Design and simulation of an integrated process for biodiesel production from waste cooking oil using supercritical methanolysis

Omar Aboelazayem [a,b], *Mamdouh Gadalla* [c,a,1] *and Basudeb Saha* [b]

[a] Department of Chemical Engineering, The British University in Egypt, Misr-Ismalia Road, El-Shorouk City 11837, Cairo, Egypt.

[b] Centre for Energy and Environment Research, School of Engineering, London South Bank University, 103 Borough Road, London, SE1 0AA, UK.

[c] Department of Chemical Engineering, Port Said University, Port Said, Egypt.

ABSTRACT

Non-catalytic transesterification has been recognised as an effective technique for biodiesel production. It has many advantages over conventional catalytic transesterification, where it eliminates the difficulties of catalysts preparation and separation. It also produces high biodiesel yield in shorter reaction time. However, it requires harsh operating conditions at high reaction temperature and pressure, in addition to using large excess of methanol. In an attempt to mitigate these problems, a process design/integration for biodiesel production has been performed. The process has been subjected to both mass and energy integration to minimise fresh methanol requirements and to minimise heating and cooling energies, respectively. A new graphical Pinch Analysis method has been used to evaluate the energy performance of a literature design for the current process. It has been subsequently used to develop an optimum heat exchanger network (HEN) for the process by matching of process streams. Also, the design made by using an automated commercial simulation (Aspen Energy Analyzer) has been evaluated using the same graphical method. The produced HEN design from graphical method has achieved the optimum results with respect to energy targets.

KEYWORDS

Biodiesel, Waste cooking oil, Graphical Pinch Analysis, Heat integration, Mass integration.

HIGHLIGHTS

- Design and simulation of an integrated process for biodiesel production.
- Heat and mass integration have been applied to improve process performance.
- An optimum HEN has been developed using graphical Pinch method.
- Previous HEN designs have been evaluated using new graphical Pinch method.

Corresponding Author

[1] (M. Gadalla). Email: mamdouh.gadalla@bue.edu.eg

1. INTRODUCTION

Diesel fuel is the most consumed fuel over other petroleum products where it is used extensively in transportation sector and energy generation [1]. Consequently, biofuels with similar properties to diesel fuel are considered the future of energy resources as they could be implemented directly to the existing diesel engines with minor or without modifications [2]. Biodiesel has been considered as the most promising substitute for petroleum diesel fuel. It is a biomass-derived fuel from vegetable oils, animal fats and recently from micro-algae. Biodiesel is defined as mono alkyl esters of long chain fatty acids derived by alcoholysis of triglycerides from different feedstocks [3]. Biodiesel is characterised by its biodegradability, low emissions of particulates, carbon monoxide and hydrocarbons, absence of sulphur emission and high cetane number [4]. The similar properties of biodiesel to petroleum diesel besides being a sustainable green fuel have promoted biodiesel as a significant alternative fuel for petroleum diesel.

Transesterification reaction has been considered the most commonly used method for biodiesel synthesis. The reaction is usually catalysed using different techniques including alkaline, acidic and biological catalysts [5]. Recently, non-catalytic transesterification has been reported using supercritical alcohols [6]. Origin and quality of the feedstock are responsible for selecting the processing technique. Presently, edible oils are considered the main feedstock for biodiesel. However, the raising competition with food industry resulted in food insecurity has boosted the research towards non-edible and waste oils (second generation feedstock). The second generation feedstock is considerably cheaper than edible oil, which contributes in lowering the overall cost of the produced biodiesel [7].

List of Abbreviations:

WCO, waste cooking oil; FFA, free fatty acid; NaOH, sodium hydroxide; KOH, potassium hydroxide; DMC, dimethyl carbonate; MTBE, methyl *tert*-butyl ether; CSTR, continuous stirred tank reactor; FAME, fatty acid methyl ester; CO_2, carbon dioxide; TAN, total acid number; TG, triglycerides; NRTL, non-random two liquids; EOS, equation of state; k, reaction rate constant; HEN, heat exchanger network; MEN, mass exchanger network; VDU, vacuum distillation column.

Waste cooking oil (WCO) has been recognised as a potential feedstock for biodiesel as it is relatively cheaper than fresh edible oils and it contributes in waste utilisation [8]. However, the high free fatty acids (FFA) and water content are the main drawbacks of using WCO as a feedstock. Alkaline homogenous catalysed technique, using either sodium hydroxide (NaOH) or potassium hydroxide (KOH), is considered the most commonly used technique for biodiesel synthesis. The high FFA content in WCO leads to saponification reaction while using alkaline homogenous catalysts, which results in lowering biodiesel yield and preventing product separation. Using heterogeneous catalyst prevents saponification side reaction. However, it is very sensitive even at low water content in addition to high cost of catalyst preparation as it requires extremely harsh conditions (700-900^{o}C). Two-steps transesterification is considered as an acceptable technique for producing biodiesel from WCO. A pre-treatment esterification of FFA step using acidic catalysts is followed by transesterification step using alkaline homogenous catalysts. Nevertheless, the lengthy process leads to increase in biodiesel cost [9].

Non-catalytic transesterification has been considered as an ideal technique for biodiesel production from WCO as it prevents all the above-mentioned problems. It tolerates both esterification of FFA and transesterification of triglycerides in a single step reaction. However, it requires high reaction temperature and pressure, where the alcohol should be at the supercritical state [10]. Several supercritical technologies have been used for non-catalytic production of biodiesel using methanol, ethanol, 1-propanol, dimethyl carbonate (DMC), methyl *tert*-butyl ether (MTBE) and methyl acetate [11,12].

West et al [13] have designed and simulated four biodiesel production processes using different techniques including homogenous alkaline catalysed, homogenous acidic catalysed, heterogeneous alkaline catalysed and non-catalytic supercritical processes. They have also performed an economic comparative analysis between the designed processes for the cost of production 8000 tonne/y of biodiesel from WCO. They have concluded supercritical process as the second most profitable process next to heterogeneous catalysed process. Lee et al [10] have simulated production process for biodiesel using both fresh and used cooking oils. They have reported that the cost of the feedstock attributes with about 64-84% of the produced biodiesel cost. They have also reported that using supercritical methanol is the most economically favourable process over alkaline catalysed processes. Manuale et al [14] have simulated an energy-integrated biodiesel production process using supercritical methanol. They have proposed that using the enthalpy content of the reactor product stream to separate most of the unreacted methanol in a flash drum decreased the process required heating energy.

