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ORIGINAL ARTICLE

Optimal synthesis and design of the number of cycles in the leaching process for surimi production

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Abstract Water consumption required during the leaching stage in the surimi manufacturing process strongly depends on the design and the number and size of stages connected in series for the soluble protein extraction target, and it is considered as the main contributor to the operating costs. Therefore, the optimal synthesis and design of the leaching stage is essential to minimize the total annual cost. In this study, a mathematical optimization model for the optimal design of the leaching operation is presented. Precisely, a detailed Mixed Integer Nonlinear Programming (MINLP) model including operating and geometric constraints was developed based on our previous optimization model (NLP model). Aspects about quality, water consumption and main operating parameters were considered. The minimization of total annual costs, which considered a trade-off between investment and operating costs, led to an optimal solution with lesser number of stages (2 instead of 3 stages) and higher volumes of the leaching tanks comparing with previous results. An analysis was performed in order to investigate how the optimal solution was influenced by the variations of the unitary cost of fresh water, waste treatment and capital investment.

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List of symbols

A	Area (m^2)
c	Concentration (mg/ml)
$CAGT$	Propeller agitator cost (U$S)
CEn_T	Total cost of the power consumption (U$S/year)
CFW	Fresh water cost (U$S/year)
$CLTK$	Leaching tank cost (U$S)
COL	Operational labor cost (U$S/year)
COM	Manufacturing cost (U$S/year)
$CSCPR$	Screw press cost (U$S)
CSP	Sanitary pump cost (U$S)
CRS	Rotary sieve cost (U$S)
CRM	Raw material cost (U$S/year)
CUT	Utility costs (U$S/year)
CWT	Waste treatment cost (U$S/year)
D_p	Particle's diameter (m)
$D_{\beta\gamma}$	Mass diffusivity (m^2/s)
FI	Investment cost (U$S)
FW	Fresh water flow stream (m^3/s)
FW_T	Total fresh water consumption (m^3/s)
J_D	Chilton and Colburn factor
k_c	Global mass transfer coefficient (m/s)
K	Distribution constant
M_w	Molecular weight (kDa)
M	Mass flow rate of minced fish (kg/s)
Q_{inl}	Inlet flow stream (m^3/s)
Q_{out}	Outlet flow stream (m^3/s)
Qr	Recycle flow stream (m^3/s)
R	Sample radius (m)
r	Variable radius (m)
S	Lost minced fish mass stream (kg/s)

T	Temperature (°C)
t	Time (s)
v	Agitation velocity (m/s)
V_{TK}	Leaching tank volume (m^3)
V_{op}	Operative volume of the leaching tank (m^3)
V_f	Volume of minced fish (m^3)
V_w	Volume of washing water (m^3)
Y	Percentage of extraction (%)

Dimensionless groups

Re	Reynolds's number
Sc	Schmidt's number

Greek symbols

ε	Volume fraction of solvent (dimensionless)
ρ	Density (kg/m^3)
μ	Viscosity (N·s/m^2)
θ	Residence time (s)
θ_T	Total residence time (s)

Subscripts

β	Proteins presented in minced fish
γ	Proteins presented in solvent phase
c	Cycle
EP	From the removable proteins
f	Minced fish
i	At interface
inl	Inlet
out	Outlet

Introduction

Surimi, minced and water-washed fish muscle tissue, has been used as a primary material for gelling foods enriched in myofibrillar proteins, such as kamaboko and shellfish analog products. The use of surimi as raw material or feedstock for further processing products is more efficient, in terms of transportation, storage, supply of raw materials, and better quality than the use of fresh fish (Fahrizal et al. 2015).

Fish sarcoplasmic proteins are a large family of proteins that include myogens and enzymes, and are soluble in water or low ionic strength solution. These components are normally washed out from surimi processing since they can lead to deterioration during surimi storage (Hemung and Chin 2013). Leaching process facilitates concentration of myofibrillar proteins by removal of soluble components (soluble proteins, other nitrogenous components, blood, etc.) from minced fish (Suzuki 1981).

The frequency of fish muscle washing is important, since the colour of surimi improves by increasing the washing frequency (number of cycles) (Park 2005), washing time and water quantity (Fahrizal et al. 2015). However, long period washing would result in high hydration of mince and degradation of myofibrillar proteins, making the subsequent dehydration process more difficult and could repress the gel-forming ability (Park 2005). Several works have been published considering the advantages and disadvantages of the washing frequency on quality properties of surimi, such us process yield, sensory evaluation, gel strength, texture analysis, water holding capacity and whiteness (Park 2005; Ng and Huda 2011; Yongsawatdigul et al. 2013; Ramadhan et al. 2014; Fahrizal et al. 2015; Fogaça et al. 2015; Hamzah et al. 2015). Despite the fact that in these articles the number of cycles is changed, the extraction time and/or water/mince ratio per cycle are maintained as constant. This operating mode could increase the loss of myofibrillar proteins and water usage because there is no control of the extraction rate with the number of cycles.

