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RESEARCH ARTICLE

OPTIMUM HEAT INTEGRATION OF CUMENE PROCESS BY HEAT EXCHANGER NETWORK

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ABSTRACT

In process industries, heat exchanger networks represent an important part of the plant structure. The purpose of the networks is to maximize heat recovery, thereby lowering the overall plant costs. This work attempt to present process design of heat exchanger network using pinch technology. Pinch Technology concept based on thermodynamic principles, offers a systematic approach to optimum energy integration in a process. The improvements in the process associated with this technique are not due to the use of advanced unit operations, but to the generation of a heat integration scheme. One of the key advantages of pinch technology over conventional design methods is the ability to set an energy target for the design. The energy target is the minimum theoretical energy demand for the overall process. After introducing cumene manufacturing process, its thermodynamic properties-molar flow rate and temperature values have been used for material balances and energy balances which further results in total energy required for the process. After getting total energy for the process we have calculated heat recovery by finding pinch point with the help of composite curve. Finally we have proposed a heat exchanger network design which shows maximum of heat recovery from different process streams.

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INTRODUCTION

The world is facing steady fuel-price challenge day-by-day. In the "World Energy Outlook 2008" report, the International Energy Agency (IEA) predicts world energy demand to increase by 45% over the next 20 years. They also predict the supply of fossil fuels will not be able to meet this demand, even when taking new, undiscovered fields into account. More and more governments around the world will probably start charging industries for emitting CO_2, with emission credits becoming more and more expensive. The result of all this will undoubtedly be increasing energy prices; just how much is hard to predict. In 2007, the IEA predicted oil prices to stay at 50-55 $ per barrel until 2030. A year later, in June 2008, it peaked at 147 $ per barrel and at the time of writing it is above 100 $ per barrel. Such a unsteadiness increases the instability for the fuel-prices. The solution is to save energy is to generate the same. But to save the energy needs a systematic approach of process design itself. Among the top commodity chemicals, taking about 7 – 8 % from the total worldwide propylene consumption, today cumene is used almost exclusively for manufacturing phenol and acetone.

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Cumene (isopropyl benzene) is produced by reacting propylene with benzene. During World War II, cumene was used as an octane enhancer for piston engine aircraft fuel. Presently, most of the worldwide supply of cumene is used as a raw material for phenol production. Typically, cumene is produced at the same facility that manufactures phenol. The cumene manufacturing is shown by process design hierarchy such as "onion diagram" shown below. The design of a process starts with the reactors (in the "core" of the onion). If feeds, products, recycle concentrations and flow rates are known, the separators (the second layer of the onion) can be designed. The basic process heat and material balance is now in place, and the heat exchanger network (the third layer) can be designed. The remaining heating and cooling duties are handled by the utility system (the fourth layer). The process utility system may be a part of a centralized site-wide utility system. Those heating and cooling duties that cannot be satisfied by heat recovery, dictate the need for external heating and cooling utilities (furnace heating, use of steam, steam generation, cooling water, air-cooling or refrigeration). Thus, utility selection and design follows the design of the heat recovery system. The selection and design of the utilities is made more complex by the fact that the process will most likely operate within the context of a site comprising a number of different processes that are all connected to a common utility system (Linnhoff and Vredevelt, 1984 and Linnhoff March, 1998).

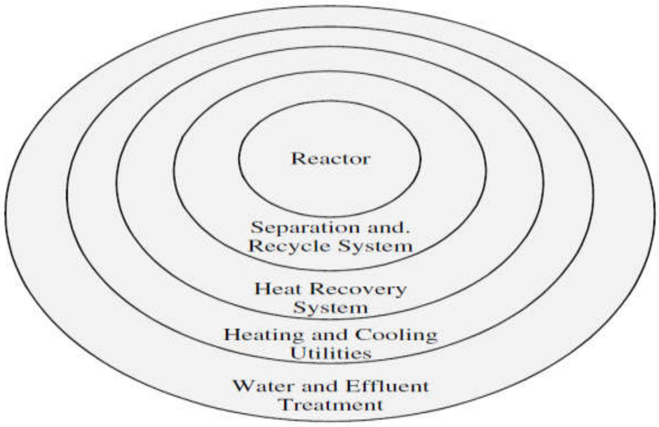

Fig. 1. The Onion model of process design

Energy saving is an important issue in the process industry. A process integration analysis is a highly suitable measure in this respect, since it relates the production, the energy efficiencies and the construction costs. All of these questions and more can be answered with a full understanding of Pinch Technology and an awareness of the available tools for applying it in practical way. This article aims to provide the basic knowledge of concepts in pinch technology and how they have been be applied across a wide range of process industries (Robin Smith, 2005).

