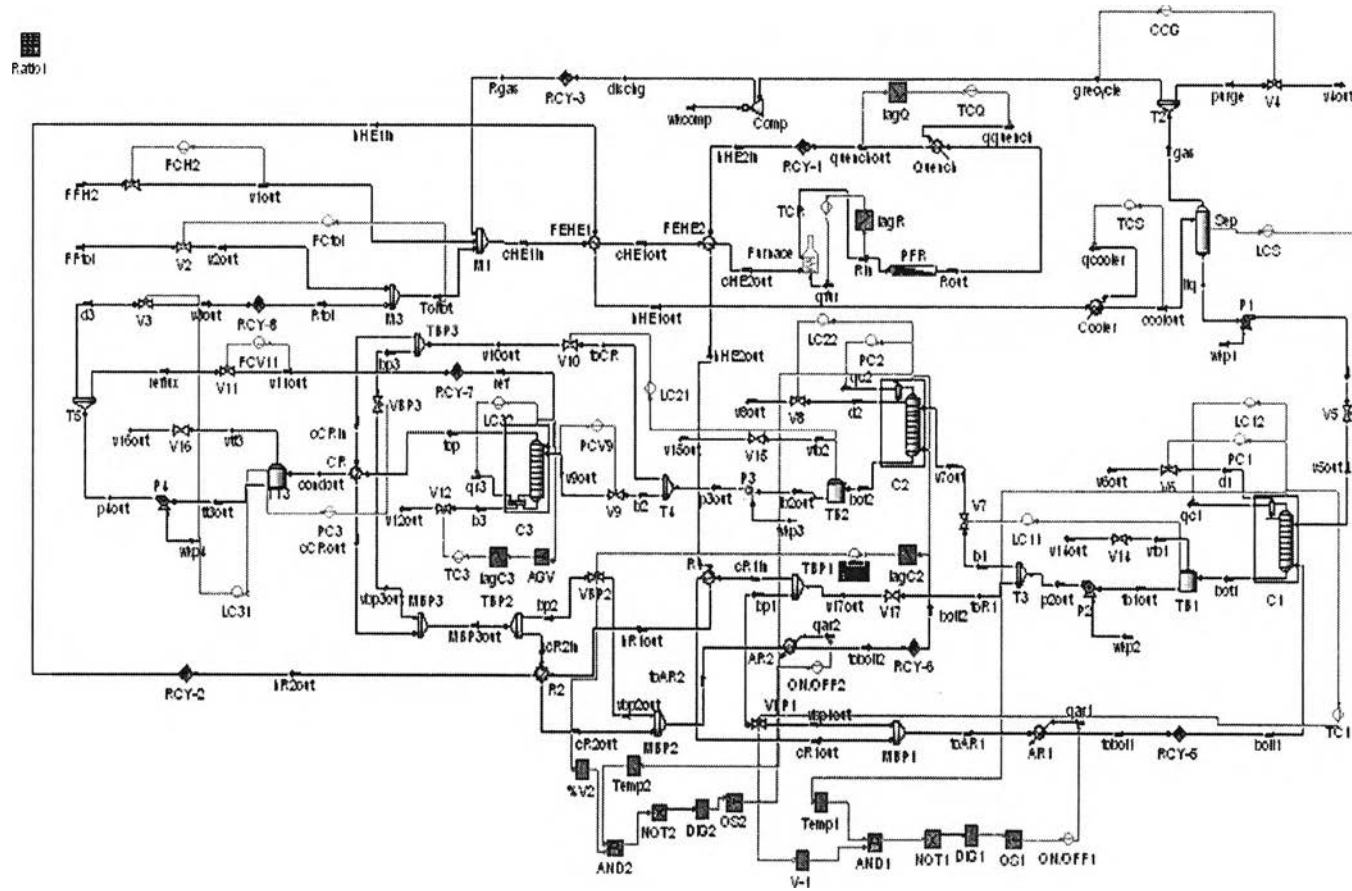


Figure 4.18 Application of control structure 3 to HDA plant alternative 5

4.6 Dynamic simulation results

In order to illustrate the dynamic behaviors of the new control structures in HDA plant alternatives 1, 2 and 5 several disturbance loads were made. The dynamic responses of our control structure are shown in Figures 4.19 to 4.24. In general, CS1 has better responses of utility consumptions are achieved here compared to CS2 and CS3. Results for individual disturbance load changes are as follows:

4.6.1 Change in the heat load disturbance of cold stream for HDA plant alternative 1

Figure 4.19 shows the dynamic responses of the HDA plant alternative 1 to a change in the heat load disturbance of the cold stream (reactor feed stream).This disturbance is made as follows: first the fresh toluene feed temperature is decreased from 30 to 20 oC at time equals 10 minutes, and the temperature is increased from 20 to 40 oC at time equals 100 minutes, then its temperature is returned to its nominal value of 30 oC at time equals 200 minutes .

The dynamic response of control structure 1 same as CS2 and CS3 but CS3 control system can handle more disturbance and faster than other. As can be seen,in our study the reactor inlet temperature (Figure4.19.a) , the reactor oulet temperature(Figure4.19.b), and the separator temperature (FIgure4.19.c) are slightly well controled. But, for CS1 control system has more oscillations occur in the tray temperature of stabilizer column (Figure4.19.e)and tray temperature of recycle column (Figure4.19.g)

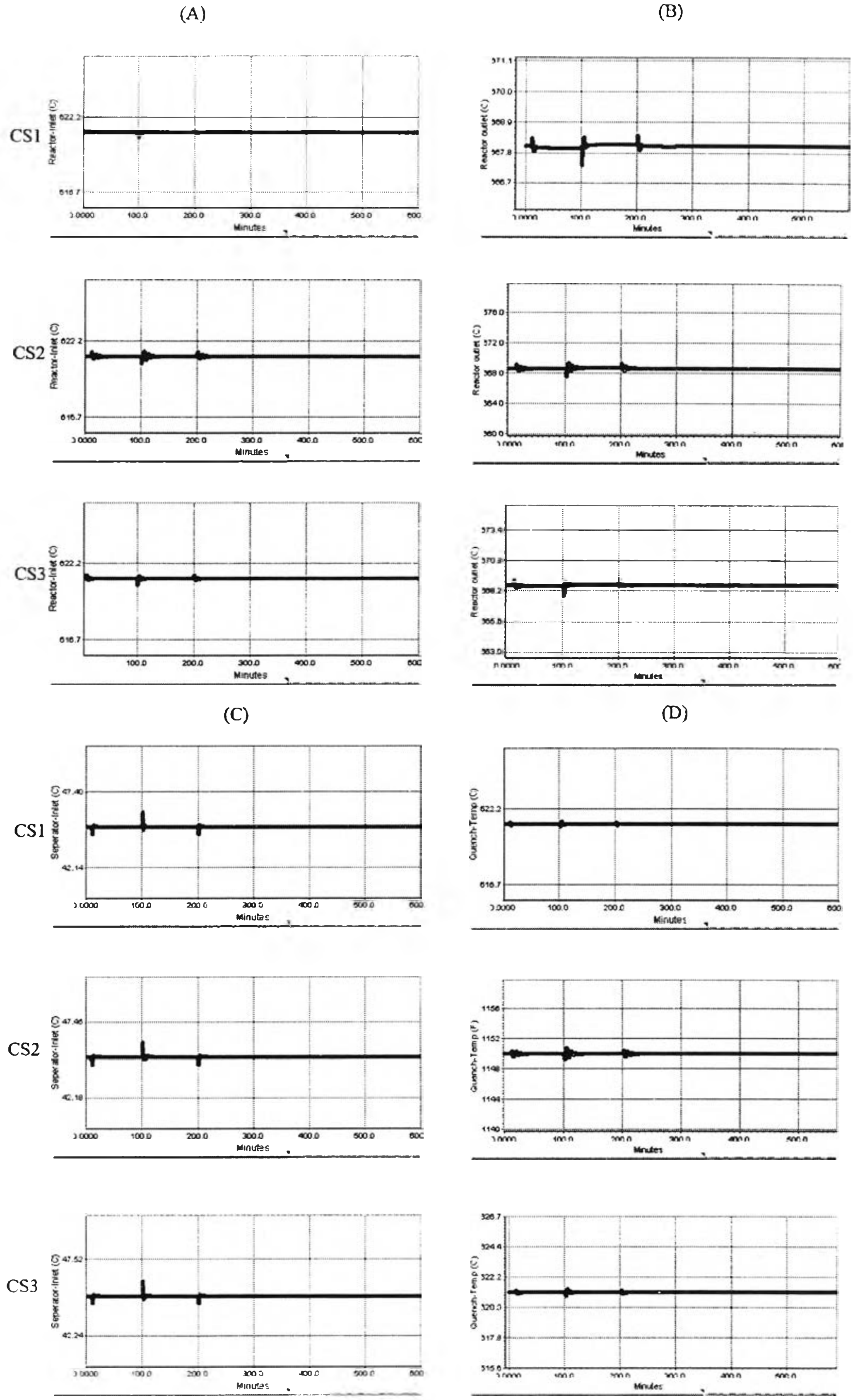


Figure 4.19: Dynamic responses of the HDA plant alternative 1 to a change in the heat load disturbance of cold stream (reactor feed stream), where (a) reactor inlet temperature, (b) the reactor oulet temperature (C) separator temperature , (d) quench temperature , (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 , CS2 and CS3.

