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Optimization of brewer's yeast spray drying process

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Abstract

A combination of simulation with experimental treatments of spray drying process were developed in order to minimize the energy resources required to obtain a production of spray dried brewer's yeast at given viability and moisture. The drying was optimized by combining a few affected experiments with a 1.2 m^3 spray dryer. Cost function, viability, output moisture and production, were related with process variables, both empirically and by simulation with a non-linear space state reported in literature. Both results were represented with response surface models (RSM). The empirical results state that the optimal operation was at 1 g grits (g yeast solids)$^{-1}$, 319 s^{-1}, spray rotor velocity, and 60 °C in output air. At these conditions it was obtained 6.86 kg of dried yeast h^{-1}, with a viability of 1.26×10^6 cfu g^{-1}, and 55.5 cost \$ kg^{-1} of product. Simulation results states optimal conditions at initial product moisture of 0.84 g g^{-1}, 214 °C at input air, 202 kg dry air h^{-1}, and 9.56 kg dry product h^{-1}. At these conditions the simulator predicts 10 kg dried yeast per hour with a viability of 1.00×10^6 cfu g^{-1}, and 26.7 cost kg^{-1} of product.

Keywords: Spray drying; Brewer's yeast; Optimization

1. Introduction

Spray drying was reported as an appropriate preservation method for yeast and other microorganisms (Kim & Bhowmik, 1990; Labuza, Le Roux, Fan, & Tannenbaum, 1970; Wan-Yin, Shing-Yi, & Mark, 1994). The effect of spray drying process variables over final moisture and biological or biochemical activity of dehydrated products was also amended (Labuza et al., 1970; Luybe, Liou, & Bruin, 1982; Ramírez, Salgado, Rodríguez, & García, 1998; Ré, 1998). More specifically, Luna, Salgado, García, and Rodréiguez (1998, 2000) reported how spray drying process variables and processing aids might improve dried brewer's yeast viability. They also showed that the dried yeast viability could be kept during storage at least 3 weeks. Recently, Luna-Solano, Salgado-Cervantes, Ramírez-Lepe, Garcia-Alvarado, and Rodríguez-Jimenes (2003) demonstrated that spray drying output temperature, rotor speed and aids concentration have effect over logistic growth specific velocity of re hydrated yeast, but have not effect over maximal growth. Therefore, spray drying is a feasible method for preserve viable yeast.

Spray drying is however a high energy demand operation, requiring at least 2500 J g^{-1} of evaporated water, in the ideal process for pure water evaporation, and up to 3 or 4 times greater values in the real processes. Despite spray drying should be optimized with respect to energy resource, several papers (Mudahar, Toledo, & Jen, 1990; Okello, Brennan, Lewis, & Gilmour, 1998; Ponciano, 1997) have so far dealt with an objective function different of energy resource. For instance, polyphenol oxidase spray drying (Okello et al., 1998), and potato fluidized bed drying (Mudahar et al., 1990)

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Nomenclature

a_w	water activity, –
a	specific surface, m^2 m^{-3}
A	external contact surface of drying chamber, m^2
C	specific total cost, \$ kg^{-1}
C_E	energy total cost, \$ kg^{-1}
C_{elec}	electric energy cost, \$ kW^{-1} h^{-1}
c_ω	specific energy cost for ω, \$ kg^{-1}
C_p	specific heat, J kg^{-1} K^{-1}
C_R	raw material cost, \$ kg^{-1}
D	mass diffusivity, m^2 s^{-1}
G	mass flow rate, kg dry matter h^{-1}
h	heat transfer coefficient, W m^{-2} K^{-1}
H	enthalpy, J kg^{-1}
I_ω	electric current for ω, A
k	heat conductivity, W m^{-1} K^{-1}
k_c	mass transfer coefficient, m s^{-1}
p	pressure, Pa
P	production, kg h^{-1}
R	droplet radio, m
T	temperature, °C
W	amount grits added, kg (kg dry matter)$^{-1}$
X	water content, kg water (kg dry matter)$^{-1}$
V	volume, m^3
E_ω	voltage applied to ω, V
$V_{\beta 1}$	output viability, cfu g^{-1}

Greek symbols

ε	volume fraction of continuous phase (air), –
ω	rotor velocity, s^{-1}
μ	viscosity, kg m^{-1} s^{-1}
ρ	volumetric concentration of dry matter, kg m^{-3}

Subscripts

air	referred to air
f	referred to fan
g	referred to grits
i	at dried product–air interface
j	at the end of a j ideal step
n	number of ideal mixed stages
0	at initial or inlet condition
out	to the exterior of drying chamber
p	referred to pump
r	referred to rotary disc
w	referred to water
wv	referred to water vapor
yc	referred to yeast cream

Greek subscripts

β	referred to dry matter in disperse phase (dried product)
γ	referred to dry matter in continuous phase (air)

Dimensionless groups

Sh	Sherwood number
Nu	Nusselt number
Re	Reynolds number
Sc	Schmidt number
Pr	Prandtl number

were optimized by maximizing the final product quality using the response surface methodology (RSM). Moreover, by using RMS, Ponciano (1997) determined the minimal drying time required to obtain dried garlic with given quality. The industrial engineering approach tends to minimize the energy resources required to obtain given quality and productivity, developed by Yunfei and Chengzhi (1996) in the care of food freeze drying. Use of response surface methodology to fit experimental drying treatments in optimization problems appears to be very useful when a mechanistic model is unavailable. Nevertheless, RSM may be applied to extract from use improved on mechanistic models to when these models are mathematically complex. For example, (Szitkai, Lelkes, Rev, & Fonyo, 2002) applied a regression model based on simulation results of pervaporation membrane networks to optimize ethanol dehydration systems. In the case of spray drying several mechanistic models representing the process dynamic have been proposed (Palencia, Nava, Herman, Rodríguez-Jimenes, & García-Alvarado, 2002; Pérez-Correa, 1995). In particular, the developed by Palencia et al. (2002) consists of a system of four differential equations coupled with three algebraic equations. Relation of input to output temperatures and moistures of the and feed streams air. These variables are directly related to energy and raw material consumption, dried product productivity and final moisture, but not to correlate to dried yeast viability. The aim of this work was to combine experimental results in a pilot plant spray drier to a series of simulated results using a well known mechanistic model, and minimize energy resources under prefixed limits involving yeast viability, moisture and productivity.

