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RESEARCH ARTICLE

Environmental and economic assessment of vegetable oil production using membrane separation and vapor recompression

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Abstract Solvent extraction of crude oil from oilseeds is widely applied for its high production capacity and low cost. In this process, solvent recovery and tail gas treatment are usually performed by adsorption, paraffin scrubbing, or even cryogenics (at low tail gas flow rates). Membrane separation, which has a lower energy consumption than these techniques, spans a broad range of admissible concentrations and flow rates, and is moreover easily combined with other techniques. Vapor recompression has potentials to reduce the heat loss in association with distillation and evaporation. In this study, we proved the possibility of combining membrane separation and vapor recompression to improve the conventional vegetable oil production, by both experiments and process simulation. Nearly 73% of energy can be saved in the process of vegetable oil extraction by the novel processing approach. By further environmental assessment, several impact categories show that the optimized process is environmentally sustainable.

Keywords vegetable oil, solvent-extraction, membrane separation, vapor recompression, environmental and economic assessment

Supplementary Material

1. Process modeling [1]

The RBO mill under scrutiny, is located in the Heilongjiang Province of China, and has a production capacity of 32 T/day RBO. The steam consumption of the current process is ~ 2.4 T/h and ~ 393 m³/h of cooling water are used. The energy reduction is

a critical issue for the process, since representing nearly 25% of the total production cost.

The process simulations are mainly based on Aspen Plus v8.2, by defining components, researching appropriate thermodynamical models, and fixing equipment and operating conditions. (Yun et al., 2013) Since Aspen Plus v8.2 is incomplete mostly towards operations involving a solid phase, additional subroutines need to be prepared, compiled and imported into the simulation. In the present study, the pressure drop by pipe flow resistance was neglected. The details of the simulation package are described below.

1.1 Key components

Insoluble solids are not defined in Aspen Plus. In this process, defatted RB is stable and no physical changes occur, except for its temperature. We experimentally determined the specific heat capacity of defatted RB to enable the calculation of the energy consumption of defatted RB in the desolventizing process. At the same time, triglycerides are not subject to physical changes in the process, and their heat capacity is close to the C_p of defatted RB. An amount of triglyceride was hence chosen to replace defatted RB to calculate the heat exchange in the process (since the Aspen supporting files do not cater for defatted RB).The amount of triglyceride can be calculated as:

$$\int_{T_0}^{T_1} m(DRB)CP(DRB)dT = \int_{T_0}^{T_1} m(TAG)CP(TAG)dT \quad \text{(S1)}$$

where *m (DRB)* is the mass flow of defatted RB meal in the RBO plant, *CP (DRB)* is the specific heat of defatted RB meal, *m (TAG)* is the amount of triglyceride that replaces defatted RB meal in the simulation, *CP (TAG)* is the specific heat of triglyceride, T_0 and T_1 are the initial and final temperature respectively of the DT.

For crude RBO, triglyceride was the main component. In the present simulation, olein was used to represent triglyceride. Since the boiling points of triglycerides are much higher than those of the other components in the miscella, different triglycerides achieve nearly the same mass and heat transfer in the process. Free fatty acid (FFA), with low boiling point, was simulated by three kinds of FFA. Since the binary interaction parameters and thermodynamical properties required for the Aspen simulation of some components were not available (triglyceride and diglyceride), UNIFAC was used to predict them by their molecular structure. (Yun et al., 2013) The molecular structure of triglyceride and diglyceride were defined by their connectivity in Aspen Plus.

1.2 Specific heat of defatted RB meal

The specific heat (CP) of defatted RB meal is important in the energy balance calculation of the desolventizer-toaster. Since no data were available, experiments were carried out by differential scanning calorimetry (DSC 204 F1, NETZSCH

Corporation, Germany) to measure the CP of defatted RB between 27 °C and 109 °C. The relation between the CP of defatted RB and temperature was studied by polynomial fit. The RB meal sample was the pretreated rice bran from the plant in Heilongjiang, and it was defatted by soxhlet extraction. Three different defatted RB samples with different particle sizes were tested by DSC. Results are given in Table S1 and fitting data are represented in Figure S1. The polynomial fitting of CP and temperature was as follows.

$$CP = -1.465 + 0.221T - 6.31 \times 10^{-3} T^2 + 8.67 \times 10^{-5} T^3 - 3.91 \times 10^{-7} T^4 \quad \text{(S2)}$$

where *CP* is the specific heat of defatted RB, *T* is the temperature (°C).