Pinch Technology

Pinch Technology Concept

Based on thermodynamic principles, pinch technology offers a systematic approach to optimum energy integration in a process. The improvements in the process associated with this technique are not due to the use of advanced unit operations, but to the generation of a heat integration scheme. One of the key advantages of pinch technology over conventional design methods is the ability to set an energy target for the design. The energy target is the minimum theoretical energy demand for the overall process. The principal objective of this technology is to match cold and hot process streams with a network of exchangers so that demands for externally supplied utilities are minimized. Pinch technology establishes a temperature difference, designated as the pinch point that separates the overall operating temperature region observed in the process into two temperature regions. Once a pinch point has been established, heat from external sources must be supplied to the process only at temperatures above the pinch and removed from the process by cooling media only at temperatures below the pinch. Such a methodology will maximize the heat recovery in the process with the establishment of a heat exchanger network based on pinch analysis principles.

The best design for an energy-efficient heat exchanger network will result in a trade-off between the energy recovered and the capital costs involved in this energy recovery. The success of pinch technology has led to more inclusive ideas of process integration in which chemical processes are examined for both mass and energy efficiency. Even though process integration is a relatively new technology, its importance in process design is continuing to grow as processes become more complex. As noted above, one concept of pinch analysis is to set energy targets prior to the design of the heat exchanger network. Targets can be set for the heat exchanger network without actually having to complete the design. Energy targets can also be set for the utility heat duties at different temperature levels such as refrigeration and steam heat supply levels. Pinch analysis provides the thermodynamic rules to ensure that the energy targets are achieved during the heat exchanger network design (Richard Turton *et al*., 2012)

Pinch Technology Analysis

The starting point for a pinch technology analysis is to identify in the process of interest all the process streams that need to be heated and all those that need to be cooled. This means identifying the streams, their flow rates and thermal properties, phase changes, and the temperature ranges through which they must be heated or cooled. This can be accomplished after mass balances have been performed and temperatures and pressures have been established for the process streams. The material balance has been carried out for each stream at 97 % conversion. Accordingly mass flow rates have been calculated for each stream. These mass flow rates values have been further used for the calculation of energy balances. The known as well as unknown temperature values calculated by energy balance are used for hot and cold streams named as hot and cold stream data. Energy quantities can be calculated conveniently by using a simulation program or by traditional thermodynamic calculations. Some heat duties may not be included in the network analysis because they are handled independently of the integration. For example, distillation column reboiler heating and condenser cooling may be treated independently of the rest of the heat duties. However, such independent duties should always be considered for inclusion in the network. All the process streams that are to be heated, their temperatures, and enthalpy change rates corresponding to their respective temperature changes or phase changes are then tabulated. The enthalpy change rate for *each* stream is obtained from

$$\Delta H = mCp\,\Delta T = CP\Delta T$$

where

ΔH the enthalpy change rate,
$\dot{m}$ the mass flow rate,
Cp the heat capacity,
ΔT the temperature change in the stream,
CP the heat capacity rate defined as the $\dot{m}\,Cp$ product.

The enthalpy change rates are then added over each temperature interval that includes one or more of the streams to be heated. The resulting values allow plotting of the temperature versus enthalpy rate to provide a composite curve of all the streams that require a heat source. The same information and procedures are followed to develop a composite curve of the streams to be cooled. The resulting diagram, shown in Fig.2, is designated as a composite diagram for the heat integration problem. The actual steps involved in preparing such a diagram are presented (Richard Turton *et al*., 2012). It must be recognized that while each temperature is a fixed value on the vertical axis, enthalpy change rates are relative quantities. Enthalpy changes rather than absolute enthalpies are calculated via thermodynamic methods. Thus, the horizontal location of a composite line on the diagram is arbitrarily fixed. For the purposes of pinch technology analysis, the composite curve for streams to be cooled is located so as to be to the left, at every temperature, of the composite curve for those streams to be heated. Fixing the location of the composite curves with respect to one another with the use of a preselected value of $\Delta T_{\min}$ completes the composite diagram. The location of $\Delta T_{\min}$ on the composite diagram is where the two curves most closely approach each other in temperature, when measured in a vertical direction. On the first plotting of these curves, the vertical distance will rarely equal the preselected $\Delta T_{\min}$. This deficiency is remedied by moving one of the two curves horizontally until the distance of closest vertical approach matches the preselected $\Delta T_{\min}$. This can be done graphically or by calculation.

composite curves can be used to evaluate the overall tradeoff between energy and capital costs. An increase in $\Delta T_{\min}$ causes the energy costs to increase, but also provides larger driving forces for heat transfer and accompanying reduced capital costs (Richard Turton *et al*., 2012 and Ray Sinnott *et al*., 2009).

