

Chapter 11

**DESIGN AND OPTIMIZATION OF REACTIVE
DISTILLATION SEQUENCES WITH THERMAL
COUPLING WITH MINIMUM NUMBER OF REBOILERS**

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ABSTRACT

The esterification of lauric acid and methanol is explored using thermally coupled distillation sequences with side columns and with minimum number of reboilers. The product of the esterification can be used as biodiesel. This is a major step forward since thermally coupled reactive distillation sequences with side columns and with minimum number of reboilers offer significant benefits, such as: reductions on both capital investment and operating costs due to the absence of the reboilers, higher conversion and selectivity as no products are recycled in the form of reflux or boil-up vapors, as well as no occurrence of thermal degradation of the products due to a lower temperature profile in the column. In the traditional process, lauric acid is fed as saturated liquid, while the methanol is fed as saturated vapor stream to the main column, then we can take heat from vapor stream to deliver it to the system without a reboiler. Rigorous simulations were performed in AspenTech AspenONE engineering suite to design this novel reactive distillation process and evaluate the technical and economical feasibility. The results indicate that the energy consumption of the complex distillation sequences with a side column can be reduced significantly by varying operational conditions, in comparison with conventional reactive distillation sequences.

NOMENCLATURE

D	≡	Diameter
P	≡	Pressure top
Q _C	≡	Condenser duty
Q _B	≡	Reboiler duty
T _C	≡	Condenser temperature
T _B	≡	Reboiler temperature
T _D	≡	Dew point temperature
P _B	≡	Pressure bottom
RR	≡	Reflux ratio
R _D	≡	Distillate rate
R _B	≡	Bottoms rate
F _{FA}	≡	Flow of fatty acid
F _M	≡	Flow of methanol
S	≡	Number of stages
S _{FA}	≡	Feed stage fatty acid
S _M	≡	Feed stage methanol
S _{II}	≡	Feed stage column II / Rectifier / Stripper
S _R	≡	Number of reactive stage
S _{RI}	≡	Starting reactive stage
S _{FV}	≡	Interconnection stage vapor phase
S _{LV}	≡	Interconnection stage liquid phase
F _{FV}	≡	Interconnection flow vapor phase
F _{LV}	≡	Interconnection flow liquid phase.

INTRODUCTION

The world is presently confronted with the twin crises of fossil fuel depletion and environmental degradation. Indiscriminate extraction and lavish consumption of fossil fuels have led to reduction in underground-based carbon resources. The search for alternative fuels, which promise a harmonious correlation with sustainable development, energy conservation, efficiency and environmental preservation, has become highly pronounced in the present context. The fuels of bio-origin can provide a feasible solution to this worldwide petroleum crisis. Gasoline and diesel-driven automobiles are the major sources of greenhouse gases emission (Johansson and McCarthy, 1999; Kesse, 2000; Cao, 2003). Scientists all around the world have explored several alternative energy resources, which have the potential to quench the ever-increasing energy thirst of today's population. Various biofuel energy resources explored include biomass, biogas (Murphy and McCarthy, 2005), primary alcohols, vegetable oils, biodiesel, etc. These alternative energy resources are largely environment-friendly but they need to be evaluated on case-to-case basis for their advantages, disadvantages and specific applications. Some of these fuels can be used directly while others need to be formulated to bring the relevant properties closer to conventional fuels (Agarwal, 2007).

Biodiesel, the common name for fatty acid methyl esters (FAMES), is a liquid fuel obtained from the transesterification of vegetable (or animal) oils. It involves simply the reaction, under very mild conditions, between vegetable oil and typically a large excess of alcohol of methanol, in the presence of an acid or basic catalyst, which produces FAMES as the main product and glycerol as a byproduct (Di Felice et al., 2008). Biodiesel has drawn significant attention due to increasing environmental concern and diminishing petroleum reserves. Biodiesel is a renewable and environmentally friendly energy. Application of this energy not only can significantly reduce the pollution generated from petroleum based diesel oil but also can lessen the dependence on petroleum. Presently, biodiesel is produced commercially in Europe and USA to reduce air pollution and the net emission of greenhouse gas. Surplus edible oils, such as rapeseed oil, soybean oil or castor oil, used cooking oil, rice bran oil, and algae, among others, are used as raw materials for biodiesel (Wahlen et al., 2008; Zhang et al., 2008; Lu et al., 2009; Peña et al., 2009). Outstanding benefits of biodiesel are the high cetane numbers that it naturally achieves and the lack of polluting heteroatoms like sulfur or nitrogen. Cetane number increase and sulfur/nitrogen content reduction are the most capital intensive hydrotreatment processes that a conventional refinery has to handle, a fact that lays the ground for the biodiesel niche in the near-future refining business (Stiefel and Dassori, 2009). Also, biodiesel is safe, renewable, nontoxic, and biodegradable; it contains no sulfur and is a better lubricant.

