STUDY OF OPERATIVE VARIABLE INFLUENCE ON THE VEGETABLE OIL EXTRACTION PROCESS

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***Abstract*— In the vegetable oil extraction process it is possible to identify a sequence of process units that deal with solids: extractor, desolventizer and meal drier, the so-called ¨solid line¨. Such unit operations are rarely included in available commercial process simulators, so it was necessary to develop a simulation tool for this particular application. The model obtained by using mass and energy conservation equations is represented by a differential equation system. In each process unit, heat and mass transfer flows were written in terms of lumped parameter constitutive equations. Industrial data were collected in a soybean extraction facility and the parameters of the model were estimated by using the Marquardt method. This simulation tool allows to analyze the influence of operative variables on solid line performance. It was found that the main variables that affect extraction efficiency are operation temperature, solid material preparation characteristics and residence time. In the desolventizer, the direct steam flow and the amount of indirect heat affect final meal solvent and moisture content. Inlet air flow, temperature and humidity have an important effect on the degree of meal moisture elimination in the drier. Effects of changes in the amount of solid material treated in the extractor is shown in this work.**

***Keywords*— Extraction, Modeling, Simulation**.

I. INTRODUCTION

Among the computational tools developed with the objective of facilitating the combined analysis of all the operative variables involved in a process it is possible to find commercial process simulators (Aspen, Pro II, etc.) which contain a wide range of simulated operations. In general, these simulators, guided mainly towards the chemical and petrochemical industry, do not have the processes of solid transformation among their basic operations. In the oilseed industry, available commercial simulators have been of great utility to optimize the process operation in which solids do not intervene, i.e. in solvent recuperation system (Ferrero, 2001). Some advances in the development of a specific simulation tool for this solid line have been carried out (Pramparo et al., 2000).

When conditioned oilseeds come into the treatment plant, they are subjected to the following main operations: extraction, desolventization and drying. In the modeling of the mentioned operations appear the contributions of Rice (1982), Majumdar et al. (1995) and Schwartzberg et al. (1987), who have modeled the extraction operation; Cardarelli (1999) and Martinello et al. (1994) have made contributions in the modeling of the desolventizer. Regarding drying, the available design methods are rather empirical. There are some references about the modeling of the drier types employed in this process (Crapiste and Rotstein, 1997), although there are not references concerning to this particular application.

In the present work it has been simulated the performance of the extraction line that is constituted by the three mentioned operations. The developed software has been adapted to the operative conditions of a soybean extraction plant by estimating the parameters involved in the mathematical models by means of the utilization of the industrial plant data. This tool has been used to study the influence of the operative variables involved in the process.

II. METHODS

A.Model Description

Extractor. The modeled extractor consists of a perforated belt (longitudinal or circular) where the solid material is placed. The extraction solvent (hexane) percolates through the bed by gravity action. Solvent/miscella is sprinkled in different stages, conforming a counter-current scheme. The mathematical model consists of three main parts. The first one is the simulation of one stage. A system of partial differential equations represents transfer phenomena in a fixed bed of particles. The second part consists of a system of algebraic equations obtained from the macroscopic mass balances for each stage. The third part of the model consists of the drainage stage simulation. This stage is located at the final part of the extractor in order to eliminate retained solvent within interstices of solid material. For this part,

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the proposal of drainage through a porous solid matrix developed by Zeitsch (1990) has been utilized. The model equations that describe the extraction process in one stage (fixed bed arrangement) are the following:

$$\frac{\partial x_o}{\partial t^*} = -\frac{k\, a_v\, H}{v_m\, \rho_m}(K\, x_o - y_o) \qquad (1)$$

$$\frac{\partial y_o}{\partial t^*} + \frac{\partial y_o}{\partial z^*} = \frac{1}{Pe}\frac{\partial^2 y_o}{\partial z^{*2}} + \frac{(1-\varepsilon) H\, k\, a_v}{\varepsilon\, v_m \rho_m}(K\, x_o - y_o) \qquad (2)$$

In these equations $z^* = z/H$ is dimensionless axial coordinate and $t^* = tv_m/H$ is dimensionless time. v_m is miscella superficial velocity, x_o and y_o are oil mass fraction in solid and liquid phases, respectively and $Pe = v_m H/D_z$ is Peclet number. In order to solve this partial differential equation system, the following boundary conditions have been used:

$$\begin{aligned} &t^* = 0 && x_o = x_o^0 \quad y_o = y_o^{inc} \\ &z^* = 0 && y_o\big|_{z^*=0} = \frac{1}{Pe}\left.\frac{\partial y_o}{\partial z^*}\right|_{z^*=0} \\ &z^* = 1 && \left.\frac{\partial y_o}{\partial z^*}\right|_{z^*=1} = 0 \end{aligned} \qquad (3)$$

where x_o^0 is initial oil mass fraction in solid phase and y_o^{inc} is oil mass fraction in incoming liquid phase.

