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Review Paper

Heat Integration of Crude Organic Distillation Unit

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Abstract

While oil prices continue to climb, energy conservation remains the prime concern for many process industries. The challenge every process engineer faced is to seek the answer to questions related to their process energy pattern. Distillation column are of great importance in process analysis as they are the most common and the most energy intensive separation systems and hence it is the first separation system to be analyzed specifically from a pinch view point. In this paper heat integration of crude organic distillation unit has been done using pinch technology. Pinch Technology involves composite curves, problem table algorithm and heat exchanger network design. Using targeting procedures, hot and cold utility reduction occurs. With this design, cost estimation has been done using heat exchanger cost equation. Although the results found can be used for development of new projects, as heuristics rules, the application has been limited due to lack of understanding of the subject

Keywords: pinch technology, composite curve, problem table algorithm, heat exchanger network.

Introduction

Pinch Analysis Techniques for integrated network design presented in this paper were originally developed from the 1970s onwards at the ETH Zurich and Leeds University (Linnhoff and Flower 1978; linnhoff 1979)[1]. This pinch analysis technique is very much useful for both heat and mass integration. To apply this technique, a systematic study is required. The stages in a process of pinch analysis of a real process plant or site are as follows[2]: i. Obtain or produce a copy of the plant flowsheet including temperature, flow heat capacity data and produce a consistent heat and mass balance. ii. Extract the stream data from the heat and mass balance. iii. Selection of initial ΔT_{min} value. iv. Construction of composite curves and grand composite curve. v. Estimation of minimun energy cost target. vi. Design of heat exchanger network. vii. Estimation of HEN capital cost targets.[2] In this paper process selected is crude organic distillation unit. Heat integration of crude organic distillation unit have been done using above which are given below.

Case Study

Process Description: Crude feed is supplied at ambient temperature and fed to distillation column at atmospheric pressure. It is splitted into three fractions middle oil, light oil, and residue. Feed has to be heated up to the operating temperature of the column. There is no separate reboiler and hence feed is partially vaporized. Some flashing also occurs as the hot liquid enters the column. Light oil comes off the top as vapour (overhead) is condensed and majority is recycled to provide the top reflux. Small amount of water is removed in gravity separator. Various products are cooled to different level depending on the viscosity. Crude feed passes through two heat exchanger i.e.by overhead heat exchanger and middle oil heat exchanger before entering a furnace which brings it up to its final feed temperature. All other heating and cooling duties are performed by utilities[3]. The overall flow sheet is shown in figure-1.

Heat and Mass Balance: Here the data for given work has been taken from[4] Information is available on the production rate of the various fractions and their specific heat capacities and several of the temperature recorded by thermocouple as shown on the process flow sheet figure-1. Hence it is possible to form preliminary heat and mass balance as shown in table-1, and data for existing heat exchanger is given in table-2. However, some information is unknown including number of temperature and exact reflux ratio (it is thought to be approximately 5:1). These unknown stream information is calculated by back calculation. The specific heat capacity for the crude feed is known to be significantly temperature dependant, as (2+0.005T) kJ/kgK. Since the mass flow rate is 10 kg/s and CP = (20+0.05T) kJ/K, therefore by integrating heat load (relative to 0^0C datum) is $(20T+0.025T^2)$ kW. No data are available on heat transfer coefficients. However the sizes of the two existing heat exchangers are known and information have to be deduced from them to obtain the HTC and also provide further information on the heat balance. The heat load on the crude feed-overhead exchanger can be calculated from the known temperatures and heat capacity flow rate of the crude feed; likewise, that for the middle oil-crude feed exchanger can be calculated from the details of the middle oil, and the exit temperature of the crude feed can then be back calculated as 92^0C. Heat loads are calculated as given below,

$Q_{CF} = m_C \times Cp \times \Delta T$

$Q_F = 10 \times 2.1 \times (60-20) = 880kW.$

Here value of Cp is taken from 2+0.005T. T is the mean temperature of inlet and outlet cold fluid temperature. Heat load for the second heat exchanger is calculated from hot fluid i.e. middle oil.

$Q_{MO} = m_h \times Cp \times \Delta T$

$Q_{MO} = 5 \times 2 \times (199-123) = 760KW.$

The overall heat transfer coefficients U for the exchangers are then calculated to be 0.2 and 0.125kW/m^{2}K, respectively. The basic relationship of U to the film heat transfer coefficients h (ignoring fouling and wall resistance) is:

$$\frac{1}{U} = \frac{1}{h_1} + \frac{1}{h_2}$$

Assuming that all organic liquids have the same film HTC, setting $h_1 = h_2$ for the crude feed/middle oil exchanger gives h as 0.25 kW/m^{2}K. Back calculation for the crude feed/overheads exchanger gives a film HTC of 1kW/m^{2}K, which is reasonable for the condensing stream as shown in Table-2. The Cp for the condensing overheads over this range can be calculated as 80kW/K.

$$CP\ (kW/K) = \frac{Q}{\Delta T}$$

$$CP\ (kW/K) = \frac{880}{(123-112)} = 80kW/K.$$

Measurements of cooling water flow and temperature drop indicate that the heat load on the following overheads cooler is 1.8 MW, and since the temperature drop is 60^{o}C this gives a CP of 30kW/K. The current level of heat recovery in these two exchangers is 1640kW .