Pinch technology is recognised as one of the most effective methods used to assess the efficiency of energy utilisation for production processes. The idea was proposed in 1978 by Umeda et al [15] which has been developed for further aspects by Linnhoff and Hindmarch [16]. The principle has been subsequently extended into several areas including mass Pinch, hydrogen Pinch and water Pinch. Smith [17] has discussed the principles for Pinch Analysis which have been implemented in mass and energy integration applications and extensively applied in heat recovery. The applications of such principles are very critical for providing energy and mass targets that should ideally be achieved in a process [18]. El-Halwagi [19] has introduced systematic and graphical procedures based on Pinch Analysis to design both mass and heat exchanger networks in complicated process industries.

Process integration for energy or materials savings can be achieved through two approaches, one which is based on insights derived from Pinch Analysis and the other is based on mathematical programming methodologies. The first approach normally comprises of two stages, first determining the energy (or mass) targets known as targeting, and then designing the heat and/or mass exchanger network to achieve these targets [20]. The mathematical programming-based approach relies on building superstructure for all alternatives and then using simultaneous optimisation and integration to explore all interconnection within the proposed superstructure. This is followed by screening of all the alternative to find the optimal combination [21,22]. The recent handbook of Klemes [23] is a good source for such literature.

Gadalla [24] has reported a novel graphical technique for HEN designs based on Pinch technology. The graphical method has been defined by plotting process hot streams *versus* process cold streams. Each process heat exchanger has been represented by a straight line where its slope the is function of the ratio between heat flows and capacities. In addition, each line is proportional to the flow of the heat transferred across the exchanger. This method could easily analyse any proposed HEN to identify inappropriate exchangers whether across the Pinch, network Pinch and improper placements. In addition, he reported that the developed method could be implemented in designing optimum HENs using numerical process streams matching technique. Gadalla [25] has also extended the same conceptual novel graphical method for mass integration applications and mass exchanger networks (MEN) designs.

In this study, a comprehensive integrated design for biodiesel production process using supercritical methanol has been simulated. The reactor has been designed based on previous experimentally reported kinetic parameters. Energy and mass integration principles have been applied to reduce the process required external energy and fresh resources, respectively. Graphical Pinch method has been applied to design and develop a new optimum HEN responsible for reduction of heating and cooling required energies. In addition, it has been used to evaluate previously reported designs.

2. MATERIALS AND METHODS

The transesterification/esterification reactions for WCO were carried out using supercritical methanol. The details about the experimental design and procedures are presented elsewhere [26]. Aspen HYSYS simulation programme version 8.8 was used for simulating the biodiesel process (Aspen Technology Inc., USA). The procedures for process simulation based on HYSYS simulator consist of several steps including selection of chemical components for the process, appropriate thermodynamic models, required process units and operating conditions. The actual existing pressure drop in different equipment was neglected in the present study.

The assumptions associated with the present simulation are as follows:

1. The transesterification reaction steps were represented by only overall step where triglycerides (TG) are converted to fatty acid methyl esters (FAME).
2. Glycerol methanol side reaction was ignored.
3. Heat exchangers were selected as counter flow type and were simulated by a means of a shortcut module.

2.1 Chemical components

Most of the required information for chemical components used in the process design were included in HYSYS data bank library. Triolein ($C_{57}H_{104}O_6$) and Trilinolein ($C_{57}H_{98}O_6$) were used to represent the triglycerides exists in the WCO as they were reported as the major compositions (~86%) based on the chromatographic analysis reported elsewhere [26]. Oleic and linoleic acids have been used to represent the FFAs exist in the WCO. Methyl oleate

($C_{19}H_{36}O_2$) and methyl linoleate ($C_{19}H_{34}O_2$) were considered as the desirable product of the reaction. Conferring to the WCO's total acid number (TAN) of 0.8 mg KOH/ g oil, the FFAs weight percentage were equivalent to 1.6%. Trilinolein component was not available in the HYSYS data bank library where it has been introduced as a hypo-component using hypo-manager tool by identifying its physicochemical properties [27].

2.2. Thermodynamic model

Owing to the presence of polar components in the process, i.e.; methanol and glycerol, non-random two liquid (NRTL) activity model was selected as the fluid thermodynamic package for the activity coefficient calculations [28]. Some binary interaction coefficients were not available in the HYSYS data bank library. Accordingly, the missing coefficients were estimated using UNIFAC liquid-liquid equilibrium and UNIFAC vapour-liquid equilibrium methods. Since the activity coefficient based model such as NRTL is not recommended to be used at pressure greater than 1000 kPa, Peng-Robinson equation of state (EOS) was used in the process streams at high pressure and at separating units [10].

2.3. Plant capacity, unit operations and operating conditions

The biodiesel plant capacity was specified by 9.2 kgmol/h of fresh WCO feed. Conversion reactor unit exists in the simulation environment, which requires only the final reaction conversion, used in most of the process designs in the literature was replaced by kinetic continuous stirred tank reactor (CSTR) reactor. Kinetic and thermodynamic data required for the reactor including reaction rate constant (k), activation energy and frequency factor, were identified based on previous reported experimental data as of 0.0006 s^{-1}, 50.5 kJ/mol and 4.05 s^{-1}, respectively [26]. Reactor operating conditions were identified based on the experimentally concluded optimum conditions reported previously, i.e. methanol to oil (M:O) molar ratio of 37:1, reaction temperature of 253ºC, reaction pressure of 198.5 bar in 14.8 minutes reaction time [26]. The process units include reactor, distillation columns, flash drum, heat exchangers and pumps.

3. PROCESS DESIGN

Biodiesel production process consists of several process stages including reactants preparation, transesterification/esterification reactions, methanol recovery and finally biodiesel purification. The process has been designed as a modified version for a previous process design reported by Lee et al [10]. Methanol and WCO have been pressurised and then heated to the specified conditions; then both reactants have been mixed and fed to the reactor. The reactor product stream has been depressurised and proceeded for further biodiesel purification units. Reactor product stream (Stream 106) has been processed to a simulation tool called “Cutter” which has changed the thermodynamic model from NRTL to Peng-Robinson. Fresh reactants streams for both WCO and methanol were labelled as 100 and 101. Products’ streams including glycerol and biodiesel were labelled as 112 and 114A, respectively. Process flowsheet is presented in Figure 1 and the properties of main streams are given in Tables 1 and 2. A summary of the units’ operating conditions is presented in Table 3.

