CHAPTER IV
RESULTS AND DISCUSSION

4.1 Review Historical Operation of Refinery Plant

The process flow scheme of the distillation of ARC is shown in Figure 4.1, consisting of 22 heat exchangers, 1 air fan cooler and 1 furnace. The overall energy usage that consume fuel oil as heat source in a furnace is 170 tsrf/day (1 tsrf is 40.5 GT) or 6885 GJ/day. Crude oil will be preheated about 5,508 GJ/day or 63,750 kW at furnace which is 80% of overall energy usage.

The cold crude from the tank farm is boosted up in pressure by P-1001 and preheated in E-1001 to 60°C. Upstream of E-1002, 200 t/sd of wash water is injected. The crude is further heated up in E-1002 and E-1003. Then the crude is split into two streams and heated up in E-1005 and E-1006. The recombined crude stream is heated up in E-1007 and in E-1008 and then routed to Crude Desalter V-1001.

Preheated crude is mixed with wash water and passes through a mixing valve and Static Mixer Wash Water/Crude M-1002 for homogenizing before entering the Crude Desalter V-1001. Facilities are present to recycle the wash water in Crude Desalter V-1001 to improve the desalting and washing efficiency. The washed and desaltered crude leaves Crude Desalter V-1001 via the top and is further preheated in crude preheat train. The desalted crude is heated up in E-1009 and E-1010 to 176°C, which is the preflash vessel inlet temperature upstream the level control valve.

In Preflash Vessel V-1002, design operating conditions 167°C and 3 barg, light hydrocarbons and the majority of the water are flash off. The preheated crude is split into three streams and heated up in E-1011, E-1015, E-1012, E-1013 and E-1014. The recombined preflashed crude picks up heat in E-1016. Finally, the crude is split into two streams in one heat exchanger take place in E-1017 and E-1019, and in the other in E-1018 and E-1020. The crude is recombined and then split over the furnace coils. The pressure level upstream the furnace flow control valves, 18 barg, is selected such that under all expected operating conditions, no vaporization occurs.

The outlet temperature of CDU Furnace F-1001 is 360˚C. The liquid and vapour coming from the furnace and vapours from Preflash Vessel V-1002 enter CDU Column C-1001. The feed stream enters CDU Column between trays 6 and 7. The column has 38 trays in total and 3 circuration sections: LCR, MCR and TCR. The amount of stripping used is 1.6 wt% on long residue product. There are two temperature controllers on C-1001. The temperature controller on tray 12 sets the HGO/LR cutpoint via the HGO-under reflux; the one on tray 31 sets the naphtha/kero cutpoint via the TCR duty.

The long residue section consists of 6 trays located below the feed inlet. The long residue is normally routed to unit HVU at 348˚C.

The middle distillate product, routed to HDS, normally consists of 4 streams, 3 originate from CDU Column C-1001 whilst the fourth stream is VGO, a liquid product from HVU unit. The lower circulating reflux section comprises trays 14 to18, and serves to condense HGO and provides under reflux for separation between HGO and LR. From total draw-off tray 14, a HGO stream is routed to Middle Distillate Vessel V-1003. It is thereafter spilt into an under reflux, which is pumped back to tray 13 while part of the HGO is pumped to HDS. The remainder is routed back as LCR to tray 18. The LCR is cooled in E-1451, E-1022, E-1011 and E-1007.

The MCR section comprises trays 19 to 22 and serves to condense a LGO stream. From total draw-off tray 19 the LGO stream is routed to Middle Distillate Vessel V-1003. Part of the LGO stream is sent to E-1022. The remainder is routed back as MCR to tray 22. The MCR is cooled in E-1452 and E-1002

The kero section comprises trays 23 to 33. From the total draw-off tray 23, the kero stream is routed hot to Middle Distillate Vessel V-1003, and pumped to E-1022

The TCR section comprises trays 34 to 38 and serves to condense kero, leaving a naphtha-minus vapor overhead fraction. The naphtha-minus leaves the top of the column. From partial draw-off tray 34 TCR is pumped around by P-1007A/B and cooled E-1001 while the remainder of duty is removed in air cooler E-1004. The cooled TCR enter column above top tray 38.

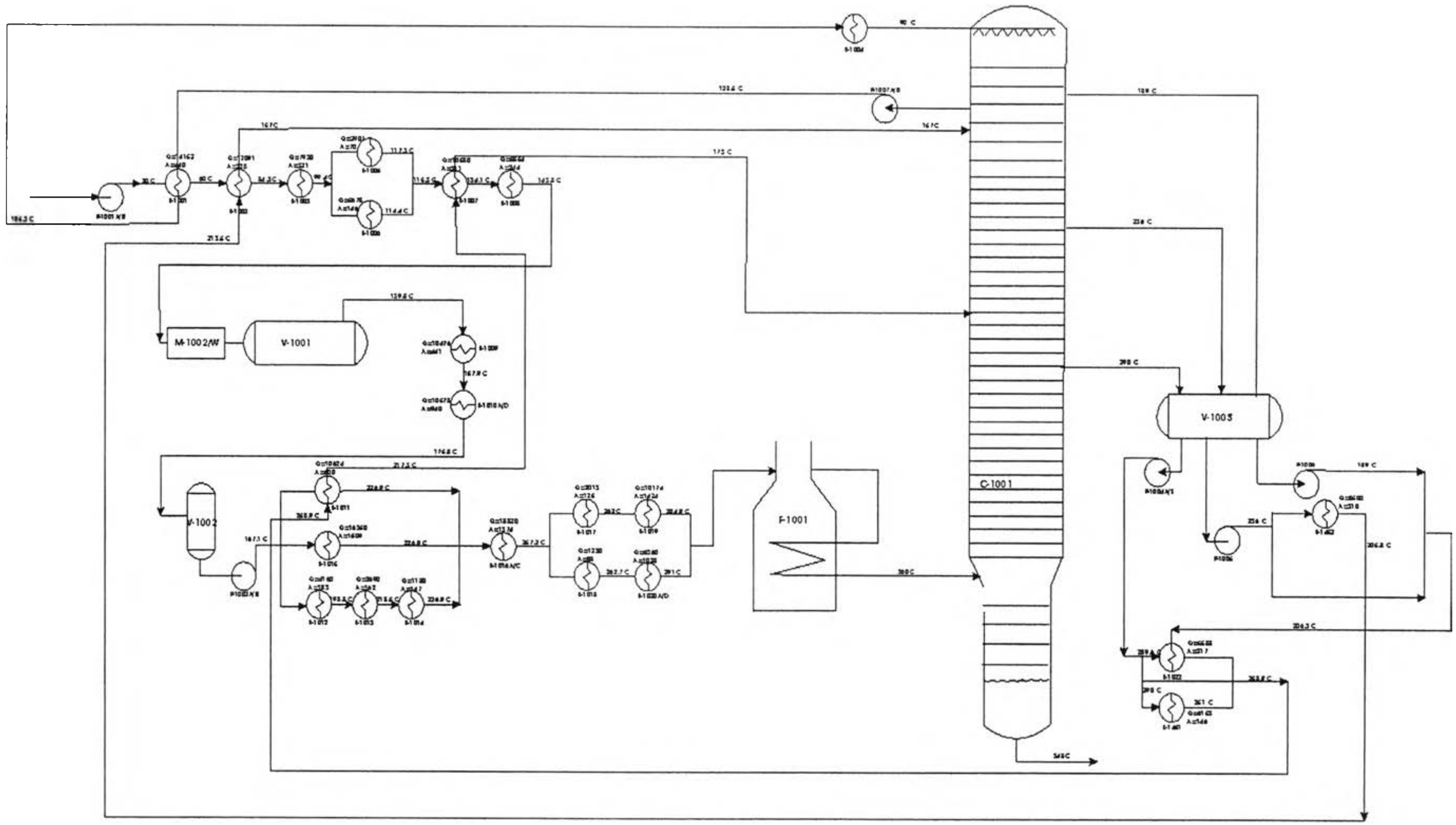


Figure 4.1 The process flow scheme of ARC atmospheric distillation unit

4.2 Operating data of Refinery Plant Performance

Extracting the stream data from the process flow sheet is absolutely crucial part of pinch analysis. The design of existing heat exchanger network data was shown in table 4.1. The temperature, stream flow rates, heat capacity the heat exchanger areas and overall heat transfer coefficients were taken from the design data (all parameters in table 4.1 are not corrected by heat balance)

Table 4.1 Heat exchanger data of ARC atmospheric distillation unit (data sheet)

	Hot fluid			Cold fluid			Overall heat transfer coef. (W/m²/°C)	Area (m²)	Duty (kW)
	Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)	Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)			
E-1001	128.5	105.2	624.42	30	60	456.92	464	440	14152
E-1002	213.5	157	223.9	60.4	84.3	493.34	484	228	12091
E-1003	216.7	143.5	114.45	84.3	99.4	520.24	304	321	7930
E-1004	105.3	90.3	654.512	35	50	-	-	-	8733
E-1005	223.1	159.9	48.29	99.4	117.3	159.64	532	70	2907
E-1006	216.9	164.1	107.54	99.4	114.4	378.5	484	146	5678
E-1007	217.3	173	247.68	115.3	134.1	549.93	506	311	10580
E-1008	233	159.4	79.76	134.1	143.8	570.80	476	244	5554
E-1009	220.4	189.8	350.14	139.8	157.9	572.09	439	441	10476
E-1010	281.6	200	137.34	157.9	175.8	580.84	171	940	10578
E-1011	258.9	217.3	262.96	167.1	226.9	169.08	474	600	10624
E-1012	253	216.9	118.26	167.1	198.8	127.81	476	183	4160
E-1013	276.7	223.1	52.15	198.8	218.5	134.02	500	162	2690
E-1014	285.4	240	26.98	218.5	226.9	139.62	233	147	1180
E-1015	256.6	211.7	376.48	167.1	226.9	260.47	313	1509	16350
E-1016	304.9	256.6	402.2166	226.9	257.2	615.45	369	1374	18820
E-1017	324	280.6	47.88	257.1	262	412.1	416	125	2013
E-1018	330	285.4	28.42	257.2	262.7	222.36	285	98.2	1230
E-1019	350	281.6	153.37	260.4	284.9	416.1	195	1424	10174
E-1020	360	280	83.76	262.7	291	220.35	192	1028	6360
E-1022	289.6	233.4	103.292	206.3	231.1	221.27	577	317	5588
E-1451	290	251	164.23	182	188	254.14	505	146	6153
E-1452	236	205.8	587.26	124.6	126.9	307.92	324	310	8500

Note: E-1004 is air fan cooler

Figure 4.1 shown the process from a crude oil distillation unit. In this study, it divided to 23 hot and 23 cold process streams including air utility. The detail and description of hot and cold stream showed in Table 4.2.

Table 4.2 Hot and cold streams data of ARC atmospheric distillation unit (From design data of ARC refining)

Hot Process Streams			
Streams	Temp. inlet (C)	Temp. outlet (C)	Description
I1	128.5	105.2	CDU-TCR from tray 34 of distillation column
I2	213.5	157	CDU-MCR from tray 23 of distillation column
I3	216.7	143.5	HCU feed : Waxy distillate from unit 1100
I4	223.1	159.9	HDF-LGO from E-1013
I5	216.9	164.1	HDF-LGO from E-1012
I6	217.3	173	CDU-LCR : LCR stream E-1011A/B
I7	233	159.4	HDF-MCR from unit 1350
I8	220.4	189.8	HVU-MCR from unit 1100
I9	281.6	200	VBU Residue from E-1019
I10	258.9	217.3	CDU-LCR from E-1022 and bypass
I11	256.6	211.7	HVU-LCR from E-1016A/C
I12	253	216.9	HDF-HGO from unit 1350
I13	276.7	223.1	HDF-LGO from unit 1350
I14	285.4	240	Wash oil quench from E-1018
I15	304.9	256.6	HVU-LCR from unit 1100
I16	324	280.6	VBU-TCR from unit 1200
I17	330	285.4	Wash oil quench from unit 1100
I18	350	281.6	VBU Residue from unit 1200
I19	360	280	SR quench from unit 11000
I20	105.3	90.3	CDU-TCR from E-1001
I21	236	205.8	CDU-MCR from V-1003
I22	289.6	233.4	LCR-MCR from V-1003
I23	290	251	LCR-MCR from V-1003

Table 4.2 (Continued)

Cold Process Streams			
Streams	Temp. inlet (C)	Temp. outlet (C)	Description
J1	30	60	Crude in E-1001
J2	60.4	84.3	Crude in E-1002
J3	84.3	99.4	Crude in E-1003
J4	99.4	117.3	Crude in E-1005
J5	99.4	114.4	Crude in E-1006
J6	115.3	134.1	Crude in E-1007
J7	134.1	143.8	Crude in E-1008
J8	139.8	157.9	Crude in E-1009
J9	157.9	175.8	Crude in E-1010
J10	167.1	226.9	Crude in E-1011
J11	167.1	226.9	Crude in E-1015
J12	167.1	198.8	Crude in E-1012
J13	198.8	218.5	Crude in E-1013
J14	218.5	226.9	Crude in E-1014
J15	226.9	257.2	Crude in E-1016
J16	257.1	262	Crude in E-1017
J17	257.2	262.7	Crude in E-1018
J18	260.4	284.9	Crude in E-1019
J19	262.7	291	Crude in E-1020
J20	30	50	Air utility
J21	124.6	126.9	Crude in E-1452
J22	206.3	231.1	Crude in E-1022
J23	182	188	Crude in E-1451

Note: J20 is cold utility

The first step necessary is to produce a heat and mass balance for the refinery. Stream matching of this refinery between hot and cold stream to make heat exchanger was presented in Table 4.3 and validated by heat balance calculation in order to confirm the design data of refinery.