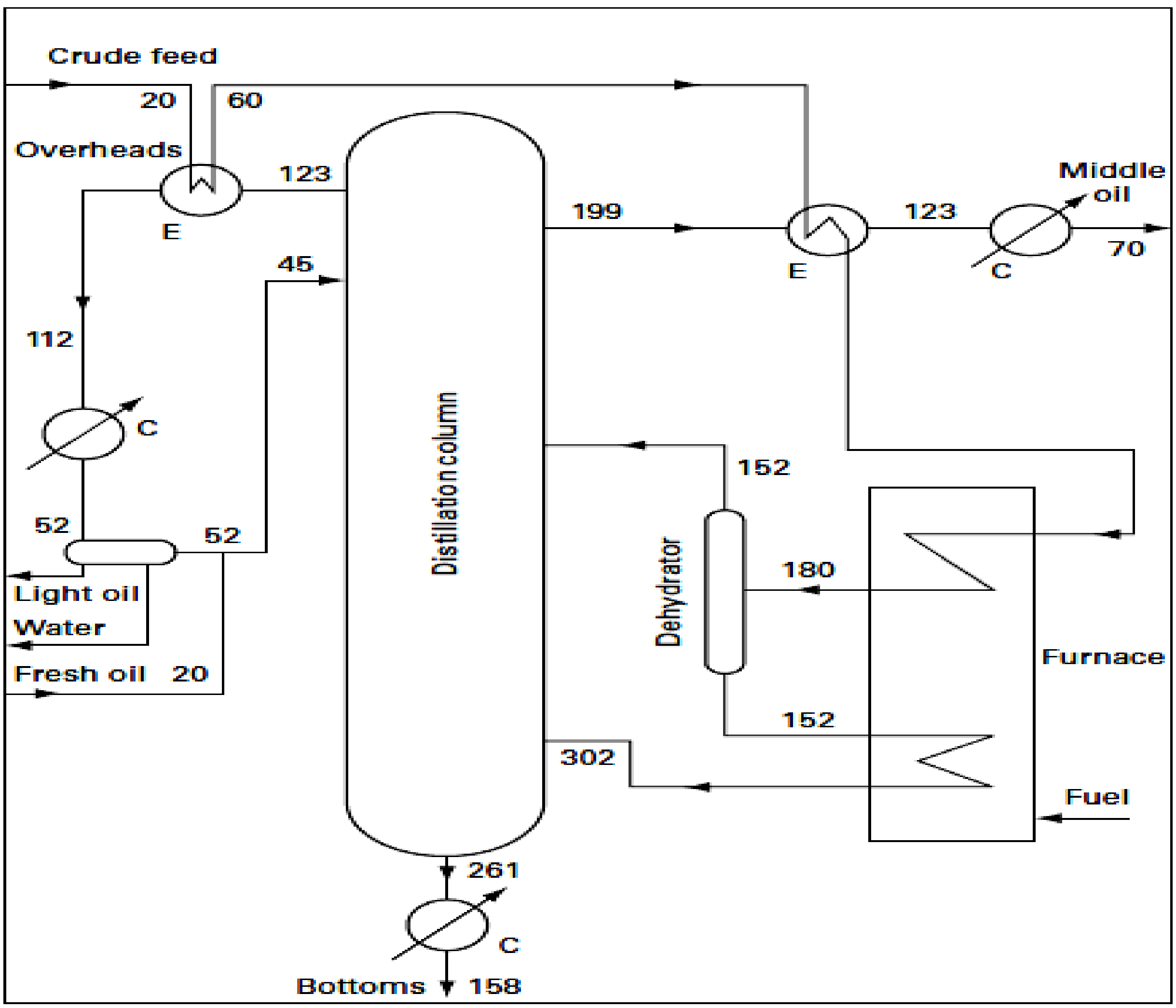


Figure-1
Existing flow sheet for Crude Distillation Unit

Stream Data Extraction: With consistent heat and mass balance available, stream data are extracted. Here only those flows which require heating or cooling are extracted from given process. Therefore, the light oil and water from the separator are ignored. The overheads emerging from the separator are mixed directly with fresh oil and direct heat exchange of 100 kW takes place. The amount of heat involved in these two streams is so small and the temperature so low that they can be safely ignored. This leaves five actual streams whose characteristics are listed in table-3. From above table it has been found that total heat load for the cold streams is 8500kW and for the hot streams is 5000kW. The current level of heat recovery observed is 1640kW. This implies that current hot utility and cold utility demands are 6860kW and 3360kW respectively. For reducing the hot utility and cold utility demand targeting procedure is applied. There are two targeting procedure applied for utility reduction. One is to draw "composite curve" and other is to form "problem table algorithm". For drawing the composite curve, understanding of pinch concept is very essential.

