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# Scale up proposal for the spray drying process of hebernem-s bionematicide

# Propuesta de escalado del proceso de secado por aspersión del bionematicida hebernem-s

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## ABSTRACT

The need for robust spray dryer scale-up methods is essential to control critical quality attributes in the biopharmaceutical industry. In this second part, various parameters are determined such as the heat capacity of the cream (value of 3.626 kJ/kg.ºC), the enthalpies and mass flowrates of the different streams involved, the Sauter diameter (88.78 µm), the drying total time (6.90 s) and the efficiencies. A mathematical model was also obtained by using MATLAB® software, in order to determine the main phenomenological characteristics of the evaluated spray drying process. In general, the increase in the cream feeding temperature, the powder temperature, the cream dry mass and the rotation speed of the atomizer decrease the final humidity of the powder, while the air inlet temperature and the absolute air humidity increase it. The temperature of the powder at the exit of the cyclone and the rotation speed of the atomizer disk are the variables that most influence the scale-up of spray drying. The new mathematical model obtained in this work quantitatively describes the spray-drying process of HeberNem-S® bionematicide.

**Keywords:** Sauter diameter; Efficiencies; Scale-up; Spray drying; Mathematical model; Enthalpies.

## RESUMEN

La necesidad de métodos robustos de escalado de secadores por aspersión es esencial para controlar atributos de calidad críticos en la industria biofarmacéutica. En esta segunda parte se determinan varios parámetros tales como capacidad calórica de la crema (valor de 3.626 kJ/kg.ºC), las entalpías y los flujos másicos de las diferentes corrientes involucradas, el diámetro Sauter (88,78 µm), el tiempo total de secado (6,90 s) y las eficiencias. También se obtuvo un modelo matemático mediante el empleo del software MATLAB®, con el fin de determinar las principales características fenomenológicas del proceso de secado por aspersión evaluado. De forma general, el incremento de la temperatura de alimentación de la crema, la temperatura del polvo, la masa seca de la crema y la velocidad de rotación del atomizador disminuyen la humedad final del polvo, mientras que la temperatura de entrada del aire y la humedad absoluta del aire la incrementan. La temperatura del polvo a la salida del ciclón y la velocidad de rotación

del disco atomizador son las variables que más influyen en el escalado del secado por aspersión evaluado en este estudio. El nuevo modelo matemático obtenido en este trabajo describe cuantitativamente el proceso de secado por aspersión del bionematicida HeberNem-S®.

**Palabras claves:** Diámetro Sauter; Eficiencias; Escalado; Secado por aspersión; Modelo matemático; Entalpías.

## NOMENCLATURE

| Symbol | Description | Unit |
|---|---|---|
| $a$ | Constant | - |
| $b$ | Constant | - |
| $c$ | Constant | - |
| $Cp_{as}$ | Heat capacity of dry air | kJ/kg.ºC |
| $Cp_{bact}$ | Heat capacity of *Tsukamurella paurometabola* bacterium | kJ/kg.ºC |
| $Cp_{c}$ | Heat capacity of cream | kJ/kg.ºC |
| $Cp_{h}$ | Heat capacity of liquid water | kJ/kg.ºC |
| $Cp_{va}$ | Heat capacity of water vapor | kJ/kg.ºC |
| $d$ | Diameter of the atomizer disc | m |
| $D_{vs}$ | Sauter diameter of the droplet | µm |
| $E_{f}$ | thermal efficiency of the drying operation | % |
| $E_{ev}$ | Evaporation efficiency | % |
| $E_{tg}$ | Overall thermal efficiency | % |
| $E_{tm}$ | Maximum thermal efficiency | % |
| $h_{a}$ | Height of the atomizer disc | m |
| $h_{a0}$ | Enthalpy of the air at the dryer inlet | kJ/kg |
| $h_{af}$ | Enthalpy of the air at the dryer outlet | kJ/kg |
| $h_{c}$ | Enthalpy of the cream at the inlet of the drying chamber | kJ/kg |
| $h_{hv}$ | Enthalpy of vaporization at the outlet of the drying chamber | kJ/kg |
| $h_{p}$ | Enthalpy of the powder at the outlet of the drying chamber | kJ/kg |
| $K$ | Constant | - |
| $K_{g}$ | Thermal conductivity of air | W/m.K |
| $m_{sc}$ | Dry matter of the cream | % |
| $m_{sp}$ | Dry matter of the powder | % |
| $M_{bact}$ | Molar weight of the bacterium | kg/kmol |
| $M_{H2O}$ | Molar weight of water | kg/kmol |
| $M_{aire}$ | Molar weight of air | kg/kmol |
| $n$ | Number of atomizer blades | - |
| $N$ | Synchronous speed of the atomizing disc | rpm |

| | | |
|---|---|---|
| $P_t$ | Ambient pressure of the room | Pa |
| $Q_{ma}$ | Mass flowrate of inlet air | kg/h |
| $Q_{maa}$ | Mass flowrate of water entering with air | kg/h |
| $Q_{ma0}$ | Mass flowrate of air at the dryer inlet | kg/h |
| $Q_{maf}$ | Mass flowrate of air at the dryer outlet | kg/h |
| $Q_{mas}$ | Mass flowrate of dry air | kg/h |
| $Q_{mc}$ | Feed flowrate of cream | kg/h |
| $Q_{mhv}$ | Mass flowrate of evaporated water | kg/h |
| $Q_{mp}$ | Outlet mass flowrate of powder | kg/h |
| $Q_{ms}$ | Mass flowrate of water in the exhaust air | kg/h |
| $Q_{va}$ | Volumetric flowrate of air entering the dryer | $m^3$/h |
| $t_d$ | Drying time in the decreasing speed period | s |
| $t_s$ | Drying time in the constant speed period | s |
| $t_t$ | Total drying time | s |
| $T_{ac}$ | Air temperature at the critical point | ºC |
| $T_c$ | Cream inlet temperature | ºC |
| $T_{ea}$ | Air inlet temperature | ºC |
| $T_L$ | Ambient temperature of the room | ºC |
| $T_{sa}$ | Air outlet temperature | ºC |
| $T_{sat}$ | Air saturation temperature | ºC |
| $T_{sc}$ | Temperature of the solid at the critical point | ºC |
| $T_{sp}$ | Powder outlet temperature | ºC |
| $T_{we}$ | Wet bulb temperature of inlet air | ºC |
| $\Delta T_{ml2}$ | log mean temperature for the decreasing speed period | K |
| $V_h$ | Wet volume | $m^3_{.mixture}/kg_{.dry\ air}$ |
| $Wc$ | Critical humidity | % |
| $We$ | Final humidity | % |
| $Y_{abs}$ | Absolute humidity of ambient air | kg de agua/kg de aire seco |
| | **Greek symbols** | |
| $\lambda$ | Latent heat of vaporization of water | kJ/kg.ºC |
| $\rho_c$ | Cream density | $kg/m^3$ |
| $\rho_p$ | Powder density | $kg/m^3$ |

## 1. INTRODUCCIÓN

Spray drying is currently one of the most interesting technologies in the biopharmaceutical industry. The process shows a remarkable ability to manipulate particle/powder attributes such as size, morphology, density, and residual solvent level. This flexibility leads to its application in a wide variety of powder formulations and advanced forms of solids: from very fine powders for inhalation to large particles for direct compression, and from solid dispersions for bioavailability to microcapsules for drug protection or controlled release (Gil *et al*., 2010).

Spray drying has evolved into a mature technique for industrial scale production of up to several tons of product per day. It represents a feasible one-step method for the production of biopharmaceutical formulations with solid particles of unique characteristics (Ameri & Maa, 2006).

Because of its reproducibility, continuous mode of operation, high capacity, fast drying ability, and short product exposure to elevated processing temperatures associated with rapid evaporative cooling, this process has become very popular in the pharmaceutical and biotechnological industries (Poozesh & Bilgili, 2019).