In particular, operating time, number of cycles, water/mince ratio, pH and temperature affect the final composition and texture analysis. Medina and Garrote (2002), Park (2005) and Reinheimer et al. (2013) exposed the importance of the mentioned variables on the final quality.

On the other hand, the surimi manufacturing process requires a large amount of wash water, which may contribute to high operating costs and environmental problems (Reinheimer et al. 2013). Therefore, surimi manufacturers have to consider minimizing water usage and reduce wastewater disposal. Analyzing quality and operating aspects evident trade-offs exist among the number of cycles, wash water ratio and leaching time per cycle. The key of the problem is to whether more water usage guarantees better surimi quality or if it is unnecessary and wasteful (Reinheimer et al. 2013). The analysis of the optimal design and operating mode of the leaching process, considering the optimal number of stages, extraction time and water/mince ratio is addressed in this work.

In previous works, a mathematical model to simulate the extraction process of soluble protein from *sábalo* (*Prochilodus platensis*) during the surimi elaboration was developed (Reinheimer et al. 2014). Indeed, the model validation using experimental data obtained at laboratory scale was presented there. The NLP optimization model with a superstructure formulation was developed considering different process arrangements (conventional, countercurrent and hybrid configurations) to minimize the fresh water requirement of the leaching process (Reinheimer et al. 2014). From several optimization results, it was concluded that the countercurrent arrangement leads to the minimal water consumption. In addition, the results also revealed that the minimum water consumption is achieved when three or more stages are used for a fixed total tank volume of the leaching process (for both cases: equal and different tank size distribution). But excessive washing not only increases the cost of water usage and wastewater treatment, but also results in a loss of myofibrillar proteins (Reinheimer et al. 2014).

The fresh water consumption decreases when the number of stages increases, due to the fact that more recycle streams are available in the countercurrent arrangement, reducing both the fresh water requirement and the liquid waste stream (Reinheimer et al. 2013). This clearly evidences a trade-off between operating costs and fixed investment costs. This point has not been deeply discussed in the literature. However, it is an important issue for processors and manufacturers. A review of the state of the art of the surimi production technology indicates a lack of works addressing the optimization of the leaching process, specifically the minimization of total annual cost (investment plus operating cost).

In this paper, the cost minimization problem, selecting the optimum number of process units and their sizes, as well as the optimum set of operating conditions (agitation speed, temperature and flowrates) has been formulated. Quality and technological constraints are simultaneously considered. To do this, a superstructure of alternative configurations for the leaching process design is proposed resulting in a Mixed Integer Non-Linear Programming (MINLP) model. Superstructure-based models (NLP and/or MINLP models) have been applied successfully in several application areas such as, among others, energy systems (Manassaldi et al. 2014, 2016; Oliva et al. 2011; Zondervan et al. 2011); wastewater treatment systems (Alasino et al. 2010); multiperiod blend scheduling problems (Kolodzieja et al. 2013); biodiesel production (Rizwan et al. 2013); desalination seawater processes (Mussati et al. 2006, 2008). In this paper, the proposed MINLP model is used to determine the optimal process configuration, sizes of each process unit and operating conditions that minimize the total annualized cost for a desired production.

The optimization model is based on previous ones including now the cost model. This is a hard challenge to solve due to the high nonlinearities and the large number of variables. The model includes both, partial differential and algebraic equations. Further difficulty encountered, is to capture product quality and operating aspects through the problem constraints.

Optimization model

For the superstructure of alternative configurations illustrated in Fig. 1, the optimization problem can be formally defined as follows: determine the optimal number of stages (size, operating temperature and agitation speed of each one), protein concentrations and flow-rates; minimizing the annualized total cost, given a minimum desired extraction yield and satisfying operating and quality constraints.

The MINLP problem is formulated defining a superstructure describing the space of feasible alternatives, according to:

$$\begin{aligned} &\min_{x,y} \; f(x,y) \\ &s.t. \quad h(x,y) = 0 \\ &\qquad\; g(x,y) \le 0 \\ &\qquad\; x \in X \subseteq \Re^n \\ &\qquad\; y \in Y \; integer \end{aligned}$$

where $f(x,y)$ is the objective function (for this case the annualized total cost). Equality constraints correspond to the process model (mass balances, physicochemical and equilibrium relationships, among others). Inequality constraints are all the technological, operating and quality constraints such as allowed maximum temperature, maximum (total and per tank) volumes, minimum extraction yield, water/mince ratios among others.

In the following sections, a detailed description of the optimization model will be presented.

As shown in Fig. 1 the superstructure embeds several stages arranged in a countercurrent pattern where the unwashed minced fish is fed at the first stage while the fresh water is fed at the last stage, according to previous results of Reinheimer et al. (2013). As indicated, a generic stage is composed by a leaching tank, a rotary sieve and a sanitary pump. A maximum number of four stages ($c = 4$) is introduced in the superstructure. The existence of each stage is modeled by using binary variables [$y(c)$], which are included in linear inequality constraints, considering that:

$$y(c) = \left\{ \begin{matrix} 1 & \textit{if the stage exists} \\ 0 & \textit{if not} \end{matrix} \right\}$$

In our case, the following constraint is imposed:

$$1 \le \left(\sum_{c=1}^{c_{\max}} y(c) \right) \le 4$$

Model assumptions

Main assumptions for this model are the same proposed and verified in earlier works (Reinheimer et al. 2013, Reinheimer et al. 2014) and are summarized as follows:

- Spherical particle (minced fish) is considered.
- Soluble proteins diffuse to the surface of each sphere according to Fick's second law.
- Model 1-D. Temporal evolution in radial direction of total protein concentration is contemplated.