Pinch Analysis Review

Introduced by Linnhoff et al. pinch analysis has a purpose to identify the optimum heat recovery of process and to establish the most promising options concerning cost of the heat exchanger network (HEN). The identification of possible opportunities of thermal integration can be visualized through the hot and cold composite curves (CCs), as shown in figure-2.[5]

Composite curve: These are combinations of the thermal streams of total process, in terms of their heat contents over each temperature level (temperature × enthalpy). Here minimum temperature difference is taken as 20^{o}C and data for actual and shifted temperature curve is given in table-4. Hot and cold composite curves are drawn using data given in table-5. Hot and cold composite curves represent the energy avaibility and the requirement of the overall process respectively. Their overlapping indicates the maximum heat recovery of process, whereas the overshoot determines the minimum hot and cold utility requirements of the process (targets). The minimum temperature difference (ΔT_{min}), with regard to capital cost is the limit for the approximation between the curves and limit imposed by the project i.e. the pinch point. In application terms it is relevant to observe that there should be no heat transfer across the pinch, because transfer of heat indicates an increase in hot and cold utilities[6].

Table-1
Heat and Mass balance for organic process

Flow stream	Production Rate (ton/hr)	Mass flow (kg/s)	Specific heat (kJ/kgK)	Heat capacity rate(kW/K)	Initial temperature (^{o}c)	Final temperature (^{o}c)	Heat Load (kW)
Crude feed	36	10	2+0.05T	25	20	180	4000
Dehydrate	?	(9.67)	3.1	30	152	302	4500
Bottoms	14.4	4	2.5	10	261	158	1030
Middle oil	18	5	2	10	199	70	1290
Light oil	9.6	2.67	2	5.33	52	52	0
Overheads	?	?	?	?	112	45	?
Fresh oil	7.2	2	2	4	20	45	100
Water	1.2	0.33	4.19	1.4	52	52	0

Here question marks denote unknown values and bracketed values were found by back calculation

Table-2
Data for existing Heat Exchangers

Streams	Exchanger area (m^2)	Hot stream temperature (^{o}C)	Cold stream temperature (^{o}C)	Log mean temperature difference (^{o}C)	Calculated heat load (kW)	Calculated overall HTC (kW/m^2K)
CF/Ohds	57.5	123-112	20-60	77	880	0.2
CF/MO	73.2	199-123	60-(92)	83	760	0.125

Table-3
Stream data extraction

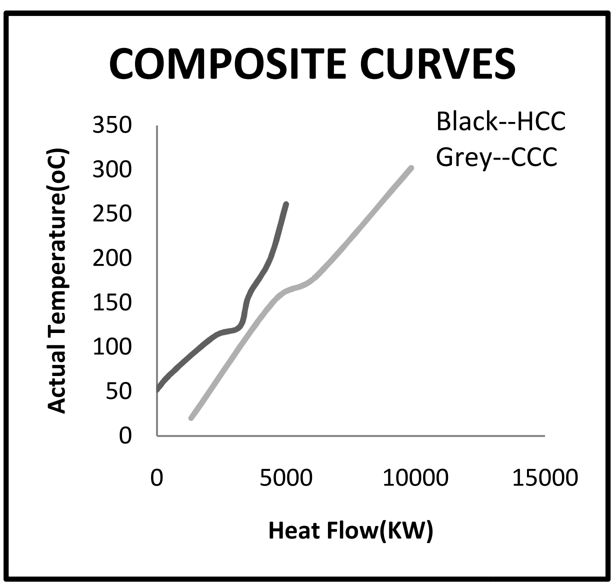
Stream name	Stream type	Initial temperature (^{o}C)	Target temperature (^{o}C)	Film HTC (kW/m^2K)	Heat capacity flow rate (kW/K)	Heat flow rate (kW)
Bottoms	Hot 1	261	158	0.25	10	1030
Middle Oil	Hot 2	199	70	0.25	10	1290
Overheads	Hot 3A	123	112	1	80	880
	Hot 3B	112	52	1	30	1800
Crude Feed	Cold 1	20	180	0.25	20+0.05T	-4000
Dehydrate	Cold 2	152	302	0.25	30	-4500

Table-4
Actual and shifted temperature for given stream extracted

Stream Name	**Actual Temperature**			**Shifted Temperature**		
	Heat capacity Cp(kW/K)	**Inlet Temperature (°C)**	**Outlet Temperature(°C)**	**Heat capacity Cp(kW/K)**	**Inlet Temperature (°C)**	**Outlet Temperature (°C)**
Hot 1	10	261	158	10	251	148
Hot 2	10	199	70	10	189	60
Hot 3A	80	123	112	80	112	102
Hot 3B	30	112	52	30	102	42
Cold 1	20+0.005T	20	180	20+0.005T	30	190
Cold 2	30	152	302	30	162	312

Table-5
Data required for plotting the composite curves.

