P-ISSN 2586-9000
E-ISSN 2586-9027
Homepage : https://tci-thaijo.org/index.php/SciTechAsia
Science & Technology Asia
Vol.24 No.1 January - March 2019
Page: [30-40]
Original research article

Pinch Analysis of a Sugarcane Wax Extraction and Purification Process

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Received 25 October 2018; Received in revised form 8 January 2019
Accepted 19 December 2018; Available online 8 February 2019

ABSTRACT

This work analyzes heat integration of sugarcane wax extraction in purification plants. It focuses on the development of sugarcane wax extraction from the mud of sugar mill presses, using supercritical CO_2 as a solvent and ethanol as a co-solvent, with further sugarcane wax purification of 25 kg per day. To improve heat recovery for the designed heat exchanger networks (HEN), pinch analysis was applied to achieve optimal, total, external hot and cold utility demands at the optimum minimum temperature difference of 15°C. Generated HEN alternatives were selected based on an incremental net present value (ΔNPV). Compared to the current HEN design, the most cost-effective HEN could decrease total annual costs by a factor of 1.2.

Keywords: Pinch analysis; Heat exchanger network; Sugarcane wax extraction; Supercritical CO_2; Simulation.

1. Introduction

Thailand is one of world's top ten sugar exporters. By 2026, the Ministry of Industry of Thailand expects to produce 180 million tons of cane, or a 70% increase. Co-products from sugarcane production are bagasse, molasses, and press mud. When one ton of cane is crushed, raw cane juice is passed through clarification steps, yielding from 36 to 40 kg of press mud or filter cake [1]. Press mud is composed of 50–65% moisture, 5–12% sugar, 20–30% fiber, and 7–15% crude wax and inorganic salts [2]. Although press mud can be used as a fertilizer and as a substrate for biogas production, it also can be extracted to produce value-added products, such as wax. Sugarcane wax is composed of numerous complicated elements, such as alkanes, hydrocarbons, fatty acids, ketones, aldehydes, alcohols, esters, and steroids [1]. Sugarcane wax, a potential substitute for carnauba wax, candelilla wax, and chitosan, is used in many applications, such as

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doi: 10.14456/scitechasia.2019.4

cosmetics, paper coating, textiles, edible coatings, leather sizings, lubricants, adhesives, polishes, and pharmaceutical industry products [3].

Conventionally, extractions use the relatively time- and solvent-consuming Soxhlet method. New extraction techniques, such as microwave-assisted extraction (MAE) [4], supercritical fluid extraction (SFE) [1], ultrasonic-assisted supercritical CO_2 extraction (USC-CO_2) [5] and accelerated solvent extraction (ASE) [6], have been developed to reduce the time and solvent consumption of conventional extraction methods. MAE depends on the absorption capacity of selected solvents under microwave irradiation. Because the efficiency of the MAE technique is enhanced at higher temperatures, it is not suitable for thermolabile substances. If solvent extracted compounds are non-polar, MAE is not efficient because it does not absorb energy [7]. SFE uses strong, non-toxic, non-flammable, CO_2 as an extracting solvent to selectively extract natural products. The use of CO_2, under supercritical conditions, has been widely expanded in the isolation of compounds from natural materials. The optimal condition of supercritical CO_2 extraction of psoralen and isopsoralen from Fructus Psoraleae (Psoralea corylitolia L.), under a pressure of 26 MPa and a temperature of 60°C, yielded 9.1% [8]. The USC-CO_2 gave higher curmin content in turmeric extraction yield than conventional Soxhlet extraction [5]. When the target substances are polar, the addition of a co-solvent, such as water or ethanol, improved extraction yield by about 5–10% [7]. The CO_2 completely miscible with ethanol gives high supersaturation and fast crystal particle formation for the precipitation of andrographolide from *Andrographis paniculata* extracts [9]. Alternatively, ASE is an environmentally friendly technique using a small amount of organic solvents at high temperature (50–200°C) and pressure (5–200 atm). To reduce the extraction time and to improve the penetration efficiency of the extraction solvent, the solvent is maintained in a liquid phase by elevating its temperature. Compared to MAE, manufacturing costs associated with SFE and ASE processes are higher, due to the required high operating pressure. Despite advantages of ASE, only preliminary results of bench-scale sugarcane wax extraction were reported in the previous study [6].

