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Heat Integrated Process Design, Simulation and Control of Polymerization and Drying Sections for HDPE Production

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Abstract
This paper describes a production model for high density polyethylene. Aspen Polymer Plus was used to model and simulate the polymerization drying and separation sections. Heat operations were improved using Aspen Energy Analyzer; The pinch analysis method generated, the heat exchanger network. The dynamic model via Aspen Plus Dynamics evaluated the stability and controllability of the models. The simulation results such as production rate, polydispersity index, weight-average molecular weight, and number-average molecular weight correponded with Hakim and Moballegh [1]. Five heat exchanger networks were proposed: the one with the lowest total index cost was selected to perform the dynamics analysis.

Keywords: High Density Polyethylene (HDPE) production process, Aspen Polymer Plus, Aspen Energy Analyzer, Aspen Plus Dynamics

1 Introduction

Currently, high density polyethylene (HDPE) is the world's third-largest commodity plastic resin with 35 million tons produced in 2010 [2]. Widespread uses of HDPE include in liquid food packaging, sterile packaging, automobiles, recycled HDPE, and innovative packaging designs. Unique features of HDPE include being environmentally friendly and cost-effective as well as having excellent flexibility and a wide range of designs.

The slurry polymerization of HDPE is widely used in HDPE plants. The major advantages of this slurry process include mild operating conditions, high monomer conversion, ease of heat removal, and relative ease of processing. Its disadvantages include long residence times (1–2.5 h per reactor), and limited production rates of polymers that have relatively low densities (lower than 0.940 g/cm^3) due to resin swelling.

In this work, the Aspen Polymer Plus was used to develop a HDPE slurry production process. The developed model consisted of polymerization, drying and separation sections. The Aspen Energy Analyzer was used for the heat exchanger network to provide recommend various economical designs.

Finally the Aspen Plus Dynamics was used to analyze the stability and controllability of each heat exchanger network design. The economics and control system were considered to pin-point the most suitable heat exchanger network design.

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1.1 *Process description*

A catalyst mixture was prepared in the catalyst feeding section by mixing catalyst powders with hexane at a specific ratio for each product grade. The catalyst slurry was fed to the polymerization section in conjunction with ethylene, co-monomer (either propylene or butane-1 depending on the desired product grade) and hydrogen. The generated PE slurry was passed through the horizontal centrifuge and rotary dryer in the separation and drying sections to separate PE wet cake from hexane and low polymer (the mixture is called mother liquor). HDPE was dried to reach a specification of about 0.5%wt of hexane content. In the pelletizing section, the molten polymer was mixed with stabilizers and passed through the extruder to become pellets. The on-spec pellets were then delivered to the packing process. The mother liquor from was sent from the hexane recovery section to recover hexane. A low polymer was obtained at the bottom of the hexane stripper while the hexane discharged at the top still needed moisture to be removed to meet the specification in the hexane dehydrator. The recoveed hexane was finally recycled back into the process.

1.2 *Polymerization section*

In the polymerization section, the two reactors were operated in either parallel or series and were fed with ethylene monomer, hexane solvent, co-monomer and catalysts.

The co-monomer for the parallel process was propylene. The slurry streams leaving the two reactors were combined and entered a flash unit for removal of light hydrocarbons. The vapor streams leaving the reactors contained hexane, monomer, and light gases present in the system. These streams were cooled and flashed into vapor and liquid streams, which recycled the monomer and solvent feed streams respectively.

For the series reactor, raw materials were fed into the first CSTR and the slurry product was then pumped to the second CSTR, which also received fresh monomer, catalyst, and solvent. The co-monomer for the series process was 1-butene, and it entered only as a feed stream in the second reactor. The vapor outlet from each reactor underwent cooling and was recycled to the reactor inlet. The slurry stream leaving the second reactor entered a flash unit for removal of volatiles. The resulting stream entered a centrifugal separator, which removed and returned hexane to the reactor inlets.

1.3 *Drying and separation sections*

The slurry was continuously fed to a horizontal-type centrifuge which revolved at a high rotating speed. The product was separated into wet cake and hexane solvent. The hexane stream containing a low polymer called mother liquor flowed into the mother liquor drum. Then it was pressurized in a pump and fed to the hexane recovery section. Meanwhile, the wet cake containing 32–35% hexane was discharged to the dryer through the wet cake screw conveyor. Hot nitrogen heated from the dryer gas heater flowed through the steam tube rotary dryer counter-currently with the product. Low pressure steam was supplied to the dryer as the heat source. The product leaving the dryer contained less than 0.35% volatile matter (as hexane). Some of the powder that contained nitrogen gas was sent to the dryer gas scrubber to separate polyethylene powder from mixed gas. Hexane in mixed gas leaving the scrubber was condensed by the dryer gas condenser supplied by cooling water and sent back to the scrubber. The non-condensable gas was cooled and condensed again with the dryer gas cooler supplied with brine. After the dryer gas cooler, mixed gas containing mostly nitrogen was heated by the dryer gas condenser and sent back to the dryer [3].

2 Methodology

The methodology consisted of 4 steps. It started with developing and verifying the HDPE production model. After that the heat exchanger network and control system design methods was applied to the HDPE production model. Finally the suitable heat exchanger network was identified.

2.1 *Develop and verify the HDPE production model*

In this step contained 2 main parts. The first part was to develop the HDPE production process model. This part started with understanding the HDPE polymerization process. Then it was necessary to learn about the Aspen Polymer Plus simulation program and determine all input data [4].

In this work, the slurry polymerization technique was selected for model development, because it can be operated at moderate temperature and pressure. In addition, Ziegler-Natta catalyst was applied in this model [5]. Ziegler-Natta catalysts produce polymers with broad molecular weight distributions because it has several sites with a difference in chemical properties at each site type. Therefore the necessary input data are feed condition, operating condition and kinetic parameters that were provided by Hakim and Moballegh [1]. For thermodynamics, properties can be obtained from Khare.

In the second part, after the polymerization section of the HDPE production process was developed, we verified this model by comparing the results from a simulation with the results provided in a previous study [1]. In this comparison step, production rate, polydispersity index (PDI), the weight-average molecular weight (Mw), number-average molecular weight (Mn), and monomer conversion corresponded with Hakim and Moballegh's [1] work.

2.2 *Establishing the heat exchanger network design*

After the HDPE production process was developed, the Aspen Energy Analyzer was used to generate the recommended design of the heat exchanger network. The Aspen Energy Analyzer uses the pinch method to calculate the pinch point and show the recommended design of the heat exchanger network automatically. Therefore the heat source and heat sink must first be identified.

To use the Aspen Energy Analyzer, we imported the backup file (.bkp) of Aspen Plus directly to the Aspen Energy Analyzer or input only process stream to create the heat exchanger network. Therefore the heat source and heat sink must first be identified.

Moreover, the Aspen Energy Analyzer generates economic results which are the capital cost (cost of heat exchanger) and operating cost (cost of utility). The sum of the capital cost and operating cost is called "total cost index". The calculation of total cost index is presented in Eqs. (1)–(3).

$$\text{Total Cost Index} = \text{Annualized Capital Cost} + \text{Operating Cost} \qquad (1)$$

$$\text{Annualized Capital Cost} = \text{Capital Cost} * \text{Annualization Factor} \qquad (2)$$

$$\text{Annualization Factor} = ((1+\text{Rate of Return}/100)^{\text{Plant Life}})/\text{Plant Life} \qquad (3)$$

In this study, the rate of return was 10%, and plant life was 5 years.

2.3 *Design the control system of each heat exchanger network design*

After obtaining the recommended designs of heat the exchanger network, Aspen Plus Dynamics was used to check the feasibility or controllability of each design [6]. In this work, inlet temperature of CSTR-1, CSTR-2, and drying and separation units should be controlled in order to maintain the production rate and to evaluate the performance of each heat exchanger network design.

Next, the controlled variables, manipulated variable, and disturbance were then identified. Controlled and manipulated variables of each design were different in the heat exchanger network configuration. The guidelines by Hougen, Newell and Lee were very acceptable for the preliminary process of control variables selection.