Identifying an Optimal Heat Exchange Network

There is not a unique network for any but a two-stream heat exchange problem. So the design engineer needs both insight and creativity, in addition to described procedures that identify an appropriate network among the many possibilities. A network is developed one section at a time. Since the minimum number of heat exchangers already has been established, the task now becomes one of identifying which streams go to which exchangers. For each heat exchanger, a heat balance must be satisfied. If it is assumed that there are negligible heat gains or losses from the exchanger, the heat balance equation is

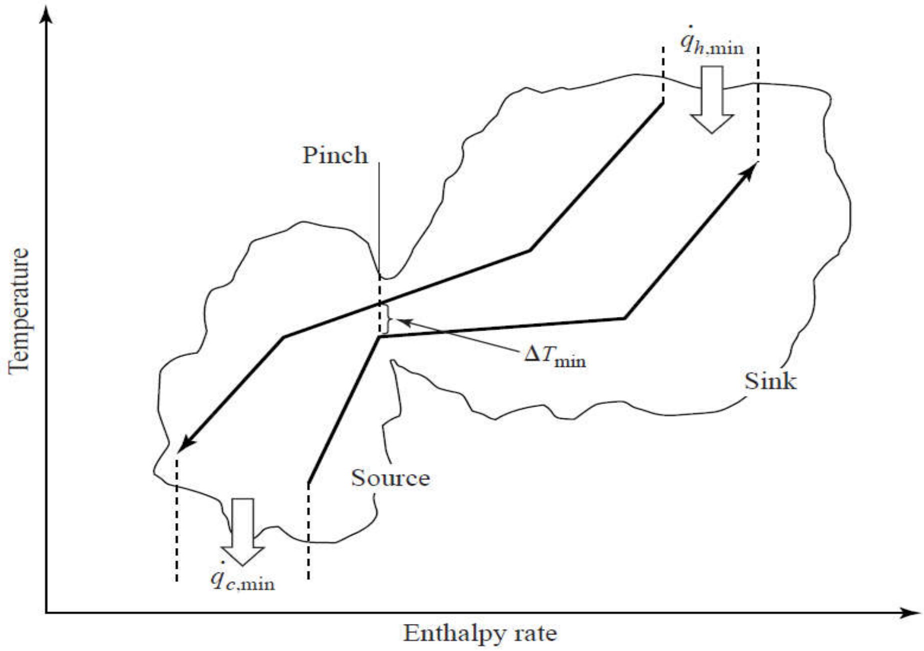

Fig. 2. Composite diagram prepared for pinch technology analysis

The optimum value for $\Delta T_{\min}$ is generally in the range of 3 to 40°C for heat exchange networks, but is unique for each network and needs to be established before the pinch technology analysis is completed. If no cooling media are required below about 10°C, the optimum $\Delta T_{\min}$ is often in the range of 10 to 40°C. For a given $\Delta T_{\min}$, the composite curves define the utility heating and cooling duties. The composite curves show the overall profiles of heat availability and heat demand in the process over the entire temperature range. These curves represent the cumulative heat sources and heat sinks in the process. The overlap between the two composite curves indicates the maximum quantity of heat recovery that is possible within the process. The overshoot of the hot composite curve represents the minimum quantity of external cooling $\dot{q}_{c,\min}$ required, and the overshoot of the cold composite curve represents the minimum quantity of external heating $\dot{q}_{h,\min}$ required for the process. Note that the

$$\Delta H = 0 = (CP(T_{out} - T_{in}))_{\text{hot stream}} + (CP(T_{out} - T_{in}))_{\text{cold stream}}$$

Some specific guidelines useful in finding good heat exchange matches are given below:

1. At the pinch point, each stream that is to be heated must enter or leave an exchanger at the pinch point, cold composite temperature; and each stream that is to be cooled must enter or leave an exchanger at the pinch point, hot composite temperature.
2. Start the analysis of exchangers in the sink and source sections at the pinch point where all temperatures are fixed.
3. A point of discontinuity in a composite curve indicates the addition or removal of a stream, or the onset of a phase change. The stream that is being added or removed must enter or leave an exchanger at the temperature where the discontinuity occurs.

4. If there are only two streams in a section, they both go to the one exchanger that is reserved for the section.
5. If there are three streams in a section, the stream with the largest change in enthalpy should be split across two exchangers to satisfy the heat duties for each of the other two streams.
6. If there are four streams in a section, three heat exchangers will be required. If three streams are either heated or cooled, then the fourth stream is split into three flows to satisfy the heat duties from the other three streams. If there are two streams that are to be heated and two streams that are to be cooled, a convenient way to allocate the streams to exchangers is to prepare a new composite diagram and use this to make the allocations.
7. If there are more than four streams in a section, attempt to follow guideline 6. The use of a computer-based algorithm is recommended for this more complicated case.
8. If the matches between heat exchange duties result in more than the minimum number of exchangers being required, try other matches. Look for loops and eliminate them.
9. If a discontinuity occurs in a process stream curve within a utility section, it may be possible by means of the adjacent process section to meet the duty of the stream by leaving the curve at the discontinuity and still not violate the ΔT_{min}. Doing so reduces the required number of exchangers by 1 without changing the utility requirements and many times is an economical choice (Ray Sinnott *et al*., 2009)

Cumene Process

Process Description

The PFD for the cumene production process is given in Figure 4. The reactants are fed from their respective storage tanks. After being pumped up to the required pressure (dictated by catalyst operating conditions), the reactants are mixed, vaporized, and heated to the temperature required by the catalyst in the fired heater. The shell-and-tube reactor converts the reactants to desired and undesired products as per the above reactions. The exothermic heat of reaction is removed by producing high pressure steam from boiler feed water in the reactor. The stream leaving the reactor enters the flash unit, which consists of a heat exchanger and a flash drum. The flash unit is used to separate the C3 impurities, which are used as fuel for a furnace in another on-site process. The liquid stream from the flash drum is sent to the first distillation column, which separates benzene for recycle. The second distillation column purifies cumene from the p-di isopropyl benzene impurity. Currently, the waste p-DIPB is used as fuel for a furnace. The pressure of both distillation columns is determined by the pressure in the flash drum, i.e., there are no pressure reduction valves downstream of the flash drum (Shenoy, 1985).