Pinch analysis, a technique for reducing the capital and utility cost, is proposed to synthesize heat exchanger networks of chemical processes [10, 11], bioprocess plants [12], and polymerization and drying plants [13].

The wax extraction process has been studied from the viewpoint of energy and exergy. Pinch techniques were applied to save USD 1.74 million in sub-cooling operating costs over a ten-year period at a gas processing plant [13]. The pinch method was applied for a thermo-economical evaluation of CO_2 recycling systems in the integration of a β-ecdysone SFE plant at a sugarcane bio-refinery. The best techno-economic operating condition was obtained at 40 bar and a temperature of 30°C for a stand-alone SFE process, using CO_2 separation in the second flash tank [15].

Therefore, the sugarcane wax extraction process was scaled up from the pilot-scale wax extracting experiment of Paoborome [16] to a commercial-scale plant, using eco-friendly SFE technology. To design more energy-efficient processes, pinch analysis is used to develop energy-saving heat systems for industrial SFE sugarcane wax extraction and purification processes, with incremental net present values.

2. Methodology

2.1 Process Description

Fig. 1 shows an industrial-scale SFE plant, using supercritical CO_2 as solvent and ethanol as co-solvent, in an Aspen plus®

commercial simulator and thermodynamic model PRMHV2. Press mud was prepared by crushing and removing its moisture content at 60°C. A mass flow of 800 kg prepared press mud/day was introduced into three extractors, operating in parallel. 2,610 kg CO_2/day and 3,520 kg ethanol/day were pumped into the extractors at the desired proportions, pressure, and temperature reported in Paoborome's pilot-scale sugarcane wax extraction [16]. After extraction, CO_2 and ethanol were separated from the extract for further recovery and recycling. The separated CO_2 through one flash tank, operating at 1 atm, 25°C, was cooled until liquefied and then recycled into the process. CO_2 was recovered by feeding 66.23% of the CO_2 into the flash tank. Ethanol was separated from the extracted sugarcane wax by vacuum evaporation, operating at 0.1 atm, 70°C, to prevent product oxidation, and then recycled into the process. Ethanol was recovered by feeding 99.61 % of the ethanol into the evaporator. It was necessary to feed makeup ethanol and CO_2 because of their loss in the extraction unit. The names of Aspen equipment, shown in Fig. 1, are: flash tanks (F101, F102, and F103), extractors (E101, E102, and E103), vacuum evaporation (H105 and F104), cooling or heating operations (H101, H102, H103, H104, H106, H107, and H108), compressors (P101, P102, and P103), and valves (V101, V102, V103, and V104).

2.2 Calculations for Pinch Analysis

Pinch analysis, proposed by Linnhoff [17], applies the first and second laws of thermodynamics to minimize the energy consumption process by calculating feasible energy targets. A minimum temperature difference between hot and cold streams determines the optimal capital cost of integrating heat exchangers and energy demands. The thermal integration calculation, based on a material and energy balance model, was performed using Aspen plus® and an Aspen Energy Analyzer™. Alternative designs of the heat exchanger network (HEN) were provided by optimizing the heat recovery system and energy supply methods.

2.3 Economic Evaluation

Table 1 shows cost used to determine the economic evaluation. Based on pinch analysis of the SPE process, the percentage of energy savings and the net present value were calculated to support decisions about suitable heat exchanger networks for the waste heat recovery system. All cost estimates were updated using the Marshall and Swift index.

The typical capital cost (C_C) of a single heat exchanger is expressed as:

$$C_C = a + bA^c \tag{1}$$

where a, b, and c are cost coefficients that depend on construction materials. For a stainless steel shell and a tube heat exchanger, the corresponding coefficients are 30800, 1644, and 0.81. A is the surface area (m^2).

