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Implementation of Heat Integration to Improve the Sustainability of an Integrated Biodiesel Biorefinery

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Recently, a great research effort has been dedicated in order to establish integrated biorefineries as the next generation plants for the production of fuels, energy and chemicals. In this work, heat integration techniques are applied to an integrated biodiesel biorefinery that as well as biodiesel it also produces succinic acid through a fermentation process in order to improve its economic efficiency. Initially, heat integration options are identified by extracting the thermodynamic results of the overall biorefinery from simulations carried out in Aspen Plus 2006.5. By implementing pinch analysis and stochastic optimisation techniques, we determine a new heat exchanger network (HEN) that shows a minimum total annualised cost (TAC) of the overall HEN. The optimisation of the HEN results in a 17.2 % reduction in the TAC and 62.3 % and 64.2 % reduction in the hot and cold utilities, respectively, compared to the initial plant.

1. Introduction

As society seeks technological alternatives to decrease dependency on petroleum and fossil fuels, the development of integrated biorefineries for the production of energy, fuels and chemicals is becoming more and more attractive. To date, this concept has had few applications due to the elevated cost of feedstocks and of processing technologies of biomass. Therefore biorefineries are mainly regarded as biofuels plants. However, there is an increasing interest in research and development for expanding their applications to the chemical market, by analogy to petroleum refineries.

Biodiesel can partially and/or entirely substitute diesel in internal combustion engines. Although this biofuel constitutes only a small fraction of the overall liquid fuel market, recently its worldwide production has been increased (Vlysidis et al., 2011a). The main drawback of the biodiesel process in comparison with petroleum-based diesel is its higher production cost due to the high vegetable oil prices (except if waste oils are used as feedstock). Therefore, biofuels, in order to be competitive, are subsidized by governments or they are exempt from excise taxes. It becomes evident that viable solutions to increase the economic sustainability of the biodiesel industry are in great demand.

In previous studies, it has been showed that the construction of an integrated biodiesel biorefinery plant that produces biodiesel and succinic acid can give a significant economic boost to the biodiesel industry (Vlysidis et al., 2011a; Vlysidis et al., 2011b; Binns et al., 2011). Succinic acid is a value added chemical that can be used as a platform monomer for the production of bio-plastics and in the biodiesel plant it is generated from fermenting the glycerol that is produced as a by-product in the transesterification reaction (Vlysidis et al., 2011c). Additional methods which are usually applied in conventional refineries like heat integration techniques will result in increasing the sustainability of the

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biorefineries as they will use energy more efficiently. A well-established method for energy recovery in chemical processes is the pinch analysis where hot and cold streams of chemical plants interact together in such a way so as to provide economic and practicable processes for sustainable industries (Smith, 2005). In this study, we design and simulate a heat exchanger network (HEN) that minimises the total annualised cost (TAC) of an integrated biodiesel biorefinery by using pinch analysis and stochastic optimisation techniques. We also compare the optimised designs with the initial network in terms of energy recovery and economics and we draw appropriate conclusions.

2. Optimisation of HEN - Methodology

The biodiesel succinic acid biorefinery has been designed and simulated in a previous study in Aspen Plus 2006.5 and has been already described extensively by Vlysidis et al. (2011a) and Binns et al. (2011). Here for completeness, we show the overall biorefinery in Figure 1 which includes the biodiesel and fermentation process and in Figure 2 which depicts the downstream process of succinic acid. From the original plant of Vlysidis et al. (2011a), we have made few modifications. Particularly, we have introduced two coolers (COOLER-3, Fig.1) and (COOLER-6, Fig.2) to cool down the biodiesel and the purified water, respectively, before they are released. In addition, we have removed two heaters which heat up the methanol/KOH and the OIL stream, as the transesterification reaction can be also performed in ambient temperatures (Komers et al., 2002).

