CHAPTER IV
RESULTS AND DISCUSSION

4.1 Targeting

The properties of the light crude, intermediate crude, and heavy crude are shown in Tables 4.1-4.3. Table 4.4 shows the specifications of the products. The withdrawn product locations are determined according to Watskin design, and the results are shown in Table 4.5.

Table 4.1 Feedstock used for the design

Crude	Density (kg/m^3)	Throughput (m^3/h)
Light Crude	845 (36.0 API)	795
Intermediate Crude	889 (27.7 API)	795
Heavy Crude	934 (20.0 API)	795

Table 4.2 TBP data

Vol %	Temperature (°C)		
	Light Crude	Intermediate Crude	Heavy Crude
5	45	94	133
10	82	131	237
30	186	265	344
50	281	380	482
70	382	506	640
90	552	670	N/A

Table 4.3 Light-End composition of crude

Compound	Vol %		
	Light Crude	Intermediate Crude	Heavy Crude
Ethane	0.13	0.1	0
Propane	0.78	0.3	0.04
Isobutane	0.49	0.2	0.04
n-butane	1.36	0.7	0.11
Isopentane	1.05	0	0.14
n-pentane	1.30	0	0.16
Total	5.11	1.3	0.48

Table 4.4 Product specifications and withdrawal tray

Product	Specification	Withdrawal Tray
Naphtha	D86 (95% point) = 182 °C	1
Kerosene	D86 (95% point) = 271 °C	9
Diesel	D86 (95% point) = 327 °C	16
AGO	D86 (95% point) = 377-410 °C	25
Overflash rate	0.03	
Kerosene-Naphtha	(5-95) gap ≥ 16.7 °C	
Diesel-Kerosene	(5-95) gap ≥ 0 °C	
AGO-Diesel	(5-95) gap ≥ -5.6 °C to -11 °C	
Feed Tray		29
Total Trays		34

Table 4.5 Tray requirements to Watskin design

Separation	Number of Trays
Light Naphtha to Heavy Naphtha	6-8
Heavy Naphtha to Light Distillate	6-8
Light Distillate to Heavy Distillate	4-6
Heavy Distillate to AGO	4-6
Flash Zone to First Draw Tray	3-4
Steam Stripping Sections	4

There are 34 trays in the main column and 4 trays in each stripper. The flow rates of stripping steam streams are estimated and adjusted to 10 lb per barrel of product, as suggested by Watskin. The total energy consumption (E) is evaluated by using the following expression:

$$E = U + 0.7\sum H_i^s$$

where U is the minimum heating utility obtained by using straight pinch analysis and $\sum H_i^s$ is the summation of energy flow of all steam streams. Because low-pressure steam is cheaper than fuel gas with the same amount of heat content, a weight factor of 0.7 is used for the steam. The total energy consumption is used as an objective function.

4.1.1 Light Crude

Simulation results for the initial scheme with no pump-around circuits are shown in Table 5.6. Note the product gaps are well above the specifications.

Table 4.6 Results of No Pump-Around Circuit Scheme for Light Crude

Product	No Pump-Around
Naphtha Flow Rate	248.73 m^3/h
Kerosene Flow Rate	144.29 m^3/h
Diesel Flow Rate	71.93 m^3/h
AGO Flow Rate	123.79 m^3/h
Residue Flow Rate	206.65 m^3/h
Kerosene Stripping Steam Ratio	9.98
Diesel Stripping Steam Ratio	9.73
AGO Stripping Steam Ratio	9.77
Residue Stripping Steam Ratio	10.21
(5-95) Kerosene-Naphtha Gap	25.50 °C
(5-95) Diesel-Kerosene Gap	5.55 °C
(5-95) AGO-Diesel Gap	2.06 °C
Kerosene Withdrawal Temperature	205.73 °C
Diesel Withdrawal Temperature	273.84 °C
AGO Withdrawal Temperature	322.69 °C
Residue Withdrawal Temperature	350.56 °C
Condenser Duty	105 MW
Condenser Temperature Range	155.25-43.3 °C
Flash Zone Temperature	359.70 °C
Energy Consumption (*E*)	100.04 MW

The heat demand-supply diagram corresponding to solution in Table 5.6 is shown in Figure 4.1. There is a huge heat surplus in the condenser region, which results in a large cooling utility. Meanwhile, a large heat deficit exists above 155.25 °C. As the total heat supply is almost constant, the way toward energy savings is to change the heat supply profile. That is, instead of supplying all heat at a low temperature, some heat can be supplied at a higher temperature where the heat demand is larger than the heat supply. In other words, transfer some heat from the condenser to a pump-around circuit.

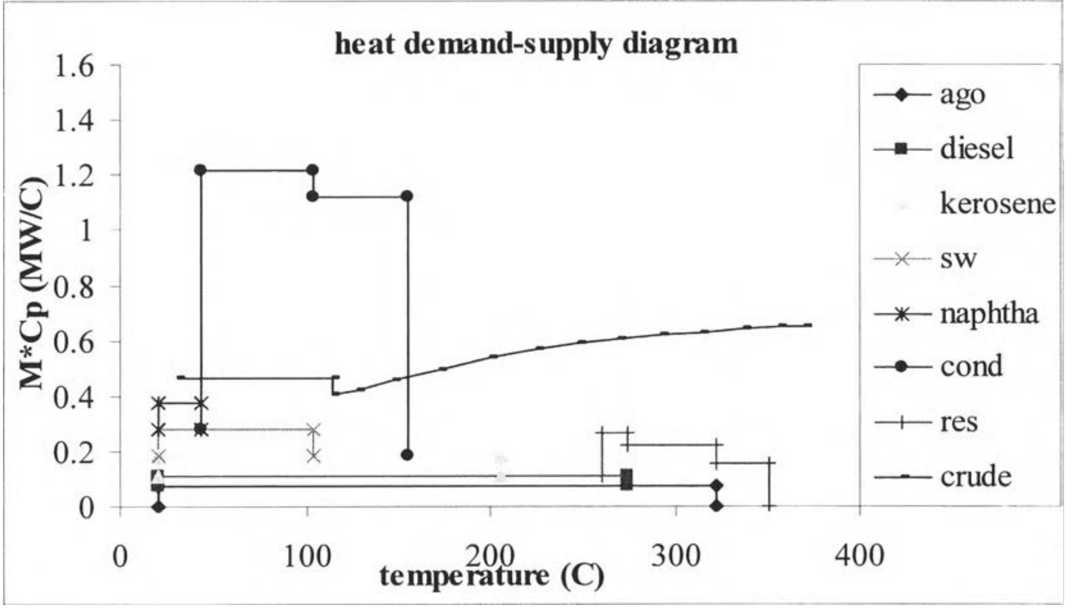

Figure 4.1 Heat demand-supply diagram for light crude distillation without pump-around circuits.

4.1.1.1 One Pump-Around Circuit

The first pump-around has to be above the kerosene withdrawal tray because the heat that can be transferred from the condenser will be the maximum when a pump-around is above all of the side-withdrawal product lines. The pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C.

The duty of the first pump-around (PA1) is increased steadily, and the product gaps are observed in each simulation. The kerosene-naphtha gap decreases when the PA1 duty increases but remains well above the specification, while the other gaps are almost unaffected. The heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on the top tray. Therefore, the limit of the heat shifting has been accomplished. The duty of 64 MW represents the total amount of heat one could obtain from all pump-around circuits. The following steps consist of distributing this amount of heat properly among several pump-around circuits. The main operation variables of the scheme with one pump-around are shown in Table 4.7.

Table 4.7 Results of one pump-around circuit scheme for light crude

Product	One Pump-Around
Naphtha Flow Rate	246.82 m^3/h
Kerosene Flow Rate	146.45 m^3/h
Diesel Flow Rate	71.62 m^3/h
AGO Flow Rate	124.28 m^3/h
Residue Flow Rate	206.28 m^3/h
Kerosene Stripping Steam Ratio	9.83
Diesel Stripping Steam Ratio	9.77
AGO Stripping Steam Ratio	9.73
Residue Stripping Steam Ratio	10.23
(5-95) Kerosene-Naphtha Gap	23.39 °C
(5-95) Diesel-Kerosene Gap	5.71 °C
(5-95) AGO-Diesel Gap	1.99 °C
Kerosene Withdrawal Temperature	204.45 °C
Diesel Withdrawal Temperature	274.00 °C
AGO Withdrawal Temperature	322.90 °C
Residue Withdrawal Temperature	350.66 °C
Condenser Duty	41.2 MW
Condenser Temperature Range	146.26-43.33 °C
Pump-Around 1 Duty	64 MW
Pump-Around 1 Temperature Range	179.54-104.44 °C
Flash Zone Temperature	359.73 °C
Energy Consumption (E)	90.38 MW

From the results, the major conclusions are as follows: the total energy consumption (*E*) decreases by 9.66 MW compared to the no pump-around scheme. The kerosene-naphtha gap is reduced from 25.50 to 23.39 °C, remaining well above the specification of 16.7 °C and the yield of naphtha decreases but the yield of kerosene increases because some light components of the vapor are absorbed by the cold pump-around stream and carried to the kerosene withdrawal tray. Note that the total yield of the two products remains constant.

The resulting heat demand-supply diagram is shown in Figure 4.2, the heat surplus in the condenser region is greatly reduced, but it is still larger than the demand. Nevertheless, it is impossible to shift more heat from the condenser to PA1. To reduce the heat surplus in the region of PA1, a second pump-around is installed at the position indicated in Figure 4.2.

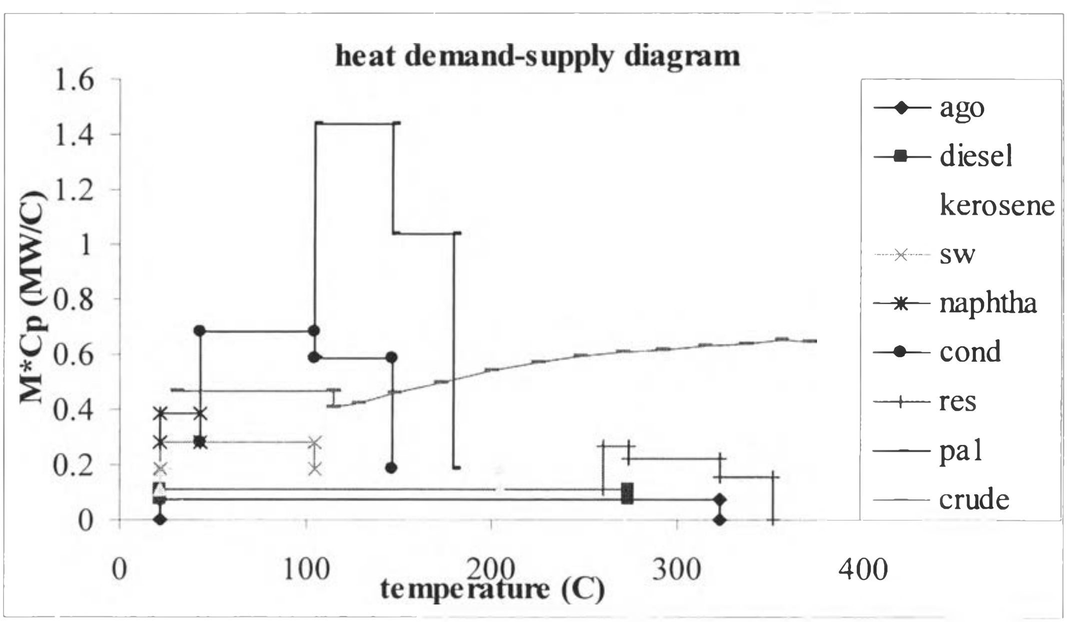

Figure 4.2 Heat demand-supply diagram for light crude distillation with a top pump-around.

4.1.1.2 Two Pump-Around Circuits

The second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is chosen to be around equal to the withdrawal temperature of PA1 that is 171.11 °C. With the increase in the PA2 duty, the kerosene-naphtha gap decreases quickly.

