Research article

An energy-friendly alternative in the large-scale production of soybean oil

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ABSTRACT

Soybean oil is widely used as cooking oil, whereas the soybean cake is a valuable ingredient for animal food. The extraction of soybean oil is an energy-intensive process, with additional significant impact on the environment via the wastewater and hexane emissions. The research investigated different ways to minimize the energy consumption. In a traditional process, both direct (live) steam and indirect steam heating (jackets, tubular exchangers) are used to deliver the required heat duty. Direct steam injection is restricted to the first evaporator and the stripper, for a total of 620 kg/h. Indirect steam is also applied in the evaporators for a total of 6.44 MW. The desolventizing process requires a steam energy input of 8.15 MW. Integration of a heat exchanger network in the evaporation and stripping part of the process reduces the amount of direct steam usage from 620 kg/h to 270 kg/h and of the indirect heat duty from 6.44 to 5.05 MW. In the cake desolventizing part of the process, the energy requirement is reduced from 8.15 to 2.12 MW. The overall gross energy saving is hence ~50%. The improvements moreover reduce both the waste water loadings by 56.5% and the CO_2 emissions by 62.5%. Hexane emissions are moreover significantly (> 90%) reduced.

1. Introduction

Vegetable oils are commonly used for different end applications. Whereas edible vegetable oils, such as soybean, peanut, palm and sunflower oils are used in food commodities. (Alam et al., 2014), they also find applications in biodiesel or even bio-aviation fuel via transesterification and other secondary transformation processes. (Issariyakul and Dalai, 2014; Lin et al., 2009; Wang et al., 2017; Yun et al., 2013). There are also a number of oils, such as rice bran oil, palm kernel oil and sheanut butter, that are used in specific markets with added value for cosmetics and pharmaceuticals. (Cargill, n.d.)

Soybeans are used worldwide for their oil and protein contents. In 2017, the world production of soybeans was 340 million ton (33.9% in America, 32.1% in Brazil, 16.2% in Argentina). The world production of soybean oil was 56 million ton. The high content of poly-unsaturated components in soya has positive effects on the human health, by e.g. reducing the blood lipids, preventing blood clots, and boosting immunity. The cultivation of soybeans moreover captures nitrogen from the air and transforms it into N-fertiliser. The main components of the soybean are proteins (36 wt%), carbohydrates (30 wt%), fat/oil (20 wt%) and water (9 wt%). The composition of the soybean oil is given in Table 1 (Carrín and Crapiste, 2008).

After an extraction with hexane, two streams are produced, i.e. a liquid phase (miscella: soybean oil and n-hexane) and a residual solid phase (soybean cake containing hexane).

The oil cake consists of proteins and carbohydrates. Soybean meal is produced from the oil cake after removing and recovering hexane. Animal feed is the main application for this meal (98%), where it is very popular for its low price, high protein content, and exceptional source of essential nutrients. During the pretreatment of the soybeans, high temperatures can cause denaturation of some proteins, and thus a loss of functionality. (Cerutti et al., 2012).

In earlier papers, Kong et al. (Kong et al., 2018, 2017a) studied the extraction and treatment processes for a small-scale Rice Bran Oil (RBO) plant. Several measures of energy reduction were investigated and a novel devolatilisation method of the cake using nitrogen was introduced and experimentally studied. The present research expands these findings to a large-scale soybean oil process, with special emphasis on the improved energy efficiency and on the pollution prevention as a result of the process improvements.

Table 1
Composition of soybean oil.

Component	Percentage (%)
Alpha- linoleic acid (polyunsaturated)	7–10
Linoleic acid (polyunsaturated)	51
Oleic acid (monounsaturated)	23
Stearic acid (saturated)	4
Palmitic acid (saturated)	10

2. Description of the traditional process

As previously detailed by Kong et al. (2018), there are two main methods for producing vegetable oils, either by pressing or extrusion at small to moderate capacities, or by solvent extraction of the pretreated oil seeds when larger capacities are needed. The solvent for extraction is usually hexane, and the first stages of extraction can use a mixture of solvent (hexane) and oil, called miscella. Fig. 1 illustrates the conventional setup for the extraction of soybean oil (SBO). The oil production process for different oil seeds is quite similar and typically consists of the following steps. First the oilseeds are delivered, and a pretreated (drying, dehulling, flaking). Subsequently the actual leaching process takes place. Crude SBO is extracted on a belt conveyor by hexane. This is followed by the removal of hexane from both the miscella and from the cake. The miscella is subjected to a two-step evaporation and a direct steam stripping. Soybean meal is thermally treated in a multiple hearth desolventizer/toaster/cooler. Solvent and water vapors are then condensed in a multi-step cooling. Residual vapors are finally stripped of hexane by absorption in paraffin. Hexane is recovered and can be reused in the leaching process (Martinho et al., 2008).

The process is very energy intensive by the large amount of steam and cooling water needed, mainly in the solvent removal/recovery sections of the process. In addition, the supply and storage of hexane as well as diffusive losses from the waste water treatment, pipes, pumps, valves and vents create considerable atmospheric emissions of hexane. The waste water still contains trace quantities of hexane (< 0.2 vol% and < 0.014 vol% respectively). At the end of the process, solvent free products must remain. Reductions in steam and cooling water consumptions as well as reducing the environmental effects of the process are important economic targets in improving solvent extraction plants. In order to meet these objectives, the present paper investigates

possible methods to improve the traditional process by: (i) examining and minimizing the steam and heat duty requirements in the evaporator-stripper part of the extraction process by performing a sensitivity analyses in Aspen Plus® V8.2; (ii) assessing a further reduction of the energy requirements by replacing steam by nitrogen in an alternative oil cake desolventizing process; and (iii) examining the environmental benefits of the process improvements.

3. Energy saving simulations and methods

3.1. Process description and operating parameters

The previous studies by Kong et al. for the RBO case study show that simulations using Aspen Plus® V8.2 provides results which approach the real process (Kong et al., 2018, 2017a; 2015a). The same simulations will therefore be used to obtain the steam requirements for the extraction of soybean oil (SBO), operated in a similar process as for RBO. In the RBO-study, the capacity was 1330 kg/h RBO. The entire process is scaled to the desired capacity of SBO (10,000 kg/h), keeping the same temperatures and pressures of the streams/bloc elements. The set-up of the three oil-hexane columns can be seen in Fig. 2, with essential parameter data given in Table 2.

The FEED consists of the oil that needs to be recovered (OIL) and the associated amount of hexane (HEXANE). The composition of OIL is given in Table 1. The mass flow of this stream is 10,000 kg/h. To achieve a complete oil leaching, the maximum mass flow rate of hexane should be 41,308 kg/h. Both streams have a temperature of 50 °C, a pressure of 1 atm, and are mixed at these conditions. Saturated steam is available at 10 bar and 179.9 °C.

To simulate the evaporators, the block Flash 2 is chosen. The choice to keep the evaporators at a constant temperature ensures that not all of the used steam condenses and dilutes the product stream. Keeping the second evaporator at 120 °C also ensures that the vapor stream from this column is usable for heat recovery. To keep these columns at their desired temperature, they require an input of energy. These heat duties are mostly provided by steam in an external jacket. No loss of energy to the environment is assumed.