The investment is estimated based on the following assumptions [18].

1. Only investment for the required additional heat exchanger area is considered as expressed in Eqs. (2) - (5) [19].

2. Piping and other costs are not taken into account for economic assessment.

3. Heat exchanger averages are estimated from the base-case heat exchanger area, and the number of shells is assumed to be one shell pass.

4. Energy prices are fixed for the entire project life.

Investment cost and cost savings calculated using corresponding Eqs. (2) – (5) and Eq. (6) are applied to measure the relative cost of alternative or new HEN design to a base-case HEN design ratio. A base-case HEN design is selected on the basis of minimum requirements of external utilities, HEN area, and the number of heat exchanging units.

$$\text{investment} = \Delta N\left(a + b\left(\frac{\Delta A}{\Delta N}\right)^c\right) \quad (2)$$

$$\Delta N = \frac{\Delta A}{avg_{\text{shell}}} \quad (3)$$

$$avg_{\text{shell}} = \frac{\Delta A}{N_{\text{shell}}} \quad (4)$$

$$\Delta A = A_{\text{new, HEN}} - A_{\text{base, HEN}} \quad (5)$$

where A is the area of the heat exchanger (m^2), $A_{\text{base, HEN}}$ is the total base-case HEN area (m^2), $A_{\text{new, HEN}}$ is the total alternative HEN area (m^2), avg_{shell} is the average size of the exchanger shell, and N_{shell} is the number of shells.

Annual cost savings and the net present value are calculated as follows:

$$S = \left(\sum \text{HU}_{\text{base cost}} + \sum \text{CU}_{\text{base cost}}\right) - \left(\sum \text{HU}_{\text{new cost}} + \sum \text{CU}_{\text{new cost}}\right) \quad (6)$$

where $\text{CU}_{\text{base cost}}$ and $\text{CU}_{\text{new cost}}$ are base-case and alternative-case cold utility costs (USD), respectively.

$\text{HU}_{\text{base cost}}$ and $\text{HU}_{\text{new cost}}$ are base-case and alternative-case hot utility costs (USD), respectively, and S is cost savings (USD).

The cost savings percentage (%S) is calculated using base-case utility costs.

$$\%S = \frac{\left(\sum \text{HU}_{\text{base cost}} + \sum \text{CU}_{\text{base cost}}\right) - \left(\sum \text{HU}_{\text{new cost}} + \sum \text{CU}_{\text{new cost}}\right)}{\left(\sum \text{HU}_{\text{base cost}} + \sum \text{CU}_{\text{base cost}}\right)} \quad (7)$$

Incremental cash flows, associated with capital budgeting and cost savings projects, evaluate economic feasibility investment decisions, among multiple alternatives.

Incremental net present values (ΔNPV) for each HEN are calculated using Eq.(8) [12].

$$\Delta\text{NPV} = S \cdot \frac{1 - (1+i)^{-n}}{i} - \text{investment} \quad (8)$$

where n is the project's life (y), and i is the annual interest rate. Moreover, investment and savings (S) are calculated from Eq. (2) and Eq.(6), respectively.