To specify disturbances, this should occur during the actual operation. Therefore, it could be the temperature deviation of input process streams.

2.4 *Identify the suitable heat exchanger network design*

The final step was to analyze results from the previous step to select the best heat exchanger network to apply to the HDPE production process model. We considered the economics and control system simultaneously. The selected design should have the lowest cost, as present in the Aspen Energy Analyzer, and be stable in the dynamics mode as shown in Aspen Plus Dynamics.

3 Results and Discussion

3.1 *HDPE production process developed model*

The model of the HDPE production process was developed in the Aspen Polymer Plus simulation program. This developed model was divided into 3 sections including polymerization, separation and drying sections, and hexane recovery sections.

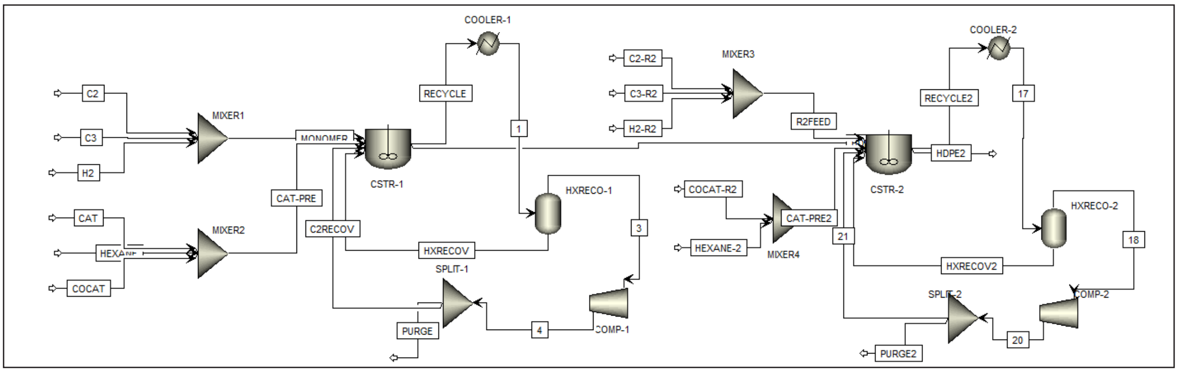

Figure 1: Developed model of polymerization section in Aspen Polymer Plus.

3.1.1 Polymerization section

In the polymerization section, ethylene, propylene, hydrogen, catalyst slurry and hexane were mixed before polymerization: CSTR-1 and CSTR-2. The model was set to simulate a series of operations as shown in Figure 1.

The process conditions which consisted of feed condition, operating condition, reaction mechanism and kinetic parameter refer to Hakim and Moballegh's [1] work. The first reactor operated at 85°C and 8.5 bar, while the second reactor operated at 80°C and 2.5 bar. The feed condition is presented in Table 1.

Table 1: Feed condition

Component	Description	CSTR-1 (kg/hr)	CSTR-2 (kg/hr)
Ethylene	Monomer	8,120	8,650
Propylene	Co-monomer	650	610
Hydrogen	Chain transfer agent	185	30
Hexane	Solvent	12,845	13,086
Titanium-tetrachloride	Catalyst	152.98	183.12
Triethyl-aluminum	Co-catalyst	44.59	47.56

In this work, we selected the reaction subset in Ziegler-Natta kinetic [7]. The mechanism started with catalyst site activation by co-catalyst in which the transition-metal catalyst (CAT) reacted with the co-catalyst to form a vacant site. Next a monomer molecule reacted with a vacant site to initiate chain growth or propagation site in the chain initiation step. The polymer chain grew rapidly with the successive addition of monomer molecules at the propagation site called the chain propagation step. Chain transfer occurs when a monomer or chain-transfer agent releases a polymer chain from the catalyst which is inactive, and initiates the growth of a new chain. Finally, the active sites on the catalyst can undergo spontaneous deactivation to form dead sites that are not active.

The kinetic parameters used in this work were provided by Hakim and Moballegh [1]. This catalyst consisted of 6 sites. Kf was the rate constant for catalyst activation, Ki was the rate constant for chain initiation, Kp was the rate constant for chain propagation, Ktm was the rate constant for chain transfer by monomer, Kth was the rate constant for chain transfer by hydrogen, and Kd was the rate constant for catalyst deactivation.

Choosing the appropriate property model for thermodynamic calculations was provided by Khare [8]. Because polymer equations of state do not normally perform as well as simple cubic equations of state for small components, we used different property methods for the units and streams that contained a polymer and those that did not. In the HDPE process, the polymer was present in the reactors and the subsequent separation units. The vapor recycle contained only the monomer, solvent, and other small molecule components, because the polymer was nonvolatile. We used the Sanchez-Lacombe equation of state for the polymer-containing sections of the plant and the Chao-Seader method for the non-polymer areas.

After the polymerization section of the HDPE production process was developed, we compared the

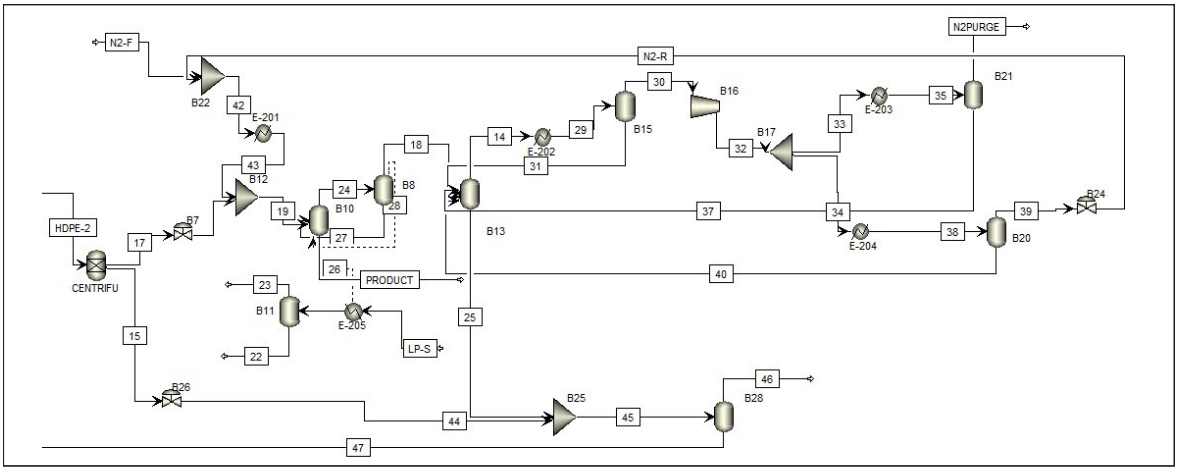

Figure 2: Developed model of dryer and separation section in Aspen Polymer Plus.

Table 2: Verification

Data	Unit	CSTR-1			CSTR-2		
		Actual	***Model***	***% deviation***	***Actual***	***Model***	***% deviation***
Production Rate	kg/hr	8,610	8,652.4	0.49	18,216	18,210	–0.03
Monomer Conversion	%	94.4	94.9	0.53	95.6	95.17	–0.45
Mw	g/mol	48,885	47,985	–1.84	238,182	235,861	–0.97
PDI	-	9	9.13	1.44	20.6	22.2	7.76

results from the simulation with the results provided by Hakim and Moballegh [1] to verify the developed model. Table 2, shows an acceptable deviation for further study using this model.

3.1.2 Dryer and separation section

From Figure 2, the slurry HDPE from CSTR-2 was continuously fed to the horizontal-type centrifuge (CENTRIFU) which was used as a separation-block. The product was pressed to the inside wall of the rotating bowl under centrifugal force and separated into wet cake and hexane solvent. The hexane stream (S-15) containing low polymer as called mother liquor, was sent to the hexane recovery section. Meanwhile, the wet cake (S-17) containing 32–35% hexane was discharged to the dryer and separation section. Hot nitrogen heated from the dryer gas heater (E-201) was mixed with the product. Low pressure steam was supplied to the dryer (B10 and B8) as the heat source.