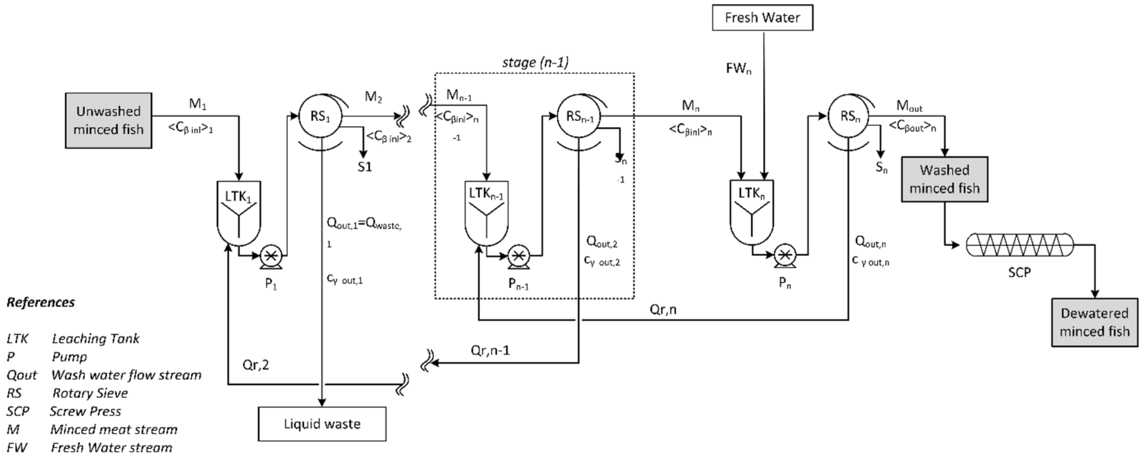

Fig. 1 Superstructure for the countercurrent leaching process in the manufacturing process of surimi

- Size and shape of spherical particles and the density of the minced fish do not change along the leaching process.
- Continuous process and perfect stirred tanks are considered.
- Only soluble proteins diffuse within the particles. At the solid–liquid interface, the transference is made by convection.

Process model

The mathematical model basically includes the kinetic model of soluble protein extraction, as well as the global and protein mass balances in each leaching stage and equilibrium and physicochemical relationships, which are described in detail in "Appendix". Operative and technological constraints are imposed by inequality restrictions and are also presented in the "Appendix" section. For more details, the reader is referred to Reinheimer et al. (2014).

Discrete decisions are handled by binary variables according to the existence (or not existence) of each stage, as mentioned in previous section.

The lower and upper bounds of operating and design constraints are closely related to the final quality aspects. Water temperature values must be in the safety operating range to retain functional properties. Protein concentration and humidity content in the final products are the main factors affecting the gel strength. Therefore, a minimum extraction percentage (Y_{EP}) is required in order to ensure an adequate and acceptable gel strength (Park 2005; Reinheimer et al. 2014).

Solvent volume fraction, residence time and number of cycles are the variables that are strongly linked to one another for the unit operation design and quality aspects. The most typical range of volume fraction used at the industrial scale was adopted for this work. Then, the combination of the optimal number of stages, residence time per cycle and solvent volume fraction must be contemplated from the quality point of view. The arrangement of these variables is critical for the loss of myofibrillar proteins.

The agitation speed range must be selected in order to obtain adequate efficiency and preventing high speed, which can result in a temperature rise as well as difficulty in dewatering by the screw press.

Several constraints are incorporated in the model to consider all the aspects aforementioned.

During surimi processing, especially when marine fish species are used as raw material, the wash water temperature ranges from 5 to 10 °C (Suzuki 1981; Sonu 1986). Also, the average soluble protein extraction of different fish species used for surimi production is about 60% (Sonu 1986; Green and Lanier 1999). Finally, water volume fractions from 0.7 to 0.9 are commonly used.

Here, wash water temperature, the average soluble protein extraction and the overall water to meat ratio are optimization variables and restricted by given operative ranges:

- Extraction Temperature: $T_{min} \leq T \leq T_{max}$
- Extraction percentage of soluble proteins: $Y_{EP} \geq Y_{EP\,min}$
- Solvent volume fraction: $\varepsilon_{w,c\ min} \leq \varepsilon_{w,c} \leq \varepsilon_{w,c\ max}$

In this model, the pH evolution and control have not been considered. However, this variable must be controlled in an adjusted range to assure the desired gel properties. Reducing the pH of the washing stage near the isoelectric point of fish proteins will facilitate dewatering and gels with higher stress characteristics (Park 2005).

