Recycling vs. reprocessing. Optimization of a gossypol production process

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ABSTRACT

Three processes to obtain GAA from the soapstock residue of oil refineries were compared in this work. The first process is the originally proposed by Dowd and Pelitire (2001. Ind. Crops Prod. 14, 113) in which the mother liquor from the last crystallization step still contains an appreciable amount of gossypol which is lost as a process residual stream. The second process recycles the mother liquor to the hydrolysis first step of the process, following the heuristic of the traditional process design procedure by Douglas (1988. Conceptual Design of Chemical Processes. McGraw Hill, New York, NY), which increases product yield. While the third process adds a new downstream processing line to reprocess the mother liquor. This last alternative renders a slightly lower product yield than Process 2 but requires a smaller investment cost, exhibiting the best economic performance.

The alternative of incorporating a recycle to reprocess unreacted material (in this case the bound gossypol present in the crystallization mother liquor) is the usual approach in traditional process design. However, in this particular study case, it does not succeed in rendering the process alternative with the best economic performance: the recycle stream flow impacts on the equipment sizes, increasing investment cost far beyond the alternative that adds smaller units to reprocess the mother liquor stream.

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1. Introduction

Gossypol, (2,2′-bis(1,6,7-trihydroxy-3-5-isopropyl-8-aldehydonaphthalene) is a yellow pigment found in cotton (Gossypium) species. It constitutes 20–40% in weight of the gland responsible for pigmentation and accounts for 0.4–1.7% of the whole kernel (Dowd and Pelitire, 2001; Xi et al., 2009). Gossypol was first discovered as a crude pigment from cottonseed oil foot, a mixture of precipitated soaps and gums produced in the refining of crude cottonseed oil with sodium hydroxide.

Gossypol exists in a number of tautomeric and isomeric forms, depending on the position of methyl and hydroxyl groups in the carbon atoms adjacent to the binaphthalene. During cottonseed oil processing, gossypol and phosphatides are extracted together with triglycerides during solvent extraction. In the oil refining process, caustics are added so that free fatty acids, phosphatides and gossypol are separated from oil as a soapstock by-product (Kuk and Bland, 2002). Some evidence indicates that most gossypol exists as Schiff bases through the condensation between the aldehydic groups of gossypol and amino groups of proteins (Xi et al., 2009).

Kuk and Bland (2002)) studied the pH influence on the Schiff base resulting from the reaction of gossypol with phosphatidiletanolamine PE, which is the largest phospholipid present in oilseed soapstock. Their work found a strong dependence on pH in the formation of the Schiff base. This reaction is readily reversible by acidic hydrolysis in aqueous solution and can be chemically stabilized by reduction.

Gossypol has important biological properties in human health and toxic properties as animal feed. There are numerous studies about its biological activity such as antifertility

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(Coutinho, 2002) and anticancer activities, as well as inhibiting the growth of a variety of cell lines including breast, colon, prostate, and leukemia cells (Xi et al., 2009). Also, gossypol has antivirus activity that inhibits the replication of human immunodeficiency virus type 1 (Lin et al., 1989, 1993; An et al., 2012). Due to these biological properties, gossypol is a compound of interest for the pharmaceutical industry. Otherwise, this compound contributes to the toxicity of soapstock used as animals feed: gossypol would need to be removed from soapstock before adding it to poultry feed (Dumont and Narine, 2007).

The production of cottonseed oil in Argentina was 12,500 ton in year 2012 (CIARA, 2014), corresponding to a soapstock production of about 750 t/yr. Soapstock contains between 0.33 and 13% of gossypol (Dowd, 1996), considering an average of 7% its potential production rate is 50 t/yr. Thus, the locally available raw material would allow industrial scale production of gossypol.

There is abundant literature about isolating gossypol from this oil refining by-product. The basic processes consist of acid treatment to hydrolyze covalently bound gossypol,

partitioning it into an organic phase, concentrating gossypol by vacuum evaporation of the organic solvent and finally crystallizing it as gossypol acetic acid (GAA), e.g. Dowd and Pelitire (2001), Pons et al. (1959), Jia et al. (2009).

The objective of this study was to optimize the structure of the process proposed by Dowd and Pelitire (2001). First, the hierarchical design concepts proposed by Douglas (1988) were applied, implementing the recycle of the liquid outlet stream from the crystallizer (the mother liquor) to the hydrolysis step of the process. This stream still contains about a 30% of the gossypol entering with the raw material. By recycling this stream the process product yield is increased, but also the size of the units involved in the recycle loop increase.

Afterwards we take a different approach by proposing to add a downstream reprocessing train, as an alternative to recycling. This alternative was analyzed and optimized. Finally, both processes were compared (mother liquor recycling and reprocessing) in order to select the GAA production process with the maximal economic performance.

2.   Process modeling and optimization

We start considering the process proposed by Dowd and Pelitire (2001), who experimentally optimized its operating variables: reaction times, concentration of reactants, etc. as shown in Table 1, aimed at improving yield, subject to not jeopardize product purity.

The process consists of an initial hydrolysis with phosphoric acid, where the released gossypol is dissolved into the organic phase of methyl ethyl ketone (MEK). Then, the residual aqueous phase is washed with MEK to recover dissolved product. The stream containing gossypol is then concentrated by vacuum evaporation of MEK, and finally converted into GAA with the addition of glacial acetic acid and allowing time for crystallization.

The process is presented in Fig. 1, stream F1 is the soapstock feed: we adopted a base case flow rate of 100 kg/h with an average composition of 50% water, 38% free fatty acids FFA, 8.3% oil and 3.7% total gossypol (Dowd, 1996), while stream F2 is a 1.4 M solution of phosphoric acid in MEK. They are mixed in a hydrolysis reactor (H1) and afterwards the light phase is separated from the heavy one in the centrifuge (CE1). Stream F4 is the light phase containing most of the gossypol and stream F5 is the (aqueous) heavy phase, which is subjected to consecutive washes with MEK to recover the gossypol contained therein. E1 is an evaporator that concentrates the light phase producing stream F19 of recovered MEK. Stream F21 is glacial acetic acid and C1 is the crystallizer. Stream F24 is the process product: crystallized GAA (a crystalline inclusion complex containing an equimolar ratio of both) and F23 is the residual mother liquor.

The mathematical model consists of mass and heat balances, equipment sizing and cost estimations implemented with algebraic equations. Thus, it is a NLP (Non Linear Program) model that was implemented in GAMS (General Algebraic Modeling System). Product yields, flow rate ratios, concentrations, etc. were taken from Dowd and Pelitire (2001). A brief outline of the model follows.

The total and component mass balances are represented by Eqs. (1) and (2). Eqs. (3) and (4) describe the hydrolysis reaction, where μ is a conversion factor given as a mass fraction of the gossypol present in the raw material, taken from the literature (Dowd and Pelitire, 2001; Pons et al., 1959).

Table 1 – Process variables optimized by Dowd and Pelitire (2001).