2. Function definition

The energy resources required to obtain a dried yeast product at a given moisture and viability may be defined in terms of cost function per unit of product defined in the following equations:

$$C = C_{\mathrm{E}} + C_{\mathrm{R}} \tag{1}$$

$$P = G_{\beta \mathrm{yc}}(1 + W_{\mathrm{g}})(1 + X_{\beta n}) \tag{2}$$

where

$$C_{\mathrm{E}} = c_{\mathrm{air}} + c_p + c_{\mathrm{r}} + c_{\mathrm{f}} \tag{3}$$

$$c_{\mathrm{air}} = G_\gamma (H_{\gamma 0} - H_{\gamma\,\mathrm{out}}) C_{\mathrm{elec}} / (3.6 \times 10^6 P) \tag{4}$$

$$c_p = E_p I_p C_{\mathrm{elec}} / (1000 P) \tag{5}$$

$$c_{\mathrm{r}} = E_{\mathrm{r}} I_{\mathrm{r}} C_{\mathrm{elec}} / (1000 P) \tag{6}$$

$$c_{\mathrm{f}} = E_{\mathrm{f}} I_{\mathrm{f}} C_{\mathrm{elec}} / (1000 P) \tag{7}$$

$$C_{\mathrm{R}} = \{ G_{\beta \mathrm{yc}} [(1 + X_{\beta \mathrm{yc}0}) C_{\mathrm{yc}} + W_{\mathrm{g}} C_{\mathrm{g}}] \} / P \tag{8}$$

Eq. (1) states that the unit total cost is the sum of the cost of energy consumed during drying process plus the cost of raw materials. The cost of yeast cream represent the energy resources (thermal energy, salaries and engines keeping) required to obtain it. Eq. (2) defines the dried yeast production. Eq. (3) states that the elements contained in energy cost are: the energy required for heating the air, the energy required for the feed pump, and the energy required for the spray rotor. Eqs. (4)–(8) describe these costs in terms of process variables. The air enthalpy is defined as

$$H_\gamma = C_{p_\gamma} T_\gamma + (H^0_{\mathrm{wv}} + C_{p_{\mathrm{wv}}} T_\gamma) X_\gamma \tag{9}$$

The correlation of the cost functions with respect of process variables was stated by two ways. One fully empirical, based in a RSM and experimental treatments; and the second as a combination of simulation with experimentation. The experimental procedure is described in the following sections.

3. Experimental procedure

Yeast cream (*Saccharomyces* sp.) as obtained from a brewer's industry, was used as raw material in this work. It had an initial viability of around 108 cfu (colony forming units) cm^{-3}, density of 1.030 $\mathrm{g\,cm}^{-3}$, and solid content of 14–18% (w/w). The experimental treatments in the spray drying were developed with dispersions of the yeast cream in water and grits. The amount of grits in dispersion was of 1, 0.75 and 0.5 g grits per g dry yeast solids, in accord with experimental design listed in Table 1. The addition of grits was suggested by Luna et al. (2000) as processing aids in order to keep the viability in the dried product.

A pilot plant scale spray drier Niro Atomizer with a 1.2 m^3 chamber and 5.5 m^2 of external area was used. The air flow is produced with 736 W centrifugal fan then was used assure a air rate constant of 674 $\mathrm{kg\,h}^{-1}$. The air heating system consists of a series of electrical resistances of 3, 4.5 and 9 kW respectively, thus allowing the air to be heated up to a maximum temperature of 145 °C. The feed spray system consisted of a 0.12 m disc rotating at 277–417 s^{-1} depending on the experimental treatment. The experimental design (often called phase center cube design) was a 2^3 factorial design implemented with six axial points and one central point. This experimental design was stated in order to obtain enough information for to relates variables by using polynomial models. The statistical significance of the process variables was demonstrated in a previous work (Luna-Solano et al., 2003).

Yeast dispersions were dried at conditions listed in Table 1. The output air temperature was fixed by controlling the yeast dispersion feed flow. The air input temperature was kept constant at 145 °C. These process variables were selected at the values in which Luna et al. (2000) reported the maximal residual viability for dried yeast. Yeast cream at input conditions (Table 1), and yeast dried moisture contents were determined by weight loss in a vacuum oven at 60 °C and 0.6 atm, and its water activity in an Aqualab CX2 (Deagon Devices) hygrometer at 25 °C. Yeast dried viability was evaluated by growth in inverted plate. Nutritive medium for yeast YPD (yeast extract, peptone and dextrose) containing 2.5 g of bactopeptone, 2.5 of yeast extract, 5 g glucose, and 250 cm^3 of distilled water, was used. The medium was sterilized at 120 °C by 15 min. The plates were incubated at 25 °C for 72 h.