Cost model

The manufacturing cost (COM, U$S/year) is calculated as suggested by Turton et al. (2009):

$$COM = 0.28 \cdot FI + 2.37 \cdot COL + 1.23 \cdot (CRM + CWT + CUT) \quad (1)$$

where FI is the investment cost. An equipment depreciation of 10% is considered for equipment investment. COL is the operational labor cost (U$S/year); CRM is the raw material cost (U$S/year); CWT is the waste treatment cost (U$S/year) and CUT the utility costs (U$S/year), defined as:

$$CUT = CEn_T + CFW \quad (2)$$

where CEn_T is the total cost of the power consumption and CFW the fresh water cost.

As mentioned earlier, each leaching stage (c) is composed by a leaching tank (LTK) equipped with a propeller agitator (AGT), a sanitary pump (SP) and a rotary sieve (RS). The equipment costs are computed using the correlations published in Seider et al. (2009). Some of them are considered as fixed cost, considering the operating condition range managed in this work, which are reported in Table 1; while others depend on the main process variables, as listed below.

For each stage, the leaching tank cost is calculated as:

$$CLTK(c) = 2500 \cdot y(c) + 36 \cdot (264.172 \cdot v_{LTK}(c))^{0.72} \quad (3)$$

where the first term of Eq. 3 corresponds to fixed tank cost comprising foundations fixed pipelines and others structural costs, which are independent of the volume. Then, the total stage cost ($CLTK_T$) considering only tank costs is:

$$CLTK_T = \sum_{c=1}^{c_{\max}} (CLTK(c)) \quad (4)$$

where $CLTK(c)$ is the cost of each leaching tank, which depends on its volume, $v_{LTK}(c)$ and $y(c)$ is a binary variable (1 or 0) associated with the existence or non-existence of each stage, and c_{max} is the maximum number of stages (in this problem four stages are adopted, as mentioned previously).

The constraints for the volume tank's distribution corresponding to a countercurrent configuration (in which the fresh water stream flow enters through the last stage) applied in this case is:

$$v_{LTK}(c) \geq v_{LTK}(c-1), \quad for\ c > 1 \quad (5)$$

where

$$v_{LTK\ \min} \leq v_{LTK}(c) \leq v_{LTK\ \max}$$

The agitator cost is computed as:

$$CAGT(c) = 2.3 \cdot P_{AGT}^{0.34} \quad (6)$$

Then, the total train cost, $CAGT_T$, is calculated as:

$$CAGT_T = \sum_{c=1}^{c_{\max}} (y(c) \cdot CAGT(c)) \quad (7)$$

where $CAGT(c)$ is the cost corresponding to the propeller agitator of each leaching tank. For the speed range considered, the propeller agitator power consumption, P_{AGT}, is assumed constant (reported in Table 1).

Each stage is also equipped by a sanitary pump and a rotary sieve. The costs of these units (CSP and CRS, respectively) are reported in Table 1. Then, the total pump and rotary sieve costs for all the train (according to the associated binary variable related) are calculated respectively as follows:

$$CSP_T = \sum_{c=1}^{c_{\max}} (y(c) \cdot CSP(c)) \quad (8)$$

$$CRS_T = \sum_{c=1}^{c_{\max}} (y(c) \cdot CRS(c)) \quad (9)$$

As can be seen in Fig. 1, a screw press is located at the exit of the last washing stage. Its cost ($CSCPR$) is estimated as:

$$CSCPR = \exp\left\{10.9733 - 0.3580 \cdot [\ln(7936.5080 \cdot M_{out})] + 0.05853 \cdot [\ln(7936.5080 \cdot M_{out})]^2\right\} \quad (10)$$

where M_{out} is the exit minced fish mass flow rate. By using the screw press the final dewatering is produced prior to blending the washed minced with cryoprotectants and freezing (Park 2005).

Then, the fixed capital investment calculated as the sum of the equipment costs involved in the process (FI, used in Eq. 1) is:

$$FI = CLTK_T + CAGT_T + CSP_T + CRS_T + CSCPR \quad (11)$$

The power consumption costs are calculated as:

$$En(c) = (P_{AGT}(c) + P_{SP}(c) + P_{RS}(c)) \quad (12)$$

$$En_T = \sum_{c=1}^{c_{\max}} (En(c)) + P_{SCPR} \quad (13)$$

where $En(c)$ is the power consumption per stage, including the following equipment: propeller agitator, P_{AGT}, sanitary pump, P_{SP} and the rotary sieve, P_{RS}. Then, En_T is the total

Table 1 Process parameters used for COM estimation

Daily operating time	[DT, h/day]	5
Operating days per year	[OD, day/year]	330
Mass flow rate of minced fish	[M_{inl}, kg/s]	1.253
Number of operators	[n_{op}, adim]	1

power consumption of the washing process. So, the total power consumption cost of the washing process, CEn_T, is calculated as:

$$CEn_T = UCEn \cdot En_T \cdot OT \tag{14}$$

$$OT = DT \cdot OD \tag{15}$$

where $UCEn$ is the unitary power cost, and OT, is the annual operating time expressed in hours, DT is the daily operating time, and OD is the annual operating days.The fresh water cost, CFW, is calculated as:

$$CFW = UCFW \cdot FW \cdot OT \tag{16}$$

where $UCFW$ and FW are the unitary cost and the fresh water flow stream requirement, respectively.