Process stage	Process variables optimized	Optimal values found
Hydrolysis	Phosphoric acid concentration in MEK	100 ml of 1.4 M solution per each 100 g soapstock feed
Hydrolysis	Hydrolysis time	2 h
Washing	Number of MEK washing steps of the aqueous phase	3 washings with 50 ml MEK per each 100 g soapstock feed
Crystallization	MEK left in the concentrated solution	50 ml concentrated solution per each 100 g soapstock feed
Crystallization	Amount of glacial acetic acid added	1:2 ratio in volumes of acetic acid: concentrate solution
Crystallization	Crystallization time	9 h
Crystallization	Crystallization temperature	Room temperature of 4 °C

Eqs. (5)–(7) describe the centrifugal separation downstream of the hydrolysis: Eq. (5) distributes hydrolyzed gossypol between both phases and Eq. (7) computes the MEK lost in the aqueous phase due to its solubility of 250 g/L of water (Perry and Chilton, 1973). The oily phase is concentrated evaporating a fraction β of the incoming MEK (Eq. (8)). The purity of the product GAA was taken to be the same as reported by Dowd and Pelitire (2001).

$$\sum_f Q_{(f)} \times W_{(i,f)} = \sum_f Q_{(f)} \times W_{(i,f)} \quad \forall \ (eq,i) \tag{1}$$

$$\sum_i W_{(i,f)} = 1 \quad \forall (f) \tag{2}$$

$$Q_{(f3)} \times W_{(gossy,f3)} \times N\mu = Q_{(f4)} \times W_{(gossynl,f4)} + Q_{(f5)} \times W_{(gossynl,f5)} \tag{3}$$

$$Q_{(f3)} \times W_{(gossy,f3)} \times (1 - N\mu) = Q_{(f4)} \times W_{(gossy,f4)} \tag{4}$$

$$\sum_{f(out_{(eq,f)})} W_{(i,f)} = (1.092 - 0.744 \times \sum_{f(in_{(eq,f)})} W_{(i,f)})$$
$$\times \sum_{f(outheavy_{(eq,f)})} W_{(i,f)} \quad \forall (F\ Goss, centrifuge) \tag{5}$$

$$\sum_{f\left(outlight_{(eq,f)}\right)} W_{(H_2O,f4)} \leq 0 \quad \forall (centrifuge) \tag{6}$$

$$\sum_{f(outheavy_{(eq,f)})} W_{(MEK,f)} = sol_{MEK} \tag{7}$$

$$Q_{(f19)} = Q_{(f18)} \times W_{(mek,f18)} \times \beta \tag{8}$$

After computing the mass balances, we proceed to size and cost the equipment. With the reactor inlet flow rate and residence time we compute the reaction volume and then the diameter and height of the vessel (Eqs. (9) and (10)). The investment cost is estimated following Douglas (1988) updated with the Marshall & Swift (M&S) index (Eq. (11)). For each centrifuge, its power requirement is estimated as a function of the
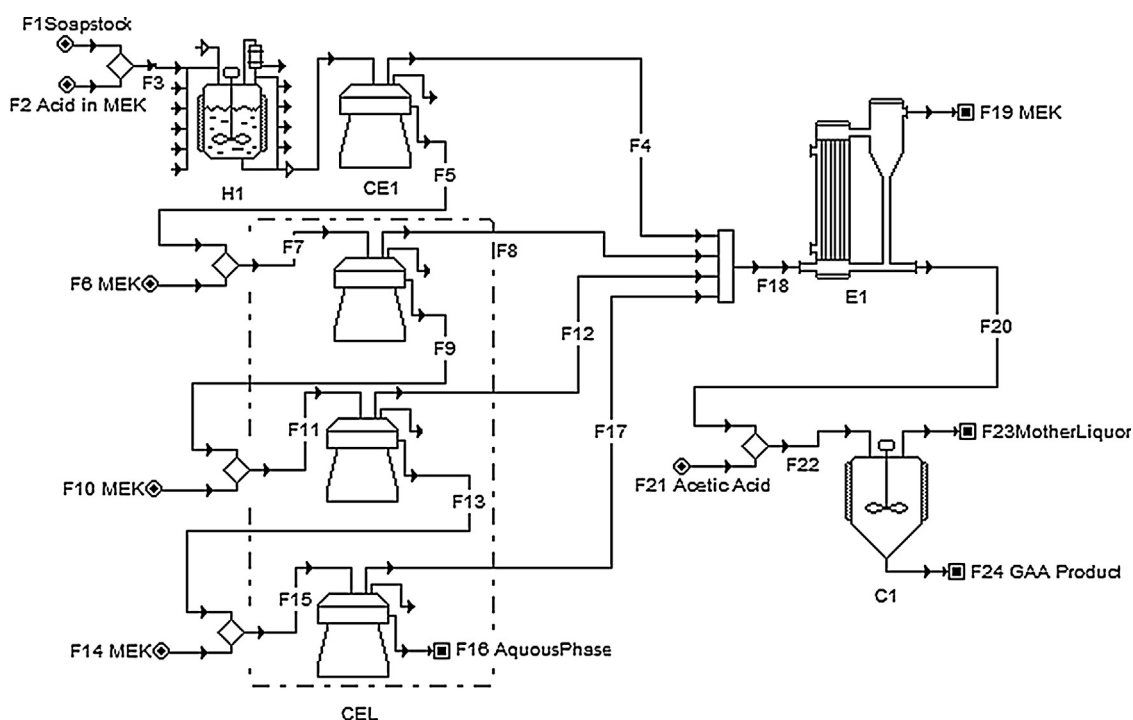

Fig. 1 – Process 1 published by Dowd and Pelitire (2001).

inlet flow rate in gal/min (Eq. (12)), and then the investment cost through Eq. (13). The areas needed by the evaporator and condenser are computed in Eq. (14) as a function of the vaporization heat of MEK with an overall heat transfer coefficient U of 1220 kcal/h m^2 °C for the evaporator and 732 kcal/h m^2 °C (Douglas, 1988) for the condenser, then these areas are used to compute the cost of equipment through Eq. (15). The investment cost of the crystallizer was obtained considering it as a reactor.

$$\frac{\left(\pi \times D^2\right)}{4 \times H} = 1.25 \times \text{Vol} \tag{9}$$

$$H = 3 \times D \tag{10}$$

$$\text{CI} = \text{M\&S} \times \left(101.9 \times D^{1.066} \times H^{0.82} \times 3.1\right) \tag{11}$$

$$P = \left(0.025 \times (\text{feed})^2\right)^{0.986} \tag{12}$$

$$\text{CI} = \text{M\&S} \times \left(5200 \times P^{0.68}\right) \tag{13}$$

$$A = \frac{\phi}{U} \times \Delta T \tag{14}$$

$$\text{CI} = \text{M\&S} \times \left(101.3 \times A^{0.65} \times 3.79\right) \tag{15}$$

2.1. Conceptual design of the process

The process information reported by Dowd and Pelitire (2001) may be viewed as the result of the laboratory research stage for developing a process, that provides the basic knowledge required to start a process design, e.g. with the hierarchical procedure proposed by Douglas (1988).

The first step in Douglas procedure is "Input Information and Batch vs. Continuous". He describes this input information as "what raw materials are fed to the process and what products and by-products are formed". Dowd and Pelitire (2001) defined a feasible product formation reaction route for producing a good quality product, optimizing the reaction operating conditions of the hydrolysis and the crystallization. With respect to the decision batch vs. continuous, we select the batch alternative for the reaction stages hydrolysis and crystallization because this is how Dowd and Pelitire (2001) did the experiments they report, while generating continuous alternatives for these stages (if worthwhile), would require considerable extra research and development effort. For these process reactors we take their size to be the inlet volumetric flow rates times the batch reaction time. This corresponds to a batch—semi continuous plant design with enough intermediate storage tanks capacity, whose cost will be neglected with respect to the cost of the process units.

The second step in Douglas procedure is "Input–Output Structure of the Process" which will furthermore define the recycle structure of the flow sheet and the structure of the separation system. Actually Dowd and Pelitire (2001) already started this step of the procedure, when deciding to separate the aqueous residual stream by centrifugation, recover the gossypol lost in it by liquid–liquid extraction with MEK, and recycling it to the concentration stage of the process. Starting at the point where they left, we will complete the design.