Actual Temperature(°C)	**Heat capacity(kW/K)**	**Enthalpy(kW)**	**Shifted Temperature(°C)**	**Heat capacity(kW/K)**	**Enthalpy(kW)**
261			251		
	10	620		10	620
199			189		
	20	820		20	820
158			148		
	10	350		10	350
123			113		
	90	990		90	990
112			102		
	40	1680		40	1680
70			60		
	30	540		30	540
52			42		



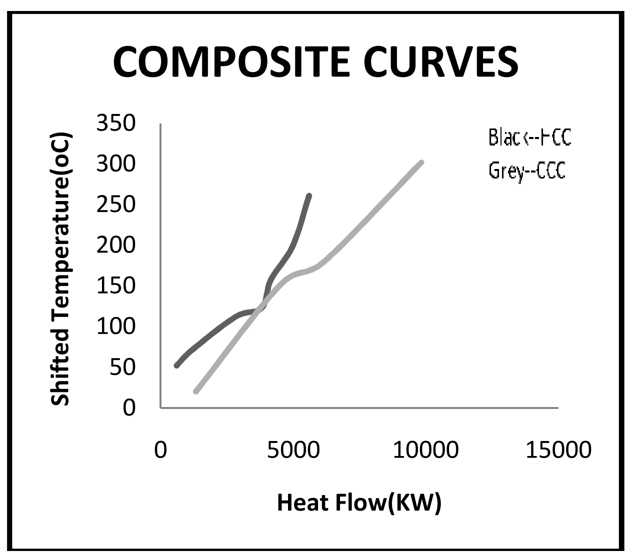


Figure-2
Composite curve

Composite curves for hot and cold fluid for actual and shifted temperature are drawn using data shown in table-4 and 5. Composite curves for actual temperature are shown in figure 2(a) and for shifted temperature are shown in figure 2(b). The grand composite curve (GCC) shown in figure-3 is another tool, also used in pinch analysis. This grand composite curve is plotted using cascade figure-5 that combines hot and cold composite curves in a single curve, also through the sum of their heat content in each temperature level. For zero value of the enthalpy (horizontal axis), the temperature of that point coincides with the pinch point. Using GCC, it is easier to observe that, in the temperature level above the pinch, the process just needs hot utility, whereas below the pinch the demand is for cold utility. In case of many utilities (multiple level utilities), it is possible to choose one of them, based on the closer temperature level to the demand, minimizing the heat transfer irreversibility. GCC also shows the use of steam turbine which is able to generate power using the degradation from high to low steam pressure. From this GCC, location of the pinch has been identified. At temperature 113^{o}C pinch temperature occurs at which there is no heat transfer. Composite curve shows the same value as in problem table algorithm. It gives simple framework for numerical analysis. The cold utility target - Hot utility target should equal the bottom line of the infeasible heat cascade which is 3535.35kW. These calculations provide useful cross-checks that the stream data and heat cascades have been evaluated correctly [7].