4. Optimization relations

The cost function (Eq. 1) was calculated directly from experimental drying results described in the precedent section. The costs are states in terms of units (these units are based in México local money: 1 Unit=0.1 US$ in the first half of 2002 year). The cost used were

$$C_{\mathrm{yc}} = 1.5\$\mathrm{kg}^{-1} \quad C_{\mathrm{g}} = 2.0\$\mathrm{kg}^{-1}$$
$$C_{\mathrm{elec}} = 0.045\$\mathrm{kW}^{-1}\,\mathrm{h}^{-1}$$

Eqs. (6)–(8) were evaluated directly by measuring electrical current, with an amperemeter, of feed pump, spray rotor and fan. All of these accessories operate at 220 V. The feed flow was estimated from the volume changes in feed tank at a given time. The air flow with an anemometer in the air input duct.

The cost function, evaluated from Eq. (1) in the experimental treatments conditions showed in Table 1, was described using the following second order polynomial:

$$\begin{aligned} C = {} & \beta_0 + \beta_1 x_1 + \beta_{21} x_2 + \beta_3 x_3 + \beta_{12} x_1 x_2 + \beta_{13} x_1 x_3 \\ & + \beta_{23} x_2 x_3 + \beta_{11} x_1 x_1 + \beta_{22} x_2 x_2 + \beta_{33} x_3 x_3 \end{aligned} \tag{10}$$

Table 1
Experimental design of empirical spray drying of brewer's yeast

Treatment	Grits content (W_g)	Rotor velocity (s^{-1})	Out temperature (°C)	G_β ($kg\,h^{-1}$)	$X_{\beta 0}$ (kg water kg dm^{-1})
1	0.50	277	60	4.25	4.09
2	1.00	277	60	5.90	2.83
3	0.50	417	60	4.70	3.82
4	1.00	417	60	5.30	3.01
5	0.50	277	70	3.08	3.98
6	1.00	277	70	3.96	2.83
7	0.50	417	70	3.57	3.85
8	1.00	417	70	4.20	3.00
9	0.75	357	65	4.28	3.17
10	1.00	357	65	4.74	2.97
11	0.50	357	65	4.47	3.25
12	0.75	417	65	3.59	3.41
13	0.75	277	65	4.48	3.29
14	0.75	357	70	3.52	3.47
15	0.75	357	60	5.60	2.57

where x_1, x_2 and x_3 are the standardized values of the independent variables taken into accent:

$$x_1 = (W_g - 0.75)/0.25 \quad x_2 = (\omega - 347)/70$$
$$x_3 = (T_{\beta n} - 65)/5$$

The optimization problem, in this first case, was stated as the determination of process variables, W_g, ω, and $T_{\beta n}$, in which the cost function (10) is minimal, at a given production, moisture and viability. Therefore it is necessary to specify production (P), moisture ($X_{\beta 1}$) and viability ($V_{\beta 1}$) as a function of the same process variables. This relationship was obtained by fitting the moisture and viability results (Table 1) to a similar polynomial of Eq. (10). Then, the optimization problem was stated as

$$\min C = f_C(x_1, x_2, x_3) \tag{11}$$

subject to

$$-1 \leqslant x_1, x_2, x_3 \leqslant 1 \tag{12}$$

$$P = f_P(x_1, x_2, x_3) \geqslant 3.0 \tag{13}$$

$$V_{\beta 1} = f_V(x_1, x_2, x_3) \geqslant 1 \times 10^6 \tag{14}$$

$$X_{\beta 1} = f_X(x_1, x_2, x_3) \leqslant 0.12 \tag{15}$$

Constriction (12) states the fact that empirical model (9) cannot be used to extrapolate. Constriction (12) is a production that can be reached in the plant pilot spray dryer. Constriction (13) is the minimal viability admitted in dried yeast, and, constriction (14) is the maximal moisture content required to preserve the dried yeast during storage (Luna et al., 2000).

The minimization was carried out with the random search method called Complex algorithm for constriction functions (Box, 1965).

4.1. Simulation relationship

The fully empirical optimization described in the past sections has the advantage of obtain the data directly from experimental results. However it requires so many experimental treatments if we need to introduce more process variables like air input temperature and air flow. In such case the experimental design is increased in a power law. A very common engineering method to save experimental runs, is the simulation of the process. Pérez-Correa (1995), and Palencia et al. (2002) reported dynamic models for the description of spray drying process. Palencia et al. (2002) model is more general because it is in terms of heat and mass transfer coefficients and equilibrium relation for water among phases. This model is represented by

$$\frac{dX_{\beta j}}{dt} = -\frac{k_{c\beta} a (X_{\beta j} - X_{\beta ji})}{1 - \varepsilon} - \frac{G_\beta (X_{\beta j} - X_{\beta j-1})}{V_j \rho_\beta (1 - \varepsilon)} \tag{16}$$

$$\frac{dX_{\gamma j}}{dt} = \frac{k_{c\gamma} a (X_{\gamma ji} - X_{\gamma j})}{\varepsilon} - \frac{G_\gamma (X_{\gamma j} - X_{\gamma j-1})}{V_j \rho_\gamma \varepsilon} \tag{17}$$

$$\begin{aligned}\frac{dT_{\beta j}}{dt} = {} & \frac{h_\beta a (T_{ji} - T_{\beta j})}{(C_{p_\beta} + C_{p_w} X_{\beta j}) \rho_\beta (1 - \varepsilon)} - \frac{C_{p_w}}{C_{p_\beta} + C_{p_w} X_{\beta j}} \frac{dX_{\beta j}}{dt} \\ & - \frac{G_\beta [(C_{p_\beta} + C_{p_w} X_{\beta j}) T_{\beta j} - (C_{p_\beta} + C_{p_w} X_{\beta j-1}) T_{\beta j-1}]}{(C_{p_\beta} + C_{p_w} X_{\beta j})(1 - \varepsilon) \rho_\beta V_j}\end{aligned} \tag{18}$$