The conceptual process design procedure asks to classify the components present in the process and assign them a destination code: primary product, valuable by product, recycle (reactants), recycle and purge (reactants plus inert that

would accumulate in the recycle), fuel and waste. In our case GAA is the primary product, and bound gossypol the reactant. The most abundant contaminants (non product components) present in the feed to the process are water, FFA and triglycerides (Dowd, 1996) presents a detailed list of compounds found in FFA and triglycerides contained in soapstock, but as we are not interested in separating among themselves, we grouped them here as pseudo components). And we also consider MEK added in the hydrolysis as a reactant: even if it does not participate in the reaction, it is an appropriate reaction solvent and permits breaking the emulsion.

Water shall be considered a waste as we don't need extra water in the process and has not economic value: we decide it should be removed and send to treatment, as already done by Dowd and Pelitire (2001). FFA may have a commercial value (depending on its quality) or, in any case, be used as boiler fuel. So, we assign it the destination code valuable by product. The same occurs with triglycerides, they have a commercial value if returned to the oil refining process, or can be used as boiler fuel. MEK can be considered a reactant, and as its separation by evaporation produces a pure product (without contaminants) should be recovered from any exit stream and recycled to the hydrolysis stage. Finally, bound gossypol still present in the mother liquor of the crystallizer is a reactant and it may be worthwhile to recycle it to the hydrolysis stage, but in this case implementing a recycle purge, because the valuable product is dissolved in an oily solution that is a mixture of other components.

Steps 3rd and 4th in Douglas procedure are "Recycle Structure of the Process" and "Structure of the Separation System". In the original procedure by Douglas (1988), he works on a process (hydro dealquilation of toluene to benzene HDA) whose reactor exit stream is a mixture of gaseous and liquid components (if cooled down to temperatures nearer atmospheric). Thus, the stream exiting the reactor is split into a gaseous stream (that contains the reactant to be recycled), and a liquid separation "process box". In his case, as gaseous recycles need recompression (a pretty expensive operation), it is true that deciding the "Recycle Structure of the Process" is a higher level of decision than deciding the "Structure of the Separation System" (has a much higher economic impact on the design). So, the recycle decisions are made first, leaving the liquid separations as a "process box" to be approached later. This is not our case, where all the components to be separated belong to the category "liquids", and the recycle has not a higher economic impact than the liquid separation decisions. As a matter of fact, the present paper is focused on reporting that recycling is not the best economic alternative. Furthermore, our process consists of two consecutive reactions instead of one, so it is not possible to draw a unique separation "process box". Thus, we modified the original procedure by making the decisions about both steps at the same hierarchical level.

The original process of Dowd and Pelitire (2001) has a low yield: the mother liquor exiting the process still contains about a 30–40% of the gossypol present in the soapstock feed. There are two reasons to explain this. One is that the hydrolysis is not complete: Jia et al. (2009) report that part of the gossypol is still bound when the yield vs. hydrolysis time reaches a peak. At larger processing times the active aldehydic groups in the free gossypol turn into an inactive substance, i.e. at the optimal hydrolysis time there is still bound gossypol in the reactor, while the already released gossypol is deactivating. The other reason is the not negligible solubility of gossypol in the concentrated mother liquor: while the solubility of

Table 2 – Physical properties of the components to be separated.

Component	Boiling point	Vapor pressure at 80 °C	Molecular weight	Density	Solubility in water
MEK	79.6 °C	760 mmHg	72.1	0.805	350 g/1000 g
Water	100.0 °C	350 mmHg	18.0	1.0	
FFA	285.0 °C	«1 mmHg	282.0	0.855	Insoluble
Gossypol			518.0	~0.8	Insoluble
Oil			862.0	~0.8	Insoluble
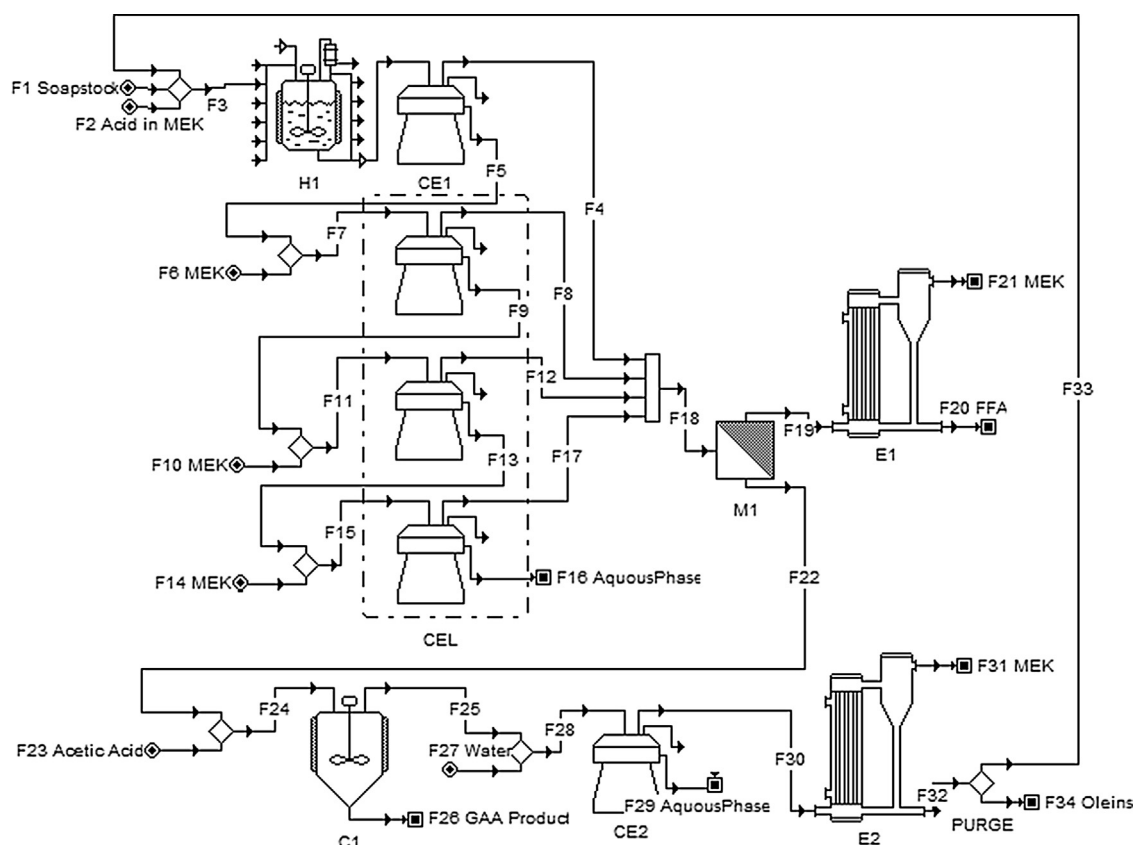
Fig. 2 – Process 2 that implements a recycle of the mother liquor.

GAA in pure MEK is very small, the presence of triglycerides and other impurities jeopardize its precipitation (Dowd and Pelitire, 2001; Jia et al., 2009). Thus, we will follow the Douglas heuristic that would recommend recycling the residual stream of the crystallization stage as it contains unreacted raw material (bound gossypol).

In addition, we start considering the other separations identified in the previous input-output level of decision, by considering the physical properties of the components involved, as presented in Table 2.

The components were ordered according to their boiling points, as in the original Douglas procedure (which implicitly assumes that their separation will be based on that property). Table spaces for components with a boiling point higher than that of FFA were not filled, we exclude distillation for separating these components: even if it would be technically feasible with molecular distillation (at very high vacuum) it is a pretty expensive operation, besides we suspect that high temperatures may jeopardize the biological activity of our product.