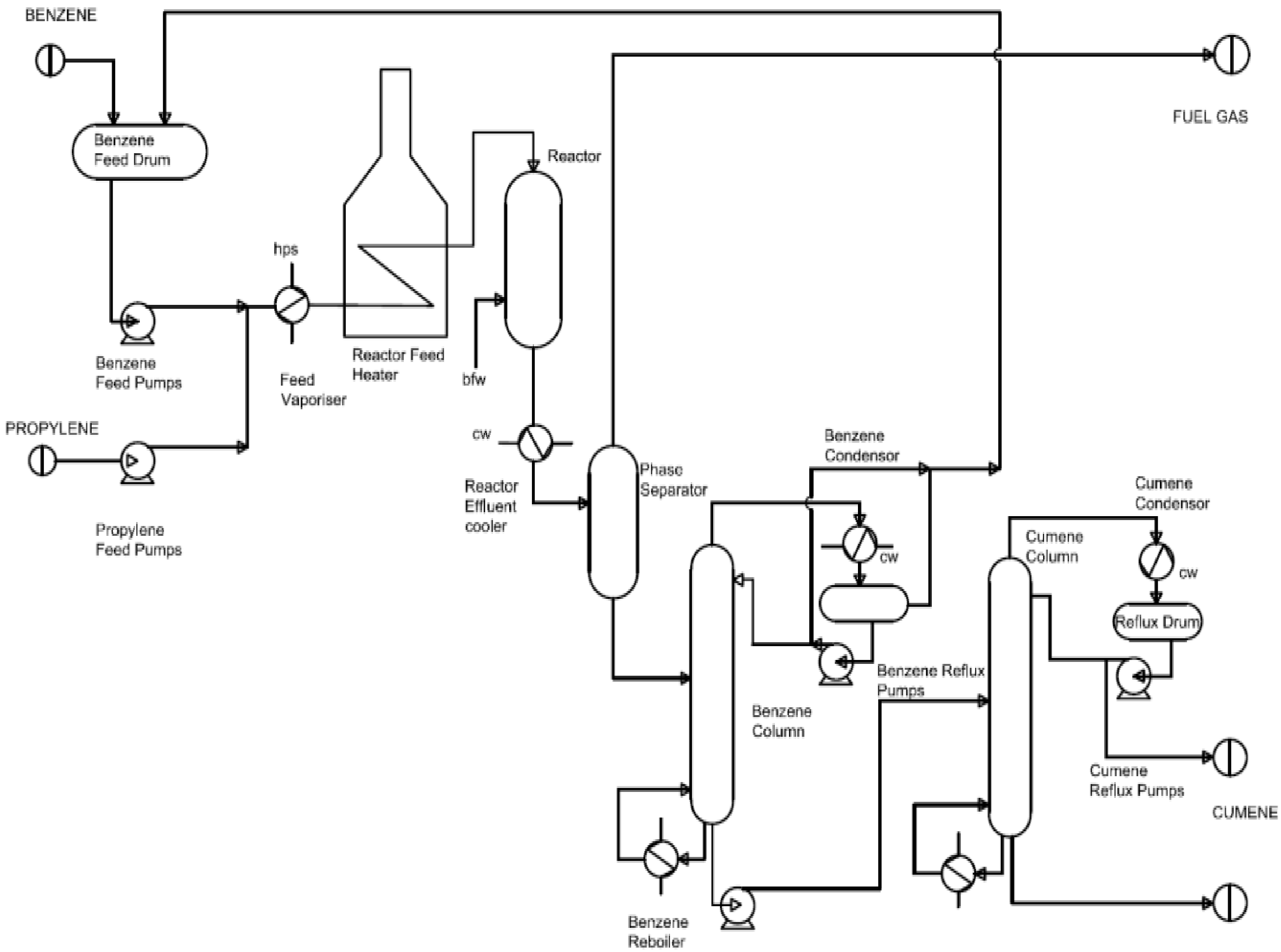

Fig. 3. Process Flow Diagram for Cumene manufacturing

Cumene Production Reactions

The reactions for cumene production from benzene and propylene are as follows:

$$\underset{\text{Propylene}}{C_3H_6} + \underset{\text{Benzene}}{C_6H_6} \rightarrow \underset{\text{Cumene}}{C_6H_5\text{-}C_3H_7}$$

$$\underset{\text{Propylene}}{C_3H_6} + \underset{\text{Cumene}}{C_6H_5\text{-}C_3H_7} \rightarrow \underset{p\text{- diisopropyl benzene}}{C_3H_7\text{-}C_6H_4\text{-}C_3H_7}$$