$$\begin{aligned}\frac{dT_{\gamma j}}{dt} = {} & \frac{h_\gamma a (T_{\gamma j} - T_{ji})}{(C_{p_\gamma} + C_{p_{wv}} X_{\gamma j}) \varepsilon \rho_\gamma} + \frac{k_{c\beta} a \rho_\beta (X_{\beta j} - X_{\beta ji}) \lambda}{(C_{p_\gamma} + C_{p_{wv}} X_{\gamma j}) \varepsilon \rho_\gamma} \\ & - \frac{H^0_{wv} + C_{p_{wv}} T_{\gamma j}}{C_{p_\gamma} + C_{p_{wv}} X_{\gamma j}} \frac{dX_\gamma}{dt} - \frac{h_{out} A_{out} (T_{\gamma j} - T_{out})}{(C_{p_\gamma} + C_{p_{wv}} X_{\gamma j}) \varepsilon \rho_\gamma V_{\gamma j}}\end{aligned}$$

$$\frac{G_\gamma}{(C_{p_\gamma}+C_{p_{\mathrm{wv}}}X_{\gamma j})\varepsilon\rho_\gamma V_j}\{[C_{p_\gamma}T_{\gamma j}+(H^0_{\mathrm{wv}}+C_{p_{\mathrm{wv}}}T_{\gamma j})X_{\gamma j}] - [C_{p_\gamma}T_{\gamma j-1}+(H^0_{\mathrm{wv}}+C_{p_{\mathrm{wv}}}T_{\gamma j-1})X_{\gamma j-1}]\} \quad (19)$$

$$X_\gamma - X_{\gamma i} - \frac{k_{\mathrm{c}\beta}\rho_\beta}{k_{\mathrm{c}\gamma}\rho_\gamma}(X_{\beta i}-X_\beta) = 0 \quad (20)$$

$$h_\gamma(T_\gamma - T_i) - h_\beta(T_i - T_\beta) - k_{\mathrm{c}\beta}\rho_\beta(X_\beta - X_{\beta i})\lambda = 0 \quad (21)$$

$$X_{\gamma i} = \frac{a_{\mathrm{w}}p_{\mathrm{w}}/p}{1-a_{\mathrm{w}}p_{\mathrm{w}}/p}\,\frac{18}{19} \quad \text{where } a_{\mathrm{w}} = f(X_{\beta i}, T_i) \quad (22)$$

For spherical particles the specific surface is (Geankoplis, 1993)

$$a = 6(1-\varepsilon)/R \quad (23)$$

The internal heat and mass transfer coefficient are in function of particle transport properties, and external ones of flow regimen. That is, for spherical particles (Palencia et al., 2002; Geankoplis, 1993)

$$k_{\mathrm{c}\beta} = \frac{\pi^2 D_{\mathrm{w}\beta}}{3R} \qquad h_\beta = \frac{\pi^2 k_\beta}{3R} \quad (24)$$

$$k_{\mathrm{c}\gamma} = \frac{Sh\,D_{\mathrm{w}\gamma}}{2R} \qquad h_\gamma = \frac{Nu\,k_\gamma}{2R} \quad (25)$$

$$Sh = 2 + 0.6Re^{0.53}Sc^{0.33} \qquad Nu = 2 + 0.6Re^{0.5}Pr^{0.33} \quad (26)$$

$$Re = \frac{2Rv\rho_\gamma}{\mu_\gamma} \qquad Sc = \frac{\mu_\gamma}{\rho_\gamma \mathrm{D}_{\mathrm{w}\gamma}} \qquad Pr = \frac{C_{p_\gamma}\mu_\gamma}{k_\gamma} \quad (27)$$

Eqs. (16) and (18) represent mass and heat balances for yeast suspension, Eqs. (17) and (19) represent the same balances for drying air. Eqs. (20)–(22) represent the interfacial phenomena between yeast suspension and drying air. The whole equations (16)–(27) represent the dynamic behavior of yeast spray drying in which the process is considered as a series of well mixed process. The subscript j in the state variables of Eqs. (16)–(19) indicates the output at any well mixed stage. Palencia et al. (2002) showed experimentally that the plant pilot spray drying dynamic was equivalent to a one well mixed stage. Then, the output temperature is equal (or approximately equal) to the temperature in the interior of dryer chamber. Solving these equations from an arbitrary initial state until a steady state was reached produces an operation state. They showed too that the process dynamic represented by Eqs. (16)–(27) is stable and has a unique solution if the initial state is in the neighborhood of positive feasible values of the four dependent variables. Some previous simulations were developed in order to state the heat, mass transfer and thermodynamic properties for which Eqs. (16)–(27) represent the experimental drying treatments of Table 1. A Runge-Kutta method was used. These previous simulations are described in the results section.

Cost function was calculated with Eq. (1) and (8) and simulation results. Although the relation between cost function and process variables are stated directly by the simulator (Eq. (16)–(27)), this type of relation requires long computer time if it is used in a random search method for optimization, like Complex. Therefore the relationship was obtained by fitting the simulator results to RSM models. In a similar way that Szitkai et al. (2002) used to pervaporate membrane networks model. Three new variables were added to the optimization problem: product and air flow, and air input temperature. The variable grits relation was changed through the input product moisture. The RSM models had the following general form:

$$\begin{aligned} C = {} & \beta_{01} + \beta_4 G_{\beta 0} + \beta_5 X_{\beta 0} + \beta_6 T_{\gamma 0} + \beta_7 G_{\gamma 0} \\ & + \beta_{45} G_{\beta 0} X_{\beta 0} + \beta_{46} G_{\beta 0} T_{\gamma 0} + \beta_{47} G_{\beta 0} G_{\gamma 0} \\ & + \beta_{56} X_{\beta 0} T_{\gamma 0} + \beta_{57} X_{\beta 0} G_{\gamma 0} + \beta_{67} T_{\gamma 0} G_{\gamma 0} \\ & + \beta_{44} G_{\beta 0} G_{\beta 0} + \beta_{55} X_{\beta 0} X_{\beta 0} + \beta_{66} T_{\gamma 0} T_{\gamma 0} \\ & + \beta_{77} G_{\gamma 0} G_{\gamma 0} \end{aligned} \quad (28)$$

As a direct result of the well mixed characteristic of the process the viability was considered only in terms of output temperature, rotor velocity and initial grits content. Then, the viability relation was the same than those used in the fully empirical optimization. Output temperature and output yeast moisture content were obtained from simulation results and related with process variables with an empirical model similar to Eq. (28).