The current manufacturing biodiesel processes, however, have several disadvantages: shifting the equilibrium to fatty esters by using an excess of alcohol that must be separated and recycled, making use of homogeneous catalysts that require neutralization (causing salt waste streams), expensive separation of products from the reaction mixture, and high costs due to relatively complex processes involving one to two reactors and several separation units. Therefore, to solve these problems, some authors (for example, Kiss et al., 2008) have developed a sustainable biodiesel production process based on reactive distillation using acid catalysts. Reactive distillation integrates reaction and separation in one unit. This intensifies mass transfer and allows in situ energy integration while simplifying the process flowsheet and operation. However, combining the two operations is possible only if the reactions show reasonable conversion and selectivity data at pressures and temperatures that are compatible with the distillation conditions. The reduction in the number of processing units and the direct heat integration between reaction and separation can reduce capital investment as well as utility costs. Increased overall conversion, as well as improved selectivity in competing reactions, can be achieved in reactive distillation by the continuous removal of products from the reaction zone of equilibrium limited reactions (Ciric and Gu, 1994). Other process improvements can be realized such as reducing byproduct formation, improving raw material usage, and overcoming chemical equilibrium limitations.

Design issues for reactive distillation systems are significantly more complex than those involved in ordinary distillation (Doherty and Buzad, 1992). Catalyst selection, liquid holdup on each tray, and position of feeds become important design considerations. Reaction often occurs in the liquid holdup so that the reaction volume is a major design parameter, and constant molar overflow cannot be assumed. Also, a single feed may not be appropriate and a distributed feed must be considered. Thermally coupled distillation systems (TCDS) are obtained through the implementation of interconnecting streams (one in the vapor phase and the other one in the liquid phase) between two columns; each interconnection replaces one condenser or one reboiler from one of the columns, thus providing potential savings in capital

investment. Furthermore, through a proper selection of the flow values for the interconnecting streams of TCDS, one can obtain significant energy savings (consequently, reductions in CO₂ emissions) with respect to the energy consumption of conventional distillation sequences. There is a considerable amount of literature on the analysis of the relative advantages of TCDS for ternary separations with equilibrium and nonequilibrium stage models (Triantafyllou and Smith, 1992; Hernández and Jiménez, 1996; Hernández and Jiménez, 1999; Dünnebier and Pantelides, 1999; Yeomans and Grossmann, 2000; Emtir et al, 2003; Olujić et al, 2003; Hernández et al, 2006; Abad-Zárate et al, 2006, among others). These studies have shown that those thermally coupled distillation schemes are capable of achieving typically 30% of energy savings compared to the conventional schemes. Recently some studies have reported the possibility of implementing the process of reactive distillation in thermally coupled distillation sequences. In these conditions it is possible to combine the kindness of the reactive distillation and the savings of energy widely reported for the systems with thermal coupling (Barroso-Muñoz et al. 2007; Wang et al. 2008; Hernández et al. 2009). Thus, in this paper, the production of biodiesel by esterification of methanol and lauric acid is studied using conventional, thermally coupled distillation sequences with side columns and thermally coupled distillation sequences with a side column with minimum number of reboilers (Figure 1 -3).

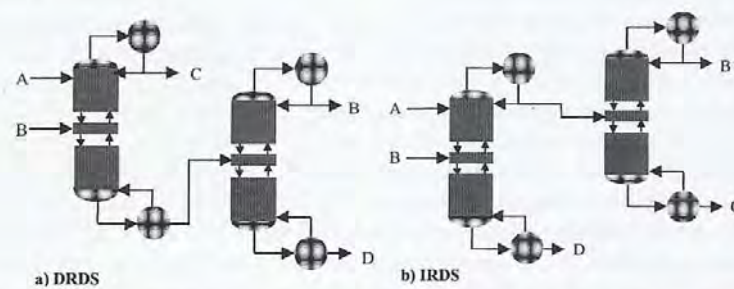


Figure 1. Conventional reactive distillation sequences: (a) direct (DRDS), and (b) indirect (IRDS).

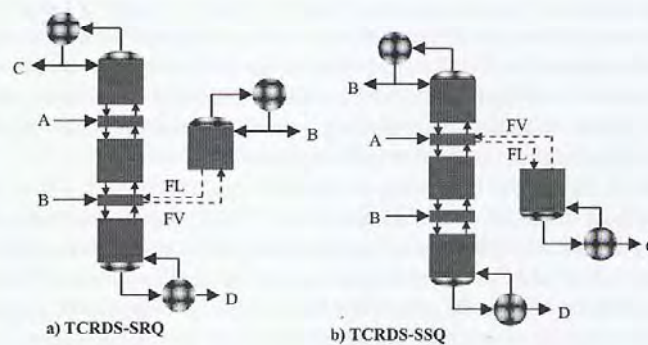


Figure 2. Thermally coupled reactive distillation sequences: (a) with side rectifier (TCRDS-SRQ) and (b) with side stripper (TCRDS-SSQ).