The stripping column can be configured as a RadFrac type. It has only two stages and neither reboiler nor condenser. At the first stage, LIQ2 enters and VAP3 leaves; at the second stage STEAM3 enters and LIQ3 leaves. No pressure drop is assumed across the stages, thus
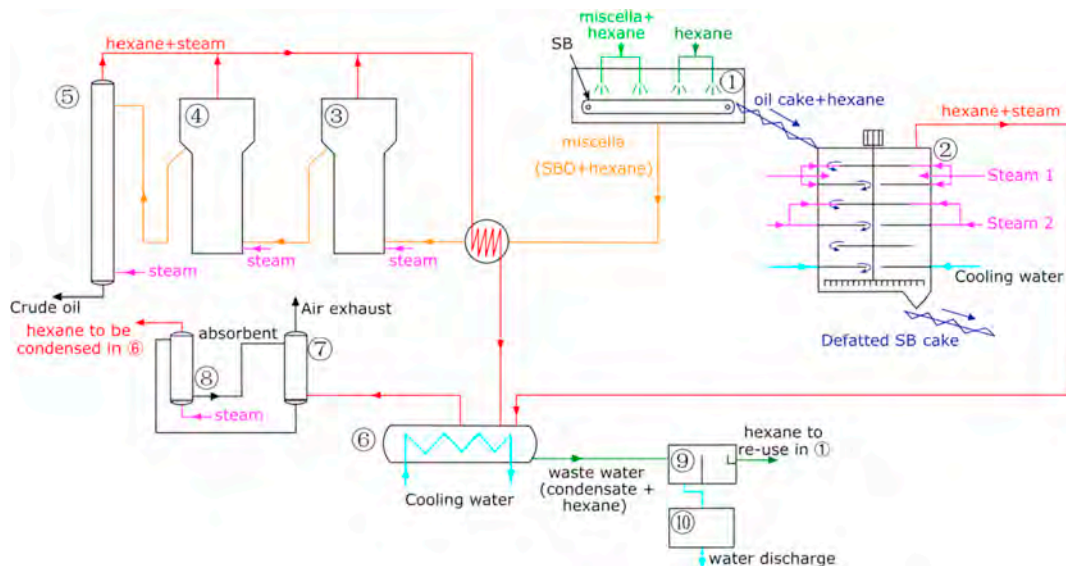
Fig. 1. Traditional SBO solvent-extraction: ① Leaching ② Desolventizer/toaster/cooler, ③ 1st evaporator, ④ 2nd evaporator, ⑤stripping tower, ⑥ cooler, ⑦ absorption ⑧, desorption, ⑨ water/hexane separator, ⑩ waste water treatment plant. Steam 1 is indirect or direct steam for hexane volatilisation, Steam 2 is indirect steam for toasting.
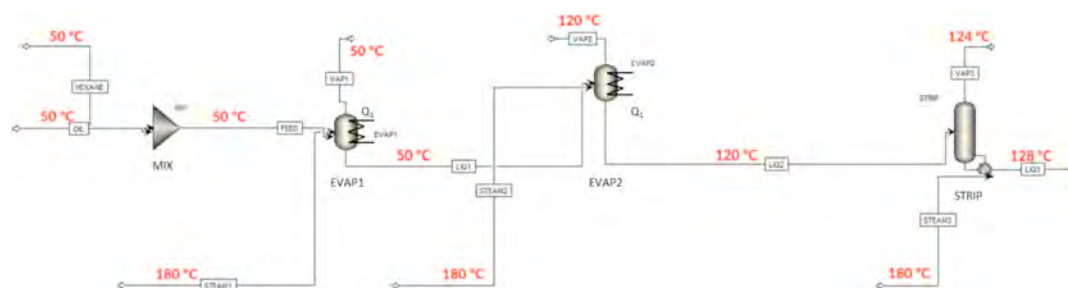
Fig. 2. Simulation set-up in Aspen Plus® 8.2.

Table 2
Configuration of both evaporators (Kong et al., 2015b).

Characteristic	1st Evaporator (EVAP1)	2nd Evaporator (EVAP2)
Temperature (°C)	50	120
Pressure (bar)	0.5	0.5
Occurring Phases	Vapor-Liquid	Vapor-Liquid

Table 3
Hexane content guidelines for different streams.

Stream	RBO extraction (Kong et al., 2015b)	SBO extraction
FEED	4122–5494 kg/h	30,992–41,308 kg/h
LIQ1	587–913 kg/h	4413–6864 kg/h
LIQ2	< 70 kg/h	< 526 kg/h
LIQ3	< 4 kg/h	< 30 kg/h

maintaining a constant pressure of 0.5 bar and operated around 120 °C, which is the temperature of the liquid input. If heat exchangers are used in further flow sheets, they are set-up with countercurrent flow and with a minimum temperature difference of 10 °C between the hot outlet flow and cold inlet flow. Non-Random Two-Liquids - Redlich-Kwong (NRTL-RK) is used in Aspen Plus® as the main method throughout the simulations. For the water phase, the Steam Tables are consulted by the simulation program.

There are three directly injected steam flow rates that need to be calculated with Aspen Plus®, one for each column and indicated as STEAM1, STEAM2 and STEAM3 in Fig. 2. The indirect heating of EVAP1 and EVAP2, indicated as Q1 and Q2 respectively, complete the required heat duty. Simultaneously varying the mass flow rates of the three live steam additions is not an efficient strategy and will significantly complicate the calculation. Therefore, the calculations are split into two parts: the first part considers both evaporators while afterwards the stripping column is taken into account.

Because the goal is to design a process that has minimum energy requirements, the heat duties of the evaporators have to be taken into account on the basis of the required energy to produce steam. Steam at 10 bar has a specific enthalpy of 2778 kJ/kg (Spirax Sarco, 2018). and is produced in a boiler, starting from condensate and make-up water at a pressure of 10 bar and a temperature of 25 °C. This water has a specific enthalpy of 105 kJ/kg. The net steam duty is hence 2673 kJ/kg. If the efficiency of the boiler is taken at 85–90%, the gross energy required to produce 1 kg of steam can be calculated:

$$\Delta h_{boiler} = (h_{steam} - h_{water})/\eta_{boiler} = 2673/0.85 \ or \ 0.9 = 3145 \ or \ 2970 \ kJ/kg \quad (1)$$

The required energy to operate the two evaporators consists of the indirect heat duties ($Q1$ [MW] and $Q2$ [MW]) and the live steam injection (m_{STEAM1} [kg/h] and m_{STEAM2} [kg/h]):

$$E_{evap} = Q_1 + Q_2 + \Delta h_{boiler} \times \frac{m_{STEAM1} + m_{STEAM2}}{3600 \ s/h} \quad (2)$$

Equation (2) has to be minimized, while also meeting the hexane content guideline from Table 3. The flow rates of STEAM1 (m_{STEAM1}) and STEAM2 (m_{STEAM2}) are varied both from 0 to 500 kg/h at an interval of 50 kg/h (11 different values), thus creating 121 (11^2) different possible combinations.

After the optimum combination of m_{STEAM1} and m_{STEAM2} is selected using Equation (2), the flow rate of the steam going into the stripper (m_{STEAM3}) is determined as injected live steam. The flow rate is varied from 100 to 500 kg/h, for intervals of 10 kg/h. The criterion to select the optimal m_{STEAM3} is that the liquid outflow of the stripping column must contain less

than 30 kg/h of hexane, while minimizing the amount of steam used. This stripper has no indirect heat duty to be taken into account.