Regardless of the numerous advantages that spray drying presents and its promising future, discrepancies in product quality across different scales has been the biggest hurdle in the development and manufacturing of this drying operation. This is primarily due to the fact that spray drying involves a combination of complex physical transformations that are multi-scale and multi-physical in nature. These include mixing of the microorganism, active pharmaceutical ingredient or excipients in a suitable solvent, atomization of the liquid, evaporation of the solvent (drying), gas-particle separation, usually with secondary drying by fluidized bed or vacuum driers. In the first stage, the liquid stream (which can be a solution, suspension, or emulsion) is sprayed in a drying chamber, by increasing the surface/volume ratio, thus increasing heat and mass transfer. Meanwhile, due to differences in the vapor pressure of the droplet and the gas, the solvent component of the droplets is gradually transferred to the gas stream. The particles formed during evaporation collide with the wall of the drying chamber or transit to the cyclone or baghouse filters where they are separated from the gas. Cyclones work based on the differences between the hydrodynamically induced forces between the drying gas and the particles, while filters use the concept of impact towards a filter or electrostatic precipitation. The collected powders can go to a secondary drying stage to remove the solvent amount to the desired levels depending on the doses to be considered for human or animal use, and physical stability (Poozesh & Bilgili, 2019). Spray dryers can operate in an open cycle mode (where the drying gas is passed through the chamber and vented to the appropriate waste stream) or in a closed loop system (where the drying gas is typically recycled via a condenser or washing unit) (Gil *et al*., 2010). The dimensions of the drying chamber determine the scale of the dryer. The higher the chamber, the longer the residence time of the droplets and the larger the particles that can be produced (Ameri & Maa, 2006).

A key factor in any drying process is the atomization system, since it controls the droplet size and therefore the particle size (Gil *et al*., 2010). In spray drying, the final particle size distribution is usually directly proportional to the droplet size distribution created by the atomizer (Thybo *et al*., 2008). The most widely atomization systems used today are rotary, pressure, two-fluid, and ultrasonic (Gil *et al*., 2010).

For rotary atomizers, atomization is achieved by centrifugal energy transmitted to the liquid stream by means of a disk or wheel rotating at high speeds (from 5,000 to 50,000 rpm). The liquid is fed at the center of the rotating wheel, moves to the center of the wheel under centrifugal force, and is disintegrated at the edge of the wheel to form droplets. These types of atomizers can be used to spray slurries, suspensions, or high-viscosity solutions. Apart from feed properties, the operating variables that influence droplet size are flow rate, rotational speed, wheel diameter, and design, and are capable of producing droplets in a wide range of sizes, from 20 to 200 μm (Gaspar *et al*., 2014).

The optimization of the spray drying process involves the evaluation of the parameters related to the process, as well as the properties of the liquid formulation to be fed to the dryer (Thybo *et al*., 2008).

A particular concern that arises during the development of spray-dried materials is the influence of scale-up on critical quality attributes. Furthermore, a less careful scale-up strategy can lead to significant losses of very expensive materials and compromise the timelines of a clinical program. In this sense, mathematical modeling is a very powerful tool to support the scale-up of spray-drying processes (Gil *et al*., 2010).

The scale-up of a spray drying process involves increasing powder production while maintaining product quality compared to the smaller scale process. The product production rate is governed by three factors: liquid feed flow rate, solid content of the solution/suspension formulation, and product recovery efficiency (Ameri & Maa, 2006). In addition, the use of a larger atomizer is often required to carry out the scale-up procedure (Thybo *et al*., 2008).

An obvious challenge for the direct application of an industrial scale spray dryer to carry out development studies is that such dryers may be economically impractical due to high material costs (Ameri & Maa, 2006).

Laboratory scale spray dryers are particularly useful for producing small quantities of prototype formulations in the early stages of development. They can process small amounts of solution (as low as 2 mL) with relatively high throughputs. Not surprisingly, units on this scale have been used to produce commercial quantities of very low volume products. A typical feature of these small-scale systems is that the drying chamber and cyclone are made of glass, allowing a privileged view of the drying and separation processes. However, the main limitation of laboratory units is the powder properties of the resulting materials, i.e. particle size. The small dimensions of the drying chambers limit the residence time, and therefore the droplets need to be smaller in order to be completely dried before leaving the drying chamber, or they may collide with the walls. Therefore, most small units produce powders with mean particle sizes below 10 μm, more often between 3 and 10 μm (Gaspar *et al*., 2014).

Pilot and commercial scale spray dryers are suitable for a wide range of batch sizes, from less than 1 kg to several metric tons. They share many similarities concerning configuration, materials of construction, ability to handle most organic solvents, and level of automation. Other features such as cleaning or drying gas recirculation may be included, as well as operation under vacuum, although they commonly operate under slightly positive pressure. Despite these options and similarities, the main difference between the scales is the evaporative capacity and productivity. In this case, the evaporation capacity depends fundamentally on the flow rate of the drying gas, solvent type, temperature profile (inlet and outlet temperature) and heat losses from the spray drying unit (Gaspar *et al*., 2014).

The scale-up of spray drying processes has been motivated mainly on experimental experience. Among the reasons for this are the complexity of the drying process (characterized by simultaneous and rapid heat and mass transfer between the droplets and the drying gas), and the unavailability of model parameters, often not easily measured. On the other hand, the whole process is extremely dependent on the properties of the feed, the production scale and the equipment design (Gil *et al*., 2010). In most cases, they offer a more efficient process from an energy point of view, less labor per kilogram produced, and greater flexibility in adjusting and optimizing process attributes. In this way, the selection of the ideal scale for a given product will be dictated by projections of commercial demand and the actual productivity of the process (Gaspar *et al*., 2014).

When scaling up a spray drying process, several areas must be analyzed since the complete drying process is made up of the following stages (Thybo *et al*., 2008):

1) Atomization
2) Mixing of droplets with the drying gas.
3) Drying (kinetics and residence time).
4) Separation.

Ideally, all four criteria should be related in order to maintain geometric, kinematic, and dynamic similarity of the process when carrying out the scale-up procedure.

If necessary, a spray drying process can be scaled up directly from laboratory scale to final production scale. However, the quantities required to carry out clinical trials are produced more efficiently on a pilot scale, where product losses and the risk of scale up are markedly lower. Therefore, these intermediate production scales are commonly used during the development of the scale-up process (Poozesh & Bilgili, 2019).

Fundamental engineering methodologies and models for the scale-up of spray drying processes can be found in the current literature. In this sense, (Berman *et al*., 1994) carried out the macroscopic application of the Laws of Conservation of Mass and Energy in steady state, to facilitate the scale-up of spray dryers, thus obtaining an equation that describes the productivity (spray rate) of the solution fed based on five factors, while the theoretical equation obtained equals favorably with the experimental data collected at pilot and production scales. Other authors (Thybo *et al*., 2008) evaluated the performance of two fluid nozzles of similar design but with different capacities, by measuring the droplet size distributions through laser diffraction, to perform the scale-up of spray dryers from pilot to industrial scale. Several liquid properties and system considerations were evaluated in this study in order to measure the droplet size distribution, showing that the two fluid nozzle atomization was dependent on the following variables: atomization gas flowrate, liquid flowrate, liquid surface tension, orifice diameter, and atomizing gas outlet area. The study showed that it is possible to scale from a liquid flow rate of 2.5 kg/h for the pilot scale to 45 kg/h for the production scale, while achieving the same droplet size distribution. Also (Thybo *et al*., 2008) evaluated the influence of five parameters to scale up a pilot-scale spray dryer to an industrial-scale one. The five parameters to be controlled were gas outlet temperature, feed flow rate, solid concentration and the drug/polymer ratio of the feeding solution, and the open/closed cycle. The parameters to be measured were the particle size, morphology, residual solvent, crystallinity, yield and stability. The results obtained in this study showed that the scale-up process based on matching the atomizer droplet size distributions was not satisfactory for the applied formulations, although it was possible to scale-up the process by varying the operating conditions. On the other hand, (Dobry *et al*., 2009) described a novel flowsheet methodology based on fundamental engineering models and advanced process characterization techniques, which ensures that the scale-up of spray-drying processes is efficient and requires a minimum time. The methodology presented by these authors offers substantial advantages compared to traditional methods, which are often empirical, require large amounts of material and longer development times. In addition, (Gil *et al*., 2010) described a pragmatic scale-up methodology for spray drying processes based on simulation models with scientific principles and process characterization techniques, more specifically based on thermodynamic, fluid dynamics and drying kinetic considerations, with the aim of ensuring that the scale-up is as direct and predictable as possible. Likewise, (Sosnik & Seremeta, 2015) discussed two main challenges in technology transfer regarding differences in product performance and particle size distributions between the different scales evaluated. These authors suggested solving these differences by using a laboratory scale B-290 spray dryer. This was established, in part, due to reports of successful scale-up results using this equipment as the small-scale dryer, taking into account its ability to capture very fine particles in order to minimize productivity losses during the separation process. In (Singh & Mooter, 2015) several formulations and process variables that influence critical quality attributes of

