

RETROFIT OF PALM OIL REFINERY TO IMPROVE ENERGY CONSUMPTION AND REDUCE FOULING COSTS

by

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ABSTRACT

This paper presents the results of a detailed process integration retrofit study done on a palm oil refinery. The study, which is based on pinch technology is aimed at optimizing the energy consumption and recovering losses from the process streams of the existing plant. Bulk of the losses can be attributed to suboptimal heat recovery network structure and heat exchanger fouling effects. Previous pinch-based studies on fouling costs reductions were focused on grassroots designs. Retrofit studies reported for palm oil refineries were slightly different in approach and did not take into account of the effects of heat exchanger fouling. The present study demonstrates that notwithstanding the existing integration scheme employed in the refinery, energy savings of 71% in hot utility and 52% in cold utility with a payback period of less than 2 years is possible. The proposed scheme eliminates the need for routine maintenance downtime dedicated for cleaning of fouled exchangers. As a result, energy consumption is further improved and productivity increased.

INTRODUCTION

Pinch Analysis has come a long way from being a powerful and efficient energy saving tool to addressing issues such as fouling control. Fouling phenomenon has caused billions of dollars in losses from add-on capital investments, excessive utility consumption and maintenance downtime. An annual survey done in 1979 indicate total annual expenditures related to heat exchanger fouling in oil refineries exceed \$4.5 billion (Kotjabasakis, 1987). The fact that refineries comprise a small fraction of chemical industries in general underlines the importance of fouling control.

In Malaysia, although no statistics on fouling related expenses are available, similar problems exist particularly in petrochemical and palm oil refineries. A lot of money is spent on heat exchanger overdesign and purchase of additives for fouling reductions. Maintenance downtime and loss of productivity has resulted in significant losses. Others dedicate a certain amount of budget for research purposes.

This paper presents the results of a process integration study completed on a palm oil refinery using the techniques of Pinch Technology. The aim of the study is to retrofit the existing refinery to reduce energy consumption and control fouling problems hence the fouling costs. The analysis covers the refinery degumming, bleaching and

deodorization sections. In order to preserve the confidentiality of the source, neither the plant structure nor the company¹ involved in the study is revealed. The paper analyses the present plant situations, the consumption, extent of fouling problems and the various losses due to inefficiencies in the present heat recovery scheme. Retrofit of the heat exchanger network is performed with the philosophy of retaining the present structure as much as possible while heat is being recovered and fouling costs reduced. At the end of the study, the economics of the project is assessed.

PROCESS DESCRIPTION

Retrofit of the palm oil refinery plant covers 3 main processes. The first is degumming process which involves mixing the crude oil with phosphoric acid acting as a solvent. This is done at a temperature in excess of 100 C in order to rid the oil of gums (phosphatide compounds). The resulting mixture forms two phases which separate by gravity difference. Next, bleaching earth is used to remove color pigments from the degummed oil. This is done under vacuum at 120 C. The slurry product from the bleaching section proceeds through a series of filter to remove bleaching earth. The final refining step involves removal of free fatty acids, aldehyde and ketone compounds from the bleached oil by steam stripping. The deodorizer column operates under high vacuum and temperature in the range of 240-270 C. This is by far the most energy intensive section of the refining process. The product from this section is known as the refined, bleached and deodorized oil (RBDPO).

PROCESS STREAM DATA

Streams having potential for heat interchange are grouped together for the purpose of the analysis. These streams are classified into hot streams (that requires cooling) and cold streams (that requires heating). Each streams are designated according to its function and nature in the process. The stream data is extracted from the nominal plant heat and material balance.

During stream data extraction, stream enthalpy refers to the enthalpy excess of the hot streams and the enthalpy requirement of the cold streams.

Stream No & Type	Stream Name	Supply Temp., T _s (C)	Target Temp., T _T (C)
1 Cold	Crude Palm Oil Feed, CPO	50	97
2 Cold	Degaser Outlet, DEGAS	104	124
3 Hot	Bleached Palm Oil, BPO	120	86
4 Cold	Dried Oil, DRIED	86	230
5 Hot	Deodorizer Outlet, DEO	260	160
6 Hot	Free Fatty Acids, FFA	83.3	70
7 Hot	Refined, Bleached & Deodorized Palm Oil, RBDPO	160	50

Table 1. Process Stream Data for the Palm Oil Refinery

BASIS OF STUDY

Throughout this paper, the term "current" and "present" refer to the nominal plant conditions in order to cater the case of maximum plant capacity. For this nominal conditions, stream data necessary for the purpose of the studies is given in Table 1.