The liquid waste treatment cost is computed as:

$$CWT = UCWT \cdot QW \cdot OT \tag{17}$$

where $UCWT$ is the waste treatment unitary cost and QW is the liquid waste flow stream.The minced fish cost is estimated as follows:

$$CRM = UCRM \cdot M_{inl} \cdot OT \tag{18}$$

where $UCRM$ is the raw material unitary cost and M_{inl} is the inlet mass flow rate of minced fish.Finally, the labor cost is computed as:

$$COL = n_{op} \cdot COP \cdot OT \tag{19}$$

where COP is the operational labor cost per working hour and n_{op} is the number of operators. Here, it is assumed that one operator is needed independently of the number of stages.

Solution strategy

MINLP problems are usually solved using decomposition methods. Binary variables are applied in linear constraints to avoid nonlinearities between discrete and continuous variables. Several ways can be applied for performing such a transformation. Widespread transformations are the so-called 'Big M', 'Multi M' and 'Convex Hull' methods (Raman and Grossmann 1991; Grossmann and Turkay 1996; Vecchietti et al. 2003), all are equally proper methodologies. In this paper, the simplest transformation of equations is used as indicated in Eq. (20):

$$x_i(c) \le M_i \cdot y(c) \tag{20}$$

where $x(c)$ is the continuous variable, M refers to a large numerical value and $y(c)$ the binary variable: 1 if the stage exists and 0 if not. These transformations are applied to the following continuous variables: residence time per stage, $\theta(c)$ leaching tank volume, $v_{TK}(c)$ volume of minced fish, $v_f(c)$ volume of washing water, $v_w(c)$ inlet flow stream, $Q_{inl}(c)$ outlet flow stream, $Q_{out}(c)$ recycle flow stream, $Q_r(c)$ and percentage of extraction (soluble proteins), Y_{EP}. The objective is to decide which stages remain in the leaching process. Here, the mentioned constraints are listed with their suggested numerical values for the M_i factor:

$$\theta(c) \le 9000 \cdot y(c) \tag{21}$$

$$v_{TK}(c) \le 200 \cdot y(c) \tag{22}$$

$$v_f(c) \le 200 \cdot y(c) \tag{23}$$

$$v_w(c) \le 200 \cdot y(c) \tag{24}$$

$$Q_{inl}(c) \le 100 \cdot y(c) \tag{25}$$

$$Q_{out}(c) \le 100 \cdot y(c) \tag{26}$$

$$Q_r(c) \le 100 \cdot y(c) \tag{27}$$

$$Y_{EP}(c) \le 100 \cdot y(c) \tag{28}$$

The following constraint ensures that the remaining stages are consecutives:

$$y(c-1) \le y(c) \tag{29}$$

The optimization model is implemented in GAMS (General Algebraic Modeling System) and involves 3709 constraints (equalities and inequalities) and 2797 variables (6 integer variables). SBB (Standard Branch and Bound) is used as solver for the resulting MINLP model (Brooke et al. 1992). It is based on a combination of the Standard Branch and Bound method known from Mixed Integer Linear Programming and some of the standard NLP (nonlinear programming) solvers already supported by GAMS. Intel Core i7 M480 2.67 GHz processor and 8 GB RAM has been used to perform the optimizations.

Results and discussion

A processing plant with a production capacity of 10 tons of frozen surimi (resulting in a surimi yield of 24% from the weight of the raw fish input) on an 8-h day basis is considered for the optimization problem. Table 1 reports the data related with the operation mode. Cost parameters are presented in Table 2. Finally, Table 3 lists lower and upper bounds used for quality and technological constraints.

The first step was to verify the proposed MINLP model comparing the output results with those obtained by the NLP model reported in Reinheimer et al. (2013). To do this, it was necessary to fix several binary variables in the MINLP model in order to make an appropriate comparison. In addition, the same objective function investigated in Reinheimer et al. (2013) (minimization of the total water consumption) was also used in the MINLP. Different optimization runs were performed varying the total tank volume (each leaching tank volume as a free variable) as in Reinheimer et al. (2013) obtaining, as expected, the same optimal results.