In this fit, R^2 is 0.99964 and the F-value is 57589.36, which means that the fit is accurate.

Table S1

Specific heat (CP) of defatted RB versus T, from DSC measurements

Temperature (℃)	CP [kJ/(kg·°C)]
27	1.269 ± 0.0321
30	1.534 ± 0.0117
35	1.659±0.0166
40	1.800 ± 0.0348
45	1.969 ± 0.0576
50	2.175 ± 0.0848
55	2.420 ± 0.1157
60	2.701 ± 0.1503
65	3.020 ± 0.1865
70	3.376 ± 0.2211
75	3.763 ± 0.2510
80	4.161 ±0.2665
85	4.526 ± 0.2616
90	4.820 ± 0.2342
95	4.996 ± 0.1775
100	5.017 ± 0.1277
105	4.901 ± 0.0789
109	4.728 ± 0.0699

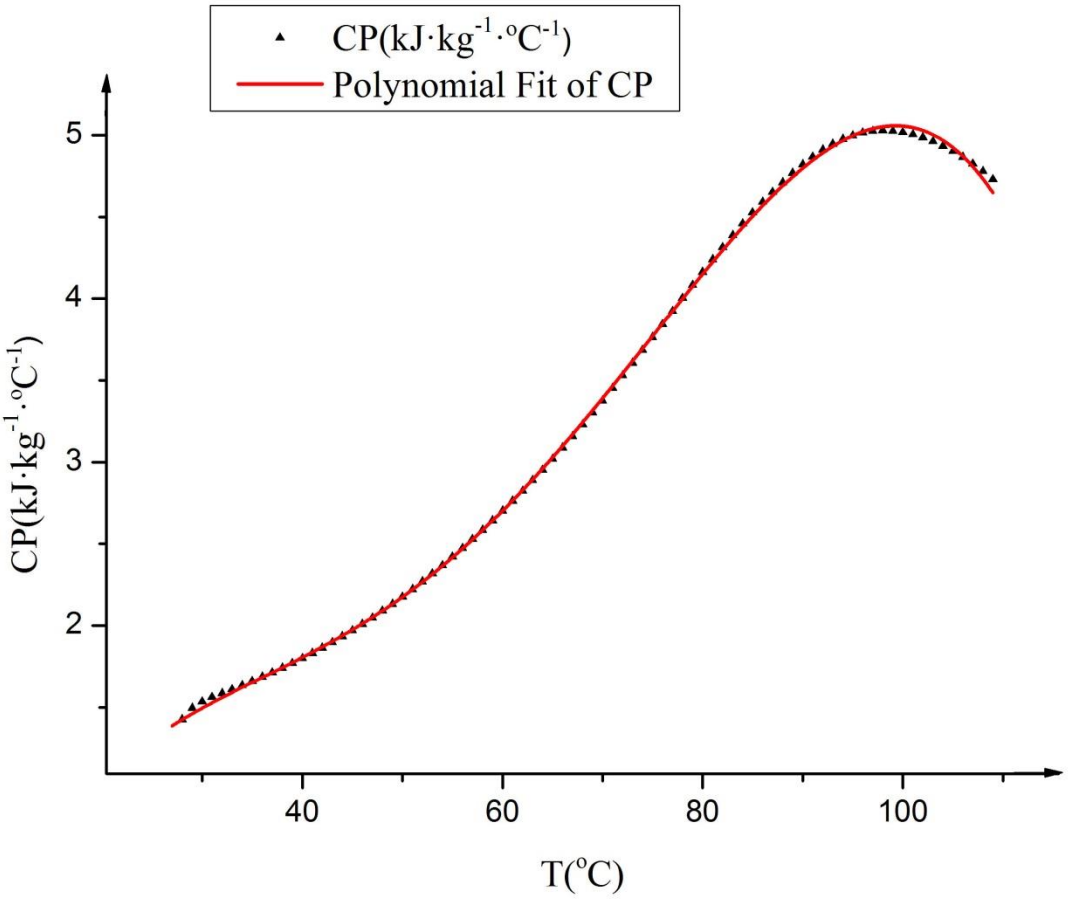

Fig.S1. Polynomial fit curve of the CP of the defatted rice bran.