The following information form a cost basis for this study. Plant annual downtime is taken as 500 hours (including routine exchanger cleaning). Equipment capital cost is taken from Douglas (Douglas, 1988), using M&S indices to update to 1992. Present process conditions and materials of construction are built in into the cost factor. *Installed* costs are as follows;

Heat exchangers

$$1054 A^{0.65}$$

Cooling is performed using cooling water with a temperature range of 25 - 30 C. Fuel cost is averaged at US\$ 0.00259/MJ. This figure is taken from a recent energy audit studies (ADB,1993) on the Malaysian industries and agrees well with the value quoted by Kotjabasakis at US\$ 0.0026/MJ. Assuming the steam generation efficiency to be 75% the utility costs are given as follows,

Type	T_s (C)	T_T (C)	h_f (kW/m ² .C)	\$/kW.yr
Cooling Water	25	30	2.5	18.2
LP Steam	143	142	4.5	106.4
MP Steam	184	183	4.5	159.6

FOULING PROBLEMS

In most edible oil refineries, fouling problems are worst in locations where heat exchangers operating at a temperature in excess of 200 C are exposed to deposition of free fatty acids. It is found that the process heat exchanger located just before the deodorizer column is badly affected by fouling. Particles deposited on the heat exchanger surface form a resistance to heat transfer and results in decreased heat transfer area. A standby heater is needed to provide the additional duty required to raise the stream temperature to the desired value. As fouling becomes worse, the utility consumption begins to increase and reaches its peak after 3 months of operations when the entire heat exchanger surface is fouled. At this point, the heater reaches its maximum capacity. The fouled exchanger is taken off-line for two days for cleaning. This incurs some downtime and loss of productivity.

PLANT PERFORMANCE INDEX

A performance index is required in order to assess and improve plant's existing performance. The current practice is to compare a plant's specific energy consumption to that of similar plants elsewhere. Such comparison provide a rough guideline of the possible scope for improvement. The reference figures should always be used with much reservation due to the differences that exist in plant designs, operating conditions, and the economics. Furthermore, it is not uncommon for the reference plant itself to undergo further improvement in terms of energy

consumption. Therefore, any comparison made on the basis of specific consumption is relative and not absolute.

It is possible to predict the best energy performance achievable by a process solely by thermodynamics considerations. All that is required is the stream data as in Table 1 which is derived from process heat and material balance data. Hot and cold streams enthalpies are combined into cumulative profiles of hot and cold curves respectively. The resulting profile are represented on a temperature versus enthalpy diagram to give the total process heat availability and requirement. The profile is known as the composite curves. The composite curves for the refinery under study are shown in figure 1. The overshoot of the cold composite defines the minimum hot utility requirement for the process. Similarly, the overshoot of the hot curve gives the minimum hot utility requirement. For a given minimum approach temperature for heat exchange (ΔT_{min}), the overlap between the curves defines the maximum possible energy recovery for the process. The grand composite curves depicted in figure 2 is essentially a profile of the horizontal separation of the process composite curves with built in ΔT_{min} . The grand composite curves is useful to provide an optimum interface between the process and utility systems. It also allows appropriate placement of distillation columns, heat engines for cogenerations and heat pumps.

