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MEASUREMENTS AND COMPUTATIONS OF THE FLOW IN FULL-SCALE SUGAR EVAPORATIVE-CRYSTALLIZERS AND IN LAB-SCALE MODELS

A Dissertation

Submitted to the Graduate Faculty of the Louisiana State University and Agricultural and Mechanical College in partial fulfillment of the requirements for the degree of Doctor of Philosophy

in

The Department of Mechanical Engineering

by

Luis Fernando Echeverri
B.S., Universidad del Valle, 1997
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LIST OF ABBREVIATIONS AND ACRONYMS

• LATIN

\( A_{HT} \): Heat-exchange area \((m^2)\)

\( A_{XS} \): Cross-section area \((m^2)\)

\( BX \): Brix, unit of concentration for soluble solids.

\( C_D \): Drag coefficient

\( C_D^o \): Drag coefficient of a single bubble rising in stagnant media.

\( C_D^M \): Drag coefficient for a multi-particle system (e.g. high void fraction).

\( CFH \): Cubic feet per hour, unit for volumetric flow of gas.

\( CFM \): Cubic feet per minute, unit for volumetric flow of gas.

\( Co \): Distribution parameter (Drift flux model)

\( C_P \): Specific heat; heat capacity at constant pressure \((kJ/kg.K)\)

\( C_R \): Circulation ratio.

\( CVP \): Continuous vacuum pan.

\( D \): Diameter \((m)\)

\( D_G \): Bubble / gas particle equivalent diameter \((m)\).

\( D_h \): Hydraulic diameter of the pipe / channel \((m)\).

\( DS \): Dry substance, unit of concentration for total content of solid matter.

\( EC \): Evaporation rate \((kg/h.m^2)\)

\( Eo \): Eotvos number.

\( f \): Drag function.

\( fps \): Frames per second, unit defining frequency of image acquisition in a camera.

\( F \): Force \((N)\).

\( F_B \): Buoyancy force \((N)\)

\( F_D \): Drag force \((N)\).

\( Fr \): Froude number.

\( g \): Acceleration of gravity \((9.81 \text{ m/s}^2)\)

\( G \): Total mass flux \((kg/s.m^2)\)

\( h \): Heat transfer coefficient \((W/m^2.K)\)
\( h_g \): Latent heat of vaporization (kJ/kg)
\( H \): Liquid head above the top calandria plate (m)
\( ID \): Internal diameter (m).
\( j \): Volumetric flux of mixture (Drift flux model, m/s)
\( J_G \): Gas phase superficial velocity – gas volumetric flux (m/s)
\( J_L \): Liquid phase superficial velocity – liquid flux (m/s)
\( K \): Momentum exchange coefficient.
\( k \): Thermal conductivity (W/m.K)
\( L \): Tube / channel length (m).
\( LPS \): Liters per second, volumetric flow units for circulation.
\( M \): Drag coefficient multiplier
\( Mo \): Morton number.
\( N_\mu \): Inverse viscosity dimensionless number.
\( \text{No. tubes} \): Number of heating tubes in the calandria.
\( Nu \): Nusselt number
\( OD \): External diameter (m).
\( P \): Pressure (N/m\(^2\) = Pa)
\( PIV \): Particle Image Velocimetry
\( Pr \): Prandtl number
\( Pty \): Purity of sugar solutions.
\( q^\prime \prime \): Heat flux (J/m\(^2\).s = W/m\(^2\))
\( Q \): Volumetric flow rate (m\(^3\)/s)
\( \dot{Q}_w \): Heat transfer rate (J/s = W)
\( R_{ij} \): Interaction force between the phases i and j.
\( Re \): Reynolds number
\( Re_{TP} \): Two-phase Reynolds number
\( R_G \): Bubble / gas particle equivalent radius (m)
\( S \): Velocity ratio \((u_G/u_L)\)
\( SRI \): Sugar Research Institute.
\( t \): Time (s)
\( t/h \): Metric tons per hour, Mass flow unit.
T: Temperature (°C, K)
ΔT: Temperature difference (°C, K)
u: Velocity (m/s).
u_G: Gas phase velocity (m/s)
u_L: Liquid phase velocity (m/s)
Δu: Slip velocity \([u_G-u_L]\) (m/s)
U_t: Terminal velocity (m/s)
U_{TB}: Velocity of Taylor bubbles (m/s)
U_{TRANS}: Superficial gas velocity at regime transition to churn-turbulent regime (m/s)
UDF: User Defined Function, Fluent.
V: Velocity number.
V_B: Volume of a bubble (m³)
VF: Void fraction, volumetric fraction of the gas-phase.
We: Weber number
x: Quality of two-phase flow.
X_{LM}: Martinelli factor (two-phase algebraic model)