By inspection of the vapor pressures at 80 °C it Table 2 (relative volatility can be approximated by the ratio of vapor pressures) we notice that the separation of MEK from FFA or heavier components is very easily accomplished by distillation. A single stage flash, or an evaporator as adopted by Dowd

and Pelitire (2001), suffices for obtaining pure MEK at a harmless temperature. So we endorse the adoption of this operation by these authors, to adjust MEK concentration of the stream entering to the crystallization reactor. Furthermore, we also adopt it to recover the MEK remaining in the mother liquor stream that exits the crystallizer. About the MEK remaining in the residual aqueous stream (notice the solubility of MEK in water in Table 2), we compute the relative volatility of MEK/water be circa 2.2 enough away from the figure of 1.1 recommended by Douglas to adopt distillation. So, in case

Table 3 – Energy costs in Processes 1 and 2.

Operating cost $/(kg/h soapstock)	Process 1	Process 2
Electric energy		
Hydrolysis reactor (H1)	0.8^{-2}	1.1^{-3}
Centrifuge (CE1)	0.1^{-4}	2.088^{-5}
Centrifuge washing (CEL)	0.9^{-5}	4.48^{-6}
Crystallization reactor (C1)	0.4^{-2}	0.5^{-3}
Centrifuge (CE2)	–	7.16^{-4}
Thermal energy		
Evaporator (E1)	0.001	8.166^{-4}
Evaporator (E2)	–	4.083^{-4}

Table 4 – Material balances for Process 1 compositions in mass fraction.									
	Flow rate (kg/h)	Water	BGoss	OIL	FFA	MEK	FGoss	ACH	GAA
F1	100.00	0.50	0.04	0.08	0.38	0.00	0.00	0.00	0.00
F2	100.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F3	200.00	0.25	0.02	0.04	0.19	0.50	0.00	0.00	0.00
F4	136.14	0.00	0.01	0.06	0.28	0.64	0.01	0.00	0.00
F5	63.86	0.78	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F6	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F7	104.11	0.48	0.00	0.00	0.00	0.51	0.01	0.00	0.00
F8	40.54	0.00	0.00	0.00	0.00	0.99	0.01	0.00	0.00
F9	63.57	0.79	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F10	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F11	103.82	0.48	0.00	0.00	0.00	0.51	0.00	0.00	0.00
F12	40.42	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F13	63.40	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F14	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F15	103.65	0.48	0.00	0.00	0.00	0.52	0.00	0.00	0.00
F16	63.30	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F17	40.35	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F18	257.45	0.00	0.01	0.03	0.15	0.81	0.01	0.00	0.00
F19	166.08	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F20	91.37	0.00	0.01	0.09	0.42	0.45	0.02	0.00	0.00
F21	20.76	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00
F22	112.13	0.00	0.01	0.07	0.34	0.37	0.02	0.19	0.00
F23	109.34	0.00	0.01	0.08	0.35	0.38	0.00	0.19	0.00
F24	2.79	0.00	0.00	0.00	0.07	0.07	0.00	0.00	0.87

BGoss: bound gossypol, FGoss: free gossypol, ACH: glacial acetic acid GAA: gossypol acetic acid.

Table 5 – Material balances for Process 2 compositions in mass fraction.									
	Flow rate (kg/h)	Water	BGoss	OIL	FFA	MEK	FGoss	ACH	GAA
F1	100.00	0.50	0.04	0.08	0.38	0.00	0.00	0.00	0.00
F2	165.36	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F3	330.72	0.15	0.02	0.19	0.14	0.50	0.00	0.00	0.00
F4	266.95	0.00	0.01	0.24	0.17	0.57	0.01	0.00	0.00
F5	63.77	0.78	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F6	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F7	104.02	0.48	0.00	0.00	0.00	0.51	0.01	0.00	0.00
F8	40.51	0.00	0.00	0.00	0.00	0.99	0.01	0.00	0.00
F9	63.52	0.79	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F10	40.25	0.00	BGoss	0.00	0.00	1.00	FGoss	ACH	0.00
F11	103.77	0.48	0.00	0.00	0.00	0.51	0.00	0.00	0.00
F12	40.40	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F13	63.37	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F14	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F15	103.62	0.48	0.00	0.00	0.00	0.52	0.00	0.00	0.00
F16	63.28	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F17	40.34	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F18	388.20	0.00	0.01	0.16	0.12	0.70	0.01	0.00	0.00
F19	255.17	0.00	0.00	0.00	0.14	0.86	0.00	0.00	0.00
F20	36.80	0.00	0.00	0.00	1.00	0.00	0.00	0.00	0.00
F21	218.37	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F22	133.02	0.00	0.02	0.48	0.07	0.41	0.02	0.00	0.00
F23	27.30	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00
F24	160.32	0.00	0.01	0.40	0.06	0.34	0.02	0.17	0.00
F25	156.07	0.00	0.01	0.41	0.06	0.35	0.00	0.17	0.00
F26	4.24	0.00	0.00	0.00	0.00	0.13	0.00	0.00	0.87
F27	46.82	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
F28	202.90	0.23	0.01	0.31	0.05	0.27	0.00	0.13	0.00
F29	86.05	0.54	0.00	0.00	0.00	0.14	0.00	0.31	0.00
F30	116.85	0.00	0.02	0.55	0.08	0.36	0.00	0.00	0.00
F31	41.73	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F32	75.12	0.00	0.03	0.85	0.12	0.00	0.00	0.00	0.00
F33	65.36	0.00	0.03	0.85	0.12	0.00	0.00	0.00	0.00
F34	9.76	0.00	0.03	0.85	0.12	0.00	0.00	0.00	0.00

BGoss: bound gossypol, FGoss: free gossypol, ACH: glacial acetic acid, GAA: gossypol acetic acid.

Table 6 – Equipment sizes in each process.

	Process 1		Process 2	
	H(m)	D (m)	H (m)	D (m)
H1	1.9	0.6	2.2	0.7
C1	2.5	0.9	2.8	0.9
	Power (HP)		Power (HP)	
CE1	0.030		0.065	
CEL	0.005		0.005	
CE2	–		0.024	
	Area (m^2)		Area (m^2)	
E1	0.7		1.0	
M1	–		6.4	
E2	–		0.5	

Table 8 – Annual production of gossypol and by-products.

Product (kg/yr)	Process 1	Process 2
GAA	22,097	33,581
FFA	–	291,456
OLEINS	–	77,299

Table 9 – Economic assessment of Process 1 and 2.

Process	NPV ($)
Process 1	−4191,000
Process 2	712,220

of deciding to recover MEK from this stream too, distillation would be the first choice when selecting an operation.

Discarding separations based on difference in boiling points for the heavier components, we resort to the procedure proposed by Steffens et al. (2000) for the synthesis of processes in biotechnology. The approach is similar to the one of Douglas, but they consider other operations: discarding beforehand the chromatographic techniques because they are too expensive, we explore membrane ultra filtration: they recommend its adoption if the ratio of molecular weights of components to be separated exceeds 3. By inspection of Table 2 we notice that the ratio for Oil/FFA accomplishes this requirement: we could use MEK (with a much smaller molecular weight) as a carrier to separate FFA from Oil. There exists literature (reporting the commercial membrane tested, permeabilities, etc.) about this operation used in deacidification of edible oils (Kumar and Bhowmick, 1996; Bhosle and Subramanian, 2005), but with other solvents (methanol, ethanol, isopropyl alcohol, hexane) but not MEK. Another concern that would need to be experimentally checked is retention of gossypol in the oily retentate, because the ratio of molecular weights gossypol/FFA is circa 1.85, smaller than the figure of 3 recommended by Steffens et al. (2000).

The last columns in Table 2 display density and solubility in water, endorsing Dowd and Pelitire (2001) choice of centrifugation to separate the aqueous phase.