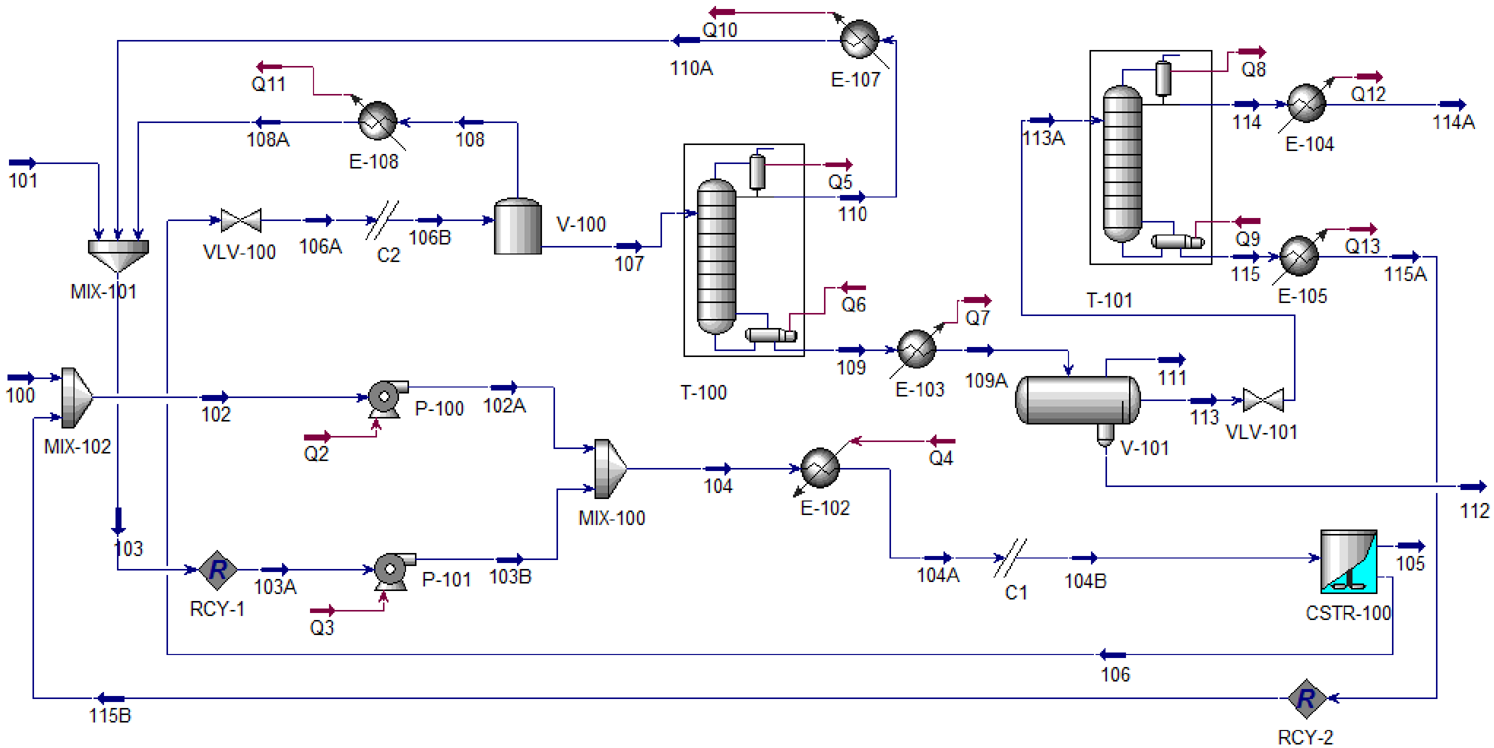

Figure 1. Process flowchart for biodiesel production (numbers below streams refer to stream names)

Table 1. Stream table for the designed process (Part 1)

Name	100	101	102	103	104A	105	106	107
Temperature (ºC)	25	25	25	62	253.5	230	230	87.92
Pressure [kPa]	101	101	101	101	19850	19840	19840	105
Molar Flow [kmol/h]	9.2	27	11.3	386	397.3	0	397.3	53.59
Mole fractions								
Triolein	0.6935	0	0.618	0	0.017	0	0.0016	0.0118
Trilinolein	0.2583	0	0.23	0	0.0066	0	0.0006	0.0044
Methanol	0	1	0	0.922	0.848	0	0.7809	0.2271
Methyl oleate	0	0	0.11	0	0.003	0	0.0522	0.3869
Methyl linoleate	0	0	0.001	0.002	0.0003	0	0.0185	0.1356
Linoleic acid	0.0131	0	0.011	0	0.0003	0	0	0.0002
Oleic acid	0.0351	0	0.03	0	0.0009	0	0.0001	0.0004
Glycerol	0	0	0	0	0	0	0.0220	0.1634
Water	0	0	0	0.076	0.123	0	0.1241	0.0701

Table 2. Stream table for the designed process (Part 2)

Name	108A	109	110	111	112	113	114A	115A
Temperature (ºC)	65	254	72.3	25	25	25	25	25
Pressure [kPa]	101	112	101	101	101	101	101	101
Molar Flow [kmol/h]	343.7	37.97	15.62	0	8.809	29.16	27	1.17
Mole fractions								
Triolein	0	0.016	0	0	0	0.0619	0	0.72
Trilinolein	0	0.006	0	0	0	0.0229	0	0.22
Methanol	0.868	0.008	0.76	0	0.006	0.0009	0.008	0
Methyl oleate	0	0.546	0	0	0	0.6775	0.722	0.04
Methyl linoleate	0	0.192	0	0	0	0.2358	0.27	0
Linoleic acid	0	0.0002	0	0	0	0.0003	0	0
Oleic acid	0	0.0006	0	0	0	0.0007	0	0.02
Glycerol	0	0.2306	0	0	0.994	0	0.0001	0
Water	0.132	0.0001	0.24	0	0	0	0	0