amorphous solid dispersions were reviewed, while briefly discussed the various challenges presented by the scale-up process. They also suggested that the change from one scale to another usually comprises adjustments in: (1) feeding system; (2) type, location, and condition of the feeding device; (3) drying gas dispensing system; (4) dimensions of the chamber; (5) exhaust gas duct and; (6) change from an open-loop system (found in the laboratory) to a closed-loop one (seen on pilot and production scales). In a recent study (Bellinghausen, 2019) suggested qualitatively that successful scale-up depends on controlling the flow configuration (cocurrent or countercurrent flow), atomization, particle separation, and the drying process. To facilitate the scale-up, this author suggested: (1) control the particle size distribution through the use of a rotary, ultrasonic or mono-dispenser atomizer in order to produce the same droplet sizes; (2) to control the drying process either by creating a laminar or counter current flow; and (3) use "flow curtains" to prevent particles from flowing into the wall regions. Similarly, (Al-Khattawi *et al*., 2017) mentioned industrial scale-up methods, offering to match key parameters between scales, including: (1) feed properties and feed solvent content; (2) size distribution of the atomized droplets; (3) the drying history of the droplet; (4) the desired droplet/particle collision history to form agglomerates; and (5) avoid contact and accumulation on the wall at all scales. These authors enunciated several basic scale-up methods such as thermodynamic (Dobry *et al*., 2009), design of experiments based on statistics (Lebrun *et al*., 2012) and fundamental models including computational methods (Li & Zbiciński, 2005).

The complexity of the physical transformations that occur in a spray dryer makes the scale-up a challenging process and requires a fundamental understanding of fluid mechanics, heat and mass transfer, and thermodynamics (Poozesh & Bilgili, 2019).

In the first part of this study (Diaz et al., 2023), the spray drying process to obtain HeberNem-S product was described; while the average diameter and area of *T. paurometabola* cells in the formulated cream before the drying step; the cream viability at different temperatures and times; the diameter of HeberNem-S powder particles after spray drying; the stability of the HeberNem-S product at temperatures near ambient without applying vacuum conditions; and the adsorption isotherms of the HeberNem-S product were determined. In this second part, the spray drying process to obtain the HeberNem-S® product is phenomenologically characterized, specifically regarding the calculation of the heat capacity of the cream, the enthalpies and the mass flowrates of the different streams involved, the Sauter diameter, drying times and efficiencies of the spray drying process. Finally, a mathematical model obtained through the use of MATLAB® software is presented, while several scale-up alternatives for the spray drying process according to existing technology are simulated, in order to determine how the increase of six operating parameters (cream feeding temperature, powder temperature, cream dry mass, air inlet temperature, air absolute humidity, rotation speed of atomizing disc) influences on the final moisture percentage of the powder.

## 2. MATERIALS AND METHODS

### 2.1. Phenomenological characterization of the drying process

By carrying out a study of the spray drying fundamentals and the transport phenomenon at the level of the droplet, it will be possible to fully understand the process. This study was carried out using an appropriate calculation sequence that allowed the solution of the different algebraic equations. Thus, the drying was characterized in a phenomenological way, thus defining the properties of the dried product.

*2.1.1. Mass and energy balances:* To carry out the mass and energy balances, the following initial data are available:

- Air inlet temperature ($T_{ea}$) = 130 ºC.

- Cream inlet temperature ($T_c$) = 37 ºC.
- Powder outlet temperature ($T_{sp}$): 60 ºC.
- Air temperature at the critical point ($T_{ac}$) = 120 ºC.
- Temperature of the solid at the critical point ($T_{sc}$) = 25 ºC.
- Cream dry matter ($m_{sc}$) = 0.22 m/m.
- Powder dry matter ($m_{sp}$) = 0.92 m/m.
- Heat capacity of liquid water ($Cp_h$): 4.182 kJ/kg.ºC (Green & Southard, 2019).
- Heat capacity of water vapor ($Cp_{va}$): 1.857 kJ/kg.ºC (Green & Southard, 2019).
- Heat capacity of dry air ($Cp_{as}$): 1.005 kJ/kg.ºC (Green & Southard, 2019).
- Latent heat of vaporization of water ($\lambda$): 2,502 kJ/kg (Green & Southard, 2019).
- Wet bulb temperature of inlet air ($Tw_e$): 26.98 ºC.
- Air outlet temperature ($T_{sa}$): 63 ºC.
- Absolute humidity of ambient air ($Y_{abs}$) = 0.019 kg $H_2O$/kg dry air.
- Ambient temperature of the room ($T_L$): 25 ºC.
- Ambient pressure of the room ($P_t$): 101,325 Pa.
- Molar weight of water ($M_{H2O}$): 18 kg/kmol
- Molar weight of air ($M_{air}$): 29 kg/kmol.
- Volumetric flowrate of air entering the dryer ($Q_{va}$): 520 $m^3$/h.
- Thermal conductivity of air ($K_g$): 0.03318 (Green & Southard, 2019).
- Cream density ($\rho_c$) = 1,035 kg/$m^3$.
- Powder density ($\rho_p$) = 806 kg/$m^3$.
- Critical humidity ($Wc$) = 33%.
- Final humidity ($We$) = 8 %.
- Air saturation temperature ($T_{sat}$) = 26.98 ºC.

*2.1.2. Calculation of the cream heat capacity:* The heat capacity of the cream was calculated using the correlation reported by (Felder & Rousseau, 2004), where it is proposed to estimate the heat capacity of a compound from Kopp's rule, based on knowledge of its structural formula. For this, the law of mixtures is used as the sum of the products and the individual contribution of each constituent element. These contributions (heat capacities) are found as a function of the state of molecular aggregation for the different compounds at room temperature, which are summarized in Table 1:

Table 1. Data to estimate the heat capacity of *Tsukamurella paurometabola* C-924 bacterium

| Element | Contribution (kJ/kmol.ºC) | |
|---|---|---|
| | Solid | Liquid |

| | | |
|---|---|---|
| C | 7.5 | 12 |
| H | 9.6 | 18 |
| N | 26 | 33 |
| O | 17 | 25 |

Source: (Felder & Rousseau, 2004).

According to (Hu, 2018), the empirical molecular formula of a bacterium is the following: $CH_2N_{0.26}O_{0.45}$, therefore the molecular coefficients will be:

$\alpha$ = 1
$\beta$ = 2
$\gamma$ = 0.26
$\sigma$ = 0.45

Then, the molecular weight of the bacterium, in units of kg/kmol, will be determined according to the following equation:

$$M_{bact} = \left[\left(\alpha \cdot M_C\right) + \left(\beta \cdot M_H\right) + \left(\gamma \cdot M_N\right) + \left(\sigma \cdot M_O\right)\right] \quad (1)$$

Where:

$M_C$ - Molar weight of carbon = 12.01 kg/kmol.