The fifth step in Douglas procedure is "Heat exchangers network". In our case the only streams liable to be heat integrated are the ones feeding and exiting the MEK evaporators (the only process sites at a temperature different from room temperature). The heat integration would consist in contacting these streams in a countercurrent arrangement. This was not

done, considering that it is expected to have a small economic impact in the process, in order to preserve clarity and brevity of exposition.

2.2. *Process 2 that recycles the mother liquor*

Implementing the recycle and separation decisions above, leads to the process flow sheet in Fig. 2 where the recycle stream is F33. As mandatory when implementing a recycle, a purge was incorporated to allow an escape route for those components not separated in the process stages, avoiding their indefinite accumulation in the recycle loop. The purge (stream F34 in Fig. 2), is a proportion of the recycle stream (the figure for this proportion is to be found by numerical optimization).

While the original process already has a stage that removes water (that will now remove it from the recycle), other components would accumulate if they can exit the process only through the purge stream. These components not only increase the flow rate to be processed by the stages involved in the recycle loop, but also jeopardize the crystallization yield by increasing the solubility of GAA in the mother liquor.

Thus, a membrane process stage was added in Process 2 to separate FFA as an olein secondary product (stream F20 in Fig. 2), which could be sold to other industries for the production of soaps, cosmetics, animal feed, etc. or as a boiler fuel (depending on its quality). While the original process evaporated MEK from the bulk mixture, we can use MEK to carry FFA through the pores of a membrane that rejects components of a molecular weight larger than FFA, prior to evaporating the MEK. This stage allows an escape route for FFA from the recycle, is a good alternative for increasing sales revenue from a secondary product, does not affect MEK recovery, improves subsequent crystallization and reduces the flow rate of the purge stream. With respect to triglycerides, we did not find

Table 7 – Investment costs for Processes 1 and 2 in US $.

Equipment	Procces 1	Procces 2
Reactor 1	11,764	15,600
Centrifuge 1	2243	4445
Wash centrifuge	2766	2397
Evaporator 1	6660	7776
Crystallization 1	21,050	25,077
Centrifuge 2	–	2286
Evaporator 2	–	2700
Membrane	–	10,729
Condenser 1	2846	2028
Condenser 2	–	700
TCI	47,329	73,738

Table 10 – Energy costs for Processes 1$^+$ and 3.

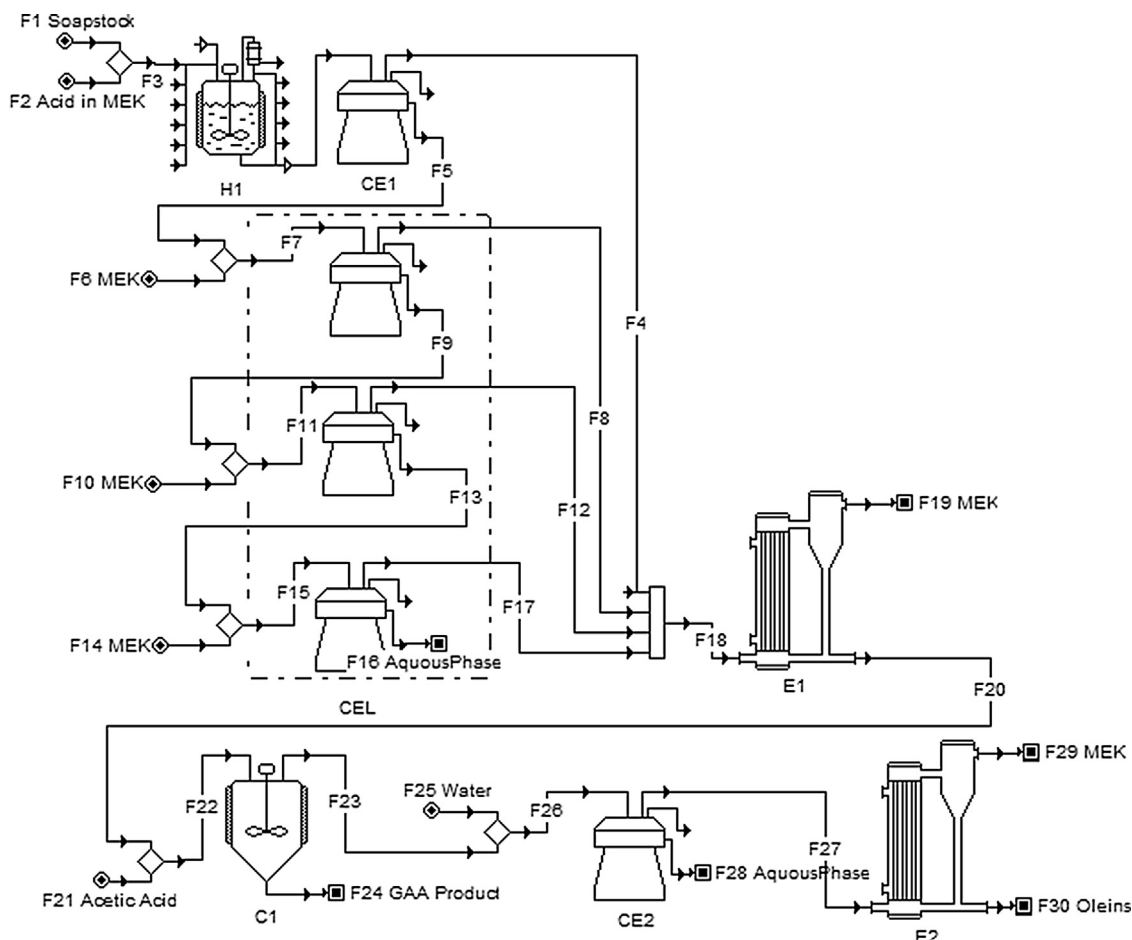| Operating cost $/(kg/h soapstock)	Process 1$^+$	Process 3
Electric energy for each equipment		
Reactor (H1)	0.8^{-2}	0.8^{-2}
Reactor (H2)	–	0.4^{-2}
Centrifuge (CE1)	0.1^{-4}	0.1^{-4}
Centrifuge washing (CEL)	0.9^{-5}	0.9^{-5}
Centrifuge (CE2)	1.3^{-3}	–
Tank crystallization (C1)	0.4^{-2}	0.4^{-2}
Tank crystallization (C2)	–	0.2^{-2}
Thermal energy		
Evaporator (E1)	0.1^{-2}	0.1^{-2}
Evaporator (E2)	1.35^{-3}	1.7^{-4}
Evaporator (E3)	–	3.5^{-4}

Fig. 3 – Process 1⁺ that improves Process 1 recovering by products.

an a priori satisfactory unit operation to separate them from the recycle loop, without losing product. Both molecular distillation and supercritical extraction were considered because they have been used to separate the triglycerides fraction from mixtures containing lower molecular weight components of oil as e.g. tocopherols. However, we lack data neither about solubility of gossypol in supercritical CO_2 nor about eventual thermal damage of gossypol at the relatively high temperatures (above 200 °C) required by molecular distillation. Thus, also considering that both operations are pretty expensive, and that the amount of oil in the feed is not that large, we allowed its accumulation in the recycle.

As the mother liquor contains acetic acid added in excess to precipitate the product, before recycling it to the hydrolysis we added a washing step to separate it: we mix the mother liquor stream F25 with a water stream F27 and afterwards place a centrifuge to separate the aqueous solution of acetic acid. Furthermore, we send the washed stream F30 to evaporator E2 to recover MEK. The stream that leaves the process is the purge, which can be given the destination "boiler fuel" which is economically more convenient than "waste".