Desolventizer. The modeled desolventizer consists of a vertical shell containing indirect-steam-heated-trays at the top (predesolventizing section) followed by perforated trays provided with sparging steam (desolventizing section). The solid phase runs down these trays and direct steam crosses solid material in ascending way conforming a countercurrent scheme. The proposed model is the result of solving material balances for solvent in solid and vapor phases and energy balance for solid phase. In the vapor phase, hexane content and temperature are related by the condition of the equality of partial pressure sum to total pressure. The model equations are:

$$\frac{dv_v}{dz^*} = \frac{n_h\, a_v\, H}{\varepsilon\; v_v^{in}\, \rho_v} - \frac{n_w\, a_v\, H}{\varepsilon\; v_v^{in}\, \rho_v} \qquad (4)$$

$$y_h \frac{dv_v}{dz^*} + v_v \frac{dy_h}{dz^*} = \frac{1}{Pe}\frac{d^2 y_h}{dz^{*2}} + \frac{n_h\; a_v}{\varepsilon\;\; v_v{}^{in}\;\; \rho_v} \qquad (5)$$

$$\frac{dx_h}{dz^*} = -\frac{n_h\, a_v\, H}{(1-\varepsilon\;)v_s\, \rho_s} \qquad (6)$$

$$\frac{dT_s}{dz^*} = \frac{(Q - n_h\, a_v\, \lambda_h)H}{\rho_s (1-\varepsilon) C_{ps}\, v_s} \qquad (7)$$

In these equations x_h and y_h are hexane mass fraction in solid and vapor phases. Boundary conditions to solve these equations are:

$$\begin{aligned} &z^* = 0 && y_h^{in} - y_h\big|_{z^*=0} = \frac{1}{Pe}\left.\frac{dy_h}{dz^*}\right|_{z^*=0} \\ & && v_v = -1\;;\; x_h = x_h^{in}\,;\, T_s = T_s^{in} \\ &z^* = 1 && y_h = 0 \end{aligned} \qquad (8)$$

where x_h^{in} is hexane mass fraction in incoming solid phase and T_s^{in} is incoming solid temperature.

The flow density of removed solvent, n_h, and the specific transferred heat, Q, are obtained by the following expressions for each of the DT sections:

- Predesolventizing section.

$$n_h = \frac{Q}{\lambda_h\, a_v} \qquad (9)$$

$$Q = A_t\, U\; (T_w - T_s) \qquad (10)$$

where T_w is indirect steam temperature.

- Desolventizing section.

$$n_h = k\left(K\, x_h - y_h\right) \qquad (11)$$

$$Q = h\left(T_v - T_s\right) \qquad (12)$$

In the desolventizing section the flow of condensing steam, n_w, is expressed by:

$$n_w = \frac{Q}{a_v\, \lambda_h} \qquad (13)$$

Vapor velocity, v_v, is defined as the ratio of vapor velocity at position z to inlet vapor velocity, v_v^{in} . v_s is solid velocity.

Drier. Meal drying is accomplished in different types of driers, such as the drying section of the DTDC (Desolventizer-Toaster-Drier-Cooler) and rotary driers. The former consists of perforated trays that allow the hot air-meal contact. In the developed software both types of driers were considered. The model of the first one is based on moisture mass and energy balances for gas and solid phases, assuming plug flow. Regarding the rotary drier, a semi empirical equation was used for residence time, τ, calculation:

$$\tau = \frac{0.2\;\, L}{D\;\, N^{0.9}\;\, \tan(\alpha)} + \frac{10\;\, D_p{}^{-0.5}\;\, L\;\, G}{F} \qquad (14)$$

here D and L are equipment diameter and length, α is drum slope, N is rotational dryer speed and D_p is particle diameter. G and F are the air and solid mass flow rates, respectively.

It was assumed that internal mass transfer is the controlling mechanism. Diffusion model was adopted for drying kinetics calculation:

$$\frac{x_w^{out} - x_e}{x_w^{in} - x_e} = \frac{6}{\pi^2} \sum_{i=1}^{n_t} \frac{1}{n_t^2} e^{-\frac{4 n_t^2 D_{eff}}{D_p^2} \tau} \qquad (15)$$

here n_t is the considered number of terms and x_e is the equilibrium moisture content, determined by using experimental data for fitting the GAB equation and thus obtaining the desortion isotherm. Experimental data were obtained for soybean meal in a Novasina apparatus. Global mass and energy balances complete the set of equations for the drier model.