3.2. Results and discussion

3.2.1. Energy consumption of the traditional process

The FEED contains 41,308 kg/h hexane. After the two evaporators, this has to be reduced to approximately 526 kg/h. The sensitivity analysis consists of 121 separate simulations, each with a different configuration of m_{STEAM1} and m_{STEAM2}. Fig. 3 displays the configuration results. Configuration N°89 minimizes the energy requirement for the system of two evaporators. The first evaporator uses 400 kg/h of steam (m_{STEAM1}) and requires 2.30 MW of additional external heating ($Q1$). The second evaporator uses no direct steam and requires 2.71 MW of additional heating ($Q2$). The liquid that exits the second evaporator to the stripping contains 434.3 kg/h of hexane, which is well below the guideline of Table 3. Using Equation (2), a net total energy requirement of 5.30 MW for the two evaporators is calculated. Using this optimal configuration for EVAP1 and EVAP2, a sensitivity analysis for m_{STEAM3} is performed. The liquid output of the stripping column has to contain less than 30 kg/h of hexane as a strict requirement. The results for the several values of m_{STEAM3} are represented in Fig. 4.

The minimum amount of steam that is needed to meet the stripper requirement is 220 kg/h. This results in a liquid outflow that contains 28.2 kg/h of hexane, 133.3 kg/h of water and 9996 kg/h of SBO. This stream has a temperature of 128 °C, which could be used for heating a colder stream.

3.2.2. Heat exchange improvements of the traditional process

Within the traditional process, there are different options to improve the energy efficiency by adding waste heat exchangers into the circuits. These different options are illustrated in Fig. 5, and involve either adding one heat exchanger (HEX1), or two heat exchangers (HEX2) in two options A and B.

3.2.2.1. One additional heat exchanger.
The vapor streams that exit the second evaporator and the stripping column are both at high temperature (120 °C and 124 °C respectively). These two are combined and sent to a heat exchanger that will heat the FEED stream. Fig. 5 displays the new flow sheet. A similar sensitivity analysis as presented in section 3.2.1 can be used to optimize the heat recovery efficiency. The energy balance and residual hexane concentration around the stripping column remain unchanged and were presented in Fig. 4 before. Results of the simulation are given in
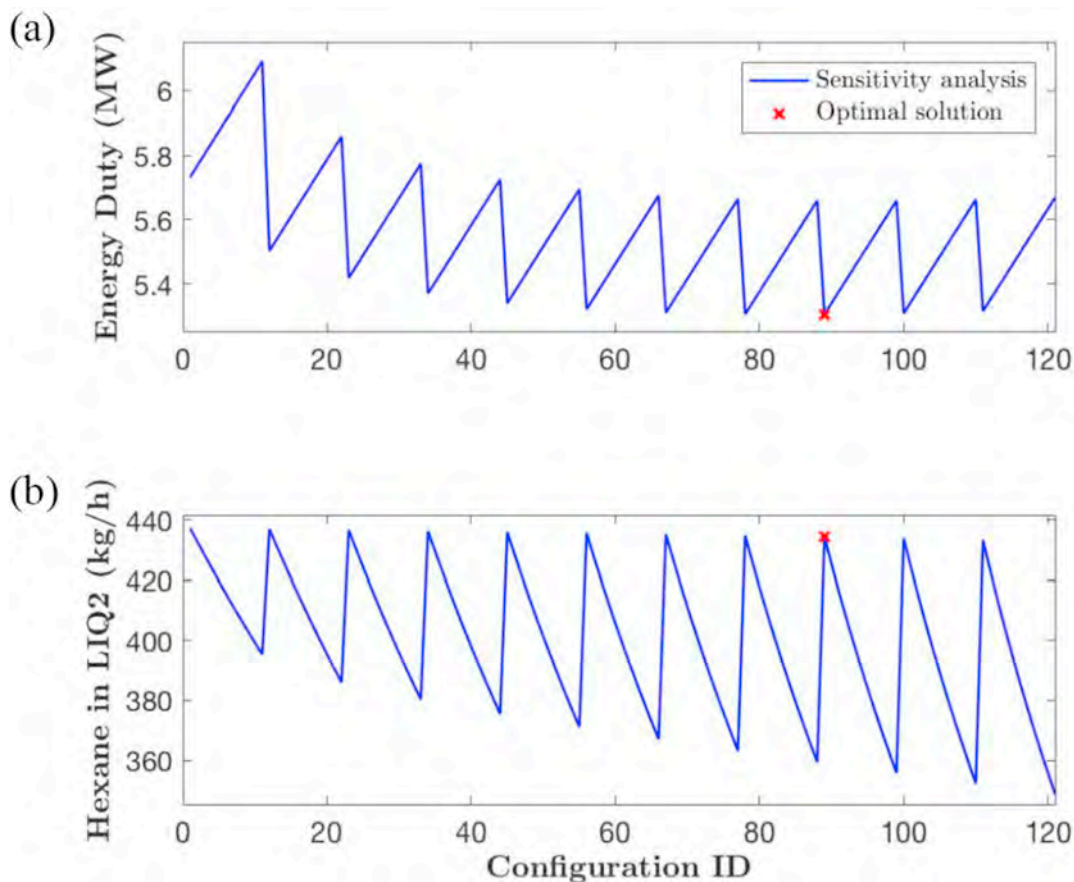
Fig. 3. (a) Energy required for different configurations; (b) Hexane remaining in the liquid outflow of EVAP2 for different configurations.

Appendix (Fig. A1). The first evaporator utilizes 50 kg/h of live steam (m_{STEAM} 1) and consumes 0.25 MW ($Q1$). The second evaporator only requires 4.26 MW ($Q2$) and no live steam. After the two evaporators, the liquid outflow still contains 436.87 kg/h of hexane. To reduce this hexane content below 30 kg/h, the stripper needs at least 220 kg/h of steam, as shown in Fig. 4. This results in a liquid outflow that contains 27.5 kg/h of hexane, 133.8 kg/h of water and 9993 kg/h of SBO.

3.2.2.2. Two additional heat exchangers (option A). An additional stream is selected to further preheat the feed. The liquid exit flow of the stripper has a temperature of 128 °C and a mass flow rate of 10,158 kg/

h. This stream, which contains the desired oil product can be used for preheating other streams. Two different streams can be preheated: HOTFEED and LIQ1. This will decrease the heat duty of the evaporators by implementing a second heat exchanger to preheat the feed. This is represented in Fig. 5. The sensitivity analysis (Appendix, Fig. A2) demonstrates that the first evaporator has a heat duty of −0.045 MW, meaning that there is a limited amount of additional heat available for a recovery. In practice, this heat can be used to preheat the boiler feed water (of 25 °C). The first evaporator uses 50 kg/h of steam (m_{STEAM1}). The second evaporator requires 4.26 MW ($Q2$) and no live steam. The liquid that exits the second evaporator to the stripping still contains
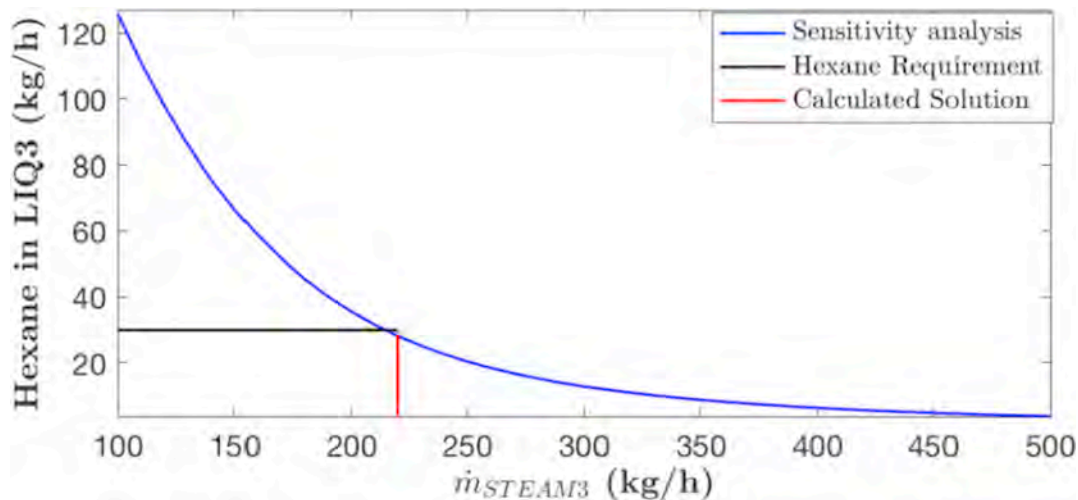