The NLP model presented in the previous section needs additional equations to consider the mass balances at the membrane, the recycle and purge, the washing step prior to recycling and the evaporator in the crystallization exit stream. The mass balance for the membrane (Eq. (16)) considers a rejection coefficient for each component i $\sigma_{(i)}$. The area required by the membrane is computed from the flow rate of permeate and the permeability of the membrane J_{mem} here estimated to be 50 L/m² h (Eq. (17)). Then, this area allows

computing the cost of investment through linear Eq. (18), with a membrane price of 550 $/m².

$$\frac{W_{(i,f19)}}{W_{(i,f18)}} = \left(\frac{Q_{(f18)}}{Q_{(f19)}}\right)^{\sigma_{(i)}} \quad \forall i \tag{16}$$

$$A = \frac{\text{Vol}_{\text{permeate}}}{J_{\text{mem}}} \tag{17}$$

$$CI = A \times \text{Price}_{\text{mem}} \tag{18}$$

The purge is generated with a flow splitter represented by Eqs. (19) and (20) that adds a degree of freedom: the variable X which is the fraction of the recycle incoming flow that is purged. This variable is subjected to optimization; it was here bounded between a value close to 0.0 and 0.4 without these bounds becoming active.

$$Q_{(f32)} \times (1 - X) = Q_{(f33)} \tag{19}$$

$$Q_{(f32)} \times X = Q_{(f34)} \tag{20}$$

To compare the two processes on an economic performance basis, we added the linear equations that compute Net Present Value NPV for a time horizon of 10 years, considering the following product prices: GAA 38 $/kg (Organic-herb, 2014), by-products oleins and FFA 0.51 $/kg (both products were priced as boiler fuel), soapstock 0.06 $/kg (Hass, 2005), phosphoric acid 1.88 $/kg (Química Agroindustrial Neo S.A., 2014), acetic acid 1 $/kg (Química Agroindustrial Neo S.A.,

Table 11 – Material balances for Process 1[+] compositions in mass fraction.								
	Flow rate (kg/h)	Water	BGoss	OIL	FFA	MEK	FGoss	ACH	GAA
F1	100.00	0.50	0.04	0.08	0.38	0.00	0.00	0.00	0.00
F2	100.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F3	200.00	0.25	0.02	0.04	0.19	0.50	0.00	0.00	0.00
F4	136.06	0.00	0.01	0.06	0.28	0.64	0.01	0.00	0.00
F5	63.94	0.78	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F6	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F7	104.19	0.48	0.00	0.00	0.00	0.51	0.01	0.00	0.00
F8	40.57	0.00	0.00	0.00	0.00	0.99	0.01	0.00	0.00
F9	63.61	0.79	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F10	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F11	103.86	0.48	0.00	0.00	0.00	0.51	0.00	0.00	0.00
F12	40.44	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F13	63.42	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F14	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F15	103.67	0.48	0.00	0.00	0.00	0.52	0.00	0.00	0.00
F16	63.31	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F17	40.36	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F18	257.44	0.00	0.00	0.03	0.15	0.81	0.01	0.00	0.00
F19	166.08	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F20	91.36	0.00	0.01	0.09	0.42	0.45	0.03	0.00	0.00
F21	20.76	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00
F22	112.12	0.00	0.01	0.07	0.34	0.37	0.02	0.19	0.00
F23	109.02	0.00	0.01	0.08	0.35	0.38	0.00	0.19	0.00
F24	3.10	0.00	0.00	0.00	0.07	0.00	0.00	0.07	0.87
F25	32.70	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
F26	141.72	0.23	0.01	0.06	0.27	0.29	0.00	0.14	0.00
F27	88.74	0.00	0.01	0.09	0.43	0.47	0.00	0.00	0.00
F28	52.98	0.62	0.00	0.00	0.00	0.00	0.00	0.38	0.00
F29	41.52	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F30	47.22	0.00	0.02	0.18	0.80	0.00	0.00	0.00	0.00

BGoss: bound gossypol, FGoss: free gossypol, ACH: glacial acetic acid AAG: gossypol acetic acid.

2014) and MEK 2.5 $/kg. In the case of MEK which is recovered, its consumption is computed as the losses in the streams exiting the process.

Utility costs were also taken into account, computing the electric and thermal energy consumption per kg/h of soap-stock feed. Electric energy was computed for the centrifuges and reactors agitation. Thermal energy consumption was calculated accounting steam usage with an addition of 3% for losses in pipes and 2% in the boiler. A price of 0.04 $/KWh energy (Douglas, 1988) and 0.005 $/kg steam (Douglas, 1988) was considered. Cooling water was considered to be implemented as a closed loop and its make up neglected. Energy costs of both processes are given in Table 3: the process that recycles the mother liquor slightly increases energy operating costs.

2.3. Economic assessment of Processes 1 and 2

Following, we report the material balances, equipment sizes, and the economic assessment of both processes. The material balances in Tables 4 and 5 display the stream flow rates and compositions for each analyzed process. Feed soap-stock/solvent ratio is 1:1 in both cases.

Consider the product streams: in Process 1 it is F24 with a flow rate of 2.79 kg/h and with a purity of 87% while in Process 2 it is F26 with 4.24 kg/h with a purity of 87%. Process 2 has two by-products: stream F20 composed of FFA and F34 mainly composed of oil, both of them priced as boiler fuels. In Process 2, the recycle stream F33 is highly concentrated in oil. As a result of the economic optimization of Process 2, the purge is a 13% of the bulk recycle stream. If a larger

price were assigned to the purge (e.g. as oil returning to the refinery instead of boiler fuel) the purge would be a larger fraction of the recycle, and the composition of oil would decrease. Anyway the oil content is large and may negatively affect the product yield in the crystallization stage. This effect could not be modeled due to lack of solubility data; however it was taken into account qualitatively in the assessment of alternative processes.

Table 6 displays the equipment sizes which determine their cost, and the total capital investment TCI for each process, presented in Table 7. This Table highlights the big difference in TCI of both processes: the implementation of recycle F33 greatly increases the size of the equipment involved in the recycle loop.

On the other hand, Table 8 displays the annual production of GAA and by-products of each Process: in this respect Process 2 largely improves the performance of the process by Dowd and Pelitire (2001) yielding a 50% larger production of gossypol, from the same raw material. Finally, Table 9 displays the Net Present Value of both processes.