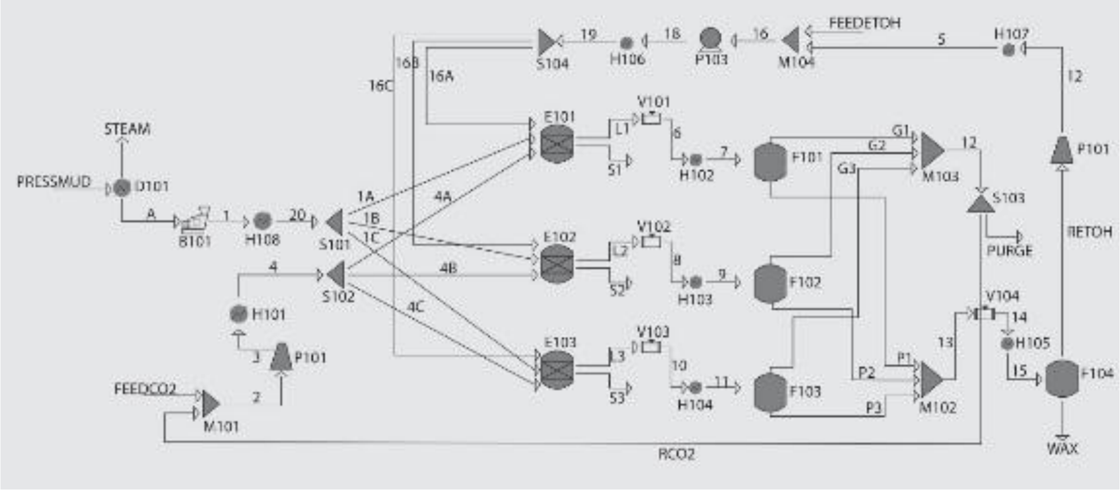

Fig. 1. A Pilot-Scale Sugarcane Wax Extraction Plant, Using Supercritical CO2 as a Solvent and Ethanol as a Co-solvent.

3. Results

3.1 Process Simulation of Supercritical Extraction and Purification of Sugarcane Wax

After sugarcane wax extraction and purification, the wax yield of 4.75% or 25.6 kg/day was achieved with an initial feed of 800 kg press mud/day, 2,610 kg CO_2/day and 3,520 kg ethanol/day. The simulated yield was consistent with the sugarcane wax yield obtained from the pilot-scale extracting experiment of Paoborome [16].

3.2 Pinch Analysis

The range target shown in Fig. 2 provides the minimum total cost of the HEN, corresponding to the optimal values of the minimum approach temperature. By varying the minimum approach temperature, the total HEN costs are calculated using the best trade-offs among utility requirements, heat exchanger areas, and unit shell numbers.

As shown in Fig. 3, the composite curve of both hot and cold streams must be constructed on a temperature-enthalpy plot before the HEN is designed to provide information about maximum heat integration and minimum hot and cold utility requirements. The closest temperature difference between the two curves is called the pinch point, indicating the minimum temperature between hot and cold streams ($\Delta T_{\min}$) for exchanging heat. Fig. 3 shows pinch points at 127.5°C and 112.5°C. The smaller the $\Delta T_{\min}$, the more the heat can be transferred in the heat exchanger, resulting in a larger required heat exchanger area. For larger values of $\Delta T_{\min}$, heat recovery in the heat exchanger decreases, while the external utility requirement increases. Fig. 2 shows insignificant changes in the total cost of HEN in the $\Delta T_{\min}$ range from 2°C to 32°C. Moreover, the choice of $\Delta T_{\min}$ should be based on economic considerations and experience [21]. For industrial chemical plants, $\Delta T_{\min}$ values typically range from 12–20°C [14]. Hence, heat integration and exchanges between hot and cold utilities of the pilot-scale SFE plant were calculated using a $\Delta T_{\min}$ of 15°C.

After identifying the pinch temperature and the minimum demand of heating and cooling utilities, a grand composite curve (Fig. 4) was generated to illustrate the excess heat available for the process, within each temperature interval for utility selection, by minimizing the use of expensive utilities. Fig. 4 shows the minimum requirement of external hot utility of 2.26×10^5 kJ/h to increase temperature from 300°C to 650°C on the hot scale while the cold utility was not externally required.

After performing a pinch analysis, the thermal optimization problem, comprising three cold streams and six hot streams, is shown in Table 2. As shown in Table 3, the HEN design of the current SFE plant shown in Fig. 1 and alternative designs (options A, B, C, and D) provided the surface area required and the annual cost estimates for the specified heat duty.