Table 2 Economic parameters used for COM estimation

Sanitary pump cost	[*CSP*(*c*), U$S]	5000
Rotary sieve cost	[*CRS*(*c*), U$S]	220,000
Propeller power consumption	[P_{AGT}, kW]	0.56
Sanitary pump power consumption	[P_{SP}, *kW*]	2,76
Rotary sieve power consumption	[P_{RS}, kW]	0.74
Screw press power consumption	[P_{SCPR}, kW]	1.64
Unitary cost of raw material	[*UCRM*, U$S/Kg]	0.7
Unitary cost of waste treatment	[*UCWT*, U$S/m^3]	0.33
Unitary cost of energy	[*UCEn*, U$S/kW h]	0.06
Unitary cost of fresh water consumption	[*UCFW*, U$S/m^3]	0.9
Operational labor cost	[*COP*, U$S/h]	16

Table 3 Lower and upper bounds for the main optimization variables

Variable	Lower bound	Upper bound
Agitation velocity, v (m/s)	0.05	0.3
Water temperature, T (°C)	5	10
Leaching tank volumes, $V_{TK,c}$ (m^3)	1	100
Percentage of extraction from soluble proteins, Y_{EP} (%)	60	100
Volume fraction of solvent, $\varepsilon_{w,c}$ (adim)	0.7	0.9

After that the MINLP models verification was carried out, the model was used to determine the optimal number of stages, sizes and the rest of optimization variables. In addition, a sensitivity analysis regarding main equipment and operating costs was also investigated. In contrast to the verifications case, all the binary variables are considered as optimization variables instead of fixed ones. Indeed, minimization of total annual cost is here used as objective function instead of the water consumption minimization.

Table 4 presents the optimal values after solving the MINLP problem (*min.COM*). For the proposed objective function and for the assumed parameter values, the optimal number of stages is two, as shown in Table 4. In contrast to previous results presented in Reinheimer et al. (2013), the trade-off between the total investment and total operating costs imposes lesser number of stages (2 vs 3). Also, both variables, water requirement and total tank volume are now different (Table 4), compared with previous results (Reinheimer et al. 2013).

As mentioned above, final quality is related to the number of stages, residence time, solvent/mince fraction, and temperature as the main variables. From the quality point of view, optimal values of temperature, residence time per stage and solvent volume fraction in each stage are according to values reported in the literature to prevent relative losses of protein functionality (Sonu 1986; Canpolar Inc 1988; Park 2005). However, temperature, solvent volume fraction and percentage of extraction take their lower or upper bounds in order to reduce water consumption, operating time and to minimize operating costs. In fact, the percentage of extraction (soluble proteins) takes its lower bound to reduce operating time and costs, but this bound ensures the final quality.

It is interesting to remark that in the last two decades, various authors incorporated quality and environmental aspects in their studies for the surimi production made by different meats (fish, chicken, duck). In the works of Ninan et al. (2004), Department of Fisheries Malaysia (2009) and Yongsawatdigul et al. (2013), two leaching stages were implemented and the gel properties have been enhanced, compared with the gel properties obtained by traditional recipes of three stages reported in the works of Suzuki (1981), Sonu 1986, Green and Lanier (1999). However, these authors made experiments with fixed residence time and solvent/mince ratio per cycle. Here, all these variables are simultaneously considered.

In order to clearly visualize the existing trade-off between investments and operating costs, a parametric study was performed in order to investigate the influence of the water consumption as a function of the number of stages and the total tank volumes. For this purpose, the MINLP model is solved by minimizing the total annual cost fixing the number of stages from 1 to 4, and simultaneously fixing the total volume of the leaching tanks in a range (60–160 m^3) varying ±40% of the total volume of the previous solution (Table 4). As described above (fixed investment costs), each point in the curves plotted in Fig. 2, corresponds to minimum water consumption.

Fresh water consumption (Fig. 2) tends to an asymptotic value: 3 stages and 120 m^3 (total tank volume), which can

Table 4 Optimal main process variable values

Variable	Value
Number of stages, *c* (adim)	2
Leaching tank volumes, $V_{TK,1,2}$ (m^3)	59.46
Total volume, $V_{TK,T}$ (m^3)	118.92
Percentage of extraction from soluble proteins, Y_{EP} (%)	60^a
Volume fraction of solvent, ε_c (dimensionless)	0.7^a
Water temperature, *T* (°C)	10^b
Agitation speed, v_c (m/s)	0.3^a
Residence time per stage, θ_c (s)	624
Fresh water consumption, FW_T (m^3/s)	0.054
Fixed capital investment, *FI* (US$)	8.4255 E+5
Manufacturing cost, *COM* (US$/year)	7.4619 E+6

[a] Ower bound

[b] Upper bound

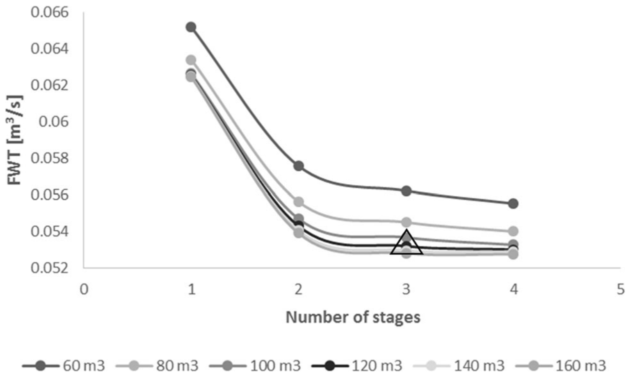

Fig. 2 Fresh water consumption for a leaching process considering from 1 to 4 stages and different total volumes

be assumed as a "practical" approximation (referenced in Fig. 2 with Δ) of the optimal values. As expected, the previous optimal solution reported in Reinheimer et al. (2013) involves greater number of stages because the model does not consider any cost. When the costs are introduced, the optimal solution moves to lesser number of stages (2 stages) and it is practically maintained above the mentioned asymptote of the total tank volume (118.92 m^3).