Fig. 4. Hexane remaining in the liquid outflow of the stripper for different values of m_{STEAM3}.
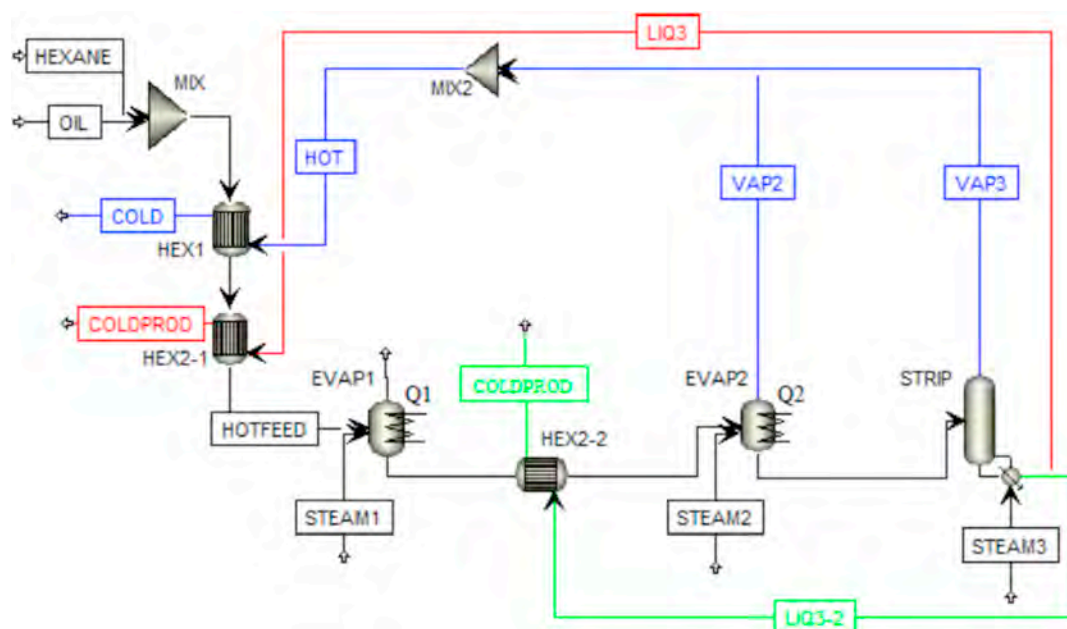
Fig. 5. Heat recovery improvements of the conventional extraction process. Red: one heat exchanger; Blue: two heat exchangers (opt A); Green: two heat exchangers (opt B). (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)

436.9 kg/h of hexane, again below the guideline. This results in a total indirect energy requirement of 4.26 MW for the two evaporators.

The stripper column employs 220 kg/h of steam to reduce this hexane content to below 30 kg/h, according to the sensitivity analysis. This results in a liquid that contains 28.4 kg/h of hexane, 133.1 kg/h of water and 9993 kg/h of SBO. This stream has a temperature of 128 °C.

COLDPROD is still at a relatively high temperature (81 °C) after leaving the heat exchanger. HOTFEED is only 0.4 °C hotter than in the previous set-up, but its vapor fraction has increased from 0% to 12%. The exchanged heat in the second heat exchanger is thus used for the partial evaporation of the feed.

3.2.2.3. Two additional heat exchangers (option B). LIQ3 will be used to preheat LIQ1 before it enters the second evaporator, which is displayed in Fig. 5. Results are illustrated in Appendix, Fig. A3. Similar to the previous set-ups, the first evaporator now uses 50 kg/h of steam (m_{STEAM1}) and has an indirect heat duty of 0.25 MW (in contrast to the negative $Q1$ in option A). The second evaporator solely requires 3.84 MW ($Q2$) (without live steam addition). A total energy requirement of 4.13 MW is calculated for the first two evaporators. This is less than the energy requirement for the previous set-up. At 220 kg/h of steam feed, the stripper column reduces the hexane content from 436.9 kg/h to below 30 kg/h. LIQ3 contains 28.4 kg/h hexane, 133.1 kg/h *and* 9993 kg/h oil.

LIQ3 is cooled from 128 °C to 60 °C, too cold for further heat recovery and COLDPROD thus serves no further use. LIQ1 is heated only from 50 °C to 51 °C, since most of the exchanged energy is used as latent heat to evaporate the hexane in LIQ1, this explains the small increase in temperature. The vapor fraction of LIQ1 increases from 0% to 13%.

Although the stream that exits HEX2 can still be further used to preheat LIQ1, the heat recovery potential is limited, and the implementation of an extra heat exchanger is neither thermally nor economically justified.

3.2.2.4. Final selection. The different set-ups that were discussed in the previous sections all have different energy consumptions. Table 4 displays these different values as well as their saved energy relative to the basic set-up. E_{evap} is calculated using Equation (3). Option 2 (B) has the highest energy saving potential and 21.2% of the energy can be

Table 4
Summary of the energy consumption for the five different set-ups.

Heat exchangers	0	1	2 (option A)	2 (option B)
$Q1$ (*MW*)	2.30	0.25	−0.045	0.25
$Q2$ (*MW*)	2.71	4.26	4.26	3.84
m_{STEAM1} (*kg/h*)	400	50	50	50
m_{STEAM2} (*kg/h*)	0	0	0	0
m_{STEAM3} (*kg/h*)	220	220	220	220
E_{evap} (*MW*)	5.47	4.71	4.42	4.29
Energy Saved	0%	13.9%	19.2%	21.2%

saved.

$$E_{evap} = Q_1 + Q_2 + \Delta h_{boiler} \times \frac{m_{STEAM1} + m_{STEAM2} + m_{STEAM3}}{3600 \ s/h} \qquad (3)$$

It should however be remembered that values indicated are net heat duties. Considering the efficiency of the boiler (85–90%), the gross heat supply to the boiler should be multiplied by 1/0.85 or 1/0.9 respectively, increasing the amount of fuel to be burned in the boiler.

4. Improvement of the desolventizer operation

4.1. Desolventizer in the traditional process

To obtain the separation between the oil cake and hexane, the aforementioned desolventizer-toaster (DT) unit is applied. The DT is conventionally of multiple-heath concept. To keep the hexane cake mixture moving downwards from tray to tray, there is a rotating central shaft with arms horizontally connected to it. Rotation of the central shaft causes the arms to sweep over the horizontal trays and thus keep the materials moving and stir the mixture. The oil cake-solvent mixture from the extraction process enters the desolventizer-toaster at the top. Steam is used to increase the wet meal temperature and evaporate the hexane and moisture from the oil cake. The steam can be introduced directly and indirectly to the mixture via the trays. One can distinguish between indirect tray steam and live sparge steam heating, the latter using steam directly introduced to the meal via holes in the desolventizing tray or the sweeping arms.