Table 4.8 Results of two pump-around circuits scheme for light crude

Product	Two Pump-Arounds
Naphtha Flow Rate	242.96 m^3/h
Kerosene Flow Rate	146.93 m^3/h
Diesel Flow Rate	75.74 m^3/h
AGO Flow Rate	123.58 m^3/h
Residue Flow Rate	206.28 m^3/h
Kerosene Stripping Steam Ratio	15.93
Diesel Stripping Steam Ratio	9.45
AGO Stripping Steam Ratio	9.79
Residue Stripping Steam Ratio	10.23
(5-95) Kerosene-Naphtha Gap	16.70 °C
(5-95) Diesel-Kerosene Gap	1.62 °C
(5-95) AGO-Diesel Gap	2.41 °C
Kerosene Withdrawal Temperature	184.44 °C
Diesel Withdrawal Temperature	271.95 °C
AGO Withdrawal Temperature	323.02 °C
Residue Withdrawal Temperature	350.66 °C
Condenser Duty	44.6 MW
Condenser Temperature Range	144.56-43.33 °C
Pump-Around 1 Duty	21.00 MW
Pump-Around 1 Temperature Range	170.86-104.44 °C
Pump-Around 2 Duty	43.00 MW
Pump-Around 2 Temperature Range	259.14-171.11 °C
Flash Zone Temperature	359.73 °C
Energy Consumption (*E*)	56.24 MW

When the duty of PA2 is larger than 36.10 MW, the kerosene-naphtha gap does not satisfy the specification. To recover this gap, one could increase the stripping steam flow rate or increase the number of trays in the naphtha-kerosene section. However, the number of trays keeps constant; therefore, the kerosene and diesel stripping steam flow rates are adjusted with a controller in which the gap specifications are defined. Heat shifting continues until the liquid reflux at the kerosene withdrawal tray is small and/or the kerosene-naphtha gap can not be recovered even by increased amounts of stripping steam. The main operation variables of the scheme with one pump-around are shown in Table 4.8 and should be compared with the Table 4.7.

From the results, the main changes from one pump-around to two pump-around circuits are: the net energy consumption decreases by 34.14 MW. The flow rate of the kerosene stripping steam is nearly doubled. The large amount of extra steam is used to strip a significant amount of light components in the kerosene withdrawal stream. The top section of the column becomes less hot because of the increased stripping steam. The kerosene withdrawal temperature drops by 20.01 °C. And the yield of diesel increases while the yield of naphtha decreases.

The heat demand-supply diagram is shown in Figure 4.3, the pinch temperature increases to the value of the PA2 withdrawal temperature. The heat surplus in the region of PA1 is still high, but further shifting would cost too much steam to be beneficial. Consequently, this remaining heat surplus is useless.

At this time, the only heat surplus transferable is located in the PA2 circuit, to make use of heat surplus, the adding a third pump-around circuit is necessary.

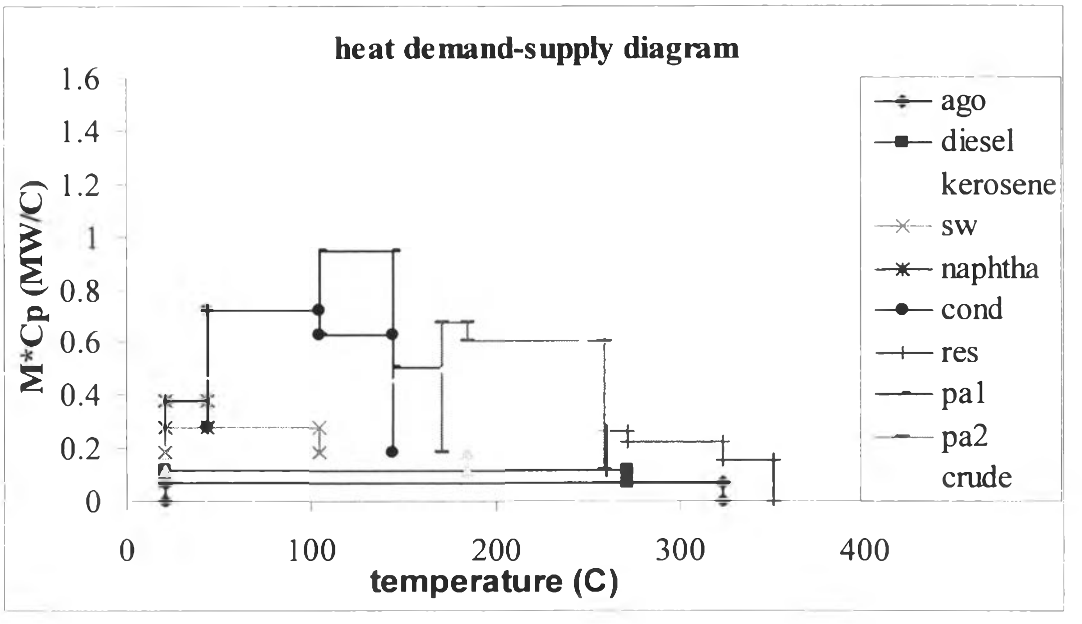

Figure 4.3 Heat demand-supply diagram for light crude distillation with two pump-around circuits.

4.1.1.3 Three Pump-Around Circuits

The third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. Heat is shifted gradually from PA2 to PA3, with the gap maintained above the specification. The effect of the duty of PA3 on energy consumption is shown in Table 4.9. A summary of all of the variables is shown in Table 4.10.

Table 4.9 Effect of the duty of PA3 on energy consumption for light crude

PA3 duty (MW)	Energy consumption (MW)
4.00	55.79
5.00	55.77
6.00	55.91
7.00	55.99
8.00	56.01
9.00	56.16

Table 4.10 Results of three pump-around circuits scheme for light crude

Product	Three Pump-Arounds
Naphtha Flow Rate	242.99 m^3/h
Kerosene Flow Rate	146.08 m^3/h
Diesel Flow Rate	75.84 m^3/h
AGO Flow Rate	124.29 m^3/h
Residue Flow Rate	206.28 m^3/h
Kerosene Stripping Steam Ratio	16.01
Diesel Stripping Steam Ratio	9.23
AGO Stripping Steam Ratio	9.74
Residue Stripping Steam Ratio	10.23
(5-95) Kerosene-Naphtha Gap	16.70 °C
(5-95) Diesel-Kerosene Gap	0.00 °C
(5-95) AGO-Diesel Gap	1.67 °C
Kerosene Withdrawal Temperature	184.34 °C
Diesel Withdrawal Temperature	270.10 °C
AGO Withdrawal Temperature	322.76°C
Residue Withdrawal Temperature	350.66 °C
Condenser Duty	44.60 MW
Condenser Temperature Range	144.62-43.33 °C
Pump-Around 1 Duty	21.00 MW
Pump-Around 1 Temperature Range	170.93-104.44 °C
Pump-Around 2 Duty	38 MW
Pump-Around 2 Temperature Range	257.70-171.11 °C
Pump-Around 3 Duty	5.00 MW
Pump-Around 3 Temperature Range	314.05-232.22 °C
Flash Zone Temperature	359.73 °C
Energy Consumption (*E*)	55.77 MW

At the beginning, the energy consumption decreases by the increase in the duty of PA3. However, when the PA3 duty exceeds 5 MW, the energy consumption stays constant in a rather wide range (Table 4.9). This is because little heat surplus exists in the region of PA2. Thus, more heat shifting makes no difference. Beyond this stable range, increase heat shifting to PA3 results in an increase in energy consumption because of the additional steam consumption outweighs the gain in energy recovery for maintaining the gap. Clearly, 5 MW is the right point at which stop. This effect can not be arrested with other design procedures.

Figure 4.4 is the heat demand-supply diagram. The heat surplus in the region of PA2 has been moved to the PA3, which accounts for the decrease in energy consumption.

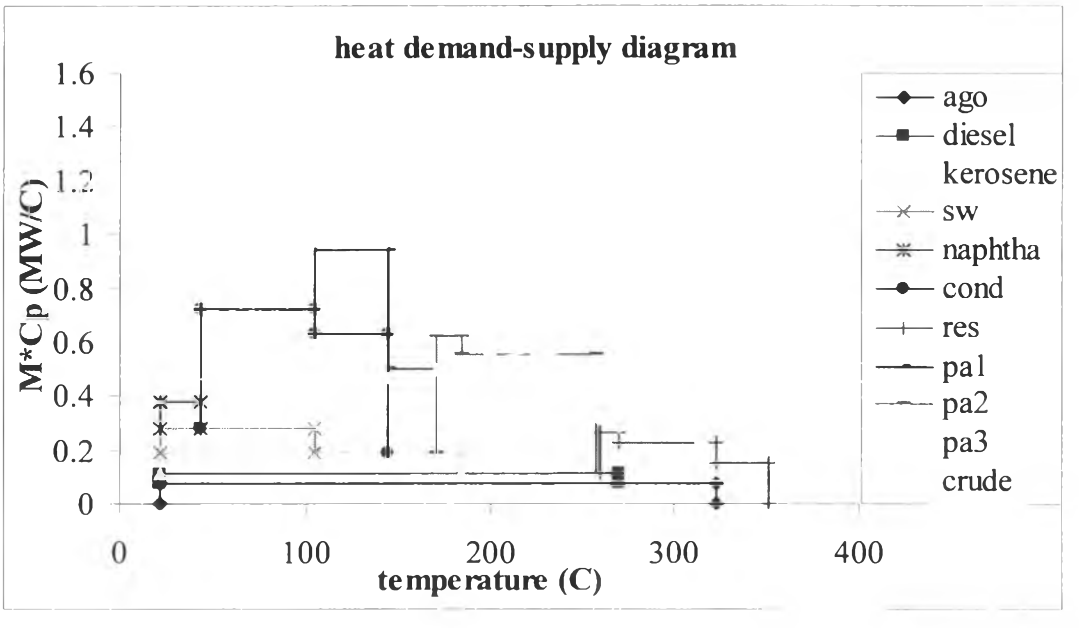

Figure 4.4 Heat demand-supply diagram for light crude distillation with three pump-around circuits.

At this point, the best scheme for the light crude has been reached. Next, the same analysis is performed for intermediate and heavy crude.

4.1.2 Intermediate Crude

Simulation results for the initial scheme with no pump-around circuits are shown in Table 4.11. Note the product gaps are well above the specifications.

Table 4.11 Results of no pump-around circuit scheme for intermediate crude

Product	No Pump-Around
Naphtha Flow Rate	155.08 m^3/h
Kerosene Flow Rate	96.26 m^3/h
Diesel Flow Rate	73.76 m^3/h
AGO Flow Rate	61.27 m^3/h
Residue Flow Rate	409.00 m^3/h
Kerosene Stripping Steam Ratio	9.86
Diesel Stripping Steam Ratio	8.23
AGO Stripping Steam Ratio	9.57
Residue Stripping Steam	10.43
(5-95) Kerosene-Naphtha Gap	23.76 °C
(5-95) Diesel-Kerosene Gap	4.89 °C
(5-95) AGO-Diesel Gap	-8.26 °C
Kerosene Withdrawal Temperature	204.51 °C
Diesel Withdrawal Temperature	275.79 °C
AGO Withdrawal Temperature	314.45 °C
Residue Withdrawal Temperature	351.26 °C
Condenser Duty	76.50 MW
Condenser Temperature Range	158.81-43.33 °C
Flash Zone Temperature	359.27 °C
Energy Consumption (*E*)	88.61 MW

The heat demand-supply diagram corresponding to the solution in Table 4.11 is shown in Figure 4.5. There is a huge heat surplus in the condenser region, which results in a large cooling utility. Meanwhile, a large deficit exists above 158.81 °C. As the total heat supply is almost constant, the way toward energy savings is to change the heat supply profile. That is, in place of supplying all heat at a low temperature, some heat can be supplied at a higher temperature where the heat demand is larger than the heat supply. In the other word, transfer some heat from the condenser to a pump-around circuit.