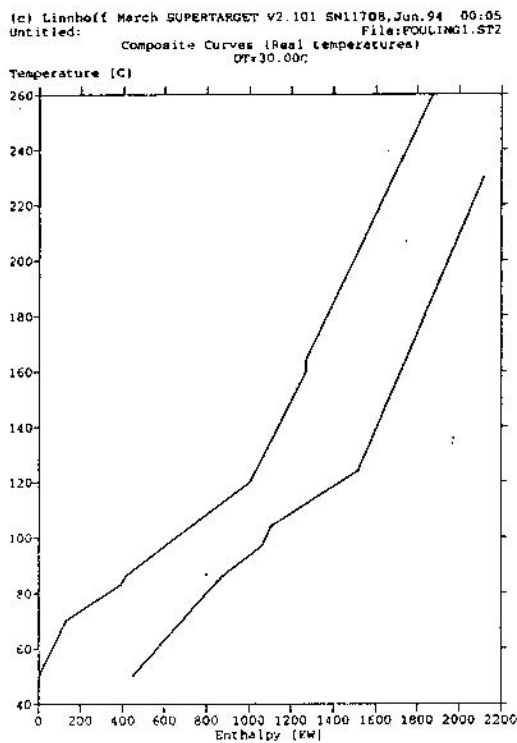


Figure 1. Process Composite Curves

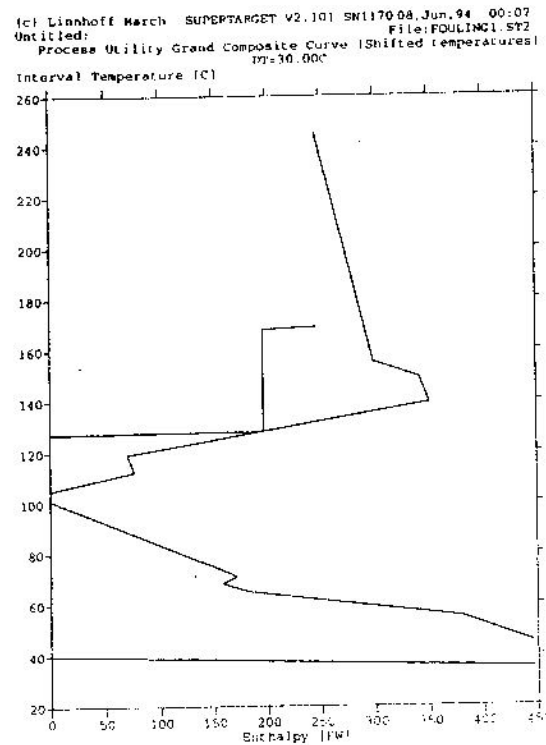


Figure 2. Grand Composite Curves

Table 2 gives a comparison between the existing plant consumption and the thermodynamically derived energy targets for the various relevant approach temperatures for heat exchange. Comparison against the target energy consumption for the case with optimum approach temperature gives an indication of how far the current design is from the optimum grassroots design. The procedure provides an absolute basis of comparison and eliminates uncertainties arising from the relative comparison done on the basis of specific consumption as described earlier.

	Targeted Consumption (Hot Utility) kW	Targeted Consumption (Cold Utility) kW	Pinch Interval Temperature (C)	% Difference with existing consumption (Hot Utility)
Current ΔT_{min} (30 C)	246.33	446.14	105	56%
Optimum ΔT_{min} (8 C)	15.50	215.31	108	97%
Retrofit ΔT_{min} (18 C)	136.28	336.09	113	76%

Table 2. Comparison between energy targets¹ and the actual consumption² at relevant ΔT_{mins}

¹Values generated by SUPERTARGET™ (Linnhoff March, UK) and verified by manual calculations.

²Current consumption; Hot utility = 559.32 kW; Cold utility : 757.79 kW

HEAT EXCHANGER NETWORK RETROFIT

Techniques for network retrofit differ from that of grassroots designs because of the need to maintain the existing plant structure as much as possible. In retrofit schemes, modifications have to be kept at a minimum in order to achieve good payback. With this target in mind, it would be irrational to aim at achieving grassroots design in any retrofit project. This philosophy affects the choice of ΔT_{min} for initialization. In any retrofit project, the grassroots optimum ΔT_{min} is no longer valid. Constraints such as plant layout, operability and company financial standings deserve due considerations.

A more practical approach begins with assessment of the existing plant heat recovery network efficiency and comparing it against the optimum. Linnhoff and Joe (Tjoe, 1986) first devised a conservative retrofit targeting method based on the surface area efficiency, α . α is defined as the ratio of minimum area at the current plant energy usage to the actual area. The area efficiency for the refinery under study is 0.51 which is rather low. This means that more effective use can be made of the existing area. A conservative target would be to assume a constant surface area efficiency for retrofit. Physically, this means that the energy consumption is to be reduced at the expense of some investment in added area, repiping works, modifications and control. Ideally, retrofit should approach the target energy consumption corresponding to the current network area as depicted by the horizontal line in figure 4. In practice, this does not happen as recovery is always accompanied by some form of network modifications and investments. Therefore, a constant α value represents a conservative assumption as the present network inefficiency is maintained while energy is recovered.

From figure 4, it is evident that different levels of savings can be achieved for different levels of investments. A savings versus investment curve (Figure 3) can be generated based on the retrofit curve to give an estimated target of the project economics before detail retrofit is carried out. This target is normally chosen from a payback criteria specified by plant managers.

49065.86, or 71% and 52% of the present hot and cold utility consumption respectively. The payback period is evaluated at 1.11 years.