The product steam (S-PRODUCT) leaving the dryer (B10 and B8) contained less than 0.35% volatile matter (as hexane) and then flowed to the pelletizing and storage sections. The B13 was used to condense some hexane from nitrogen. The non-condensable gas was cooled and condensed again with E-204 supplied with brine. After E-204, mixed gas containing mostly nitrogen was sent back to the dryer. Some of the mixed gas was sent to the purge gas condenser (E-203) to remove more hexane before venting to the flare system.

As seen in Table 3, the centrifuge can reduce the hexane to 32–35% in the wet cake stream product (S-17). After the wet cake was passed through the dryer and separation sections, the product contained less than 0.35% hexane.

Table 3: Mole fraction of S-17 and product stream

Component	S-17	Product
C2	0.0017	0.0000
C3	0.0005	0.0000
H2	0.0000	0.0000
C6	0.3482	0.0053
CAT	0.0000	0.0000
COCAT	0.0000	0.0000
HDPE	0.6496	0.9945
N2	0.0000	0.0002

3.2 *Heat exchanger network design*

After the HDPE production process was developed, we found that the heat in the process could be improved in the heat exchanger network design. In this work, the Aspen Energy Analyzer is used as a tool to create the heat exchanger network.

In this case, the Aspen Energy Analyzer generated five recommended designs for the heat exchanger network. Each design had a different number of heat exchangers, area and amount of utility, because of each configuration. Therefore, each design needed individual operating costs and capital costs.

In an economic analysis in Table 4, we only considered the total cost index. The total cost index was the sum of the operating cost and annualized capital cost with a five year plant life as described in the methodology. The total cost index indicated that design 2 should be selected because it had the lowest investment cost.

Table 4: The cost of each design of heat exchanger network

Design	Total cost index (cost/year)
1	9.400E5
2	9.383E5
3	9.390E5
4	9.427E5
5	9.388E5

3.3 *Control system design*

In the previous part, we already developed the model for the HDPE production process and designed the heat exchanger network. From the heat exchanger network design using Aspen Energy Analyzer, we obtained 5 recommended designs and design 2 had the lowest total index cost. However, we conducted a control system design to check the feasibility or controllability of each design.

In this work, the Aspen Plus Dynamics was applied to design the control system [9]. The objective was to control the temperature of CSTR-1, CSTR-2, and drying and separation units. Therefore the process streams that input to CSTR-1, CSTR-2 and drying and separation units were controlled variables. The process streams which were controlled are shown in Table 5.

Table 5: Controlled variables and set pionts

CVs	Process stream	Set-point (°C)
CV1	Monomer entering CSTR-1	85
CV2	Catalyst entering CSTR-1	85
CV3	C2recycle entering CSTR-1	85
CV4	HXrecycle entering CSTR-1	85
CV5	Monomer entering CSTR-2	80
CV6	Catalyst entering CSTR-2	80
CV7	C2recycle entering CSTR -2	80
CV8	HXrecycle entering CSTR -2	80
CV9	N2 HXrecycle entering the drying section	200

In this work, the manipulated variables were the duty of utility such as heater and cooler located before or after the heat exchanger. Utility was selected as a manipulated variable because it has large and rapid effects on the controlled variable and avoids recycling the disturbances.

The disturbances that could occur in the actual operation should be specified. For this case, it was assumed that there were 4 disturbances: +/–10°C of temperature of input process streams. It may be caused by the previous section such as the monomer or catalyst preparation.

The disturbances were;

D-1: Temp. of monomer stream entering polymerization section (CSTR-1) +10°C

D-2: Temp. of monomer stream entering polymerization section (CSTR-1) –10°C

D-3: Temp. of catalyst stream entering polymerization section (CSTR-1) +10°C

D-4: Temp. of catalyst stream entering polymerization section (CSTR-1) –10°C

D-5: Temp. of monomer stream entering polymerization section (CSTR-2) +10°C

D-6: Temp. of monomer stream entering polymerization section (CSTR-2) –10°C

D-7: Temp. of catalyst stream entering polymerization section (CSTR-2) +10°C

D-8: Temp. of catalyst stream entering polymerization section (CSTR-2) –10°C

3.3.1 Control system of heat exchanger network design 1

For the control system of design 1, Figure 3 depicts 7 controllers in which 2 points could not be controlled because if they were controlled, the manipulated variables that comprise the recycling disturbance must

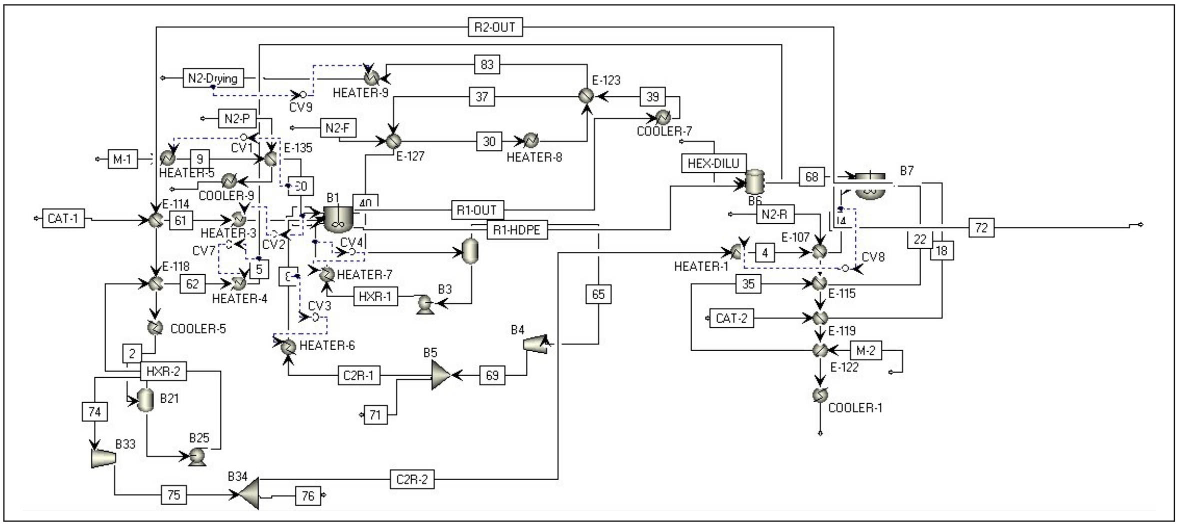

Figure 3: The control system of heat exchanger network design 1.

be selected. The controlled and manipulated variables are presented in Table 6.

Table 6: Controlled and manipulated variables of heat exchanger network design 1

CVs	Controlled variable	Manipulated variable
CV1	Temp. of S80	Duty of HEATER-5
CV2	Temp. of S3	Duty of HEATER-3
CV3	Temp. of S8	Duty of HEATER-6
CV4	Temp. of S7	Duty of HEATER-7
CV5	-	-
CV6	-	-
CV7	Temp. of S5	Duty of HEATER-4
CV8	Temp. of S14	Duty of HEATER-1
CV9	Temp. of N2Drying	Duty of HEATER-9

From Figure 4, the results showed that all of the controlled variables could return to their set-point. The controlled variables CV3, CV4, CV8 and CV9 did not receive any effect from the disturbances. The controlled variable CV1 received an effect from the temperature deviation of the monomer entering the section (D-1 and D-2). CV2, had a large effect from the temperature change of the catalyst entering the section (D-3 and D-4). In addition, CV2 received small effects from D-5, D-6, D-7 and D-8 which were disturbances to the monomer and catalyst entering CSTR-2, because CV2 was selected to control the temperature of the catalyst entering CSTR-1 after exchanging heat with CSTR-2 outlet stream. The controlled variable CV7 responded to disturbances D3 and D4, because it used the same hot stream as the catalyst feed stream.

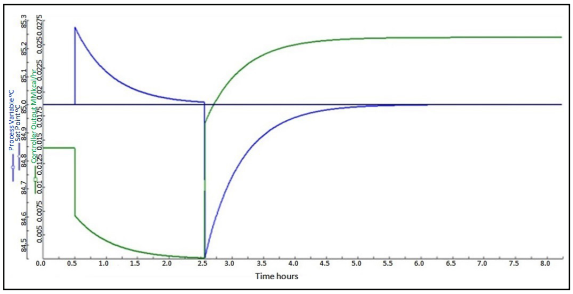

Figure 4a: The test of CV1 of heat exchanger network design 1.