The optimization problem was stated similarly to Eqs. (11)–(15), with the variables $G_{\beta 0}$, $X_{\beta 0}$, $T_{\gamma 0}$, and $G_{\gamma 0}$ instead of x_1, x_2, and x_3. The lower and upper limits for these variables were stated as simulator results, and therefore were shown in result section. Simulations were developed at product flow ($G_{\beta 1}$) between 2.25 and 10.25 kg dry matter h^{-1}; input product moisture ($T_{\gamma 1}$) between 0.84 and 5.84 kg water (kg dry matter)$^{-1}$, input air temperature ($T_{\gamma 1}$) between 145 and 280 °C; and air flow ($G_{\gamma 1}$) between 75 and 2400 kg dry air h^{-1}. In the initial moisture content a same relation of yeast dry matter: grits dry matter that those obtained in fully empirical optimization, was assumed. Therefore, the initial product moisture depends upon of the initial yeast cream moisture ($X_{\beta\mathrm{yc}0}$) and of the amount of grits added (W_{g}) by the relation

$$X_{\beta 0} = X_{\beta\mathrm{yc}0}/(1+W_{\mathrm{g}}) \quad (29)$$

5. Fully empirical optimization

The experimental yeast cream spray drying at different conditions is listed in Table 2. Previous researches (Luna et al., 2000) had reported similar residual viability results at similar conditions of drying and grits concentrations.

Table 2
Empirical yeast spray drying results at conditions listed in Table 1

Treatment	T_{out} (°C)		X_β (g g^{-1})		Viability (cfu g^{-1})	Production (kg h^{-1})	Cost (kg^{-1})
	Experimental	Simulation	Experimental	Simulation			
1	60	60.7	0.093	0.0040	4.03×10^6	4.815	71.625
2	60	61.5	0.118	0.0039	6.88×10^6	6.935	57.060
3	60	62.0	0.104	0.0055	8.13×10^5	5.705	70.455
4	60	63.6	0.077	0.0028	2.19×10^6	5.240	62.130
5	70	71.4	0.052	0.0090	1.88×10^6	3.395	78.325
6	70	73.6	0.051	0.0007	4.17×10^6	4.345	61.405
7	70	69.6	0.063	0.0014	9.01×10^5	3.835	72.690
8	70	70.7	0.057	0.0010	1.00×10^6	4.480	62.815
9	65	69.3	0.067	0.0013	8.18×10^5	4.835	64.180
10	65	67.5	0.051	0.0014	1.85×10^6	5.215	63.190
11	65	67.1	0.084	0.0018	3.91×10^5	5.045	63.840
12	65	68.5	0.053	0.0014	1.66×10^6	4.490	69.530
13	65	65.0	0.084	0.0019	1.00×10^7	4.870	65.385
14	70	72.3	0.040	0.0008	2.05×10^5	3.735	71.745
15	60	67.1	0.105	0.0018	4.80×10^5	6.465	55.620

At the final moisture obtained it was reported that dried yeast may be stable during 3 month in storage at temperature lower than 25 °C (Luna et al., 1998).

The empirical models obtained from these results are

$$\begin{aligned} C = {} & 63.776 - 5.085x_1 + 0.382x_2 + 3.0328x_3 \\ & + 1.8214x_1x_2 - 0.4881x_1x_3 - 0.832x_2x_3 \\ & - 0.324x_1x_1 + 3.75x_2x_2 - 0.156x_3x_3 \end{aligned} \tag{30}$$

$$\begin{aligned} P = {} & 4.99 + 0.3525x_1 - 0.061x_2 - 0.9415x_3 \\ & - 0.369x_1x_2 - 0.0075x_1x_3 + 0.1566x_2x_3 \\ & + 0.119x_1x_1 - 0.347x_2x_2 + 0.0889x_3x_3 \end{aligned} \tag{31}$$

$$\begin{aligned} \log(V_{\beta 1}) = {} & 5.785 + 0.1754x_1 - 0.306x_2 - 0.1415x_3 \\ & + 0.0072x_1x_2 - 0.0136x_1x_3 \\ & + 0.0107x_2x_3 + 0.084x_1x_1 + 0.7812x_2x_2 \\ & - 0.3462x_3x_3 \end{aligned} \tag{32}$$

$$\begin{aligned} X_{\beta 1} = {} & 0.066773 - 0.003847x_1 - 0.004295x_2 \\ & - 0.023554x_3 - 0.007455x_1x_2 \\ & - 0.00054375x_1x_3 + 0.0053833x_2x_3 \\ & + 0.001477x_1x_1 + 0.001874x_2x_2 \\ & + 0.0067277x_3x_3 \end{aligned} \tag{33}$$

Eqs. (30)–(33) had a generalized correlation coefficient with respect to experimental results greater than 0.9. There is not statistical analysis for equations parameters (β_{ij}) because the objective of this model is the representation of the data with a mathematically simple model, which can be used easily in the complex algorithm for optimization. In previous works (Luna et al., 1998, 2000), it had been shown that the variables involved have significant effect over viability.