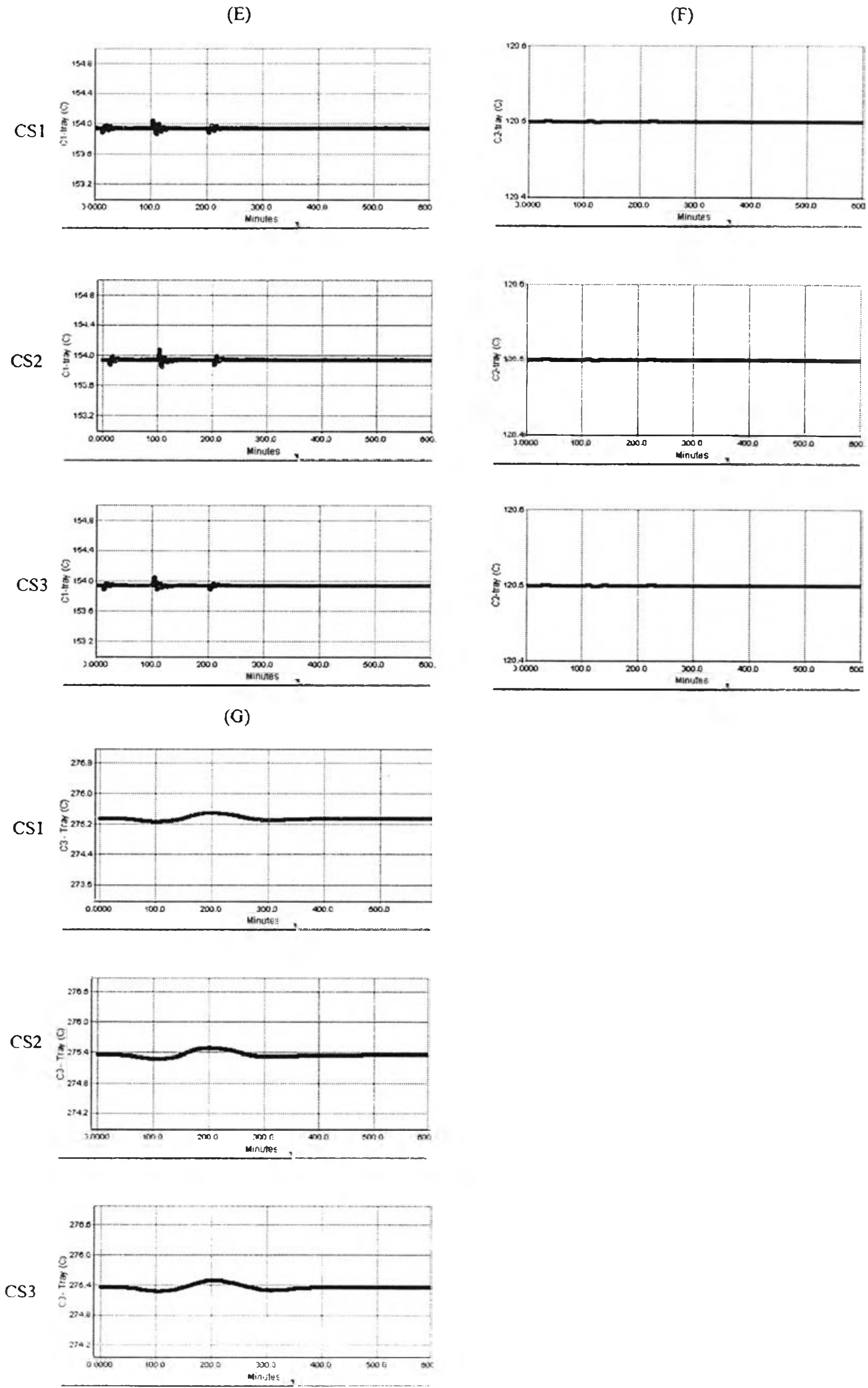


Figure 4.19: Continued.

4.6.2 Change in the heat load disturbance of cold stream for HDA plant alternative 2 and 5

Figure 4.20 and 4.21 shows the dynamic responses of the HDA plant alternative 2 and 5 to a change in the heat load disturbance of the cold stream (reactor feed stream).This disturbance is made as follows: first the fresh toluene feed temperature is decreased from 30 to 20 oC at time equals 10 minutes, and the temperature is increased from 20 to 40 oC at time equals 100 minutes, then its temperature is returned to its nominal value of 30 oC at time equals 200 minutes.

As can be seen,The dynamic response of HDA process alternative 2 are slower than those in HDA process alternative 1 .In our study the reactor inlet temperature (Figure4.20.a) , the reactor oulet temperature(Figure4.20.b), and the separator temperature (FIgure4.20.c) are slightly well controled. But, for tray temperature of recycle column (Figure4.20.g) has more oscillations occur.

In HDA process alternative 4, the dynamic response are slower than previous case (i.e. HDA process alternative 3 ,2 , and 1). For CS1 and CS2 control system most temperature loop are slightly well controled, for CS1 control system has more oscillations occur in reactor inlet temperature , the reactor oulet temperature , the separator temperature , tray temperature of product column , the tray temperature of stabilizer , and tray temperature of recycle column .

In HDA process alternative 5, the dynamic response aore slower than previous case (i.e. HDA process alternative 2 and 1). For CS1 control system has more oscillations occur in reactor inlet temperature and the reactor oulet temperature . The dynamic response of tray temperature of recycle column for CS1 similary as CS2 structure and CS3, this control loop the response has more oscillations occur and slower to returned its setpoint.

For complex heat integration plant more oscillations occur in the tray temperature of stabilizer column, tray temperature of product column and tray temperature of recycle column. Those results indicate that the implementation of complex energy integration to the process deteriorates the dynamic performance of the process. CS3 control system can handle more disturbance and faster than other, but for first control system has better responses of utility consumptions are achieved here compared to CS2 and CS3 because in CS2 and CS3 control system as modified from the first control system by adding a cooling unit to control the outlet temperature from reactor, instead of using internal process flow. So, first control system require less furnace utility consumtions are achieved compare to other control system.

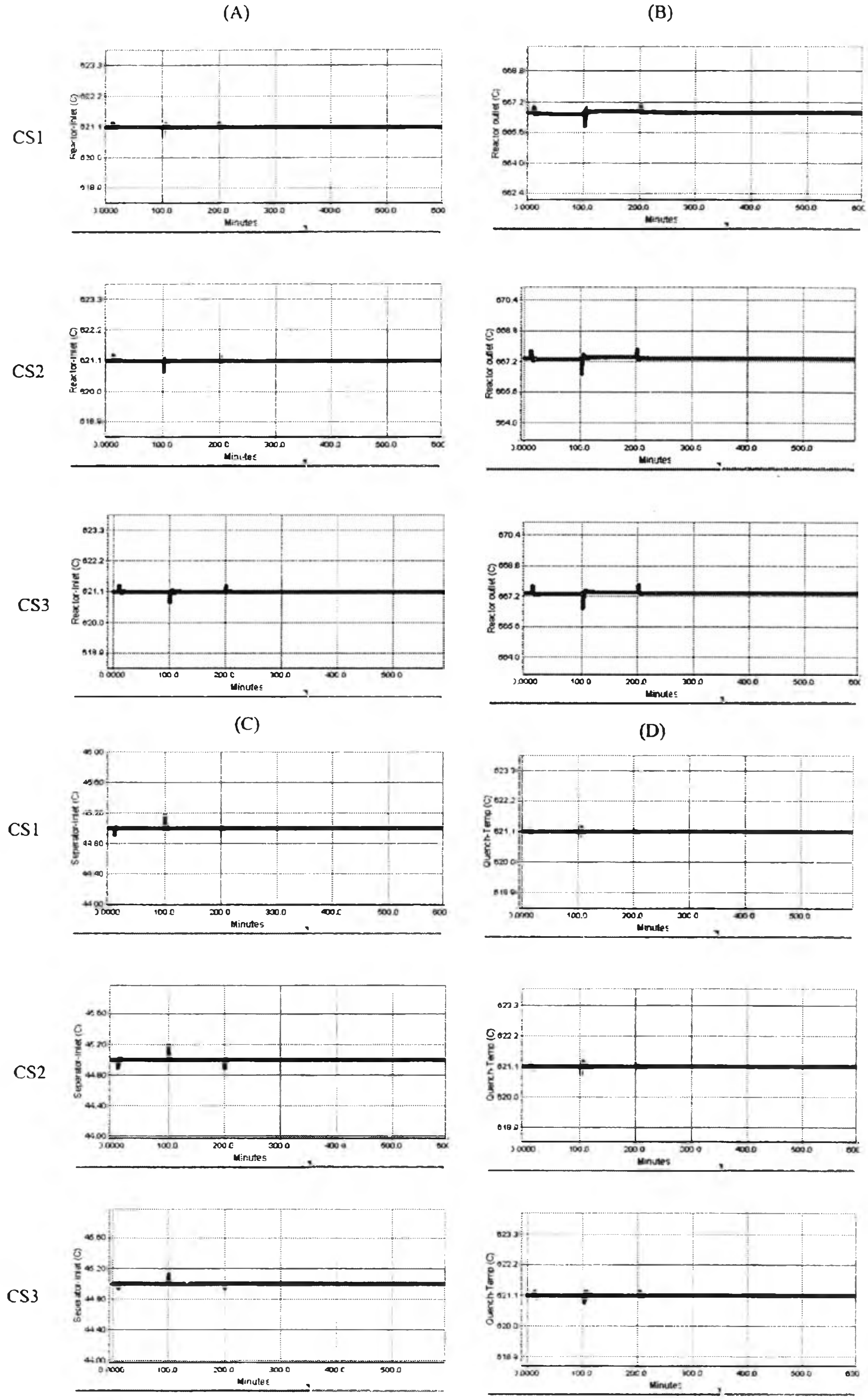


Figure 4.20: Dynamic responses of the HDA plant alternative 2 to a change in the heat load disturbance of cold stream (reactor feed stream), where (a) reactor inlet temperature, (b) the reactor oulet temperature (C) separator temperature , (d) quench temperature , (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 , CS2 and CS3.