The results indicate that the energy consumption of the complex distillation sequences with a side column and minimum number of reboilers can be reduced significantly by varying operational conditions, in comparison with conventional and thermally coupled systems when reactive systems are included.

PROBLEM STATEMENT

There are three basic methods to produce fatty esters from oils/fats: 1) base catalyzed trans-esterification, 2) acid catalyzed esterification, and 3) enzymatic catalysis. The first method is the most frequently used. However, due to the escalating costs of fatty raw materials, the current trend is to use less expensive alternatives such as animal fat, waste cooking oil from catering premises, or waste vegetable oil. The problem with waste oils is the very high content of free fatty acids (FFA) that lead to soap formation in a conventional base catalyzed process. Therefore, in order to avoid production loss and soap associated problems, the FFA's must be completely converted first to fatty esters by esterification. Moreover, the conventional biodiesel processes employ liquid catalysts, such as H_2SO_4 , NaOH or methoxides. The problem is that homogeneous catalysts require neutralization, washing, separation, recovery, and waste disposal operations with severe economical and environmental penalties. To solve this problem we propose an alternative fatty esterification process based on thermally coupled reactive distillation with minimum number of reboilers using acid as catalysts and therefore eliminating the additional separation steps and the salt waste streams, thus simplifying the downstream processing.

The alternative technology proposed in this work offers advantages compared to conventional reactive distillation, as for example diminution in the use of energy in the system and low temperature profile in the reactive separation column, due to the elimination of reboilers to avoid the thermal degradation of the fatty esters products. As a result, the process presents reductions in both capital investment and operating costs due to the absence of a reboiler (no product vapors return to the column). This process approach based on reactive distillation is particularly suitable for treating waste oil or animal fat, including triglycerides with up to 100% free fatty acids (FFA).

In this work we analyzed three different types of reactive distillation systems: the conventional case, the thermally coupled arrangement and the thermally coupled arrangement with minimum number of reboilers. Each of these cases is analyzed in the direct and indirect structures.

We took as base designs of the distillation sequences for conventional and thermally coupled reactive systems designs, those obtained by Miranda-Galindo et al., 2011, see Figures 1 and 2. Table 1 shows the design parameters of the configurations of conventional and thermally coupled distillation schemes. Based on the thermally coupled systems we proceeded to remove the maximum amount of reboilers to generate alternative distillation systems. In the case of thermally coupled distillation sequence with side rectifier was possible to remove all the reboilers of the system. In the case of thermally coupled distillation sequence with side stripper is only feasible to eliminate the reboiler of the main column and must remain the reboiler in the side stripper (Figure 3).

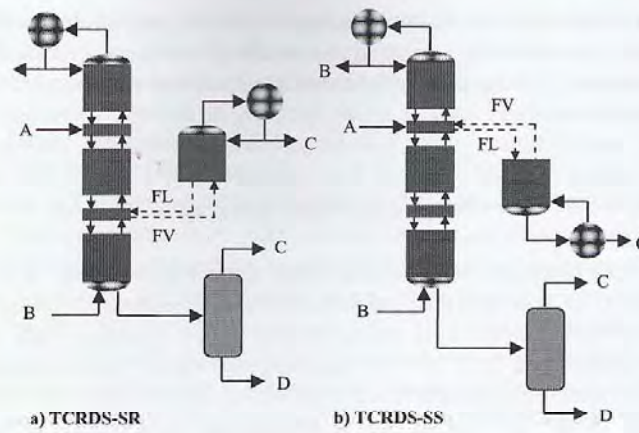


Figure 3. Thermally coupled arrangement with minimum use of reboilers: (a) with side rectifier (TCRDS-SR) and (b) with side stripper (TCRDS-SS).

Table 1. Parameters of the configurations of conventional and Thermally Coupled Distillation Schemes

Sequences	DRDS		IRDS		TCRDS-SRQ		TCRDS-SSQ	
	Column I	Column II	Column I	Column II	Column main	Rectifier	Column main	Stripper
RR	2.6665	0.06072	2.3100	1.6953	2.0568	0.3468	19.5116	6.1894
R_D [lbmol/h]	100.0240	19.3662	119.3477	19.8149	101.0767	18.2670	19.0155	19.2013
R_B [lbmol/h]	119.9760	100.6098	100.6523	99.5328	100.6563	1.5974	100.6398	100.3446
F_{FA} [lbmol/h]	100.0000	-	100.0000	-	100.0000	-	100.0000	-
F_M [lbmol/h]	120.0000	-	120.0000	-	120.0000	-	120.0000	-
S	21	7	10	15	16	5	22	2
S_{FA}	4	-	3	-	3	-	10	-
S_M	20	-	10	-	15	-	22	-
S_{II}	-	4	-	9	-	-	-	-
S_{FV}	-	-	-	-	12	5	7	1
S_{LV}	-	-	-	-	11	5	7	1
F_{FV} [lbmol/h]	-	-	-	-	19.8644		19.2013	
F_{LV} [lbmol/h]	-	-	-	-	1.5974		119.5460	
S_R	20	-	7	-	15	-	21	-
S_{RI}	2	-	3	-	3	-	3	-