- **GREEK:**
  α_i: Volumetric fraction of the i-phase.
  α: Volumetric fraction of the gas-phase, void fraction.
  θ: Slope, inclination of the tube calandria plates.
  ρ: Density (kg/m³)
  Δ ρ: Density difference (kg/m³)
  F: Flow number
  λ_{RT}: Raleigh – Taylor wavelength (m)
  μ: Dynamic viscosity (Pa.s)
  σ: Surface tension (N/m)
  τ: Circulation period, ratio between nominal massecuite volume and circulation rate (s)
  τ_{ji}: Particulate relaxation time
  χ: Viscosity ratio \(\mu_G/\mu_L\)

- **INDEXES:**
  *: Dimensionless form.
B: Buoyancy
D: Drag

FILM: Layer at the interface between a fluid and a solid surface.

G: Gas phase
h: Hydraulic (diameter of the channel)
L: Liquid phase

LIFT: Lift force

M: Mixture liquid and gas
NB: Nucleate boiling

SAT: Saturation

TB: Taylor bubble.

TP: Two-phase

VM: Virtual mass.

W: Wall

x: Component in the horizontal plane.

y: Component in the vertical plane.
ABSTRACT

The circulation of massecuite is a key factor in achieving efficient heat transfer and crystallization in sugar evaporative crystallizers, and should be as high as practically possible for recovery, quality, and capacity reasons. This research report presents results on the circulation obtained applying modern experimental and numerical techniques. The main goals are contributing to expand the understanding of the process in sugar crystallizers; developing realistic models for the simulation of the circulation; and studying the effect of different design parameters.

The circulation of massecuite is driven by buoyancy force due to density difference between the vapor generated and the surrounding liquid in calandria tubes, where the momentum exchange models normally used for flow simulation cannot predict correctly the complex interfacial interactions. To address this problem, the exchange of momentum in buoyancy-driven gas-liquid vertical channel flows has been investigated experimentally from a fundamental perspective, focusing particularly on the complex regimes associated with high void fractions and with highly viscous media. A drag model has been developed from the experimental results, which represents the transfer of momentum in gas-liquid multiphase flows under adiabatic conditions.

A flow boiling instability has been identified in the calandria tubes, causing intermittent vaporization and pulsating circulation. This boiling instability leads to higher frictional resistance than in corresponding continuous adiabatic gas-liquid flows, and affects the transfer of momentum to the liquid phase.

Experimental results on the flow in sugar evaporative crystallizers have been obtained using a lab-scale model, where the major features of the fluid flow were replicated and
studied applying Particle Image Velocimetry. Field measurements of the flow in a full-scale continuous crystallizer have also been performed, where hot anemometers were used to determine the massecuite velocity and circulation.

Numerical results obtained applying Computational Fluid Dynamics (CFD) are presented and compared with the measurements performed in the lab-scale and full-scale crystallizers, confirming that the CFD solutions developed represent reasonably the flows studied. The CFD model developed has been applied to investigate numerically the effect of different design parameters on circulation, identifying potential alternatives for improving the hydraulic design and performance of sugar crystallizers through enhanced circulation.
CHAPTER 1 - INTRODUCTION

1.1 EVAPORATIVE CRYSTALLIZATION

Crystallization is essentially a purification process widely used in diverse industries because of several distinctive advantages. The solute that comes out of solution solidifies forming crystals that exhibit high purity, even when the process starts from extremely impure solutions. According to Mullin (2001) crystallization is “one of the best and cheapest methods available for the production of pure solids from impure solutions”. The energy requirements are normally lower than in other purification techniques such as distillation, and it can be performed at low temperatures (Perry, 1997), which is a desirable feature when dealing with biological systems for example. Ease in handling and packing of crystals, and its good appearance, are additional reasons for a large number of products being commercialized as crystals, as in the case of the sugar.