Table 3. Summary of units operating conditions of each process

Operating parameter	**Value**
Reactor (CSTR-100)	
Temperature (^{o}C)	253.5
Pressure (bar)	198.5
Methanol:Oil molar ratio	37:1
Residence time (min)	14.8
Conversion (%)	91.7
Methanol Separating Column (T-100)	
Reflux ratio	1
Number of stages	10
Condenser pressure (kPa)	101
Reboiler pressure (kPa)	112
Methanol recovery	97.8%
Distillate flowrate (kgmol/h)	16.57
Distillate purity (wt%)	84.5
FAME Separating Column (T-101)	
Reflux ratio	1
Number of stages	10
Condenser pressure (kPa)	2
Reboiler pressure (kPa)	5
Distillate flowrate (kgmol/h)	27
Distillate purity (wt %)	99.89

3.1. Non-catalytic reactor

The reactor feed stream (Stream 104A) has been pre-processed to the reaction conditions i.e. temperature of 253.5°C and pressure of 198.5 bar. The volume of the reactor has been identified based on the experimental optimum time of reaction and the flow rate of the reactants where the residence time of the reaction has been adjusted at 14.8 minutes. Consequently, the reactor has resulted in 91.7% conversion of both triolein and trilinolein to methyl oleate and methyl linoleate as similarly reported elsewhere [26]. Esterification reactions of FFAs i.e. oleic and linoleic acids to methyl oleate and methyl linoleate, respectively, have been included to the reaction set. Reaction product stream (Stream 106) has been processed for further separation unit to separate methyl oleate from unreacted components and side products.

In addition, a sensitivity analysis has been performed to investigate the effect of the variation of k on the simulated conversion in the kinetic reactor. It has been reported previously [26] that the reaction is pseudo first order and the value of k is 0.0006 s^{-1} at 253.5°C. Moreover, frequency factor and activation energy have been reported as 4.05 s^{-1} and 50.5 kJ/mol, respectively within the temperature range between 240°C and 280°C. Accordingly, a variation of +/- 0.0001 of the value k has been applied, where new values of activation energy and frequency factor have been determined. Using the new values of k, activation energy and frequency factor have been varied within ranges of 44.26-58.97 kJ/mol and 3.01-6.08 s^{-1}, respectively resulting in a significant variation of the simulated conversion results between ranges of 70.1-97.2 %. The results of this analysis could highlight the high sensitivity of the simulated conversion based on the kinetic data. Hence, it is highly recommended to perform accurate experimental kinetic calculations.

3.2. Separation of unreacted methanol

The actual M:O molar ratio (37:1) used in the reactor is much higher than the stoichiometric requirements for both transesterification of triglycerides (3:1) and esterification of FFAs (1:1). Accordingly, the product stream includes huge excess of unreacted methanol. Reactor product stream (Stream 106) has been de-pressurised to the atmospheric pressure using an expansion valve (VLV-100), where the enthalpy difference of the mixture has converted some of the liquid methanol to the vapour state. The de-pressurised product stream (Stream 106D) has been fed to a flash drum (V-100) which has separated different liquid and gas phases. The top product stream (Stream 108) composed mainly from methanol in addition to water. However, the bottom liquid stream (Stream 107) contains mixture of reactions products and unreacted reactants as shown in Table 1. The adiabatic flash drum has separated 96% of the unreacted methanol from the reactor product stream (Stream 106).

Further methanol separation has been carried out using a distillation column (T-100) with 10 stages to provide sufficient separation [10]. Using distillation column, 97.8% of the unreacted methanol in the feed stream (Stream 107) has been separated in the top product stream (Stream 110). The bottom product (Stream 109), which mainly consists of unreacted triglycerides, produced methyl esters, fatty acids and glycerol, as shown in Table 1, has left the column at 253.9^oC and cooled to 25^oC for further separation processes. The unreacted methanol could be completely separated at temperatures higher than 278^oC. However, the column's reboiler temperature has not exceeded 253.9^oC for several reasons including avoiding thermal degradation of FAMEs that shows only stability up to 270^oC [29] and avoiding having traces of vaporised glycerol at the top stream where its boiling temperature is 280^oC. In addition, increasing the temperature from 253.9^oC to 270^oC has no significant increase in methanol recovery.

3.3. Glycerol separation

Separation of glycerol from biodiesel is considered as an essential purification step as the high content of glycerol could lead to storage problems due to phase separation, higher emission of aldehyde in combustion process and clogging of the fuel injector [30]. The separation processes that have been reported in previous studies involved several techniques including gravity settling and washing with water [13]. In this work, gravity settling using phase separator has been applied. The cooled bottom product stream from distillation column (Stream 109A) has been fed to the settling unit (phase separator). Glycerol has been separated in the bottom product stream (Stream 112), where biodiesel associated with the unreacted triglycerides has been separated in the middle product stream (Stream 113). About 99.9% of glycerol in the feed stream (Stream 109A) has been separated in bottom stream (Stream 112) associated with traces of unreacted methanol. Finally, as the influent stream to the separator does not include any gases, nothing has been reported at the top product stream (Stream 111).

3.4. Biodiesel purification

According to the European standard for biodiesel specifications, EN14214, maximum concentration of triglycerides in the pure biodiesel is 0.2% by weight [10]. In this study, the glycerol free biodiesel mixture stream (Stream 113) contains 8.38% by weight of triglycerides, where it exceeds the specification of EN14214. Accordingly, further purification process has been applied for biodiesel mixture stream in order to separate the residuals of triolein. Vacuum distillation unit (VDU) has been used to avoid any thermal cracking or degradation of FAMEs. Imahara et al [29] have reported that at high temperature, FAMEs show stability up to 270^{o}C, while beyond this temperature FAME starts to decompose due to isomerisation from *cis*-form to *trans*-form.

The feed stream has been de-pressurised using vacuum pump, which has been represented in the simulation environment as an expansion valve tool (VLV-101). Ten stages column has been used for the separation process [10]. The purified biodiesel stream (Stream 114) exits the column with less than 0.02% by weight of triolein, which is in agreement with the European standard biodiesel specifications, EN14214.