Table 4.3 Hot and cold streams data from validation by heat balance calculation

	Matching between hot and cold streams	Hot fluid			Cold fluid				
		Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)	Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)	Overall heat transfer coef. (W/m²/°C)	Area (m²)
E-1001	I1-J1	128.5	105.3	624.42	30	61.71	456.92	464	440
E-1002	I2-J2	213.5	158.66	223.9	61.71	86.597	493.34	484	228
E-1003	I3-J3	216.7	140.46	114.45	86.597	103.37	520.24	304	321
E-1005	I4-J4	223.1	163.03	48.29	103.37	123.57	143.59	532	70
E-1006	I5-J5	216.9	165.10	107.54	103.37	118.09	378.533	484	146
E-1007	I6-J6	217.3	175.10	247.68	119.62	138.62	550	506	311
E-1008	I7-J7	233	162.78	79.76	138.62	148.44	570.41	476	244
E-1009	I8-J8	220.4	192.95	350.14	148.44	165.24	572.09	439	441
E-1010	I9-J9	281.6	205.48	137.34	165.24	183.24	580.84	171	940
E-1011	I10-J10	258.9	217.73	262.96	167.1	231.13	169.077	474	600
E-1015	I11-J11	256.6	212.71	376.48	167.1	230.81	259.33	313	1509
E-1012	I12-J2	253	215.97	118.26	167.1	201.31	128.02	476	183
E-1013	I13-J13	276.7	222.3	52.15	201.31	222.48	134.02	500	162
E-1014	I14-J14	285.4	242.06	26.98	222.48	230.85	139.62	233	147
E-1016	I15-J15	304.9	259.54	402.21	230.85	260.48	615.79	369	1374
E-1017	I16-J16	324	282.98	47.88	260.48	265.25	412.096	416	125
E-1018	I17-J17	330	287.64	28.42	260.48	265.42	222.36	285	98.2
E-1019	I18-J18	350	284.58	153.37	265.25	289.36	416.1	195	1424
E-1020	I19-J19	360	280.34	83.76	265.42	295.7	220.35	192	1028
E-1022	I22-J22	289.6	227.41	103.29	206.3	235.33	221.267	577	317
E-1451	I23-J23	290	254.65	164.23	182	207.84	254.14	505	146
E-1452	I21-J21	236	206.88	287.26	124.6	151.76	307.923	324	310
E-1004	I20-J20	105.3	90.3	654.51	35	50	643.33	-	-

Note: J20 is cold utility

For heat exchanger network, the most helpful representation is "grid diagram" introduced by Linhoff and Flower (1978). The grid diagram is much easier to draw than flow sheet, especially as heat exchangers can be placed in any order redrawing the stream system. Also, the grid diagram represents the countercurrent nature of heat exchange. Finally, the pinch is easily represented in the grid as shown in Figure 4.7

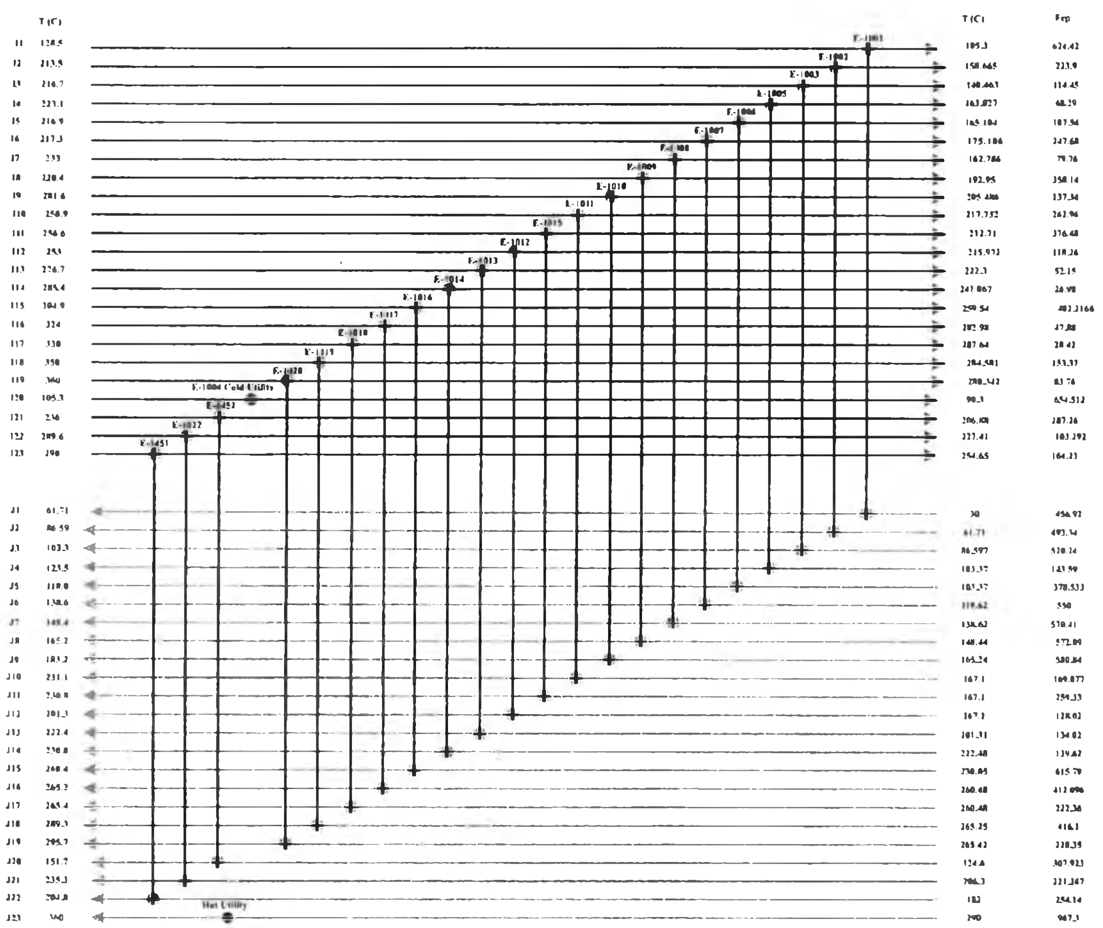


Figure 4.2 The initial grid diagram of hot and cold matching streams data (from validation by heat balance calculation).

Following the grid diagram of refinery in Figure 4.2, cold stream J18 and J19 routed to furnace and to distillation column. The outlet temperatures of J18 and J19 from heat balance calculation were 289.36°C and 295.7°C respectively, compared with design data sheet was 284.9°C of J18 and 291°C of J19. The result showed the difference from heat balance calculation and design data were less than 2%. The base case diagram showed ΔT_{min} of this system was 33.5°C which located from 305°C to 271.5°C as shown in Figure 4.3. There was 3 heat exchanger E-1016, E-1017, and E-1018 which heat transfer inside the heat exchanger crossed the pinch point.

4.3 Retrofit Potential

The problem table analysis and composite curve of the existing network had been done in order to check the retrofit potential of the existing network. The data extracted from composite curve gave the utility consumption, minimum temperature difference (ΔT min), and pinch point of the process.

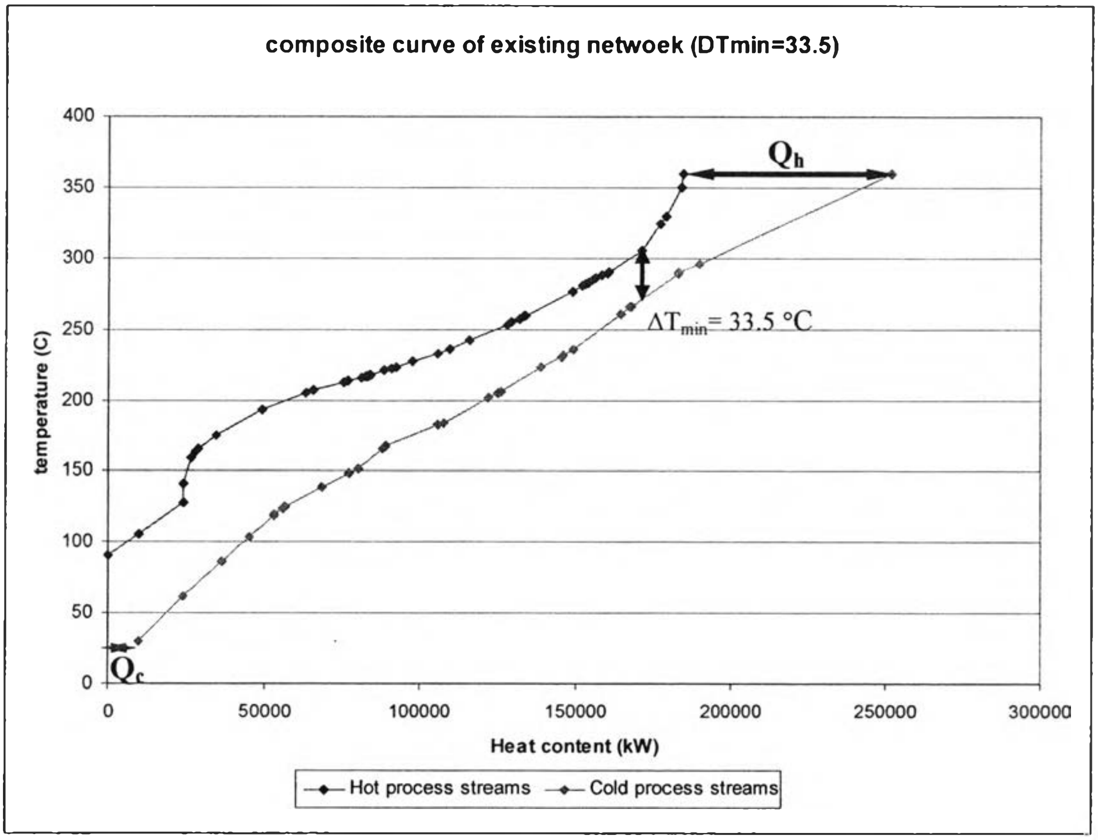


Figure 4.3 The composite curve of existing network (ΔT_{min} = 33.5 °C)

From stream data in Table 4.3, The overlap between the composite curves represented the maximum amount of heat recovery in this process. The overshoot at the bottom of the hot composite represented the minimum amount of external cooling required (Qc) which consumed 9,818 kW and the overshoot at the top of the cold composite represented the minimum amount of external heating required (Qh) which consume 67,536 kW in process and ΔT_{min} of the process was 33.5 °C.

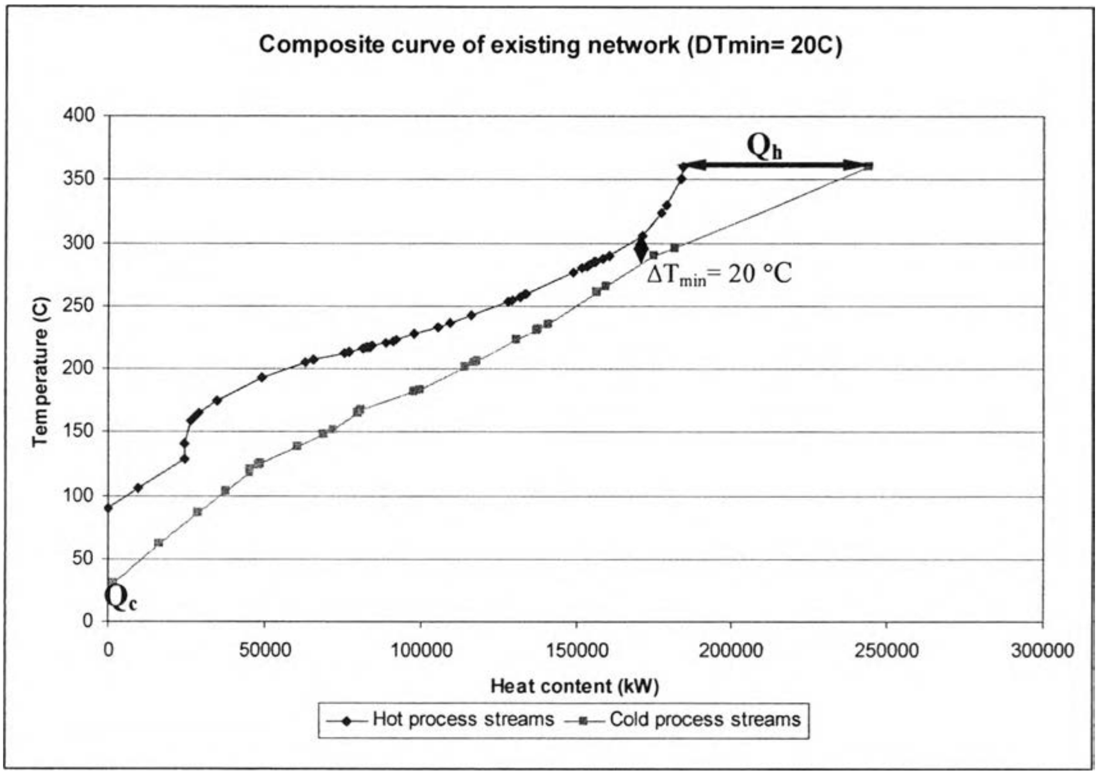


Figure 4.4 The composite curve of existing network when ΔT_{min} = 20°C

The shifted composite curve was re-plotted in Figure 4.4 showed the retrofit potential of the network at ΔT_{min} = 20°C. Compared to the existing network of ΔT_{min} = 33.5°C, hot utility can be reduced from 67,536 kW to 59,571 kW and cold utility can be reduced from 9,818 kW to 1,853 kW. The utility consumption was saved about 12% of hot utility and 81% of cold utility by adding more heat exchanger area.