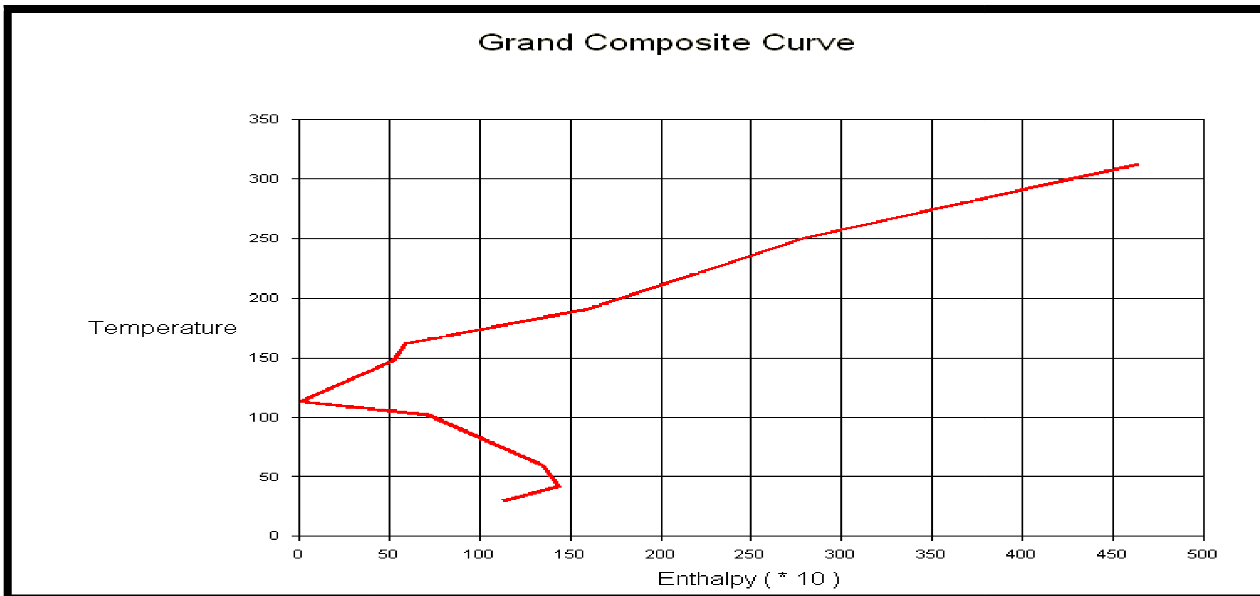


Figure-3
Grand Composite Curve

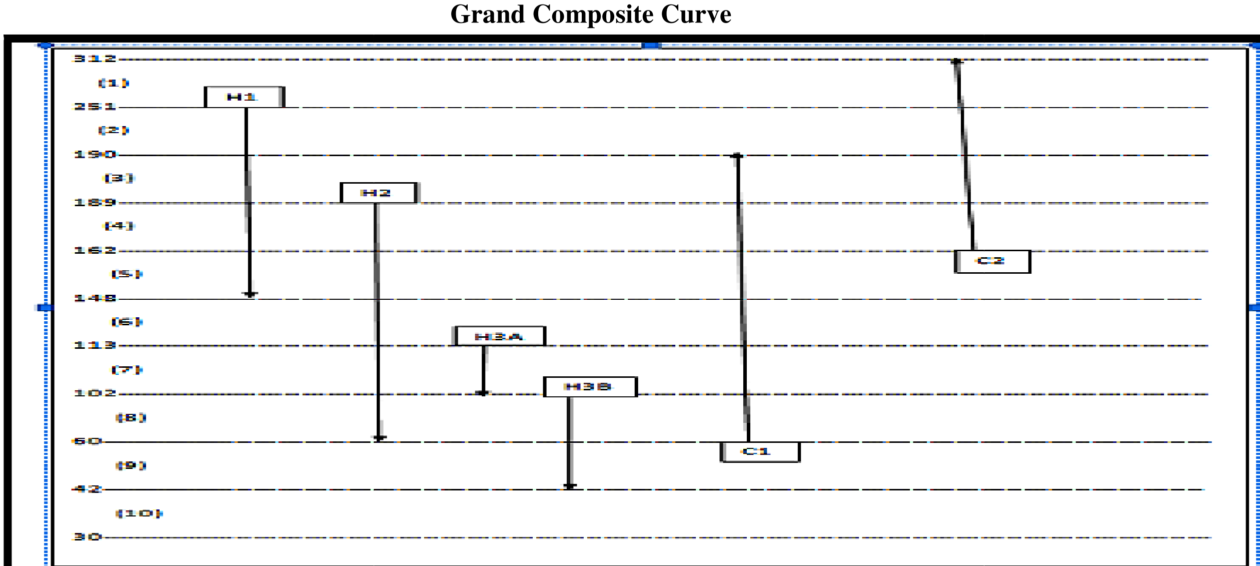


Figure-4
Stream and Temperature Intervals

The Problem Table: Composite curve requires a graph paper and scissor approach (for sliding the graph relative to one another which would be messy and imprecise. Therefore, we are using an algorithm for setting the target algebraically "The problem table". Following steps are to be done for the formation of the problem table [8]. i. Enthalpy balance interval was set up based on stream supply and target temperature shown in figure-4, ii. The same is to be done for hot and cold stream together to allow for the maximum possible amount of heat exchange with each temperature interval. iii. The modification needed here is that within any interval hot and cold stream are at least ΔT_{min} apart. iv. This is done by using shifted temperature which is set at $1/2\Delta T_{min}$ below hot stream temperature and $1/2\Delta T_{min}$ above cold stream temperature. v. Now set the shifted temperature of stream in a descending order from (higher to lower). vi. Setting up the interval in this way guarantees that full heat interchange within any interval is possible. vii. Each interval will have either a net surplus or net deficit of heat as dictated by enthalpy balance. viii. Enthalpy balance can easily be calculated for each according to $\Delta Hi = (S_{\mathrm{i}} - S_{i+1})(\sum \mathrm{CP_{hot}} \leq \sum \mathrm{CP_{cold}})_i$.