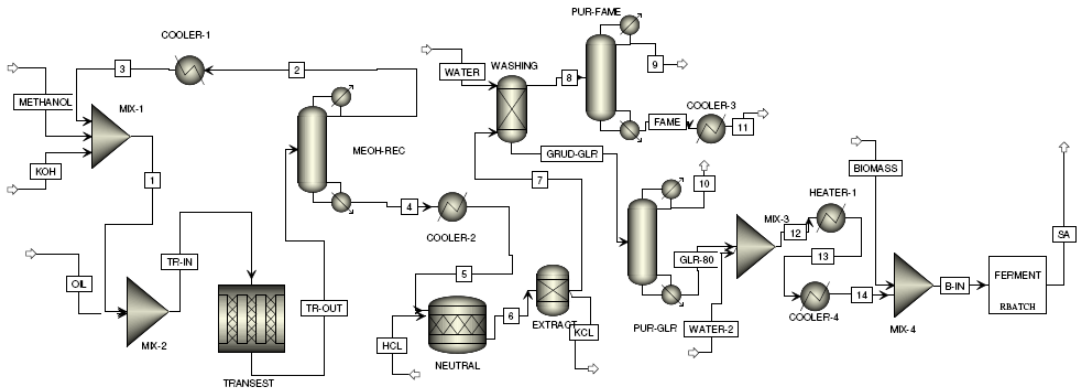

Figure 1: Flowsheet of integrated biorefinery for biodiesel production from vegetable oils and production of succinate through fermentation from crude glycerol (Binns et al., 2011).

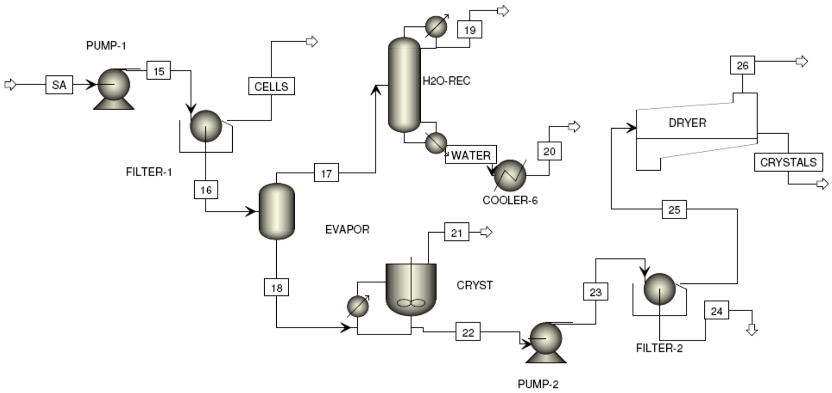

Figure 2: Flowsheet of the downstream process of succinic acid after the fermentation procedure (Vlysidis et al., 2011a).

In this work, we design a new HEN (i.e. not a retrofit of an existing design) that minimises the total annualised cost (TAC) of the process. The procedure that we follow is presented in Figure 3. Initially, the basic stream data are extracted from the material and energy balances from Aspen Plus 2006.5, before any heat integration analysis is performed. The initial assumed ΔT_{min} is 10^{o}C. The heating and cooling duties that are not serviced by heat recovery and should be provided by external utilities have to be considered. The appropriate use of utilities, that provide the energy requirements of the plants at the cheapest operating cost, are determined by the examination of Grand Composite Curves (GCC) of the processes (Smith, 2005). Medium (MP, 28 bar) and low pressure (LP, 3.5 bar) steam levels and hot oil are used as hot utilities and cold water is used for cooling. The supply temperature of the latter is based on the plant location (UK) and the average ambient temperature is assumed to be 10 ^{o}C. The target temperature of CW is assumed to be 25 ^{o}C. In order to determine the optimum ΔT_{min} for the process, the energy-capital trade-off has to be examined.