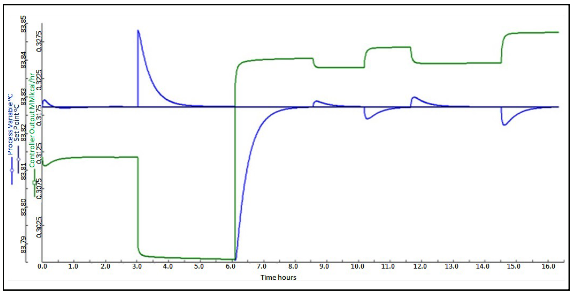

Figure 4b: The test of CV2 of heat exchanger network design 1.

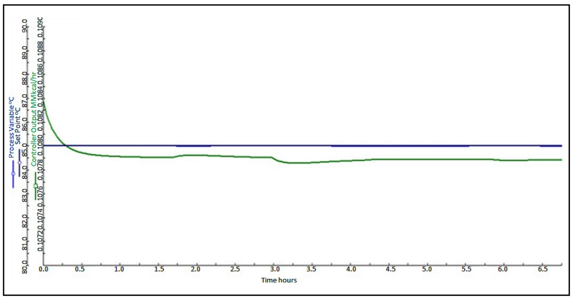

Figure 4c: The test of CV3 of heat exchanger network design 1.

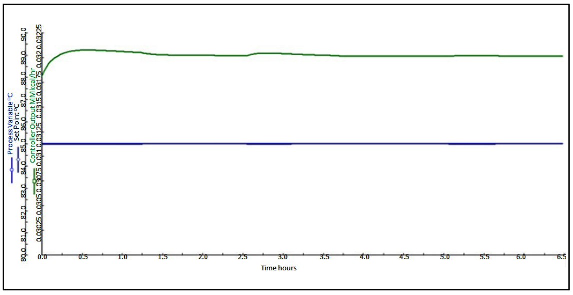

Figure 4d: The test of CV4 of heat exchanger network design 1.

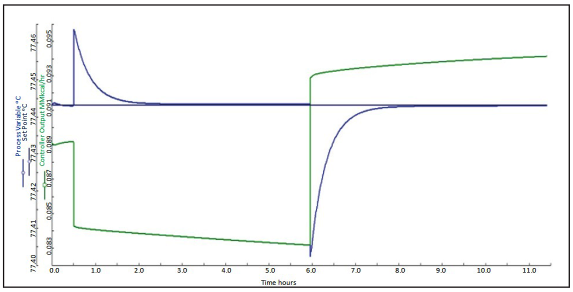

Figure 4e: The test of CV7 of heat exchanger network design 1.

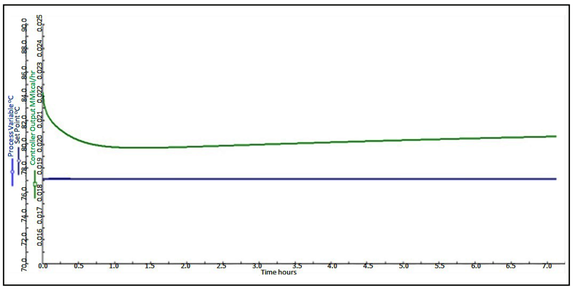

Figure 4f: The test of CV8 of heat exchanger network design 1.

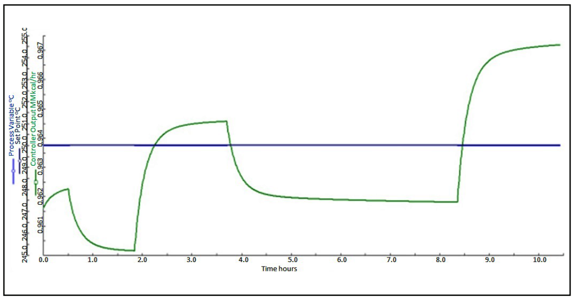

Figure 4g: The test of CV9 of heat exchanger network design 1.

3.3.2 Control system of heat exchanger network design 2

For the control system in design 2, Figure 5 depicts 8 controllers according to which only one point could not be controlled because if it were controlled, the manipulated variables in the recycling disturbance must be selected. The controlled and manipulated variables are presented in Table 7.

Table 7: Controlled and manipulated variables of heat exchanger network design 2

CVs	Controlled variable	Manipulated variable
CV1	Temp. of S80	Duty of HEATER-5
CV2	Temp. of S3	Duty of HEATER-3
CV3	Temp. of S8	Duty of HEATER-6
CV4	Temp. of S7	Duty of HEATER-7
CV5	Temp. of S22	Duty of HEATER-10
CV6	-	-
CV7	Temp. of S14	Duty of HEATER-1
CV8	Temp. of S5	Duty of HEATER-4
CV9	Temp. of N2-Drying	Duty of HEATER-9

For the heat exchanger network design 2, according to Figure 6, the results showed that all the controlled variables could return to their set-point. The controlled variable CV2 and CV8 received an effect from the disturbances. The major effect to CV2 was caused by disturbance from the temperature deviation of catalyst feed to CSTR-1 (D-3 and D-4). CV-8, also responded to D-3 and D-4. This was becauses CV2 and CV8 used the same heat source.

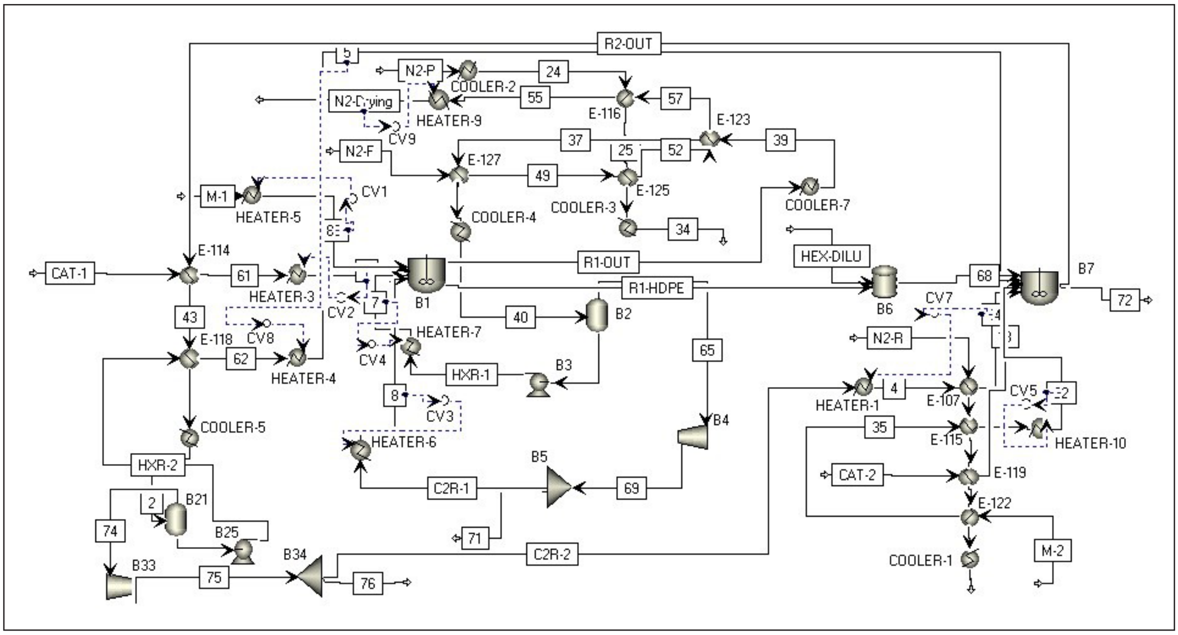

Figure 5: The control system of heat exchanger network design 2.