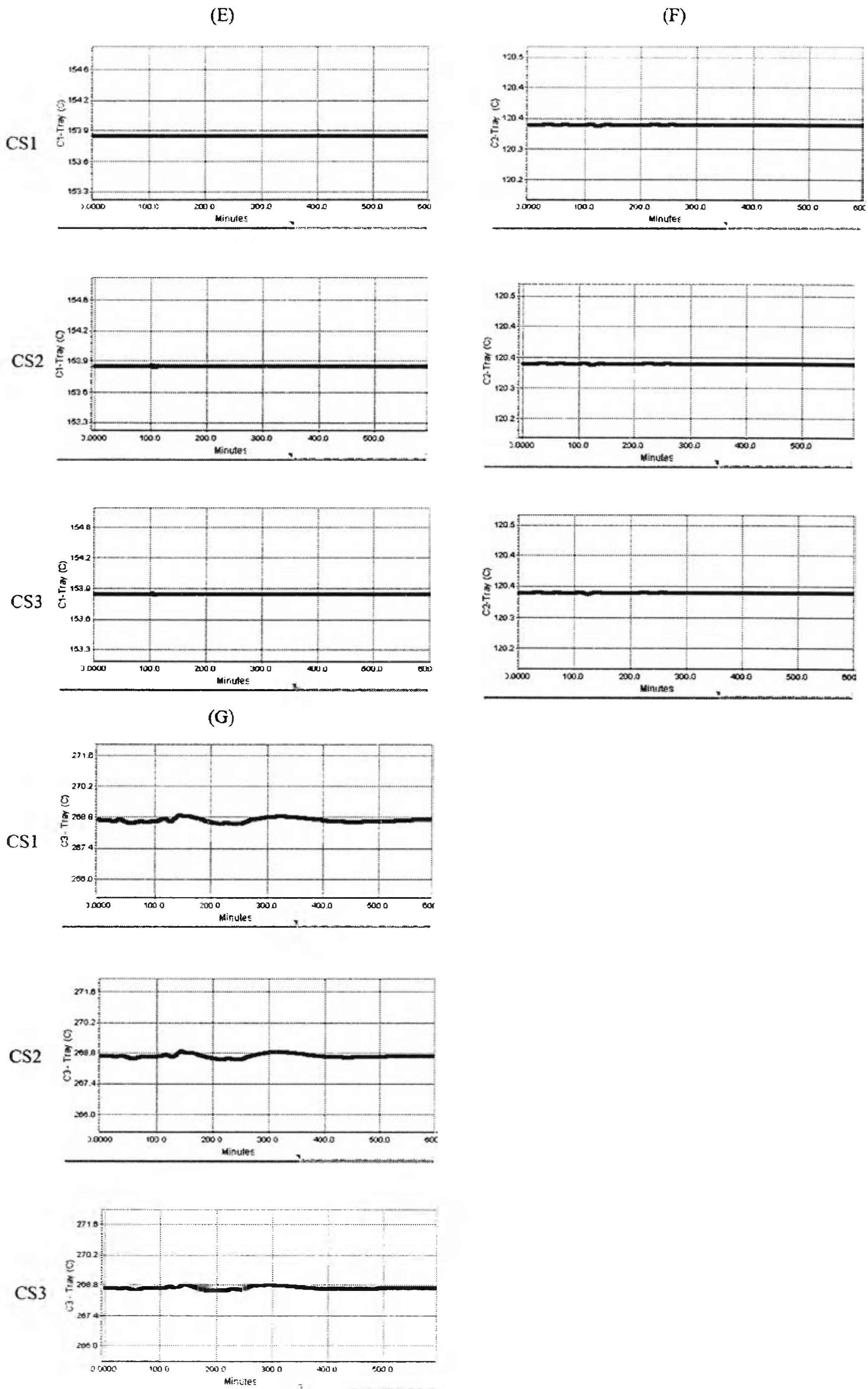


Figure 4.20: Continued.

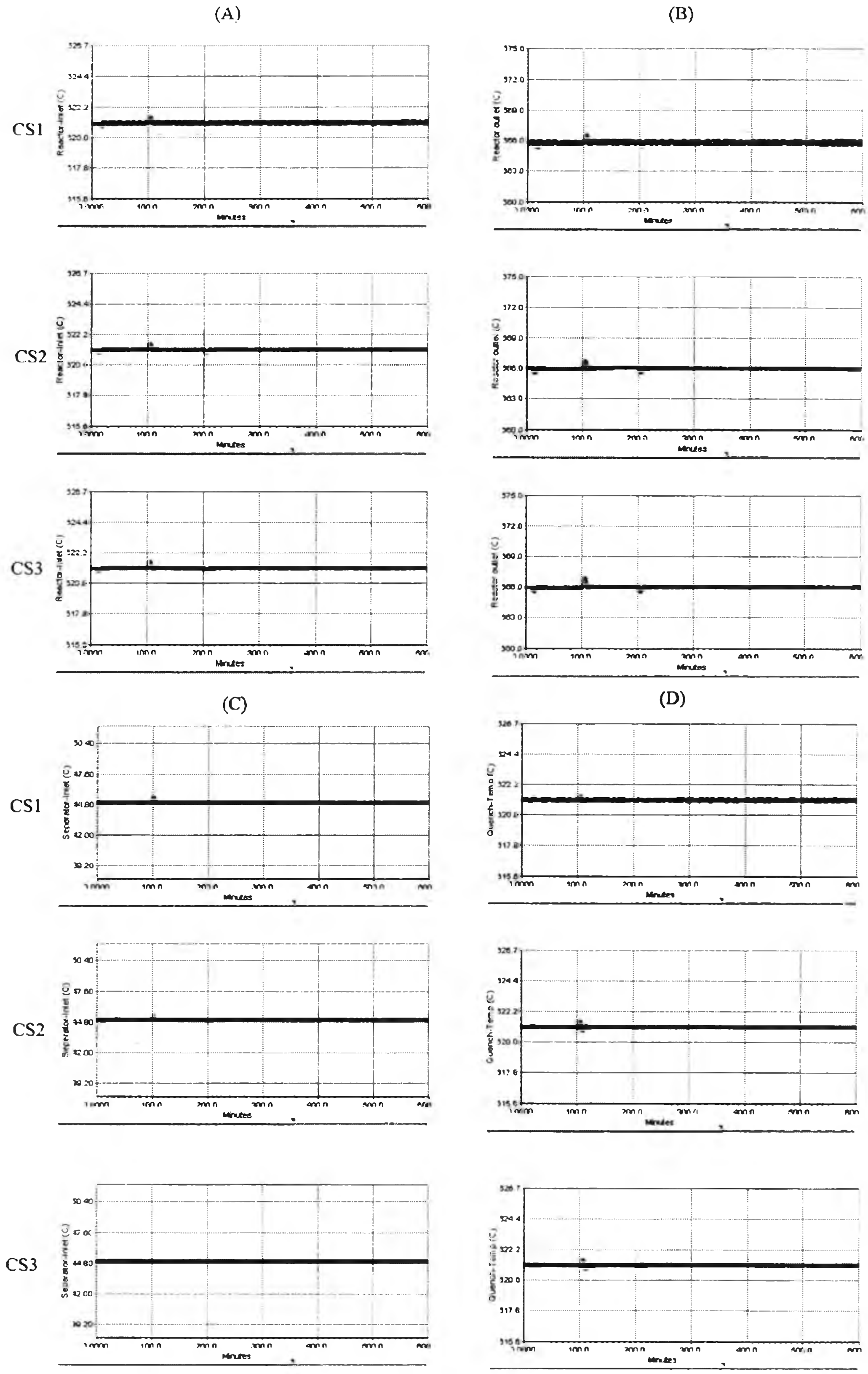


Figure 4.21: Dynamic responses of the HDA plant alternative 5 to a change in the heat load disturbance of cold stream (reactor feed stream), where (a) reactor inlet temperature, (b) the reactor oulet temperature (C) separator temperature , (d) quench temperature , (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 , CS2 and CS3.

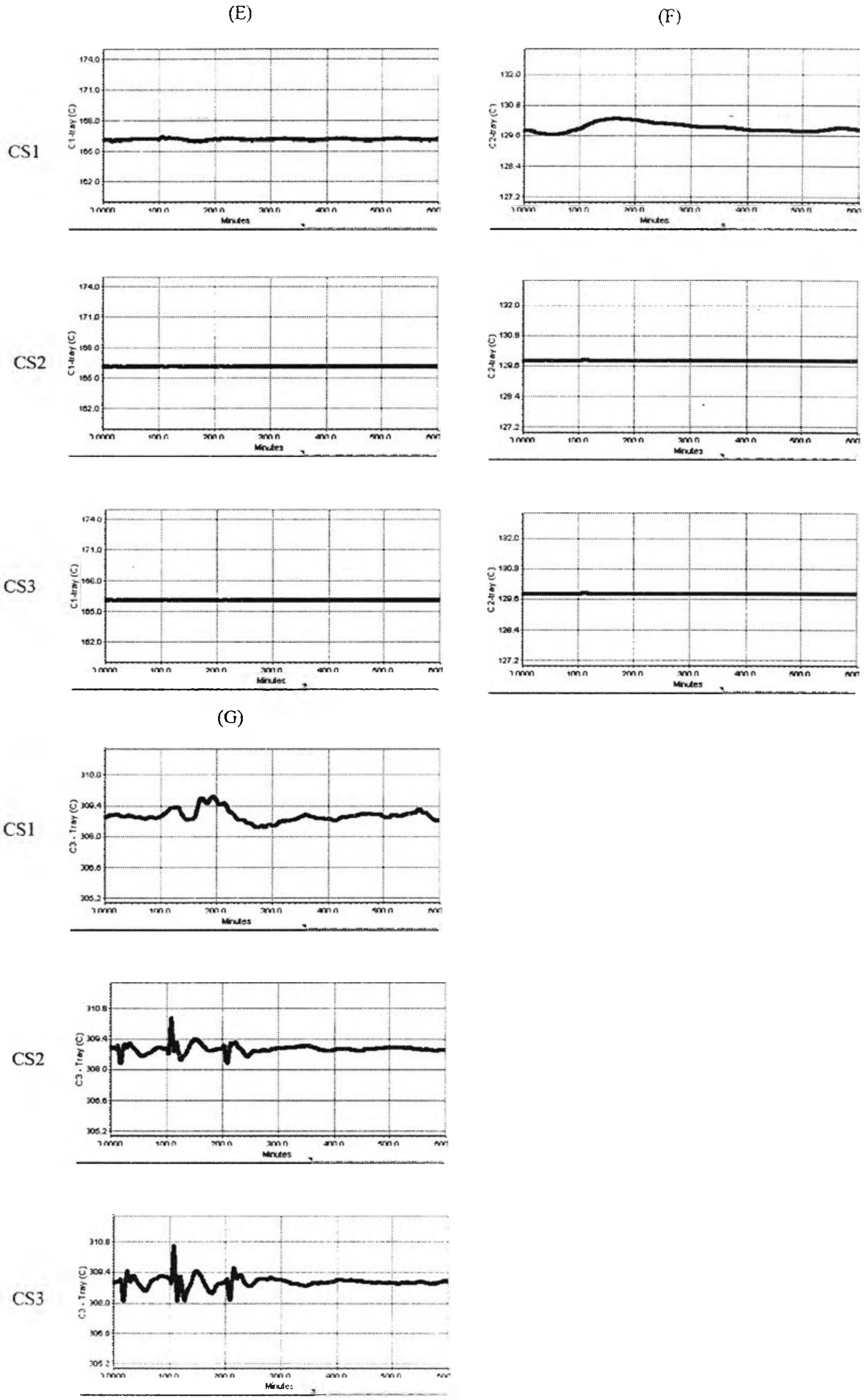


Figure 4.21: Continued.

