Castor Oil Drying Plant

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1 Executive Summary

Neglecting capital expenditure and considering only operating costs, the most economically preferential process conditions under which the plant should be run are using a large reactor and 300°C operating temperature. These conditions yield an annual profit of 3.3 million USD, and an annual yield is 12,501 tons of improved dehydrated castor oil. This is a viable business venture if the market has a bright future for the time between design and production and the CAPEX is not prohibitive.

Considering the process uses no specialised machinery, and the castor oil is already in storage in the company in abundance – it is recommended that production starts as soon as possible in order to capitalise on current market conditions.

These calculations are based off operating 8,000 hours per annum, this allows for down time and maintenance of the plant as required. With the current information, it seems the operation is initially profitable – although further investigation and calculation is required in order to quantify if the yearly profit outweighs the initial investment required given the short lifetime of the product due to alternatives in development.

2 Introduction

2.1 Product usage

Drying oils such as Castor oil are products used primarily in the finishing industry to create durable finishes with unique properties. Using an oil based solvent instead of an aqueous base allows painting on a wider range of surfaces such as metal where aqueous base paints and finishes do not easily adhere to the surface. The current process of removing hydrogen from castor oil leaves it in a less than optimal state to be used as a drying oil, the novel process discussed within this report creates a more desirable end product – and hence will be appealing to the markets.

2.2 Product market

Due to the toxicity of evaporating solvents, the market is moving away from solvent based finishes as a whole. They still have their uses as aqueous solutions are not yet able to adhere to some metals – but by time the process plant is finalised and commissioned, the market may have suffered even further due to advancements in alternatives. Short term the market for drying oils is expected to grow, hence the profitability of the operation is highly time dependant. A recent financial report [2] reports the current annual growth rate as 5 %, with the market expected to reach 2 Billion USD by 2028. Combining this with the improved product and a suitable patent, this plant could have potential to generate a large profit.

2.3 Location research

India has a large amount of castor seed farms, and hence castor seeds and oil retail at a much lower price per kilogram than other nations. The excess of oil held by the company could be utilised via the processes outlined below into a useful product – but it may depend on the timescale in which the plant can become operational to define the profitability of the operation.

2.4 Process assumptions

The process calculation technique employed within this report has a few key limitations. One of which is the lack of consideration for CAPEX – the initial investment cost in the plant equipment or design consultancy fees. The main consideration mathematically for the calculation was simple mass balances with reactions being considered in terms of molar equivalence. This is an acceptable assumption to employ as details of the conversion/selectivity and separation efficiencies of each of the separation columns are given – and hence energy considerations on this level do not need to be made.

The analysis does not take into account variations in feed stock, as castor oil is assumed to be pure methylated ricinoleic acid – these delivery to delivery variations could cause issues with the process not considered within this report.

3 Analysis Results

3.1 Most Economically Favourable Case (OPEX only)

For the process highlighted in the below process flow diagram, manipulation of each stream in an iterative program (full details shown in Appendix A was undertaken to determine the most economically favourable process was able to be determined. This yielded the following results as shown in table 1. Other considerations aside, this is the case recommended as 'best' with the information provided in this brief.

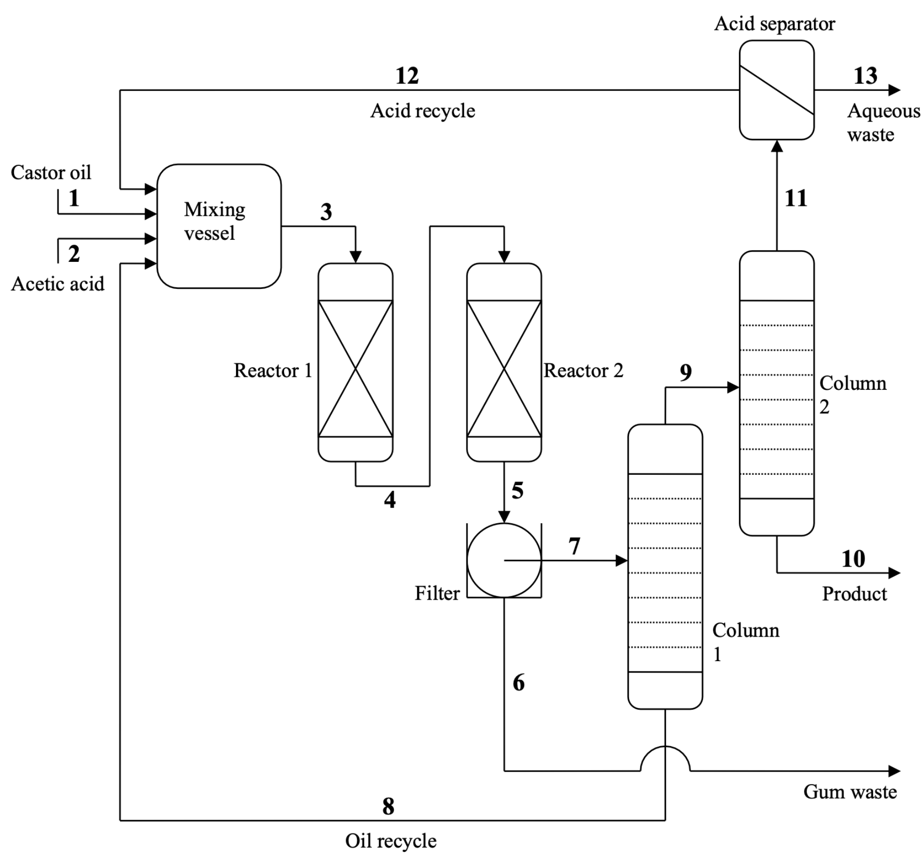

Figure 1: This process diagram is from reference [1], it shows the novel process used as reference to create the most economically viable scenario.