Various articles have investigated the desolventizer and all its

aspects, as one of the problems that industry encounters is the high energy requirements. This high energy requirement originates from the production of steam that is used for hexane removal from the meal. Alternative toasting techniques to lower the energy usage in a desolventizer include the mitigation of additional heat that is required for drying of the flakes (Ong, 1984). Using the energy requirement of the pilot 1330.00 kg/h RBO process, an estimation can be made for the 10,000 kg/h SBO process (Kong et al., 2018). On average, the amount of steam needed to desolventize and toast the cake is 0.195 MW/kg cake. Depending on the operating steam conditions (pressure, temperature) this results in 265 kg steam/ton cake for 6 bar saturated steam, and 263 kg steam/ton cake for 10 bar saturated steam.

Transforming this to the soybean case, with 35,500 kg/h of cake produced, and using 10 bar saturated steam, the total amount required is 9.34 ton steam/h or 7.7 MW at a boiler efficiency of 90%.

4.2. Fluidized bed desolventizer

The most promising alternative replaces the live and indirect steam heating of the flakes in the desolventizer by a nitrogen driven fluidized bed. Fluidization is an interesting technique that is already applied in conventional food processing for drying of various solids. Its popularity originates from the high mass and heat transfer achieved in these beds by continuously pumping a gas through different gaps in the distributor plate. This initiates the agitation of the particles present in the bed. As soon as the gas velocity is high enough to carry the weight of the particles (= minimum fluidization velocity), fluidization sets in, which enhances the mixing of solids and results in a uniform temperature throughout the bed.

Experiments have shown that compared to the conventional techniques, a high hexane removal can be achieved with a low residence time (τ) of the flakes by making use of a fluidized bed desolventizing unit. The usage of nitrogen gas also allows the bed to be operated at a lower temperature compared to the conventional technique. Nitrogen lowers the partial pressure of the hexane which increases the volatility

of the compound as can be seen in Fig. 6 (Kong et al., 2018). There is however the need for on-site production of nitrogen (95–99%) at a medium production capacity. This can be done by installing either a pressure swing adsorption system or a membrane unit. Membrane systems can be useful due to their robustness but often cannot deliver the nitrogen gas capacity needed to sustain the fluidized bed operation. Since most of the N_2 is recovered in the desolventizer membrane separation module, only minor flow rates of make-up N_2 are required. Hence O_2/N_2 membrane separation is the evident choice.

When applying the fluidized bed as a desolventizer, a constant feed stream is transported from the hexane extractor to the fluidized bed (Erickson, 1995). Here, heat is provided to the flakes in two ways. First of all heat is transferred in a forced convective way by the high temperature nitrogen gas flowing around the particles. Secondly conductive transport of heat occurs by contacting the particles against the surrounding heated walls. When operating at a bed temperature of 50 °C, the hexane content of the meal is lowered beneath the threshold value of 0.1%. This value is reached in about 150 s (Kong et al., 2018) when the fluidized bed is operated at superficial gas velocities in excess of 0.04 m/s. There are two exit streams in the fluidization process: the gas exiting the top is a mixture of hexane and nitrogen (which can still be used for the preheating of other streams), and the meal itself, of which the residual heat will be regarded as a heat loss due to the low efficiency and practical difficulty of energy recovery from the low temperature meal. The layout of the fluidized bed desolventizer is illustrated in Fig. 7.

The cross-flow fluidized bed will consist of 3 sections, being the desolventizing section operated with N_2 as fluidizing gas, and the toaster/cooler sections operated with air. Baghouse filters are installed for dust abatement. The possibility to use O_2-lean carrier gas, as a mixture of air and N_2 is further discussed after having established the volumetric ratio of hexane and carrier gas, and considering the explosion hazards of hexane/air mixtures.

As soybeans consist for 71% of solids, extraction of the oil leaves a solid residue. For the extraction of 10,000 kg/h SBO, a residual solid
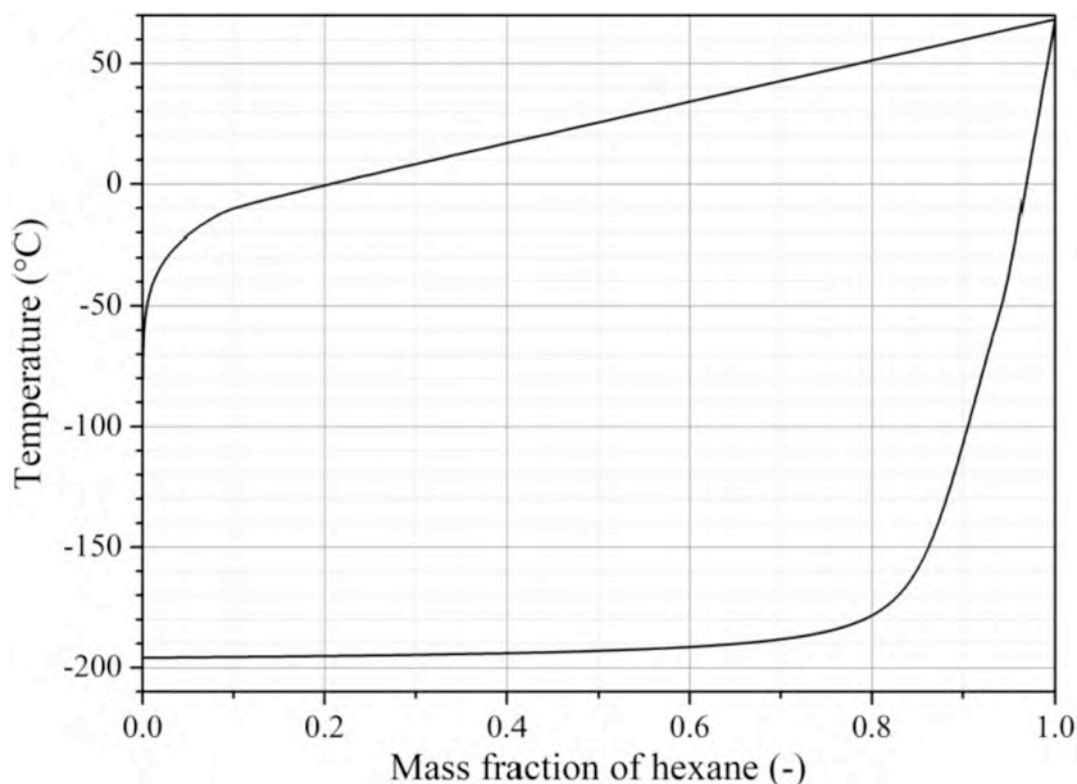