4.6.3 Change in the recycle toluene flowrates for HDA plant alternative 1

On the other case, a disturbance in the production rate is also made for this study. Figure 4.22 shows the dynamic responses of the HDA plant alternative 1 to a disturbance in the recycle toluene flowrates from 168.6 to 158.6 kgmole/h at time equals 10 minutes, and the flowrates is increased from 158.6 to 178.6 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 168.6 kgmole/h at time equals 200 minutes.

The dynamic response of control structure 1 when change in the recycle toluene flowrates for HDA plant alternative 1 are similar with those to change in the heat load disturbance of cold stream case.As can be seen, in this case has more oscillations occur in the most of temperature control loop are compare with previous case . The tray temperature of recycle column (Figure4.22.g) has a large deviation.

The dynamic response of control structure 1 same as CS2 and CS3 but CS3 control system can handle more disturbance and faster than other. As can be seen,in our study the reactor inlet temperature (Figure4.22.a) , the reactor oulet temperature(Figure4.22.b), and the separator temperature (FIgure4.22.c) are slightly well controled.

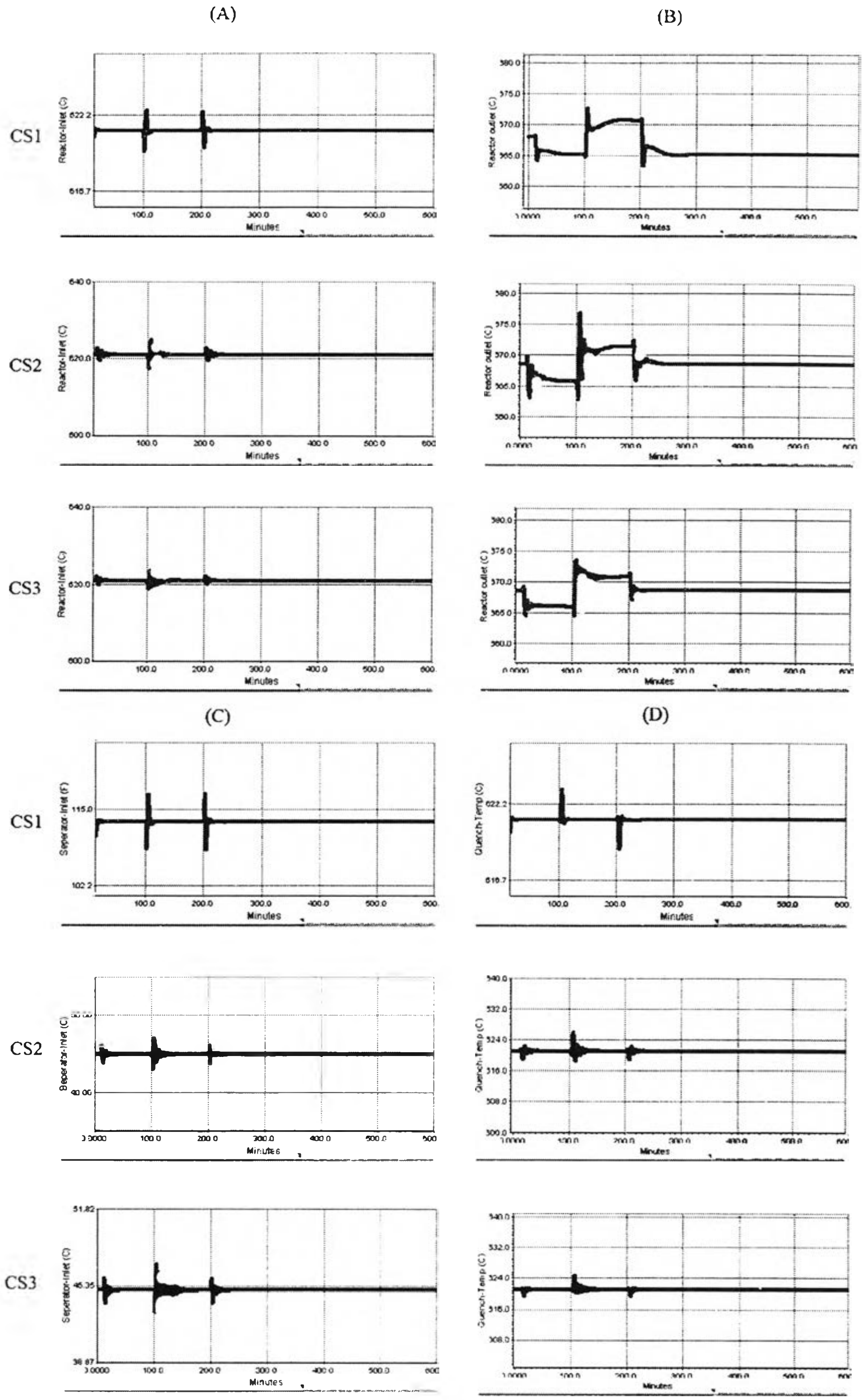


Figure 4.22: Dynamic responses of the HDA plant alternative 1 to a change in the recycle toluene flowrates. where (a) reactor inlet temperature. (b) the reactor oulet temperature (C) separator temperature , (d) quench temperature , (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 , CS2 and CS3.

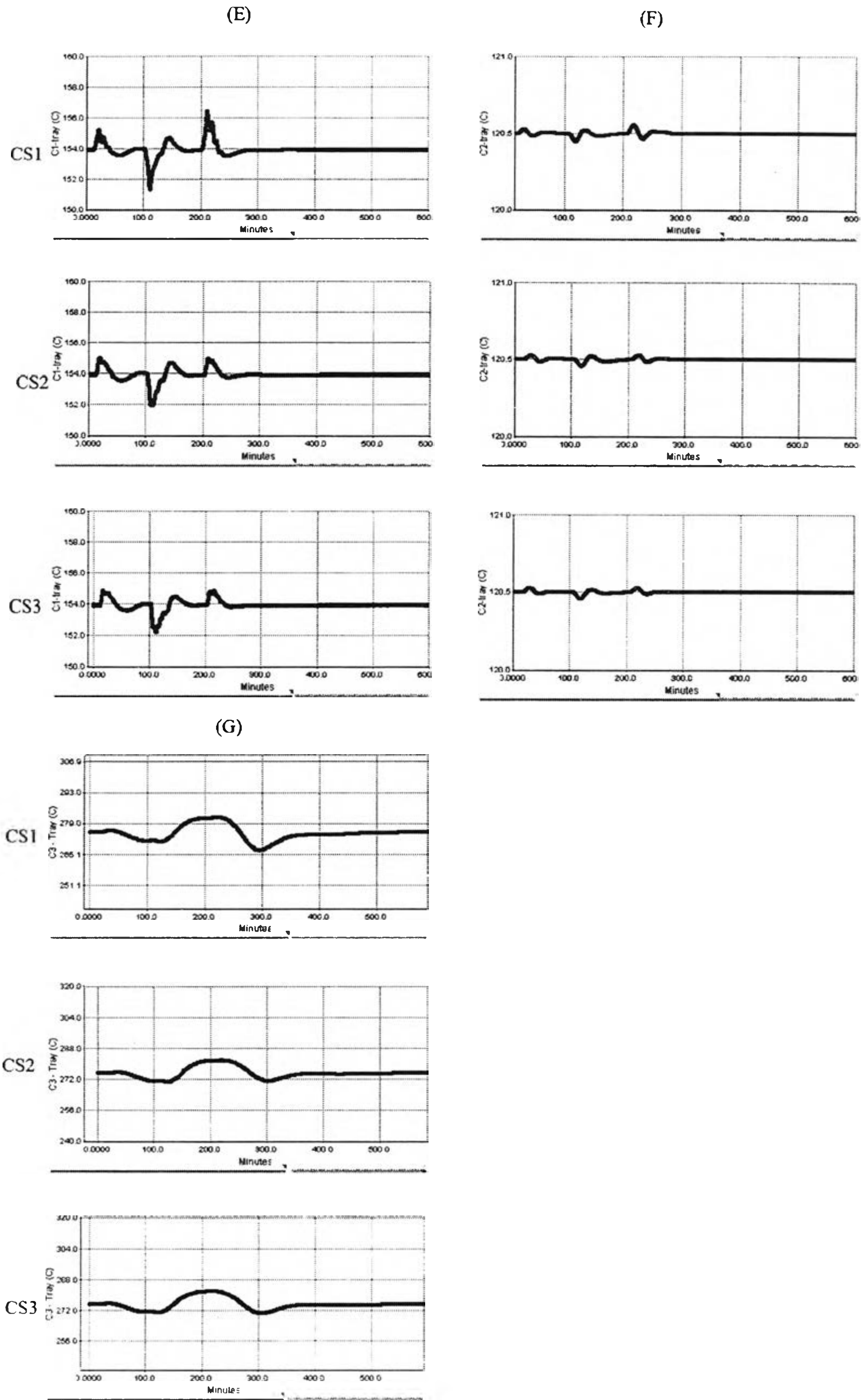


Figure 4.22: Continued.

4.6.8 Change in the recycle toluene flowrates for HDA plant alternative 2 and 5

On the other case, a disturbance in the production rate is also made for this study. Figure 4.23 and 4.24 shows the dynamic responses of the HDA plant alternative 2,3,4,5 and 6 to a disturbance in the recycle toluene flowrates from 168.6 to 158.6 kgmole/h at time equals 10 minutes, and the flowrates is increased from 158.6 to 178.6 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 168.6 kgmole/h at time equals 200 minutes.