Crystallization is a complex process, and the evolution of the technology has involved a large empirical component. Some consider that crystallization remains “more an art than a science”. Although there is some true in this old adage, nowadays there is a considerable amount of science associated with the art (Mullin, 2001), and this research project intends to apply modern experimental and numerical techniques in the analysis of the fluid flow in sugar evaporative-crystallizers.

1.1.1 Supersaturation

Typically the temperature and the content of impurities (e.g. organic and inorganic non-sugar components) determine largely the maximum amount of solute that can be present in a thermodynamically stable solution. This limit is known as ‘solubility’ and corresponds to the saturation state, where the free Gibbs energy of the solid-and-liquid phases is the same,
so they can coexist in equilibrium. In the case that the solute concentration is below the saturation point, the solution is unsaturated, and any crystal present will tend to dissolve. If the solute concentration exceeds the saturation limit the solution is called supersaturated, and solidification may occur through crystallization and/or precipitation.

During crystallization processes the solute concentration (C) is forced to be higher than the saturation concentration, leading to an excess with respect to the solubility that is denoted as ‘supersaturation’, and is the driving force for nucleation and crystal growth. In the case of low supersaturation the solution will be in the metastable region (Fig 1.1), and crystal growth will occur, but not nucleation. If the supersaturation is higher, the labile zone is reached, where spontaneous nucleation will take place, as well as crystal growth.

![Typical phase diagram for a binary solid-liquid system.](image)

Fig 1.1 Typical phase diagram for a binary solid-liquid system.

Isothermal evaporative crystallization can be obtained by evaporation of the solvent (normally water), correspondingly increasing the solute concentration, while the solubility remains the same since it is a function of temperature and purity. The increase in solute
concentration produces the supersaturation required to drive the crystallization process. Isothermal evaporative crystallization follows the path 1-2 in Fig 1.1

Cooling crystallization is obtained by reducing the temperature of the solution. Most substances exhibit a positive temperature coefficient of solubility \( \partial C_{\text{SAT}} / \partial T > 0 \), indicating that the solubility is higher as the temperature is increased. Therefore, a decrease in the temperature of the solution has the effect of lowering the saturation concentration, which can reach values below the actual concentration of the solution, producing the supersaturation required to drive the crystallization process. Cooling crystallization follows the path 1-3 in Fig 1.1

1.1.2 Crystallization of Sugar

Two techniques are commonly used to crystallize sugar; a relatively fast evaporative crystallization that separates sucrose from concentrated cane juice (syrup); and a slower cooling crystallization that recovers as much sugar as possible from the molasses.

The removal of water during the evapo-crystallization results in a dense and highly viscous mixture of sugar crystals and mother liquor called ‘massecuite’. The viscosity of massecuites increases drastically with the crystal content and the amount of impurities, leading to a practical limit for the maximum crystallization by evaporation when the flow resistance becomes too high and the heat transfer and crystallization rate are slowed down (usually after three stages). Then crystallization by cooling continues the process as long as economically justifiable.

Most of the sucrose crystallization in sugar mills and refineries is achieved through the use of evaporative-crystallizers, best known as ‘vacuum pans’, which are the main subject of interest in this research. The water in the sugar cane juice is evaporated to produce
the supersaturation required to drive the growth of the sugar crystals. Spontaneous nucleation is avoided to favor a good crystal size distribution, and as normal in industrial practice seeding is employed, so the crystals are formed around injected small grains with a controlled and regular size. The process is carried under vacuum (~14-16 kPa abs) to reduce the boiling temperature \(T_{\text{SAT}} \sim 55 \, ^{\circ} \text{C}; T_{\text{M ASS SEC UITE}} \sim 65-70 \, ^{\circ} \text{C}\), preventing the occurrence of caramelization, sucrose losses, and color formation associated with temperatures above 110 - 130 \(^{\circ} \text{C}\). Additionally, the low boiling temperature makes it possible to use as heating source low-pressure vapors bled from the evaporation series, allowing the configuration of efficient thermal systems.