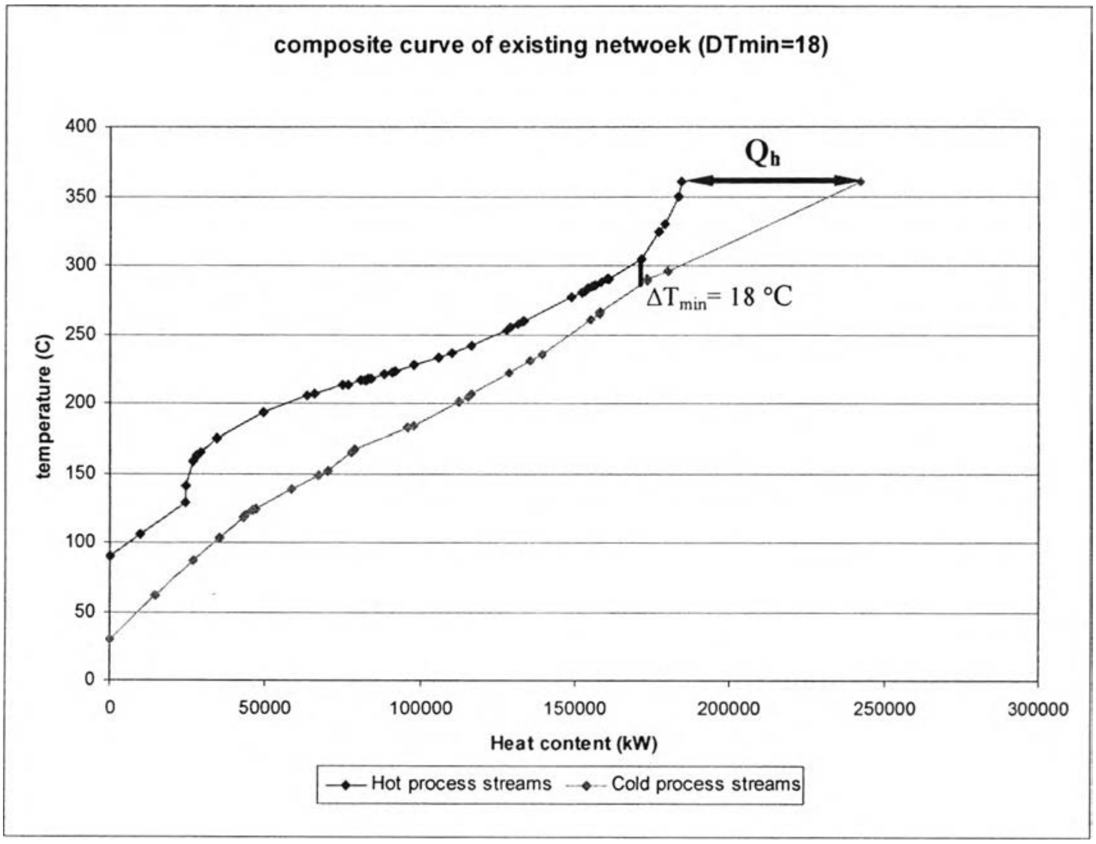


Figure 4.5 The composite curve of existing network when reduce $Q_C = 0$ kW

The shifted composite curve in Figure 4.5 showed ΔT_{min} was reduced, a point was reached at $\Delta T_{min} = 18°C$ where no cold utility was required at all lower values of ΔT_{min}, the only utility needed was 57,718 kW. Compared to the existing network of $\Delta T_{min} = 33.5°C$, hot utility can be reduced from 67,536 kW to 57,718 kW and cold utility can be reduced from 9,650 kW to 0 kW. The utility consumption was saved about 14.5% of hot utility and 100% of cold utility by adding more heat exchanger area.

Table 4.4 The result from shifted composite curve in various ΔT_{min}

ΔT min (°C)	Hot utility consumption (kW)	Cold utility consumption (kW)	% Hot utility consumption	% Cold utility consumption
35.5	67,536	9,818	0	0
30	65,818	8,100	2.54	17.5
20	59,571	1,853	12	81
18	57,718	0	14.5	100

Table 4.4 summarized the utility consumption when ΔT_{min} was varied from base case (ΔT_{min}= 35.5°C). The maximum retrofit potential (ΔT_{min} = 18°C) can reduce 14.5 % of hot utility and 100% of cold utility.

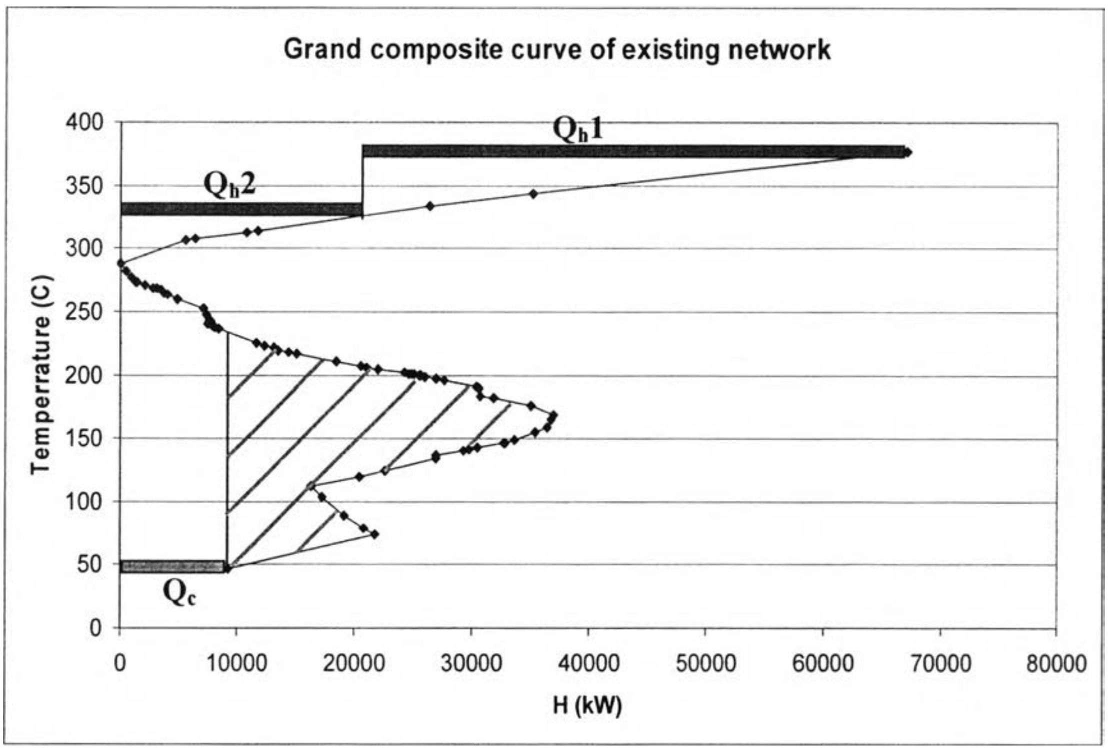


Figure 4.6 The grand composite curve of existing network.

In Figure 4.6, the grand composite curve (GCC) was showed the heat available in various temperature intervals and also represented the net heat flows in the process which was zero at pinch. The GCC helped to better utilize utility resources when design HEN. The maximum cold utility (Q_c) 9,818 kW was located at temperature 50°C which was the low temperature cold utility. A process sink segment as show in Figure 4.6 was Q_h1 plus Q_h2. The process operation in refinery used the hot utility from furnace which was high temperature level as in Q_h1 to heat up the crude to reach the target temperature. To reduce the furnace duty, the stream in lower temperature level than furnace should be applied as shown in Figure 4.6; the Q_h2 can reduce the high level utility usage. When 67,536 kW were provided at 360°C, a furnace was required that burn fuel. Alternatively, the GCC showed that up to 21,000 kW (Q_h2) can be provided at 330°C using less expensive high-pressure steam. This left only 46,536 kW (Q_h1) to be supply by furnace.

4.4 Retrofit by Pinch Analysis

This introduces a technique to retrofit network. It was found on the same thermodynamic principles that underline established pinch technology. In the context of retrofitting, this implies the setting of targets for energy saving, capital cost. The targets recognize the specifics of the existing design. Apply pinch principles and incorporate process insight during the design. Although this approach has been used industrially with some success, it is, strictly speaking, an improvisation on methodology aimed grassroots design. User experience is crucial for a good result.

The retrofit by pinch analysis was applied to the base case refinery. The grid diagram of existing network in Figure 4.7 show the heat exchanger crossing the pinch and eliminating cross-pinch heat exchanger. Position new heat exchanger where possible and a possible network show in Figure 4.8. The new heat exchanger network can save about 1% of hot utility compared with base case refinery and new adding area was 3,323 m^2.