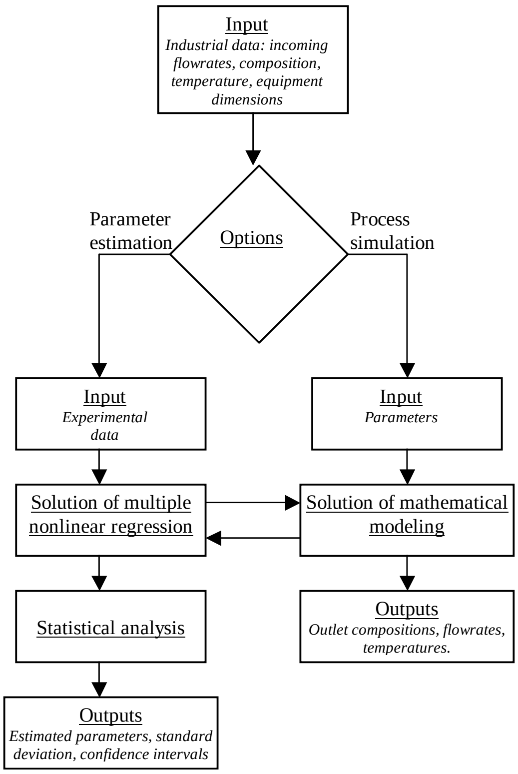

Fig.1. Structure of computational code.

B. Mathematical Solving

Each one of the developed models were solved by using adequate numerical methods: finite differences, Newton Raphson and Runge Kutta (Finlayson, 1980). The equation coefficients considered as parameters to be estimated were k, K, D_z and D_{eff}. Marquardt method was employed for parameter estimation (Constantinides and Mostoufi, 1999). The software used for programming the algorithms was Matlab 5.3 for Windows. The structure of the computational code is shown in **Figure 1.**

III.RESULTS AND DISCUSSION

The simulation program was applied to a specific industrial case. Several sets of operative variable values were collected in an industrial plant. Their average values and the equipment dimensions are shown in **Table 1**

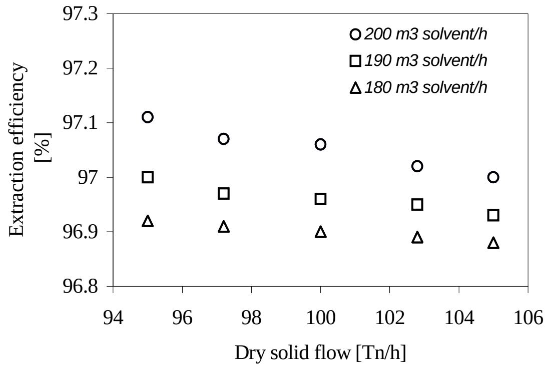

Fig.2. Effect of solid flow on extractor efficiency.

The estimated parameter values and the statistical analysis are presented in **Table 2.**

The simulation program with the appropriate parameter values can be employed in the analysis of the process variable influence. In this work a production increment is analyzed. If production increases so does the solid flow that enters to the extractor. Consequently the line equipment efficiency changes; so it is necessary to look for another set of operating conditions for improving equipment performance. The results obtained for the extractor efficiency with different solvent and solid flows are presented in **Figure 2.** It can be observed that the efficiency can be enhanced by using greater solvent flows.

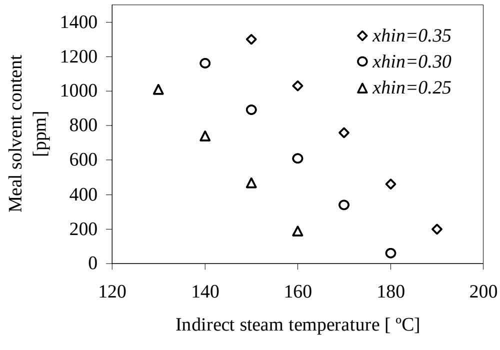

Fig.3. Final meal solvent content vs. indirect steam temperature. (*xhin* is hexane mass fraction in incoming solid phase)

This greater solvent flow causes an increment in solvent retention in the extractor. So, the desolventizer performance is affected. Indirect steam temperature can be modified for improving the desolventizing operation. Meal solvent content at the desolventizer outlet as a function of indirect steam temperature is shown in **Figure 3**, for different solvent contents of the material coming from the extractor.

The moisture content of the desolventized meal increases due to additional sparge steam condensation, as the required degree of desolventization increases. Also, the solid flow increment affects the drier performance. Indirect heat in the drier can be varied, within certain limits, for restoring the design values (see **Figure 4**).