Table 1: Table to show the conditions of the processing plant under optimum operating conditions as determined by calculation

Condition	Value
Reactor Size	Large
Temperature / °C	300
CO Feedrate / kmol·hr^{-1}	5.33
AA Feedrate / kmol·hr^{-1}	0.47
DCA / tons yearly	12501
Profit / $millions yearly	3.35

3.2 Process Flow Diagram and Full Stream Tables

Through the process of evaluating all possible combinations of selectivity and conversion in each of the reactor sizes, the maximum profit scenario was reached by picking the maximum value of profit for all those which were evaluated. The component mass flowrates could then be read for each of the streams and were compiled in the table below. They can be verified by the reader by viewing Appendix A.

Table 2: Stream tables showing the mass flow rates of each component as well as the total mass and molar flow rates within each stream in the processing plant as outlined in the above PFD

Stream Number	1	2	3	4	5	6	7
Total Massflow ($Mg \cdot hr^{-1}$)	1.66	$2.80{\cdot}10^{-2}$	4.25	4.25	4.25	$7.30{\cdot}10^{-4}$	4.25
Total Molar Flowrate ($kmol \cdot hr{-}1$)	5.33	0.47	20.20	20.20	25.50	$1.24{\cdot}10^{-3}$	25.50
CO $\dot{m}$ ($Mg \cdot hr^{-1}$)	1.66	0.00	1.85	0.19	0.19	0.00	0.19
ACO $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.00	1.78	3.67	1.79	0.00	1.79
DCO $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.00	$8.23{\cdot}10^{-2}$	$8.23{\cdot}10^{-2}$	1.64	0.00	1.64
AA $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	$2.80{\cdot}10^{-2}$	0.53	0.21	0.53	0.00	0.53
H_2O $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.0	$1.50{\cdot}10^{-2}$	$9.74{\cdot}10^{-2}$	$9.74{\cdot}10^{-2}$	0.00	$9.74{\cdot}10^{-2}$
Gum $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.00	0.00	0.00	$7.30{\cdot}10^{-4}$	$7.30{\cdot}10^{-4}$	0.00

Stream Number	8	9	10	11	12	13
Total Massflow ($Mg \cdot hr^{-1}$)	2.05	2.20	1.57	0.63	0.51	0.12
Total Molar Flowrate ($kmol \cdot hr{-}1$)	5.90	19.60	5.33	14.28	8.49	5.79
CO $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.19	$9.24{\cdot}10^{-4}$	$9.24{\cdot}10^{-4}$	0.00	0.00	0.00
ACO $\dot{m}$ ($Mg \cdot hr^{-1}$)	1.78	3.57	3.57	0.00	0.00	0.00
DCO $\dot{m}$ ($Mg \cdot hr^{-1}$)	$8.22{\cdot}10^{-2}$	1.56	1.56	0.00	0.00	0.00
AA $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.53	0.00	0.53	0.50	$27.7{\cdot}10^{-2}$
H_2O $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	$9.74{\cdot}10^{-2}$	0.00	$9.74{\cdot}10^{-2}$	$1.53{\cdot}10^{-3}$	$9.59{\cdot}10^{-2}$
Gum $\dot{m}$ ($Mg \cdot hr^{-1}$)	0.00	0.00	0.00	0.00	0.00	0.00

3.3 Relative Performance with Varied Conditions

It was possible using the same code utilised for the optimal scenario to plot a graph relating profit to temperature for each of the two reactor sizes. There is an optimal spot around 320 degrees for the small reactor where the profit becomes only marginally different than the large reactor. Given the chance to explore pricing further, it would be possible to quantify if the change to the small reactor at a slightly higher temperature would be justified.