After determining the energy targets of the plant, we continue with the design of the new HENs. The strategy that is adopted for the maximum energy recovery (MER) HEN designs is based on the pinch analysis (Linnhoff et al., 1983). After the construction of MER HEN designs, the optimisation procedure follows. Considering new designs (no retrofit of existing designs), the main focus of the optimisation is to minimise the TAC which is the objective function by taking into account the energy targets of the examined streams which are the constraints. The optimisation of a MER HEN design can be achieved through the redistribution of the exchanger duties. In the current study, the optimisation procedure is carried out through stochastic optimisation (simulated annealing) (Kirkpatrick et al., 1983). The degrees of freedom are the heat loads of the heat exchangers and the branches of the flowrates for the examined streams and the objective is to minimise the total annual cost for the chosen ΔT_{min}. After stochastic optimisation, the TAC is further optimised by using a non linear optimisation method that fine tunes the stochastic solution.

Here, the TAC is equal to the annualised capital cost and the operational cost. The capital cost is estimated from Smith (2005). In this study, we used shell and tube heat exchangers and the cost of the exchangers are adjusted to today' prices by using the cost index of 2010 (CHE, 2010). The capital cost has to be annualised in order to be able to be summed with the operating cost. The annualised capital cost is given in equation 1 (Holland et al., 1983).

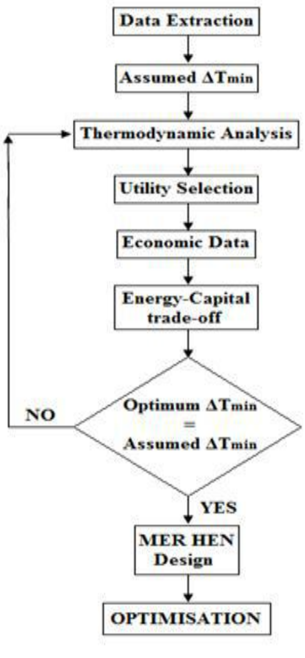

Figure 3: Methodology procedure for designing and optimising HENs

$$Annualised \quad Capital \quad Cost = Installed \quad Capital \quad Cost \; * \; \frac{i(1+i)^n}{(1+i)^n - 1} \tag{1}$$

Where i is the interest rate (7 %) and n is the lifespan of the investment (20 years), as originally selected by Vlysidis et al., 2011a, while the annual operating time of the plant is 7920 h/y. The operational cost is calculated by considering the cost of each utility and the amount of energy used from that utility. For MP steam, LP steam, CW and Hot Oil the utility costs are 90.1 (OGJ, 2011), 13.8 and 7.13 (Spirax Sarco, 2011), and 121.6 $/kWy (Vlysidis et al., 2011a).

3. Optimisation of HEN - Results and Discussion

The stream data of the biorefinery are extracted from the simulations conducted in Aspen Plus 2006.5 and presented in Table 1 and Figures 1 and 2. In the first column of Table 1, we have listed all the units where heat transfer takes place, and thus, we have included all the coolers and heaters, the reboilers and condensers from the distillation processes and the evaporator from the downstream process. The T_S is the supply temperature, the T_T is the target temperature, ΔH is the energy duty of each unit, CP is the heat capacity and h is the film transfer coefficient. The total energy requirement (e.g. the summation of all the duties) is 8737.8 kW. The annual energy cost, the heat exchange area and the annual capital cost are 144,068.6 $/y, 101.31 m^2 and 104,119.4 $/y, respectively. The composite curves of this process are constructed by using a global ΔT_{min} =10 ^oC (see Figure 4). From the pinch analysis, we have found out that the system has a process pinch on 97.5 ^oC and that there are both hot (Q_{Hmin}) and cold (Q_{Cmin}) utility requirements. It becomes obvious from Table 1 that units which are connected with the fermentation process like the heat sterilisation units (Heater 1 and Cooler 4) or with the downstream process like the evaporator and the water purification provide high opportunities for energy recovery and heat integration as they have high heat loads.