Table 1. Cost Data Used for the Economic Evaluation

Economic data	Value	Unit
Project lifetime	20	Years
Construction and start up	2	Years
Interest rate	10	% per annum
Annual working day	320	days/year
Marshall and Swift index	1593.7	
Hot utility cost (Lower steam pressure)	6.84×10^{-3}	USD/kWh
Hot utility cost (Fired heat)	1.53×10^{-3}	USD/kWh
Cold utility cost (Refrigerant 1)	9.85×10^{-3}	USD/kWh (−25°C)
Cold utility cost (Air)	3.6×10^{-6}	USD/kWh
Capital cost of heat exchanger [20]	$C_C = 30800 + 1644A^{0.81}$	

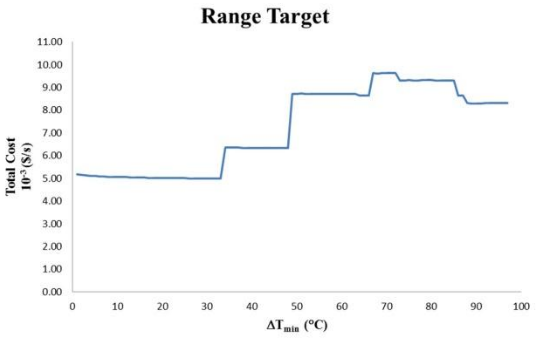

Fig. 2. A Plot of Range Target.

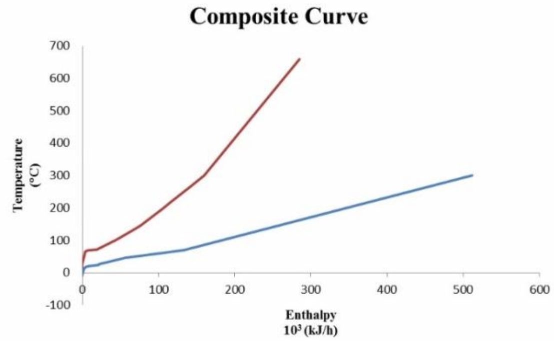

Fig. 3. Composite Curve for the Sugarcane Wax Extraction and Purification Process.

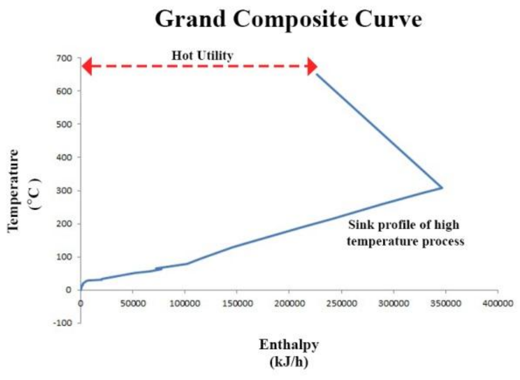

Fig. 4. Grand Composite Curve for the Sugarcane Wax Extraction and Purification Process.

3.3 Thermal Process Integration

As shown in Figs. 5(a) - 5(e), HEN options were designed to reach individual energy targets using pinch analysis, with a temperature difference of 15°C. External utilities are represented by thin lines. External cold utilities are Air and Refrigerant 1, and external hot utilities are Fired Heat and LP steam. Figs. 5(a) - 5(e) show that two heat exchangers performed internal heat recovery (white matches), two heat exchangers used external cold utility to adjust the temperature of two hot streams (blue matches), and two heat exchangers used external hot utility to adjust the temperature of two cold streams (red matches). The amount of hot and cold utilities in the current HEN design was 142.0 kW and 79.2 kW, respectively, while that of the other alternatives were 62.9 kW and 0 kW, respectively.