The optimal solution strongly depends on the adopted economic parameters. The sensitivity of the optimal solution is here discussed varying the costs in a ±30% range. Figure 3 shows the relative influence of the economic parameter variations over the total annual cost. It is shown that the fresh water cost (UCFW) is the major contribution to the total annual cost, followed in order of importance by rotary sieve investment (CRS), waste treatment cost (UCWT) and finally, the leaching tank cost (CLTK). As shown in Fig. 3, the two last mentioned cost variations have no significant impact on the total annual cost.

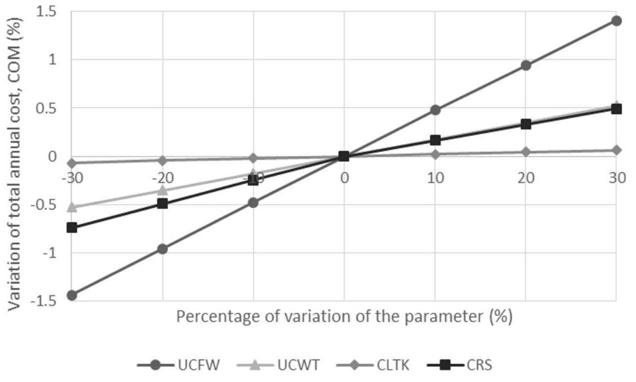

Fig. 3 Relative influence of cost variations over the total annual cost

As explained before, each stage involves units assumed with fixed costs, as the rotary sieve and sanitary pump. The rotary sieve represents the highest fixed cost associated to each stage. Hence, this cost impacts in the decision (reduction) of the number of stages in the optimal solution, which is close to the minimal fresh water consumption as can be observed in Fig. 3. This is because the fresh water cost is the more sensitive cost. In addition, the total tank volume corresponds to the lesser sensitive due to its relative minor cost.

Despite the above results, at the industrial scale operation, the leaching process in surimi manufacturing plants typically are conformed by three stages (Sonu 1986; Green and Lanier 1999; Park 2005). The configuration process with three instead of two stages could be justified from the point of view of the operating flexibility and control. For example, in case of operative or mechanical failures in an equipment unit at any stage. However, aspects like operability and controllability have not been taken into account in the optimization problem.

Conclusion

A mathematical optimization model for the optimal design of the leaching process in surimi production process was presented. A detailed Mixed Integer Nonlinear Programming (MINLP) model including operational and geometric constraints has been developed to minimize the total annual cost for a given production level. The proposed model includes discrete decisions associated with the number of stages (leaching tanks and auxiliary equipment units, such as sanitary pumps and rotary sieves), which have been modeled by using integer variables. Continuous variables are used for process conditions (temperatures, flow-rates, tank volumes, speeds, extraction rate, among others). From a computational cost point of view, the model resulted to be enough flexible to perform optimizations and sensitivity analyses. An optimal solution considering one leaching stage less than traditional recipes of three stages was obtained maintaining the composition constraints associated with quality aspects.

The optimal total tank volume and the optimal water consumption required to treat a mass flow rate of minced fish of 1.253 kg/s are 118.92 m^3 and 0.054 m^3/s, respectively.

Therefore, this paper presents two important contributions, one in the field of the synthesis and process design of soluble protein extraction, using phenomenological models and advanced mathematical programming tools and the second one for surimi manufacturers considering that the total costs can be minimized maintaining quality's target.

As future works, aspects such as final quality parameters (gel strength, water holding capacity, among others), operating flexibility and control will be further included in the MINLP model.

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Appendix

The kinetic model of soluble protein extraction from spherical particles of minced fish is described as follows:

$$(1-\varepsilon_{w,c})\cdot\frac{1}{D_{\beta\gamma}}\frac{\partial c_{\beta,c}(r,t)}{\partial t}=\frac{\partial^2 c_{\beta,c}}{\partial r^2}+\frac{2}{r}\frac{\partial c_{\beta,c}}{\partial r},\quad 0<r<R \tag{30}$$

$$c_\beta=\langle c_{\beta\ inl}\rangle_c,\quad t=0,\quad \forall\, 0\le r\le R \tag{31}$$

$$\frac{\partial c_{\beta,c}(r,t)}{\partial r}=0,\quad r=0,\quad \forall t>0 \tag{32}$$

$$-D_{\beta\gamma}\frac{\partial c_{\beta,c}}{\partial r}=k_{c\gamma,c}\left(c_{\gamma i,c}-c_{\gamma\ out,c}\right),\quad r=R,\quad \forall t>0 \tag{33}$$

Equations (30), (32) and (33) were discretized using the central finite difference method (CFDM) using the second-order accurate method in both space and time. The number of discretization nodes used for the time-domain and spatial-domain were, respectively 50 and 10.