4. PROCESS INTEGRATION

Conservation of mass and energy in the developed industries has been considered as the most effective approach for sustainable design. Hence, implementation of HEN and MEN has gained a great interest in process engineering research through the last decades. The highlights of these researches are to minimise the external usage of energy, minimise waste discharge, minimise purchasing of fresh resources and to maximise the production of the desired product. All of these aspects are implemented through both energy and mass integration for the designed processes [23]

4.1. Mass integration

In the present study, the designed process has been subjected to different mass integration aspects. Firstly, optimising the reaction conditions has been applied experimentally as reported previously [26] by maximising the desired product and minimising reaction conditions. In addition, mass integration principles have been applied for the developed process. As the designed process did not include any mass exchanging units, mass integration would be only highlighted through minimising waste and fresh resources. The fresh resources used for this process are WCO and methanol. Methanol is considered as a major reactant, which is used in large excess in the non-catalytic transesterification reaction. Hence, minimising fresh and waste methanol is considered as an essential requirement for biodiesel integrated process.

In the existing process, two available sources streams for methanol have been observed including streams separated from both adiabatic flash drum unit and distillation column unit, i.e.; streams 108 and 110. On the other hand, there is only one sink that require fresh methanol, which is the reactor (CSTR-100). The required flowrate of fresh methanol for the process sink is 386 kgmol/h which is a massive amount to be purchased. Moreover, the reactor requires huge excess of methanol where waste methanol is considerably high. Consequently, using simple source-sink mapping shown in Figure 2, a proposed scheme for methanol recycling has been developed. The reactor required methanol with maximum composition of impurities of 5% where the available sources are having much lower impurities (<1%). Accordingly, simple recovery for both available sources has been implemented as shown in the process flow chart shown in Figure 1, where both sources streams have been mixed directly with the minimum required fresh methanol stream to be fed to the reactor. After applying this mass integration recycling, the actual fresh methanol used is only 27 kgmol/h (Stream 101) instead of 386 kgmol/h in case of having no recycling approach which represents 93% savings for the fresh methanol requirements.

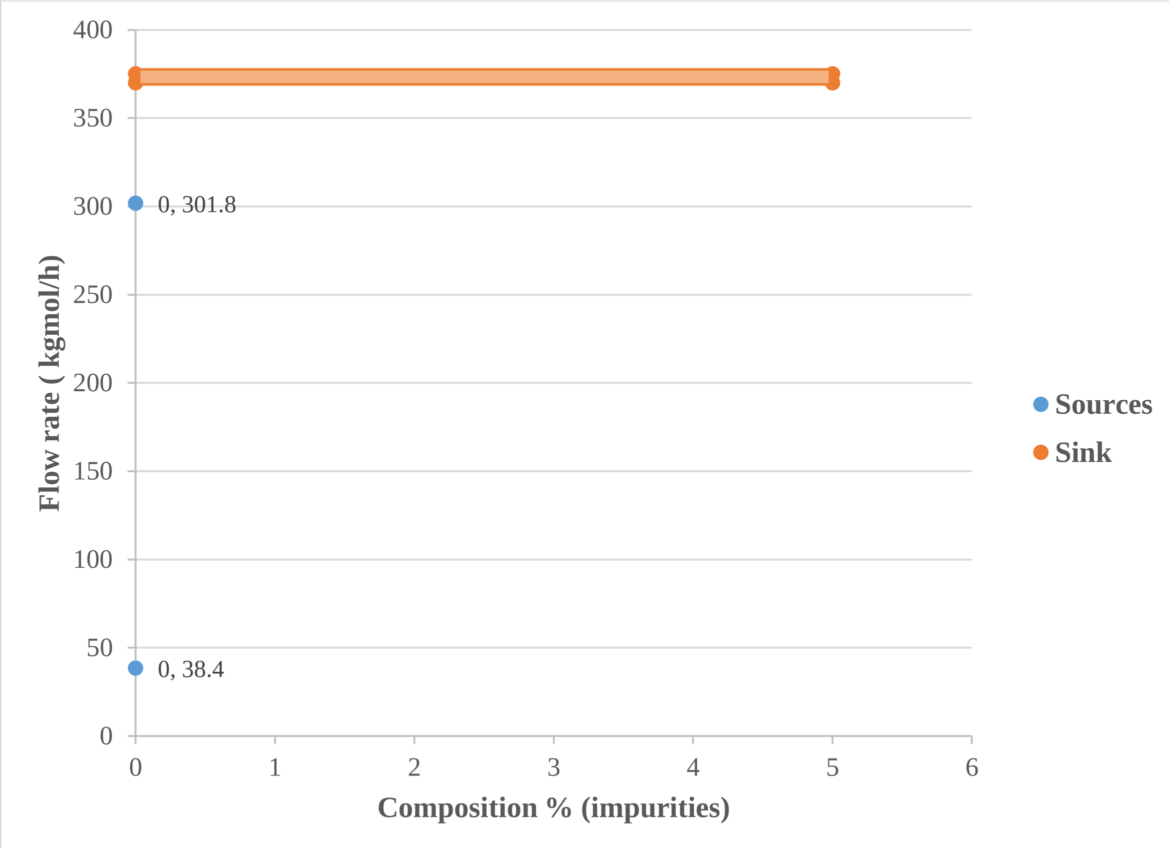

Figure 2. Source-sink mapping

4.2. Heat integration

Pinch technology has been used to integrate the energy required for both heating and cooling for all process streams. The list of the process hot and cold streams has been presented in Table 4. An assumption of ΔT_{min} of $10^{o}C$ has been proposed. Identifying Pinch temperatures would be proceeded using either problem table algorithm and/or heat composite curve. In the present study, the Pinch temperatures have been identified using the second method. Aspen Energy Analyzer® V8.8 simulation software has been used in identify the Pinch temperatures, minimum heating and cooling energy requirements and plotting composite curve for the process streams.

Table 4. Process hot and cold streams

Stream	Hot/Cold	Inlet T (ºC)	Outlet T (ºC)	C_p (kJ/kg.ºC)	Heat duty ($\times 10^6$ kJ/h)
104	↖	72.6	253.5	6.076	10.550
REB1	↖	124.1	256.4	3.053	4.9560
REB2	↖	224.5	226	176.660	10.960
108	↘	89	65	1,633.100	12.800
109	↘	253.9	25	2.416	5.858
110	↘	66.5	65	3.687	0.002
114	↘	80.4	25	2.009	0.622
115	↘	241.3	25	1.510	0.441
COND1	↘	66.5	66.4	9,012.800	0.961
COND2	↘	214.1	63.7	4.195	10.100

A composite curve for the process streams has been developed as shown in Figure 3. The overlap between hot and cold composite curves represents the prospective integration between hot and cold streams according to Pinch rules [31]. The minimum energies required for both heating (Q_h) and cooling (Q_c) have been observed from Figure 3 as 4,108 kW and 5,400 kW, respectively.