In table-6, it is indicated that last column indicates whether an interval is in heat surplus or heat deficit. Therefore it would be possible to produce a feasible network design based on the assumption that (a) all surplus intervals rejected heat to cold utility. (b) all deficit intervals took heat from hot utility. However this would not be very sensible because it would involve rejecting and accepting heat at inappropriate temperature. We have to exploit a key feature for the temperature intervals. From figure-5(a), it is clearly seen that from start there is negative flow of 1830kW, 1220 kW, 49.975 kW, 1046 kW.925 kW, 108.85 kW between interval 1-2,2-3,3-4,4-5 is thermodynamically infeasible. To make it just equal to zero, 4833.625kW of heat must be added from hot utility as shown in figure-5(b) and cascaded right through the system. The net result of this operation is that the maximum utilities requirements have been predicted (i.e.4834kW hot and 1253kW cold) [9].

Heat Exchanger Network Design

For designing a heat exchanger network the most useful representation is the "grid diagram" introduced by Linnhoff and flower. The process streams are drawn as horizontal lines with high temperature on the left and hot stream at the top.Cold stream are drawn at bottom and flow from right to left. Streams are represented by square boxes. Coolers and Heaters are represented by circles (C) and (H). Heat exchangers which are used to exchange the heat between the two process streams are marked by two circles and two circles are connected by vertical line. Two circles with vertical line connect the two streams between which heat is being exchanged[10].

Table-6
Problem Table Algorithm (Temperature intervals and heat loads for above process)

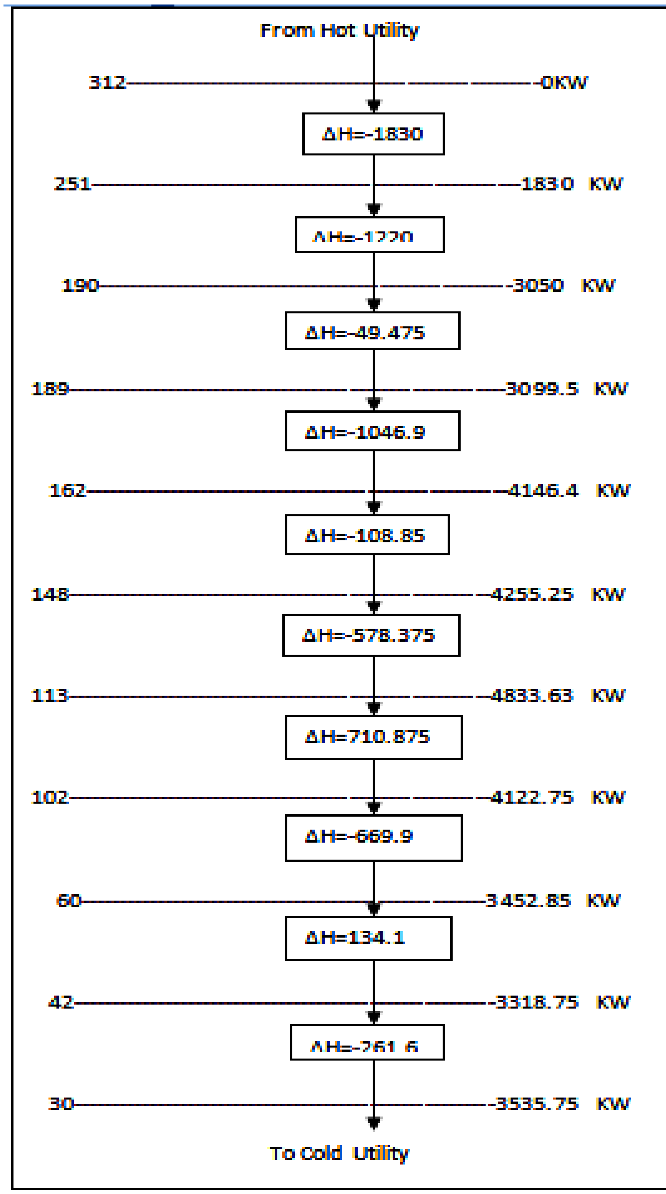
S_i (°C)	Interval No.	Si-Si+1(°C)	∑CPhot-∑CPcold (kW/°C)	ΔHI(kW)	Surplus or Deficit
S1=312					
	1	61	-30	-1830	Deficit
S2=251					
	2	61	-20	-1220	Deficit
S3=190					
	3	1	-49.475	-49.475	Deficit
S4=189					
	4	27	-38.775	-1046.925	Deficit
S5=162					
	5	14	-7.775	-108.85	Deficit
S6=148					
	6	35	-16.525	-578.375	Deficit
S7=113					
	7	11	64.625	710.875	Surplus
S8=102					
	8	42	15.95	669.9	Surplus
S9=60					
	9	18	7.45	134.1	Surplus
S10=42					
	10	12	-21.8	-261.6	Surplus
S11=30					