The semi-empirical equation of Polson (1950) was used to estimate the protein diffusion coefficient, $D_{\beta\gamma}$, which is recommended for biological solutes:

$$D_{\beta\gamma}=\frac{9.40e-15\cdot T}{\mu_w\cdot\left(Mw_\gamma\right)^{1/3}} \tag{34}$$

where Mw_γ is the protein's molecular weight, T is the system's temperature and μ_w is the fluid viscosity.

The overall mass transfer coefficient was calculated using the correlation proposed by Geankoplis (1993) for fixed beds and also valid for fluidized beds of spheres in the Reynolds number range of 10–4000:

$$J_D=\frac{0.4548}{\varepsilon_c}\cdot \mathrm{Re}^{-0.4069} \tag{35}$$

$$kc_{\gamma,c}=\frac{J_D\cdot v_c}{Sc^{2/3}} \tag{36}$$

where

$$\mathrm{Re}_c=\frac{Dp\cdot\rho_\gamma\cdot v_c}{\mu_\gamma} \tag{37}$$

and

$$Sc=\frac{\mu_\gamma}{D_{\beta\gamma}\cdot\rho_\gamma} \tag{38}$$

The equilibrium of soluble proteins concentration under diluted assumption is expressed as:

$$c_{\gamma i}=K\cdot c_{\beta i} \tag{39}$$

The average concentration of total proteins in phase β, after the leaching process, is computed as follows:

$$\langle c_{\beta\ out}\rangle_c=3A\int_0^R c_\beta\Big/(AR)\ ,\quad t=\theta_c \tag{40}$$

The initial concentration of protein in the washed minced fish is equal to the final protein concentration of the previous cycle (Fig. 3). Then, the following constraints are considered:

$$\langle c_{\beta\ out}\rangle_c=\langle c_{\beta\ inl}\rangle_{c+1}\ ,\quad c=1,2,\ldots(c_{\max}-1) \tag{41}$$

The maximum percentage of extraction $[Y_{EP}]$ is defined as the ratio of the amount of proteins extracted after washing and the maximum amount of proteins that can be extracted, according to the following constraints:

$$Y_{EP}\%=\frac{\langle c_{\beta\ inl}\rangle_{EP,1}-\langle c_{\beta\ out}\rangle_{EP,c}}{\langle c_{\beta\ inl}\rangle_{EP,1}}\% \tag{42}$$

where:

$$\langle c_{\beta 0}\rangle_{EP,1}=0.25\cdot\langle c_{\beta 0}\rangle_1 \tag{43}$$

The leaching process model in the countercurrent configuration is described as follows:

It is considered that the operative volume of the leaching tanks is 70 percentage of the total volume:

$$V_{op,c}=0.7\cdot V_{TK,c} \tag{44}$$

where the total volume ($V_{TK,T}$) is given by:

$$V_{TK,T}=\sum_{c=1}^{3}V_{TK,c} \tag{45}$$

During the washing process in each cycle, the operative volume of the tank is filled by minced fish and washing water and is computed as follows:

$$V_{op,c} = V_{f,c} + V_{w,c} \tag{46}$$

The residence time of the minced fish during each washing stage is calculated as:

$$\theta_c = \frac{V_{f,c} \cdot \rho_f}{M_c} \tag{47}$$

The volume fractions of minced fish and solvent are calculated as follows:

$$\varepsilon_{f,c} = \frac{V_{f,c}}{V_{op,c}} \tag{48}$$

$$\varepsilon_{w,c} = \frac{V_{w,c}}{V_{op,c}} \tag{49}$$

where:

$$\varepsilon_{f,c} + \varepsilon_{w,c} = 1 \tag{50}$$

The global and protein mass balances in each leaching stage (leaching tank and rotary sieve, as control volume) is given by:

$$Qr(c+1) = Q_{inl}(c), \quad c = 1, 2, \ldots (c_{\max} - 1) \tag{51}$$

$$FWT = Q_{inl}(c), \quad c = c_{\max} \tag{52}$$

$$Qr(c+1) \cdot c_{\gamma}(c+1) = Q_{inl}(c) \cdot c_{\gamma\ inl} \tag{53}$$

where

$$Qout(c) = Qr(c), \quad c > 1 \tag{54}$$

$$Qout(c) = QwasteT, \quad c = 1 \tag{55}$$

$$Qout(c) = E(c), \quad \forall c \tag{56}$$

The mass balance at each leaching tank is given by:

$$M(c) \cdot \langle c_{\beta\ inl} \rangle_c + Q_{inl}(c) \cdot c_{\gamma\ inl,c} = Q_{out}(c) \cdot c_{\gamma\ out,c} + M(c+1) \cdot \langle c_{\beta\ out} \rangle_{c+1} + S(c) \cdot \langle c_{\beta\ out} \rangle_c \tag{57}$$

where:

$$M(c+1) = 0.98 \cdot M(c), \quad c = 1, \tag{58}$$

$$M(c+1) = 0.99 \cdot M(c), \quad c > 1 \tag{59}$$

and

$$M(c) = M(c+1) + S(c), \quad c < c_{\max} \tag{60}$$

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