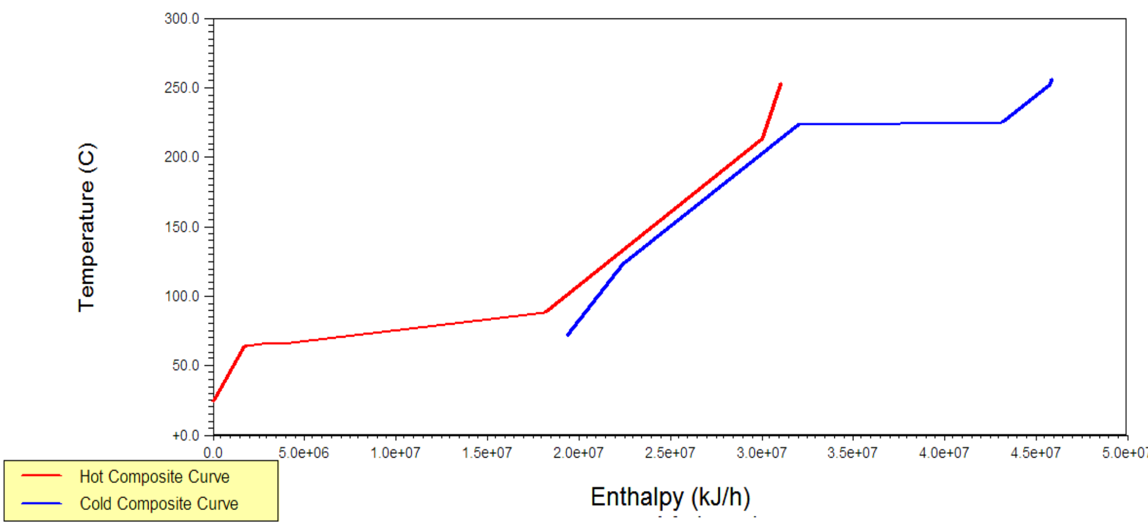

Figure 3. Composite curve of the process streams

In an attempt to minimise the process energy consumption and increase energy integration between process streams, design of a new HEN design has been developed using graphical Pinch Analysis method using only 5 heat exchangers as shown in Figure 4. Using numerical matching in graphical method eases the process of exchanger streams' selection and streams splitting. In addition, it investigates the validity of the exchangers according to Pinch rules. The designed exchangers have been analysed graphically where the exchangers fulfil the method guidelines as shown in Figure 5. The graphical method shortened the trial procedures that would be applied to achieve the optimum network using conventional Pinch methods. Consequently, the developed HEN has resulted in achieving 100% of both minimum heating and cooling energies requirements.