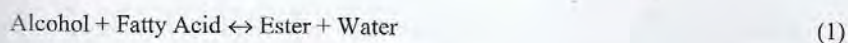
Strictly, the design of the conventional and reactive thermally coupled distillation sequences could be modeled through superstructures suitable for optimization procedures with mathematical programming techniques. However, the task is complicated and is likely to fail to achieve convergence. In this case, to overcome the complexity of the simultaneous solution of the tray arrangement and energy consumption within a formal optimization algorithm, we decoupled the design problem in two stages: (1) tray configuration; (2) energy efficient design (optimal energy consumption). The first stage of our approach begins with the development of preliminary designs for the alternative systems with minimum number of reboilers based on the designs proposed by Miranda-Galindo et al., 2011.

After the tray arrangement for the alternative coupled designs have been obtained, an optimization procedure is used to minimize the total heat duty supplied to the schemes, taking into account the constraints imposed by the required purity of the product streams. The degrees of freedom that remain after the design specifications and tray arrangement are used to obtain the operating conditions that provide minimum total energy consumption. Three degrees of freedom that remain for each integrated sequence are the interconnecting flow (vapor or liquid, depending of the scheme) and both pressures of the feed flowrates. The search procedure provides the optimal values of the interconnecting flows and pressures to minimize the energy consumption for the reaction-separation system. The design is successful if it meets the product specifications.

The optimization strategy can be summarized as follows: (a) A base design for the complex schemes is obtained with a value of pressure for each feed flowrate (alcohol and fatty acid). (b) A value for each interconnecting flow (vapor or liquid) is assumed. (c) A rigorous model for the simulation of alternative coupled schemes with the proposed tray arrangement is solved. In this work Aspen Plus was used for that purpose. If the product compositions are obtained, then the design is kept; otherwise, proper adjustments must be made. (d) One value of interconnecting flow is changed, going back to step (c) until a local minimum in energy consumption for the assumed values of the pressure for each feed flowrate is detected. (e) The value of a pressure of alcohol is modified (the pressure of the fatty acid is constant), going back to step (c) until the energy consumption is minimum. (f) The value of the pressure of the fatty acid is modified, going back to step (c) until the minimum energy consumption supplied to the system is obtained. This result implies that an optimum value has been detected for the design of the complex schemes.

CASE OF STUDY

The esterification process, studied in this work, can be represented conceptually by equation 1.



This equilibrium reaction can be favored if the products are removed as the reaction proceeds. An additional problem may present itself, depending on the acid and the alcohol used, binary or ternary homogeneous azeotropes can be formed in the reactive system. For highly non-ideal systems, heterogeneous azeotropes can be formed. These key factors must be

considered in order to select the appropriate thermodynamic model when the system is studied with process simulators. For this class of reactive systems, thermodynamic models such as NRTL can be used to calculate vapor-liquid or vapor-liquid-liquid equilibrium. For this study (esterification of methanol and lauric acid), we selected the NRTL model.

Table 2. Kinetic parameters for the pseudohomogeneous kinetic model of the esterification reaction

Reaction	k^0 [mol/g.s]	E_A [kJ/mol]
Esterification	9.1164×10^5	68.71
Hydrolysis	1.4998×10^5	64.66

The systems include two feed streams; the first is lauric acid with a flow of 45.0 kmol/h as saturated liquid, and the second is methanol with a flow of 54.48 kmol/h as saturated vapor. Through an optimization study where the objective was to reduce the total energy load of the system, we obtained the optimum operating pressure of the feed flowrates. The design objective is a process for high-purity fatty ester, over 99.9% mass fraction, suitable for applications in cosmetics, detergents, surfactants or biodiesel application. It is important to highlight that this equilibrium reaction is usually catalyzed using sulfuric acid or p-toluensulfonic acid. The kinetic model (see Table 2) reported in Steinigeweg and Gmehling (2003) was used.

RESULTS

The first stage of our approach begins with the development of preliminary designs for the alternative systems with minimum number of reboilers based on the designs proposed by Miranda-Galindo et al., 2011 (Table 1).

The strategy for the design of thermally coupled sequences with minimum number of reboilers, followed by a search procedure on the interconnecting streams to optimize the design for energy consumption was conducted. A typical optimization search for the TCRDS-SR is shown in Figure 4, where the combination of interconnecting flowrate and pressure that provide the minimum energy consumption can be detected. The tray arrangements and some parameters for that sequence after the optimization task are given in Table 3. The results obtained for energy consumption and total annual cost for each case study are summarized in Tables 4-6. In general, the TCRDS-SR presented energy savings in the range between 15-30% in contrast to conventional and thermally coupled distillation sequences (Table 4). Importantly, the TCRDS-SR does not have any energy as vapor, as the scheme has no reboiler.