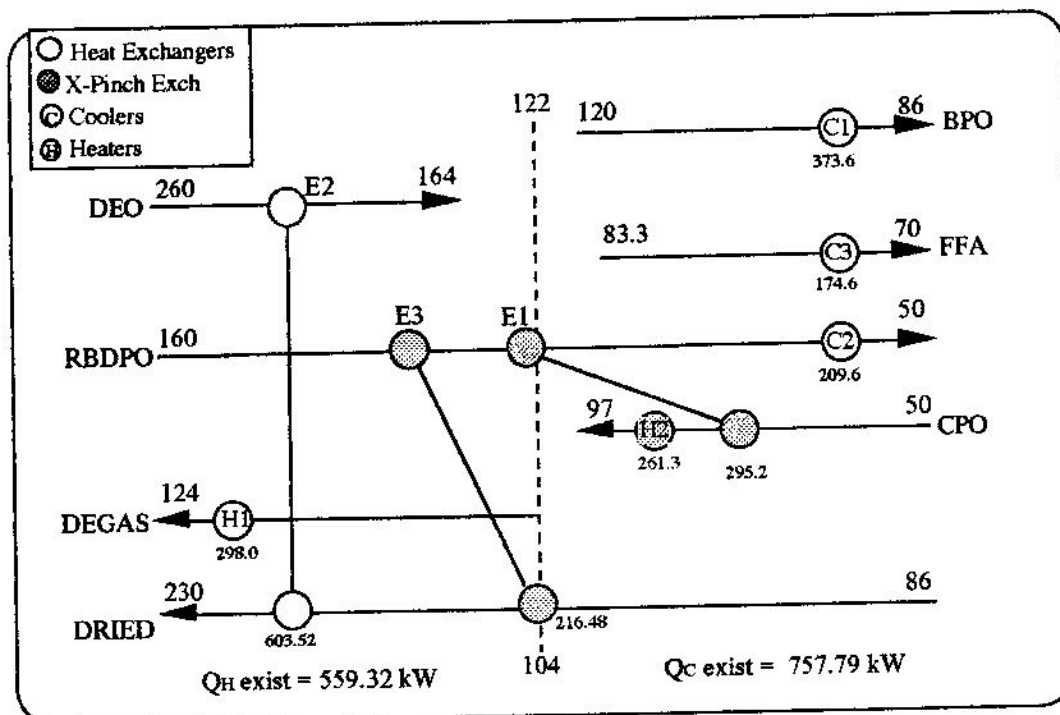


Figure 5. Current Heat Exchanger Network Showing Cross-Pinch Heat Transfer

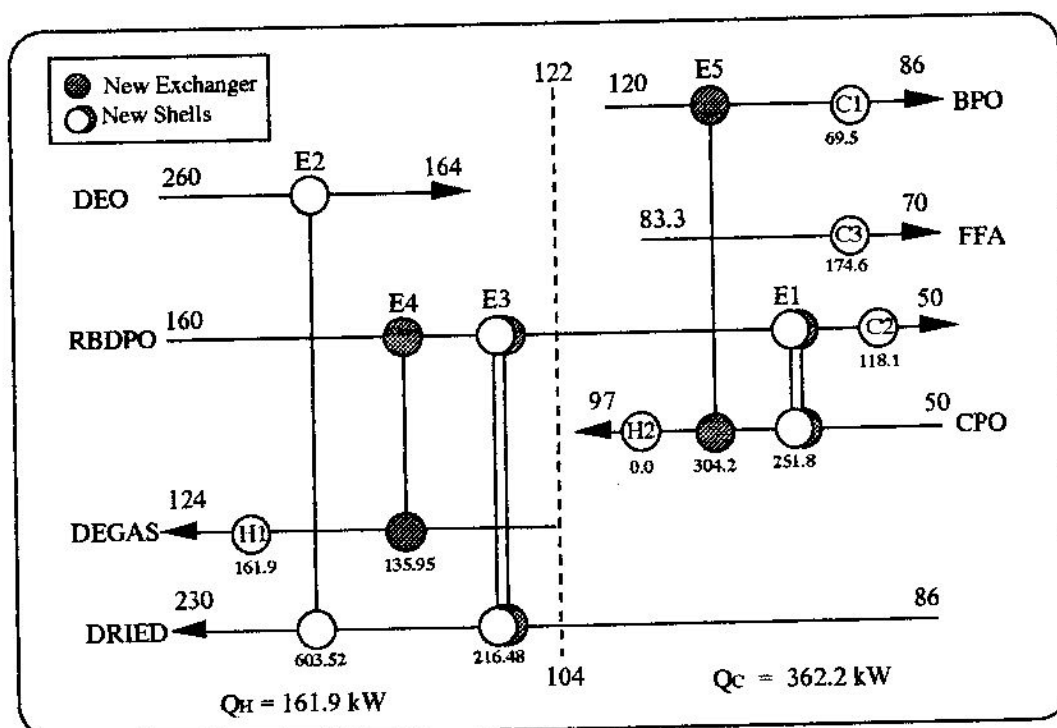


Figure 6. Final Network After Retrofit of Cross-Pinch Exchangers

FOULING ANALYSIS

The refined, bleached and deodorized oil (RBDO) contains trace amount of free fatty acids acting as a foulant once the temperature exceeds 200 C. Because fouling is temperature sensitive, only exchanger E2 is affected.