The NPV is negative for Process 1. Actually, assessing the process by Dowd and Pelitire (2001) as presented by them, with NPV as the metrics, is unfair: adding the operations to convert their waste mother liquor into olein and recovering its content of MEK is inexpensive and would largely improve the cash flow and thus, it's NPV. This will be done next. First, we modified Process 1 by processing the mother liquor: we wash it from the excess acetic acid of crystallization by adding water followed by a centrifuge to separate the acid aqueous phase, and incorporate an evaporator to recover its content of MEK to be recycled to the hydrolysis. The by-product obtained is an acid
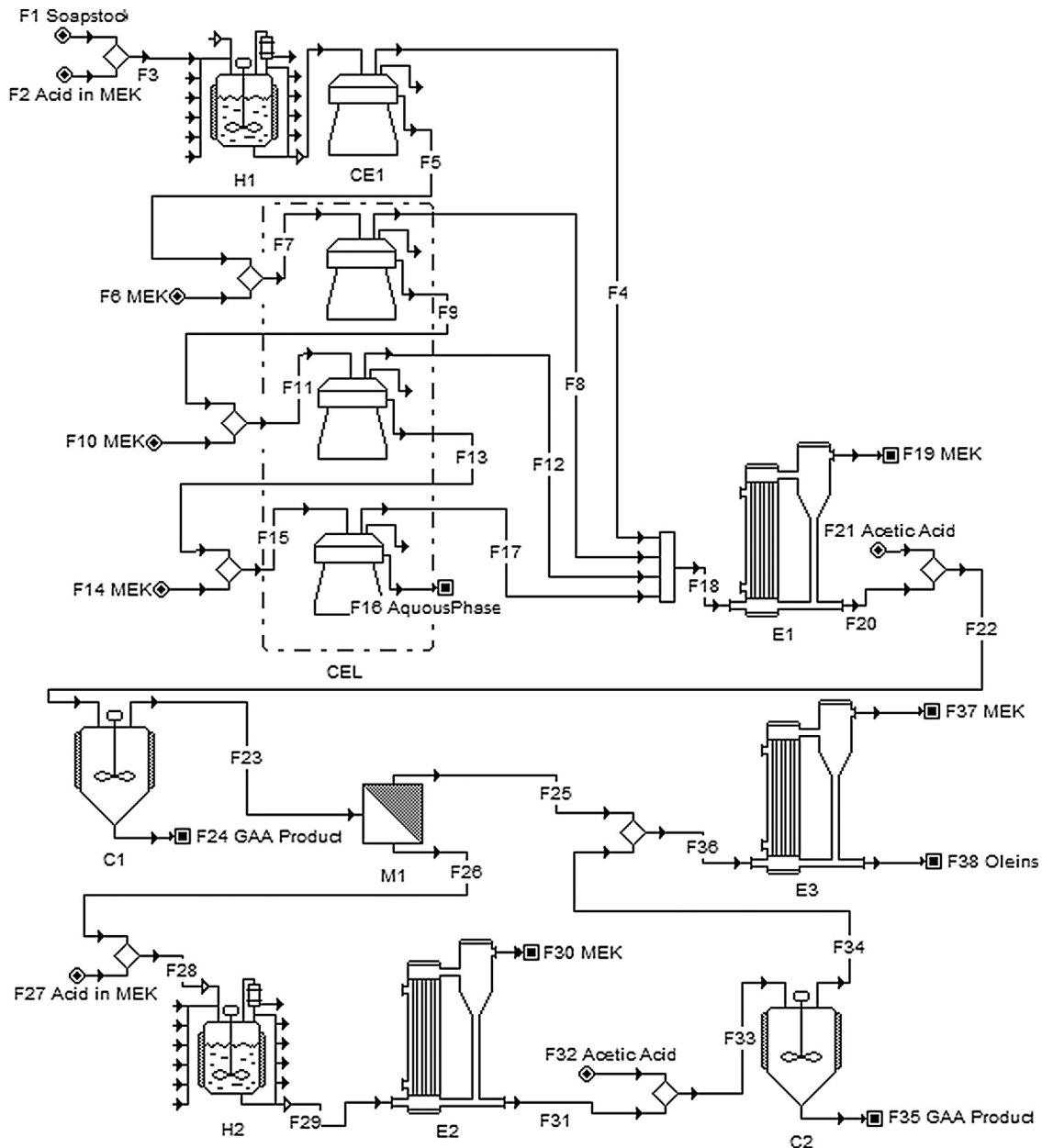
Fig. 4 – Process 3 that implements a reprocessing of the mother liquor.

free mixture of FFA and oil, which we priced as boiler-fuel. The modified process is presented in Fig. 3 as Process 1⁺.

2.4. Process 3 that reprocesses the mother liquor

But our main concern is still the problem of high oil concentration in the recycle stream in Process 2. Not finding a suitable and cost effective operation for the removal of oil, we propose a third alternative process, presented in Fig. 4 as Process 3, which adds a reprocessing line for recovering the gossypol still present in the crystallization mother liquor, instead of recycling this stream. This process duplicates all the stages, but avoids the accumulation of non product components which increase the size of all the stages involved in the recycle, and jeopardize crystallization. This process resembles the procedure that would be intuitively adopted if working batch wise in the laboratory.

This process also incorporates a membrane separation step: while in Process 2 the purpose was to allow an escape route for FFA from the recycle, now we implement it to increase the concentration of gossypol which is smaller than in the soapstock feed. Also, reducing non MEK components in the feed to the new process line will favor the second crystallization. These processes were also modeled in GAMS to obtain the mass balances, sizes of equipment, capital investment and finally NPV. Also the corresponding energy balances were performed to compute the operating costs, presented in Table 10 as $ per kg/h of soapstock feed accounting the electric and thermal energy consumptions.

Comparing also with the figures in Table 3 we notice that Process 1⁺ is only slightly more demanding than the original process by Dowd and Pelitire (2001) while processes 2 and 3 exhibit a small increase, in this order: Process 3 more than Process 2.

Table 11 presents the mass balances for Process 1⁺ and it can be seen that the olefin by-product now generated by the process (stream F30) contains mainly FFA.

Table 12 – Material balances for Process 3 compositions in mass fraction.

	Flow rate (kg/h)	Water	BGoss	OIL	FFA	MEK	FGoss	ACH	GAA
F1	100.00	0.50	0.04	0.08	0.38	0.00	0.00	0.00	0.00
F2	100.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F3	200.00	0.25	0.02	0.04	0.19	0.50	0.00	0.00	0.00
F4	136.06	0.00	0.01	0.06	0.28	0.64	0.01	0.00	0.00
F5	63.94	0.78	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F6	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F7	104.19	0.48	0.00	0.00	0.00	0.51	0.01	0.00	0.00
F8	40.57	0.00	0.00	0.00	0.00	0.99	0.01	0.00	0.00
F9	63.61	0.79	0.00	0.00	0.00	0.21	0.01	0.00	0.00
F10	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F11	103.86	0.48	0.00	0.00	0.00	0.51	0.00	0.00	0.00
F12	40.44	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F13	63.42	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F14	40.25	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F15	103.67	0.48	0.00	0.00	0.00	0.52	0.00	0.00	0.00
F16	63.31	0.79	0.00	0.00	0.00	0.21	0.00	0.00	0.00
F17	40.36	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F18	257.44	0.00	0.00	0.03	0.15	0.81	0.01	0.00	0.00
F19	166.08	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F20	91.36	0.00	0.01	0.09	0.42	0.45	0.03	0.00	0.00
F21	20.76	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00
F22	112.12	0.00	0.01	0.07	0.34	0.37	0.02	0.19	0.00
F23	109.12	0.00	0.01	0.07	0.35	0.38	0.00	0.19	0.00
F24	3.00	0.00	0.00	0.10	0.00	0.00	0.00	0.00	0.90
F25	80.00	0.00	0.00	0.00	0.38	0.42	0.00	0.20	0.00
F26	29.12	0.00	0.04	0.27	0.26	0.29	0.00	0.14	0.00
F27	25.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F28	54.12	0.00	0.02	0.15	0.14	0.62	0.00	0.08	0.00
F29	54.12	0.00	0.01	0.15	0.14	0.62	0.01	0.08	0.00
F30	26.64	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F31	27.48	0.00	0.01	0.29	0.28	0.24	0.03	0.15	0.00
F32	3.33	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00
F33	30.81	0.00	0.01	0.26	0.25	0.22	0.03	0.24	0.00
F34	29.81	0.00	0.01	0.27	0.25	0.22	0.00	0.24	0.00
F35	1.00	0.00	0.00	0.07	0.00	0.00	0.00	0.07	0.87
F36	109.81	0.00	0.00	0.07	0.35	0.36	0.00	0.22	0.00
F37	39.88	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00
F38	69.93	0.00	0.00	0.11	0.54	0.00	0.00	0.34	0.00

BGoss: bound gossypol, FGoss: free gossypol, ACH: glacial acetic acid GAA: gossypol acetic acid.

Table 12 displays the mass balances for process 3 represented in Fig. 4. The flow rate of product is the summation of F24 and F35 amounting to 4 kg/h of GAA. The stream of by-product is F38 which is used as a boiler fuel. Streams F19, F30 and F37 are recovered MEK that is recycled to the hydrolysis reactors.