As can be seen,The dynamic response of HDA process alternative 2 are slower than those in HDA process alternative 1 .In our study the reactor inlet temperature (Figure4.23.a) , the reactor oulet temperature(Figure4.23.b), and the separator temperature (FIgure4.23.c) are slightly well controled. But, for tray temperature of recycle column (Figure4.23.g) has more oscillations occur.The tray temperature of column it takes long time to return to it nominal value.

In HDA process alternative 5,the dynamic response are slower than previous case (i.e. HDA process alternative 2 and 1), for CS1 control system has more oscillations occur in reactor inlet temperature , the separator temperature , the tray temperature of stabilizer , tray temperature of product column. The dynamic response of tray temperature of recycle column for CS1 similary as CS2 structure and CS3, this control loop the response has more oscillations occur and slower to returned its setpoint.

In this case has more oscillations occur in the most of temperature control loop are compare with change in the heat load disturbance of cold stream case . For complex heat integration plant more oscillations occur in the tray temperature of stabilizer column, tray temperature of product column and tray temperature of recycle column. In tray temperature of recycle column has a large deviation and it takes long time to return to it nominal value. Those results indicate that the implementation of complex energy integration to the process deteriorates the dynamic performance of the process. CS3 control system can handle more disturbance and faster than other, but for first control system has better responses of utility consumptions are achieved here compared to CS2 and CS3 because in CS2 and CS3 control system as modified from the first control system by adding a cooling unit to control the outlet temperature from reactor, instead of using internal process flow. So, first control system require less furnace utility consumtions are achieved compare to other control system.

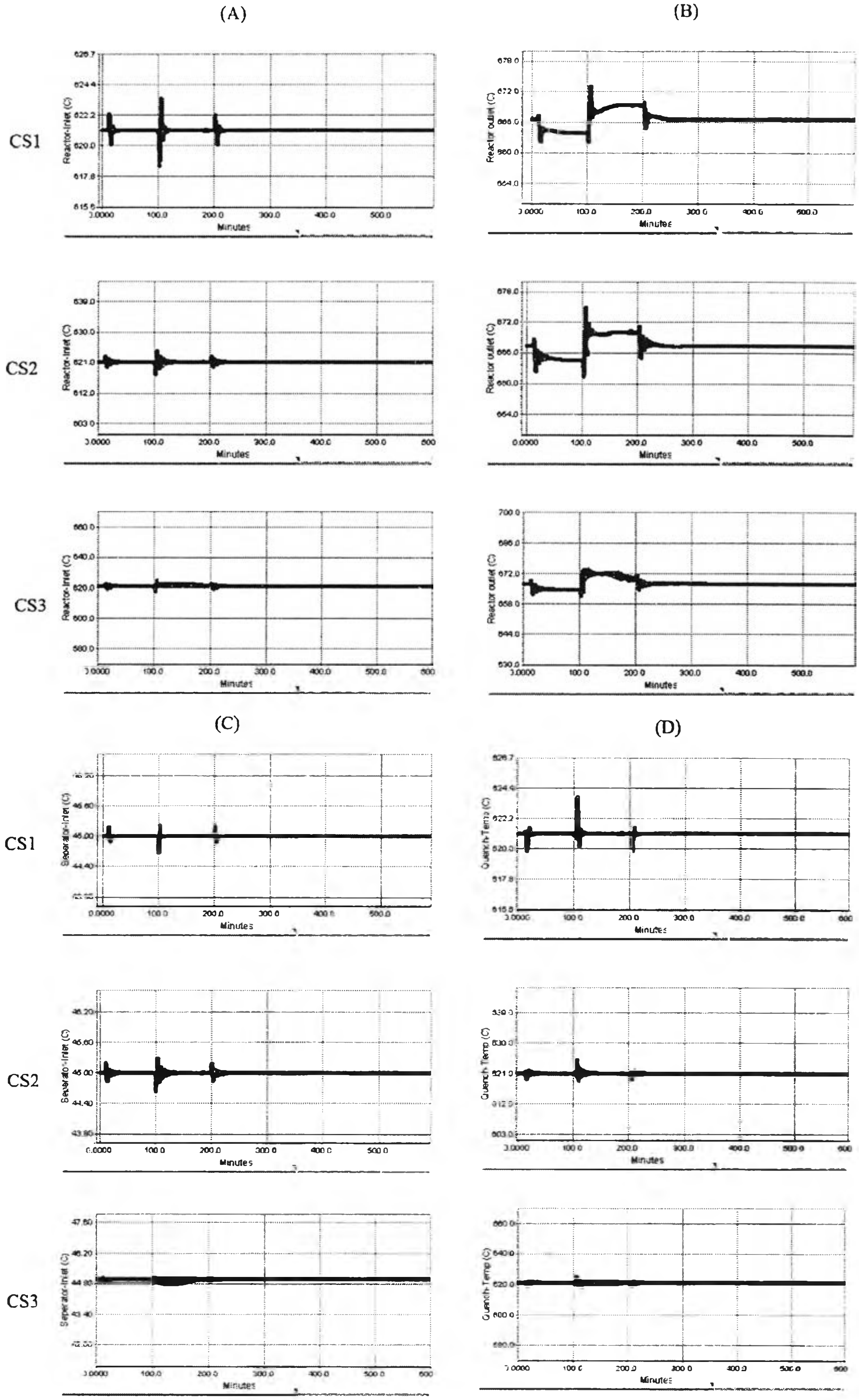


Figure 4.23: Dynamic responses of the HDA plant alternative 2 to a change in the recycle toluene flowrates. where (a) reactor inlet temperature. (b) the reactor oulet temperature (C) separator temperature . (d) quench temperature , (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 . CS2 and CS3.

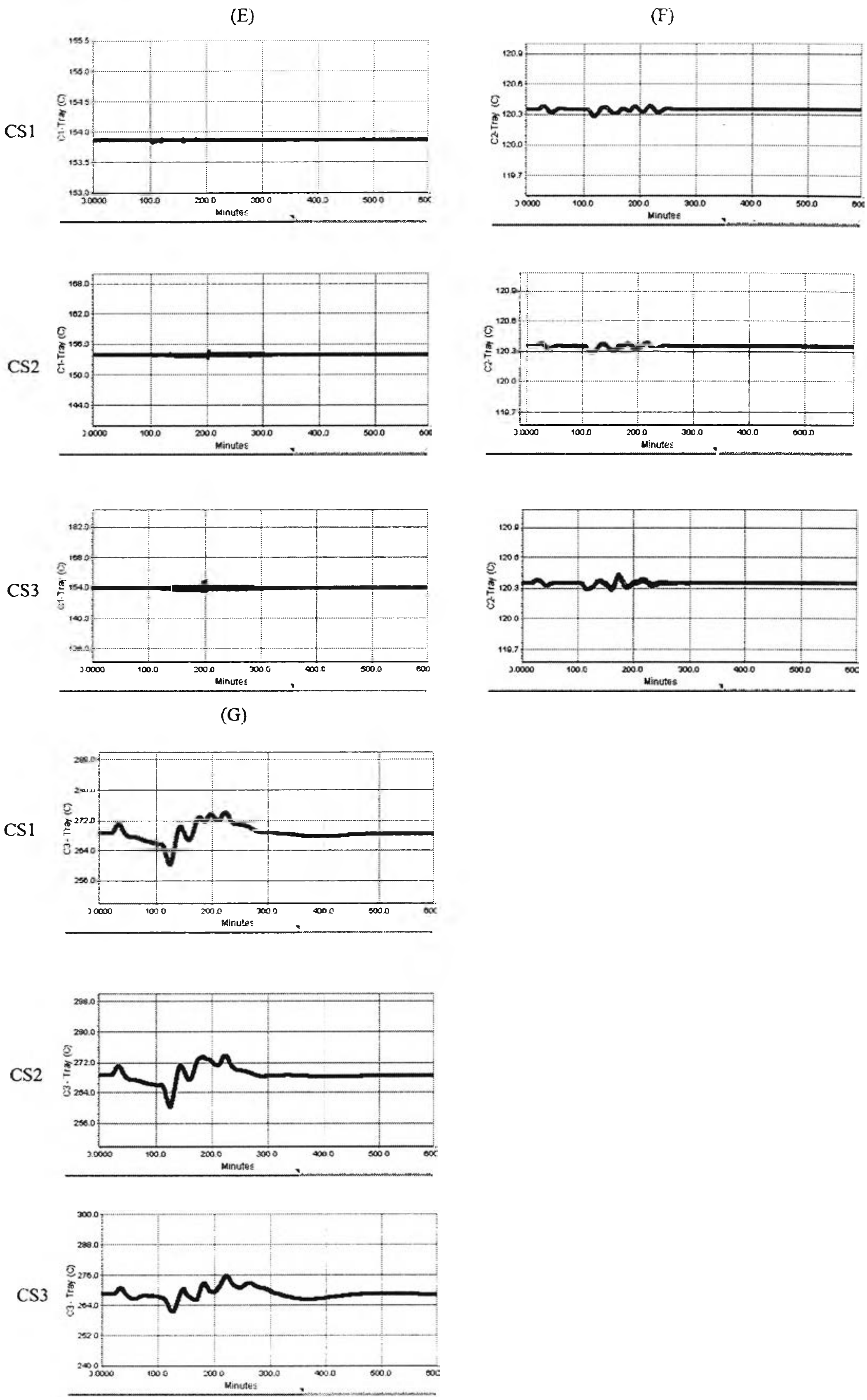


Figure 4.23: Continued.