$M_H$ - Molar weight of hydrogen = 1.01 kg/kmol.

$M_N$ - Molar weight of nitrogen = 14.01 kg/kmol.

$M_O$ - Molar weight of oxygen = 16 g/mol.

The heat capacity of the bacteria, in units of kJ/kg.ºC, is calculated according to the following equation:

$$Cp_{bact} = \frac{\alpha \cdot (7.5) + \beta \cdot (9.6) + \gamma \cdot (26) + \sigma \cdot (17)}{M_{bact}} \quad (2)$$

Finally, the heat capacity of the cream is calculated through the following equation:

$$Cp_c = m_{sc} \cdot Cp_{bact} + (1 - m_{sc}) \cdot Cp_h \quad (3)$$

*2.1.3. Calculation of enthalpies:*

Enthalpy of the air at the dryer outlet ( $h_{af}$ ):

$$h_{af} = Cp_{as} \cdot T_{sa} + Y_{abs} \cdot \left(Cp_{va} \cdot T_{sa} + \lambda\right) \quad (4)$$

Enthalpy of the air at the dryer inlet ( $h_{a0}$ ):

$$h_{a0} = Cp_{as} \cdot T_{ea} + Y_{abs} \cdot \left(Cp_{va} \cdot T_{ea} + \lambda\right) \tag{5}$$

Cream enthalpy at the inlet of the drying chamber ( $h_c$ ):

$$h_c = T_c \cdot \left[\left(m_{sc} \cdot Cp_c\right) + \left(1 - m_{sc}\right) \cdot Cp_h\right] \tag{6}$$

Powder enthalpy at the outlet of the drying chamber ( $h_p$ ):

$$h_p = T_{sp} \cdot \left[\left(m_{sp} \cdot Cp_c\right) + \left(1 - m_{sp}\right) \cdot Cp_h\right] \tag{7}$$

Enthalpy of vaporization at the outlet of the drying chamber ( $h_{hv}$ ):

$$h_{hv} = Cp_{va} \cdot T_{ea} + \lambda \tag{8}$$

*2.1.4. Calculation of mass flowrates:* The following are the general equations in accordance with a mass balance around the system without the occurrence of losses or accumulation:

$$Q_{mc} \cdot m_{sc} = Q_{mp} \cdot m_{sp} \tag{9}$$

$$Q_{mc} \cdot h_c + Q_{ma} \cdot h_{a0} = Q_{mp} \cdot h_p + Q_{maf} \cdot h_{af} + Q_{mhv} \cdot h_{hv} \tag{10}$$

$$Q_{ma} = Q_{ma0} = Q_{maf} \tag{11}$$

Wet volume (Ángeles, 2009) ( $V_h$ ):

$$V_h = 8{,}315 \cdot \left(\frac{1}{M_{air}} + \frac{Y_{abs}}{M_{H2O}}\right) \cdot \frac{\left(T_L + 273\right)}{P_t} \tag{12}$$

Mass flowrate of dry air ( $Q_{mas}$ ):

$$Q_{mas} = \frac{Q_{va}}{V_h} \tag{13}$$

Mass flowrate of water entering with air ( $Q_{maa}$ ):

$$Q_{maa} = Q_{mas} \cdot Y_{abs} \tag{14}$$

Feed flowrate of cream ($Q_{mc}$):

$$Q_{mc} = \frac{Q_{mas} \cdot \left(h_{af} - h_{a0}\right)}{h_c - \left[\left(\frac{m_{sc}}{m_{sp}}\right) \cdot h_p\right] - \left\{\left[1 - \left(\frac{m_{sc}}{m_{sp}}\right)\right] \cdot h_{hv}\right\}} \tag{15}$$

Outlet mass flowrate of powder ($Q_{mp}$):

$$Q_{mp} = \frac{Q_{mc} \cdot m_{sc}}{m_{sp}} \tag{16}$$

Mass flowrate of evaporated water ($Q_{mhv}$):

$$Q_{mhv} = Q_{mc} - Q_{mp} \tag{17}$$

Mass flowrate of water in the exhaust air ($Q_{ms}$):

$$Q_{ms} = Q_{mhv} + Q_{maa} \tag{18}$$

*2.1.5. Sauter Diameter:* The Sauter diameter ($D_{vs}$) is defined as the diameter of a droplet that has the same surface-volume ratio as that of all the droplets formed by the spray of particles (Masters, 1991).

There are various correlations that allow predicting the characteristics of the sprayed cloud obtained by rotary atomizers, which depend on the peripheral speed of the disc, feed flowrate, density, diameter of the atomizing disc, among others (Ángeles, 2009).

The following equation is used to determine the Sauter diameter of droplets sprayed with small diameter rotary atomizers (5 to 20 μm) and high peripheral velocities (Masters, 1991).

$$D_{vs} = \frac{14{,}000 \cdot Q_{mc}^{a}}{N^{b} \cdot d^{0.6} \cdot \left(n \cdot h_a\right)^{a}} \tag{19}$$

The synchronous speed of the atomizing disk ($N$) is applied depending on the atomizing disk and the specific characteristics of the fluid to be sprayed. The diameter ($d$) and the number of blades ($n$) of the rotating disk are constant for each dryer, and depend on its construction parameters. The height of the atomizing disc ($h_a$) is in correspondence to the physical construction characteristics of the dryer.

Table 2 shows the coefficients to be used in equation (19), which depend on the synchronous speed of the atomizer disc (Masters, 1991).

Table 2. Values of the constants *a*, *b*, *c* and *K* depending on the synchronous speed of the atomizing disc.

| Synchronous speed (rpm) | Classification | Constants | | | |
|---|---|---|---|---|---|
| | | *a* | *b* | *c* | *K* |
| 25 | Low | 0.24 | 0.82 | 0.24 | 1.14 |
| 250-1,500 | Normal | 0.2 | 0.8 | 0.2 | 1.6 |
| 1,000-3,000 | Normal High | 0.12 | 0.77 | 0.12 | 1.25 |
| 3,000-60,000 | Very High | 0.12 | 0.7 | 0.12 | 1.2 |

Source: (Masters, 1991).

In the case of the spray dryer evaluated in this work, the following data is available:

- Synchronous speed of the atomizing disc ( $N$ ) = 25,000 rpm.
- Number of atomizer blades ( $n$ ) = 8.
- Height of the atomizer disc ( $h_a$ ) = 0.012 m.
- Diameter of the atomizer disc ( $d$ ) = 0.10 m.

*2.1.6. Drying time:* As previously mentioned, there are two drying periods, that is, the constant speed and the decreasing speed periods. The drying time in the constant rate period can be derived from a heat balance over the sprayed droplet by assuming a dynamic equilibrium, where the heat transfer rate is equal to the mass transfer rate multiplied by the latent heat of vaporization (Masters, 1991), thus obtaining the following final equation:

$$h_c \cdot A \cdot \Delta T_{ml1} = -\lambda \cdot \frac{dm_d}{dt} \tag{20}$$

Thus, the drying time in the constant speed period ( $t_s$ ) is determined according to the following equation:

$$t_s = \frac{\rho_c \cdot h_{hv} \cdot 1{,}000 \cdot \left( \frac{D_{vs}}{1{,}000{,}000} \right)^2}{8 \cdot K_g \cdot \left( T_{ea} - T_{we} \right)} \tag{21}$$

The log mean temperature for the decreasing speed period ( $\Delta Tml_2$ ) is given by:

$$\Delta Tml_2 = \frac{\left( T_{ea} - T_{sc} \right) - \left( T_{sa} - T_{sp} \right)}{\ln \frac{\left( T_{ea} - T_{sc} \right)}{\left( T_{sa} - T_{sp} \right)}} \tag{22}$$

The drying time in the decreasing speed period ($t_d$), considering that the relative speed is negligible, is defined as:

$$t_d = \frac{h_{hv} \cdot \rho_p \cdot \left( \frac{D_{vs}}{1{,}000{,}000} \right)^2 \cdot 1{,}000}{12 \cdot K_g \cdot \Delta T_{ml2} \cdot (Wc - We)} \tag{23}$$

Where $Wc$ and $We$ have values of 0.33 and 0.08, respectively.