The grid is much easier to draw than a flow sheet, especially as heat exchangers can be placed in any order without redrawing the stream system. Also, the grid represents the countercurrent nature of the heat exchange, making it easier to check exchanger temperature feasibility. Finally, the pinch is easily represented in the grid, whereas it cannot be represented on the flow sheet. For drawing a grid diagram, above the pinch, $\sum CP_{hot} \leq \sum CP_{cold}$ below the pinch, $\sum CP_{hot} \geq \sum CP_{cold}$. Heat exchanger network design for given process is shown in figure-6. From above grid diagram, heat exchanger, heater, cooler summery is given in table-7 and 8. From grid diagram according to rules, we need total 4 heat exchangers, 2 heaters and 2 coolers. Initially when two heat exchangers have been used heat recovery was around 1640kW but by using 4 heat exchangers in this modified process, heat recovery is about 3860kW. And for modified process area requirement is 545.6m^2. Percentage reduction in hot utility and cold utility has been calculated as given below.

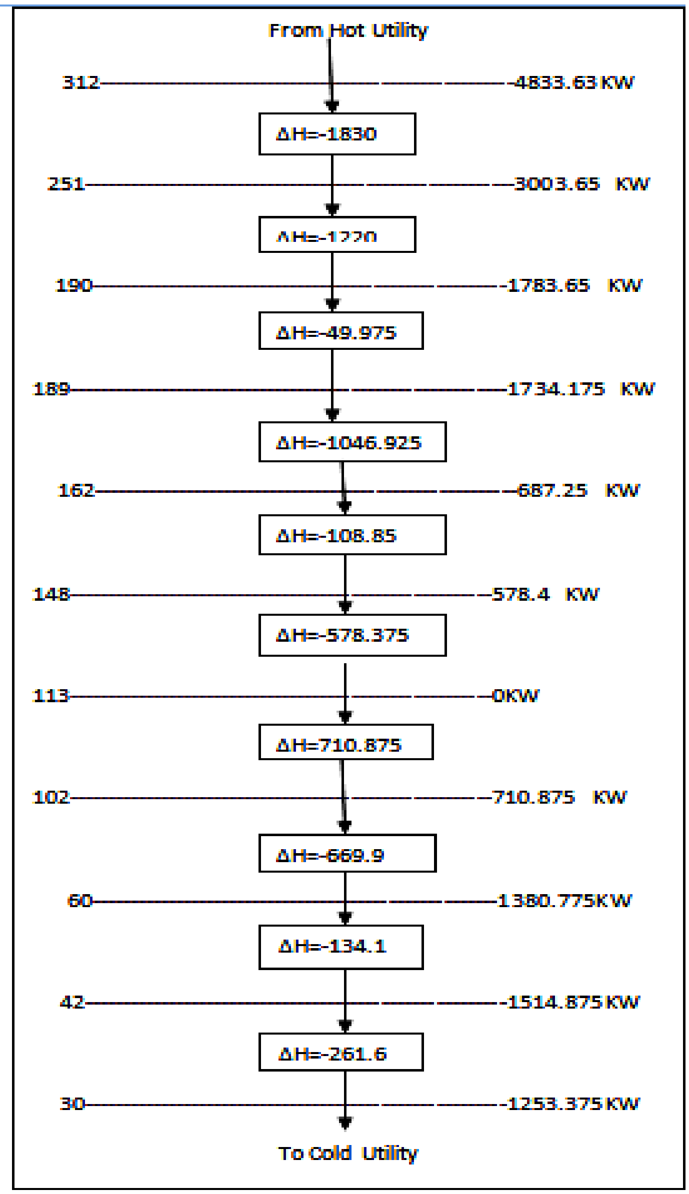
$$\%HUR = \left(\frac{6860-4833.63}{6860}\right)\times 100 = 29.5\%$$

$$\%CUR = \left(\frac{3360-1253}{3360}\right)\times 100 = 62.7\%$$

In these existing organic distillation processes, hot and cold utility requirement decreases but number of exchanger increases as shown in following grid figure-6.