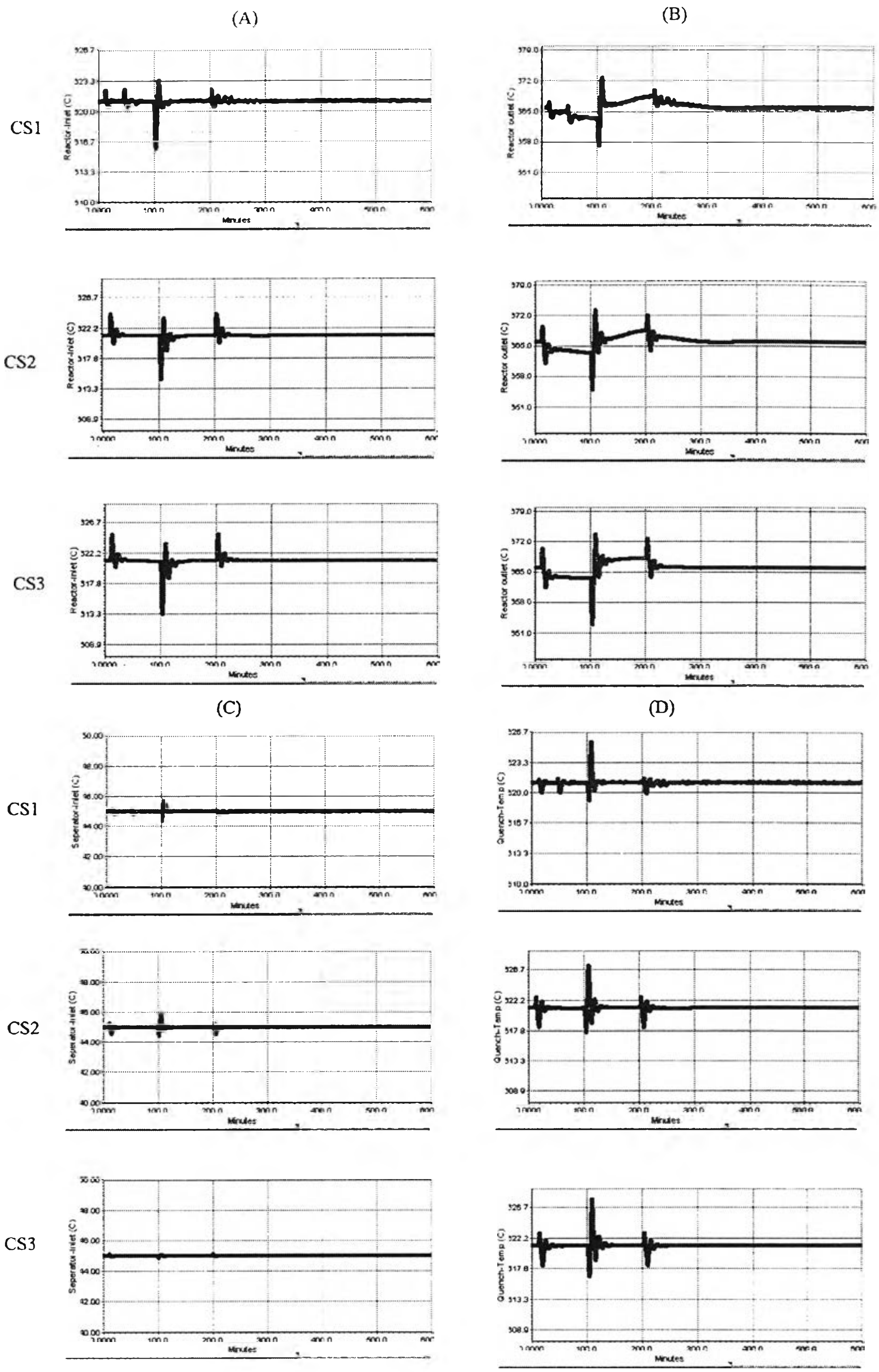
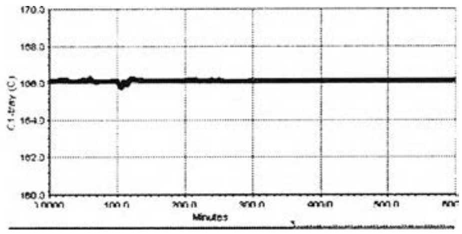


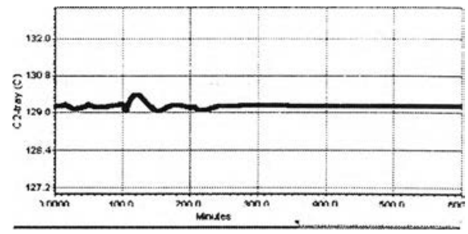
Figure 4.24: Dynamic responses of the HDA plant alternative 5 to a change in the recycle toluene flowrates. where (a) reactor inlet temperature. (b) the reactor oulet temperature (C) separator temperature . (d) quench temperature . (e)tray temperature of stabilizer column , (f)tray temperature of product column , (g) tray temperature of recycle column; comparison between CS1 , CS2 and CS3.

(E)

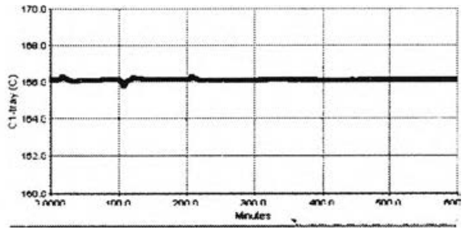
(F)

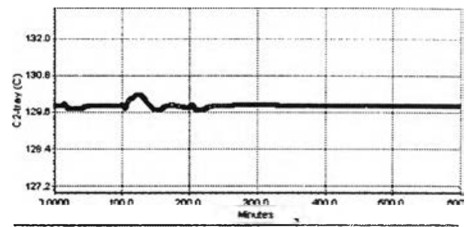
CS1



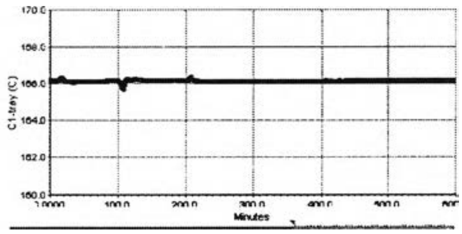


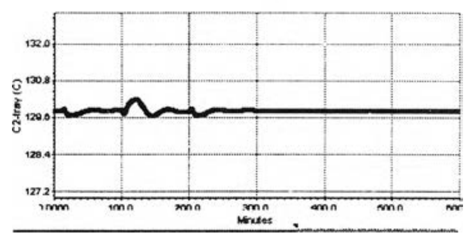
CS2





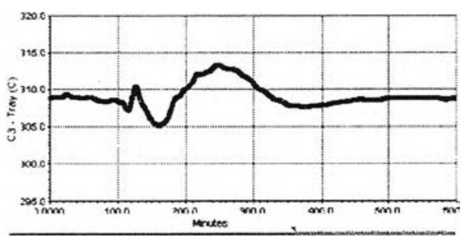
CS3



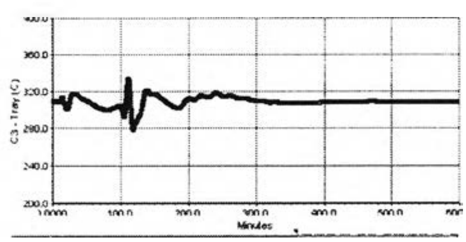


(G)

CS1



CS2



CS3

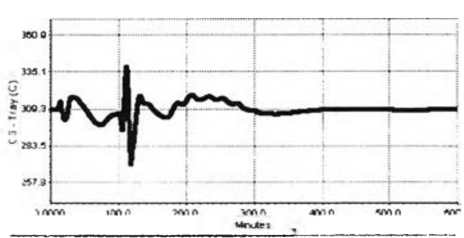


Figure 4.24: Continued.

4.7 Evaluation of the dynamic performance

The estimation of the minimum achievable variance of SISO controlled variable from 'normal' closed- loop data. Since then, minimum variance control has been widely used as a benchmark for assessing control loop performance. However, minimum variance control based performance assessment methods cannot adequately evaluate the performance for controllers with constraints explicitly incorporated or for controllers where transient response and deterministic disturbance regulation are concerned. For assessing constrained control loop performance the proposed dynamic performance index is focused on time related characteristics of the controller's response to set-point changes or deterministic disturbances. There exist several candidate performance measures such as settling time and integral absolute error (IAE). Integral absolute error is widely used for the formulation of a dynamic performance as written below:

$\text{IAE} = \int |\epsilon(t)|\, dt$

In this study, IAE method is used to evaluate the dynamic performance of the designed control system. Table 4.3a to 4.5a shows the IAE results for the change in the disturbance loads of cold steam in HDA process with different energy integration schemes (alternative 1, 2 and 5) for CS1 control structure to CS3 control structure respectively, table 4.3b to 4.5b shows the IAE results for the change in the total toluene feed flowrates in HDA process with different energy integration schemes (alternative 1, 2 and 5) for CS1 control structure to CS3 control structure respectively.

4.7.1 Evaluation of the dynamic performance for CS1 control structure case

Table 4.3a and 4.3b shows the IAE results for the change in the disturbance loads of cold steam in HDA process and the IAE results for the change in the disturbance loads of cold steam in HDA process the IAE results for the change in the total toluene feed flowrates in HDA process respectively.

For the change in the disturbance loads of cold steam on HDA process case the control system of HDA process alternative 1 for CS1 control structure case is the most effective on compared with those in HDA process alternatives 2 and 5 i.e. the value of IAE in HDA process alternative 1 is smaller than those in alternatives 2 and 5.

As can be seen the similarity result between the change in the total toluene feed flowrates on HDA process case and change in the disturbance loads of cold steam on HDA process case, the value of IAE in HDA process alternative 1 is smaller than another alternatives.

Table 4.3a The IAE results of the CS1 control structure to a change in the disturbance load of cold stream (reactor feed stream)

	alternative 1	alternative 2	alternative 5
FCtol	3.9322	2.7887	0.983
TC1	1.6467	0.3617	4.689
TC2	0.1349	0.0678	3.497
TC3	12.099	23.814	47.402
TCS	2.1409	0.21913	5.363
TCQ	0.9391	1.45875	7.452
TCR	0.93475	1.0316	2.5871
sum	21.82755	29.74168	71.9731

Table 4.3b The IAE results of the CS1 control structure to a change the total toluene feed flowrates

	alternative 1	alternative 2	alternative 5
FCtol	40.944	35.989	44.452
TC1	69.1	44.4627	47.0152
TC2	2.1064	2.3572	12.393
TC3	533.316	568.12	1303.37
TCS	17.464	11.2492	7.2534
TCQ	11.463	13.079	40.488
TCR	9.8217	15.124	43.56
sum	684.2151	690.3811	1498.5316

4.7.2 Evaluation of the dynamic performance for CS2 control structure case

Table 4.4a and 4.4b shows the IAE results for the change in the disturbance loads of cold steam in HDA process and the IAE results for the change in the disturbance loads of cold steam in HDA process the IAE results for the change in the total toluene feed flowrates in HDA process respectively.