Simulation

In process industries, during operation of any heat exchanger network (HEN), the major aim is to focus on the best performance of the network. Frequently one encounters problems that degrade the HEN performance, like heat exchanger fouling, leakage in tubes, changes in process stream conditions (flow rate, temperature), frequent changes in arrangement of utility streams to optimize heat recovery in the network, shutdown of heat exchangers for maintenance, etc. Since the changes can take place in any of the heat exchangers in the network, a complete analysis of the network in an integrated approach is required (Morgan, 1992 and Rossiter *et al*., 1993). In order to handle these issues a good understanding of modeling and simulation of HENs in a simultaneous approach is necessary (Smith and Linnhoff, 1988). The simulation software such as CHEMCAD after trying the iterations results in the following optimum parameters which have seems to be good to propose the design. The temperature values mentioned in the streams tables have been further used for the calculation of cold and hot stream data.

Hot and cold stream data

A review of the stream information in the table shows that stream 1and 2 is to be heated over the temperature interval from 150°C to 236.6 °C and 90 °C to 198 °C respectively. The minimum energy requirements are Q_h = 9143.4 kW and Q_c = 11 063.9 kW. By taking advantage only from process/process heat exchange a saving in heat up to 43% in hot utility and 40% in cold utility can be achieved. Since the reaction is highly exothermal, we expect possible export of energy too. The examination of composite curves indicates that the pinch is situated between the reactor and the benzene column. The largest energy consumer is by far the benzene column, with reboiler and condenser duties of more than 7.5 MW. By exchange between the reactor outlet and inlet streams considerable energy is saved, but only a modest amount of about 1 MW can be used for steam generation (Alexandre C. Dimian and Costin Sorin Bildea, 2008). The target temperatures of the hot streams (condensers) of the columns C - 2 and C - 3 shows that most of the energy is rejected in the environment, at a temperature slightly below 100 °C. The simulation shows that about 2660 kW may be saved by using the condenser of (C - 2B) as the reboiler for (C - 2A). Since the net distillate flow of C - 2A is larger, a supplementary reboiler duty of about 560 kW is necessary. The net hot - utility consumption of benzene distillation drops from 7870 to 3794 kW, representing a saving of 51.8 %. Furthermore, by slightly increasing the pressure in the column (C - 3) allows the generation of low - pressure steam. For both columns (C - 2) and (C - 3) the hot utility can be ensured by Dowtherm A or another similar thermal fluid (Shenoy, 1985 and Obata and Shibuya, 1993). The minimum energy requirements are now Q h = 5330 kW and Q c = 8005 kW, much lower than before. But the salient element is that an amount of 2000 kW can be exported as process steam with a pressure of about 5 bar.