Therefore the total drying time ($t_t$) will be:

$$t_t = t_s + t_d \tag{24}$$

In the case of this study, the droplets to be dried contain insoluble solids, which normally occur when biological fluids are processed.

*2.1.7. Spray dryer efficiencies:* Spray drying is a type of unit operation with high energy consumption and relatively low energy utilization, so the problem of energy saving has attracted the attention of researchers and engineers (Cheng *et al*., 2018).

Energy efficiency is usually considered as a cumulative parameter, calculated from the initial-final or inlet-outlet data. For batch drying, the energy efficiency is given as an average value over a drying time, while for continuous drying the energy efficiency is averaged over the range of moisture content or the dryer length. However, energy efficiency depends on the moisture content of the material at any instant because thermal energy is used not only for the evaporation of free moisture, but also for the removal of bound moisture, eventually breaking the material-moisture bonds (Strumiłło *et al*., 2006).

According to (Selvamuthukumaran *et al*., 2020), spray drying is an intense process from an energy point of view, where thermal efficiency is relatively low because most of the supplied heat is lost with the discharge gases.

The thermal efficiency of the drying operation ($E_f$) is defined as the ratio between the heat used for evaporation and the total heat available if the exhaust air is saturated (Gauvin & Katta, 1976), in such a way that:

$$E_f = \frac{h_c}{h_{a0}} \cdot 100 \tag{25}$$

Other efficiencies that are determined for a spray drying process are shown below (Strumiłło *et al*., 2006).

Overall thermal efficiency ($E_{tg}$):

$$E_{tg} = \frac{(T_{ea} - T_{sa})}{(T_{ea} - T_{sat})} \cdot 100 \tag{26}$$

Evaporation efficiency ( $E_{ev}$ ):

$$E_{ev} = \frac{\left(T_{ea} - T_{sa}\right)}{\left(T_{ea} - T_{L}\right)} \cdot 100 \tag{27}$$

Maximum thermal efficiency ( $E_{tm}$ ):

$$E_{tm} = \frac{\left(T_{ea} - T_{sat}\right)}{\left(T_{ea} - T_{L}\right)} \cdot 100 \tag{28}$$

From the above equations, it can be observed that the thermal efficiency of spray dryers can be increased by increasing the temperature of the drying gas and operating the dryer at a minimum exhaust gas temperature. However, this might require operating the drying tower at a higher feed rate, which could lead to more depositions on the wall. Operating the spray drying tower at a higher gas inlet temperature could also make the product more susceptible to thermal degradation and increase heat loss from the equipment. Efficiency can also be improved by minimizing heat losses to the surroundings (Ali, 2014).

2.2. Mathematical model of the spray-drying process of the bacterium *Tsukamurella paurometabola* C–924

The mathematical model focused on simulating the spray drying process, having the cream inlet temperature, the synchronous speed of the atomizing disc, the air inlet temperature, the cream dry matter and the powder dry matter as the main inlet parameters. This was done on the basis of the phenomenological equations that represent the spray drying process, which were explained in previous sections.

These parameters were combined and simulated in MATLAB® 7.10.0 software, through a programming code that allows the user to enter the real parameters of a spray dryer (the aforementioned variables) while returning the most important operating data of the evaluated spray dryer, such as feed mass flowrate, powder mass flowrate, atomized particle diameter, total drying time, and powder moisture. The necessary data for a psychometric chart were also introduced in the programming code, with the objective of calculating the wet bulb and dry bulb temperatures by the program, which are also necessary for the calculations.

Figure 1 shows the illustration of the spray dryer used in this study (Bylund, 1995), where the atomizer is of rotary type, the drying air and the feed cream enter through the upper part of the drying chamber (in cocurrent flow), so that a homogeneous temperature profile is obtained inside the chamber in tangential form.

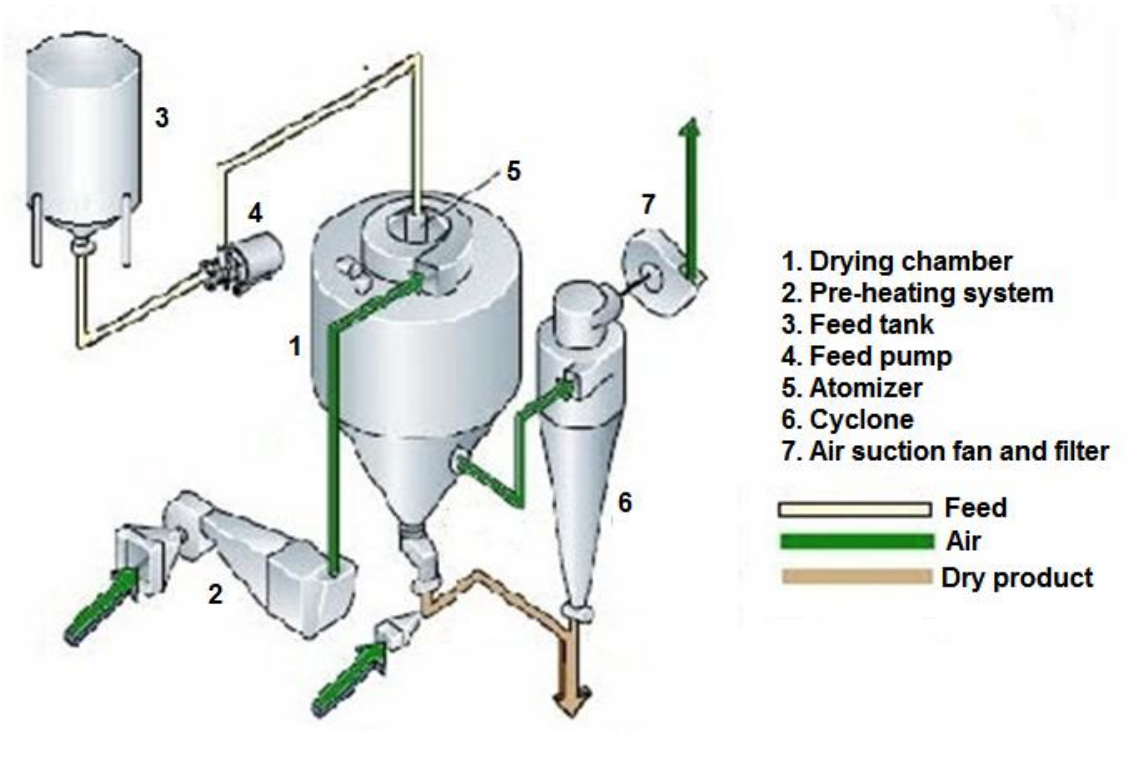



Figure 1. Diagram of the spray dryer used in this study.
Source: Adapted from (Bylund, 1995)

The modeling algorithm follows the logic of the calculations developed in the design of each of the spray dryer elements. It is worth mentioning that there is a subprogram for each of the equipment involved, which works independently. The code, elaborated in MATLAB®'s own language, was programmed using the "*Guide*" tool.

2.3. Scale-up criteria for spray drying according to existing technology

To simulate the scale-up of a spray dryer, the most important variables that have a direct influence on the final humidity of the powder were identified, which will be the response parameter and the criterion to be analyzed in this study, since the main principle of the spray dryer is to eliminate the residual humidity of the product to values below 12% (CIGB, 1992). In this case, only a partial study was carried out on the spray dryer, since the powder viability was not included as a response parameter. To analyze the proposed scale-up criterion, six alternatives were considered in order to determine the influence of each of them on the final moisture of the powder.

- Alternative 1: The feed temperature of the cream was varied from 10 to 100 °C while the air inlet temperature, the cream dry mass, the air mass flowrate, the cream mass flowrate, the cream feed temperature and the powder dry mass were kept constant.