For the change in the disturbance loads of cold steam on HDA process case the control system of HDA process alternative 1 for CS2 control structure case is the most effective on compared with the others. the value of IAE in HDA process alternative 1 is smaller than those in alternatives 2 and 5. As can be seen the similarity result between the change in the total toluene feed flowrates on HDA process case and change in the disturbance loads of cold steam on HDA process case, the value of IAE in HDA process alternative 1 is smaller than another alternatives.

As can be seen that the IAE results for CS3 control structure look just the same as CS2 control structure results, but IAE results for CS1 control structure are larger than CS2 control structure.

Table 4.4a The IAE results of the CS2 control structure to a change in the disturbance load of cold stream (reactor feed stream)

	alternative 1	alternative 2	alternative 5
FCtol	4.0015	2.8313	0.0768
TC1	1.5725	0.124	3.2326
TC2	0.13815	0.0585	1.64105
TC3	12.099	22.981	40.588
TCS	2.1523	0.27362	3.7012
TCQ	0.8261	1.2175	5.6097
TCR	0.945	1.0869	1.0518
sum	22.73455	28.57282	55.90115

Table 4.4b The IAE results of the CS2 control structure to a change the total toluene feed flowrates

	alternative 1	alternative 2	alternative 5
FCtol	48.126	50.539	53.2618
TC1	56.213	46.82	45.8657
TC2	1.6549	2.358	11.056
TC3	417.45	448.41	1200.1
TCS	15.323	28.521	5.9338
TCQ	27.67	23.025	58.35
TCR	26.217	28.521	52.823
sum	593.6539	628.194	1427.3903

4.7.3 Evaluation of the dynamic performance for CS3 control structure case

Table 4.5a and 4.5b shows the IAE results for the change in the disturbance loads of cold steam in HDA process and the IAE results for the change in the disturbance loads of cold steam in HDA process the IAE results for the change in the total toluene feed flowrates in HDA process respectively.

For the change in the disturbance loads of cold steam on HDA process case the control system of HDA process alternative 1 for CS3 control structure case is the most effective on compared with those in HDA process alternatives 2 and 5, i.e. the value of

IAE in HDA process alternative 1 is smaller than those in alternatives 2 and 5. As can be seen the similarity result between the change in the total toluene feed flowrates on HDA process case and change in the disturbance loads of cold steam on HDA process case, the value of IAE in HDA process alternative 1 is smaller than another alternatives.

As can be seen that the IAE results for CS3 control structure look just the same as CS2 control structure results, but IAE results for CS2 control structure are larger than CS1 control structure. The performance of these control structures can be arranged from the best to lowest performance (error of controllability point of view) as the following sequences: CS3, CS2, and CS1.

Table 4.5a The IAE results of the CS3 control structure to a change in the disturbance load of cold stream (reactor feed stream)

	alternative 1	alternative 2	alternative 5
FCtol	4.0092	2.8765	0.0835
TC1	1.3124	0.11795	3.21765
TC2	0.13285	0.07645	1.6568
TC3	12.094	21.874	38.758
TCS	2.1534	0.12116	3.7443
TCQ	0.8838	1.2901	5.3879
TCR	1.0957	0.8968	1.10621
sum	22.68135	27.25296	53.95436

Table 4.5b The IAE results of the CS3 control structure to a change the total toluene feed flowrates

	alternative 1	alternative 2	alternative 5
FCtol	64.719	74.44	62.3759
TC1	50.608	47.82	45.8267
TC2	1.3617	2.678	11.042
TC3	406.532	423.87	1163.5
TCS	19.411	14.0951	3.67616
TCQ	22.812	23.7	55.773
TCR	24.764	22.77	56.103
sum	591.2077	609.3731	1398.29676

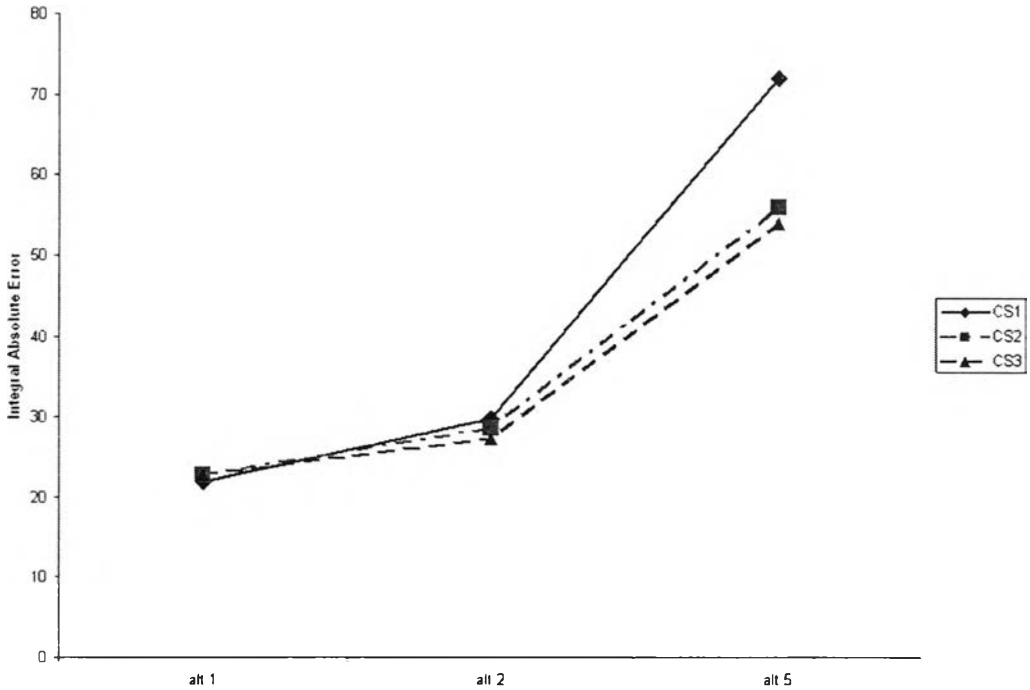


Figure 4.25: The IAE results of a change in the disturbance load of cold stream (reactor feed stream).

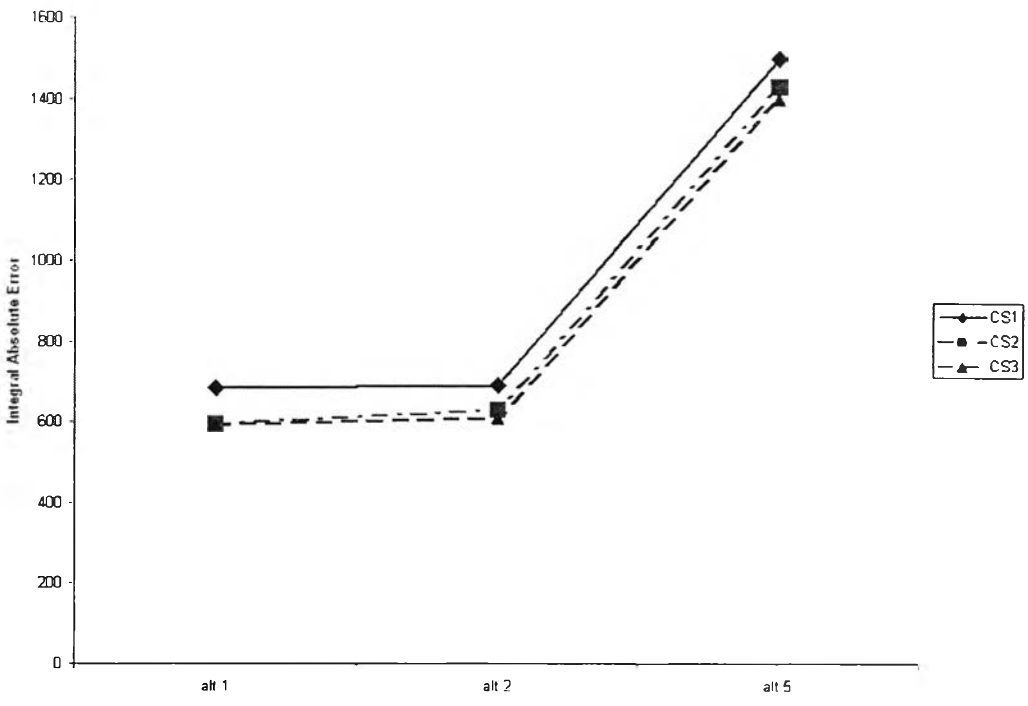


Figure 4.26: The IAE results of a change the total toluene feed flowrates.

4.8 Economic analysis for HDA process

A first study of the total processing costs to heat-exchanger network alternatives was undertaken by Terrill and Douglas (1987). They developed a Heat exchanger network for a base-case design for the HDA process. They also developed six alternative heat-exchanger networks.

From steady state pointview, on the evaluation of the economics of a HDA process. The term economics refers to the evaluation of capital costs and operating costs associated with the construction and operation of a HDA process. The methods by which the one-time costs associated with the construction of the plant and the continuing costs associated with the daily operation of the process are combined into meaningful economic criteria are provided. The benefit obtained from energy integration with the alternatives 1 to the others is given in Table 4.6. The energy cost savings from the energy integration fall between 4.68 and 22.66 %, but the capital cost rising are in the range from 2.20 to 25.56%.

Table 4.6 Results of cost estimation for HDA process with different energy integration schemes

Process alternative	Grass Roots Cost (US dollar)	Capital Cost increasing from alternative 1 (%)	Annual Utility Cost (US dollar)	Utility Cost Saving From Alternative 1 (%)
alternative 1	9,550,000	0.00	2,780,000	0.00
alternative 2	9,760,000	2.20	2,650,000	4.68
alternative 5	11,300,000	18.32	2,264,900	18.53

For evaluation of operating cost of control system are show in figure 4.47 - 4.50 control system has better responses of utility consumptions are achieved here compared to CS2 and CS3.Because both CS2 and CS3 control system require more furnace and quench utility compared to CS1. Economic analysis shows that the improved energy integration has allowed us to increase the recycle flows. The increased recycle flows actually decrease the utilities consumption.