Fig. 6. Vapor-liquid equilibrium diagram of the nitrogen-hexane mixture at 1 bar.
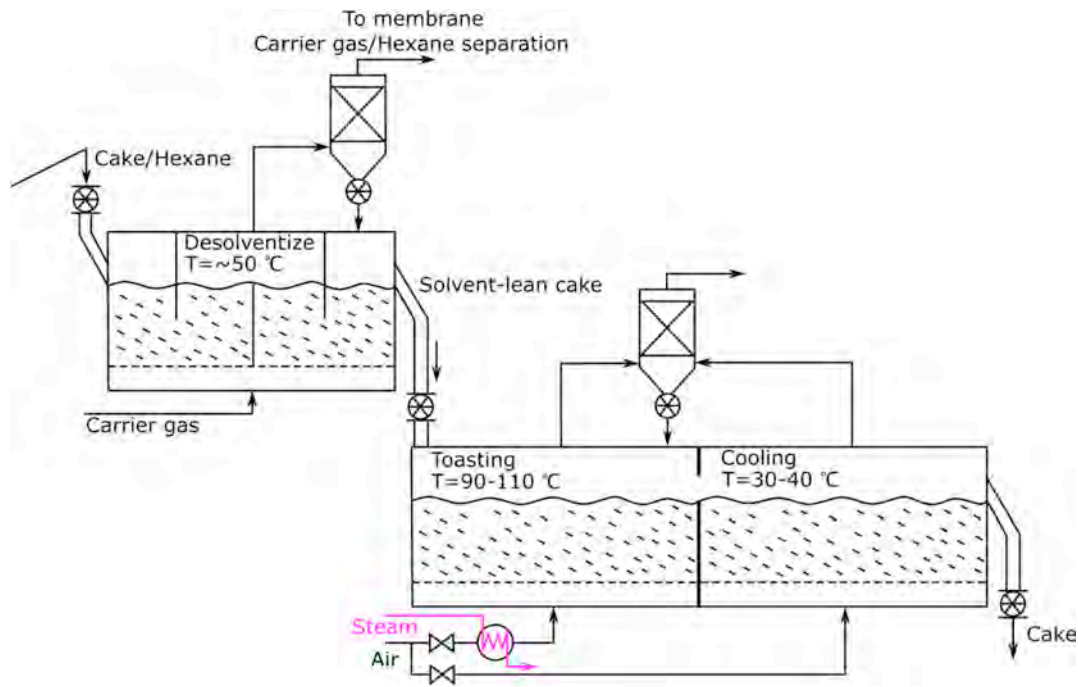
Fig. 7. Design of fluidized bed desolventizer/toaster/cooler.

stream of 35,500 kg/h is produced. It contains 35 wt% of hexane after the extraction process. The total mass of the feed can be calculated by the simple mass balance.

$$0.35 \times m_{tot} + m_{solid} = m_{tot} \tag{4}$$

The total mass of the feed is therefore 54,615 kg/h containing hexane, m_{hex}, equal to 19115 kg/h. m_{hex} is used to calculate the necessary heat requirements for the fluidized bed. Before the total heat requirement of the bed can be calculated, the mass flow rate of nitrogen should be determined. Knowing that using nitrogen gas alters the partial pressure of hexane, three separate mass percentages can be found in Fig. 6 for temperatures of 30, 40 and 50 °C respectively, all values chosen between the dew point at 50 °C and the experimental result (Kong et al., 2018). The necessary mass flow rate of nitrogen gas is determined using the equation below:

$$m_{N2} = \frac{m_{hex} \times wt\%_{N2}}{(1 - wt\%_{N2}) \times 3600} \tag{5}$$

Then the heat required in desolventizer/toaster is established.

heat required = heat for evaporatation of heaxane + heating of solids

+ heat loss to the environment

= heat provided by the internal or external heating system

+ heat from N_2

Neglecting heat losses, and with a heat of hexane evaporation of 350 kJ/kg, the heat required is determined for any selected operating temperature, e.g. between 30 and 50 °C. Results are given in Table 5.

To achieve this operation, a minimal flow of 1.42 m³/s (1.50 kg/s) of nitrogen is used at 50 °C to extract 1.85 m³/s (5.31 kg/s) of hexane from the bed.

Table 5
Heat requirements of the bed for the different weight percentages of N_2.

T (°C)	Wt % of N_2	m_{N2} (kg/s)	v_{N2} (m³/s)	Q (kW)
30	42.6	3.94	3.49	1873
40	32.5	2.56	2.34	1831
50	22.0	1.50	1.42	1799

Operating at 22 wt% N_2 reduce the production necessity of N_2 and lowers the membrane area in the subsequent hexane separation from the exhaust stream of the desolventizer. In addition, it has lower hexane losses. All leads to a reduction in capital cost. Further reduction in the process heat balance is achieved by preheating the nitrogen stream. The membrane separation module is discussed below.

The lower (LEL) and Upper (UEL) explosion limits in air are 1.2 and 7.4 vol%, respectively. Since the exhaust of the desolventizer contains over 50% vol% hexane, the use of an O_2-lean mixture of air and N_2 could be imagined, thus further limiting the size of the N_2 membrane production unit. This is however not recommended since a hexane depletion will occur in the membrane separation unit (see 4.3) and its concentration will progressively decrease and achieve values between its UEL and LEL. Since only make-up N_2 is required, safe operation imposes its use throughout the desolventizing process.

4.3. Gas membrane separation

For separation of the hexane-N_2 mixture, a gas membrane system is applied. The membrane can guarantee high recovery of hexane for reuse in the extraction process as well as for N_2 which can be reused in the fluidization process. A decrease in temperature of the gas entering the membrane system, gives a decreased efficiency of separation and thus higher losses of hexane in the process. At higher temperatures membrane swelling needs to be taken into account. Laboratory results determined the optimal performance of the membrane at 39 °C with a recovery of 96.8% of hexane. A permeability of 2690 NL/m²·h·*bar* is reached at this temperature (Degrève et al., 2001; Kong et al., 2018, 2017b).

$$A = \frac{Q_{perm, hex}}{L_{hex} \times (P_{feed} \times y_{mean,hex} - P_{perm} \times \frac{Q_{perm,hex}}{Q_{perm,tot}})} \tag{6}$$

with

$$y_{mean,hex} = [\frac{Q_{feed, hex} - Q_{perm, hex}}{Q_{feed, tot} - Q_{perm,tot}} + \frac{Q_{feed, hex}}{Q_{feed, tot}}]/2 \tag{7}$$

Equation (6) is used to determine the total area of membranes necessary to separate the N_2-hexane gas mixture. The selectivity,
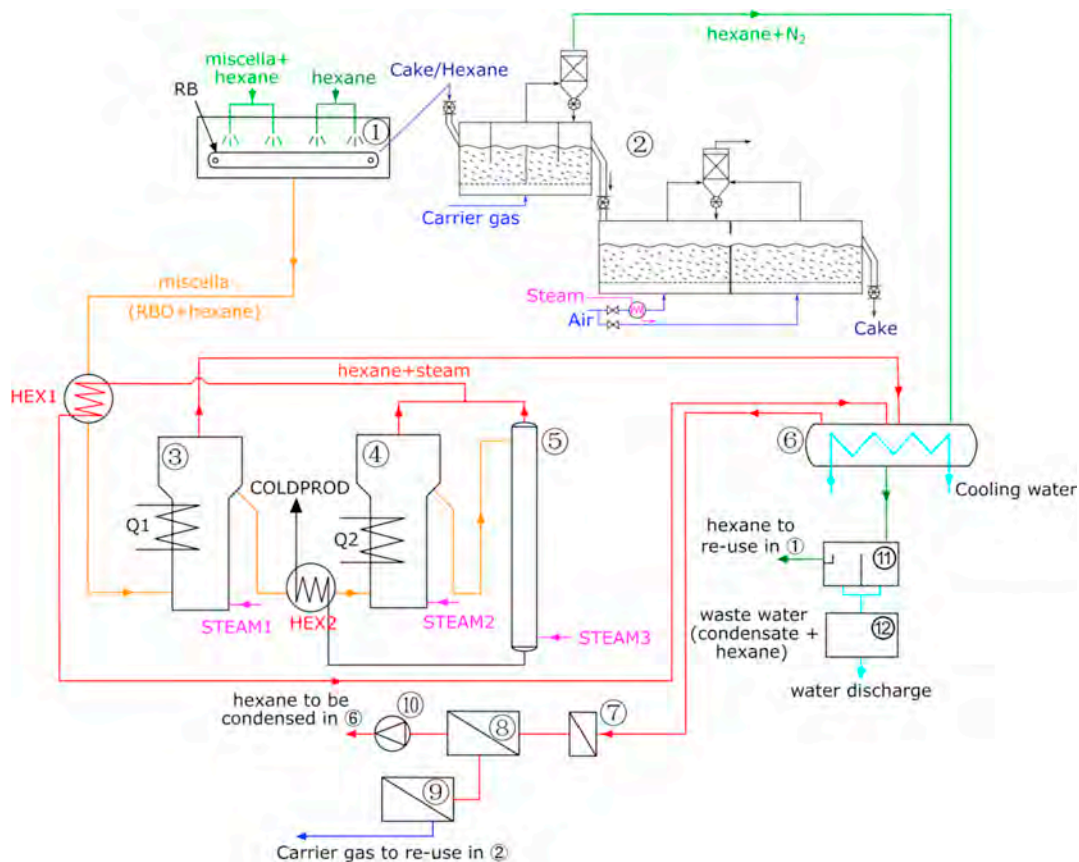
Fig. 8. Detailed flow sheet of the entire production process of SBO: 1. Leaching; 2. Fluidized bed desolventizer/toaster/cooler; 3. EVAP1; 4. EVAP2; 5. STRIP; 6. Condenser; 7. Filter; 8. Cascade gas membrane separator; 9. Cascade gas membrane separator; 10. Vacuum pump; 11. water/hexane separator; 12. Waste water treatment.