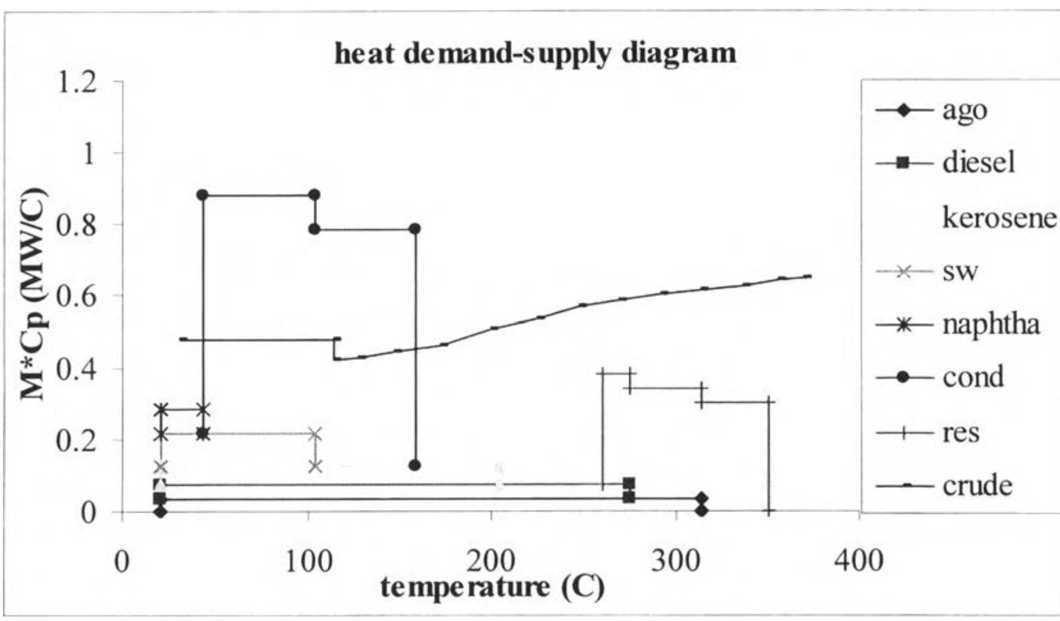

Figure 4.5 Heat demand-supply diagram for intermediate crude distillation without pump-around circuits.

4.1.2.1 One Pump-Around Circuit

The pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C.

The duty of the first pump-around (PA1) is increased steadily, and the product gaps are observed in each simulation. The kerosene-naphtha gap decreases when the PA1 duty increases but remains well above the specification, while the other gaps are almost unchanged. The heat shift continues without violating the gap specifications until heat shift would result in liquid drying up on the top tray. Therefore, the limit of the heat shifting has been reached. The duty of 40

MW represents the total amount of heat one could obtain from all pump-around circuits. The following steps consist of distributing this amount of heat properly among several pump-around circuits. The main operation variables of the scheme with one pump-around are shown in Table 4.12.

Table 4.12 Results of one pump-around circuit scheme for intermediate crude

Product	One Pump-Around
Naphtha Flow Rate	153.62 m^3/h
Kerosene Flow Rate	97.97 m^3/h
Diesel Flow Rate	73.47 m^3/h
AGO Flow Rate	61.32 m^3/h
Residue Flow Rate	409.00 m^3/h
Kerosene Stripping Steam Ratio	9.69
Diesel Stripping Steam Ratio	8.26
AGO Stripping Steam Ratio	9.56
Residue Stripping Steam Ratio	10.43
(5-95) Kerosene-Naphtha Gap	21.33 °C
(5-95) Diesel-Kerosene Gap	5.05 °C
(5-95) AGO-Diesel Gap	-8.29 °C
Kerosene Withdrawal Temperature	203.00 °C
Diesel Withdrawal Temperature	275.82 °C
AGO Withdrawal Temperature	314.45 °C
Residue Withdrawal Temperature	351.26 °C
Condenser Duty	36.50 MW
Condenser Temperature Range	147.61-43.33 °C
Pump-Around 1 Duty	40 MW
Pump-Around 1 Temperature Range	179.40-104.44 °C
Flash Zone Temperature	359.27 °C
Energy Consumption (E)	79.76 MW

From the results, the total energy consumption (E) decreases by 8.85 MW compared to the no pump-around scheme. The kerosene-naphtha gap is reduced from 23.76 to 21.33°C, remaining well above the specification of 16.7 °C and the yield of naphtha decreases but the yield of kerosene increases because some light components of the vapor are absorbed by the cold pump-around stream and carried to the kerosene withdrawal tray. Note that the total yield of the two products remains constant.

The resulting heat demand-supply diagram is shown in Figure 4.6, the heat surplus in the condenser region is greatly reduced, but it is still larger than the demand. However, to reduce the heat surplus in the region of PA1, a second pump-around is added.

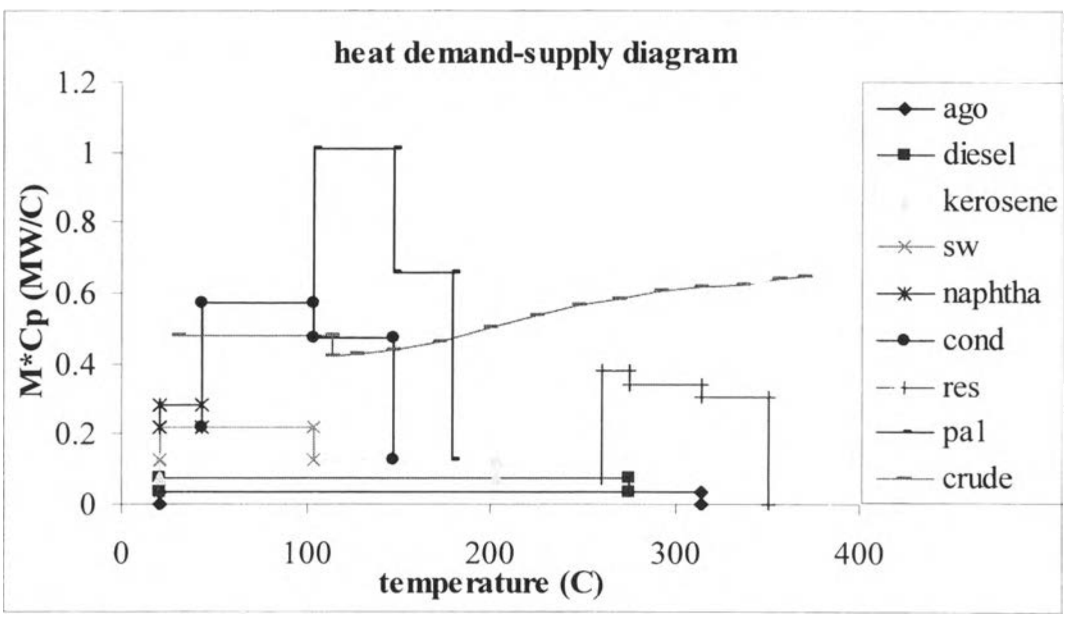

Figure 4.6 Heat demand-supply diagram for intermediate crude distillation with a top pump-around.

4.1.2.2 Two Pump-Around Circuits

The second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is 171.11 °C. With the increase in the PA2 duty, the kerosene-naphtha gap decreases quickly. Table 4.13 shows the variable of the scheme with two pump-arounds. Heat shifting continues until the liquid reflux at the kerosene withdrawal tray is small.

Table 4.13 Results of two pump-around circuits scheme for intermediate crude

Product	Two Pump-Arounds
Naphtha Flow Rate	152.50 m^3/h
Kerosene Flow Rate	96.53 m^3/h
Diesel Flow Rate	76.55 m^3/h
AGO Flow Rate	61.07 m^3/h
Residue Flow Rate	408.78 m^3/h
Kerosene Stripping Steam Ratio	9.83
Diesel Stripping Steam Ratio	7.93
AGO Stripping Steam Ratio	9.60
Residue Stripping Steam Ratio	10.44
(5-95) Kerosene-Naphtha Gap	18.22 °C
(5-95) Diesel-Kerosene Gap	2.17 °C
(5-95) AGO-Diesel Gap	-8.00 °C
Kerosene Withdrawal Temperature	198.30 °C
Diesel Withdrawal Temperature	274.83 °C
AGO Withdrawal Temperature	314.72 °C
Residue Withdrawal Temperature	351.36 °C
Condenser Duty	36.80 MW
Condenser Temperature Range	147.32-43.33 °C
Pump-Around 1 Duty	25.00 MW
Pump-Around 1 Temperature Range	176.31-104.44 °C
Pump-Around 2 Duty	15.00 MW
Pump-Around 2 Temperature Range	265.44-171.11 °C
Flash Zone Temperature	359.31 °C
Energy Consumption (*E*)	68.74 MW

From the results, the net energy consumption decreases by 11.02 MW and the yield of diesel increases while the yield of naphtha decreases.

The heat demand-supply diagram is shown in Figure 4.7, the heat surplus in the region of PA1 is still high, but further shifting would cost too much steam to be beneficial. Therefore, this remaining heat surplus is useless.

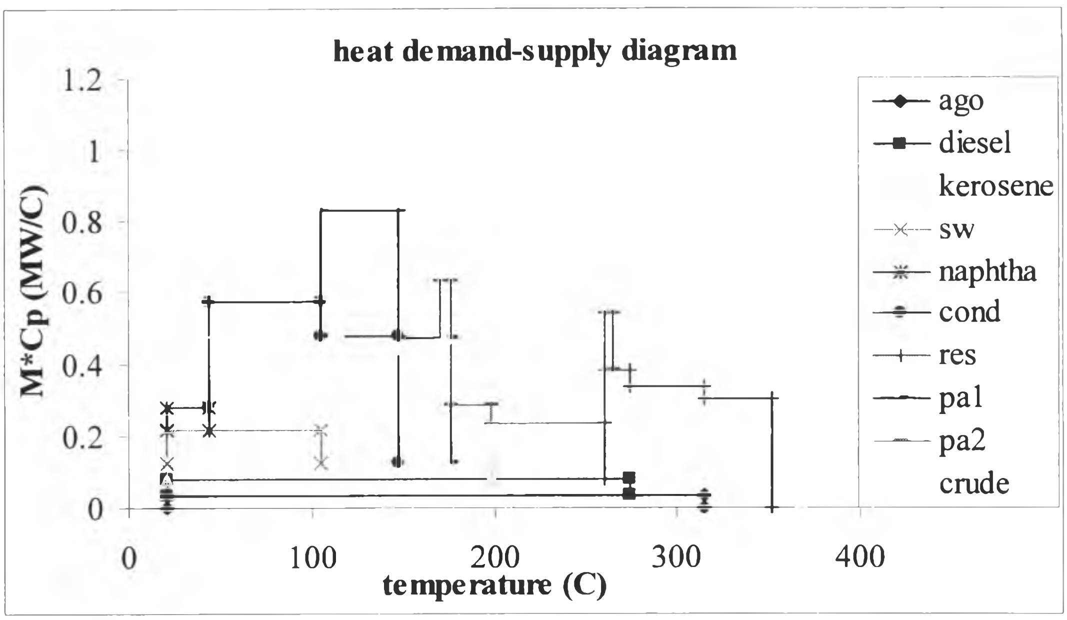

Figure 4.7 Heat demand-supply diagram for intermediate crude distillation with two pump-around circuits.

Now, the only heat surplus transferable is located in the PA2 circuit, to make use of heat surplus by adding a third pump-around circuit.

4.1.2.3 Three Pump-Around Circuits

The third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. Heat is shifted gradually from PA2 to PA3. The effect of the duty of PA3 on energy consumption is shown in Table 4.14. A summary of all of the variables is shown in Table 4.15.