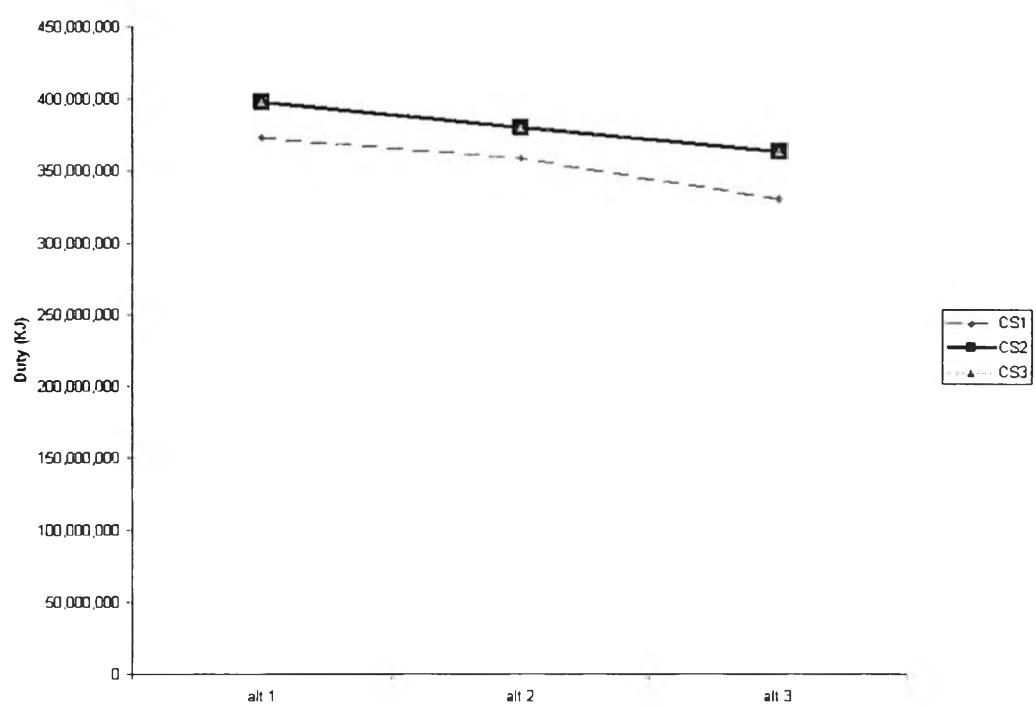


Figure 4.27: The utility consumptions (exclude cooler and quench duty) of HDA process when change in the disturbance load of cold stream (reactor feed stream).

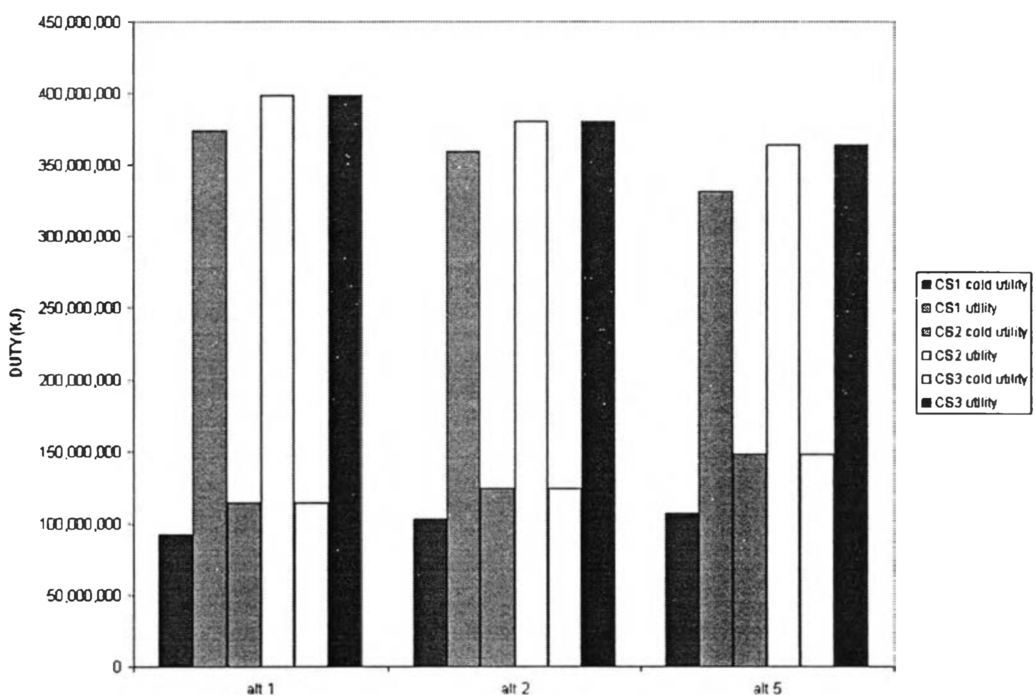


Figure 4.28: The utility consumptions of HDA process when change in the disturbance load of cold stream (reactor feed stream).

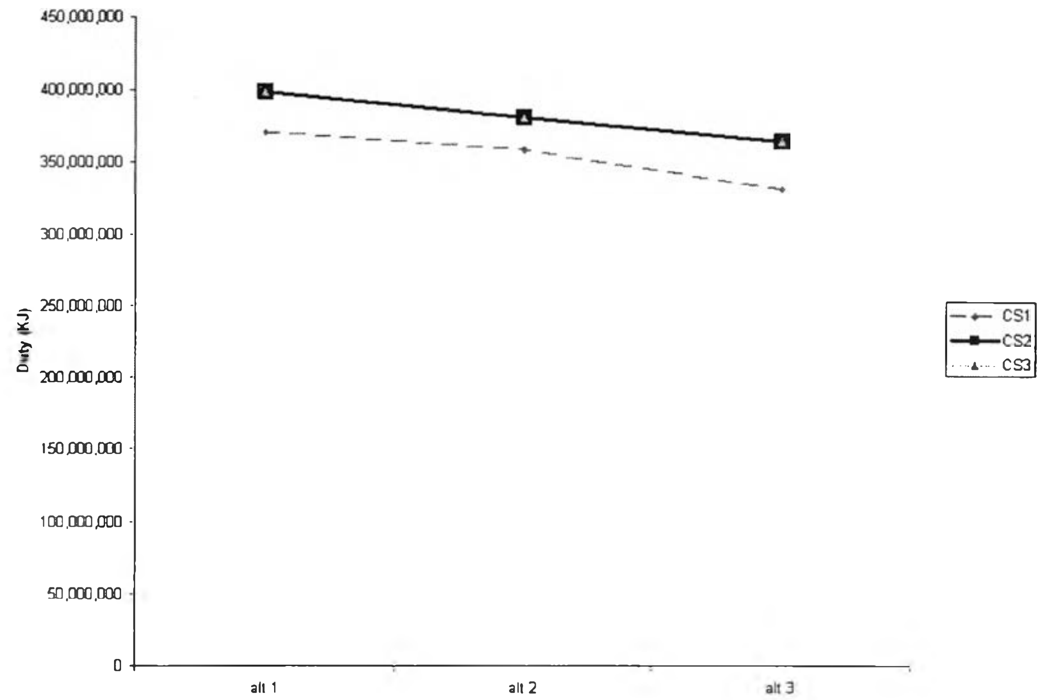


Figure 4.29: The utility consumptions (exclude cooler and quench duty) of HDA process when change the total toluene feed flowrates.

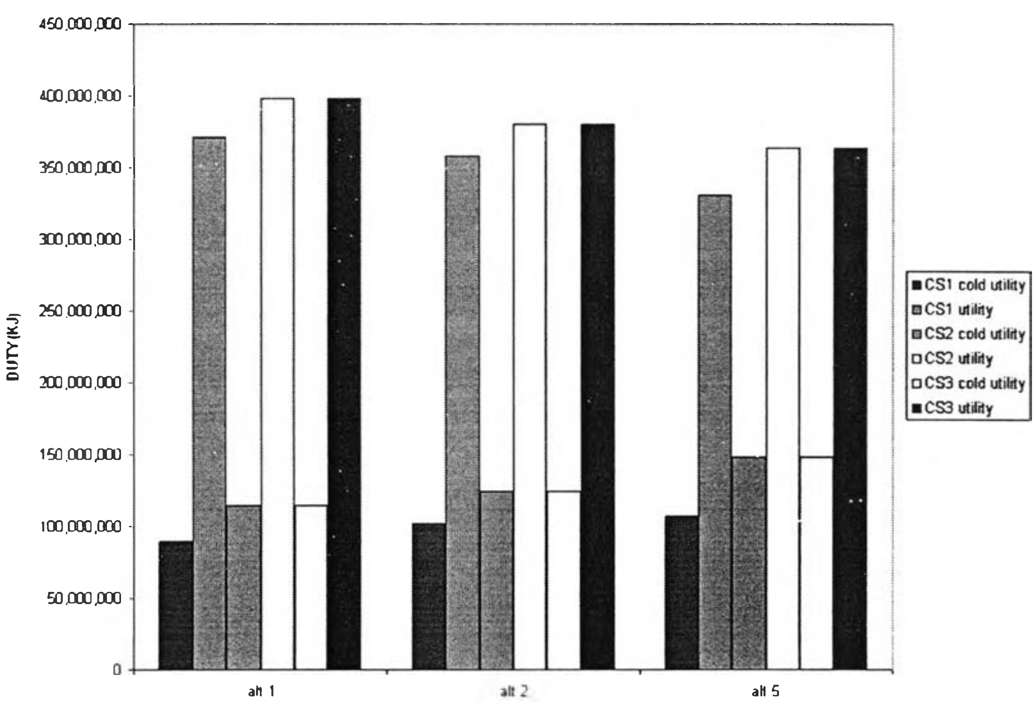


Figure 4.30: The utility consumptions of HDA process when change the total toluene feed flowrates.