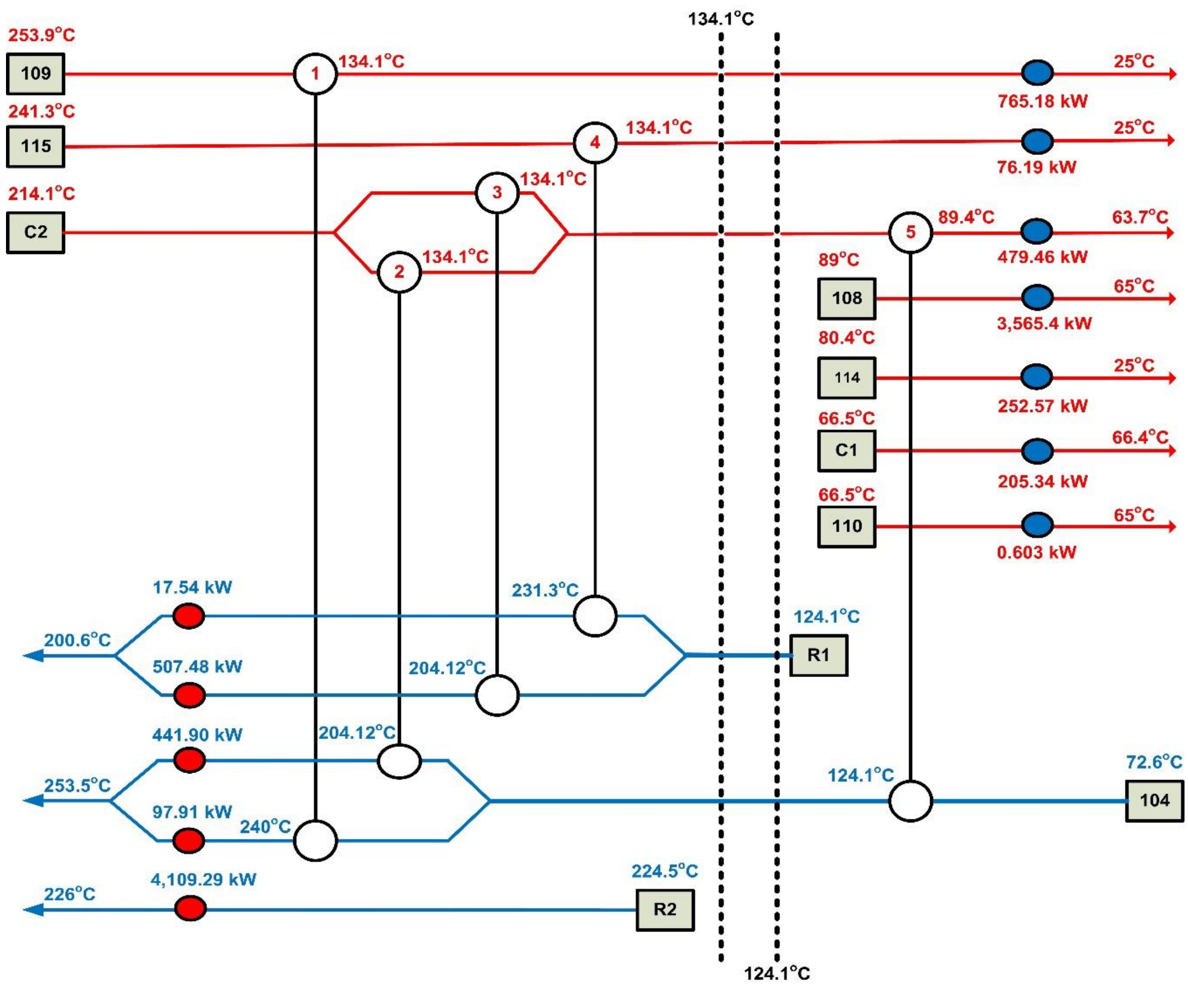

Figure 4. Heat exchanger network designed for the integrated process

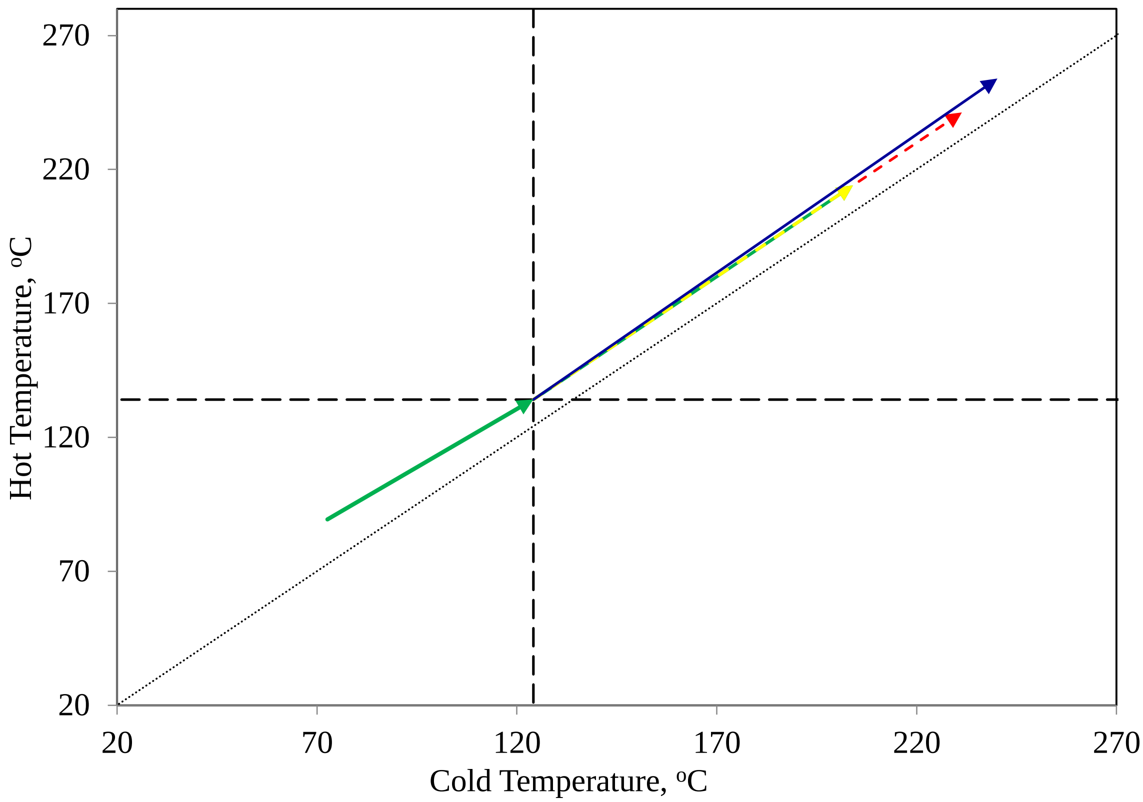

Figure 5. Graphical representation of the designed HEN on T-T diagram

These results have been compared with the automated designs developed by Aspen Energy Analyzer software. It has been observed that the optimum automated design has used 6 heat exchangers and achieved 118% and 113.9% of the minimum heating and cooling energies requirements, respectively. This implies that the automated design consumes more energy than the targets. Graphical Pinch method has been used to investigate the proposed automated design and to highlight the problems associated with the design using simple and quick observations. Figure 6 illustrates a graphical representation of the proposed design using graphical Pinch method on T-T plot. It could be easily observed that 3 of the proposed automated exchangers are existing within the non-optimum integration area where a revamping design is required to relocate the exchangers within the optimum integrating areas.