The design of batch, and more recently continuous, sugar evapo-crystallizers has evolved through an empirical process because of a lack of understanding of the fundamental heat transfer and fluid flow phenomena and the intrinsic difficulty in performing useful experimental measurements. Batch sugar evapo-crystallizers are typically designed as cylindrical vessels, using a calandria constructed with vertical tubes that provide the heat exchange area, and using steam or saturated vapor as the heating source that condenses at the outer-wall of the tubes (~ 80-130 kPa abs). Boiling of the sugar cane juice occurs inside the heated calandria tubes driving the upward circulation of the flow. A downtake path is left, typically in the center, to allow the liquid phase to flow downward to the bottom before entering into the tubes again (Fig 1.2).

1.1.3 Circulation in Sugar Evaporative Crystallizers

In a sugar evapo-crystallizer the massecuite flows upwards in the calandria tubes driven by the rise of the gas phase, above the top calandria plate it flows in direction to the downtake, it flows then downward through the downtake to reach the bottom and enters into
the tubes again (Fig 1.2). The internal flow of the massecuite described above is known simply as ‘circulation’.

![Diagram of batch sugar evaporative crystallizer](image)

**Fig 1.2** Typical design of a batch sugar evaporative crystallizer.

High circulation has proved to be indispensable for the successful design and operation of sugar evaporative crystallizers. Some benefits attributed to good circulation rates are:

- **Higher recovery**: A higher massecuite dry substance concentration is possible, leading to a higher crystal yield and a lower purity of molasses. This means that more crystal sugar can be produced and less sugar would be lost with the molasses.

- **Quality**: Enhanced crystal size distribution and lower sugar color are obtained. A vigorous circulation prevents stagnant regions or cold or hot zones that can induce differences in the supersaturation level and crystal growth rate. It also prevents settling of the crystals.
- Capacity: Higher throughput is achieved, resulting from the enhanced heat transfer and the higher crystallization rate.

- Energy efficiency: Use of lower temperature differences becomes possible, leading to potential reductions in energy costs through the use of lower pressure heating vapor.

Traditionally sugar evaporative crystallizers have relied on the natural circulation generated by boiling as the only driving force for the fluid flow. The boiling process generates vapor that has a density considerably lower than the surrounding media ($\rho_G \sim 0.2$ kg/m$^3$; $\rho_L \sim 1300-1500$ kg/m$^3$; $\rho_L/\rho_G \sim 7000$ times!), and consequently large buoyancy forces are generated by the density difference between the two phases. The gas phase flows upwards in the calandria tubes due to the buoyancy force, and in doing so it transfers momentum to the liquid phase, constituting the driving force for the natural circulation of the massecuite. Circulation assistance is provided in some cases either by using a large impeller installed in the downtake or by injecting a gas under the calandria (e.g. vapor).

The circulation in evaporative crystallizers is determined by the balance between the buoyancy forces in the calandria tubes and the frictional flow resistance. This balance is influenced by many parameters that include the regime, void fraction, vapor nucleation, coalescence and breakage of bubbles, liquid properties, etc, but the lack of detailed information has not enabled a clear understanding. It is a fact however that the circulation plays an important role in the crystallization process and the convective-boiling in the calandria tubes, where a mutual interaction occurs as the circulation depends on the flow of the evaporated gas phase. Therefore, any factor affecting the circulation will be detrimental for the heat transfer, and vice versa.
1.2 MULTIPHASE GAS-LIQUID VERTICAL CHANNEL FLOW

In sugar evaporative crystallizers heat is transferred to the massecuite contained inside the calandria tubes to produce boiling, generating the vapor bubbles and therefore the buoyancy forces responsible for the circulation. The multiphase flow and heat transfer processes are highly coupled in a phenomenon that can be described in general terms as a ‘buoyancy-driven convective-boiling vertical channel flow’. For the numerical analysis of the flow it is critical to understand the interactions between the gas and liquid phases, and consequently a part of this research addresses the study of multiphase buoyancy-driven vertical channel flows.