The objective function was set to minimize the consumption of cooling water in condensers. For the TCRDS-SR, optimal interconnecting flowrate is 16.0 lbmol/h, while operation pressure is 1 atm (Figures 4-5). Design objective was considered a mass fraction purity of 99.9 % for biodiesel and 99.6% for methanol and water.

Table 3. Parameters of Thermally Coupled Distillation Schemes with minimum number of reboilers

Secuencias	TCRDS-SR		TCRDS-SS	
	Column I	Rectifier	Column I	Stripper
RR	1.4720	0.7432	36.0000	2.2482
R_D [lbmol/h]	100.0835	13.0986	11.0302	78.8609
R_B [lbmol/h]	106.8180	2.9014	107.8308	101.1391
F_{FA} [lbmol/h]	100.0000	-	100.0000	-
F_M [lbmol/h]	120.0000	-	120.0000	-
S	16	5	23	3
S_{FA}	3	-	10	-
S_M	16	-	23	-
S_{FV}	11	5	7	1
S_{LV}	4	5	7	1
F_{FV} [lbmol/h]	16.0000		78.8610	
F_{LV} [lbmol/h]	2.9014		180.0000	
S_R	16	-	18	-
S_{RI}	2	-	6	-

Table 4. Conventional reactive distillation sequences: information about total annual cost

Sequences	DRDS		IRDS	
	Column I	Column II	Column I	Column II
D [m]	0.5480	0.3984	1.0990	0.3360
S	21	7	10	15
P [atm]	1.0000	1.0000	1.0000	1.0000
Q_C [kW]	-1886.8014	-94.0816	-2052.4111	-238.6789
Q_B [kW]	1.6437	919.9382	962.8183	258.5297
T_C [°C]	99.9766	64.5559	83.7987	65.0037
T_B [°C]	125.1838	245.3968	243.6071	103.2329
T_D [°C]	124.9446	145.1890	242.9673	102.5260
P_B [atm]	1.2160	1.1374	1.1291	1.1300
	Costs [\$/year]			
Equipment	119,659.57	60,550.05	126,742.78	71,793.34
Utility	13,526.45	206,907.34	230,321.80	37,779.40
TOTAL	400,643.41		466,637.32	

Table 5. Thermally Coupled Reactive Distillation Sequences: information about total annual cost

Sequences	TCRDS-SRQ		TCRDS-SSQ	
	Column I	Rectifier	Column I	Stripper
Data				
D [m]	0.8294	0.2047	0.6047	0.1922
S	16	5	22	2
P [atm]	1.0000	1.0000	1.0000	1.0000
Q _C [kW]	-1596.6677	-109.5481	-1738.1438	0.0000
Q _B [kW]	628.1472	0.0000	626.4624	96.4604
T _C [°C]	98.0707	64.5418	64.8576	97.3713
T _B [°C]	243.1699	144.6501	250.0000	98.8525
T _D [°C]	167.9259	71.6979	177.1609	97.3713
P _B [atm]	1.1180	1.1975	1.3010	1.0077
	Costo [\$ /year]			
Equipment	128,919.28	18,710.47	150,345.81	15,205.18
Utility	152,078.88	772.02	152,698.20	13,468.33
TOTAL	300,481		331,718	

Table 6. Thermally Coupled Reactive Distillation Sequences with minimum number of reboilers: information about total annual cost

Sequences	TCRDS-SR			TCRDS-SS		
	Column I	Rectifier	Flash	Column I	Stripper	Flash
Data						
D [m]	0.7063	0.2034	1.0631	0.9262	0.3167	1.0670
S	16	5	-	23	3	-
P [atm]	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Q _C [kW]	-1281.3391	-101.85112	-	-1812.5305	0.0000	-
Q _B [kW]	0.0000	0.0000	611.01	0.0000	407.8718	578.92
T _C [°C]	99.9070	64.6788	-	64.6771	97.4002	-
T _B [°C]	177.1936	131.6704	240.0000	171.6022	114.4295	240.0000
T _D [°C]	193.3727	78.0306	142.6200	186.2989	106.8437	171.6000
P _B [atm]	1.6807	1.6807	-	1.6807	1.6807	-
	Costo [\$ /year]					
Equipment	102,040.66	20,271.42	12,772.21	178,862.12	27,749.92	12,826.07
Utility	9,030.04	717.78	136,984.81	12,773.53	56,949.32	129,789.67
TOTAL	281,817.91			426,963.00		

Figure 4 shows the dependence of energy consumption with respect to the operating pressure of the column and the pressure of the feed streams. This figure indicates that the minimum energy consumption is achieved when the pressures of the feed streams are the same as the operational pressure of the column. Now, when the operational pressures of the feed streams are different than the operating pressure of the column, the energy consumption does not change significantly. The same procedure was repeated for operational pressures of 0.5, 1.0, 1.5 and 2.0 atm, similar results to those mentioned previously are obtained. Figure 4 shows that the pressure that minimizes the energy consumption is 0.5 atm.