4. Discussion

For the result of HEN designed by Aspen Energy Analyzer™ shown in Table 3, considering all five designs, option D showed the smallest heat transfer area and the lowest required annual cost. Therefore, as shown in Table 4, option D was used as the base case for calculating cost savings in Eq. (7) and ΔNPV in Eq. (8). A positive cost saving percentage indicates that the base case D is less expensive than that of current and other alternative HEN designs. On the other hand, a negative cost savings percentage indicates that the base case costs more than other HEN designs. Option D spends 8.28% more on utilities than option A, and 7.16% more than option B. Utilities for Option D cost 121.27% less than the current design, while options D and C spend equally on utilities cost. The NPV difference between base case (option D) and an individual HEN design option allows the selection of a HEN design with maximum net cost savings. Table 4 shows all positive ΔNPV for 2017, when option D is used as a base case. This suggests that the investment using HEN design D is more economically feasible.

Table 2. Stream Data

Steam Name	Steam	Condition	T_{in} (°C)	T_{out} (°C)	Enthalpy (kW)
3 to 4	C1	Cold	658.2	70.0	226,392.30
1 to 20	C2	Cold	300.0	70.0	33,366.16
17 to 5	C3	Cold	198.0	25.0	25,262.65
D101_heat	H1	Hot	25.0	300.0	452,447.49
14 to 15	H2	Hot	11.2	70.0	20,727.50
18 to 19	H3	Hot	44.9	70.0	33,526.12
6 to 7	H4	Hot	−7.8	25.0	1,537.30
10 to 11	H5	Hot	−8.7	25.0	1,600.93
8 to 9	H6	Hot	−7.8	25.0	1,537.30

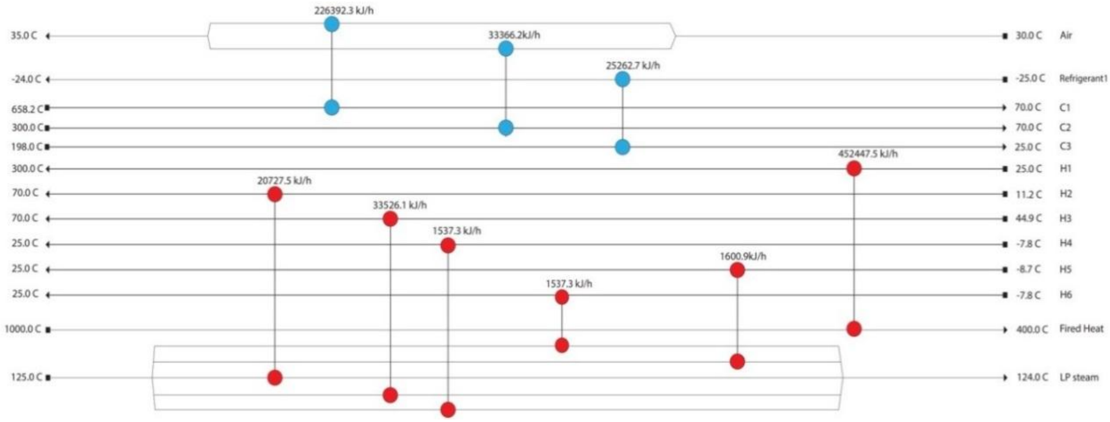

Fig. 5(a). HEN Configuration of the Current Design.

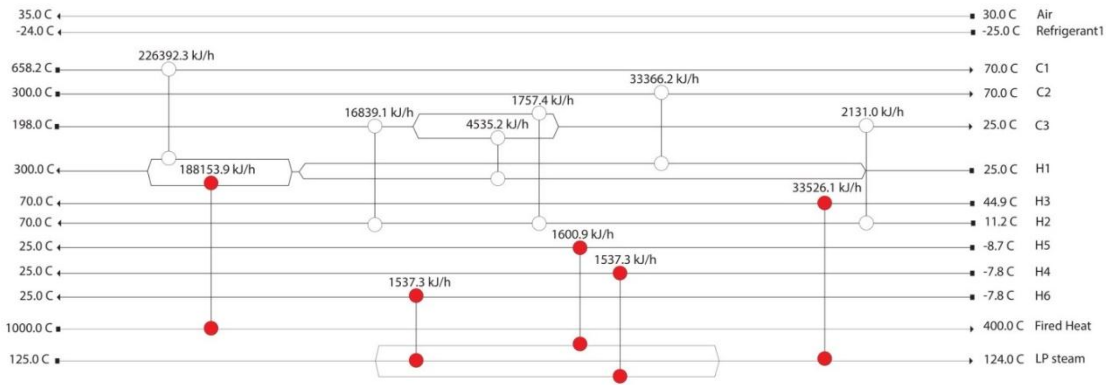

Fig. 5(b). HEN Configuration of Option A.