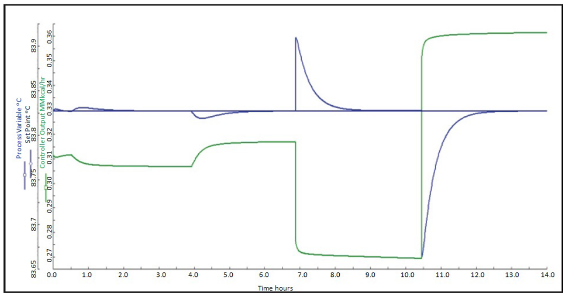

Figure 6a: The test of CV1 of heat exchanger network design 2.

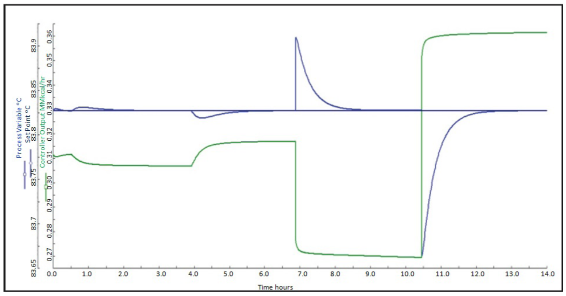

Figure 6b: The test of CV2 of heat exchanger network design 2.

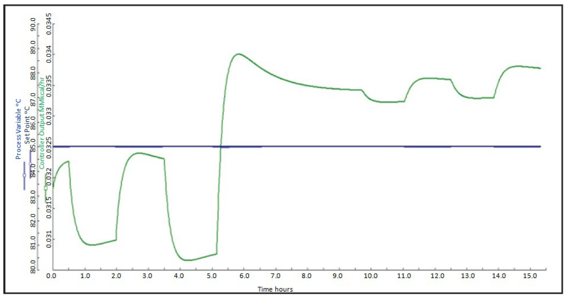

Figure 6c: The test of CV3 of heat exchanger network design 2.

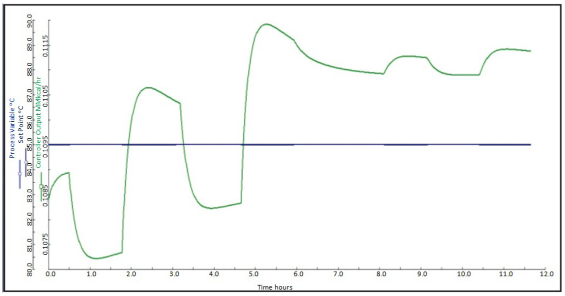

Figure 6d: The test of CV4 of heat exchanger network design 2.

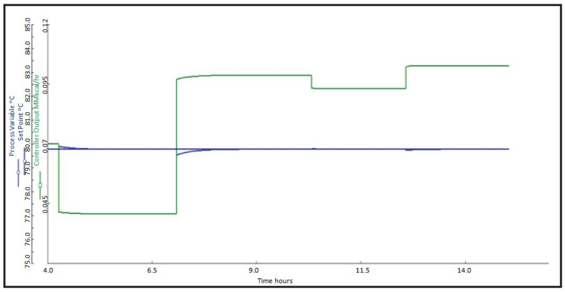

Figure 6e: The test of CV5 of heat exchanger network design 2.

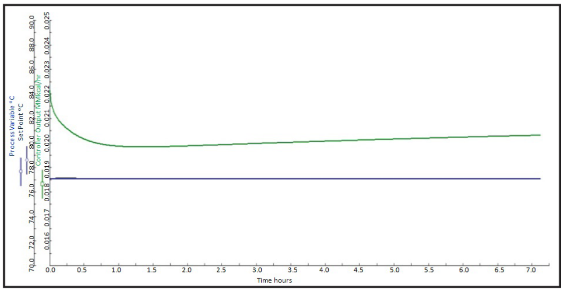

Figure 6f: The test of CV7 of heat exchanger network design 2.

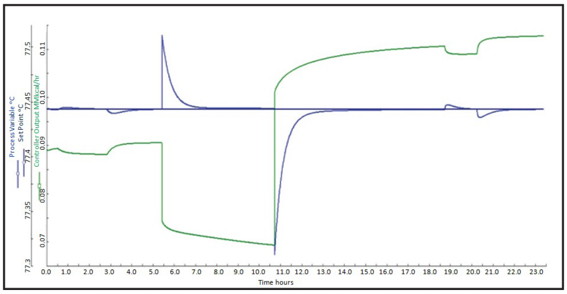

Figure 6g: The test of CV8 of heat exchanger network design 2.

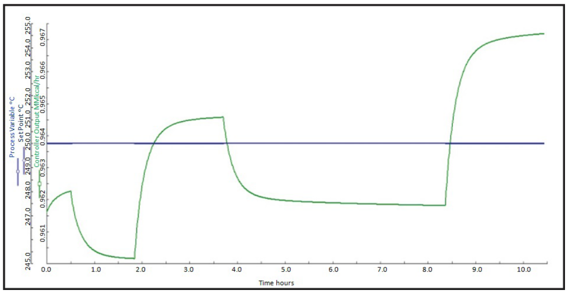

Figure 6h: The test of CV9 of heat exchanger network design 2.

3.3.3 Control system of heat exchanger network design 3

For the control system of design 3, Figure 7 depicts 7 controllers. The controlled and manipulated variables are presented in Table 8.

Table 8: Controlled and manipulated variables of heat exchanger network design 3

CVs	Controlled variable	Manipulated variable
CV1	Temp. of S80	Duty of HEATER-5
CV2	Temp. of S3	Duty of HEATER-3
CV3	Temp. of S8	Duty of HEATER-6
CV4	Temp. of S7	Duty of HEATER-7
CV5	-	-
CV6	-	-
CV7	Temp. of S14	Duty of HEATER-1
CV8	Temp. of S5	Duty of HEATER-4
CV9	Temp. of N2-Drying	Duty of HEATER-9

For the heat exchanger network design 3, in Figure 8, the results showed that all controlled variables tend to returned to their set-point. The trend of responding to the disturbance was similar to the result of design 2. Comparing the models in designs 2 and 3, they only differed in the heat exchanger configuration on monomer feed to CSTR-2 line. Therefore, in this design, CV2 and CV8 were 2 controlled variables that responded to the same disturbances as design 2. However design 3 had uncontrollable points than design 2.

3.3.4 Control system of heat exchanger network design 4

For the control system of design 4, Figure 9 depicts 8 controllers and 1 point was not controlled to avoid the cycling disturbance The controlled and manipulated variables are presented in Table 9.

Table 9: Controlled and manipulated variables of heat exchanger network design 4

CVs	Controlled variable	Manipulated variable
CV1	Temp. of S9	Duty of HEATER-5
CV2	Temp. of S3	Duty of HEATER-3
CV3	Temp. of S7	Duty of HEATER-7
CV4	Temp. of S8	Duty of HEATER-6
CV5	Temp. of S14	Duty of HEATER-1
CV6	-	-
CV7	Temp. of S22	Duty of HEATER-9
CV8	Temp. of S5	Duty of HEATER-4
CV9	Temp. of N2-Drying	Duty of HEATER-10