Fouling occurs at a rate such that after every 3 months, the performance of heat exchanger E2 has deteriorated to such an extent that the required duty on the heater H3 exceeds the maximum capacity of 602 kW. At this point, the plant can no longer operate. It is also clear that the plant cannot operate with E2 removed for cleaning. The current practice to overcome this problem is either to add a new heater or oversize the fouling exchanger, E2 in order to extend the fouling cycle. Providing a new heater would further increase utility consumption hence the operating cost. Besides, maintenance downtime may be delayed but not avoided. Oversizing the fouling exchanger may lead to accelerated fouling due to increase in exchanger tube diameter and reduced fluid velocity.

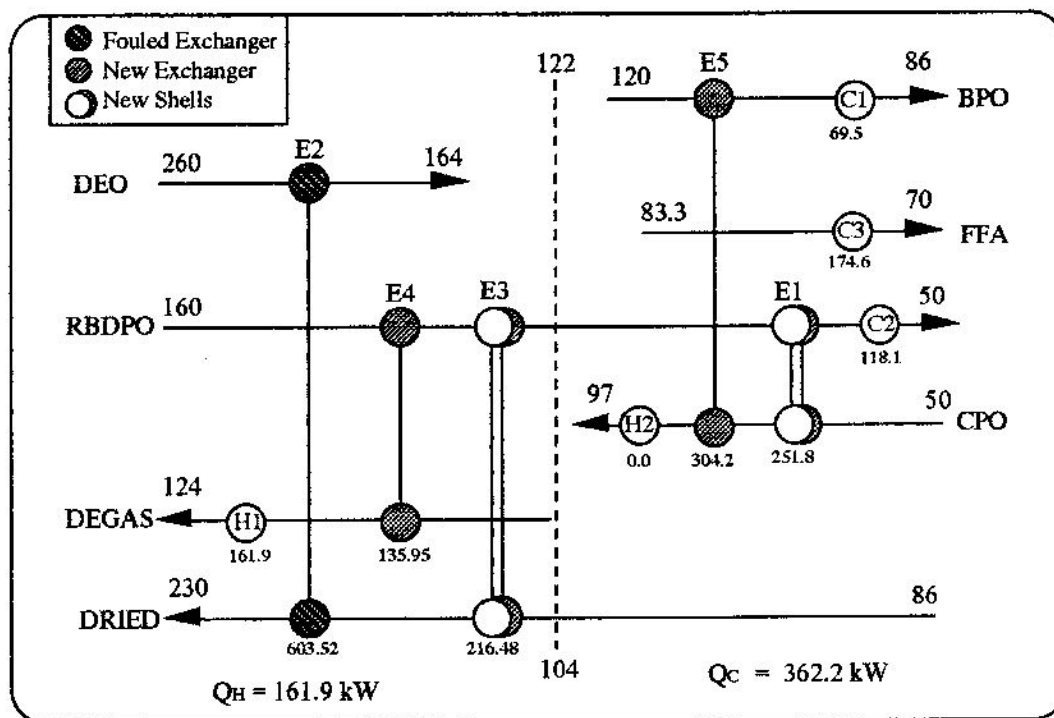


Figure 8. Retrofitted Network Showing the Fouled Exchanger, E2

Linnhoff and Kotjabasakis (Kotjabasakis, 1987) maintained that normally the best approach would be to oversize a nonfouling exchanger that exert downstream effect on the one that is fouled. If there are more than one exchangers affecting the fouling exchanger, the choice might not be so obvious. In such a case, choosing the most cost-effective exchanger to oversize requires analysis of the overall network interactions with heat exchanger E2. Such analysis will lead to the most cost-effective location for overdesign.

From figure 8, it is evident that E2 is only directly affected by E3. This happens to be the only nonfouling exchanger which has a downstream effect to E2 hence, appear to be the most logical candidate for overdesign as suggested by heuristics proposed by Kotjabasakis and Linnhoff.

A closer observation on the network interactions in Figure 8 suggests that providing overdesign on E3 would result in an inoperable network as indicated by figure 9.

This is because stream DEO has no "gearing" to utilities to enable it to reach the target temperature of 164 C as E2 is removed for cleaning. Such phenomenon appear to contradict the proposed heuristics. The only alternative left is to overdesign close to the fouling exchanger. Although the additional area is not fully utilized in this case, the network is operable and the plant can operate without shutdown with E2 removed for cleaning once every 3 months. The next logical question is how much additional area is needed to maintain the network integrity?