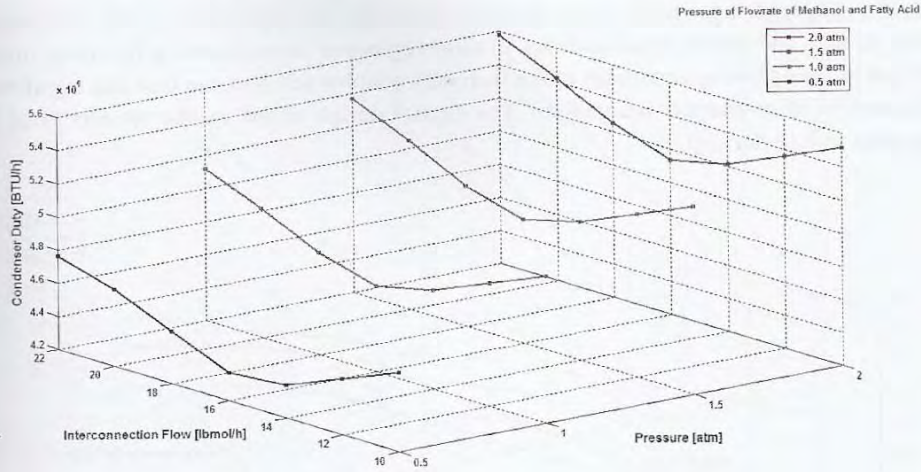


Figure 4. Condenser duty as a function of the interconnecting flow, operating pressure and pressure of feed flowrate of methanol, case TCRDS-SR.

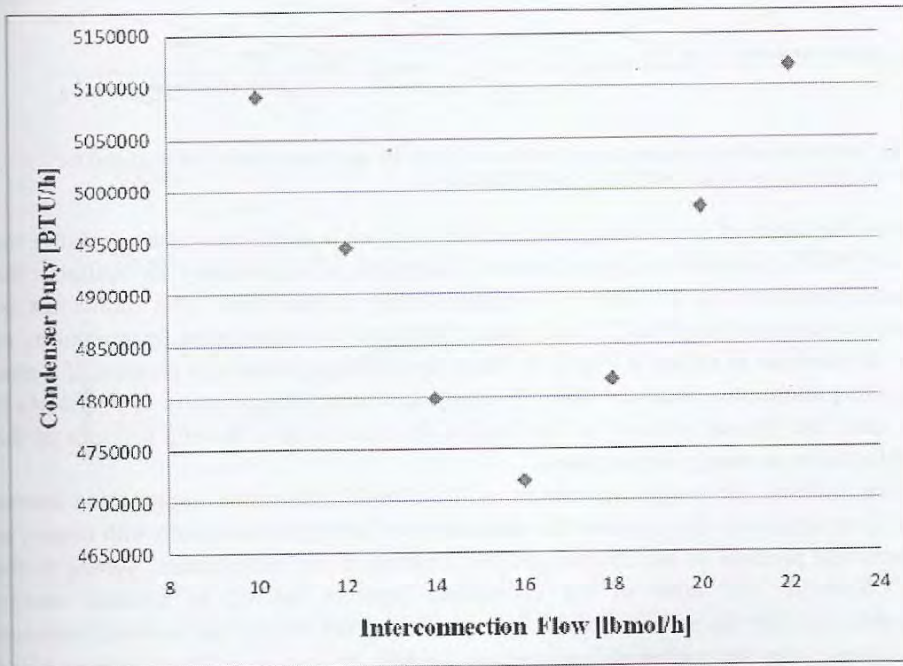


Figure 5. Condenser duty as a function of the interconnecting Flow, case TCRDS-SR.

However, it is well known that the operation in vacuum conditions substantially increases the total annual cost of the distillation column. As a result, the optimum operational pressure of 1 atm was chosen.

Figure 5 displays the optimization search of the TCRDS-SR option. The curve shows an interesting effect of the search variables. The design is sensitive, in terms of its energy

consumption, to changes in the interconnecting flowrates. An implication of this observation has to do with operational considerations. In some regions of interconnecting flowrates, minor changes in the operating conditions of the thermally coupled schemes can lead to a significant deterioration of its energy consumption. The control design of this system appears to be an important task to develop.