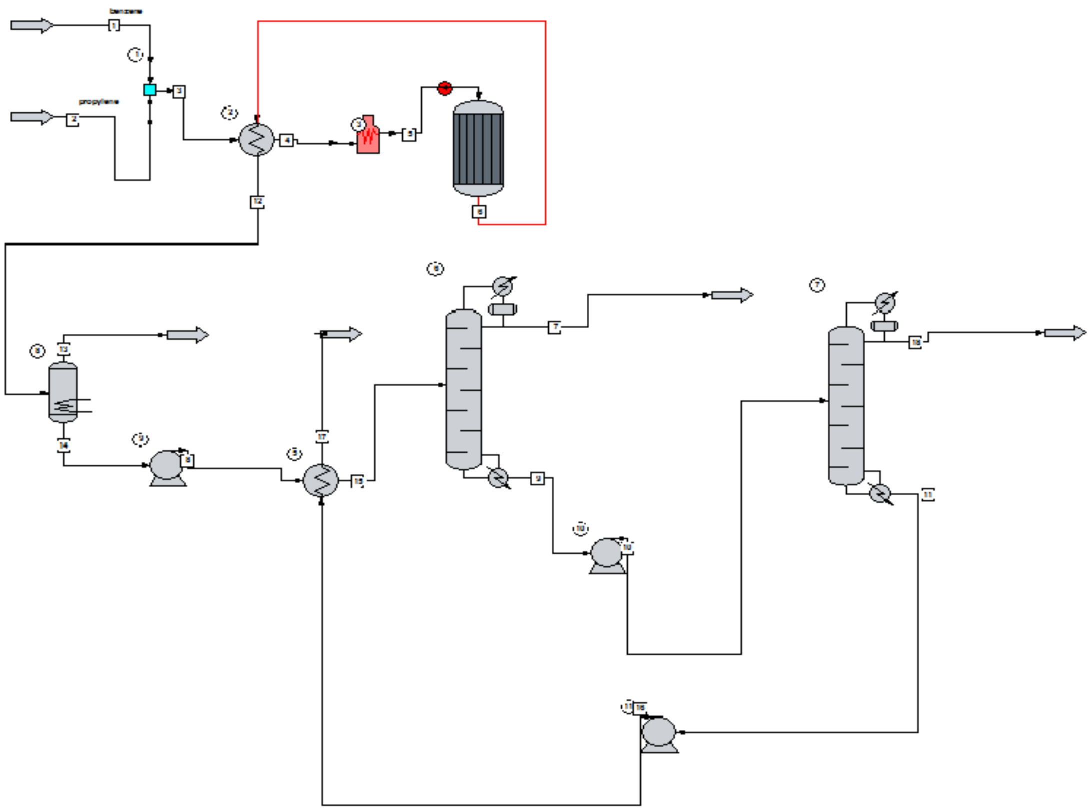

Fig. 4. Simulated PFD for Cumene manufacturing process

Simulation results

Table 7.1 Simulation result

Stream No.	1	2	3	4
Stream Name	benzene	Propylene	mixed feed-1	mixed feed-2
Temperature ^{0}C	25.0000*	25.0000*	72.0000*	72
Pressure bar	35.5000*	35.5000*	35.5	35.3
Enthalpy kJ/h	5.16E+06	4.22E+05	6.85E+06	5.66E+06
Vapor mole fraction	0	0	0	0
Total kmol/h	107.5269	117.902	223.9869	223.99
Total kg/h	8387.098	4963.68	17564.4754	17564
Total std L ft^3/hr	327.5076	315.0147	642.5224	642.5223
Total std V scfh	83109.98	87577.14	170687.13	170687.11
Flow rates in kg/h				
Benzene	107.5269	0	8201.8989	221.77
Propylene	0	112.007	4650.9014	0
M-Diisopropylbenzene	0	0	0	0
Cumene	0	0	0	2.2127
Propane	0	5.8951	3000	5.3

Stream No.	5	6	7	8
Stream Name	mixed feed-3	mixed feed-4	mixed feed-5	mixed feed-6
Temperature ^{0}C	170	200	200	236
Pressure bar	34	33.8	35.2	35.2
Enthalpy kJ/h	1.41E+07	2.29E+07	1.35E+07	4.35E+07
Vapor mole fraction	1	1	1	1
Total kmol/h	117.9022	341.88914	341.88914	231.001992
Total kg/h	4963.683	22527.6	22527.6	22527.6
Total std L ft^3/hr	642.5223	1186.0301	642.5224	400.642
Total std V scfh	170687.11	257549.36	170687.13	178954.92
Flow rates in kg/h				
Benzene	0	221.7742	221.7742	114.2473
Propylene	112.0071	112.0071	112.0071	1.1197
M-Diisopropylbenzene	0	0	0	3.360215
Cumene	0	2.21274	2.21274	106.379407
Propane	5.8951	5.8951	5.8951	5.8915

Stream No.	9	10	11	12
Stream Name	mixed feed-7	mixed feed-8	mixed feed-9	mixed feed-10
Temperature ^{0}C	170	120	150	34.3
Pressure bar	35.5000*	35.5000*	33.8	12
Enthalpy kJ/h	3.26E+06	2.45E+05	3.11E+06	1.63E+06
Vapor mole fraction	0	0	0	0
Total kmol/h	231.001992	231.001992	231.001992	7.01507
Total kg/h	25527.6	25527.6	25527.6	306.42
Total std L ft^3/hr	327.5076	315.0147	642.5224	642.5223
Total std V scfh	83109.98	87577.14	170687.13	170687.11
Flow rates in kg/h				
Benzene	114.2473	114.2473	114.2473	0
Propylene	1.11997	1.1197	1.1197	1.11997
M-Diisopropylbenzene	3.360215	3.360215	3.360215	0
Cumene	106.379407	106.379407	106.379407	0
Propane	5.8951	5.8915	5.8915	5.8951

Stream No.	13	14	15	16
Stream Name	Fuel Gas	mixed feed-12	mixed feed-13	mixed feed-14
Temperature ^{0}C	90	82.4	163.1	162.1
Pressure bar	12.1	1.1	1.3	0.4
Enthalpy kJ/h	2.14E+06	1.87E+05	3.24E+06	3.24E+06
Vapor mole fraction	0	0	0	0
Total kmol/h	223.986922	116.46	107.526882	104.270927
Total kg/h	22221.17	9177.377	13044.35	12516.99
Total std L ft3/hr	327.5076	315.0147	642.5224	642.5223
Total std V scfh	83109.98	87577.14	170687.13	170687.11
Flow rates in kg/h				
Benzene	114.2473	114.2473	0	0
Propylene	0	0	0	0
M-Diisopropylbenzene	3.360215	0	3.360215	0.10426
Cumene	106.379407	2.2127	104.166667	104.166667
Propane	0	0	0	0