- Alternative 2: The temperature of the powder was varied from 10 to 100 °C while keeping constant the cream dry mass, the air mass flowrate and the cream feed mass flowrate.

- Alternative 3: To vary the cream dry mass from 5 to 50% while keeping constant the cream feed temperature, the powder temperature, the air inlet temperature, the powder dry mass, the cream feed mass flowrate and the air inlet mass flowrate.

- Alternative 4: To vary the air inlet temperature from 50 to 150 °C, and keeping constant the cream dry mass, the air inlet mass flowrate, the cream feed mass flowrate and the powder dry matter constant.

- Alternative 5: To vary the absolute humidity of the air from 0.005 to 0.0250 kg $H_2O$/kg.dry air, and keeping constant the powder temperature, the air inlet temperature, the cream dry mass, the powder dry mass, the air inlet mass flowrate and the cream feed mass flowrate.

- Alternative 6: To vary the synchronous speed of the atomizing disk from 5,000 to 50,000 rpm and keep constant the temperature of the cream, the powder and the inlet air, the dry mass of the cream and the powder, the cream feed mass flowrate and air inlet mass flowrate.

## 3. RESULTS AND DISCUSSION

The results of the different parameters calculated to carry out the scale-up proposal of the spray dryer are shown below.

### 3.1. Heat capacity of the cream

Table 3 shows the values of the parameters molar weight and heat capacity of the bacteria *Tsukamurella paurometabola* C-924, as well as the heat capacity of the cream, determined through equations (1) to (3).

Table 3. Results of the molar weight and heat capacity of *Tsukamurella paurometabola* C-924 bacterium, and the heat capacity of the cream.

| Equation | Parameter | Symbol | Value | Unit |
|---|---|---|---|---|
| (1) | Molar weight of bacterium | $M_{bact}$ | 24.87 | kg/kmol |
| (2) | Heat capacity of bacterium | $Cp_{bact}$ | 1.653 | kJ/kg.ºC |
| (3) | Heat capacity of the cream | $Cp_c$ | 3.626 | kJ/kg.ºC |

Source: Own elaboration.

The molar weight of the bacteria had a value of 24.87 kg/kmol, while the heat capacity of the bacteria *Tsukamurella paurometabola* C-924 and the cream was 1.653 kJ/kg.ºC and 3.626 kJ/kg.ºC, respectively. In (Popovic *et al*., 2021) the molecular weight and standard heat capacity of five microorganisms were reported, which are the Gram-negative bacteria *Escherichia coli* (molecular weight of 27.039 kg/kmol and heat capacity of 1.413 kJ/kg.K), *Gluconobacter oxydans* (28.789 kg/kmol and 1.882 kJ/kg.K) and *Pseudomonas fluorescens* (26.264 kg/kmol and 1.198 kJ/kg.K), the Gram positive bacterium *Streptococcus thermophilus* (28.482 kg/kmol and 1.338 kJ/kg.K), and the penicillin-producing fungus *Penicillium chrysogenum* (25.425 kg/kmol and 1.406 kJ/kg.K).

### 3.2. Enthalpies

Table 4 presents the calculated enthalpy values for the different streams considered.

Table 4. Results of the enthalpies calculated for the different streams considered.

| Equation | Parameter | Symbol | Value | Unit |
|---|---|---|---|---|

| | | | | |
|---|---|---|---|---|
| (4) | Enthalpy of the air at the dryer outlet | $h_{af}$ | 113.08 | kJ/kg |
| (5) | Enthalpy of the air at the dryer inlet | $h_{a0}$ | 182.77 | kJ/kg |
| (6) | Cream enthalpy at the inlet of the drying chamber | $h_c$ | 150.21 | kJ/kg |
| (7) | Powder enthalpy at the outlet of the drying chamber | $h_p$ | 220.23 | kJ/kg |
| (8) | Enthalpy of vaporization at the outlet of the drying chamber | $h_{hv}$ | 2,743.41 | kJ/kg |

Source: Own elaboration.

The enthalpy of the air at the inlet and outlet of the dryer had values of 182.77 kJ/kg and 113.08 kJ/kg, respectively, while the enthalpies of the cream and the powder were 150.21 kJ/kg and 220.23 kJ/kg, respectively. Finally, the enthalpy of vaporization at the exit of the drying chamber was 2,743.41 kJ/kg.

### 3.3. Mass flowrates

Table 5 expresses the calculated values of the mass flowrates of the different inlet and outlet streams of the dryer, as well as the wet volume.

Table 5. Results of the inlet and outlet mass flowrates of the spray dryer, and the wet volume.

| Equation | Parameter | Symbol | Value | Unit |
|---|---|---|---|---|
| (12) | Wet volume | $V_h$ | 0.869 | $m^3$.mixture/kg.dry air |
| (13) | Mass flowrate of dry air | $Q_{mas}$ | 598.39 | kg/h |
| (14) | Mass flowrate of water entering with air | $Q_{maa}$ | 11.37 | kg/h |
| (15) | Feed flowrate of cream | $Q_{mc}$ | 20.95 | kg/h |
| (16) | Outlet mass flowrate of powder | $Q_{mp}$ | 5.01 | kg/h |
| (17) | Mass flowrate of evaporated water | $Q_{mhv}$ | 15.94 | kg/h |
| (18) | Mass flowrate of water in the exhaust air | $Q_{ms}$ | 27.31 | kg/h |

Source: Own elaboration.

A wet volume of 0.869 $m^3$ mixture/kg dry air was obtained, while the mass flowrates of dry air and water entering with the air presented values of 598.39 and 11.37 kg/h, respectively. Likewise, the cream feed mass flowrate had a value of 20.95 kg/h, which is close to the actual mass flowrate applied in the spray-drying stage (18.5 kg/h) of the HeberNem-S® production process. Finally, the mass flowrates of the outlet powder (throughput), the evaporated water, and water in the outlet air had values of 5.01, 15.94, and 27.31 kg/h, respectively.

### 3.4. Sauter diameter

By means of equation (19), and taking into account that $a$ = 0.12 and $b$ = 0.7, the calculated Sauter diameter presented a value of 88.78 μm.

### 3.5. Total drying time

Table 6 shows the calculated values of the parameters drying time in the constant speed period [equation (21)]; the log mean temperature for the decreasing speed period [equation (22)]; drying time in the decreasing speed period [equation (23)] and total drying time [equation (24)].

Table 6. Results of the parameters determined by equations (21) - (24).

| Equation | Parameter | Symbol | Value | Unit |
|---|---|---|---|---|
| (21) | Drying time in the constant speed period | $t_s$ | 0.82 | s |
| (22) | Log mean temperature for the decreasing speed period | $\Delta Tml_2$ | 28.69 | ºC |
| (23) | Drying time in the decreasing speed period | $t_d$ | 6.08 | s |
| (24) | Total drying time | $t_t$ | 6.90 | s |

Source: Own elaboration.

The drying time in the constant speed period had a value of 0.82 s, while the drying time in the decreasing speed period was 6.08 s, for a total drying time of 6.90 s, which is consistent with what actually happens in the dryer. This is due to the large contact surface between the droplets and the dry air inside the spray dryer. It is worth noting that $t_d$ is 88.12% higher than $t_s$, which corresponds with the reported in the literature for this type of drying (Masters, 1991; Yanza, 2003).

3.6. Efficiencies

Table 7 shows the calculated values of the different efficiencies involved in the spray drying process.

Table 7. Results of the efficiencies involved in the spray drying process.

| Equation | Parameter | Symbol | Value | Unit |
|---|---|---|---|---|
| (25) | Thermal efficiency of the drying operation | $E_f$ | 82.18 | % |
| (26) | Overall thermal efficiency | $E_{tg}$ | 65.04 | % |
| (27) | Evaporation efficiency | $E_{ev}$ | 63.81 | % |
| (28) | Maximum thermal efficiency | $E_{tm}$ | 98.11 | % |

Source: Own elaboration.