A complete interpretation of the variables effect over cost, production, viability and moisture is not simple because some interactions are present. In fact, the use of Eqs. (30)–(33) in a Complex algorithm of random search avoids the necessity of an interpretation. However a general perspective of the responses behavior with respect to the variables is shown in Fig. 1. It is possible to appreciate that cost function is inverse to residual viability function. This is practical because the variable values that predict a greater viability, predict smaller cost. It is evident that an increase of temperature produces a decrease in residual viability. This is expected because the yeast cells have thermal sensitivity. Other researchers had reported the viability loss at spray drying higher output temperatures in yeast and other microorganisms (Adamiec & Strumillo, 1998; Daemen & Van der Stege, 1982; Labuza et al., 1970; Wan-Yin et al., 1994). With respect to aspersion speed, at higher values smaller residual viabilities were obtained. Luna et al. (1998) reported similar results, but Labuza et al. (1970) did not found an effect of aspersion speed over viability. However, they dried the yeast at speeds greater than 417 s^{-1}. If we considerate that the aspersion speed has effect due to shear stress during spray process, this effect may be asymptotic. The results showed that residual viability was increased at higher solid content. At greater solid content the drop is greater and therefore the shear stress is smaller. Luna et al. (2000) found the greater residual viabilities at the higher initial solid concentration. Production of dried yeast is increasing at smaller output temperatures. This is expected because at smaller output temperatures, greater final moisture, and therefore in agreement with Eq. (2), the production is increasing. This may have some problem in the process, however the constriction (15) in optimization definition avoids that moisture could be greater than a given value.

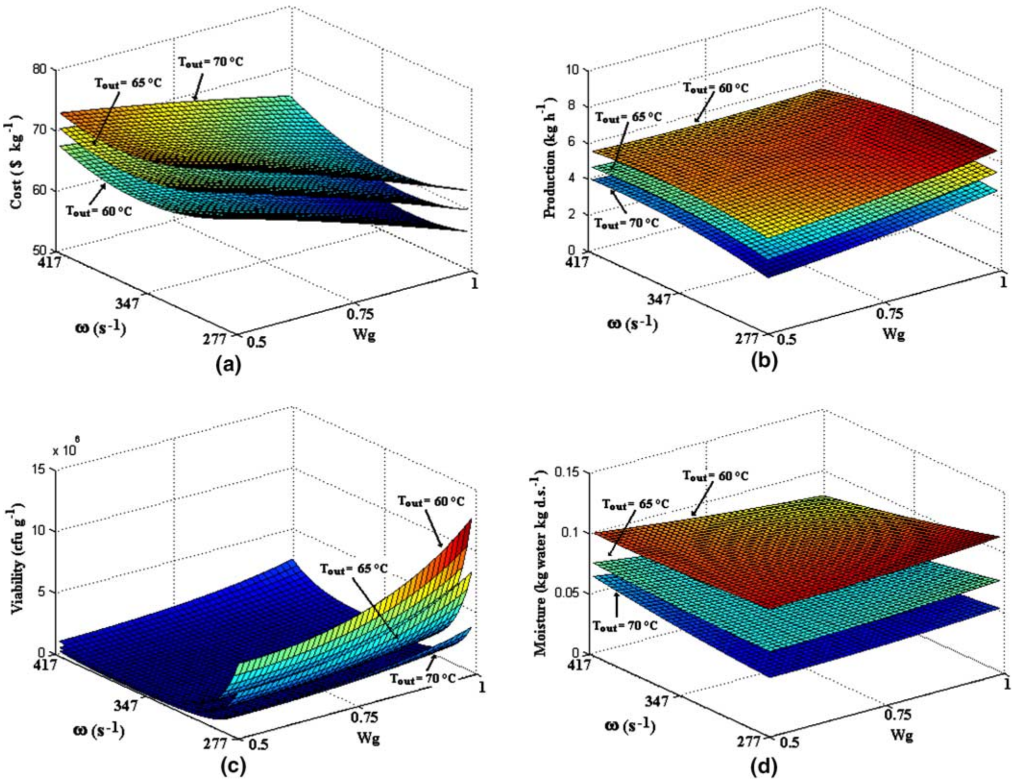


Fig. 1. General perspective of the responses: (a) cost, (b) production, (c) viability and (d) moisture with respect to the variables.

The optimization results of the problem stated in Eq. (11) and constrictions (12)–(15) were

$$x_1 = 0.9999 \quad x_2 = -0.398 \quad x_3 = -0.9999$$

and with the following cost, viability, production and moisture

$$C = 55.5\$\text{kg}^{-1} \quad V_{\beta 1} = 1.26 \times 10^{6}\ \text{cfu}\,\text{g}^{-1}$$

$$P = 686\ \text{kg h}^{-1}$$

6. Optimization by simulation

Eqs. (16)–(27) solved by Runge-Kutta method was used for predict the spray drying of yeast cream at different conditions. These equations will be called simulator in the rest of manuscript. The fact that the simulator reproduces the experimental behavior at different conditions depends of the adequate selection of thermophysical properties for air and yeast. Air properties are well known (Perry & Chilton, 1973) and they are listed in Table 3. However there is not reported the thermophysical properties required for dried and cream yeast. In the case of thermal properties: specific heat, and heat conductivity the general expression reported for foods (Mujumdar, 1995) may be used. These thermal properties are listed too in Table 3. The ε value (0.96) was calculated by the experimental dynamic behavior reported by Palencia et al. (2002) for the same plant pilot dryer. The droplet radius was evaluated with the expression suggested by Perry and Chilton (1973). The average value obtained was $R = 0.0005$ m.