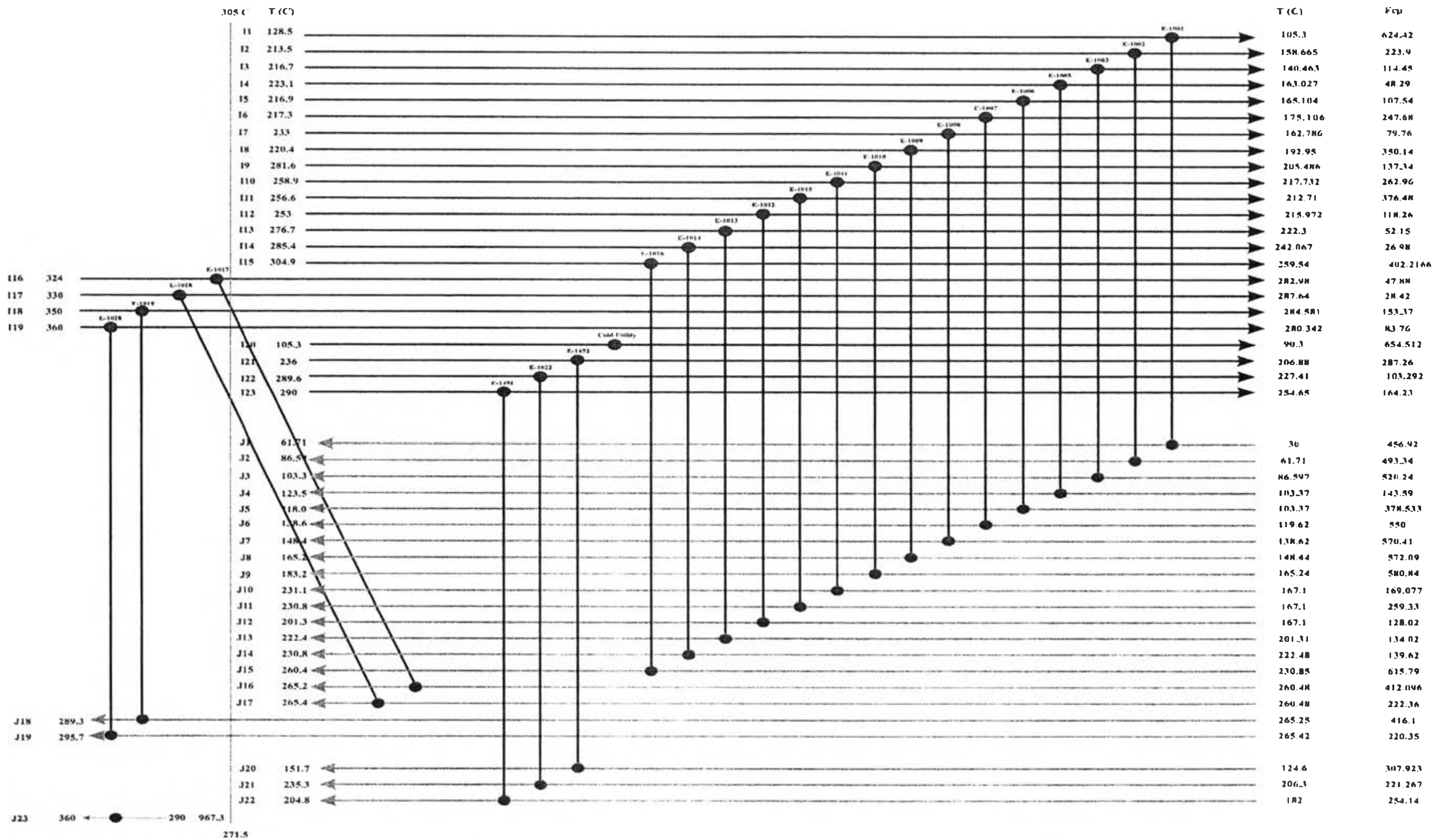


Figure 4.7 The grid diagram of base case refinery and pinch point

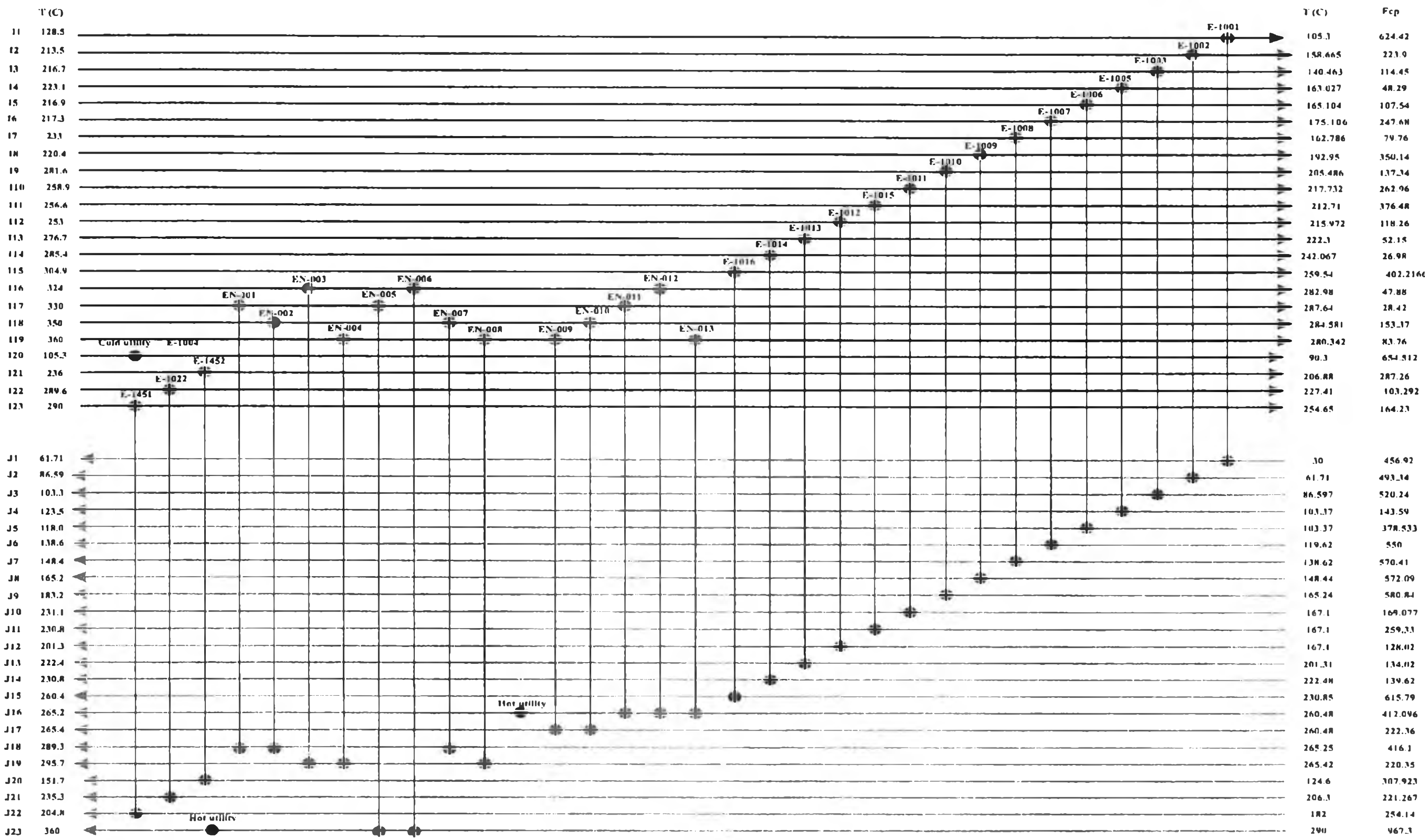


Figure 4.8 The grid diagram of the retrofit potential by pinch analysis

Table 4.5 The result of retrofit potential by pinch analysis

	Hot fluid			Cold fluid					
	Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)	Temp. inlet (°C)	Temp. outlet (°C)	F*Cp (kW/°C)	Overall heat transfer coef. (W/m²/°C)	Area (m²)	Duty (kW)
E-1001	128.5	105.3	624.42	30	61.71	456.92	464	440	14152
E-1002	213.5	158.6	223.9	61.71	86.597	493.34	484	228	12091
E-1003	216.7	140.46	114.45	86.59	103.37	520.24	304	321	7930
E-1005	223.1	163.03	48.29	103.37	123.57	143.59	532	70	2907
E-1006	216.9	165.1	107.54	103.37	118.09	378.53	484	146	5678
E-1007	217.3	175.1	247.68	119.62	138.62	550	506	311	10580
E-1008	233	162.78	79.76	138.62	148.44	570.41	476	244	5554
E-1009	220.4	192.95	350.14	148.44	165.24	572.09	439	441	10476
E-1010	281.6	205.48	137.34	165.24	183.24	580.84	171	940	10578
E-1011	258.9	217.73	262.96	167.1	231.13	169.07	474	600	10624
E-1015	256.6	212.71	376.48	167.1	230.81	259.33	313	1509	4160
E-1012	253	215.97	118.26	167.1	201.31	128.02	476	183	2690
E-1013	276.7	222.3	52.15	201.31	222.48	134.02	500	162	1180
E-1014	285.4	242.06	26.98	222.48	230.85	139.62	233	147	16350
E-1016	304.9	259.54	402.21	230.85	260.48	615.79	369	1374	18820
EN-001	330	311.35	28.42	288.1	289.36	416.1	231	74	530
EN-002	350	305	153.37	305	288.1	416.1	195	1233	6901
EN-003	324	308.8	47.88	292.4	295.4	220.35	263	126	726
EN-004	360	305	83.76	305	292.4	220.35	192	494	4607
EN-005	311.3	305	28.42	290.2	290.4	967.3	266	38	180.5
EN-006	308.8	305	47.88	290	290.2	967.3	312	35	184
EN-007	305	298.75	153.37	265.25	305	416.1	195	480	2600
EN-008	305	289	83.76	265.42	305	220.35	192	247	1340
EN-009	289	282.2	83.76	262.87	265.42	222.36	229	116	567
EN-010	298.7	284.58	153.37	260.48	262.87	222.36	232	93	531
EN-011	305	287.64	28.42	263.42	264.6	412.09	338	46	493
EN-012	305	282.98	47.88	260.86	263.42	412.09	416	82	1054
EN-013	282.2	280.34	83.76	260.48	260.86	412.09	262	29	158
E-1022	289.6	227.41	103.29	206.3	235.33	221.26	577	317	5588
E-1451	290	254.65	164.23	182	207.84	254.14	505	146	6153
E-1452	236	206.88	287.26	124.6	151.76	307.92	324	310	8500
E-1004	105.3	90.3	654.51	35	50	643.33	-	-	8733

4.5 Grassroots Model by Mathematical Programming

4.5.1 Grassroots Model 1

This model consist 23 hot and 23 cold processes with one hot and one cold utility. This model was considered only the utility consumption that means heat exchanger cost was neglected. The details of hot and cold streams cost were shown in table 4.6

Table 4.6 Cost data for model 1 (A. Babaro and M.J. Bagajewicz 2005)

Utilities	Cost \$/ (MJ/hr-year)	Cost \$/(kW/year)
I24	19.75	71.1
J24	1.861	6.7
Heat Exchanger Cost	Neglected	

This model consisted of 23 hot and 23 cold processes streams with one heating and cooling utility. The model 1 was solved by one heat transfer zone.

Table 4.7 The result of model 1 from GAMS

ΔT min (°C)	Hot utility (kW)	Hot utility Cost (\$/kW-year)	Cold utility (kW)	Cold utility cost (\$/kW-year)	Number of heat ex-changer
21.5	59798.34	4251661.856	2006.01	13440.299	65

The result of grassroots model in Table 4.7 showed the retrofit potential of new network at ΔT = 21.5°C. Compared to the existing network of ΔT = 33.5°C, hot utility can be reduced from 67536 kW to 59798.34 kW and cold utility can be reduced from 9818 kW to 2006.01 kW. The utility consumption was saved about 11.46% of hot utility and 79.57% of cold utility compared to the existing heat exchanger network with ΔT = 33.5°C.

Table 4.8 The result of model 1

No.of heat ex-changer	Duty (kW)	Area (m2)	No.of heat ex-changer	Duty (kW)	Area (m2)
E1	6677.268	196.2	E34	4229.439	216.6
E2	7809.276	313	E35	1022.225	61
E3	3136.184	218	E36	2189.466	183
E4	5925.851	159	E37	2189.466	188.1
E5	399.166	8.565	E38	1418.48	73.76
E6	2816.355	88.85	E39	1418.48	80.86
E7	4468.477	145.17	E40	584.562	44.67
E8	4256.848	142.17	E41	584.562	48.37
E9	1450.463	39.43	E42	10436.733	883
E10	1450.463	52.65	E43	1965.698	200
E11	2785.071	82.74	E44	1098.458	118.67
E12	2785.071	101.25	E45	4743.656	663
E13	1450.055	47	E46	981.997	275.4
E14	2387.769	68.13	E47	982.041	81.93
E15	1387.48	42.32	E48	1057.47	147.71
E16	5225.305	218.1	E49	146.401	8.84
E17	2800.134	89.66	E50	3462.637	1194
E18	2800.134	149.86	E51	3209.517	212.3
E19	4805.671	155.97	E52	4306.518	1781.4
E20	2615.889	124.1	E53	1170.05	258.28
E21	2189.782	115.95	E54	4556.743	367.83
E22	1363.954	81.34	E55	2006.015	125
E23	3862.795	375	E56	7811.665	365
E24	5226.748	227.7	E57	3863.051	170.75
E25	771.302	50.7	E58	319.455	15.3
E26	712.813	37.74	E59	4182.506	381.4
E27	1418.602	150.7	E60	3211.865	291.3
E28	6753.823	433.28	E61	3211.865	357
E29	1168.997	47	E62	1168.619	125.5
E30	2409.82	259.37	E63	624.213	44.29
E31	2902.279	231.72	E64	4012.698	543.6
E32	1418.602	330	E65	59798.338	373.7
E33	4541.343	1008	-	-	-

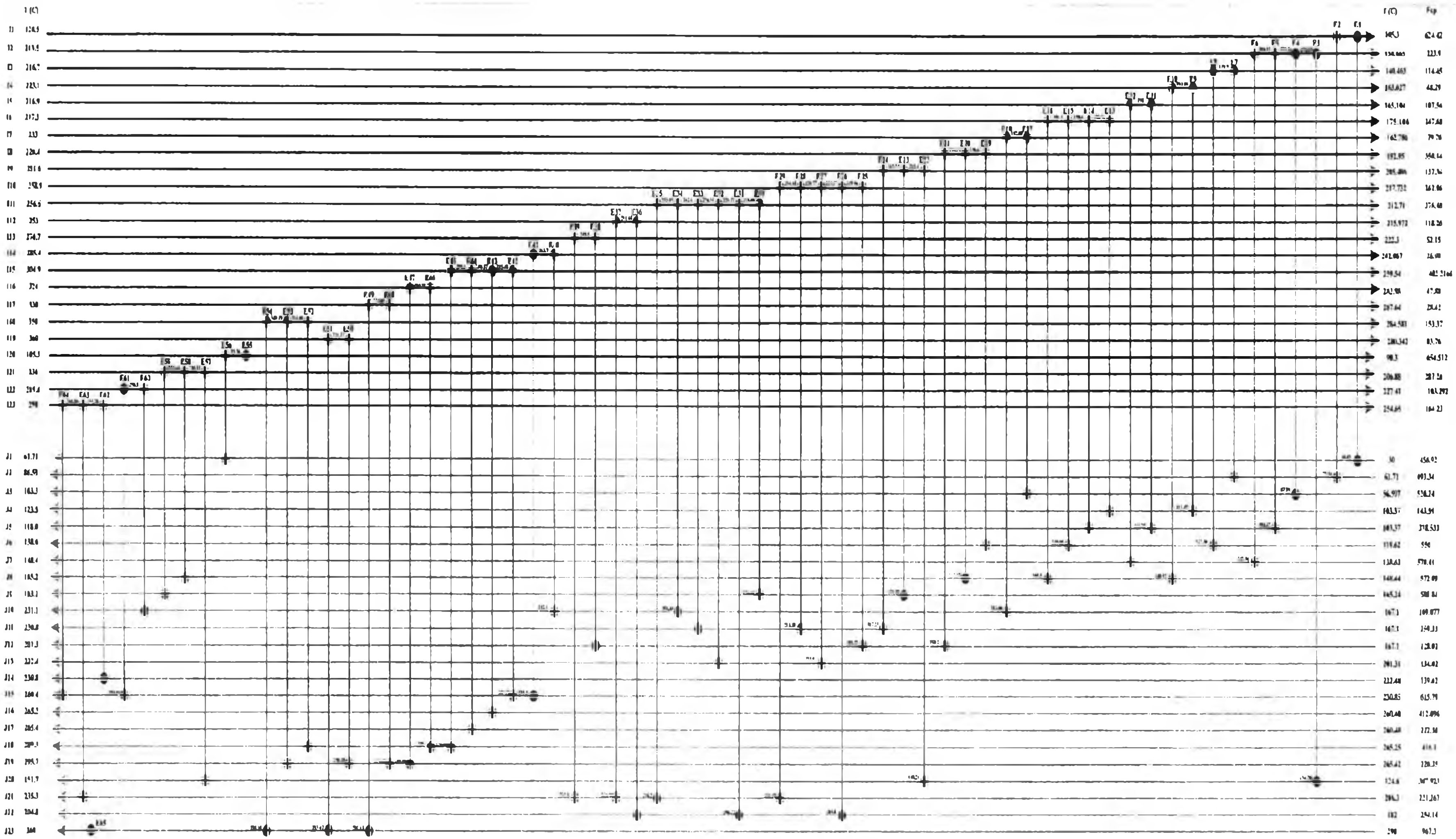


Figure 4.9 The grid diagram of grassroots model 1 from Table 4.6.

4.5.2 <u>Grassroots Model 2</u>

For this model, both cost of heat exchanger and utility was considered. The details of hot and cold streams cost were are shown in table 4.9

Table 4.9 Cost data for model 2 (A. Babaro and M.J. Bagajewicz 2005)

Utilities	Cost \$/ (MJ/hr-year)	Cost \$/(kW/year)
I24	19.75	71.1
J24	1.861	6.7
Heat Exchanger Cost	5291.9+77.79A (\$/year)	

This model consists of 24 hot and 24 cold streams, one heating and cooling utility. The model 1 was solved by one heat transfer zone.

Table 4.10 The result of model 2 from GAMS

ΔT min (°C)	Hot utility (kW)	Hot utility Cost (\$/kW-year)	Cold utility (kW)	Cold utility cost (\$/kW-year)	Number of heat ex-changer
22	59944.74	4262071	2152.41	14421.2	43

The result of grassroots model in Table 4.10 showed the retrofit potential of new network at ΔT = 22°C. Compared to the existing network of ΔT = 33.5°C, hot utility can be reduced from 67536 kW to 59944.74 kW and cold utility can be reduced from 9818 kW to 2152.41 kW. The utility consumption was saved about 11.24% of hot utility and 70.7% of cold utility compared to the existing heat exchanger network with ΔT = 33.5°C.