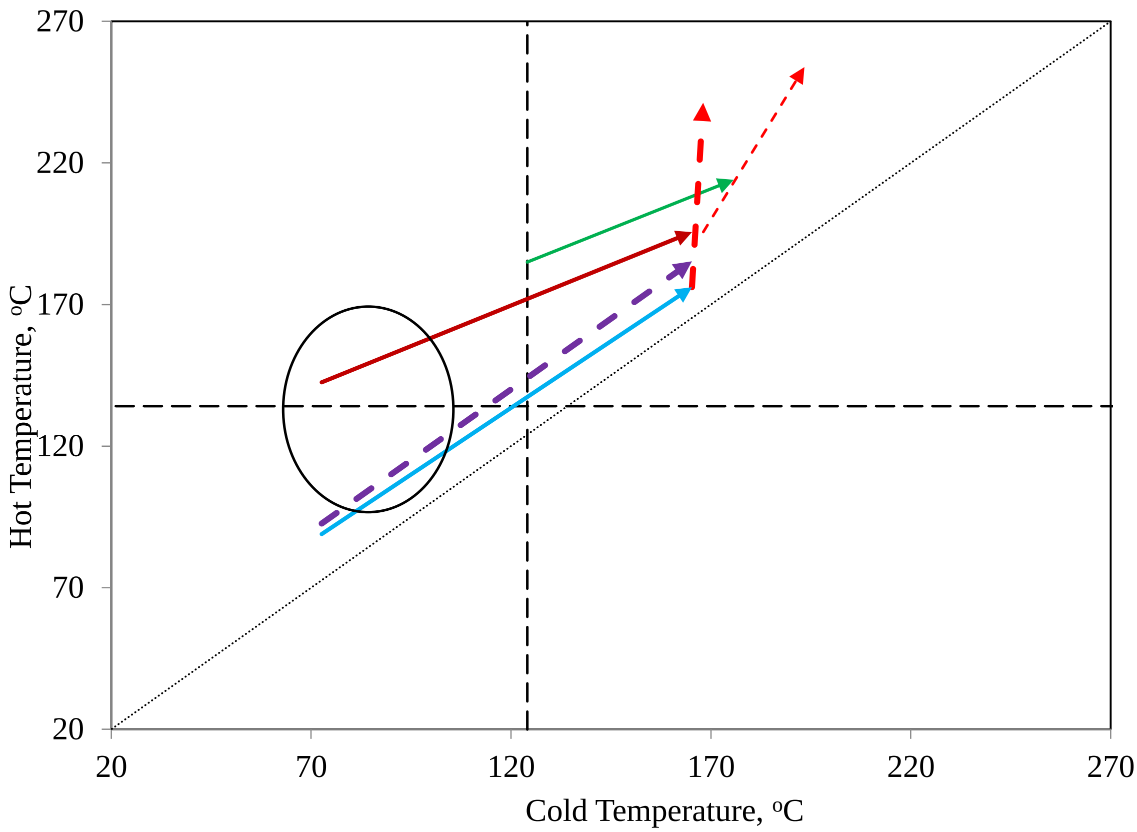

Figure 6. Graphical representation of the automated proposed HEN on T-T diagram

Lee et al [10] have designed an energy integrated process for biodiesel using supercritical methanol. They have included only 2 heat exchangers to the process HEN. Their developed HEN has been also analysed using graphical Pinch method as shown in Figure 7. The simple designed HEN includes two exchangers with major problems. The first exchanger is a network Pinch exchanger where the developed straight line representing the exchanger touches the Pinch line as shown in Figure 7. This elaborates that the exchanger is not fulfilling Pinch rules with insufficient minimum heat transfer temperature difference. When an exchanger touches the Pinch line, it indicates that the process streams temperature difference is equal to zero and accordingly, an inefficient exchanger. On the other hand, the second exchanger has been included within the non-optimum integration area and the temperatures are crossing the Pinch line.

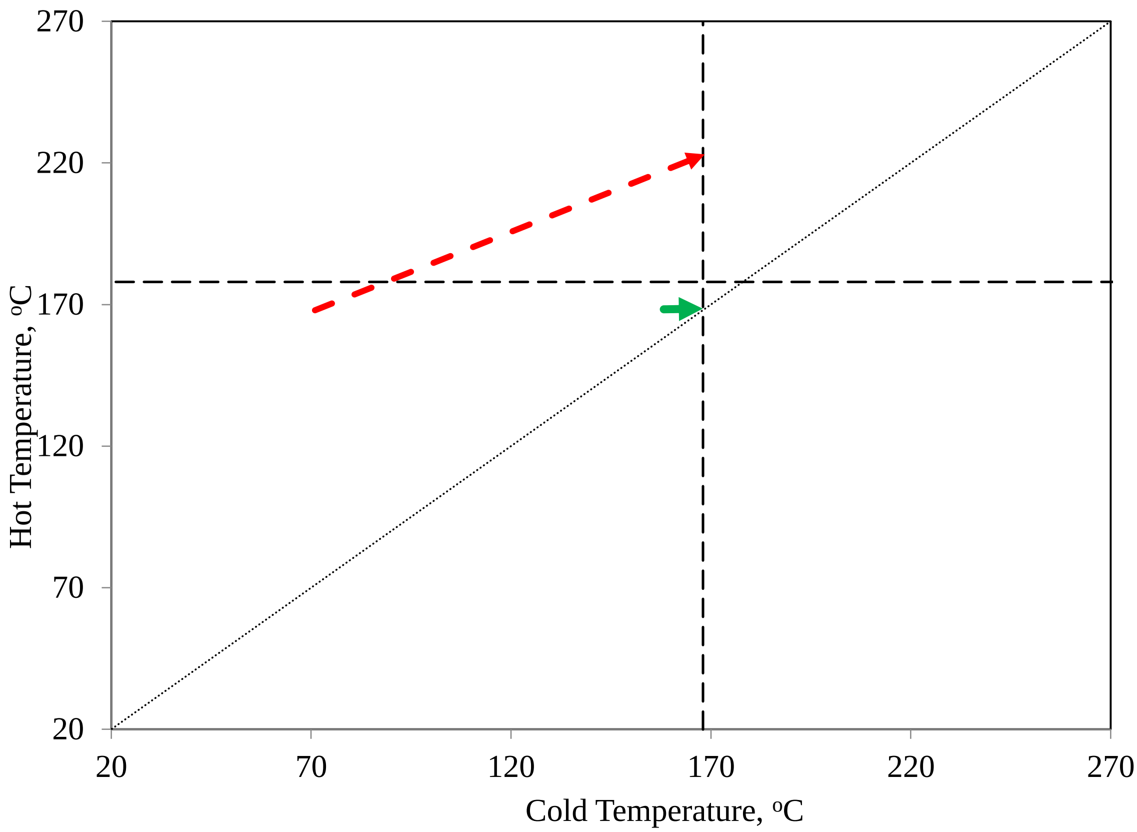

Figure 7. Graphical representation of the literature proposed HEN on T-T diagram

These results exemplify the significance of using graphical Pinch method in analysing existing HENs where the inefficient exchangers would be easily observed. In addition, using graphical matching technique for process streams simplify the integration procedures where it could be implemented to match many streams in relatively short time compared with the conventional methods.

5. CONCLUSIONS

In this work, an integrated process for non-catalytic biodiesel production from WCO using supercritical methanol has been simulated. The process has been designed where the produced biodiesel relies in agreement with the European Standard for biodiesel specifications, EN14214. The developed process has been subjected to both mass and energy integration to minimise the fresh methanol requirements and to minimise the external required energies for heating and cooling, respectively. Methanol recycling strategies have contributed to minimise fresh required methanol. Graphical Pinch method has been implemented to design n optimum HEN using numerical matching strategy. The designed HEN has achieved 100% of the targeted optimum required energies.

ACKNOWLEDGMENT

The authors would like to acknowledge The British Council UK and STDF Egypt through the Institutional Links Newton-Mosharafa Programme (Project IDs 261862377 and 27738) for funding this research.

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