The thermal efficiency of the drying operation had a value of 82.18%, while the global thermal efficiency, the evaporation efficiency and the maximum thermal efficiency were 65.04%, 63.81% and 98.11%, respectively, which can be considered acceptable and reasonable (Masters, 1991; Strumiłło *et al*., 2006; Woo *et al*., 2010). As reported by (Woo *et al*., 2010) the parameters generally monitored during spray drying include the inlet and outlet temperature of the drying gas, its moisture, as well as the feed flowrate of gas and liquid material. These data can be used to predict and improve the efficiency of the drying process, through the performance of mass and energy balances, statistical analysis and/or process simulations. Thermal efficiency is a parameter commonly used to carry out the optimization of a process of this type. Higher thermal efficiencies are obtained when the exhaust gas temperature is close to its saturation temperature. However, in most situations the process is guided by the need to obtain a product with the required quality and in the desired quantities, instead of knowing the efficiency of the equipment.

According to (Selvamuthukumaran *et al*., 2020), the air inlet temperature has to be as high as possible to maximize thermal efficiency and obtain final products with low residual moisture content. In this case, the increase in the air inlet temperature simultaneously increases the air outlet temperature.

The energy efficiency of the spray dryer can be increased through the unit operation. The higher the solids content and the temperature of the liquid to be fed, as well as the higher the air inlet temperature and the lower the air outlet temperature, the greater the energy efficiency, thus benefit the spray drying process. In addition, the use of a suitable atomizer and the implementation of a heat recovery system are important ways to increase energy efficiency (Cheng *et al*., 2018).

In (Toneli *et al*., 2013) it was verified that the air inlet temperature and the air mass flow had a negative effect on the thermal efficiency, while the reduction of the drying temperatures and the air mass flowrate resulted in lower energy requirements to heat the inlet air, which causes a higher energy efficiency. On the other hand, increased feed flowrate led to higher energy efficiency. In this study green banana biomass was processed in a pilot scale spray dryer equipped with a rotary atomizer.

3.7. Mathematical model of the spray drying process of the bacterium *Tsukamurella paurometabola* C-924

A mathematical model was obtained using the MATLAB® 7.10.0 program (see Figure 2), through which a number of phenomenological characteristics of the spray-drying process, described in the Materials and methods section, are calculated.

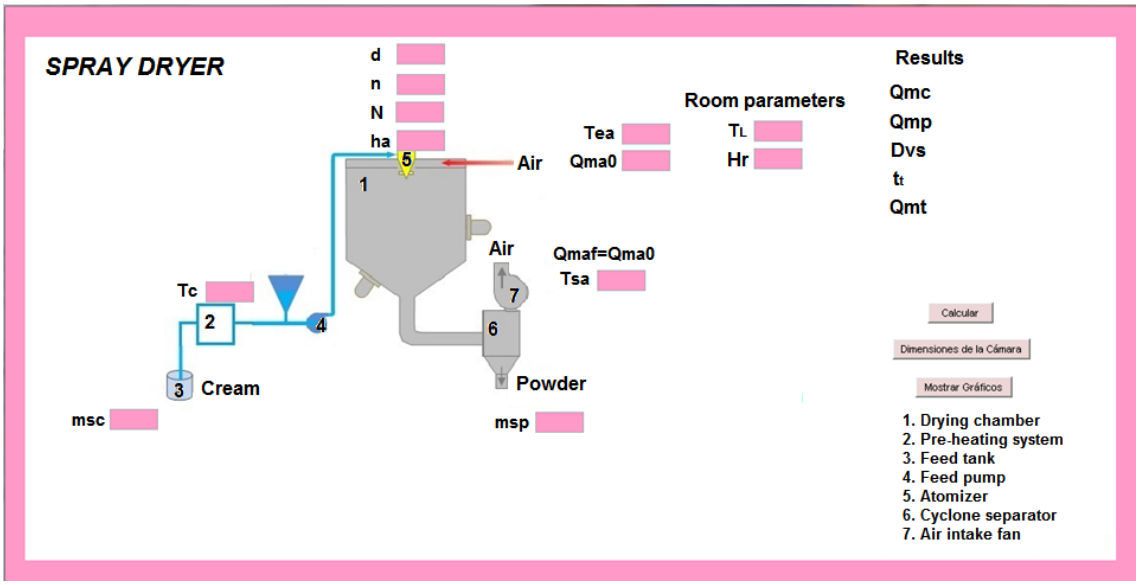



Figure 2. Main window of the mathematical modeling of the spray-drying process of *Tsukamurella paurometabola* C-924 obtained using MATLAB®.

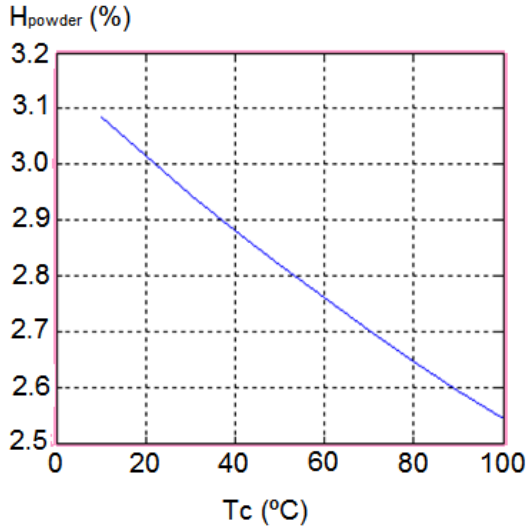
Source: Own elaboration.

The inlet flowrate is calculated based on the dry matter of the feed stream (cream), the residual moisture of the product stream (powder) and the enthalpies of the inlet and outlet streams. The evaporative capacity of the dryer is also calculated, as well as the moisture of the outlet air by using equations described in the literature by (Ángeles, 2009; Masters, 1991) and verified by (Paneque, 2010).

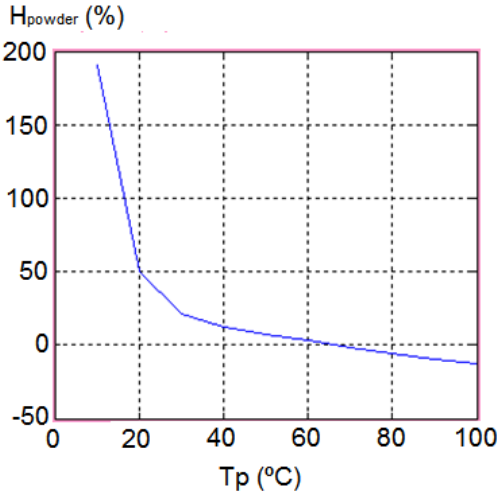
The residence time calculation of the atomized particle inside the dryer was not taken into account since the technique reported in the consulted literature (Angeles, 2009) does not share similar criteria and principles compared with the spray drying technology evaluated in this study.

3.8. Simulation of scale-up alternatives for spray drying according to the existing technology

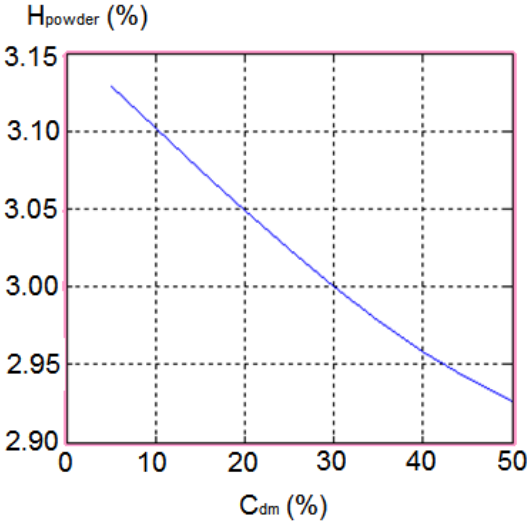
The scale-up alternatives were simulated in the MATLAB® 7.10.0 program, thus obtaining the results exposed in Figure 3.