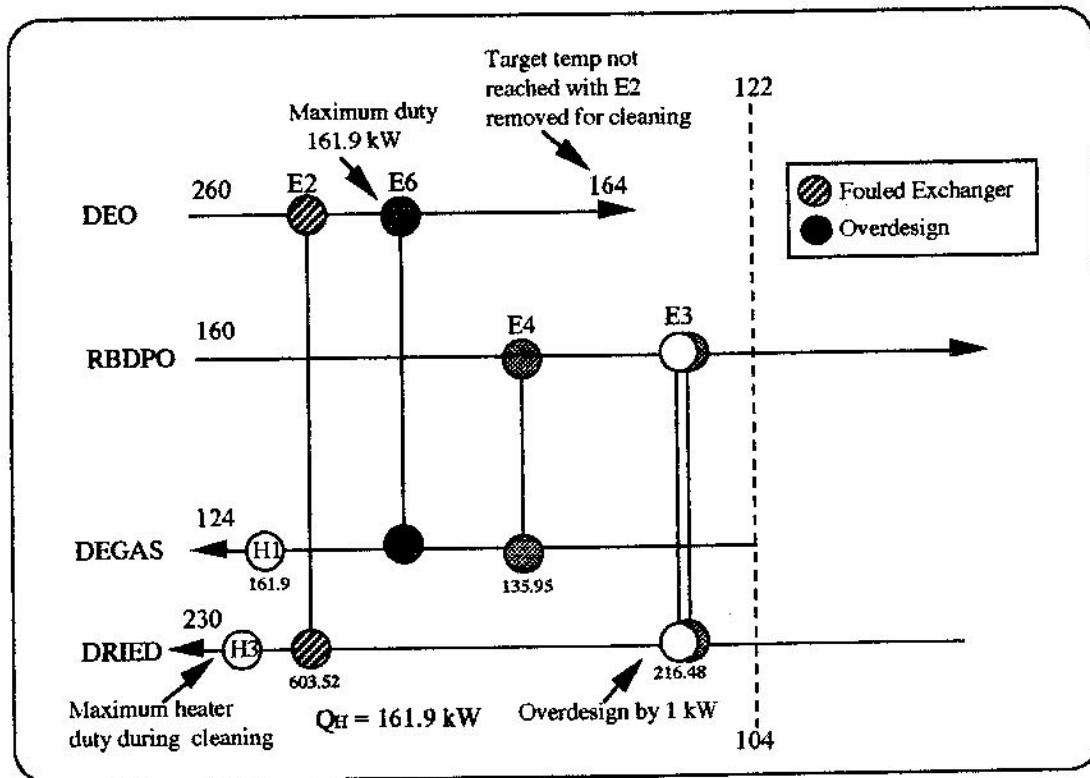


Figure 9. Overdesign Of Non-Fouling Exchanger Results in an Inoperable Network

The maximum heater duty is related to the minimum additional area required by the exchanger to be oversized. This is in order for the plant to operate without shutdown as the fouled exchanger is removed for cleaning. The maximum heater duty of 602 kW would satisfy most of the heating requirement for stream DRIED. However, the minimum additional area would only satisfy heating requirement for DRIED. Stream DEO will not reach its target temperature of 164 C unless an overdesign equal to the total duty of E2 is provided. Investment analysis shows that adding an exchanger parallel to the one that is fouled is the best option from the point of view of economics, controllability and operability.

The final network is shown in figure 12. An operable design that can operate without any cleaning downtime has been achieved and the utility consumption is maintained at the retrofitted value. Design below the pinch follows the original retrofitted scheme. The network results in overall investment of \$ 80980.94 and savings of \$ 49065.86 representing 71% savings from the annual steam consumption and 52% process water consumption. This gives a payback period of 1.65 years