(a)Infeasible



(b) Feasible

Figure-5
Infeasible and feasible heat cascades

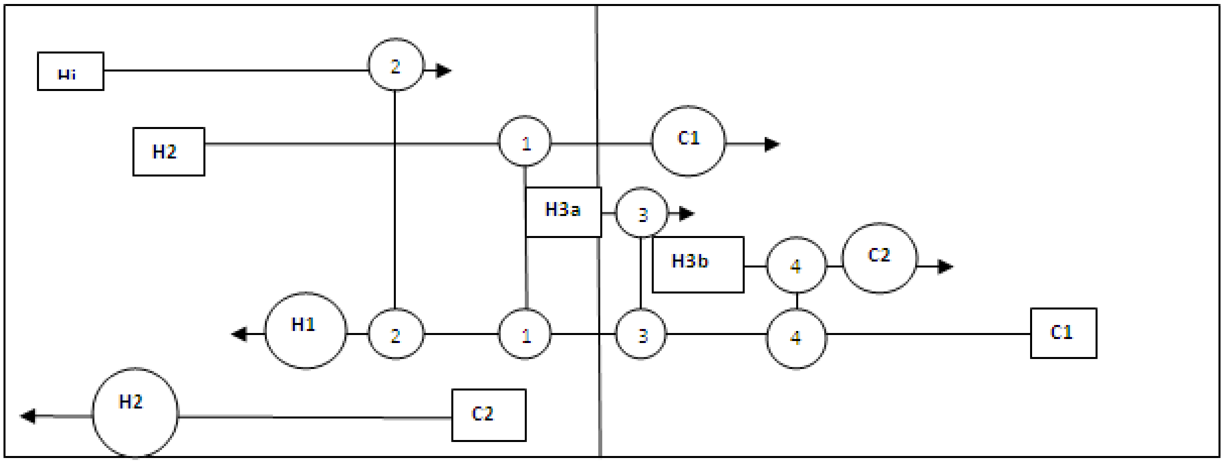


Figure-6
Combined heat exchanger grid diagram

Tabel-7
Heat Exchanger Summery

HEX No.	Thi (°C)	Tho (°C)	Tci (°C)	Tco (°C)	Heat Load (KW)	Area (m2)
1	199	123	103	133.4	760	158.4
2	261	158	103	174.6	1030	118.5
3	123	112	67.8	103	880	144.2
4	112	72	20	67.8	1195	124.5

Table-8
Heaters and Coolers summery

Heater No.	Tci(°C)	Tco (°C)	CPc	Heat Load(KW)
1	174	180	25	135
2	152	302	30	4497
Cooler No.	Thi(°C)	Tho(°C)	CPh	Heat Load(KW)
1	123	70	10	530
2	72	52	30	604

Table-9
Comparison of targets with current energy use

Situation	ΔTmin value (°C)	HU(KW)	CU(KW)	Heat recovery(KW)	Exchanger Area(m2)
Current	63	6860	3360	1640	130.7
Target	20	4834	1253	3865	545.6

Cost Estimation

The final information is to be collected is the cost of heating and cooling and the capital cost of new heat exchangers[11]. In this process, the hours worked per year are also needed i.e. here the figure is 5000. Heating is provided in the coal fired furnace whose mean temperature is approximately 400°C. The fuel costs $72/ton. Gross calorific value is 28.8 GJ/ton. Gross efficiency is 75%. Useful heat delivered cost is $3.33/GJ or $12 /MW. Cooling is much cheaper, cooling water is recirculated to a cooling tower which works in the range 25-35°C. Now exchanger cost is given by the following formula,

$$Exchanger\ Cost(\$) = d + aAb$$

The coefficients for heat exchanger cost are a=300, b=0.95, d=10000 for cooler d=5000and are given in dollar($).Total energy cost is given by,

$$TEC(\$) = \sum_{U=1}^{U} Q_U \times C_U$$

Hen Total Capital Cost Targeting

Here the targets for the minimum surface area (A_{min}) and the number of units (N_{min}) can be combined together with the heat exchanger cost law to determine the targets for the HEN capital cost $C_{HEN.}$ Capital cost is annualized using an annualization factor that takes into account interest payments on borrowed capital. The equation used for calculating the total capital cost and exchanger cost law is given below [12]

$$C(\$)_{HEN} = \left[N_{min}\left(a + b\left(\frac{A_{min}}{N_{min}}\right)^{c}\right)\right]_{AP} + \left[N_{min}\left(a + b\left(\frac{A_{min}}{N_{min}}\right)^{c}\right)\right]_{BP}$$

Where a, b, c are constants in exchanger cost law. N_{min} is calculated by,

$$N_{min} = \lfloor N_h + N_c + N_u \ - 1\rfloor_{AP} + \lfloor N_h + N_c + N_u \ - 1\rfloor_{BP}$$

N_{min} is found to be 8 and A_{min} is 546m^2. Consider carbon steel shell and tube heat exchangers. For this, constants are a=16000; b=3200; c=0.7. and $C(\$)_{HEN}$ is found to be 619626.5$. Generally installed cost can be considered 3.5 times the purchase cost. In this way heat exchanger capital cost and installed cost is to be calculated before and after heat integration.

Conclusion

In present study using pinch technology for heat integration of organic distillation unit, heat recovery of 3865kW occurs. Hot utility reduction is around 30% of the total heat load while the cold utility reduction is around 63% of the initial cold utility demand. Minimum number of units for heat transfer i.e. heater, cooler and heat exchangers increase, thereby capital cost increase is around 58% of the initial capital cost. As capital cost investment is one time investment, it is beneficial to use pinch technology. Table-8 shows the result for cost of heat exchanger before and after heat integration in the present study.

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