Table 4.14 Effect of the duty of PA3 on energy consumption for intermediate crude

PA3 duty (MW)	Energy consumption (MW)
1	68.73
2	68.70
3	68.77
4	68.78
5	68.79
6	68.88
7	68.98
8	69.11
9	69.27

Table 4.15 Results of three pump-around circuits scheme for intermediate crude

Product	Three Pump-Arounds
Naphtha Flow Rate	152.51 m^3/h
Kerosene Flow Rate	96.29 m^3/h
Diesel Flow Rate	76.38 m^3/h
AGO Flow Rate	61.48 m^3/h
Residue Flow Rate	408.78 m^3/h
Kerosene Stripping Steam Ratio	9.85
Diesel Stripping Steam Ratio	7.94
AGO Stripping Steam Ratio	9.54
Residue Stripping Steam Ratio	10.44
(5-95) Kerosene-Naphtha Gap	18.24 °C
(5-95) Diesel-Kerosene Gap	1.50 °C
(5-95) AGO-Diesel Gap	-8.44 °C
Kerosene Withdrawal Temperature	198.26 °C
Diesel Withdrawal Temperature	273.59 °C
AGO Withdrawal Temperature	314.58 °C
Residue Withdrawal Temperature	351.36 °C
Condenser Duty	36.90 MW
Condenser Temperature Range	147.35-43.33 °C
Pump-Around 1 Duty	25.00 MW
Pump-Around 1 Temperature Range	176.35-104.44 °C
Pump-Around 2 Duty	13.00 MW
Pump-Around 2 Temperature Range	264.83 -171.11°C
Pump-Around 3 Duty	2.00 MW
Pump-Around 3 Temperature Range	312.96-232.22 °C
Flash Zone Temperature	359.31 °C
Energy Consumption (*E*)	68.70 MW

At the beginning, the energy consumption decreases by the increase in the duty of PA3. However, when the PA3 duty exceeds 2 MW, the energy consumption stays constant in a rather wide range (Table 4.14). This is because little heat surplus exists in the region of PA2. Thus, more heat shifting makes no difference. Beyond this stable range, increase heat shifting to PA3 results in an increase in energy consumption. Clearly, 2 MW is the right point at which stop. This effect can not be arrested with other design procedures.

Figure 4.8 is the heat demand-supply diagram. The heat surplus in the region of PA2 has been moved to the PA3, which accounts for the decrease in energy consumption.

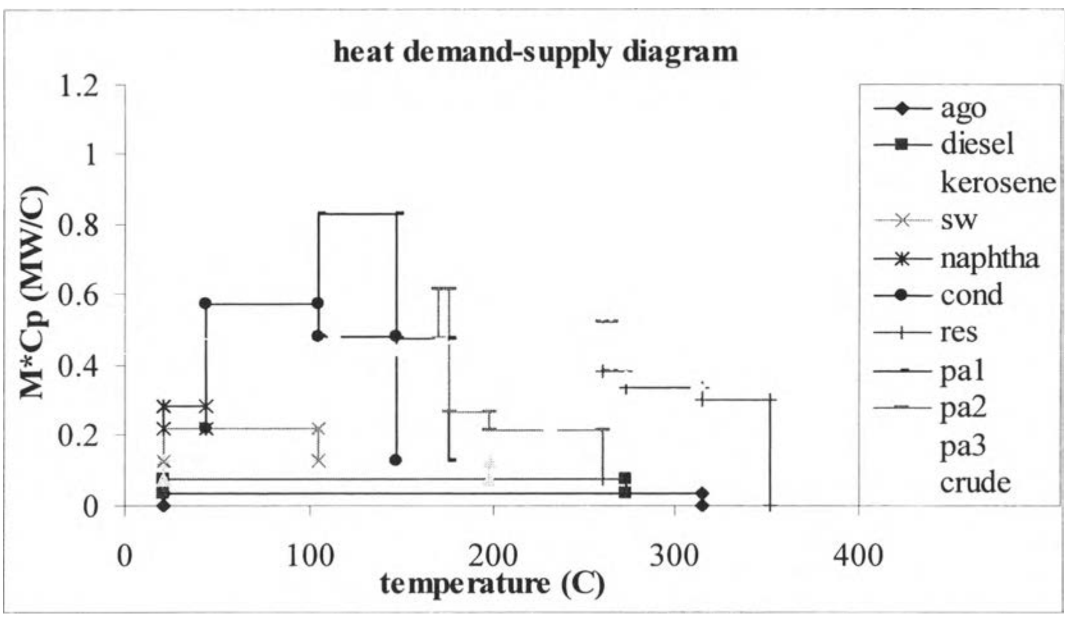

Figure 4.8 Heat demand-supply diagram for intermediate crude distillation with three pump-around circuits.

4.1.3 Heavy Crude

The total energy consumption and the operation variables for a scheme with three pump-around circuits are shown in Table 4.16. The heat demand-supply diagram is shown in Figure 4.9.

Table 4.16 Results of no pump-around circuits scheme for heavy crude

Product	No Pump-Around
Naphtha Flow Rate	56.47 m^3/h
Kerosene Flow Rate	51.47 m^3/h
Diesel Flow Rate	92.51 m^3/h
AGO Flow Rate	62.45 m^3/h
Residue Flow Rate	532.51 m^3/h
Kerosene Stripping Steam Ratio	9.48
Diesel Stripping Steam Ratio	6.46
AGO Stripping Steam Ratio	13.32
Residue Stripping Steam Ratio	11.38
(5-95) Kerosene-Naphtha Gap	34.23 °C
(5-95) Diesel-Kerosene Gap	8.55 °C
(5-95) AGO-Diesel Gap	-11.00 °C
Kerosene Withdrawal Temperature	212.92 °C
Diesel Withdrawal Temperature	284.27 °C
AGO Withdrawal Temperature	306.57 °C
Residue Withdrawal Temperature	348.80 °C
Condenser Duty	54.40 MW
Condenser Temperature Range	135.35-21.11 °C
Flash Zone Temperature	358.58 °C
Energy Consumption (E)	90.44 MW

The heat demand-supply diagram corresponding to the solution in Table 4.16 is shown in Figure 4.9. There is a huge heat surplus in the condenser region, which results in a large cooling utility. Meanwhile, a deficit exists above 135.35 °C. As the total heat supply is almost constant, the way toward energy savings is to change the heat supply profile. That is, in place of supplying all heat at a low temperature, some heat can be supplied at a higher temperature where the heat demand is larger than the heat supply. In the other word, transfer some heat from the condenser to a pump-around circuit.

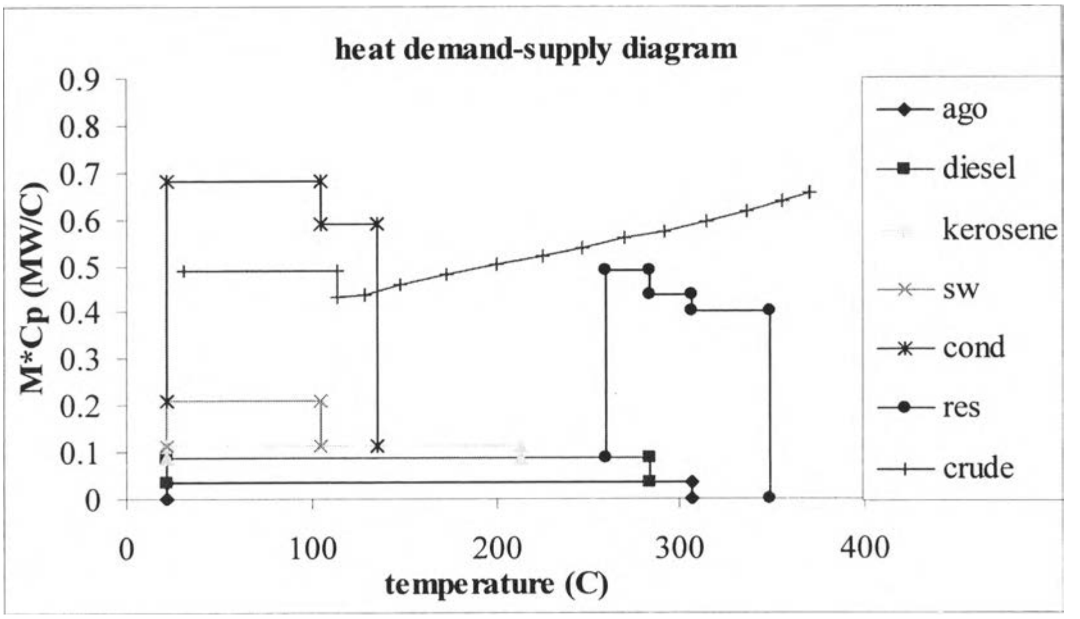

Figure 4.9 Heat demand-supply diagram for heavy crude distillation without pump-around circuits.

4.1.3.1 One Pump-Around Circuit

The pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C.

The duty of the first pump-around (PA1) is increased steadily, and the product gaps are observed in each simulation. The kerosene-naphtha gap decreases when the PA1 duty increases but remains well above the specification, while the other gaps are almost unchanged. The heat shift continues without violating the gap specifications until heat shift would result in liquid drying up on the top tray. Therefore, the limit of the heat shifting has been reached. The duty of 28 MW represents the total amount of heat one could obtain from all pump-around circuits. The following steps consist of distributing this amount of heat properly among several pump-around circuits. The main operation variables of the scheme with one pump-around are shown in Table 4.17.

Table 4.17 Results of one pump-around circuit scheme for heavy crude

Product	One Pump-Around
Naphtha Flow Rate	56.08 m^3/h
Kerosene Flow Rate	51.96 m^3/h
Diesel Flow Rate	92.40 m^3/h
AGO Flow Rate	62.48 m^3/h
Residue Flow Rate	532.51 m^3/h
Kerosene Stripping Steam Ratio	9.39
Diesel Stripping Steam Ratio	6.47
AGO Stripping Steam Ratio	13.32
Residue Stripping Steam Ratio	11.38
(5-95) Kerosene-Naphtha Gap	32.12 °C
(5-95) Diesel-Kerosene Gap	8.63 °C
(5-95) AGO-Diesel Gap	-11.00 °C
Kerosene Withdrawal Temperature	211.81 °C
Diesel Withdrawal Temperature	284.29 °C
AGO Withdrawal Temperature	306.57 °C
Residue Withdrawal Temperature	348.80 °C
Condenser Duty	26.40 MW
Condenser Temperature Range	115.75-21.11 °C
Pump-Around 1 Duty	28.00 MW
Pump-Around 1 Temperature Range	167.83-104.44 °C
Flash Zone Temperature	358.58 °C
Energy Consumption (E)	76.35 MW

From the results, the total energy consumption (E) decreases by 14.09 MW compared to the no pump-around scheme. The kerosene-naphtha gap is reduced from 34.23 to 32.12°C, remaining well above the specification of 16.7 °C and the yield of naphtha decreases but the yield of kerosene increases because some light components of the vapor are absorbed by the cold pump-around stream and carried to the kerosene withdrawal tray. Note that the total yield of the two products remains constant.

The resulting heat demand-supply diagram is shown in Figure 4.10, the heat surplus in the condenser region is significantly reduced, but the region of PA1 is still larger than the demand. However, to reduce the heat surplus in the region of PA1, a second pump-around is added.

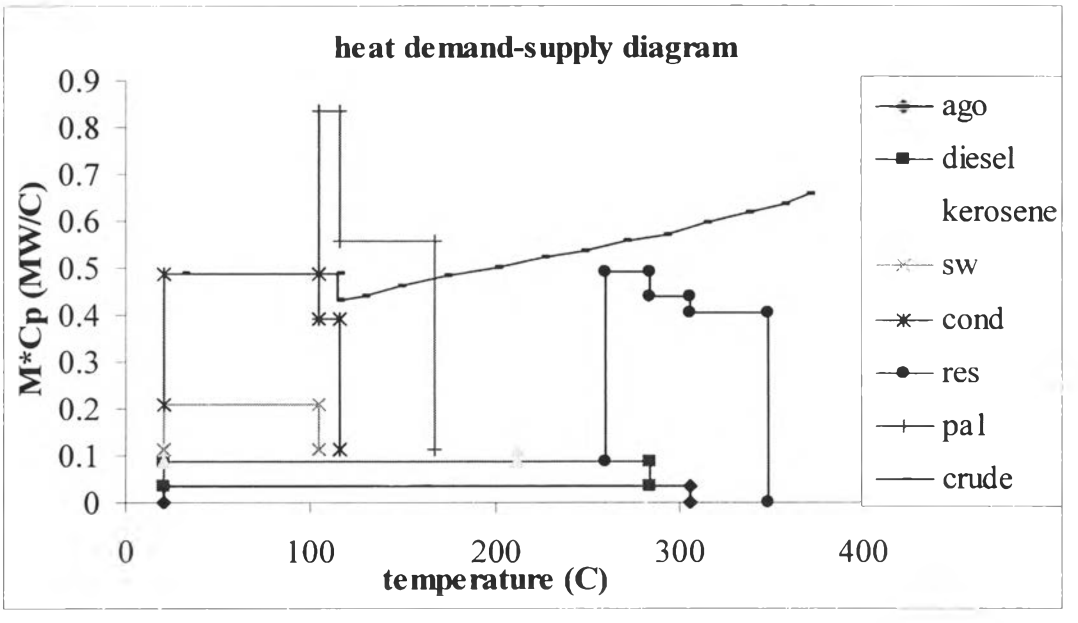

Figure 4.10 Heat demand-supply diagram for heavy crude distillation with a top pump-around.