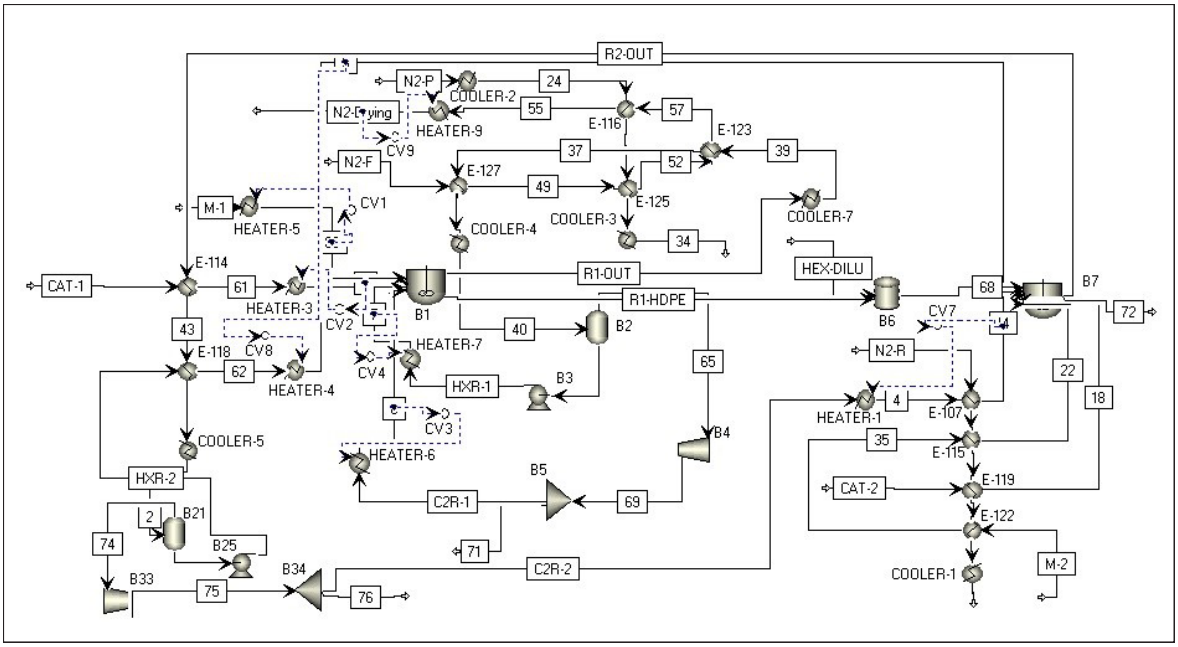

Figure 7: The control system of heat exchanger network design 3.

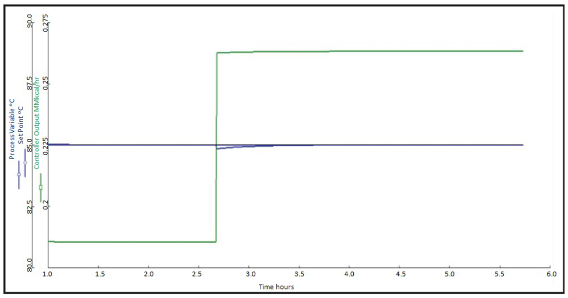

Figure 8a: The test of CV1 of heat exchanger network design 3.

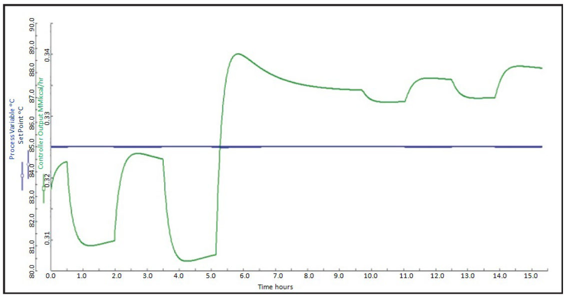

Figure 8c: The test of CV3 of heat exchanger network design 3.

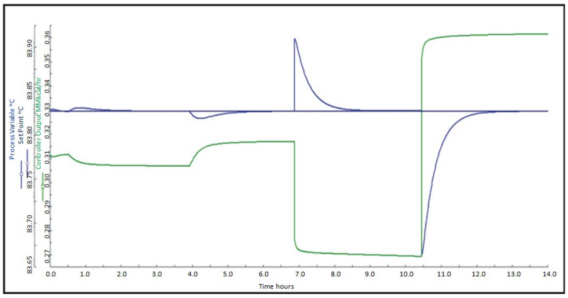

Figure 8b: The test of CV2 of heat exchanger network design 3.

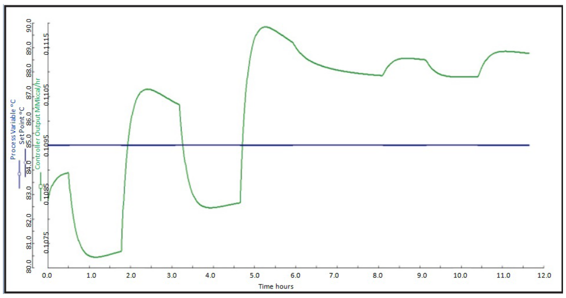

Figure 8d: The test of CV4 of heat exchanger network design 3.

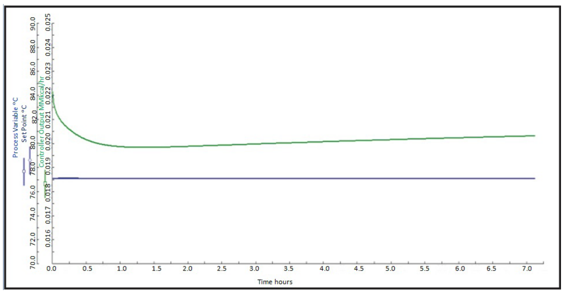

Figure 8e: The test of CV7 of heat exchanger network design 3.

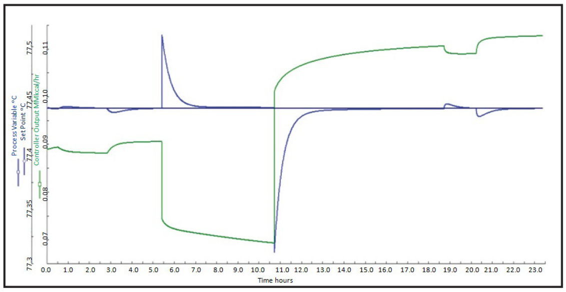

Figure 8f: The test of CV8 of heat exchanger network design 3.

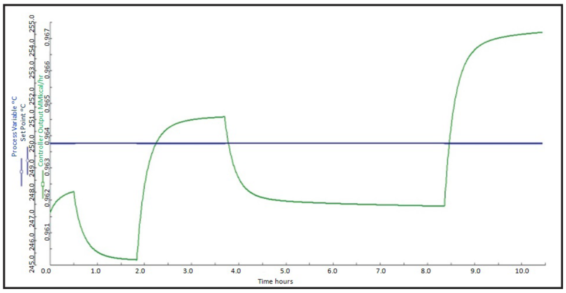

Figure 8g: The test of CV9 of heat exchanger network design 3.

For the heat exchanger network design 4, from Figure 10, the results showed that all controlled variables tended to return to their set-point. Disturbances D1 and D2 affected V1, D3 and D4 affected CV2, and D1, D2, D3 and D4 responded to CV8. CV8 received the effect from CV1 and CV2 because the three controlled variables were located on same heat exchanger train and used the same heat source.

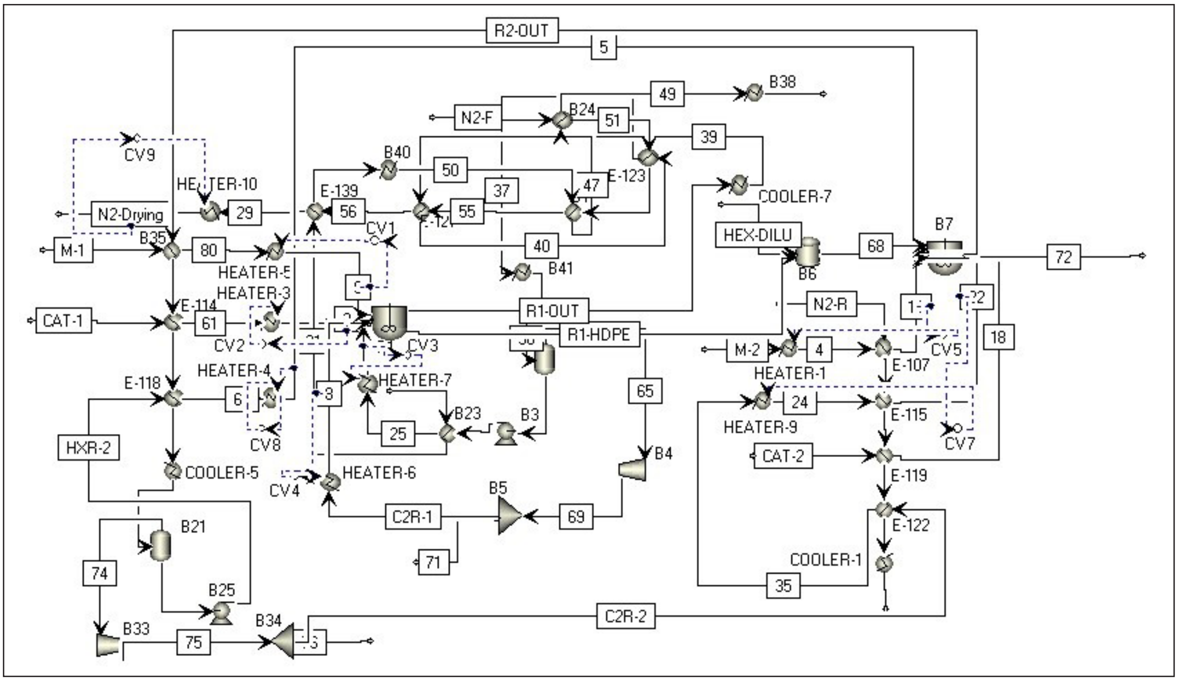

Figure 9: The control system of heat exchanger network design 4.