separation factor and the intrinsic hexane permeability of the membrane can be found in Kong et al. (2018) The amount of N_2 and hexane on the permeate/retentate side can be calculated.

$$Q_{feed,hex} = \frac{m_{feed,hex}}{\rho_{hex}} = 716 \, Nm^3/h \tag{8}$$

$$Q_{feed,tot} = \frac{m_{feed,hex}}{\rho_{hex}} + \frac{m_{feed,N2}}{\rho_{N2}} = 1966 \, Nm^3/h \tag{9}$$

$$Q_{perm,hex} = \frac{m_{perm,hex} \times Wt_{perm,hex}}{\rho_{hex}} = 693 \, Nm^3/h \tag{10}$$

Before the total volumetric flowrate at the permeate side can be determined, the amount of N_2 at the permeate side has to be calculated.

$$n_{perm,N2} = \frac{n_{perm,hex} \times y_{feed,N2}}{y_{feed,hex} \times \beta}$$

For the total calculated volumetric flowrate of hexane on the permeate and feed side, the membrane area is $1412 \, m^2$. A permeate pressure of 1000 Pa was chosen because it gives the highest membrane area necessary for separation (worst case scenario). The permeate stream is consecutively send to a cooler for further condensation. Condensed hexane can then be reused in the leaching process. The high purity N_2 retentate is reused in the fluidized bed desolventizer. By subtracting the recovered amount of N_2 by the total amount necessary of the fluidized bed, the total production quantity of N_2 necessary for operation can be determined. This helps in choosing the most adequate production technique for the make-up N_2.

It should moreover be remembered that a higher degree of separation is obtained in a membrane cascade. In gas permeation application, where the permeate stream of module 1 has to be compressed before

feeding module 2, the cascade is often limited to 2 stages. The higher separation efficiency is certainly favored in the hexane/N_2 case, facilitating the recovery and re-use of both separate streams, although gas processing costs must be optimized (Ahsan and Hussain, 2016; Aliaga-Vicente et al., 2017; Pathare and Agrawal, 2010). The total flowsheet of Fig. 8 includes the cascade membrane module.

The flowsheet of Fig. 8 includes the proposed improvements, and should be compared with the conventional process of Fig. 1. The addition of the heat exchanger network, and the implementation of a N2-driven desolventizer with gas membrane separation and reuse of hexane contribute to considerable energy and environmental benefits, as explained before.

5. Energy and environtal impacts of the improved process

The previous sections calculated the net amounts of live steam injection and closed circuit (indirect) steam uses. To convert them to operational values including the boiler efficiency (85%), gross energy inputs are obtained.

The comparative energy picture is illustrated in Table 6.

The electricity consumption in the crude oil extraction processes is on average 125 kW/ton of crude oil (Kong et al., 2018). It is not expected that major reductions are achieved since the reduction through a lower boiler capacity will be outbalanced by the required power for fluidization and gas membrane separation. A 10 ton/h SBO plant will hence consume about 1.25 MW of electric power.

Based upon these energy figures and their equivalent CO_2 emission potential, it is clear that the improved SBO process will have a lower
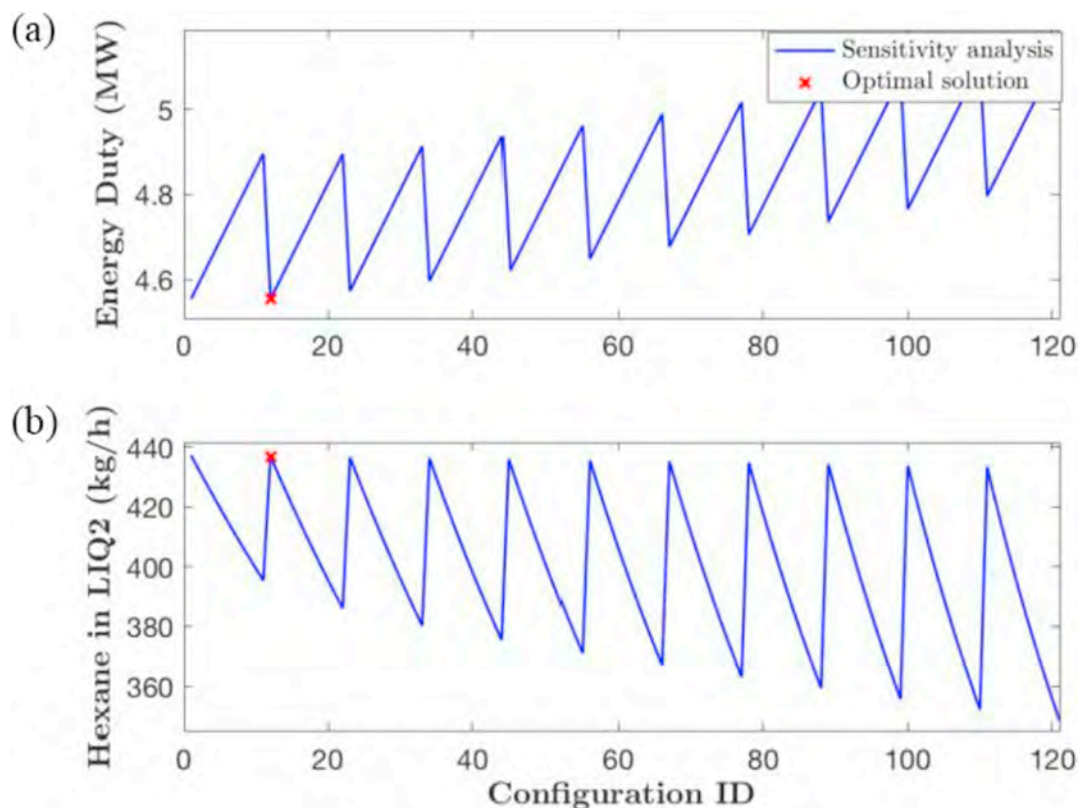
Table 6
Conventional energy consumption in SBO process (at 85% boiler efficiency).

	Conventional SBO (MW)	Improved SBO process
Leaching	0.17	0.17
Evaporators	6.44	5.05
Stripping	0.18	0.18
Desolventizer	8.15	2.12
Tail gas stripper	2.93	–
Total	17.87	7.52

CO_2 footprint than the conventional process. For the conventional process, CO_2 emissions are calculated at 506,000 ton/year against 189,570 ton/year for the improved process. A reduction of 62.5% of CO_2 emission is achieved.