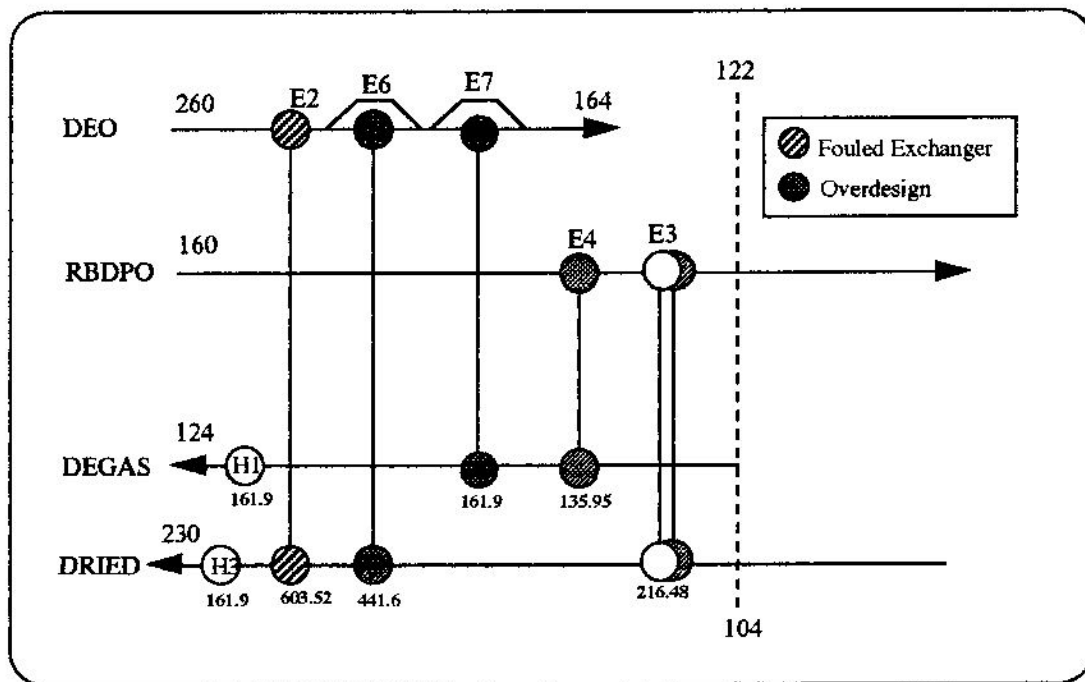


Figure 10. Oversize Close to the Fouling Exchangers - Alternative 1

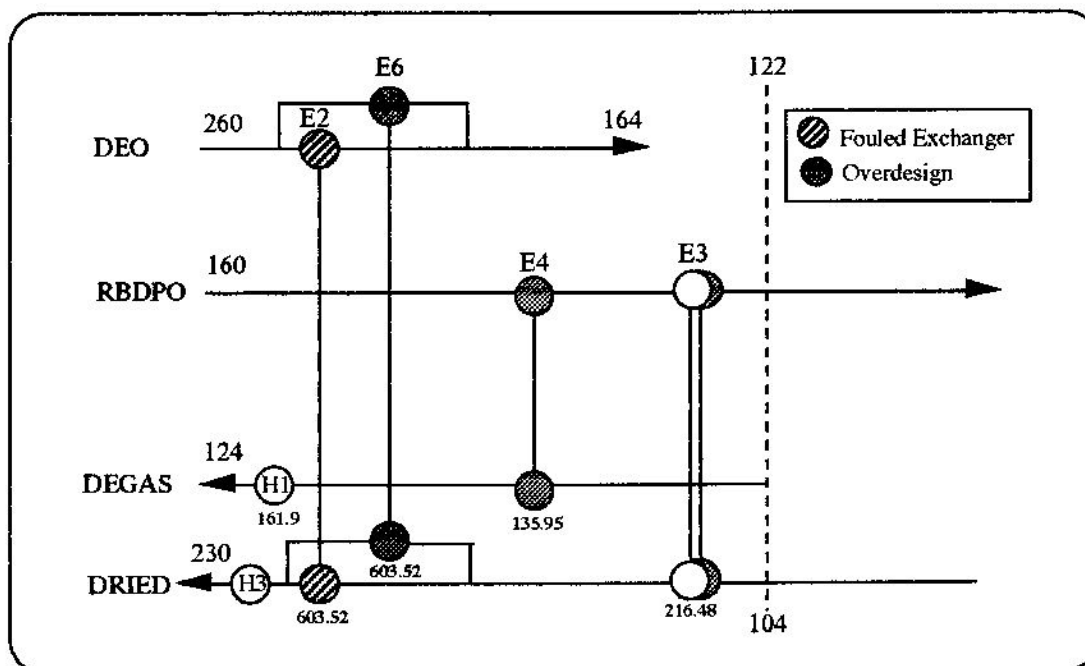


Figure 11. Oversize Close to the Fouling Exchangers - The Better Option

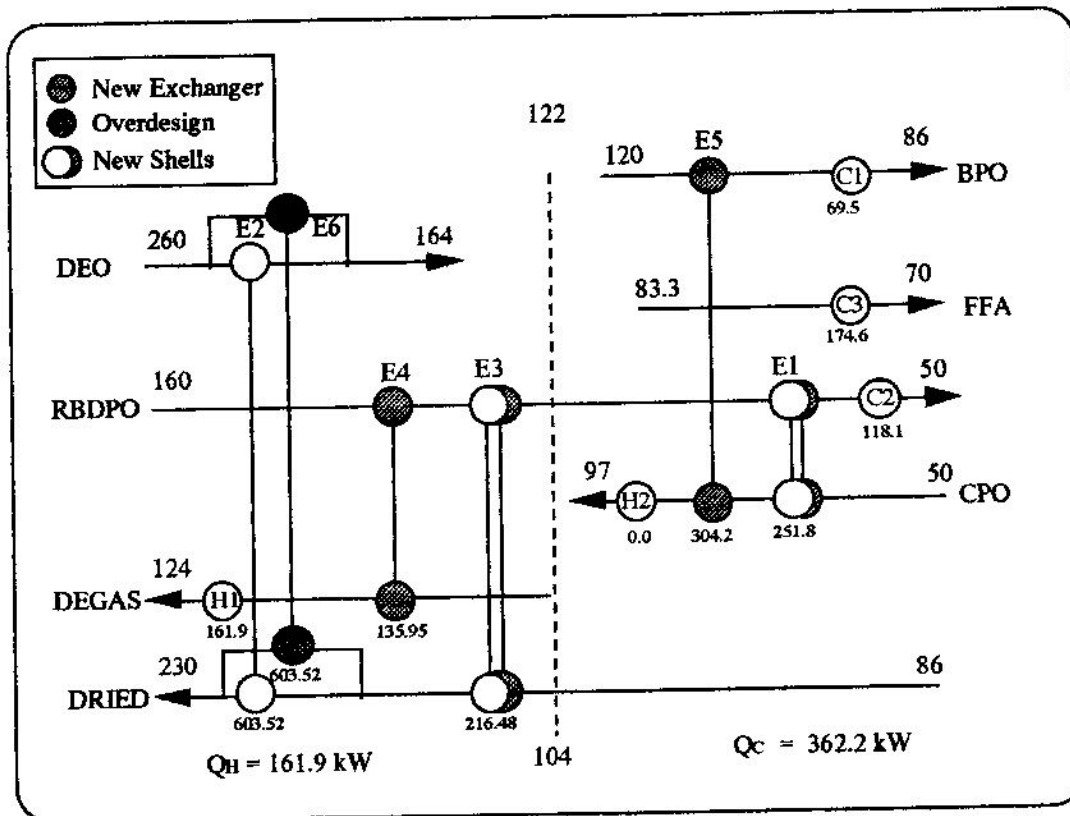


Figure 12. Final Network After Retrofit With Fouling Considerations

Heat Exchanger	Existing Area (m ²)	Retrofitted Area (m ²)	Added Area (m ²)	Additional Investment \$
E1	59.69	81.96	22.27	7922.44
E2	144.72	maintained	54.20	14123.43
E3	46.93	101.13	27.78	9146.74
E4	-	27.78	115.13	23046.93
E5	-	115.13	144.72	26741.40
E6	-	144.72	-	-
H1	30.00	maintained	-	-
C1	72.52	maintained	-	-
C2	108.28	maintained	-	-
C3	46.44	maintained	-	-
Total			364.10	80980.94

Table 3. Capital Investment For the Proposed Scheme- Installed Costs

SIGNIFICANCE

A lot of effort has been directed towards studying the mechanism of fouling and controlling heat exchanger fouling. To date, very few works are focused on exploiting heat exchanger network sensitivities. Kotjabasakis and Linhoff first studied the effects of heat exchanger network sensitivities on fouling control and

concluded that in general, more savings can be realised through oversize of non-fouling exchangers. The heuristics may not apply to all processes especially for existing installations. As illustrated in this study, there are cases where oversize of nonfouling exchangers may prove to be inoperable or more expensive to implement. The study done by Kotjabasakis and Linnhoff was aimed at improving grassroots designs. As far as is known, this is the first published work on retrofit involving fouling exchangers. Palm oil and oleochemicals are areas having tremendous potential for pinch applications. The foregoing case study proved that savings in the order of 71% in hot utility and 52% in cold utility is possible notwithstanding the present heat integration employed by the plant. Such savings results in good return on investment. Constraints such as plant layout, fouling downtime, operability are taken into considerations. However, implementation of the project, calls for input from experts on the particular process under investigation. As pointed out by Eric Petela, a pinch expert from Linnhoff March of UK, Pinch Technologists invariably works with client process specialists to ensure that only practical projects are developed (Petela, 1994). The outstanding track records of Linnhoff March speak for themselves. To date, LM inc. has completed over 1000 projects worldwide with 47% so far being implemented. The recipe is the simplicity and practicality of the methodology itself; and finally, the faith and the willingness on the part of the client to accept change.

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