4.1.3.2 Two Pump-Around Circuits

The second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is 171.11 °C. With the increase in the PA2 duty, the kerosene-naphtha gap decreases. Table 4.18 shows the variable of the scheme with two pump-arounds. Heat shifting continues until the liquid reflux at the kerosene withdrawal tray is small.

Table 4.18 Results of two pump-around circuits scheme for heavy crude

Product	Two Pump-Arounds
Naphtha Flow Rate	55.95 m^3/h
Kerosene Flow Rate	49.57 m^3/h
Diesel Flow Rate	95.55 m^3/h
AGO Flow Rate	61.83 m^3/h
Residue Flow Rate	532.51 m^3/h
Kerosene Stripping Steam Ratio	9.84
Diesel Stripping Steam Ratio	6.26
AGO Stripping Steam Ratio	13.46
Residue Stripping Steam Ratio	11.38
(5-95) Kerosene-Naphtha Gap	30.21 °C
(5-95) Diesel-Kerosene Gap	5.86 °C
(5-95) AGO-Diesel Gap	-10.80 °C
Kerosene Withdrawal Temperature	206.76 °C
Diesel Withdrawal Temperature	283.22 °C
AGO Withdrawal Temperature	306.51 °C
Residue Withdrawal Temperature	348.80 °C
Condenser Duty	26.50 MW
Condenser Temperature Range	115.82-21.11 °C
Pump-Around 1 Duty	18.00 MW
Pump-Around 1 Temperature Range	163.52-104.44 °C
Pump-Around 2 Duty	10.00 MW
Pump-Around 2 Temperature Range	266.39-171.11 °C
Flash Zone Temperature	358.58 °C
Energy Consumption (E)	74.66 MW

From the results, the net energy consumption decreases by 1.69 MW and the yield of diesel increases while the yield of naphtha decreases.

The heat demand-supply diagram is shown in Figure 4.11, the heat surplus in the region of PA1 is still high, but further shifting would cost too much steam to be beneficial. Therefore, this remaining heat surplus is useless.

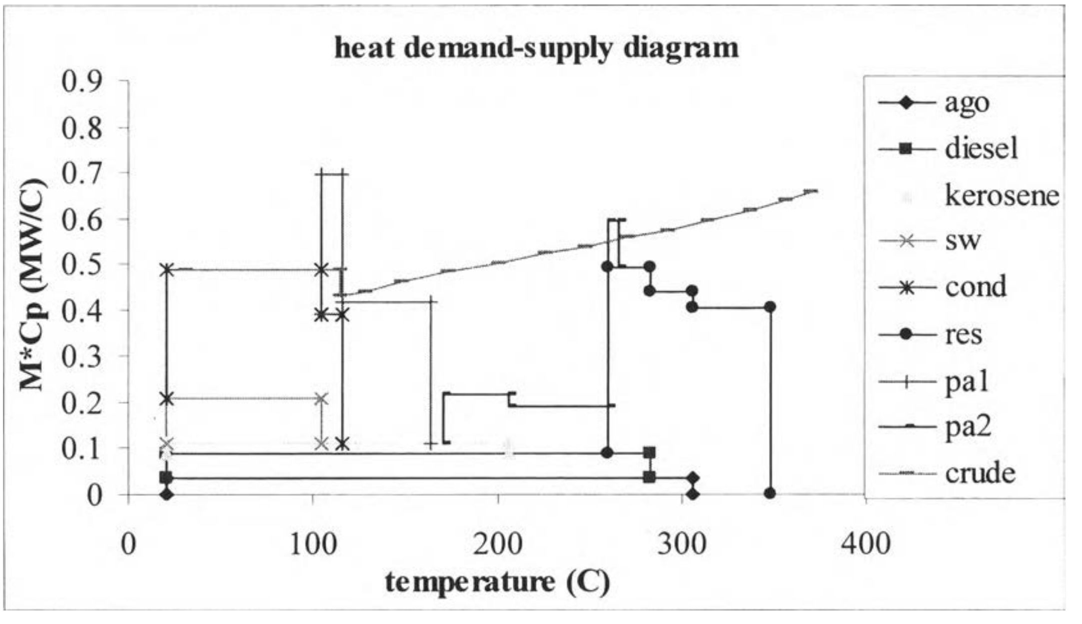

Figure 4.11 Heat demand-supply diagram for heavy crude distillation with two pump-around circuits.

Now, the only heat surplus transferable is located in the PA2 circuit, to make use of heat surplus by adding a third pump-around circuit.

4.1.3.3 Three Pump-Around Circuits

The third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. Heat is shifted gradually from PA2 to PA3, with the gap maintained above the specification. The total energy consumption and the third pump-around (PA3) duty distribution are shown in Table 4.19. A summary of all of the variables and the heat demand-supply diagram are shown in Table 4.20 and Figure 4.12.

Table 4.19 Effect of the duty of PA3 on energy consumption for heavy crude

PA3 duty (MW)	Energy consumption (MW)
1	74.58
1.5	74.65
5	74.70
7	74.88

At first, the energy consumption decreases with the increase in the duty of PA3. However, when the PA3 duty exceeds 1 MW, the energy consumption increases because of the increased use of steam to recover the AGO-diesel gap. Clearly, 1 MW is right point at heat shifting to PA3.

Table 4.20 Results of three pump-around circuits scheme for heavy crude

Product	Three Pump-Arounds
Naphtha Flow Rate	55.95 m^3/h
Kerosene Flow Rate	49.51 m^3/h
Diesel Flow Rate	95.29 m^3/h
AGO Flow Rate	62.16 m^3/h
Residue Flow Rate	532.51 m^3/h
Kerosene Stripping Steam Ratio	9.86
Diesel Stripping Steam Ratio	6.27
AGO Stripping Steam Ratio	13.39
Residue Stripping Steam Ratio	11.38
(5-95) Kerosene-Naphtha Gap	30.23 °C
(5-95) Diesel-Kerosene Gap	5.56 °C
(5-95) AGO-Diesel Gap	-10.95 °C
Kerosene Withdrawal Temperature	206.77 °C
Diesel Withdrawal Temperature	282.47 °C
AGO Withdrawal Temperature	306.50 °C
Residue Withdrawal Temperature	348.81 °C
Condenser Duty	26.50MW
Condenser Temperature Range	115.89-21.11 °C
Pump-Around 1 Duty	18.00 MW
Pump-Around 1 Temperature Range	163.67-104.44 °C
Pump-Around 2 Duty	9.00 MW
Pump-Around 2 Temperature Range	265.95-171.11°C
Pump-Around 3 Duty	1.00 MW
Pump-Around 3 Temperature Range	316.89-232.22 °C
Flash Zone Temperature	358.58 °C
Energy Consumption (E)	74.58 MW

The following results are monitored: The energy consumption changes very little when heat is shifted from the PA2 to PA3. This is because there is a little heat surplus in the PA2 region. However, because the light crude and the intermediate crude require the PA2 and PA3 heat exchangers, shifting heat from PA1 to PA2 and PA3 in heavy crude design might be necessary. The AGO stripping steam flow rates for designs with one, two and three pump-around circuits are 2373.19kg/hr, respectively because the AGO-diesel maintained product quality above the specification.

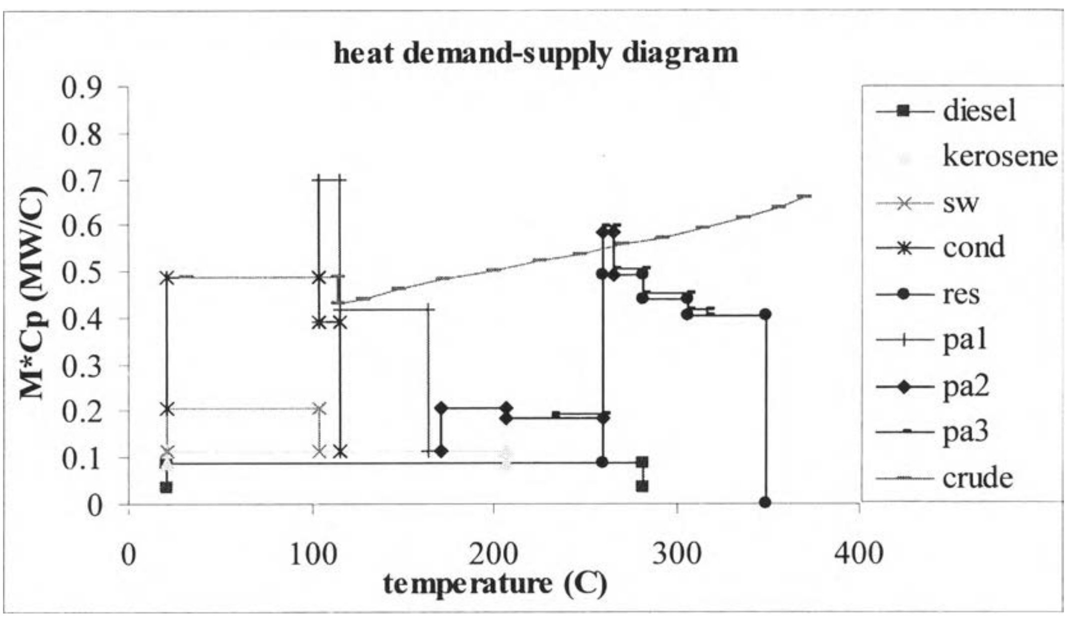

Figure 4.12 Heat demand-supply diagram for heavy crude distillation with three pump-around circuits.

4.2 The Relationship of the Steam Consumption of the Side Strippers and the Duty of Pump-Around Circuits

The steam consumption relates to the loads of the pump around circuits.

$$H_s^c = H_{s,1}^c + H_{s,2}^c + H_{s,3}^c + H_{s,M}^c$$

Where $H_{s,1}^{c}, H_{s,2}^{c}, H_{s,3}^{c}$ and $H_{s,M}^{c}$ are the steam consumption of the side-strippers and the main bottom steam stream. These values depend on the distribution of the pump around loads. As these increase, the steam consumption increases, because of the need to maintain the gap. The relationship can be obtained from regression by assuming a linear relationship proposing the coefficients to be obtained by fitting the results and by another assumption of non linear relationship. Thus,

Linear relationship

$$H_{s,r}^{c} = \sum_{z \in Z} \sum_{i \in PA^{z}} \lambda_{i}^{c} Q_{i}^{c}$$

Non linear relationship

$$H_{s,r}^{c} = \sum_{z \in Z} \sum_{i \in PA^{z}} \lambda_{i}^{c} {Q_{i}^{c}}^{2}$$

Where Q_{i}^{c} are the heat duties of pump-around circuits i and λ_{i}^{c} are constant values.

4.2.1 <u>Light Crude</u>

Assuming the total load of heat removed by the pump-around circuits and the condenser are roughly constant. In addition, a minimum amount of reflux has been used. Lower values will dry trays out with the near zero liquid flow rates. Which is a small fraction (1%) of the feed or a small number in this circumstance.