Table 4.11 Hot and cold streams data from grassroots mode2

	Matching between hot (I) and cold (J)	Hot fluid		Cold fluid				
		Temp. inlet (°C)	Temp. outlet (°C)	Temp. inlet (°C)	Temp. outlet (°C)	Overall heat transfer coef. (W/m²/°C)	Area (m²)	Duty (kW)
E1	I1-J1	112.88	105.3	30	40.37	464	138	4738.135
E2	I1-J2	128.5	112.88	61.71	81.47	473.8	419	9748.409
E3	I2-J6	205.34	158.665	119.62	138.62	494.7	409	10450
E4	I2-J7	213.5	205.34	138.62	141.82	480	55	1827.556
E5	I3-J3	216.7	141.05	86.597	103.24	340	317	8658.466
E6	I4-J4	223.1	163.027	103.37	123.57	532	70	2900.518
E7	I5-J24	189.5	165.104	35.466	50	329.7	48	2085.557
E8	I5-J20	216.9	189.5	124.6	135.9	400.7	124	3484.585
E9	I6-J5	197.6	175.106	103.37	118.09	494.7	149	5572.006
E10	I6-J20	217.3	197.6	135.9	151.76	408.1	187	4878.604
E11	I7-J2	194.5	162.786	81.47	86.597	480	56	2529.343
E12	I7-J7	215.13	194.5	141.82	144.7	476	56.5	1645.804
E13	I7-J22	233	215.13	182	187.6	490.1	75	1425.121
E14	I8-J8	220.4	192.95	148.44	165.24	439	441	9610.705
E15	I9-J7	220.98	205.486	144.7	148.44	251.6	127	2128.066
E16	I9-J11	275.98	220.98	167.1	196.23	211.7	541	7553.465
E17	I9-J12	281.6	275.98	167.1	173.13	251.6	28	771.966
E18	I10-J11	239.62	217.732	196.23	218.43	377	630	5756.585
E19	I10-J21	258.9	239.62	206.3	229.21	520.4	310	5068.953
E20	I11-J9	218.26	212.71	165.24	168.84	221.17	195	2090.109
E21	I11-J10	247	218.26	167.1	231.13	377	952.5	10826
E22	I11-J12	256.6	247	173.13	201.31	377.6	148	3607.598
E23	I12-J22	253	215.972	187.6	204.84	490	239	4378.931
E24	I13-J13	276.7	222.3	201.31	222.48	500	162	2836.96
E25	I14-J14	285.4	242.06	222.48	230.85	233	147	1168.619
E26	I15-J15	285.85	259.54	230.85	248.03	369	868	10582.89
E27	I15-J16	290.74	285.85	260.48	265.25	391.1	197	1965.698
E28	I15-J18	304.9	290.74	265.25	278.94	255.16	868	5695.956
E29	I16-J19	324	282.98	265.42	274.33	262.7	242	1964.038
E30	I17-J19	330	287.764	274.33	279.8	229.4	189	1203.628
E31	I18-J18	312.85	284.581	278.94	289.36	195	1777	4336.215
E32	I18-J19	320.3	312.85	279.8	285	193.5	172	1140.354
E33	I18-J23	350	320.3	290	294.7	326.36	356	4556.743

Table 4.11 (Continued)

	Matching between hot (I) and cold (J)	Hot fluid		Cold fluid				
		Temp. inlet (°C)	Temp. outlet (°C)	Temp. inlet (°C)	Temp. outlet (°C)	Overall heat transfer coef. ($W/m^2/°C$)	Area (m^2)	Duty (kW)
E34	I19-J17	293.46	280.342	260.48	265.42	229.4	202	1098.458
E35	I19-J19	321.68	293.46	285	295.7	192	788	2364.179
E36	I19-J23	360	321.68	294.7	298	322	236	3209.517
E37	I20-J1	105.2	90.3	40.37	61.71	464	451	9750.16
E38	I21-J9	236	206.88	168.84	183.24	223.8	830	8365.011
E39	I22-J11	258.47	227.41	218.43	230.81	405.8	477	3211.865
E40	I22-J15	289.6	258.47	248.03	253.25	450.1	344	3211.865
E41	I23-J21	262.9	254.65	229.21	235.33	474.2	107.8	1354.428
E42	I23-J15	290	262.9	253.25	260.48	385	651	4451.102
E43	I24-J23	500	480	35.466	50	1000	375	59944.74

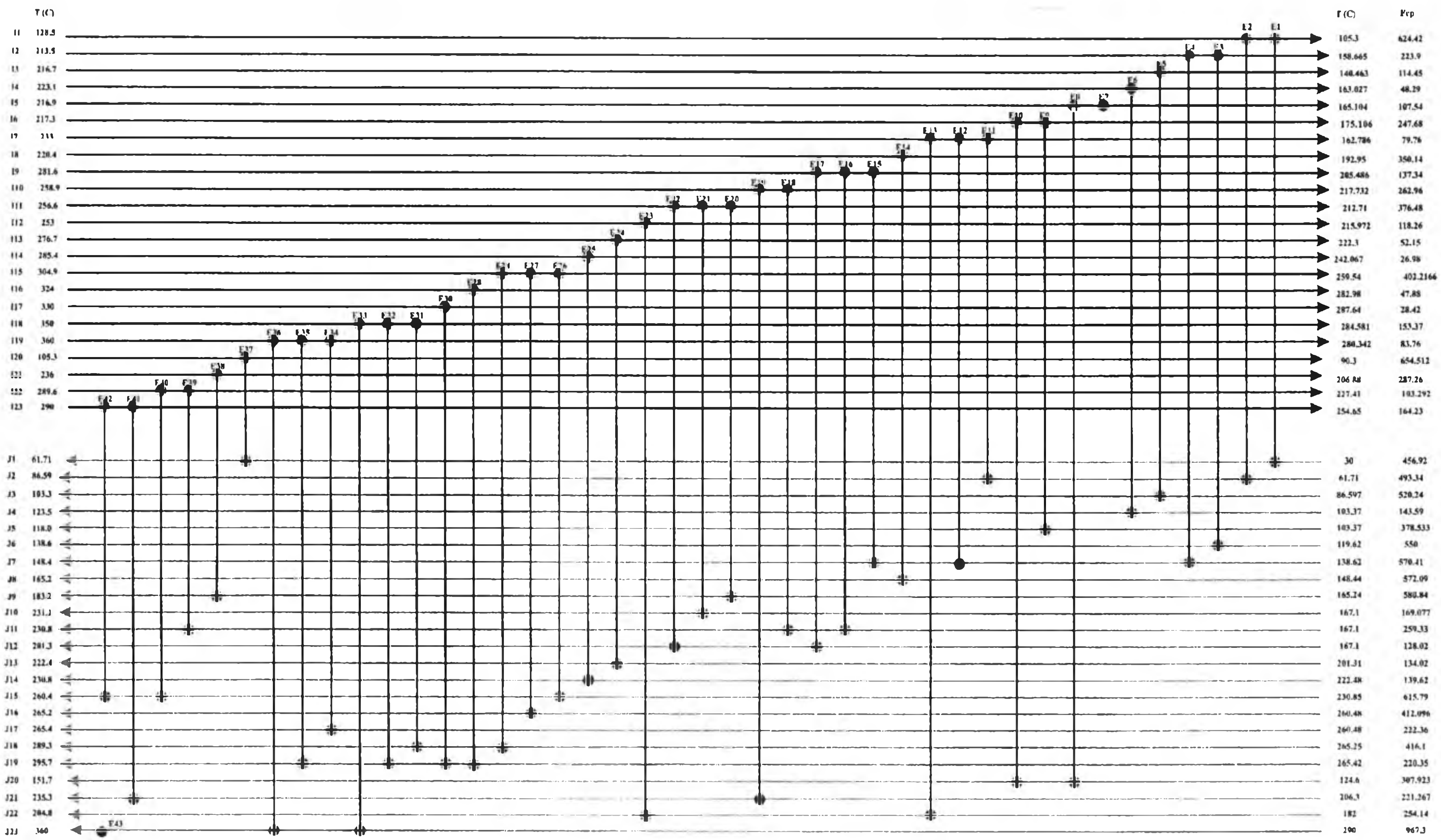


Figure 4.10 The grid diagram of grassroots model 2 from Table 4.9.

4.6 Heat Exchanger Network Design of a Crude Refinery with Multiple Types of Crude

The crude true boiling point (TBP) or ASTM D-86 of three types investigated crude oil was applied in petroleum assay in PRO II.

Table 4.12 The crude true boiling point (TBP)

Crude	Fraction, %							
	5	10	30	50	70	90	95	98
Light Arabian	89	130	243	338	444	610	694	790
Rebco	92	138	266	377	503	703	806	924
Syrian	102	153	298	425	570	801	920	1058

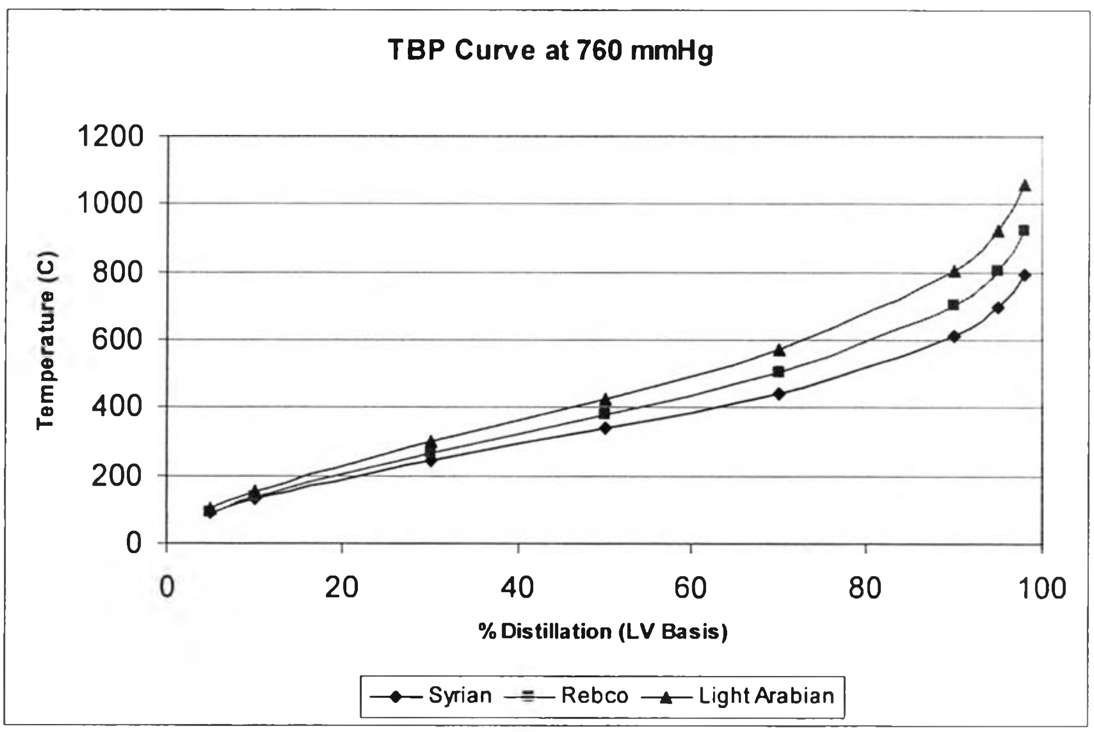


Figure 4.11 True boiling point (TBP) curve at 760 mmHg.

Table 4.13 Product Specifications (Watkins R. N.,1979)

Product	Specification
Naphtha	D86 (95% point) = 182 °C
Kerosene	D86 (95% point) = 271 °C
Diesel	D86 (95% point) = 327 °C
Gas oil	D86 (95% point) = 377-410 °C

Table 4.14 Gap Specifications (Watkins R. N.,1979)

Products	Specification
Kerosene-Naphtha	(5-95) Gap $\geq$ 16.7 °C
Diesel-Kerosene	(5-95) Gap $\geq$ 0 °C
Gas oil-Diesel	(5-95) Gap -5.6°C to -11°C

From the structure of base case refinery, the structure was applied for 3 types of crude represented as the crude used in refinery. The product specification of operation in refining set D86 (95% point) of naphtha was 182°C and Gap which is used to quantify the overlap between adjacent fraction of naphtha and mixed kerosene was 16.7°C.

4.6.1 Heat exchanger network design of a crude refinery with Light Arabian Crude

4.6.1.1 Base case simulation with Light Arabian crude

From the base case structure, the simulation was applied with Light Arabian crude. The product specification of operation in refining set D86 (95% point) of naphtha was 182°C and Gap between adjacent fraction of naphtha and mixed kerosene was 16.7°C.

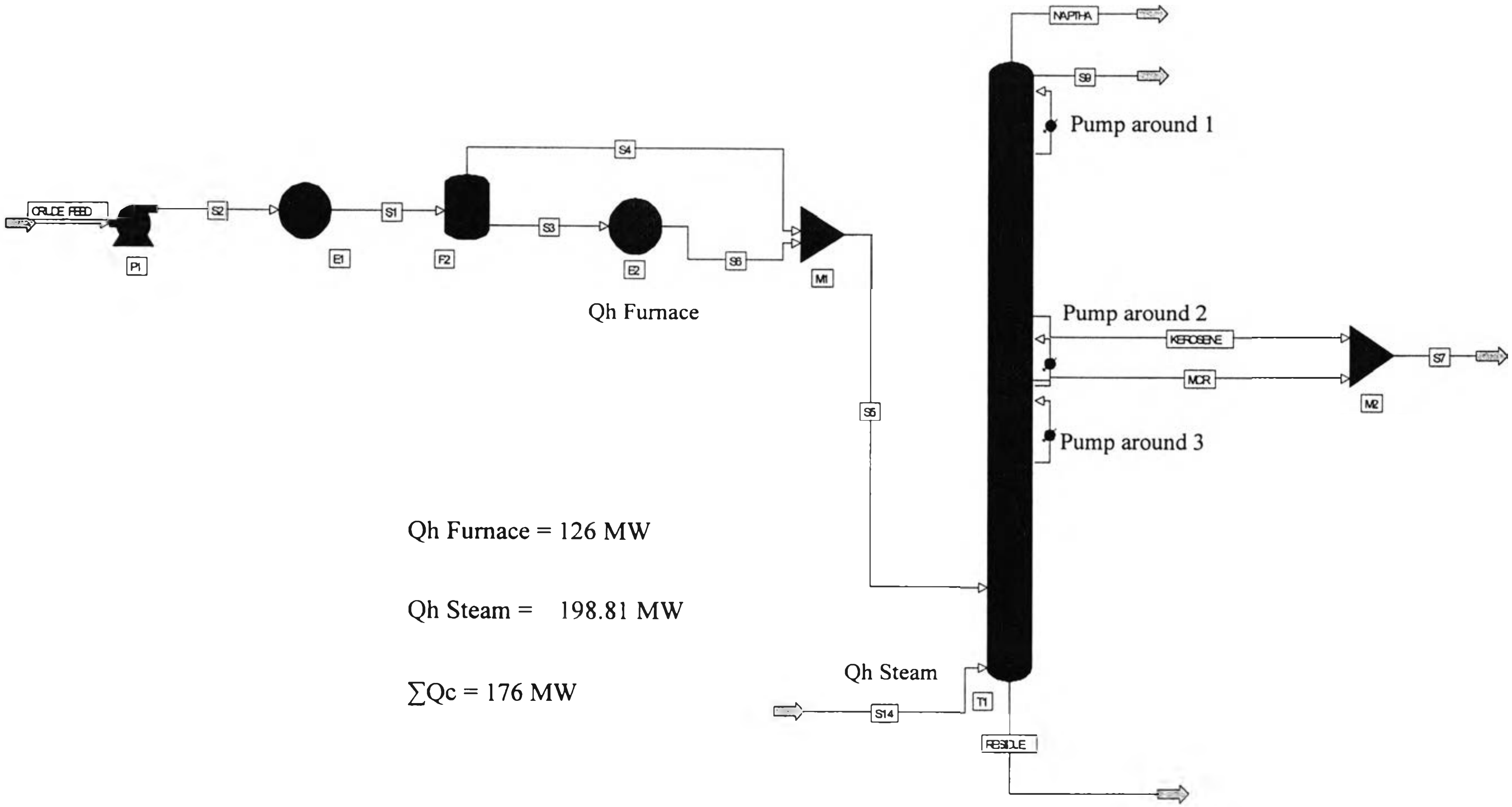


Figure 4.12 The base case refining with feed of Light Arabian crude.