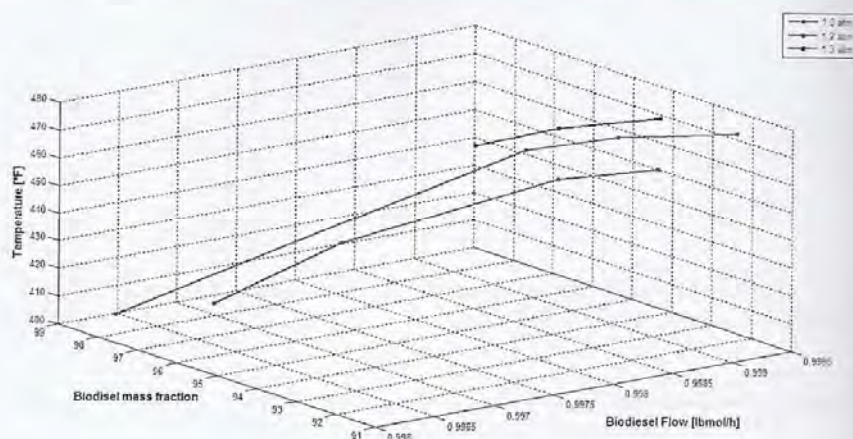


Figure 6. Temperature flash operation as a function of the flow and mass fraction of biodiesel for different operational pressures in flash, case TCRDS-SR.

Given the nature of the system, it is necessary to place an additional flash to achieve the purity of 99.9% biodiesel in mass fraction. Therefore it is necessary to optimize the operational conditions of the flash to minimize energy requirements. The flash must be operated at a temperature of 464 °F and 1 atm of pressure to obtain a mass fraction purity of 99.9 % for biodiesel as shown in Figure 6. When the flash is operated at a pressure of 1 atm, the operating temperature is lower (464 °F) obtaining a mass fraction purity of 99.9%. As it can be seen, the optimal pressure of the flash is identical to the operating pressure of the column to minimize energy consumption.

Table 3 shows the design parameters of TCRDS-SS. When the optimization curves (Figure 7) are analyzed, it is observed the dependence of energy consumption with respect to the operational pressure of the column and the pressure of the feed streams. Similar to the case TCRDS-SR, the value of the operational pressure leading to minimal energy consumption is when the pressures of the feed streams are the same as the operating pressure of the column; when the operational pressures of the feed streams are different to those of the column, the energy consumption does not change significantly. The optimal operational conditions found for this configuration are: operation pressure of the column, 1 atm; interconnecting flowrate, 180 lbmol/h. Figures 8 and 9 show the dependence of the energy consumption in the reboiler and condenser with the interconnecting streams. Clearly, the shape of the optimization curves show sensitivity of the energy consumption to changes in operating conditions. It is important to emphasize that the value of the interconnecting flowrate that minimizes energy consumption in the condenser and reboiler is the same.

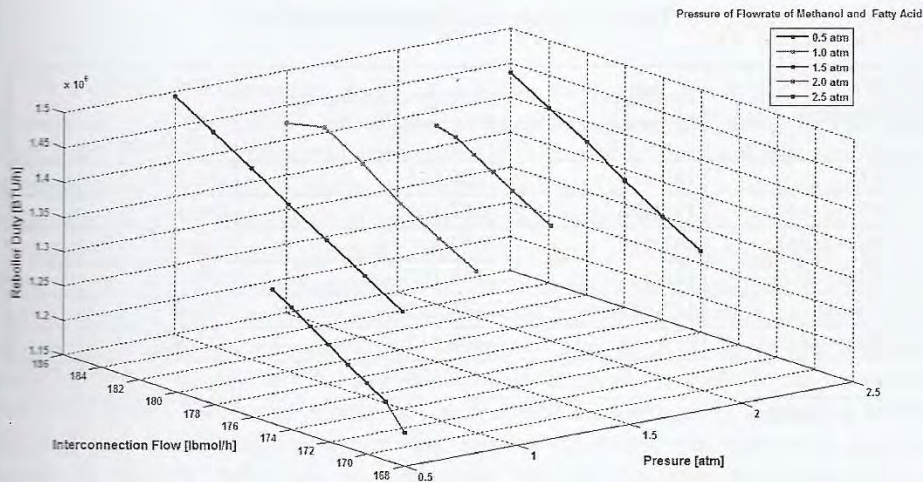


Figure 7. Reboiler duty as function of the operating pressure and interconnection flow, case TCRDS-SS.