First, adding the top pump-around circuit which pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C. The heat shift continues until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on the top tray.

Then, the second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is chosen to be around equal to the withdrawal temperature of PA1 that is 171.11 °C. Heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 9.

Finally, the third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. The heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 16. However the steams at each side strippers were used for maintaining all of the gap specifications to keep the product on specification. The results of each pump-around circuit's duties and steam consumption at each side strippers are shown in Table 4.21.

Table 4.21 The duties of each pump-around circuits with steam consumption at side stripper and product gap for light crude

Light Crude	Results
PA1 duty (M*KJ/hr)	79.20
PA2 duty (M*KJ/hr)	61.20
PA3 duty (M*KJ/hr)	111.60
Kerosene stripping steam (Kg/hr)	8242.00
Diesel stripping steam (Kg/hr)	11031.80
AGO stripping steam (Kg/hr)	3452.15
Residue stripping steam (Kg/hr)	6020.07
Kerosene-naphtha (5-95)gap (°C)	16.70
Diesel-kerosene (5-95)gap (°C)	0.00
AGO-diesel (5-95)gap (°C)	0.26

To obtain the relationship, the expression is needed to fit all possible pump-around loads. The results of variable of pump-around circuit's duties and steam consumption of each side strippers from simulation are shown in Table 4.22.

Table 4.22 The results of stripping steam consumption of various pump-around loads for light crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 79.20 PA2 = 61.20 PA3 = 111.60	8242.00	11031.80	3452.15
PA1 = 77.40 PA2 = 59.40 PA3 = 109.80	7650.00	10486.00	3452.15
PA1 = 75.60 PA2 = 57.60 PA3 = 108.00	7106.00	9971.00	3452.15
PA1 = 72.00 PA2 = 54.00 PA3 = 104.40	6141.00	9017.00	3452.15
PA1 = 70.20 PA2 = 52.20 PA3 = 102.60	5713.00	8573.00	3452.15
PA1 = 68.40 PA2 = 50.40 PA3 = 100.80	5316.50	8155.00	3452.15
PA1 = 66.60 PA2 = 48.60 PA3 = 99.00	4949.50	7756.50	3452.15
PA1 = 64.80 PA2 = 46.80 PA3 = 97.20	4608.00	7380.00	3452.15

Table 4.22 (Continue) The results of stripping steam consumption of various pump-around loads for light crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 63.00 PA2 = 45.00 PA3 = 95.40	4291.50	7025.00	3452.15
PA1 = 61.20 PA2 = 43.20 PA3 = 93.60	4108.32	6705.00	3452.15
PA1 = 57.60 PA2 = 39.60 PA3 = 90.00	4108.32	6120.00	3452.15
PA1 = 46.80 PA2 = 28.80 PA3 = 79.20	4108.32	4658.00	3452.15
PA1 = 39.60 PA2 = 21.60 PA3 = 72.00	4108.32	3880.00	3452.15
PA1 = 36.00 PA2 = 18.00 PA3 = 68.00	4108.32	3540.00	3452.15
PA1 = 0 PA2 = 0 PA3 = 0	4108.32	1997.12	3452.15

For the expression, the least linear regression method is used to obtain the relationship between the load of pump-around circuits and steam consumption at each side strippers. The expressions are

for linear relationship assumption

$$H^{c}_{s,1} = 4108.313 + 1622614.602Q^{c}_{1} - 1042972.256Q^{c}_{2} - 579555.316Q^{c}_{3}$$

R Squared = 0.721

$$H^{c}_{s,2} = 1997.120 + 742686.005Q^{c}_{1} - 477227.924Q^{c}_{2} - 265287.167Q^{c}_{3}$$

R Squared = 0.978

$$H^{c}_{s,3} = 3452.1526$$

for non linear relationship assumption

$$H^{c}_{s,1} = 4108.368 - 21.456(Q^{c}_{1})^{2} + 22.077(Q^{c}_{2})^{2} + 4.487(Q^{c}_{3})^{2}$$

R Squared = 0.979

$$H^{c}_{s,2} = 1997.131 - 4.307(Q^{c}_{1})^{2} + 5.821(Q^{c}_{2})^{2} + 1.138(Q^{c}_{3})^{2}$$

R Squared = 1.000

$$H^{c}_{s,3} = 3452.1526$$

The major conclusions from the expression are as follows: the relationship between the load of pump-around circuits and steam consumption at each side strippers are non linear relationship.

4.2.2 Intermediate Crude

Assuming the total load of heat removed by the pump-around circuits and the condenser are roughly constant. In addition, a minimum amount of reflux has been used. Lower values will dry trays out (the definition of dry here is not zero liquid flow rates). Rather, a small fraction (1%) of the feed or any small number in this circumstance.

First, adding the top pump-around circuit which pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C. The heat shift continues until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on the top tray.

Then, the second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is chosen to be around the withdrawal temperature of PA1 of 171.11 °C. Heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 9.

Finally, the third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. The heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 16. However the steam at each side strippers were used for maintaining all of the gap specifications. The results of each pump-around circuit's duties and steam consumption at each side strippers are shown in Table 4.23.

Table 4.23 The duties of each pump-around circuits with steam consumption at side stripper and product gap for intermediate crude

Intermediate Crude	Results
PA1 duty (M*KJ/hr)	50.40
PA2 duty (M*KJ/hr)	54.00
PA3 duty (M*KJ/hr)	64.80
Kerosene stripping steam (Kg/hr)	8529.00
Diesel stripping steam (Kg/hr)	5569.00
AGO stripping steam (Kg/hr)	1672.89
Residue stripping steam (Kg/hr)	12173.05
Kerosene-naphtha (5-95)gap (°C)	16.70
Diesel-kerosene (5-95)gap (°C)	0.00
AGO-diesel (5-95)gap (°C)	-9.74

To obtain the relationship, the expression is needed to fit all possible pump-around loads. The results of variable of pump-around circuit's duties and steam consumption of each side strippers from simulation are shown in Table 4.24.

Table 4.24 The results of stripping steam consumption of various pump-around loads for intermediate crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 50.40 PA2 = 54.00 PA3 = 64.80	8529.00	5569.00	1672.89
PA1 = 48.60 PA2 = 52.20 PA3 = 63.00	7700.00	5275.00	1672.89
PA1 = 46.80 PA2 = 50.40 PA3 = 61.20	6965.00	4995.00	1672.89
PA1 = 45.00 PA2 = 48.60 PA3 = 59.40	6305.00	4732.00	1672.89
PA1 = 43.20 PA2 = 46.80 PA3 = 57.60	5713.00	4482.00	1672.89
PA1 = 41.40 PA2 = 45.00 PA3 = 55.80	5180.00	4246.00	1672.89
PA1 = 39.60 PA2 = 43.20 PA3 = 54.00	4700.00	4023.00	1672.89

Table 4.24 (Continue) The results of stripping steam consumption of various pump-around loads for intermediate crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 37.80 PA2 = 41.40 PA3 = 52.20	4265.00	3812.00	1672.89
PA1 = 27.00 PA2 = 30.60 PA3 = 41.40	2707.1731	2766.00	1672.89
PA1 = 25.20 PA2 = 28.80 PA3 = 39.60	2707.1731	2626.00	1672.89
PA1 = 21.60 PA2 = 25.20 PA3 = 36.00	2707.1731	2367.00	1672.89
PA1 = 18.00 PA2 = 21.60 PA3 = 32.40	2707.1731	2128.50	1672.89
PA1 = 14.40 PA2 = 18.00 PA3 = 28.80	2707.1731	1911.00	1672.89
PA1 = 12.60 PA2 = 16.20 PA3 = 27.00	2707.1731	1810.50	1672.89
PA1 = 10.80 PA2 = 14.40 PA3 = 25.20	2707.1731	1731.00	1672.89

Table 4.24 (Continue) The results of stripping steam consumption of various pump-around loads for intermediate crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 9.00 PA2 = 12.60 PA3 = 23.40	2707.1731	1731.00	1672.89
PA1 = 7.20 PA2 = 10.80 PA3 = 21.60	2707.1731	1731.00	1672.89
PA1 = 0 PA2 = 0 PA3 = 0	2707.1731	1731.00	1672.89

For the expression, the least linear regression method is used to obtain the relationship between the load of pump-around circuits and steam consumption at each side strippers. The expressions are

for linear relationship assumption

$$H_{s,1}^{c} = 2707.173 + 10232.752 Q_1^c - 13315.064 Q_2^c + 3200.730 Q_3^c$$

R Squared = 0.819

$$H_{s,2}^{c} = 1730.998 + 4939.097 Q_1^c - 6367.908 Q_2^c + 1518.200 Q_3^c$$

R Squared = 0.974

$$H_{s,3}^{c} = 1672.890$$

for non linear relationship assumption

$$H_{s,1}^{c} = 2707.173 + 71.647(Q_1^c)^2 - 78.470(Q_2^c)^2 + 12.451(Q_3^c)^2$$

R Squared = 0.989

$$H_{s,2}^{c} = 1730.998 - 1.369(Q_1^c)^2 + 3.800(Q_2^c)^2 - 0.904(Q_3^c)^2$$

R Squared = 1.000

$$H_{s,3}^{c} = 1672.890$$

The major conclusion from the expression is as follows: the relationship between the load of pump-around circuits and steam consumption at each side strippers are non linear relationship.

4.2.3 Heavy Crude

Assuming the total load of heat removed by the pump-around circuits and the condenser are roughly constant. In addition, a minimum amount of reflux has been used. Lower values will dry trays out that the definition of dry is not zero liquid flow rates. Rather, a small fraction (1%) of the feed or any small number in this circumstance.

First, adding the top pump-around circuit which pump-around stream is withdrawn from tray 4, cooled in the heat exchangers, and returned to tray 2. The return temperature is 104.4 °C. The heat shift continues until the reflux ratio reaches 0.1. Further heat shift would result in liquid drying up on the top tray.

Then, the second pump-around (PA2) is located between tray 10 and tray 12, just below the kerosene withdrawal tray. The return temperature is chosen to be around the withdrawal temperature of PA1 of 171.11 °C. Heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 9.

Finally, the third pump-around (PA3) is located between tray 17 and tray 19. The return temperature is 232 °C. The heat shift continues without violating the gap specifications until the reflux ratio is around 0.1. Further heat shift would result in liquid drying up on tray 16. However the steam at each side strippers were used for maintain all of the gap specifications. The results of each pump-around circuit's duties and steam consumption at each side strippers are shown in Table 4.25.

Table 4.25 The duties of each pump-around circuits with steam consumption at side stripper and product gap for heavy crude

Heavy Crude	Results
PA1 duty (M*KJ/hr)	28.80
PA2 duty (M*KJ/hr)	43.20
PA3 duty (M*KJ/hr)	32.40
Kerosene stripping steam (Kg/hr)	1392.26
Diesel stripping steam (Kg/hr)	1705.23
AGO stripping steam (Kg/hr)	2373.78
Residue stripping steam (Kg/hr)	17284.12
Kerosene-naphtha (5-95)gap (°C)	26.48
Diesel-kerosene (5-95)gap (°C)	0.66
AGO-diesel (5-95)gap (°C)	-11.00

To obtain the relationship, the expression is needed to fit all possible pump-around loads. The results of variable of pump-around circuit's duties and steam consumption of each side strippers from simulation are shown in Table 4.26.