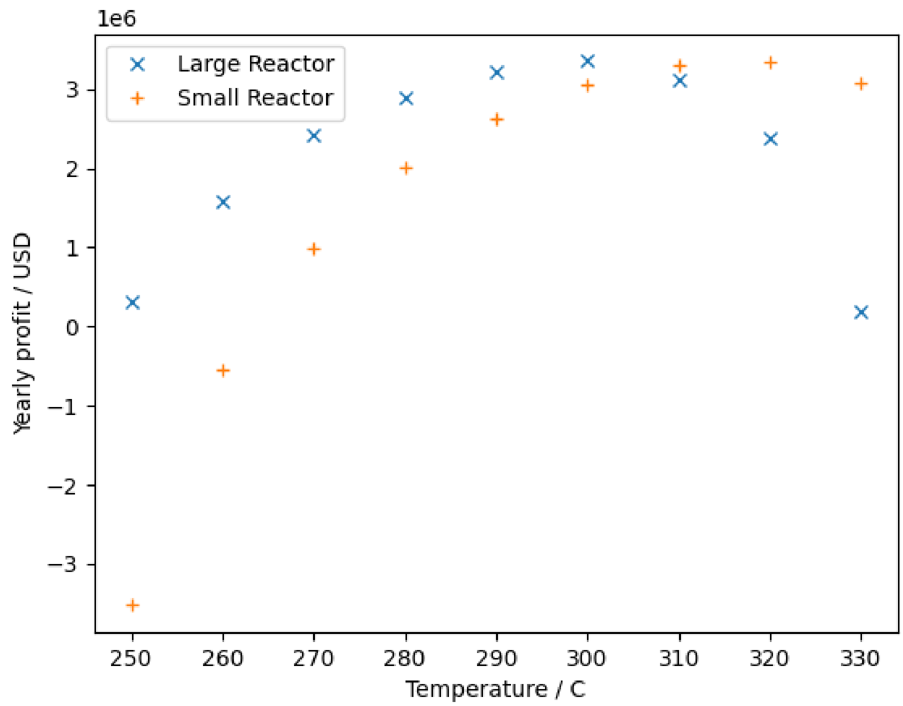

Figure 2: Profit plotted against temperature for each of the two reactor sizes

4 Discussion

Based on the figures calculated and displayed in the earlier most economic conditions section, there is enough evidence at first to suggest that it is feasible to make a profit from this process. However there is no consideration on costs other than OPEX and current sale rates – so it is entirely possible that when market changes and commissioning and construction costs are considered that the plant is considered as unfeasible. There needs to be a process and instrumentation diagram drawn up to allow for full costing and hence factor in all initial and long term costs.

As discussed briefly within the above section, given the relative closeness of the 320 degree small reactor and the 300 degree large reactor profit wise, it may be sensible to pick the smaller reactor when trying to minimise the original capital expenditure. Given the current energy crisis, it may be wise to spend more on material cost to have a reactor capable of producing the same amount of profit at a lower temperature.

Smaller scale tests of the difference in efficacy of the new drying oil in practical applications may be in order to be able to determine if the difference is quantifiable. It would be easier to sell the product in an industrial setting if the novel process produced a product which was quantifiable different in terms of time, money, or maintenance time. This report is only a preliminary glimpse into what needs to be reviewed prior to moving forward with any part of the process plant facility.

This analysis fails to take into account anything about the flow capacity of the unit operations, nor the pipes that would be required in order to safely transport fluids at the flowrates and pressures aforementioned in the analysis.

5 Conclusion

Using computing methods, it was possible to determine the optimal operating conditions for the plant to maximise profit considering only material and operating costs. Neglecting any further considerations, it is possible to conclude that the optimal conditions as determined above are 300 degrees Celsius in a large reactor. A cost analysis would have to be performed on if the added material cost would outweigh the energy saving over the lifetime of the plant.

Overall a healthy product and a growing market over the coming 5 years suggests that if the plant is able to become operational relatively soon, it would be a good venture. More information is needed to set the unit operation sizes and operating conditions properly. There may be other considerations such as safety and storage, as it may not be feasible to store large amounts of precursor, or maintaining the temperature in the exothermic acid oil reaction may not be controllable and safe with the novel process.

Nevertheless, with the information given and under initial consideration – this process is worth investing more time into at the minimum in order to properly cost everything up and get a long term insight into the project's future. The limitations of the investigation method are too great to be able to objectively state if the project is feasible.

References

[1] CET Part I Exercise Sheet 1. Process Calculations. Drying Oil Exercise. KY & MT. University of Cambridge, 2022, P2.

[2] Link. At 5% CAGR, Global Castor Oil Derivatives Market Size to Hit USD 2 Billion by 2028 - Exclusive Report by Facts & Factors. Castor Oil Global Market Report. 2021. GlobeNewsWire -Author Unknown. Accessed 3/11/22.

A Process Simulation Program / Calculation Methodology

A python program was used in order to set up streams then feed them into functions which represented different unit operations on the PFD. The program iterated from an initial feedrate until steady state is reached, changing the feedrate based on requirement to produce 12,500 tonnes per year. The initial excess to reach steady state may reduce the profit, but it was important to simulate as close to reality as possible. The full codebase can be found here – the unit operations are in the functions folder, the streams and chemical classes can be found in the classes folder – and the main method represents the runscript. PEP-8 was used where appropriate, and due to the nature of how the code was written it is able to be unit tested & can be easily debugged. An appropriate number of iterations were used, where it was found that after that number the result of the algorithm did not change. The codebase contains all the nuances of the calculations, and should be relatively easy to follow for the average reader.(Aside: I hope the code is interesting to whomever is marking this, I thought it was a more interesting avenue to explore instead of assuming steady state and doing overall mass balances.)