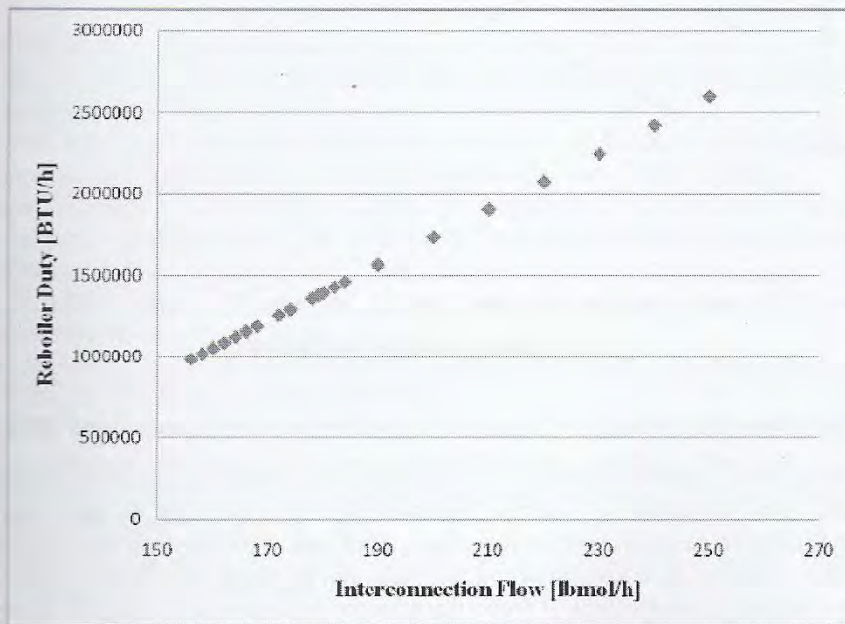


Figure 8. Reboiler duty as function of the interconnecting flow, case TCRDS-SS.

In this particular case, TCRDS-SS only presents savings in energy consumption in comparison with conventional reactive distillation sequences (~10–25%), but the configuration does not present energy savings when it is compared to thermally coupled columns.

Table 7. Total annual cost of sequences

Sequences	Equipment [UDS/year]	Energy consumption [UDS/year]	Total annual cost [UDS/year]
DRDS	180,210	220,434	400,643
IRDS	198,536	268,101	466,637
TCRDS-SRQ	147,630	152,851	300,481
TCRDS-SSQ	165,551	166,167	331,718
TCRDS-SR	135,084	146,733	281,817
TCRDS-SS	219,438	199,513	418,951

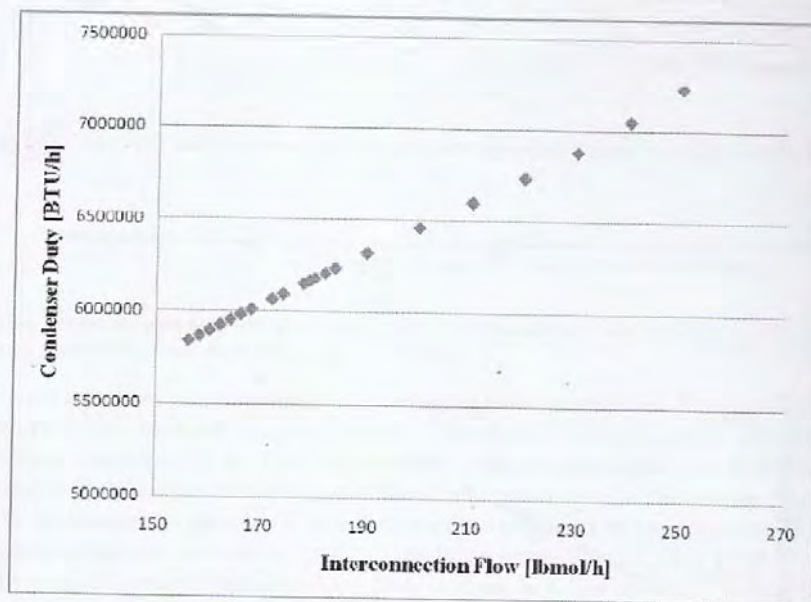


Figure 9. Condenser duty as function of the interconnection flow in the main column, case TCRDS-SS.

Finally, Table 7 summarizes the total annual costs of all systems tested. These costs were calculated using the method of Guthrie (Turton et al., 2004). As it can be seen, only the TCRDS-SR system has advantages in reductions of both energy consumption and equipment costs when compared to conventional and thermally coupled (without removal of reboilers) reactive distillation systems. Table 7 shows that a second option would be TCRDS-SR design which also presents significant savings in energy consumption. This is an interesting situation because the elimination of reboiler in a thermally coupled distillation column would seem that would lead to a design for minimum energy consumption and total annual cost. However according to the results of this work, the presence of a rectifier or side stripper influences in the energy consumption and equipment costs even when carrying out the decrease of the reboilers in the design.

CONCLUSIONS

This study uses computer aided engineering tools such as Aspen Plus for the development of an alternative biodiesel process based on reactive distillation. Two thermally coupled distillation sequences with side columns (stripper or rectifier) and minimum number of reboilers were utilized to carry out the production of the fatty ester compound. The parametric studies showed that the alternative coupled reactive arrangements column requires a lower amount of energy in contrast to the conventional and traditional coupled systems. However, only the TCRDS-SR presents advantages in the total annual cost in comparison with all sequences studied. This process improves the biodiesel production and dramatically reduces the total energy consumption of the process. The major benefit of this alternative process is low operating costs due to the integrated design with no reboilers in comparison with conventional and thermally coupled reactive distillation sequences. According to these results, it can be concluded that thermally coupled distillation sequences with minimum number of reboilers and a side rectifier option presents significant advantages over the classical process used to produce biodiesel.

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