Figure 5 illustrates the optimised HEN of the process. In this figure, hot stream lines start from the left side of the figure and end up on the right side after heat transfer and vice versa for the cold streams. The last four stream lines represent the utilities that we use (cold water is represented on the last cold line, 19). The total number of heat exchangers is eighteen. Six HEs utilise hot utilities (No: 2, 5, 7, 9, 11 and 14) while seven of them utilise cold utilities (No: 3, 8, 15, 17, 18, 19 and 20). There are also five HEs which combine cold and hot streams from the process (No: 1, 6, 10, 12 and 13). From the optimisation results we have also extracted the total heat exchanger area which is 131.3 m^2. The duties from each heat exchanger are shown in boxes below each HE (see Figure 5).

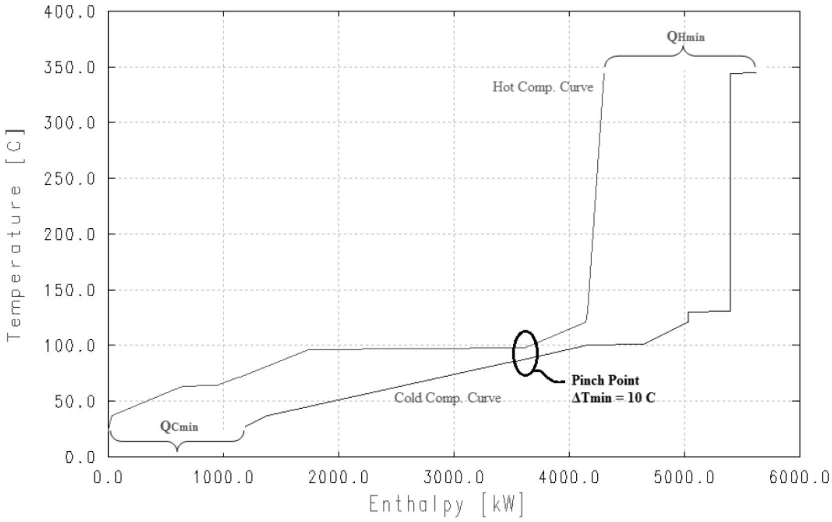

Figure 4: Composite curves of the overall biorefinery

Table 1: Stream data of the process

Process Unit	Stream No.	T_S [°C]	T_T [°C]	*ΔH* [kW]	*CP* [kW/°C]	*h* [kW/°C m²]
Cooler 1	1	64.5	22.7	11.8	0.282	2.0
Cooler 2	2	130.0	60.0	50.2	0.716	0.8
Cooler 3	3	344.0	25.0	219.3	0.687	0.6
Heater 1	4	27.0	121.0	1788.5	19.032	3.7
Cooler 4	5	121.0	37.0	1760.8	20.962	3.7
Cooler 6	6	100.0	25.0	108.6	1.448	4.0
MEOH-REC (Reb.)	7	130.0	131.0	362.8	362.845	0.8
MEOH-REC (Cond.)	8	64.5	63.5	277.4	277.415	2.0
PUR-FAME (Reb.)	9	344.0	345.0	227.3	227.350	0.6
PUR-FAME (Cond.)	10	71.5	70.5	8.8	8.769	3.0
PUR-GLR (Reb.)	11	84.7	85.7	23.1	23.119	1.0
PUR-GLR (Cond.)	12	65.3	64.3	14.6	14.563	3.0
H2O-REC (Reb.)	13	100.0	101.0	444.9	444.877	4.0
H2O-REC (Cond.)	14	97.5	96.5	1852.3	1852.267	2.5
Evaporator	15	37.0	101.0	1587.4	24.803	2.0

The optimisation results are then compared with the corresponding results from the initial network (without heat integration) in terms of economics and energy requirements (see Table 2). The optimisation procedure is applied in order to reduce the heat exchanging area (capital cost) by redistributing the heat loads of the existing heat exchanger through utility paths and loops. This is achieved by adding three more heat exchangers in the final network (18 instead of 15). As a result, the capital cost is increased by 23.3 % from the initial plant. Comparing the initial network and the optimised one, the hot requirements of the process are reduced by 62.3 %, while the cold utilities are decreased by 64.2 % due to the heat integration. Hence, the operating cost is significant lower than that of the initial network (46.4 %). Based on these changes, the final network is less expensive by 17.2 % or 42,560 $/y than the initial HEN. The above results indicate that applying heat integration techniques in biorefineries can be very vital to the economic sustainability and the overall efficiency of the plant.