A series of previous simulations were developed at different values of water mass diffusivity in the interior of yeast drops ($D_{w\beta}$) and external heat transfer coefficient (h_{out}). h_{out} was selected in a manner that simulator reproduces the experimental output temperature of Table 2 conditions. The best value obtained is listed in Table 3. Fig. 2 show the output air temperature and product moisture at different $D_{w\beta}$. The base values for diffusivity (listed in Table 3) were taken from the values reported for various foods products. It is evident that the diffusivity between 10^{-11} and 10^{-10} m^{2}s^{-1} has not a great effect over responses. The moistures obtained are significantly slower than experimental ones (Table 2) and therefore it is not necessary to use diffusivities greater than 10^{-10} m^{2}s^{-1}. On the other hand, diffusivities slower than 10^{-11} m^{2}s^{-1} are not found in literature for porous powder particles. Therefore the moisture difference must be explained in terms of product re-hydration during experimental treatments. The experimental moistures were evaluated from the powder collected in the separation unit after an air exposition. Like the air in our location has a relative humidity greater than 80%, the dry powder is fine and hot (closer than output

Table 3
Thermophysical properties used for simulator

Property	Value or expression	Reference
a_w	$1-\exp\{-\exp[-57.651]T^{10.41}X^{0.6942}\}$	Obtained in our laboratory
C_{p_γ}	$1000\ \mathrm{J\,kg^{-1}\,K^{-1}}$	Geankoplis (1993)
$C_{p_{wv}}$	$1608.92\ \mathrm{J\,kg^{-1}\,K^{-1}}$	Geankoplis (1993)
C_{p_β}	$1657\ \mathrm{J\,kg^{-1}\,K^{-1}}$	Mujumdar (1995)
C_{p_w}	$4185\ \mathrm{J\,kg^{-1}\,K^{-1}}$	Geankoplis (1993)
$D_{w\beta}$	$6.67\times10^{-10}\ \mathrm{m^2\,s^{-1}}$	Mujumdar (1995)
H_{wv} (at 273 K)	$2501\times10^3\ \mathrm{J\,kg^{-1}}$	Geankoplis (1993)
k_β	$0.1418+0.00493\ X_\beta/(1+X_\beta)\ \mathrm{W\,m^{-1}\,s^{-1}}$	Mujumdar (1995)
k_γ	$8.4044\times10^{-5}\ (T+273.15)+4.63\times10^{-5}\ \mathrm{W\,m^{-1}\,s^{-1}}$	Geankoplis (1993)
μ	$4.25\times10^{-8}\ (T+273.15)+5.87\times10^{-6}\ \mathrm{kg\,m^{-1}\,s^{-1}}$	Geankoplis (1993)
ρ_β	$800\ \mathrm{kg\,m^{-3}}$	Estimated
ε	0.96	Palencia et al. (2002)
h_{out}	$8.7\times10^4\ \mathrm{J\,m^{-2}\,K^1\,h^{-1}}$	Fitted

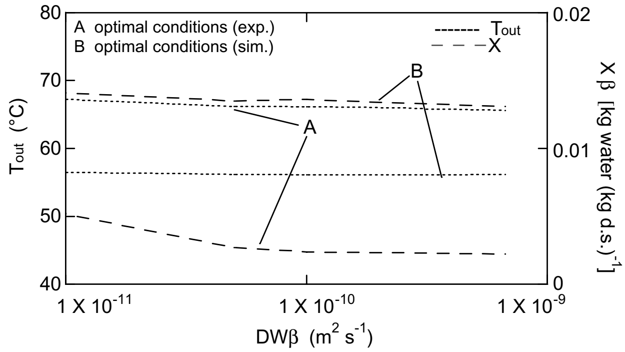
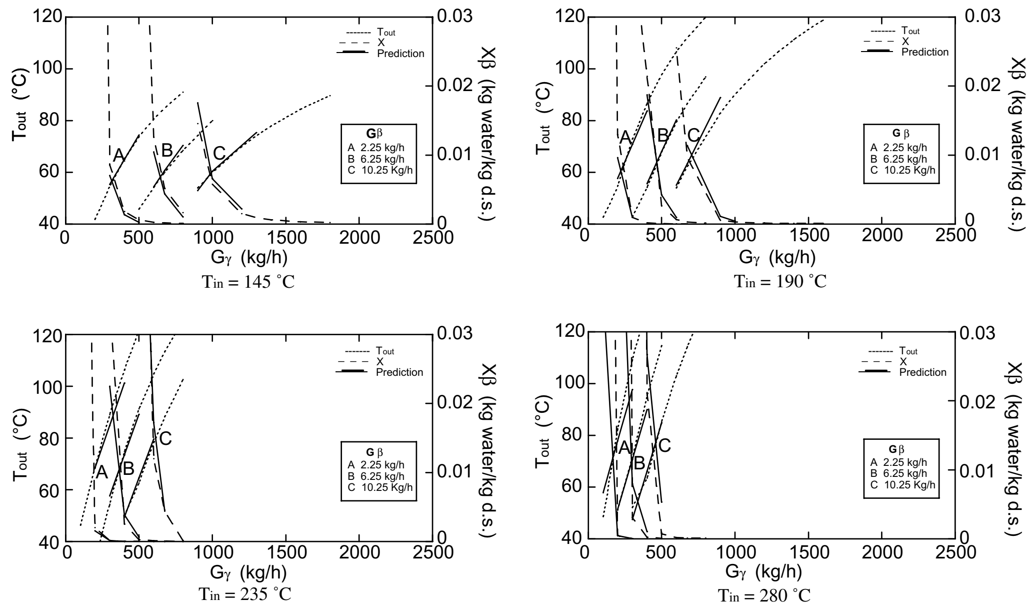


Fig. 2. Output air temperature and output product moisture results from simulation at different water diffusivities. Drop radio 0.00001 m; (A) $X_{\beta0}=2.84$, $G_\beta=5.9\ \mathrm{kg\,h^{-1}}$, $G_\gamma=674\ \mathrm{kg\,h^{-1}}$, $T_{\gamma0}=145$ °C; (B) $X_{\beta0}=0.84$, $G_\beta=9.56\ \mathrm{kg\,h^{-1}}$, $G_\gamma=202\ \mathrm{kg\,h^{-1}}$, $T_{\gamma0}=214$ °C.

temperature), and its moisture is slower than $0.005\ \mathrm{g\,g^{-1}}$, exists a significant re-hydration before the moisture evaluation.