Table 13 displays the size of equipment in each process, which determines the investment cost required by each process, shown in Table 14. Process 1+ requires a larger TCI than Process 1 as expected because it adds extra processing units, but Process 3 is less expensive in capital investment than

Table 13 – Size of equipment involved in each process.

	Process 1+		Process 3	
	H (m)	D (m)	H (m)	D (m)
H1	1.9	0.6	1.9	0.6
H2	–	–	1.0	0.4
C1	2.5	0.9	2.5	0.8
C2	–	–	1.6	0.5
	Power (HP)		Power (HP)	
CE1	0.030		0.024	
CEL	0.005		0.006	
CE2	0.012		–	
	Area (m²)		Area (m²)	
E1	0.7		0.7	
M1	–		2.0	
E2	1.0		0.3	
E3	–		0.2	

Table 14 – Total capital investment in Processes 1+ and 3.

Equipment	Procces 1+	Procces 3
Reactor 1	11,764	11,764
Centrifuge 1	2243	2243
Wash centrifuge	2766	2770
Evaporator 1	6660	6660
Crystallization 1	21,050	21,047
Centrifuge 2	1400	–
Evaporator 2	4431	3213
Membrane	–	3363
Reactor 2	2846	3429
Crystallization 2	–	9343
Evaporator 3	–	2635
Condenser 1	1737	1738
Condenser 2	450	529
Condenser 3	–	990
TCI	55,347	69,724

Table 15 – Annual production of gossypol and by-products.

Product (kg/year)	Process 1[+]	Process 3
GAA	24,552	31,680
FFA	–	–
OLEINS	538,560	554,400

Table 16 – Economic assessment of Processes 1[+] and 3.

Process	NPV ($)
Process 1[+]	900,420
Process 3	1649,200

Process 2: it requires a larger number of units, but smaller enough to compensate for the effect of economy of scale.

Table 15 shows the annual production of GAA and by-products, obtained from the material balances for each process. Process 1[+] has a lower production of GAA than Process 3 as expected because Process 3 implements an extra recovery line that increases production in about a 30% with respect to that of Process 1[+].

As regards FAA, Process 2 is the only that produces it, due to the incorporation of a membrane ultra-filtration to remove FFA from the process to avoid an excessive accumulation in the recycle. Comparing also with the figures in Table 8 we notice that both Processes 2 and 3 exhibit a much higher production of the valuable product GAA than the process proposed by Dowd and Pelitire (2001) (either the original Process 1 or the modified Process 1[+] that processes the waste stream), with Process 2 having a slightly larger production than Process 3.

Finally, Table 16 presents the economic evaluation of the alternative Processes 1[+] and 3. Comparing also with the figures in Table 9 for Processes 1 and 2, we notice that Process 3 has the better economic performance.

3.   Conclusions and further work

We thought it would be of interest to perform an economic evaluation of this process, as the only other we found in open literature was done by Kultun et al. (1959). They did a preliminary analysis of the investment cost for obtaining gossypol, in a work that differs with ours in the raw material used, the products obtained and process design. Furthermore, we thought that we could take advantage of process design tools not available at that time.

One intended use of this information (mass and energy balances, and economic assessment) is to be the input information for optimizing oil refineries, as in the paper by Martín and Grossmann (2013). They systematically evaluate a large number of alternative technologies by optimally integrating the use of raw materials and energy, forecasting that integration of processes is the future of biorefineries, exploiting synergies that reduce costs.

We took the process proposed by Dowd and Pelitire in 2001 as the base case because it processes soapstock, a readily available waste stream of oil refineries, with a simple process for which they experimentally optimize process variables and report abundant data. We expect that newer, more sophisticated approaches like the one reported by Jia et al. (2009) who use ultrasound to improve hydrolysis and crystallization yields, would in any case enhance the economic performance of this process.

Three alternatives of the process for obtaining GAA were compared in this work. Process 1 is the original one by Dowd and Pelitire (2001), in which the mother liquor from the crystallizer still contains an appreciable amount of gossypol which is lost in this residual stream. Process 1[+] is a modification of the former that processes the waste stream to produce an olefin byproduct, to increase its economic performance measured as NPV. Process 2 recycles the mother liquor to the hydrolysis step of the process, following the heuristic of the traditional process design procedure by Douglas (1988). While Process 3 adds a new processing line to reprocess the mother liquor.

The three processes 1+, 2 and 3 proved feasible, i.e. exhibit positive NPV. The new alternatives 2 and 3 have the aim of increasing product recovery, which they do, with a significant increase of NPV with respect to the original process. Even though Process 2 has the higher product yield, Process 3 is the one that exhibits the best economic performance. This outcome is in line with the concept highlighted by Luyben (2014) that there is an optimum recovery that depends on the difficulty of separation and the value of the product.

The alternative of incorporating a recycle to reprocess unreacted material (in this case the bound gossypol present in the crystallization mother liquor) is usually effective in traditional process design. However, in this case it does not succeed in pointing the process alternative with the best economic performance. In Process 2, the recycle stream flow impacts on the reactor and crystallizer sizes, increasing investment costs much beyond the alternative that adds smaller units to reprocess the mother liquor stream.

A by-product of this work is to point future experimental research needed to develop this process: data we needed much to model the process to allow optimization of operating conditions, but was scarce or not available. First, a more detailed tracking of bound and free gossypol in the hydrolysis: the experimental data reported by Dowd and Pelitire (2001) measure gossypol obtained at the end of the process, where it depends on the performance of other operations. Then, also better data about the solubility of gossypol as a function of oil, ACH, MEK and FFA content in the oily phase to describe the crystallization, and partition coefficients for gossypol, MEK and ACH to describe their distribution between the oily and aqueous phases in the washing steps. Another result of this work which we think deserves further work is to study the convenience of recycling or reprocessing, as depending on the difficulty of the separations.

A preliminary hypothesis about why the recycle heuristic failed in this case follows: The area of traditional chemical process design is biased towards the petrochemical industry processes, where the separation of components among themselves can generally be accomplished with cost effective operations, typically distillation and absorption. Actually, one of the steps in the process design procedure by Douglas (1988) is to order the components of each stream by increasing boiling points, implicitly assuming that separations will be based on their relative volatility.

In contrast, when processing natural substances of biological source, operations are often more expensive and/or based on different physical properties. In this case, water is separated by difference in density with the organic phase in a centrifugation unit and FFA by difference in molecular weight with other components of larger molecules in a membrane unit. The separation of triglycerides (the third most abundant non-product in the feed) without thermally damaging the biological properties of the product to be recovered, may

be carried out by e.g. molecular distillation or liquid–liquid supercritical extraction, which are sophisticated and costly operations. Therefore, recycling streams containing the components to be retrieved may not always be convenient in bioprocesses, opening an alternative to be considered: reprocessing. The recirculation of a stream to the process, with no available cost-effective separation of some component, causes an increase in the recycle flow rate due to the accumulation of not separated components, with a significant influence over investment costs due to large equipment sizes. This buildup of non product components may also have negative effects on both process variables and product yield: in the process here studied, oil accumulation can adversely affect the performance at the crystallization step of GAA.

The present analysis is intended to explore the application of process design rules, actually we compare the well established hierarchical process design procedure by Douglas (1988) with a more intuitive solution that arose naturally to solve the particular problem we had at hand: the accumulation of a non product component in the recycle, that negatively affected the process we were trying to design. Eventually, this can lead to the generation of a design heuristic specific for bioprocesses (but this would need much more study cases to back it), that should take into consideration other contributions in the field of design rules for bioprocesses, as the work by Steffens et al. (2000) who select separation operations based on physical properties data of the components involved (other than boiling points).

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