There are numerous practical applications involving boiling flow in vertical tubes and gas-liquid vertical channel flows, where the buoyancy forces provide totally or partially the driving force for circulation. Therefore, studying the flow in calandria tubes is important from the fundamental point of view, since it would provide information useful to enhance the understanding and predictions of the flow and heat transfer in analogous applications. Examples of some important buoyancy-driven internal flows are:

- Steam boilers: Employed for thermal energy conversion from fuels, they typically include arrays of vertical tubes known as ‘water walls’, where a large part of the heat released by the combustion is transferred to the water inside the tubes. Except for boilers operating close to the critical point, it is common to rely on buoyancy forces to drive the flow along the water walls. Due to the necessity of understanding the cooling limits in nuclear reactor cores and steam generators a large amount of information has been reported on water boiling in tubes (Dhir, 1998).
- **Bubble columns**: This is a reactor design widely used in chemical processes. The liquid media is contained within a column or recipient, where gas bubbles generated at the bottom rise due to the buoyancy forces, and come into close contact with the liquid phase. The simplicity, low cost, and absence of moving parts are reasons for the widespread utilization of bubble columns.

- **Air-lift reactors**: Similar in principle to the bubble columns, the air-lift reactor includes a disengagement region and a downcomer channel. The gas phase rises in the bubble column or ‘riser’, driving the upward two-phase flow; a disengagement volume is available above the riser, where the gas phase crosses the free surface and leaves; then the liquid phase flows through the downcomer to the enter into the riser again. Application of this technology can be found in the chemical industry, biotechnology applications, and in the treatment of wastewater (Duff, 2005).

- **Oil wells**: The density of the oil column can be reduced by means of injection of air bubbles, decreasing the hydrostatic head and therefore easing pumping operations. This ‘airlift’ effect is used by the oil industry for aged wells, where the down-hole pressure becomes too low, so air is injected to facilitate the oil extraction and enhance the production rate (Duff, 2005). Similar applications are found for the pumping of substances with high content of abrasives, where the absence of moving parts represents an obvious advantage.

### 1.2.1 Boiling in Vertical Tubes

The study of the fluid flow and heat transfer during boiling processes involves, in addition to the single-phase convective transport problem, other components related to the
dynamic mass, momentum and energy interactions between the phases and non equilibrium effects that add considerable complexity to the analysis (Kreith et al., 1999).

Elaborate theories have been proposed on the nucleation of the vapor phase during boiling, after which the nuclei particles would detach, grow and flow upwards while they may merge with other bubbles or break. It is believed that the movement of the gas phase agitates the surroundings, mixing the hot liquid close to the exchange surface with the rest of the stream, and triggering the high heat transfer coefficients typically encountered in boiling applications. In the case of interest, boiling in vertical channels, significant interactions between the fluid flow and the heat transfer occur continuously in the vertical direction as more vapor is generated.

The best-known description of boiling in vertical tubes presents a progressive change in the regime as the content of vapor is increased. The regimes are defined utilizing the same terms used for adiabatic gas-liquid flows. At the tube entrance only liquid is present, observing the classic single-phase convection problem; then as more vapor is generated and the void fraction\(^1\) increases, transition to the bubbly, slug, churn, annular, mist, and finally the single gas phase regimes will occur. The fluid flow and the heat transfer mechanisms undergo radical variations along the vertical boiling tube, including severe transitions between different regimes.

### 1.2.2 Theoretical Analysis of Gas-liquid Flow

The analysis of adiabatic multiphase gas-liquid flows has been traditionally approached through the use of empirical correlations and semi-empirical or mechanistic models; this as a consequence of the complicated interactions and the difficulty in performing experimental measurements, particularly as the void fraction gets higher, when

---

\(^1\) Void fraction \((\alpha)\): Refers to the volumetric fraction of the gas phase in the flow
most of the standard experimental fluid mechanics techniques, such as laser velocimetry, are incapable of giving accurate information. When boiling is involved the complexity is even higher, and in spite of the progress in understanding the physics behind the process, it is still necessary to rely on dimensionless groups and empirical constants for developing engineering correlations (Dhir, 1998).