The waste water discharge is mostly determined by the steam condensate. The steam condensate has an average COD-load of ~4.6 g/L, an average concentration of soap, oils and grease of 630 mg/L, and an aerobic activated sludge treatment is mostly used. Reverse osmosis is being introduced to reduce sulphate contents (Cheng et al., 2018). Due to the reduction of direct steam injection in the process from 620 to 270 kg/h, the water treatment plant will see a 56.5% reduction in hydraulic and pollutant loads.

6. Conclusions

The entire improved process was displayed in Fig. 8. By designing a new configuration for the evaporator-stripper part of the process, the live steam usage is reduced from 620 kg/h to 270 kg/h, and the total heat duty is reduced by 21.9%. The original desolventization process required an energy input of 8.15 MW. At 100% heat recovery, the application of the fluidized bed desolventization with preheating of nitrogen reduces this requirement to 2.12 MW which results in a saving of 74%. The improvements moreover reduce both the waste water loadings by 56.6% and the CO_2 emissions by 62.5%. Hexane emissions are moreover significantly reduced by over 90%.

Appendix. Sensitivity analysis and iteration results with heat recovery.

Fig. A1. Set-up with one heat exchanger (a) Energy required for different configurations (b). Hexane remaining in liquid outflow of 2nd evaporator for different configurations.
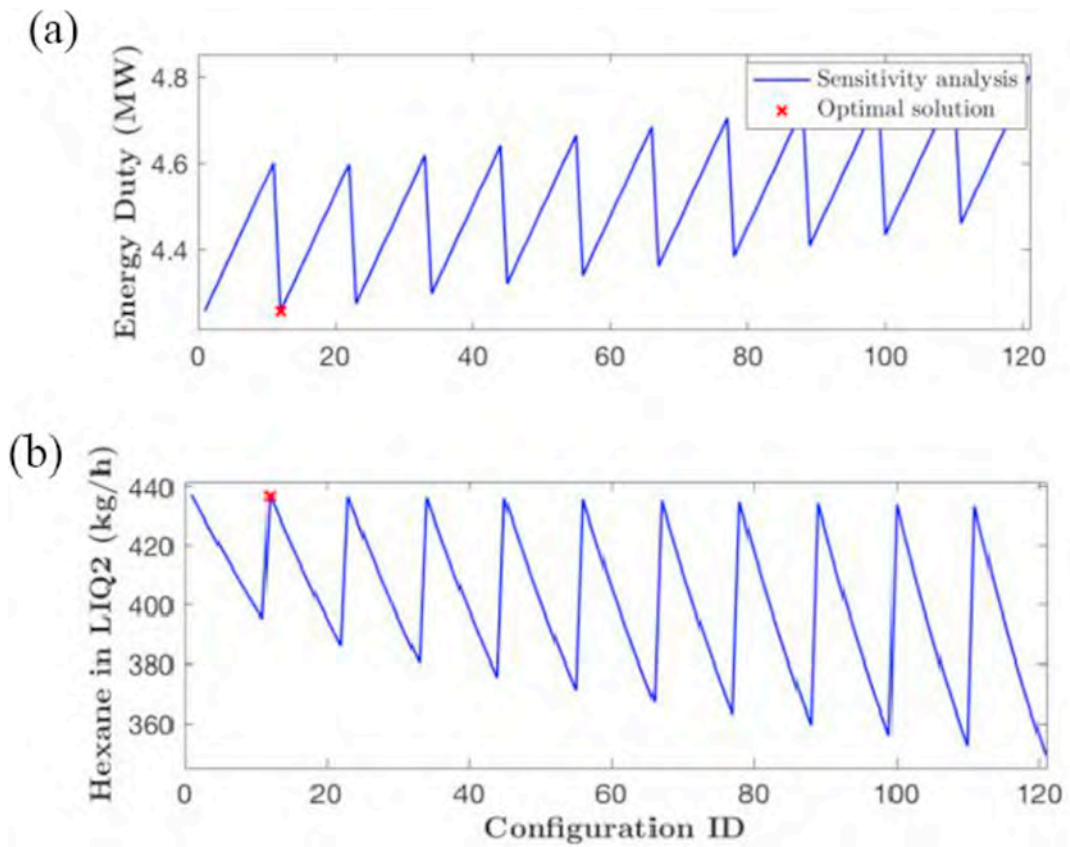
Fig. A2. Set-up with two heat exchangers, option A (a) Energy required for different configurations (b) Hexane remaining in liquid outflow of 2nd evaporator for different configurations.

Fig. A3. Set-up with two heat exchangers, option B (a) Energy required for different configurations (b) Hexane remaining in liquid outflow of 2nd evaporator for different configurations.
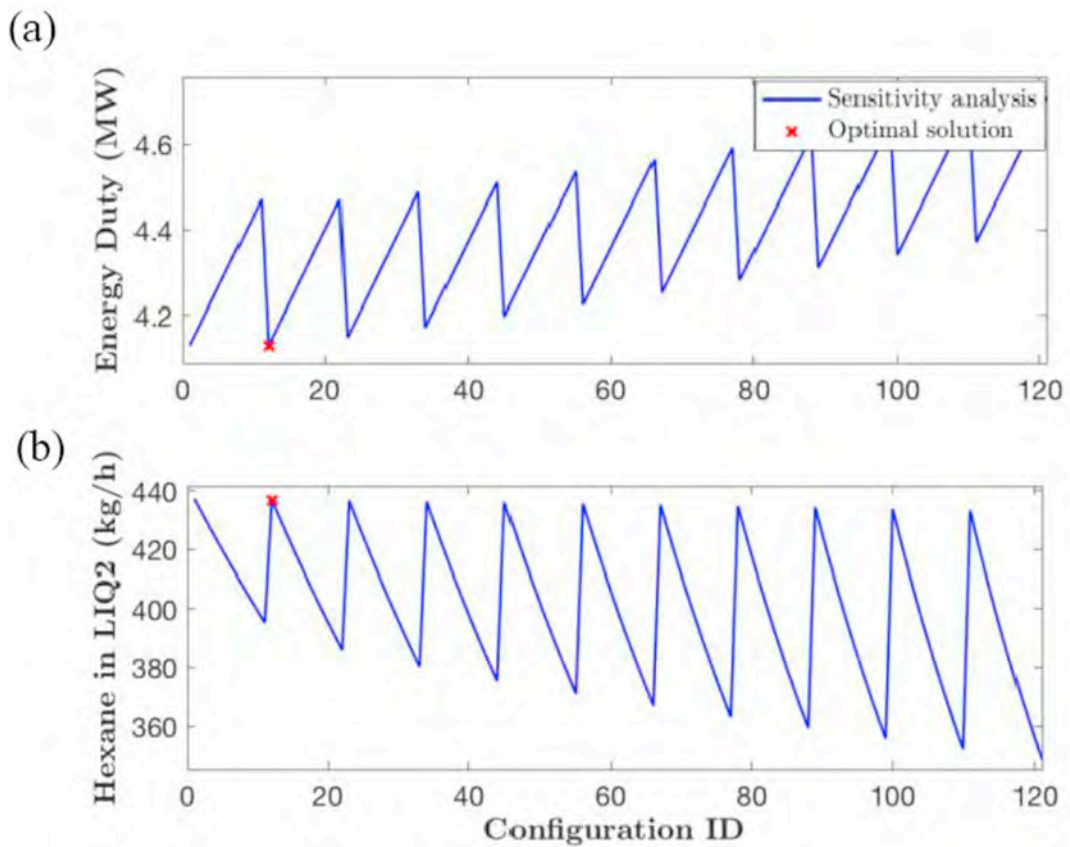
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