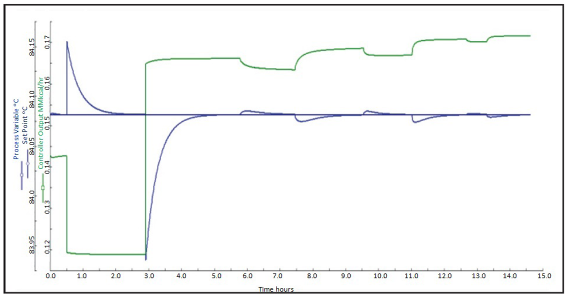

Figure 10a: The test of CV1 of heat exchanger network design 4.

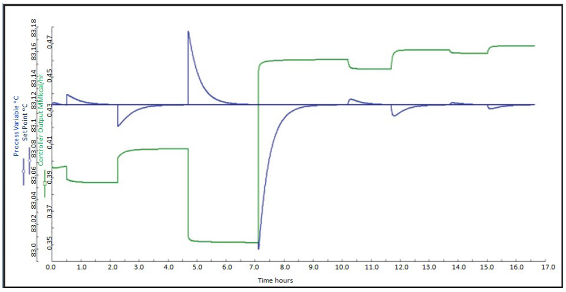

Figure 10b: The test of CV2 of heat exchanger network design 4.

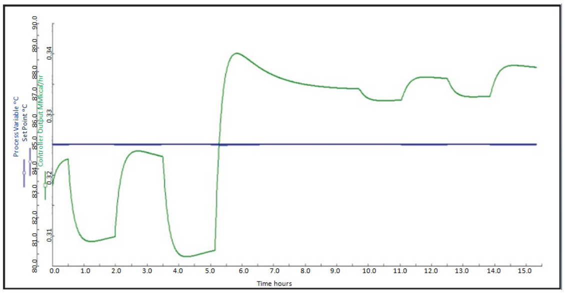

Figure 10c: The test of CV3 of heat exchanger network design 4.

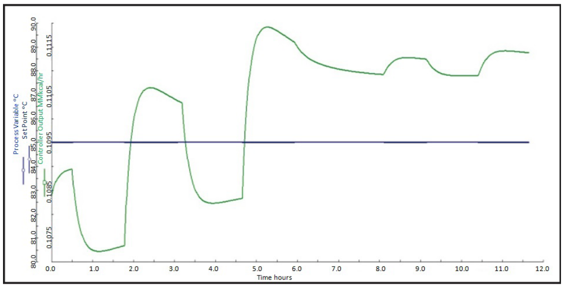

Figure 10d: The test of CV4 of heat exchanger network design 4.

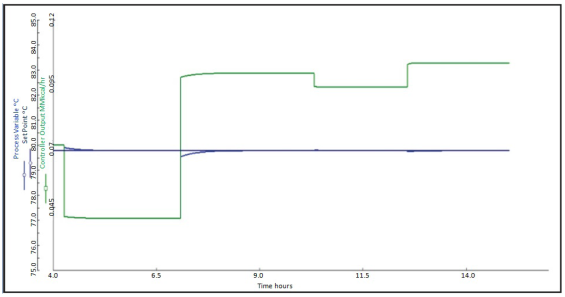

Figure 10e: The test of CV5 of heat exchanger network design 4.

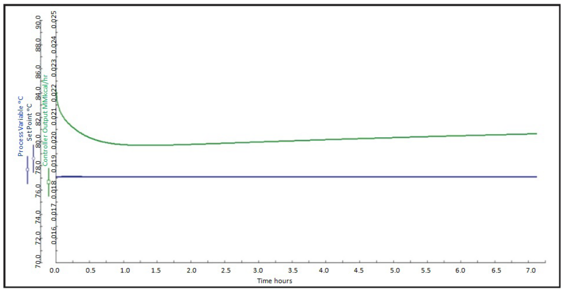

Figure 10f: The test of CV7 of heat exchanger network design 4.

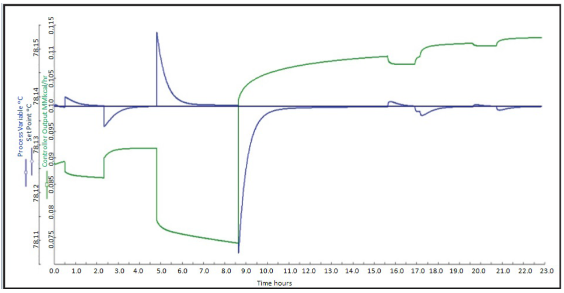

Figure 10g: The test of CV8 of heat exchanger network design 4.

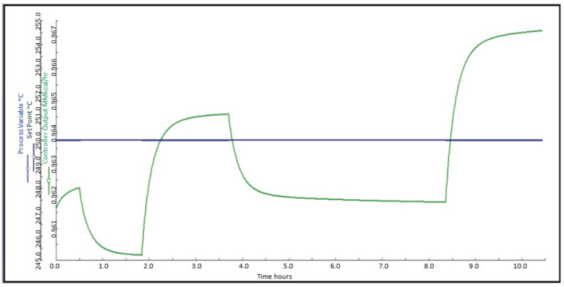

Figure 10h: The test of CV9 of heat exchanger network design 4.

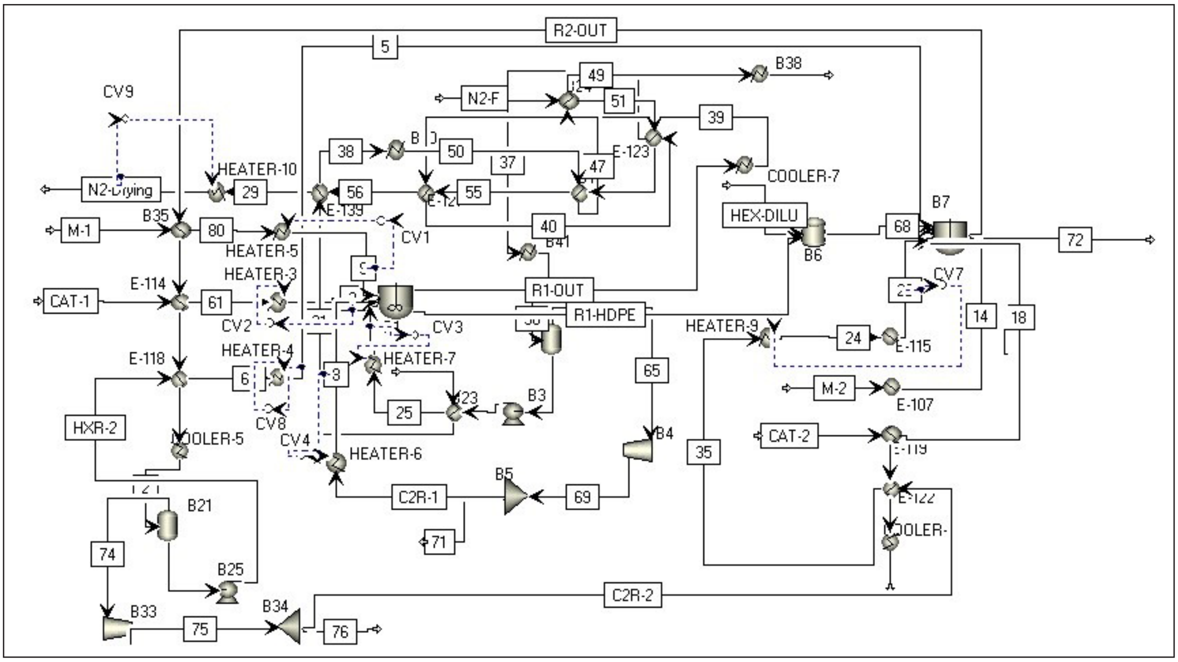

Figure 11: The control system of heat exchanger network design 5.