Table 4.26 The results of stripping steam consumption of various pump-around loads for heavy crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 28.80 PA2 = 43.20 PA3 = 32.40	1392.26	1705.23	2472.00
PA1 = 27.00 PA2 = 41.40 PA3 = 30.60	1392.26	1705.23	2470.00
PA1 = 25.20 PA2 = 39.60 PA3 = 28.80	1392.26	1705.23	2467.60
PA1 = 23.40 PA2 = 37.80 PA3 = 27.00	1392.26	1705.23	2465.00
PA1 = 21.60 PA2 = 36.00 PA3 = 25.20	1392.26	1705.23	2462.50
PA1 = 19.80 PA2 = 34.20 PA3 = 23.40	1392.26	1705.23	2457.80
PA1 = 14.40 PA2 = 28.80 PA3 = 18.00	1392.26	1705.23	2444.50
PA1 = 10.80 PA2 = 25.20 PA3 = 14.40	1392.26	1705.23	2432.00

Table 4.26 (Continue) The results of stripping steam consumption of various pump-around loads for intermediate crude

Pump-around duty (M*KJ/hr)	Kerosene stripping steam (Kg/hr)	Diesel stripping steam (Kg/hr)	AGO stripping steam (Kg/hr)
PA1 = 7.20 PA2 = 21.60 PA3 = 10.80	1392.26	1705.23	2417.40
PA1 = 5.40 PA2 = 19.80 PA3 = 9.00	1392.26	1705.23	2407.50
PA1 = 3.60 PA2 = 18.00 PA3 = 7.20	1392.26	1705.23	2398.00
PA1 = 0 PA2 = 0 PA3 = 0	1392.26	1705.23	2373.78

For the expression, the least linear regression method is used to obtain the relationship between the load of pump-around circuits and steam consumption at each side strippers. The expressions are

for linear relationship assumption

$$H^c_{s,1} = 1392.255$$

$$H^c_{s,2} = 1705.234$$

$$H^c_{s,3} = 2373.783 + 45.368Q^c_1 + 16.202Q^c_2 - 58.659Q^c_3$$

R Squared = 1.000

for non linear relationship assumption

$$H_{s,1}^{c} = 1392.255$$

$$H_{s,2}^{c} = 1705.234$$

$$H_{s,3}^{c} = 2373.783 - 0.999(Q_1^c)^2 - 0.033(Q_2^c)^2 + 0.941(Q_3^c)^2$$

R Squared = 1.000

The major conclusion from the expression is as follows: the relationship between the load of pump-around circuits and steam consumption at each side strippers are non linear relationship.

4.3 Multipurpose/Multiperiod HEN Model

The data presented in part of targeting that are used for a plant processing 795 m^3/h (120000 barrels per day).

The flowsheet of heat exchanger network is constructed by making the crude get in contact with increasingly hot streams. The same sequence model is used for all crudes. The results show the base case model for the light, intermediate and heavy crude in Figure 4.13.

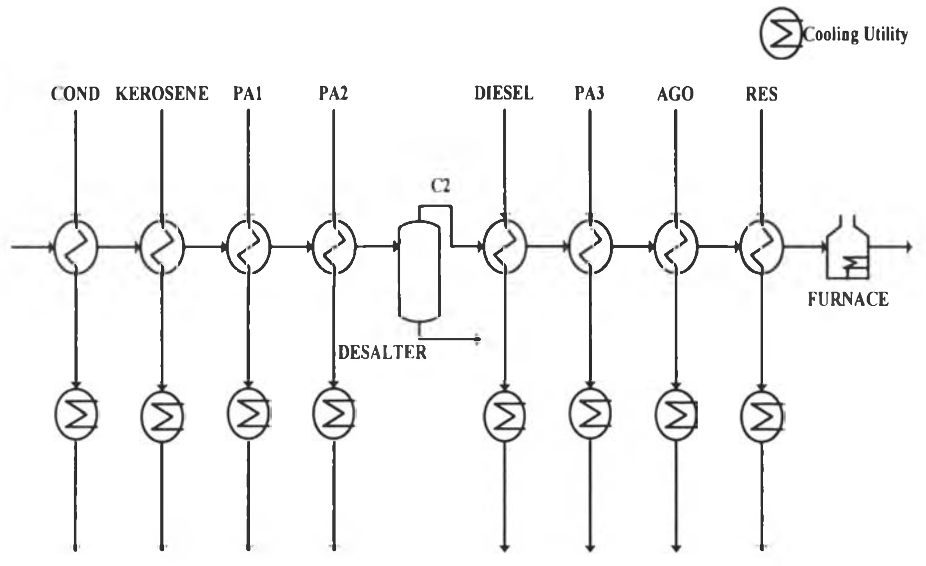

Figure 4.13 The base case of heat exchanger network model.

The optimization is applied to obtain the areas of heat exchanger network for all crudes by using optimizer in PROVISION II, before and after the desalter. The results show that, for the light crude, the network with areas and costs are required in Table 4.27 and 4.28, respectively.

Table 4.27 Areas for light crude

unit	Area (m^2)
Condenser – C1	217.828
PA1 – C1	452.104
Kerosene – C1	647.802
PA2 – C1	602.102
Diesel– C2	1925.51
PA3 – C2	247.907
AGO – C2	4010.68
Residue – C2	848.133

Table 4.28 Area and costs for light crude

	HEN model for light crude
Total area, m^2	8952.066
Fixed Cost, $/year	43289.12881
Operating Cost, $/year	7195220.096
Total cost, $/year	7238509.129

The results of areas and costs for intermediate crude are shown in Table 4.29 and 4.30, respectively.

Table 4.29 Areas for intermediate crude

unit	Area (m^2)
Condenser – C1	241.455
PA1 – C1	690.235
Kerosene– C1	588.137
PA2 – C1	631.056
Diesel – C2	1674.16
PA3 – C2	97.1352
AGO – C2	1916.39
Residue – C2	1665.89

Table 4.30 Area and costs for intermediate crude

	HEN model for intermediate crude
Total area, m^2	7504.4582
Fixed Cost, $/year	38599.90978
Operating Cost, $/year	6243003.598
Total cost, $/year	6281603.508

The results of areas and costs for heavy crude are shown in Table 4.31 and 4.32, respectively.

Table 4.31 Areas for heavy crude

unit	Area (m^2)
Condenser – C1	910.108
PA1 – C1	991.138
Kerosene – C1	613.065
PA2 – C1	448.693
Diesel – C2	1987.30
PA3 – C2	49.1488
AGO – C2	1575.02
Residue – C2	2356.44

Table 4.32 Area and costs for heavy crude

	HEN model for heavy crude
Total area, m^2	8930.9128
Fixed Cost, $/year	43222.61302
Operating Cost, $/year	5309554.816
Total cost, 10^6 $/year	5352777.429

The multiperiod heat exchanger network model is designed by assuming that the largest area of three types of crudes for each heat exchanger is chosen to use in the multiperiod heat exchanger network model. The model is solved by using PROVISION II. Figure 4.14 shows the heat exchanger network consisting of bypasses exists and using controller to control the bypasses for achieving the target temperature. The areas are presented in Table 4.33, and costs of network are shown in Table 4.34.

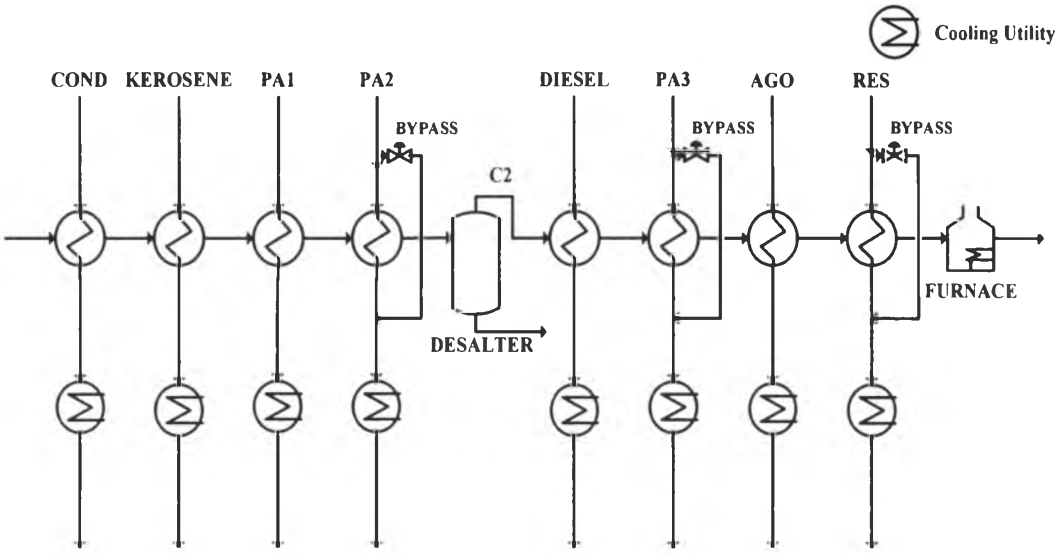

Figure 4.14 The multipurpose/multiperiod heat exchanger network model.

Table 4.33 Areas for multiperiod model

unit	Area (m^2)
Condenser – C1	910.108
PA1 – C1	991.138
Kerosene – C1	647.802
PA2 – C1	631.056
Diesel– C2	1987.30
PA3 – C2	247.907
AGO – C2	4010.68
Residue – C2	2356.44

Table 4.34 Area and costs for multiperiod model

	HEN model for multiperiod model
Total area, m^2	11782.431
Fixed Cost, $/year	51752.00
Operating Cost, $/year	7196300.00
Total cost, $/year	7248052.00

4.3.1 Retrofit

In multipurpose heat exchanger network, when the sequence of heat exchanger network is discussed, an existing plant of heat exchanger network is shown in Figure 4.14. The additional one heat exchanger is required in the network model. The possibility of adding one heat exchanger will be evaluated to improve existing plant. The condenser, kerosene, PA1, diesel, PA3 and residue streams are omitted because their supply amounts are relatively smaller.

First, adding a heat exchanger to the PA2, in the position of desalter-diesel, diesel-PA3, PA3-AGO, AGO-residue and residue heat exchanger-furnace is shown in Figure 4.15 – 4.19, respectively and using by pass PA2 stream to improve energy efficiency. The table 4.35 is based on the profitability criteria with optimization by PROVISION II.

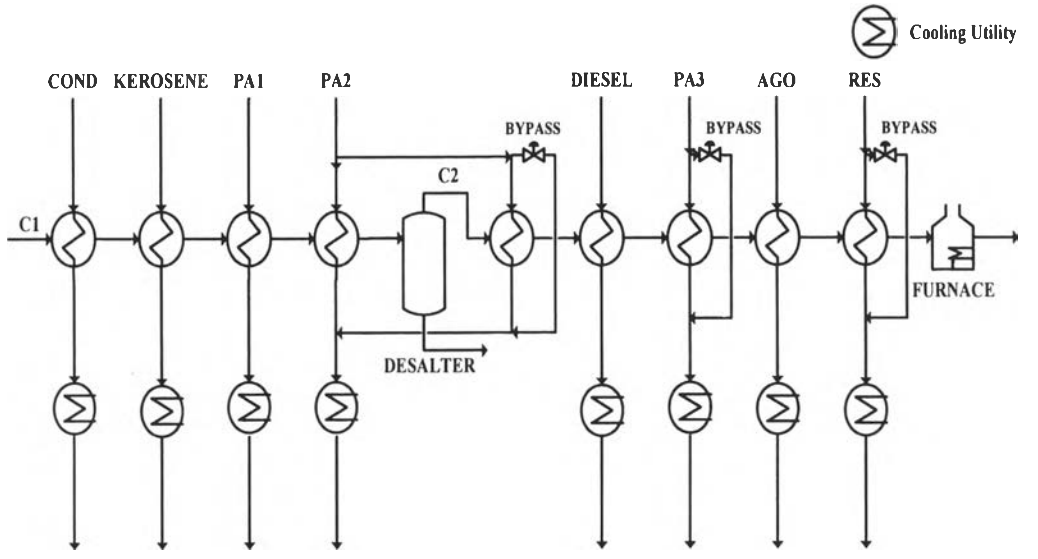

Figure 4.15 The additional one heat exchanger to PA2 in the position of desalter and diesel heat exchanger.