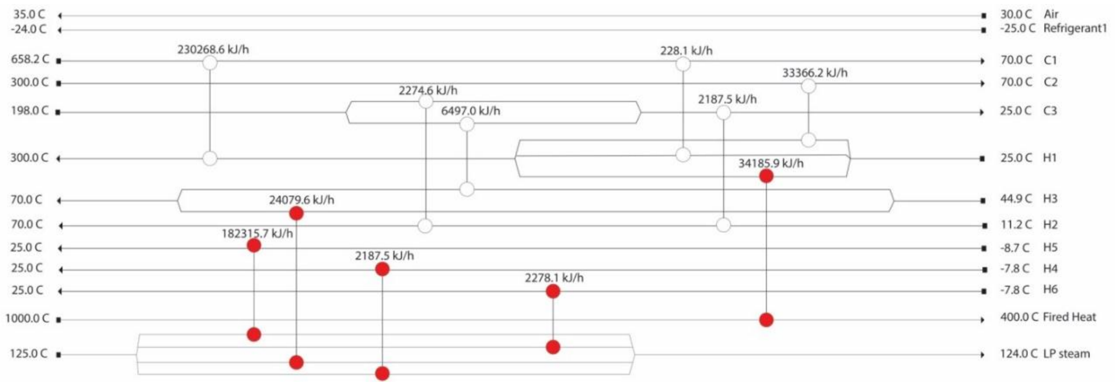

Fig. 5(c). HEN Configuration of Option B.

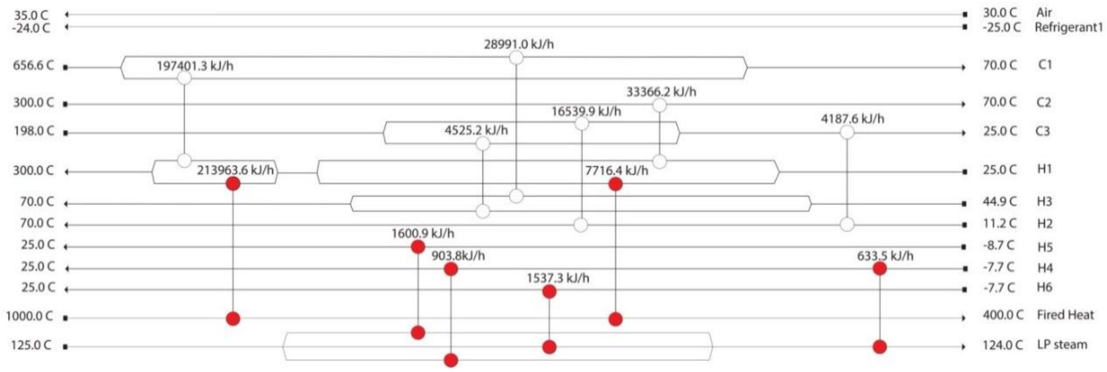

Fig. 5(d). HEN Configuration of Option C.