Table 4.15 and 4.16 show the simulation details of base case refinery with Light Arabian crude. The detail was used to be parameters in grassroots design by GAMS.

Table 4.15 Stream properties of Light Arabian crude by PRO II

Stream	Phase	Total Mass Rate (kg / sec)	Temperature (C°)	Total Specific Enthalpy (kW-hr / kg)	Total Cp (kcal/kg-K)
Crude	Liquid	231.94	29.999994	0.016773224	0.44293865
Kerosene	Liquid	42.12990751	196.725573	0.107400785	0.5954481
MCR	Liquid	35.17008672	240.761750	0.14540579	0.63587909
Naphtha	Vapor	68.2375031	113.431541	0.397021351	0.48403882
Residue	Liquid	114.6224743	325.331192	0.202449499	0.68478607
S1	Liquid	231.9400024	174.99999	0.103356897	0.58349216
S2	Liquid	231.9400024	31.32123	0.017166776	0.44514133
S3	Liquid	231.9400024	166.99999	0.09796075	0.57645895
S4	N/A	N/A	N/A	N/A	N/A
S5	Mixed	231.9400024	360.00002	0.249273225	0
S6	Mixed	231.9400024	360.00002	0.249273225	0
S7	Mixed	77.29999424	209.322170	0.1246923676	0
S9	Water	29.77998475	113.431541	0.1321116	1.01152915
S14	Vapor	58	475.00002	0.952164939	0.50549288

Table 4.16 Pump around details of Light Arabian crude by PRO II

PRO II	GAMS	Tray		Temperature (C°)		RATES	Duty
		From	To	From	To	(K*Kg/sec)	(Mw)
Pump around 1 (PA1)	I1	5	1	121	90.3	1.49	103
Pump around 2 (PA2)	I2	20	17	240.8	157	0.135	28.45
Pump around 3 (PA3)	I3	25	21	292	173	0.146	45.55

4.6.1.2 Network form grassroots model of Light Arabian crude by GAMS

The result network by using grassroots model with Light Arabian crude was showed in Figure 4.13.

Table 4.17 Stream details list for using in grassroots model

Stream Number	Description
I1	Stream line of pump around 1 (PA1) from Figure 4.8
I2	Stream line of pump around 2 (PA2) from Figure 4.8
I3	Stream line of pump around 3 (PA3) from Figure 4.8
I4	Hot utility
J1	Stream line S2 to S1 from Figure 4.8
J2	Stream line S3 passed S6 and to S5 from Figure 4.8
J3	Stream line S7 to 240°C from Figure 4.8
J4	Cold Utility

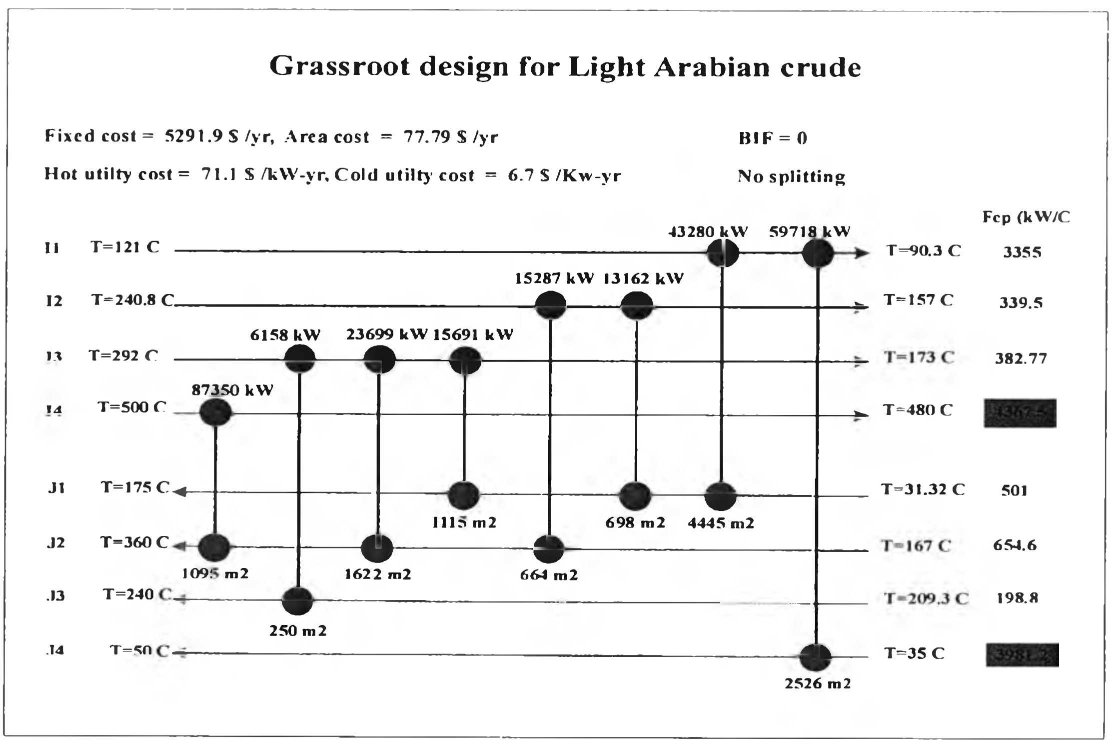
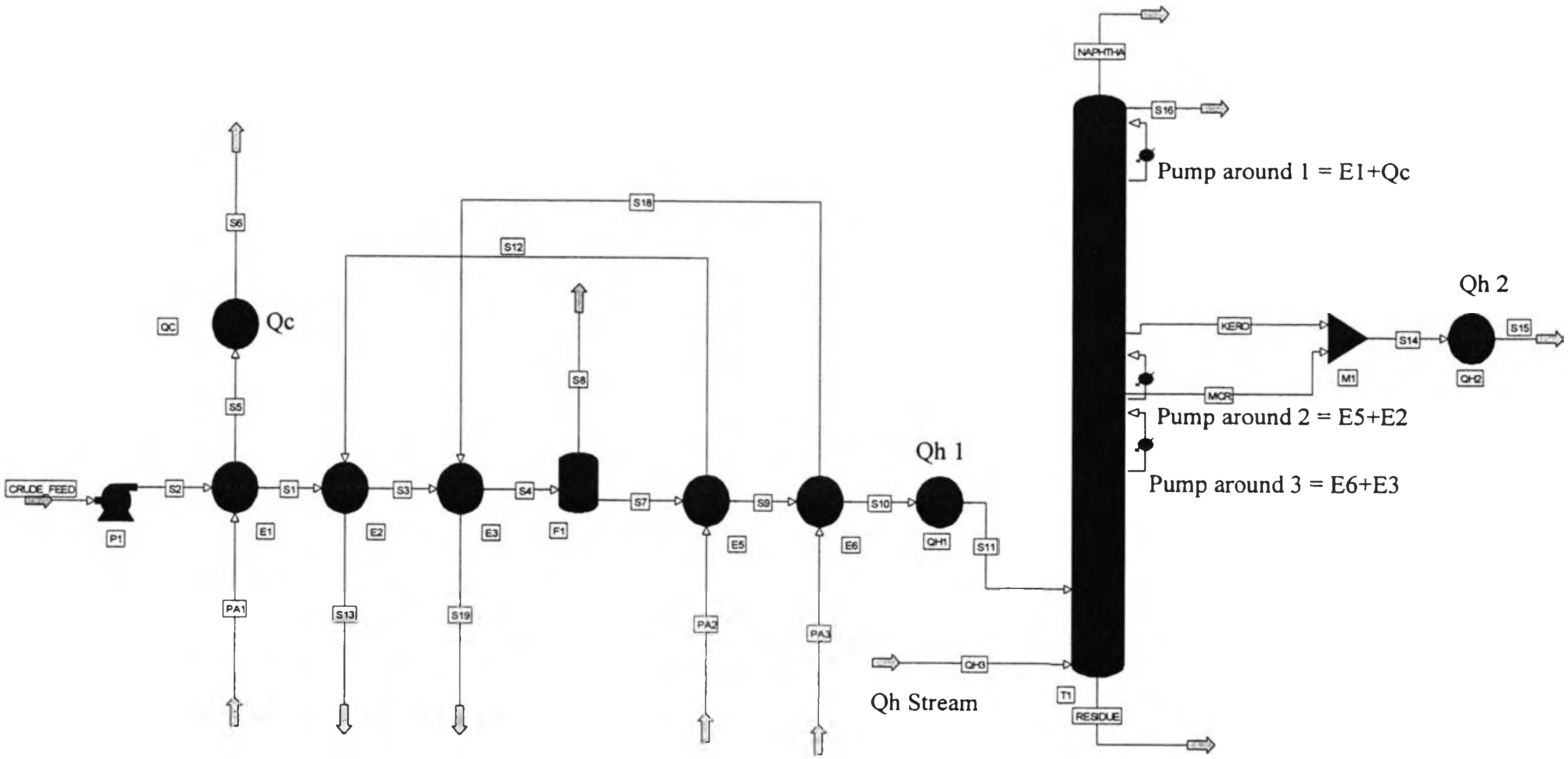


Figure 4.13 The result from grassroots model of Light Arabian crude by GAMS.

4.6.1.3 Validation of new network by PRO II

From Figure 4.13, the simulation network and the details of energy consumption was showed in Figure 4.14.



∑Qh Steam = 137.112 ∑Qh Furnace = 87.36 MW

∑Qc = 11.41 MW

Figure 4.14 The validation network (HEN1) by PRO II .

4.6.2 Heat exchanger network design of a crude refinery with Rebco crude

4.6.2.1 Base case structure of studying Rebco crude

From the base case structure, the simulation was applied with Rebco crude. The product specification of operation in refining set D86 (95% point) of naphtha was 182°C and Gap between adjacent fraction of naphtha and mixed kerosene was 16.7°C.

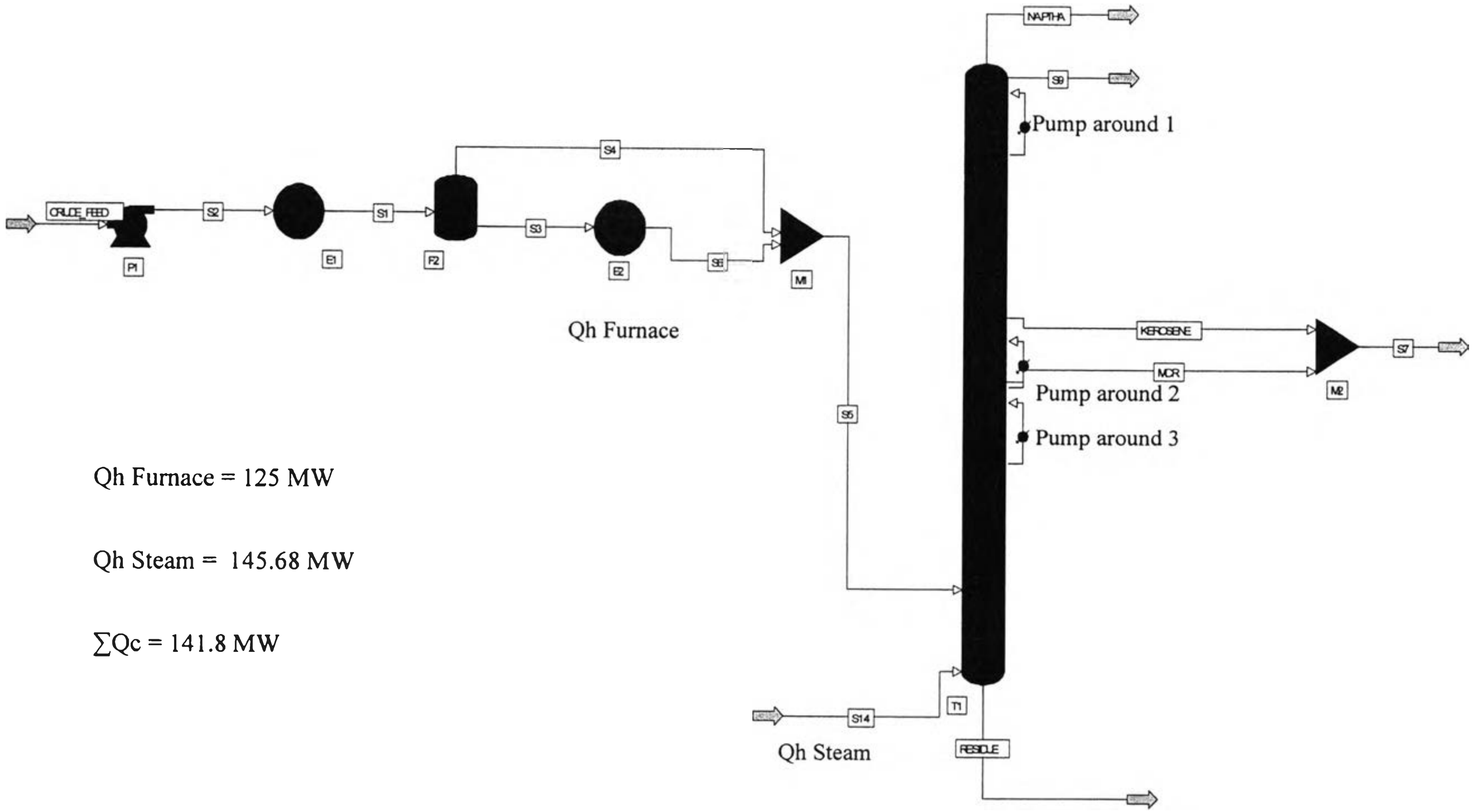


Figure 4.15 The base case refinery with feed of Rebco crude.