Computational Fluid Dynamics (CFD) is a powerful tool for the analysis of heat transfer and fluid mechanics problems, which has become particularly popular during the last decades thanks to more affordable computing power, development of efficient solvers, and the spreading of CFD commercial codes. The CFD analysis essentially comprises the discretization of the flow geometry and the application of numerical techniques to find solutions for the governing Navier-Stokes equations (conservation of mass, momentum, and energy). There are two main approaches in CFD for the analysis of multiphase flows:

- The Euler-Lagrangian approach treats the primary phase as a continuum, solving the time-averaged Navier-Stokes equations for this phase. The secondary phase is solved tracking the particles, bubbles, or droplets through the flow field. This model is based on the assumption that the dispersed secondary phase occupies a low volume fraction and has a small effect on the flow field of the primary phase, and is therefore unsuitable for the modeling of buoyancy-driven flows at high void fraction.

- The Euler-Euler approach, also known as the two-fluid model, assumes that the secondary phase(s) can be treated as a continuum, and then the governing equations are solved individually for each phase. This approach involves the solution of more equations, which are coupled to each other, introducing additional terms for the volume fraction and for the interactions between the phases, resulting in a very
expensive model from a computational point of view. However, this approach is the most suitable for the analysis of multiphase flows involving complex interactions and high void fraction, such as the case in evaporative crystallizers and buoyancy-driven gas-liquid channel flows.

The equations used are:

Conservation of mass for phase $i$ (continuity):

$$\frac{\partial}{\partial t}(\alpha_i \rho_i) + \nabla \cdot (\alpha_i \rho_i \bar{u}_i) = \sum_{j=1}^{n} \dot{m}_{ji}$$

Conservation of momentum for phase $i$:

$$\frac{\partial}{\partial t}(\alpha_i \rho_i \bar{u}_i) + \nabla \cdot (\alpha_i \rho_i \bar{u}_i \bar{u}_i) = -\alpha_i \nabla p + \nabla \cdot \tau_i + \sum_{j=1}^{n} \left( R_{ji} + \dot{m}_{ji} \bar{u}_j \right) + \alpha_i \rho_i \left( F_i + F_{\text{lift},i} + F_{\text{vm},i} \right)$$

Where:

- $\alpha_i$: Volume fraction of the $i$ phase
- $n$: Number of phases
- $\bar{u}_i$: Velocity of $i$ phase
- $\dot{m}_{ji}$: Mass transfer from $j$ to $i$ phase
- $\bar{u}_{ji}$: Interphase velocity phases $i$ and $j$
- $\tau_i$: Stress-strain tensor of $i$ phase
- $F_i$: External body force
- $F_{\text{lift},i}$: Lift force
- $F_{\text{vm},i}$: Virtual mass force
- $R_{ji}$: Interaction force between phases $i$-$j$

The interphase force $R_{ji}$ depends on the cohesion between the phases, friction, pressure, and other effects. Since the buoyancy forces acting on the gas phase drive the flow studied, representing correctly the exchange of momentum or interaction between the gas and liquid phases is crucial for the numerical analysis of the flow patterns and circulation.

To compute the momentum interaction between phases, the Eulerian-Eulerian CFD approach uses an inter-phase momentum exchange coefficient ($K_{ji}$), which allows the calculation of the local interaction forces ($R_{ji}$). The model for interphase exchange of momentum and the conditions applied are:
\[ \sum_{j=1}^{n} R_{ji} = \sum_{j=1}^{n} K_{ji} (\bar{u}_j - \bar{u}_i) \]

\[ R_{ij} = - R_{ji} \]

\[ R_{ii} = R_{jj} = 0 \]

The momentum exchange coefficient \((K_{ji})\) is customarily defined with base in the bubble Reynolds number \((Re)\), drag coefficient correlations \((C_D)\) introduced in the drag function \((f)\), the void fraction \((\alpha)\), and the particulate relaxation time \((\tau_{ji})\):

Bubble Reynolds number:

\[ Re = \frac{\rho_L \cdot D_G \cdot (u_G - u_L)}{\mu_L} \]

Drag function:

\[ f = \frac{C_D \cdot Re}{24} \]

Particulate relaxation time:

\[ \tau_{ji} = \frac{(\alpha_j \rho_j + \alpha_i \rho_i) \cdot \left(\frac{\rho_j}{2} + \frac{\rho_i}{2}\right)^2}{18 \cdot (\alpha_j \mu_j + \alpha_i \mu_i)} \]