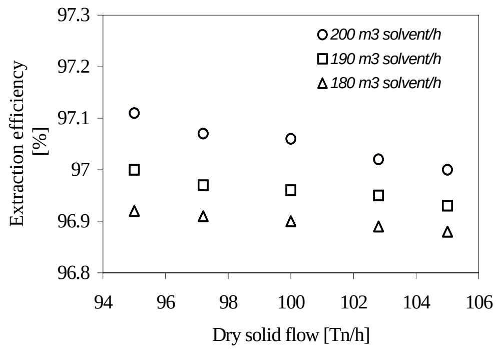

Fig.4. Effect of solid flow on outlet meal moisture content for different indirect heats.

Table 1. Industrial data.

Extractor		Desolventizer		Rotary Drier	
Length [m]	30.6	Height [m]	12.7	Length [m]	18.0
Wide [m]	3.2	Diameter [m]	4.5	Diameter [m]	2.8
Stage number	8	Tray number	9		
Dry meal flowrate [Tn/h]	98	Sparge steam flowrate [Kg/h]	20000	Air flowrate [Kg/h]	42000
Solvent flowrate [m^3/h]	188				
Average collets dimensions:					
Diameter [m]	0.012				
Length [m]	0.025				
	0.024	x_h^{in}	0.24	x_w^{in}	0.18
y_o^{out}	0.260	x_h^{out}	5 10^{-4}	x_w^{out}	0.13
x_o^{in}	0.200	T_v^{in} [ºC]	100	T_s^{in} [ºC]	95.7
x_o^{out}	0.005	T_v^{out} [ºC]	75	T_s^{out} [ºC]	83.8

Table 2. Estimated parameters and results of statistical analysis.

Parameter		Standard deviation	95% confidence interval for the parameters (test t Student)	
			Lower value	Upper value
Extractor				
k [kg/m^2 s]	1 10^{-3}	2.2 10^{-4}	5.6 10^{-4}	1.4 10^{-3}
K	0.55	8.1 10^{-2}	3.9 10^{-1}	7.1 10^{-1}
D_z [m^2/s]	6 10^{-3}	3.1 10^{-4}	5.4 10^{-3}	6.6 10^{-3}
Desolventizer				
$k.a_v$[molhex/m^3s]	9.8 10^{-4}	4.5 10^{-5}	8.9 10^{-4}	1.1 10^{-3}
K	1.1 10^{3}	2.3 10^{2}	6.4 10^{2}	1.5 10^{3}
D_z [m^2/s]	0.02	4.1 10^{-3}	1.2 10^{-2}	2.8 10^{-2}
Drier				
D_{eff} [m^2/s]	5 10^{-9}	4.7 10^{-10}	4.1 10^{-9}	5.9 10^{-9}

IV. CONCLUSIONS

A software for the simulation of the operations that deals with solids in the vegetable oil extraction process has been developed. The model parameters were calculated by using industrial data. This software can be adapted in different plants by determining for each case the equipment dimensions and the model parameters. The developed software has been run with industrial data which were collected in a soybean extraction facility and a good parameter estimation has been achieved with the proposed model. The model has yet to be validated in different operative conditions and in different plants.

The analysis of the results shows that the main variables that affect the extractor efficiency are operating temperature, solid material preparation characteristics and residence time. In the desolventizer, the more relevant variables that affect meal residual solvent content and moisture content are the sparge steam flow and the amount of indirect heat transferred. In the drier, the amount of moisture reduction can be varied by changing air conditions, especially temperature

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NOMENCLATURE

a_v :Specific direct heat and mass transfer area, m^2/m^3
A_t : Specific indirect heat transfer area, m^2/m^3
C_p : Specific heat, Kcal/kg ºC
D_{eff}:Effective diffusion coefficient, m^2/sec
D_z :Axial dispersion coefficient, m^2/sec
h : Film heat transfer coefficient, Kcal/m^2 sec ºC
H : Height, m
k : Mass transfer convective coefficient, kg/m^2 sec
K : Equilibrium constant
n : Flow density, kg/m^2 sec
Q : Specific transferred heat, Kcal/m^3 sec
t : Time, sec
T : Temperature, ºC
U :Overall indirect heat transfer coefficient, Kcal/m^2 sec ºC
v : Velocity, m/sec
x : Mass fraction in solid phase
y : Mass fraction in fluid phase
z : Axial coordinate, m
ε : Bed porosity
λ : Heat of vaporization, Kcal/kg
ρ: Density, kg/m^3
τ : Residence time, sec
subscripts:
h : hexano s : solid
m : miscella v : vapor
o : oil w : water

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