Table 4.18 and 4.19 show the simulation details of base case refinery with Light Arabian crude. The detail was used to be parameters in grassroots design by GAMS.

Table 4.18 Stream properties of Rebco crude by PRO II

Stream	Phase	Total Mass Rate (kg / sec)	Temperature (C°)	Total Specific En-thalpy (kW-hr / kg)	Total Cp (kcal/kg-K)
Crude	Liquid	231.9400024	29.9999939	0.017388703	0.439954508
Kerosene	Liquid	42.12939424	184.6271529	0.108403228	0.596831323
MCR	Liquid	35.16991374	240.1958825	0.145727016	0.637360976
Naphtha	Vapor	56.13526672	113.1127276	0.395495109	0.485114218
Residue	Liquid	121.5790409	321.4883349	0.198565575	0.679003876
S1	Liquid	231.9400024	174.9999939	0.103478408	0.580946877
S2	Liquid	231.9400024	31.3462459	0.017775869	0.44222167
S3	Liquid	231.9400024	166.9999939	0.098106008	0.573899453
S4	N/A	N/A	N/A	N/A	N/A
S5	Mixed	231.9400024	360.0000244	0.247948116	0
S6	Mixed	231.9400024	360.0000244	0.247948116	0
S7	Mixed	77.29930797	210.1238725	0.125384938	0
S9	Water	19.42635956	113.1127276	0.131736627	1.011429297
S14	Vapor	42.5	475.0000244	0.952164939	0.505492884

Table 4.19 Pump around details of Rebco crude by PRO II

PRO II	GAMS	Tray		Temperature (C°)		RATES	Duty
		From	To	From	To	(K*Kg/sec)	(Mw)
Pump around 1 (PA1)	I1	5	1	121.8	90.3	0.558	69.77
Pump around 2 (PA2)	I2	20	17	240.2	157	0.135	28.34
Pump around 2 (PA2)	I3	25	21	287	173	0.146	43.67

4.6.2.2 Network form grassroots model of Rebco crude by GAMS

The result network by using grassroots model with Rebco crude was showed in Figure 4.16.

Table 4.20 Stream details list for using in grassroots model

Stream Number	Description
I1	Stream line of pump around 1 (PA1) from Figure 4.11
I2	Stream line of pump around 2 (PA2) from Figure 4.11
I3	Stream line of pump around 3 (PA3) from Figure 4.11
I4	Hot utility
J1	Stream line S2 to S1 from Figure 4.11
J2	Stream line S3 passed S6 and to S5 from Figure 4.11
J3	Stream line S7 to 240°C from Figure 4.11
J4	Cold Utility

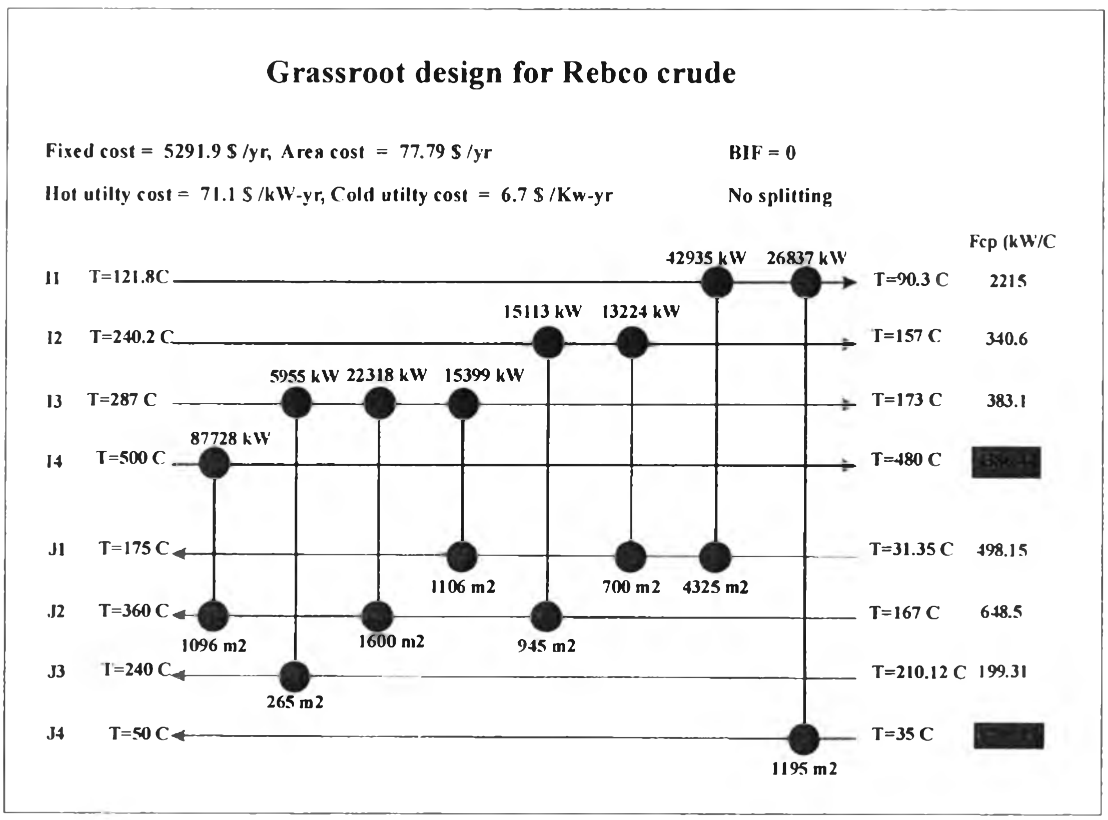
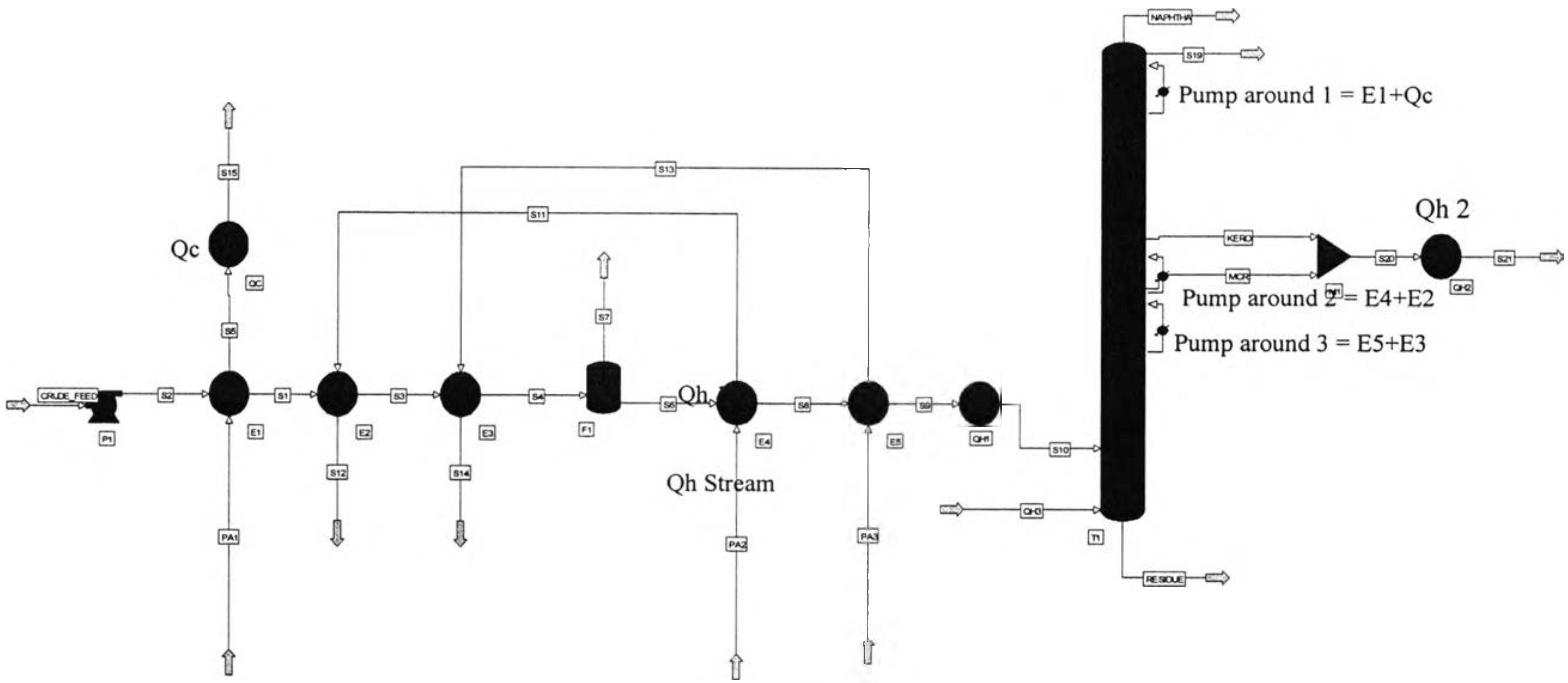


Figure 4.16 The result from grassroots model of Rebco crude by GAMS.

4.6.2.3 Validation of new network by PROII

From Figure 4.16, the simulation network and the details of energy consumption was showed in Figure 4.17.



$\sum$Qh Steam = 137.11 MW

$\sum$Qh Furnace = 87.68 MW

$\sum$Qc = 20.68 MW

Figure 4.17 The new network (HEN2) by PRO II.

4.6.3 Base case structure of studying Syrian crude

4.6.3.1 *Base case structure of studying Syrian crude*

From the base case structure, the simulation was applied with Syrian crude. The product specification of operation in refining set D86 (95% point) of naphtha was 182°C and Gap between adjacent fraction of naphtha and mixed kerosene was 16.7°C.

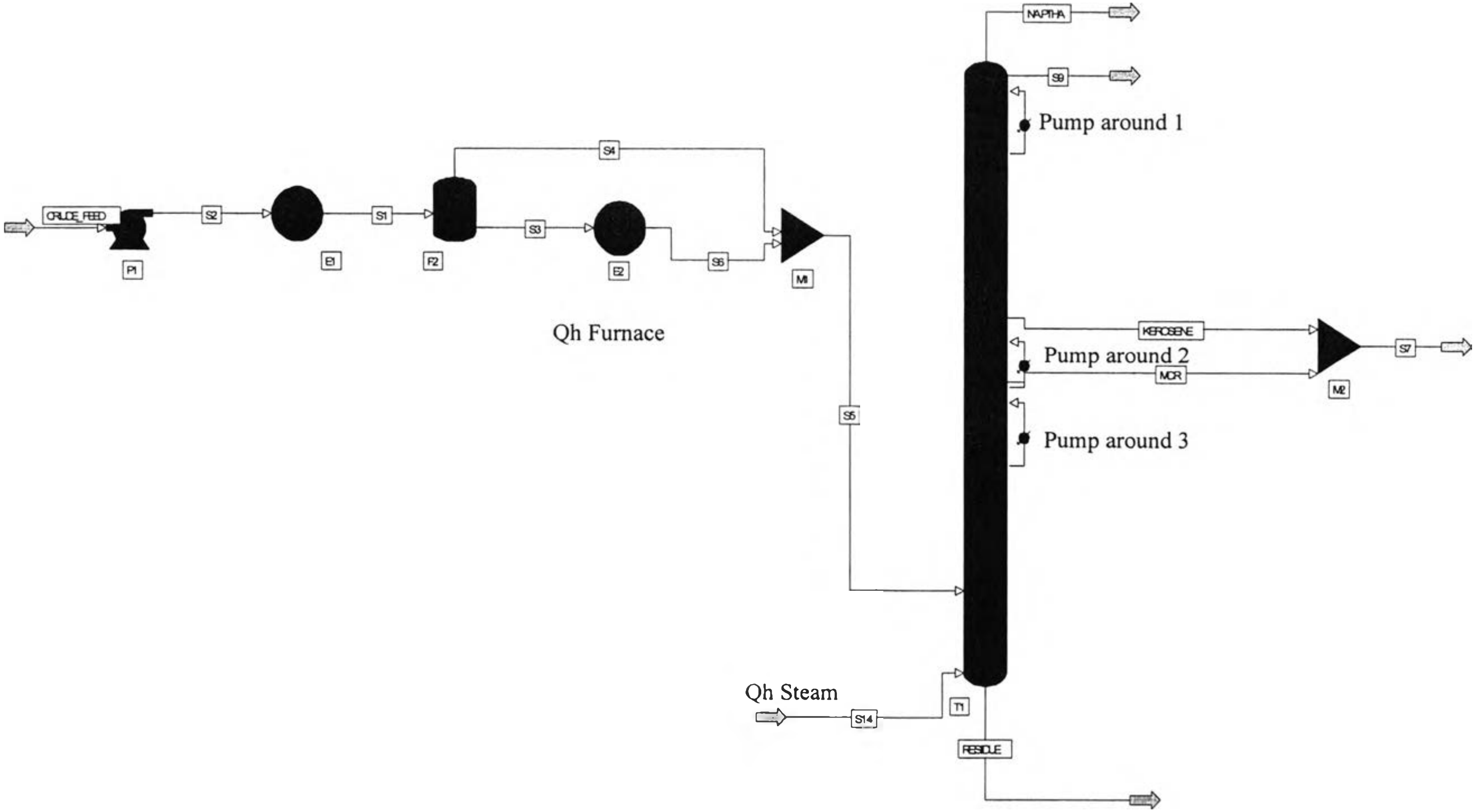


Figure 4.18 The base case refinery with feed of Syrian crude.