3.3.5 Control system of heat exchanger network design 5

For the control system of design 5, Figure 11 depicts 7 controllers and 2 points were not controlled to avoid the cycling disturbance. The controlled and manipulated variables are presented in Table 10.

Table 10: Controlled and manipulated variables of heat exchanger network design 5

CVs	Controlled variable	Manipulated variable
CV1	Temp. of S9	Duty of HEATER-5
CV2	Temp. of S3	Duty of HEATER-3
CV3	Temp. of S7	Duty of HEATER-7
CV4	Temp. of S8	Duty of HEATER-6
CV5	-	-
CV6	-	-
CV7	Temp. of S22	Duty of HEATER-9
CV8	Temp. of S5	Duty of HEATER-4
CV9	Temp. of N2-Drying	Duty of HEATER-10

For the heat exchanger network design 5, from Figure 12, the results showed that all controlled variables tended to return to their set-point. The trend of responding to the disturbance was similar to the result of design 4. Comparing the models of designs 4 and 5, the only difference was the heat exchanger configuration on monomer feed to CSTR-2 and recycle stream to CSTR-1. Therefore, in this design, CV1, CV2 and CV8 were 3 controlled variables that responded to the same disturbance as design 4. However design 5 had more uncontrollable points than design 4.

In conclusion for the dynamics part, all heat exchanger designs could reach the set-point with disturbances. When considering the uncontrollable heat exchanger in each design, designs 2 and 4 had only 1 point. There were 2 controlled variables in design 2 that responded to disturbances; on the other hand, disturbances affected 3 controlled variables in design 4 as shown in Table 11. Therefore, heat exchanger network design 2 had the highest performance in dynamics mode.

Table 11: Number of uncontrollable heat exchanger of each design

Design	Number of uncontrollable HX
1	2
2	1
3	2
4	1
5	2

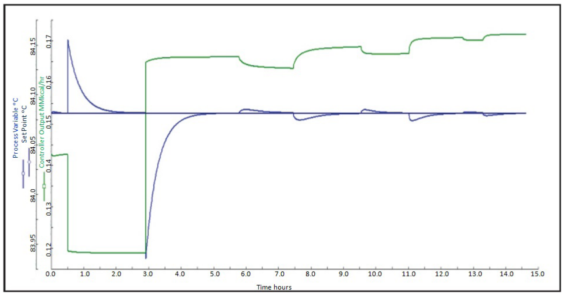

Figure 12a: The test of CV1 of heat exchanger network design 5.

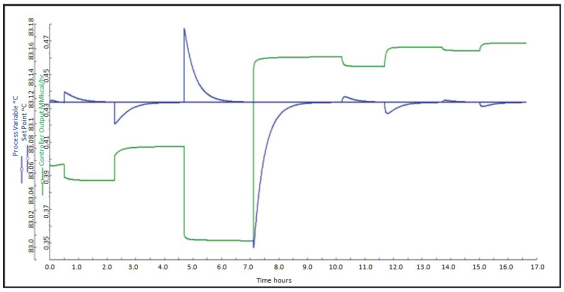

Figure 12b: The test of CV2 of heat exchanger network design 5.

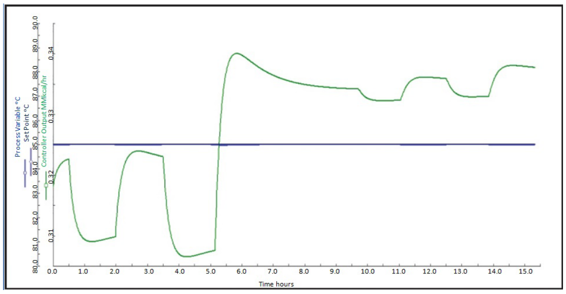

Figure 12c: The test of CV3 of heat exchanger network design 5.

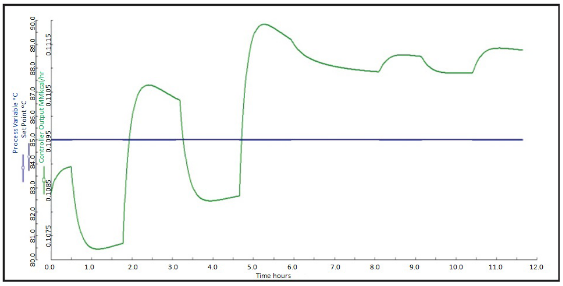

Figure 12d: The test of CV4 of heat exchanger network design 5.

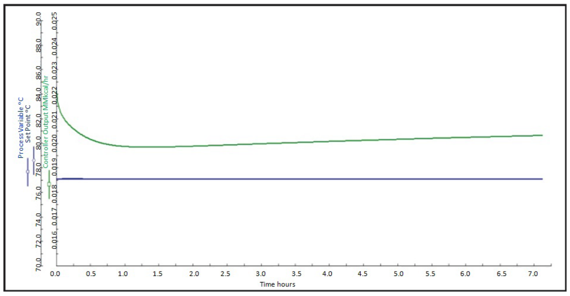

Figure 12e: The test of CV7 of heat exchanger network design 5.

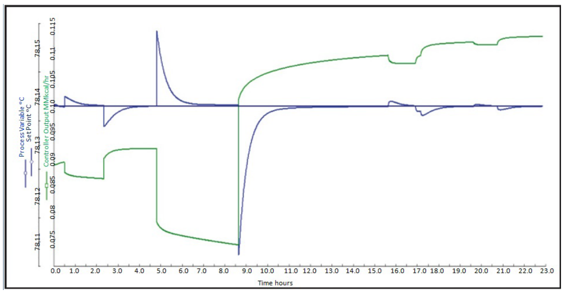

Figure 12f: The test of CV8 of heat exchanger network design 5.

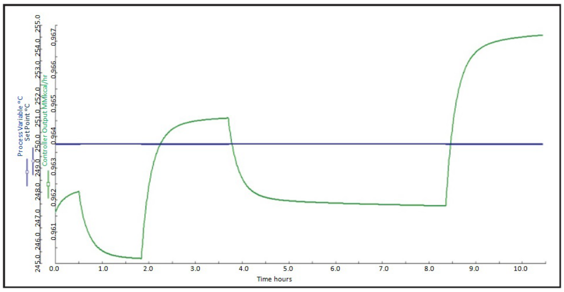

Figure 12g: The test of CV9 of heat exchanger network design 5.

4 Conclusions

This work used the Aspen Polymer Plus to develop the HDPE production model consisting of the polymerization section, drying and separation sections, and hexane recovery section. The process conditions consisted of feed condition, operating condition, reaction mechanism and kinetic parameter refering to Hakim and Moballegh's [1] work. After the polymerization section of the HDPE production process was developed, the results from the Aspen Polymer Plus and results provided in a past study [1] had little deviation. Therefore this developed model was verified. For drying and separation sections, the centrifuge could reduce the hexane to 32–35% in a wetcake stream product. After the wetcake was passed through the dryer and separation sections, the product contained less than 0.35%hexane.

After the HDPE production process was developed, the Aspen Energy Analyzer was used as a tool to create the heat exchanger network. It generated 5 recommended designs for the heat exchanger network and showed the total cost index which was the summation of capital cost and operating cost. From the results, the heat exchanger network design 2 had the lowest total cost index. The Aspen Plus Dynamics was used to create the control system. The results revealed the points that were difficult to control. From the results of the control system, the heat exchanger network design 2 was also the most stable design.

After considering the economics and control system, the heat exchanger network design 2 was the most suitable for this model because of the lowest total cost index and stability in control system.

Acknowledgements

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