The lack of sensibility of output air temperature and product output moisture with respect to water diffusivity suggest that the 1.2 m^3 dryer chamber is over specified for the air and product flow. This over specification of install capacity may be desirable because produces that the process would be robust with respect to outlet moisture of the product. Then, for the spray dryer used the outlet air temperature must be defined by h_{out}. The simulation results obtained are listed in Table 2. For temperature an average deviation of 3.6% between experimental and simulated were obtained.

Fig. 3 shows some of the simulator results. The RSM obtained from these results were



Tin = 145 °C

Tin = 190 °C

Tin = 235 °C

Tin = 280 °C

Fig. 3. Simulation and prediction for output air temperature and product moisture at different conditions. Broken lines represent the simulator results, and continuous lines represent Eqs. (35) and (36).

$$\begin{aligned} C = {} & 84.41 - 1.42x_1 + 14.94x_2 + 44.4x_3 + 12.65x_4 \\ & + 0.27x_1x_2 - 9.5x_1x_3 + 11.5x_1x_4 + 1.0x_2x_3 \\ & + 15x_2x_4 + 20.3x_3x_4 + 0.4x_1x_1 + 0.89x_2x_2 \\ & - 2.3x_3x_3 - 19.4x_4x_4 \end{aligned} \tag{34}$$

$$\begin{aligned} T_{\gamma 1} = {} & 169.8 - 34.4x_1 + 69.15x_2 - 39.2x_3 + 148.7x_4 \\ & + 39x_1x_2 - 19.5x_1x_3 - 6.1x_1x_4 + 0.1x_2x_3 \\ & + 59x_2x_4 - 5.3x_3x_4 + 1.9x_1x_1 - 1.44x_2x_2 \\ & + 1.28x_3x_3 - 9.75x_4x_4 \end{aligned} \tag{35}$$

$$\begin{aligned} \log(X_{\gamma 1}) = {} & -5.13 + 2.58x_1 - 2.21x_2 - 0.01x_3 \\ & - 5.82x_4 + 0.23x_1x_2 - 0.4x_1x_3 \\ & + 1.82x_1x_4 + 0.14x_2x_3 - 1.84x_2x_4 \\ & + 0.99x_3x_4 - 0.50x_1x_1 + 0.14x_2x_2 \\ & - 2.37x_3x_3 + 0.94x_4x_4 \end{aligned} \tag{36}$$

$$\begin{aligned} P = {} & 5.11 + 3.6x_1 - 1.55x_2 - 0.13x_3 - 1.01x_4 \\ & + 0.003x_1x_2 + 1.03x_1x_3 - 1.9x_1x_4 - 0.015x_2x_3 \\ & - 1.67x_2x_4 - 1.6x_3x_4 + 0.22x_1x_1 - 0.09x_2x_2 \\ & - 0.04x_3x_3 + 2.04x_4x_4 \end{aligned} \tag{37}$$

where

$$\begin{aligned} & x_1 = (G_{\beta 0} - 6.25)/4 \quad x_2 = (T_{\gamma 0} - 212.5)/67.5 \\ & x_3 = (X_{\beta 0} - 3.34)/2.5 \quad x_4 = (G_{\gamma 0} - 1235.5)/1162.5 \end{aligned} \tag{38}$$

Eq. (35) and (36) are plotted in Fig. 3. It is evident that empirical equations may be used in order to avoid the simulator resolution in each optimization step.

The optimal operation point for the problem defined in Eq. (11)–(15) was

$$\mathrm{X}_{\beta 0} = 0.84 \quad \mathrm{G}_{\beta 0} = 9.56\ \mathrm{kg\,h^{-1}} \quad \mathrm{G}_{\gamma 0} = 202\ \mathrm{kg\,h^{-1}}$$
$$\mathrm{T}_{\gamma 0} = 214\ ^{\circ}\mathrm{C}$$

and with the following cost, output temperature, output product moisture, viability, and production

$$\mathrm{C} = 26.7\$\mathrm{kg^{-1}} \quad \mathrm{X}_{\beta 1} = 0.013\ \mathrm{g\,g^{-1}} \quad \mathrm{T}_{\gamma 1} = 59\ ^{\circ}\mathrm{C}$$

$$\mathrm{V}_{\beta n} = 1.00 \times 10^{6}\ \mathrm{cfu\,g^{-1}} \quad \mathrm{P} = 10\ \mathrm{kg\,h^{-1}}$$

The whole of these results at the same grits: yeast solid relation and rotor speed reported as optimal in Section 5.

7. Conclusions

Powder brewer's yeast has application in beer industry, and therefore it is important increase its production efficiency. In this paper, a methodology that combines empirical with mechanistic modeling jointly with RSM, was proposed for the solving of brewer's yeast drying optimization problem. The variables optimal values reported are particulars for the stated constrictions, but the models are general for yeast spray drying. Therefore new optimal values may be obtained for other particular requirements, by using the same models.

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