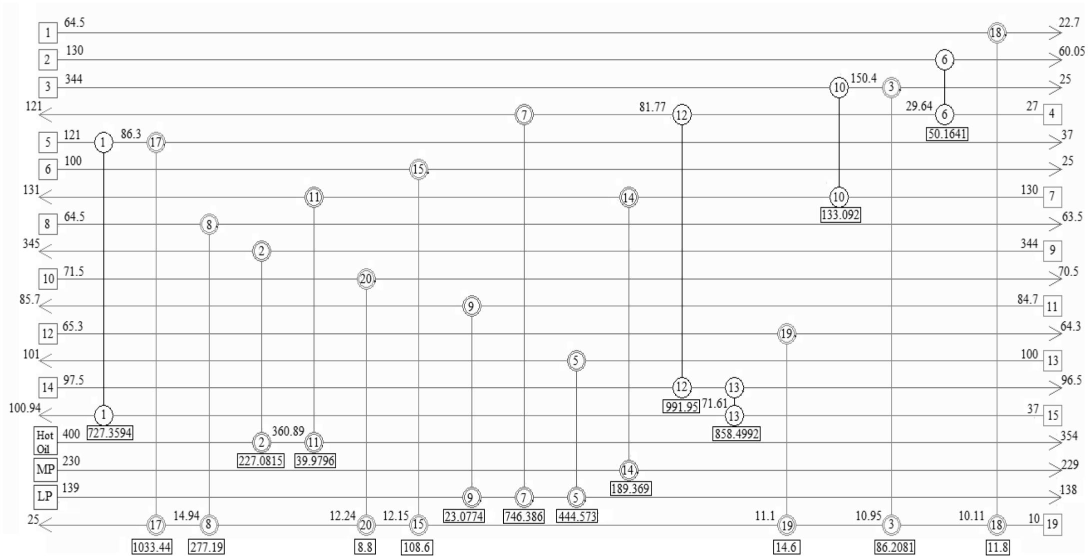

Figure 5: Final HEN design for the biodiesel succinic acid biorefinery

Table 2: Comparison of the initial and final HEN

HEN	Units	HE Area [m^2]	Q_{REC} [kW]	Hot and Cold Utilities [kW]	Capital Cost [$/y]	Energy Cost [$/y]	TAC [$/y]
Initial HEN	15	101.3	-	8,737.8	104,119	144,069	248,188
Final HEN	18	131.3	2,761.1	3,211.1	128,353	77,275	205,628

4. Conclusions

In this study, the design and optimisation of heat exchanger networks in an integrated biorefinery with minimum total annualised cost has been carried out. High opportunities for heat integration were found on the biodiesel succinic acid biorefinery due to the sterilisation and the downstream processes which have high energy requirements. Here, pinch analysis and stochastic optimisation techniques were followed in order to design and optimise the new HEN. The heat that is recovered in the optimised network is 2,761 kW which led to an important decrease in operating cost of the network (46.4 %). Although the capital cost is increased by 23.3 %, the total annualised cost of the optimised network is reduced by 17.2 % compared with the initial network. These results illustrate how the energy requirements of a biorefinery can be re-distributed by implementing heat integration techniques in order to enhance the sustainability of the plant by reducing the utility consumption and the total annualised cost of the plant. Hence, heat integration methodologies are useful for the efficient design of the integrated biorefineries of the future as they result in more sustainable and economical competitive plants when compared to conventional refineries.

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