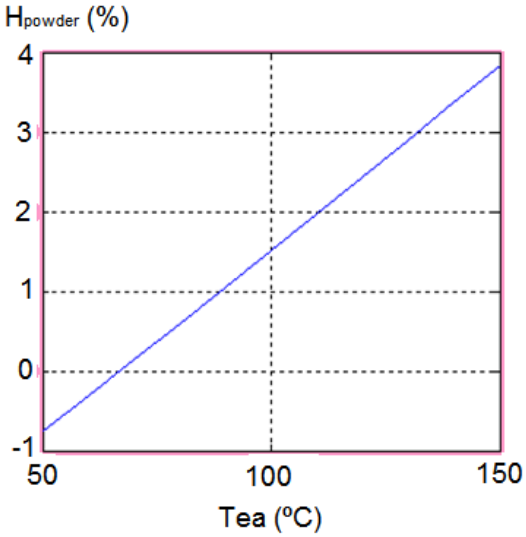

(a)



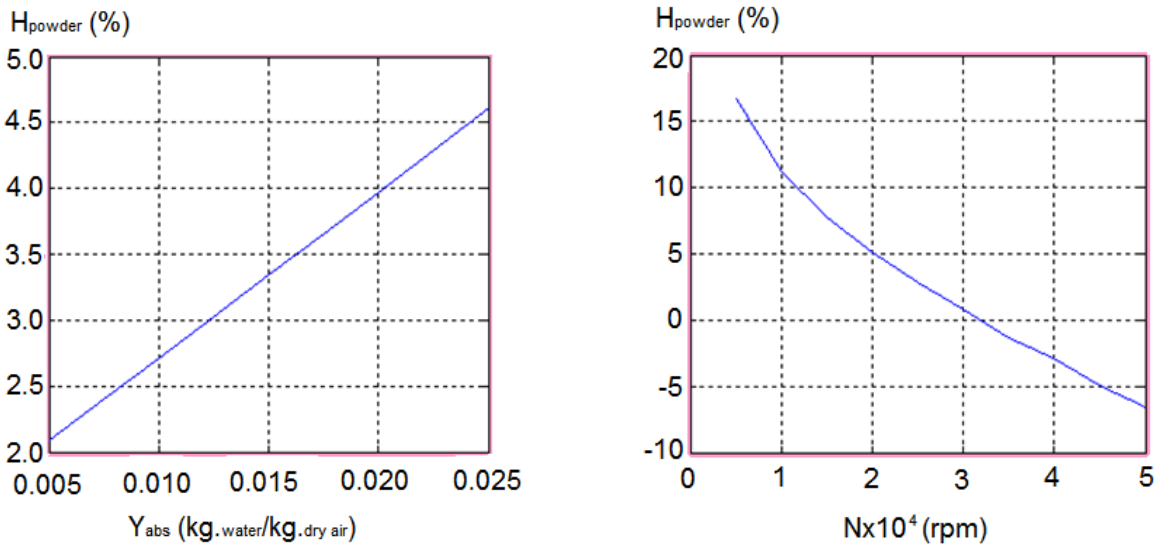

(b)




(c)




(d)




(e) (f)

Figure 3. Relation of the final moisture of the powder with the following scale-up parameters: (a) Cream feed temperature. (b) Powder temperature. (c) Cream dry mass. (d) Air inlet temperature. (e) Absolute moisture of the air. (f) Synchronous speed of the atomizer disc.
Source: Own elaboration.

The first scale-up alternative consisted of varying the cream temperature from 10 to 100 ºC, while the other parameters remained constant. When this criterion was simulated, a graph was generated where it can be observed the reduced influence that the cream temperature has on the powder moisture, which ranged from 2.5 to 3.1 %. The consulted literature (Yanza, 2003) states that this variable has more influence on the cream feed flowrate than on the final product (powder) moisture.

In the second alternative, by varying the powder outlet temperature between 10 and 100 ºC, it can be perceived that there is a strong dependence of this parameter on the final powder moisture, agreeing with what has been reported in the literature (Pita & Rojas, 2001; Yanza, 2003; Angeles, 2009). In this case, when the powder temperature varied from 10 to 100 ºC, the powder moisture ranged between 200 to -10 %. The strong dependence between these two parameters is due to the direct intense contact between the drying air and powder at the evaluated temperature, thus being a variable that directly influences moisture. Such high moisture values in the product mean that it is no longer in the solid state or did not dry out. In addition, mathematical tools were not used in the experimentally equation to restrict these excessive moisture values, since they were not reached experimentally, while negative moisture values mean that the product continues to dry until it has zero water content. In the studies carried out, such small values of residual humidity were not obtained, so there are no mathematical restrictions for these very small values, thus representing a weakness of the models obtained.

In the third alternative, the dry matter of the cream does not produce a high variation in the powder moisture because it is a variable related indirectly to the powder. However, according to (Ángeles, 2009) and (Yanza, 2003) this variable has its greatest influence on the cream feed flowrate. For this criterion, a cream dry matter between 5 and 50% was used because with values above 40%, the cream is no longer in a true liquid state, and its drying is not possible due to the high viscosity of the feed. In this alternative, for a variation in dry matter between 0 - 50%, a powder moisture between 2.80 and 3.10% was obtained.
For the fourth alternative, the air inlet temperature did not significantly affect the powder moisture, hence causing a minimal variation but more significant than what is caused by the cream feed temperature and dry matter. This is because it acts directly with the product evaporation rate, which increases the amount

of evaporated water. However, the decrease of the powder temperature in the later process stages causes a small increase in its moisture. The final powder moisture ranged from -0.6 to 4 % with a variation of the air inlet temperature from 50 to 150 ºC, which coincides with the consulted literature (Yanza, 2003; Ángeles, 2009).

In the fifth alternative, the increment of the inlet air absolute moisture causes a variation in the enthalpy of the inlet air, which also produces a decrease in the evaporation rate. This reduction of the evaporation rate, due to the very small values in which the absolute moisture can vary at controlled temperatures, causes an increase in the relative moisture of the air at the dryer outlet. This produces an increase in the powder moisture that varies from 2.1 to 4.60 %. In this alternative, the absolute humidity of the air was varied from 0.005 to 0.025 kg water/kg dry air.

In the sixth alternative, the synchronous speed of the atomizer (rpm) was varied from 5,000 to 50,000 rpm, causing that the powder moisture oscillates from 17 to -6%, which therefore caused a decrease in the size of the sprayed particle. This, in turn, directly affects the reduction of drying time both at a constant speed and at a decreasing speed, thus increasing the outlet temperature of the air and the powder, the latter being the variable that directly influences the final powder moisture.

## 4. CONCLUSIONS

1. A value of the heat capacity of the bacterium *Tsukamurella paurometabola* C-924 and the cream of 1.653 kJ/kg.ºC and 3.626 kJ/kg.ºC, were obtained respectively.

2. The mass flowrates of feed cream, outlet powder and evaporated water reached values of 20.95 kg/h, 5.01 kg/h and 15.94 kg/h, respectively.

3. The Sauter diameter had a value of 88.78 μm.

4. The drying time in the constant speed period was 0.82 s; the drying time of the decreasing speed period was 6.08 s, while a value for the total drying time of 6.90 s was obtained.

5. The thermal efficiency of the drying operation, the overall thermal efficiency, the evaporation efficiency and the maximum thermal efficiency had values of 82.18%, 65.04%, 63.81% and 98.11%, respectively, which can be considered reasonable and typical.

6. The increase in the cream feed temperature, the powder temperature, the cream dry mass and the atomizer rotation speed decrease the powder final moisture, while the air inlet temperature and the absolute moisture of the air increase it.

7. The outlet temperature of the powder and the synchronous speed of the atomizer disc are the variables that most influence the scale-up of the spray dryer in the evaluated scale.

8. The new mathematical model obtained in this work using the MATLAB® software describes the spray-drying process of HeberNem-S® bionematicide both phenomenologically and quantitatively.

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