Table 4.21 and 4.22 show the simulation details of base case refinery with Light Arabian crude. The detail was used to be parameters in grassroots design by GAMS.

Table 4.21 Stream properties of Syrian crude by PRO II

Stream	Phase	Total Mass Rate (kg / sec)	Temperature (C°)	Total Specific Enthalpy (kW-hr / kg)	Total Cp (kcal/kg-K)
Crude	Liquid	231.9400024	29.9999939	0.017828547	0.42731553
Kerosene	Liquid	42.12986106	170.5210319	0.09681989	0.57434482
MCR	Liquid	35.17029111	246.0209111	0.14449172	0.62634241
Naphtha	Vapor	49.157.99018	113.2485426	0.401626439	0.48290941
Residue	Liquid	126.2004538	373.5807269	0.23485873	0.69961827
S1	Liquid	231.9400024	174.9999939	0.10162183	0.56737629
S2	Liquid	231.9400024	31.4336393	0.018199787	0.42965744
S3	Liquid	231.9400024	166.9999939	0.096375009	0.560389743
S4	N/A	N/A	N/A	N/A	N/A
S5	Mixed	231.9400024	360.0000244	0.241479995	0
S6	Mixed	231.9400024	360.0000244	0.241479996	0
S7	Mixed	77.30015313	204.7280965	0.118509785	0
S9	Water	87.28131875	113.2485426	0.131896361	1.01147181
S14	Vapor	108	475.0000244	0.952164939	0.50549288

Table 4.22 Pump around details of Syrian crude by PRO II

PRO II	GAMS	Tray		Temperature (C°)		RATES	Duty
		From	To	From	To	(K*Kg/sec)	(Mw)
Pump around 1 (PA1)	I1	5	1	116	90.3	0.66	235.22
Pump around 2 (PA2)	I2	20	17	246	157	0.135	29.63
Pump around 3 (PA3)	I3	25	21	317.4	173	0.146	54.77

4.6.3.2 Network from grassroots model of Syrian crude by GAMS

The result network by using grassroots model with Syrian crude was showed in Figure 4.19.

Table 4.23 Stream details list for using in grassroots model

Stream Number	Description
I1	Stream line of pump around 1 (PA1) from Figure 4.14
I2	Stream line of pump around 2 (PA2) from Figure 4.14
I3	Stream line of pump around 3 (PA3) from Figure 4.14
I4	Hot utility
J1	Stream line S2 to S1 from Figure 4.14
J2	Stream line S3 passed S6 and to S5 from Figure 4.14
J3	Stream line S7 to 240°C from Figure 4.14
J4	Cold Utility

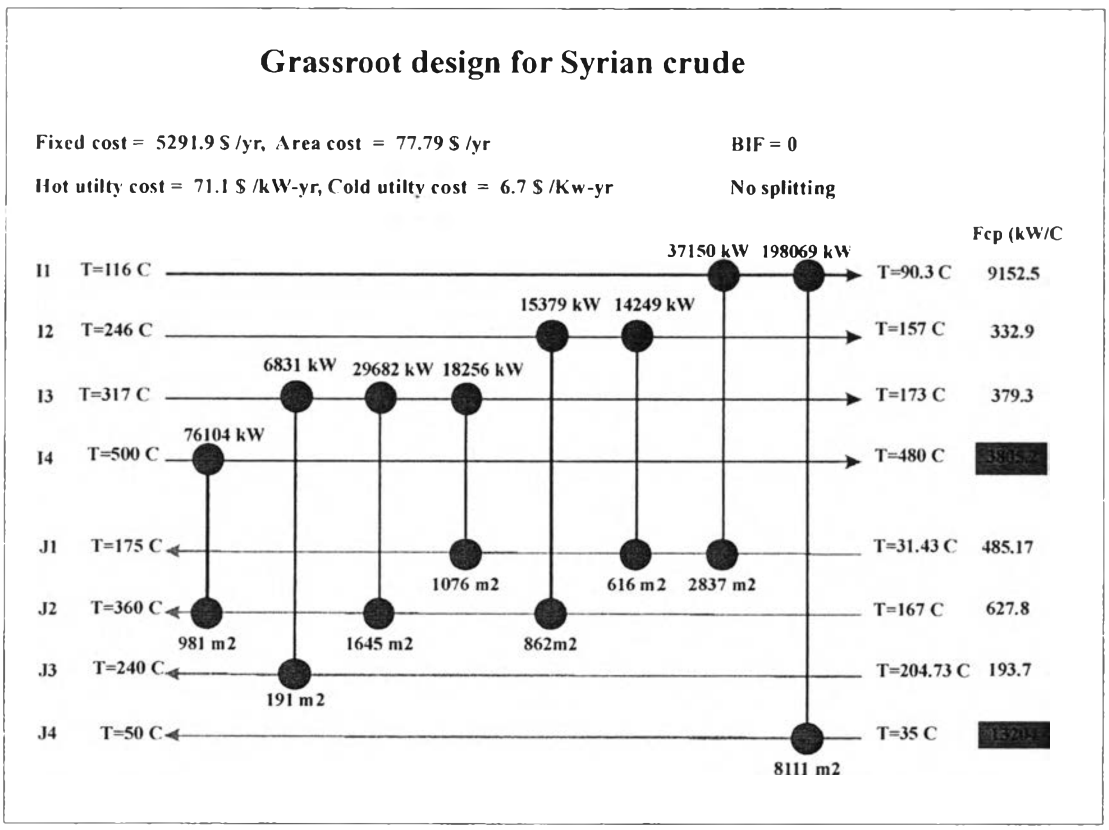
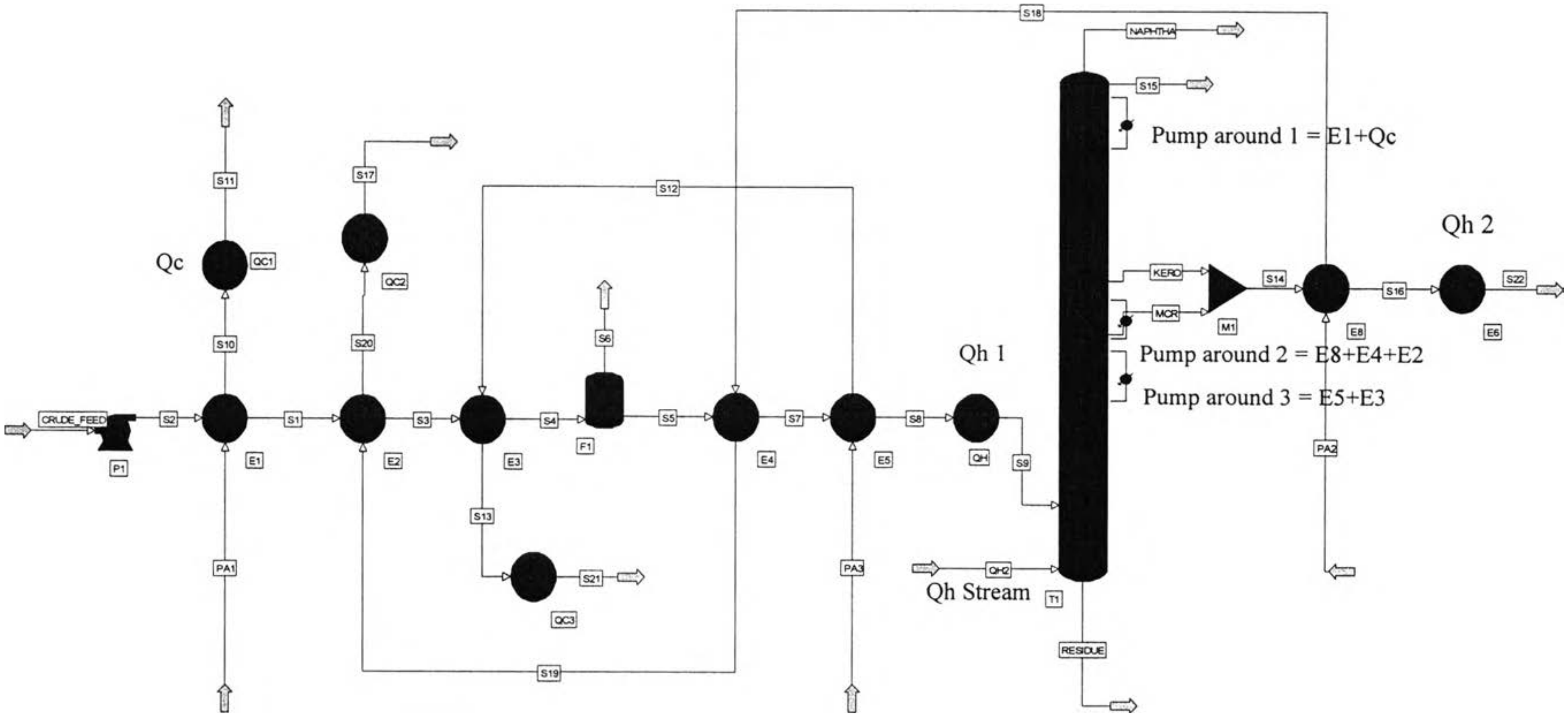


Figure 4.19 The result from grassroots model of Syrian crude by GAMS.

4.6.3.3 Validation of new network by PROII

From Figure 4.19, the simulation network and the details of energy consumption was showed in Figure 4.20.



Qh Furnace = 76.06 MW Qh Steam = 335.92 MW $\sum$Qc = 187.28 MW

Figure 4.20 The new network (HEN3) by PRO II.

The grassroots design of HEN1, HEN2, and HEN3 was simulated in PROII software by changing types of crude oil—Light, Arabian, Rebco, and Syrian crude—and result showed the hot and cold utility consumption of each network in Table 4.24.

Table 4.24 Variation of crude oil type in each heat exchanger network (HEN)

	5% mixed kerosene (C°)	95% naphtha (C°)	Specification GAP≥16.7	Qh Furnace (Mw)	Qh steam (Mw)	Qc (Mw)
HEN1-Arabian	202.36	182	20.36	87.36	137.11	11.41
HEN1-Rebco	216.08	182	34.08	86.21	222.80	77.61
HEN1-Syrian	222.90	182	40.90	82.13	335.92	167.02

	5% mixed kerosene (C°)	95% naphtha (C°)	Specification GAP≥16.7	Qh Furnace (Mw)	Qh steam (Mw)	Qc (Mw)
HEN2-Arabian	202.36	182	20.36	88.91	137.112	10.46
HEN2-Rebco	211.988	182	29.988	87.68	137.112	20.68
HEN2-Syrian	221.90	182	39.90	83.71	335.9238	166.70

	5% mixed kerosene (C°)	95% naphtha (C°)	Specification GAP≥16.7	Qh Furnace (Mw)	Qh steam (Mw)	Qc (Mw)
HEN3-Arabian	202.36	182	20.36	81.28	137.11	16.25
HEN3-Rebco	216.08	182	34.08	80.14	222.80	83.7493
HEN3-Syrian	222.90	182	40.90	76.06	335.92	187.28

4.7 The Optimization to Find the Best Network for Handling Different Crude Types

The concepts for using mathematical programming in optimization of crude uncertainty. The algebra representation of crude oil type was presented in a format following.

Indices:

i = utility of heat exchanger network

j = crude types

Given data:

b_{ij} = cold utility

c_{ij} = hot utility from furnace

d_{ij} = hot utility from steam

CFU = cost of hot utility from furnace, 71.1$/(kW/year)

CST = cost of hot utility from steam, 57.6 $/(kW/year)

CCU = cost of cold utility, 6.7 $/(kW/year)

$x(j)$ = period in operation for crude oil

Decision variable:

z = total cost

w_i = binary variable

Constraints:

Constraint for choosing HEN i: $\Sigma\, w_i = 1$

Objective function: $z = \Sigma\, ((CFU^* c_{ij}) + (CST^* d_{ij}) + (CCU^* b_{ij}))$

4.7.1 Example of master plan in operation in crude refinery

Table 4.25 showed the operation plan for 5 years. The operation period in 5 years was divided into various months to study the effect of period time in operation.

Table 4.25 The 5 years master plan in many scenarios of operation

Crude	Time (month)	Hot utility from furnace cost ($)	Hot utility stream cost ($)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	20	126,438	157,953	1,402	1,099,565	HEN2
Rebco	20	124,687	157,953	2,772		
Syrian	20	119,037	386,984	22,338		

Crude	Time (month)	Utility from furnace cost ($)	Utility stream cost ($)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	30	189,656	236,929	2,104	1,039,019	HEN2
Rebco	15	93,515	118,465	2,079		
Syrian	15	89,278	290,238	16,753		

Crude	Time (month)	Utility from furnace cost ($)	Utility stream cost ($)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	15	94,828	118,465	1,052	1,038,733	HEN2
Rebco	30	187,031	236,930	4,158		
Syrian	15	89,278	290,238	16,754		

Crude	Time (month)	Utility from furnace cost ($)	Utility stream cost ($)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	15	94,828	118,465	1,052	1,220,944	HEN2
Rebco	15	93,516	118,465	2,079		
Syrian	30	178,556	580,476	33,507		

Table 4.25 (Continued)

Crude	Time (month)	Utility from furnace cost ($)	Utility stream cost $)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	40	252,875	315,906	2,805	978,472	HEN2
Rebco	10	62,344	78,976	1,386		
Syrian	10	59,519	193,492	11,169		

Crude	Time (month)	Utility from furnace cost ($)	Utility stream cost $)	Cold utility cost ($)	Total cost ($)	The best HEN
Light Arabian	10	63,219	78,976	701	1,342,322	HEN2
Rebco	10	62,344	78,976	1,386		
Syrian	40	238,075	773,968	44,676		

The optimization model was applied in 5 years operation plan. The data from simulation was used as the parameter in optimization model. From the result in Table 4.25, network HEN 2 was suitable to handle the three types of crude which the lowest utilities cost.