Table 8.1 Hot and cold stream data

Cold streams				Hot streams			
Name	Ts (°C)	T_t (°C)	Duty (kW)	Name	T s (°C)	T_t (°C)	Duty (kW)
Reactor in	72	170	3657.5	Reactor out	236.6	150	-3781.4
				Feedc-2	198	90	-4116.9
Reb c-1	199	198	2318.1	Cond -1	36.3	35.3	-399
Reb c-2	162	163	7870.8	Cond -2	83.4	82.4	-7572.9
Reb c-3	161	162	1979	Cond -3	99.3	98.3	-2413.6
Total			15826.6				-18283

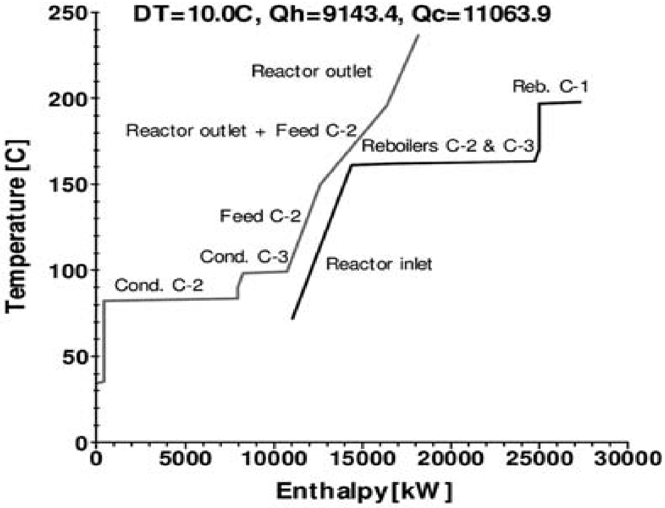

Fig. 5. Composite curve before pinch

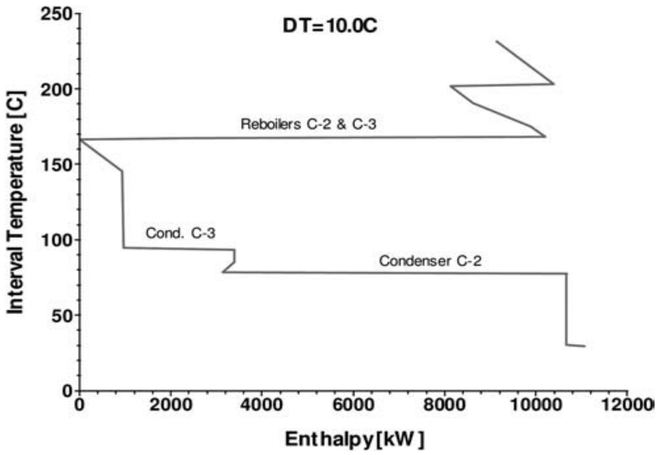

Fig. 6. Grand composite curve before pinch

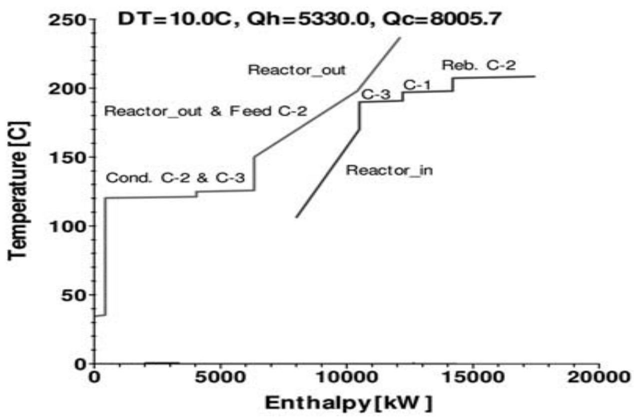

Fig. 7. composite curve after pinch

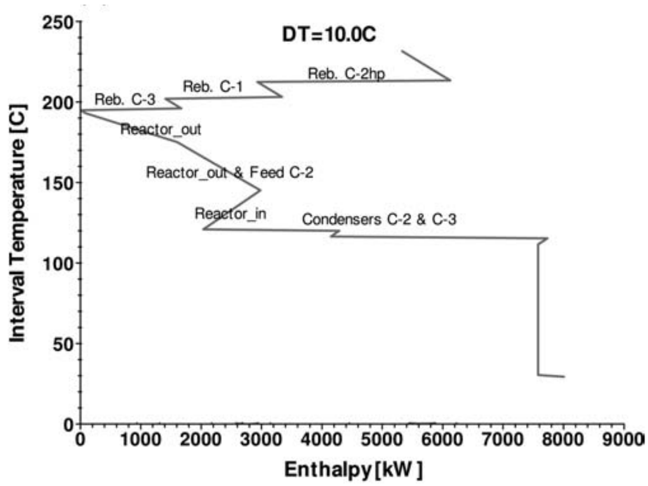

Fig. 8. Grand composite curve after pinch

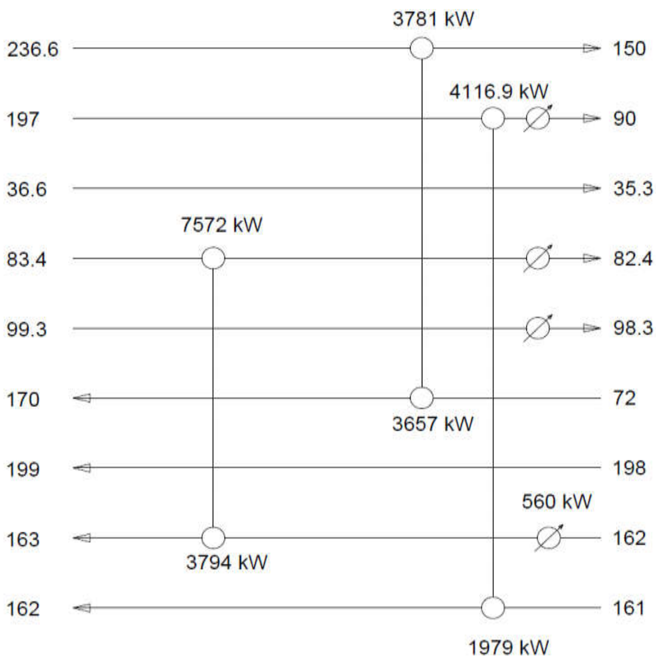

Fig. 9. Heat exchanger Network diagram

RESULTS

Proposed Heat exchanger network diagram

Cost Estimation

The cost estimation calculation is performed on the basis of 1 ton consumption. The interest and depreciation cost is considered as 30 % of capital tabulated here (Warren D. Seider *et al*., 2004).

Table 9.1. Cost of Cumene Production per Metric Ton Cumene

Factor	Consumption, t	Typical cost, $	Manufacturing cost, %
benzene charge	(0.651)	226	53
propylene charge, 95% purity	0.351	167	39
by-product (credit at $250/t)	(0.002) (1) (10)	(1)	(10)
Utilities			
heat and electricity	1600 MJ[a]	5	1
steam credit	(1.1) (6) (1)	(6)	(1)
catalyst and chemicals		3	1
labor, insurance, maintenance, etc		4	1
interest and depreciation at 30% of capital		37	7
Total		430	

[a]To convert MJ/t cumene to Btu/lb, divide by 2.324

Heat recovery cost

Before application of pinch

Consider, the cost of the hot oil is $2.25/GJ. The cooling utility is cooling water available at 10°C with an allowable temperature rise of 10°C and a cost of $0.25/GJ. The minimum energy requirements are Q_h = 9143.4 kW and Q_c = 11 063.9 kW. Therefore, Hot utility cost = (2.25 / 10^6) 9143.4 = $ 0.02057 \sec. Cold utility cost =(0.25 / 10^6)11063.9 =$ 0.002765 /sec. Total utility cost= $ 0.023335 \sec =0.023335 $\times$ 330 $\times$ 24$\times$ 60 $\times$ 60 = 6.6 $\times$ 10^5 /yr

Consider, $ 1 = ₹ 64 which gives total utility cost = ₹ 422.4 $\times$ 10^5 /yr

After application of pinch