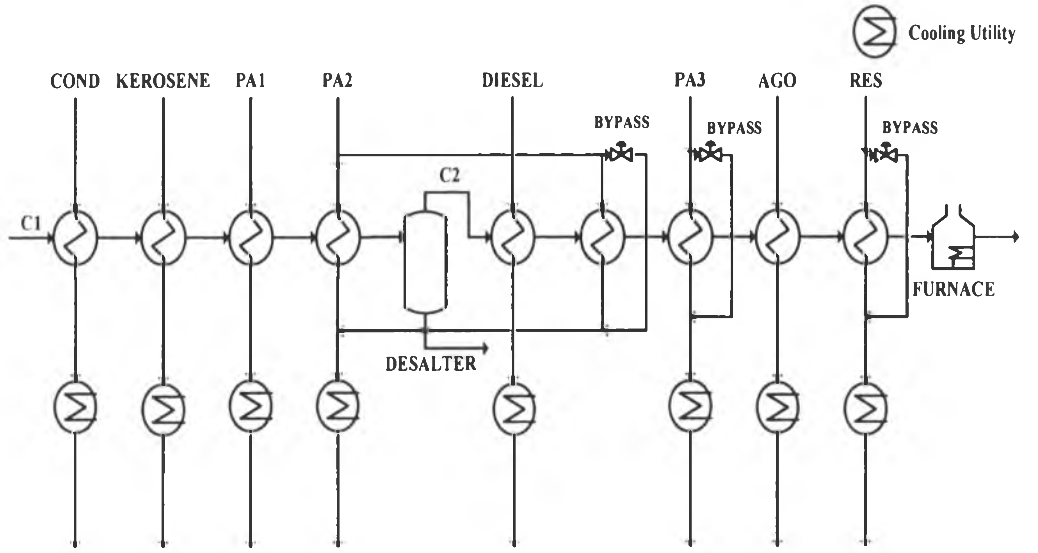

Figure 4.16 The additional one heat exchanger to PA2 in the position of diesel and PA3 heat exchangers.

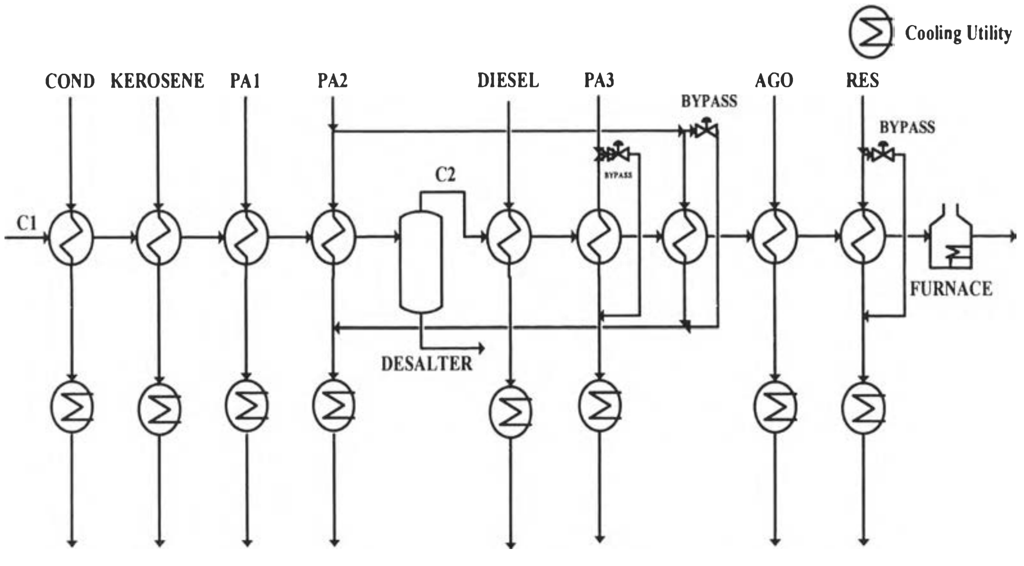

Figure 4.17 The additional one heat exchanger to PA2 in the position of PA3 and AGO heat exchangers.

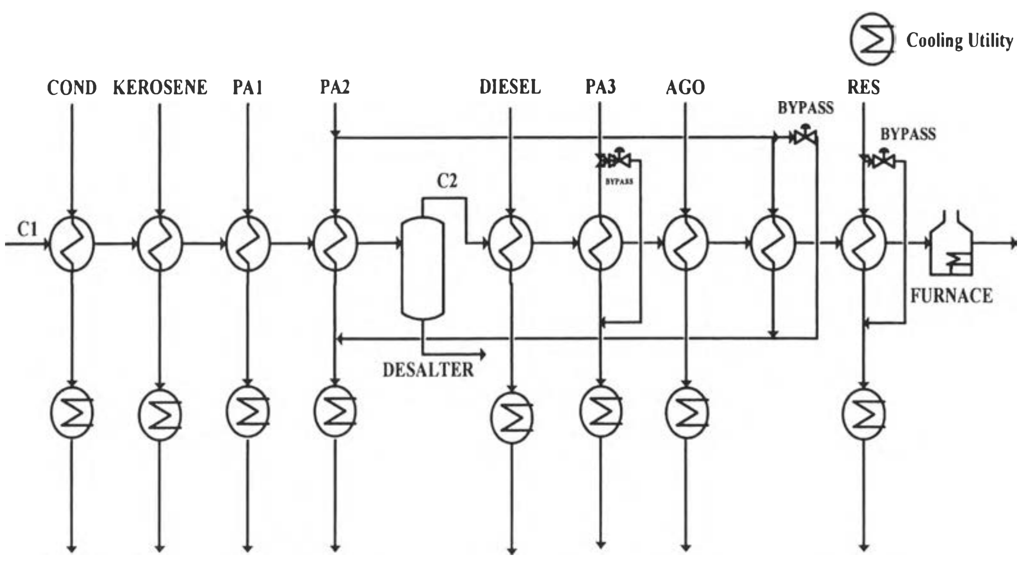

Figure 4.18 The additional one heat exchanger to PA2 in the position of AGO and residue heat exchangers.

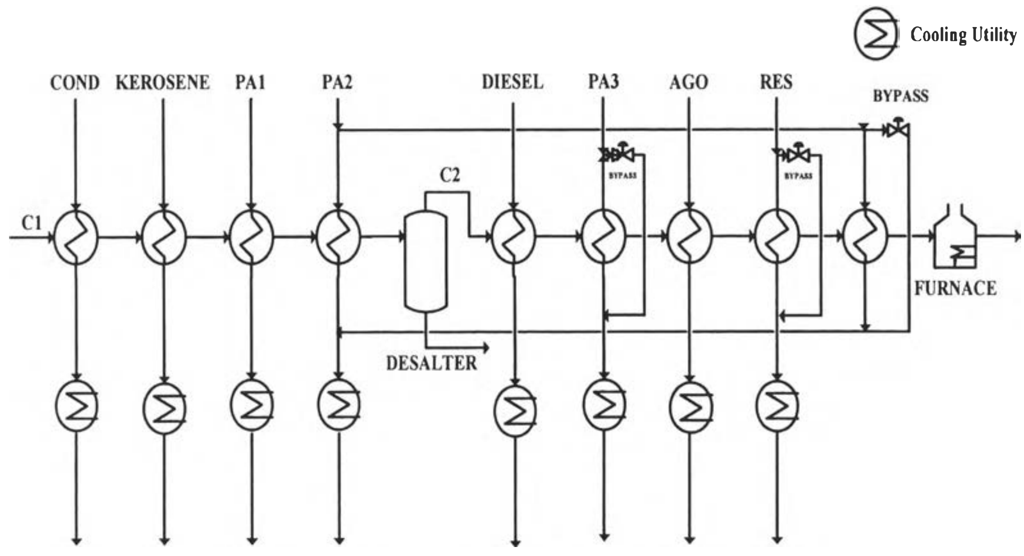

Figure 4.19 The additional one heat exchanger to PA2 in the position of residue heat exchanger and furnace.

Table 4.35 Economics of additional one heat exchanger to the PA2 in any positions

Position	Area added (m^2)	Cost of HEN, ($)	Depreciation Cost ($)	Utility saving cost ($)	Net Present Value ($)
Desalter-diesel	4004.97	625940	194910	6290672	5859642.30
Diesel-PA3	8561.96	738100	229840	6877041	6368780.70
PA3-AGO	11807.90	812650	253050	6815264	6255664.30
AGO-residue	11649.50	809100	251950	6509306	5952156.30
Residue-furnace	11635.40	808790	251850	4695610	4138670.00

The major results from simulation are as follows: when adding PA2 heat exchanger in the position of diesel and PA3 heat exchangers, the utility saving cost is the largest that makes the NPV is also the largest.

Consider adding a heat exchanger to the PA2, adding a heat exchanger between diesel and PA3 heat exchangers is shown in Figure 4.16 and using by pass PA2 stream to get more energy efficiency. The Table 4.36 is based on the profitability criteria, heat exchangers with area of 4.86535 m^2, 840.983 m^2 and 8561.96 m^2 can be chosen.

Table 4.36 Economics of additional one heat exchanger to the PA2

Area added	4.86535 m^2	840.983 m^2	8561.96 m^2
Cost of HEN	\$ 517660	\$ 541240	\$ 738100
Depreciation Cost	\$ 161200	\$ 168540	\$ 229840
Utility saving cost	\$ 46162.33	\$ 3835671.00	\$ 6877040.67
Net Present Value	\$ -310597.67	\$ 3462971.00	\$ 6368780.70

The major results from simulation are as follows: when the area increases from 4.86535 m^2 to 8561.96 m^2, the utility saving cost increases, causing this option more attractive for retrofit. Indeed, adding 8561.96 m^2, one obtains a NPV of 6.37 millions dollas. Figure 4.16 shows the retrofit heat exchanger network design optimal solution.

Addition of an exchanger to AGO of similar sizes produces poorer results at the same cost. This is not considering further.

To consider AGO, the resulting of economics analysis is shown in Table 4.37 and the model is shown in Figure 4.20. The major results from simulation are as follows: when adding AGO heat exchanger in the position of diesel and PA3 heat exchangers, the utility saving cost is smaller than adding PA2 in the same position that makes the NPV smaller.

Table 4.37 Economics of additional one heat exchanger to the AGO

Area added	8561.96 m^2
Cost of HEN	$ 738100
Depreciation Cost	$ 229840
Utility saving cost	$ 346596
Net Present Value	$ -161664

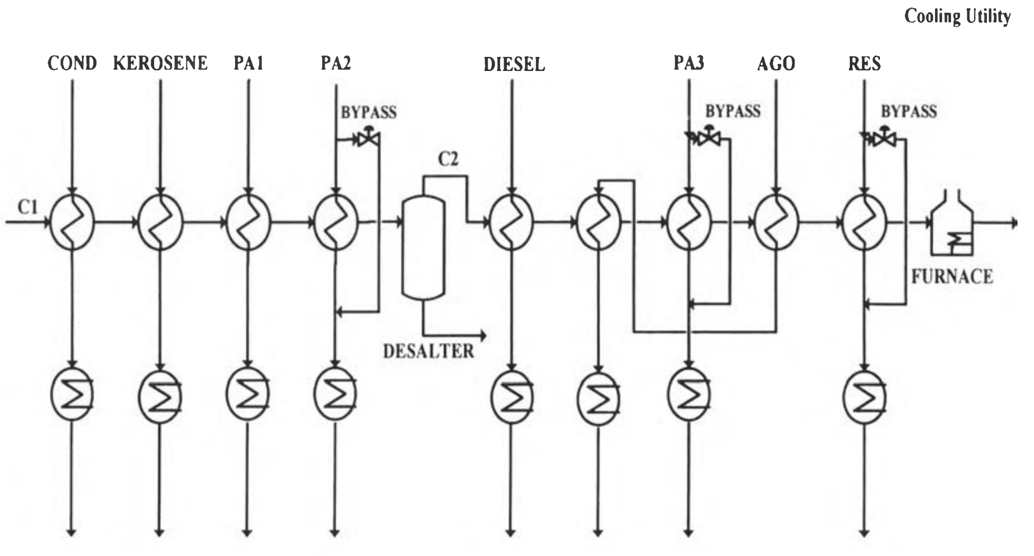

Figure 4.20 The additional one heat exchanger to AGO.

From all of possibility designs, the best solution for retrofitting is adding one heat exchanger to PA2 in the position of diesel and PA3 heat exchangers.