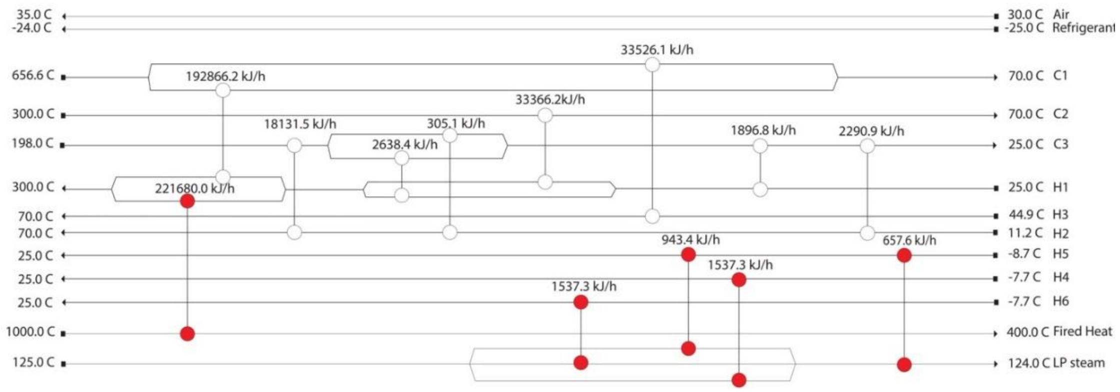

Fig. 5(e). HEN Configuration of Option D.

Table 3. Heat Exchanger Network Designs Using a Pinch Analysis.

Design	N_{shell}	$N_{\text{heat exchanger}}$	$\sum A_i$ (m^2)	Annual utility cost (USD/y)
Current design	11	9	9.69	18,430.28
Option A	17	11	9.91	7,639.33
Option B	23	11	16.52	7,732.65
Option C	19	12	9.48	8,329.24
Option D	19	13	9.27	8,329.24

Table 4. NPV Comparison of Design Options for Option D.

Comparison of options:	Investment (USD) using Eq. (2)	Savings (%) using Eq.(7)	ΔNPV (2017) (USD) using Eq.(8)
Option D versus current design	17,750.97	121.27	103,746.81
Option D versus option A	40,765.07	−8.28	34,317.56
Option D versus option B	623,608.03	−7.16	618,032.68
Option D versus option C	14,990.93	0	14,990.93

Note. HEN design of option D was selected as the base case HEN design.

Based on our economic analysis, option D was the most economically feasible investment. As shown in Fig. 5e, the available C1 stream used a portion of its flow rate capacity to partially heat the H1 and H2 streams. Similar to the C3 stream, it split to supply a portion of the capacity flow rate of the H3 stream to cool down temperatures to 25.0°C. The capacity flow rate of the H2 stream was distributed to the C2 and C3 streams to raise the temperature to 11.2°C and 70.0°C, respectively.

5. Conclusion

Heat exchanger networks (HEN) in a developed, commercial-scale sugarcane wax extraction and purification plant, using supercritical CO_2 as the solvent and ethanol as the co-solvent, can be successfully designed and achieve energy targets obtained by heat integration analysis of the pinch method, using Aspen plus® embedded with an Aspen Energy Analyzer. To achieve minimum surface area and total costs required for option D, the current HEN was improved using a stream-splitting approach.

Acknowledgement

The authors are grateful to the Faculty of Engineering, Thammasat University, Thailand for their financial support.

Nomenclature

A	the surface area of the heat exchanger (m^2)
$A_{\text{base,HEN}}$	the total base-case heat exchanger network (HEN) area (m^2)
$A_{\text{new,HEN}}$	the total alternative HEN area (m^2)
avg_{shell}	the average size of the exchanger shell
ASE	accelerated solvent extraction
C_C	capital cost of a single heat exchanger
$CU_{\text{base cost}}$	base-case cold utility cost
$CU_{\text{new cost}}$	alternative-case cold utility cost
HEN	heat exchanger network
$HU_{\text{base cost}}$	base-case hot utility cost
$HU_{\text{new cost}}$	alternative-case hot utility cost
N_{shell}	the number of shells
i	annual interest rate
investment	investment cost of HEN

LP	lower steam pressure
n	the project's life (year)
ΔNPV	incremental net present value
MAE	microwave-assisted extraction
S	annual cost savings
%*S*	cost savings percentage
SFE	supercritical fluid extraction
ΔT_{min}	the minimum temperature between hot and cold streams

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