The minimum energy requirements are Q_h = 5330 kW and Q_c = 8005kW

Therefore, Hot utility cost = (2.25 / 10^6) 5330 = $ 0.01199 \sec

Cold utility cost = (0.25 / 10^6)8005 =$ 0.00200 /sec

Total utility cost= $ 0.01399 /sec =0.01399 $\times$ 330$\times$ 24 $\times$ 60 $\times$ 60 = 3.9 $\times$ 10^5 /yr

Consider, $ 1 = ₹ 64 which gives total utility cost = ₹ 249.6 $\times$ 10^5 /yr

DISCUSSION

Pinch analysis can help reduce operating costs and capital costs while increasing plant capacity and energy efficiency. Pinch technology also eliminates costly mistakes of purchasing equipment too large, too small, or not necessary. With the application of pinch technology, both capital investment and operating cost can be reduced. Emissions can be minimized and throughput maximized. Many plants can be benefiting from a pinch analysis. It can be applied at both industrial and research. In our case the results obtained for total utility cost and cost of energy saved per year are

Before Application of pinchAfter application of pinch

Total utility cost = ₹ 422.4 $\times$ 10^5 /yr Total utility cost = ₹ 249.6 $\times$ 10^5 /yr

The amount of energy saved per year (cost) = ₹ 172.8 $\times$ 10^5 /yr

The design and optimization of HENs has been extensively studied over years and significant progress has been achieved in the development of robust methods for design of cost-optimal networks. A thorough review of these methodologies is presented. The major complexities in HEN synthesis are handling the combinatorial nature of the problem and finding a feasible and optimum solution using simultaneous synthesis methods. HEN design and optimization is largely handled using traditional optimization methods, which are robust and rigorous but are often restricted in handling large-scale problems due to the size of the problem and exponentially growing demand for computational work. Recently, non-traditional optimization methods (evolutionary methods) like genetic algorithm (GA), differential evolution have been catching up the popularity for handling large-scale combinatorial problems. These algorithms have been extensively applied to HEN synthesis problems which promised simple, fast and robust optimization solutions.

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