1.3 Thermodynamic models

In this system, components of RBO were generally polar compounds, expect for n-hexane, a non-polar compound. Since the process was completely run at a low pressure (50kPa~100kPa), both a Non-random-two liquids (NRTL) model (for polar system) and the Peng-Robinson (PENG-ROB) model (for non-polar system) were tested to simulate the process. Finally, the NRTL model proved to be better for this process because of the better convergence and more exact results.

1.4 The leaching column

Aspen Plus has the flexibility to allow the insertion of FORTRAN blocks. (Abdelouahed et al., 2012) Since leaching is not provided in Aspen Plus v8.2, a user model linked with the script compiled in FORTRAN was used to simulate the leaching column. The model assumes that the leaching column operates in steady-state; that all components are entirely transferred into the DT unit and the 1st evaporator; and that the difference in temperature between the inlet and outlet flow is negligible.

The mass balance for component *i* between inlet and outlet flow is given by:

$$\sum_{1}^{j} L(IN)_j x_{i,j} = \sum_{1}^{k} L(OUT)_k x_{i,k} \quad \text{(S3)}$$

With *j* as the number of inlet flows and *k* as the number of outlet flows (*j* = 2, *k* = 2). Heat transfer was neglected in this specific module, but was accounted for in the subsequent units.

1.5 Results of process simulation

To evaluate the accuracy of the simulation, essential results were compared with the field data. Operational temperatures, pressures, the quality of the products, and the steam and cooling water consumptions were included. The comparison between simulation results and operational data is shown in Table S2 (operational units) and Table S3 (key flows). Despite some expected differences between the average simulation values and real-plant data, both field and simulation data are in good agreement. The total steam consumption in this simulation is 2.426 ton/h, close to the real situation (2.4 ton/h), whilst the total cooling water consumption was 392.6 m^3/h in this simulation. The excellent agreement confirms that the simulation approach was effective and valid, thus stressing the fact that a simulation package, widely used in chemical engineering processes, can be successfully applied and upgraded towards the use in more complex biochemical processes. The simulation approach was further applied to investigate possible energy savings. The simulated steam consumption is 2426kg/h, very closed to the real ~2400kg/h. The simulated cooling water consumption is 392.6 m^3/h, in agreement with the plant data.

Table S2

Comparison between simulation results and real operational data (Operational units)

Unit		Temperature(°C)		Pressure (kPa)		Steam consumption in simulation (kg/h)	
		Simulation	Real	Simulation	Real	Simulation	Real
Leaching column		50	47~53	100	101	32	/
DT column	Top	80.7	80~85	100	101	1491.2	/
	Bottom	106.4	105~107	100	101		/
1st Evaporator		52.5	50-60	50	50~70	277.5	/
2nd Evaporator		120	115-120	50	50~70	191.7	/
Stripping column	Top	121.3	120	50	50~70	50.0	/
	Bottom	122.7	125~130	50	50~70		
Absorption column		40	40	100	101		
Stripping column (tail gas)		121.5	120	100	101	384.0	/

Table S3

Comparison between simulation results and real operational data (Key flows)

Flow	RB (kg/h)		Miscella after the leaching (kg/h)		Miscella after 1st stage of evaporator (kg/h)		Miscella after 2nd stage of evaporator (kg/h)		RBO (kg/h)	
	Real	Simulation	Real	Simulation	Real	Simulation	Real	Simulation	Real	Simulation
Defatted RB	5670	5670*								
Palmitic acid	29.87	29.87		29.9		29.9		29.9		29.9
Oleic acid	95.58	95.58		95.6		95.6		95.6		95.6
Linoleic acid	77.66	77.66		77.7		77.6		77.6		77.6
Monoglyceride	71.69	71.69		71.7		71.7		71.7		71.7
Diglyceride	41.82	41.82		41.8		41.8		41.8		41.8
Triglyceride	943.86	943.86		939.1		911.9		911.9		903.6
alpha-Tocopherol	11.95	11.95		11.9		11.9		11.9		11.9
beta-Sitosterol	83.63	83.63		83.6		83.6		83.6		83.6
Octacosanol	17.92	17.92		17.9		12.5		12.5		10.8
H2O	527~692	610								4.7
C6H14-01			4122~5496	4600.0	587~913	838.4	<70	21.9	<4	trace
Total	7527~7692	7654		5974		2207.2		1358.5	1333	1331.3

Refrence

1. Kong, W.; Kang, Q.; Feng, W.; Tan, T., Improving the solvent-extraction process of rice bran oil. *Chemical Engineering Research and Design* **2015,** 104, 1-10.