The relaxation time concept is used in physics to describe the time required in a system to reach equilibrium after a disturbance has occurred. For the specific case of particles, the relaxation time describes the period required for the particles to relax or achieve equilibrium with respect to a new dynamic state, and depends mainly on the mass of the particles and their mobility. From the definition presented above it can be seen that the relaxation time increases with the size of the particles, and goes down as the viscosity increases, indicating that particles reach an equilibrium state faster as their size is smaller and the viscosity of the media is higher.
Momentum exchange coefficient:

\[ K_{ji} = \frac{\alpha_j (\alpha_j \rho_j + \alpha_i \rho_i)}{\tau_{ji}} * f \]

The exchange coefficient should become negligible as the secondary phase tends to be absent from the computational domain. To impose this condition the volumetric fraction of the liquid phase \( (\alpha_i) \) is used as multiplier of the drag function in the calculation of the exchange of momentum, resulting in a linear reduction in the magnitude of the interphase forces as the local void fraction gets higher.

It can be appreciated that the momentum exchange is computed essentially as a function of the fluid and flow properties, except for the drag coefficient \( (C_D) \) that is used to calculate the drag function \( (f) \). Therefore, the Eulerian-Eulerian approach relies ultimately on a drag coefficient correlation to couple the momentum equations of the different phases and obtain closure for the system of equations, and the accuracy of the numerical prediction depends to a great extent on having precise drag correlations, able to represent the momentum interaction under a wide range of practical conditions and for fluids with rheological properties encountered in practice.

Most available drag correlations for gas-liquid systems have been developed from tests with single bubbles rising in stagnant media, and are the ones normally encoded in the interaction models included in CFD commercial codes. It is accepted however that many factors influence the behavior of single rising bubbles, such as size, temperature, shape, rising pattern, age, power law index of the liquid, contamination, surface tension, and presence of surfactants. For example, significant differences exist between the rise of bubbles in ideal systems (spherical bubbles, pure liquid) and non-pure systems (distorted
bubbles, impurities present), where the bubble terminal velocity may be only half as much as that under ideal conditions.

Normally the drag coefficient determined for bubbles is correlated with the Reynolds number of the bubble, following the same approach used to describe drag forces around solid spheres, but for the case of bubbles a large dispersion and scatter exists between the reported data and drag correlations. The uncertainty about the drag coefficient in gas-liquid flows is evidenced by the large number of drag correlations that have been proposed, while no accepted generalized expression is available for the drag coefficient of even the simplest basic case: single bubbles rising in stagnant media.

• **High Void Fraction**

For bubbly flows it has been reported that the concentration of the gas phase, or void fraction, has a significant effect on the exchange of momentum increasing the drag interaction. The change in the balance buoyancy-drag would be caused probably by the proximity between the bubbles, favoring the interaction with other bubbles, or by the reduction in the buoyancy forces as a consequence of the lower density in the surrounding gas-liquid mixture. Conversely, when the void fraction exceeds $\alpha \sim 0.20-0.30$ and transition to more complex regimes such as the slug and churn flow has taken place, it has been determined the drag interaction tends to reduce as the gas content increases.

Rusche and Issa (2000) proposed for bubbly flows a power and exponential correction for the drag coefficient that increases the momentum interaction between the phases as the void fraction becomes larger. By contrast, Isshi and Zuber (1979) proposed a correlation for the churn-turbulent regime that indicates a decrease in the drag coefficient as the void fraction goes up. The opposite behavior is evidence of fundamental differences
between the physics at low and high void fractions, which would affect significantly the flow properties and the level of interaction between the phases.

Drag coefficient for bubbly flow, Rusche and Issa (2000)

\[ C_D = \left[e^{3.644\alpha} + \alpha^{0.86\alpha}\right] C_D^0 \quad \text{- Drag increases with void fraction} \]

Drag coefficient for churn-turbulent flow, Isshi and Zuber (1979)

\[ C_D = \frac{8}{3} (1 - \alpha)^2 \quad \text{- Drag decreases with void fraction} \]

The drag correlations presented in the literature correspond usually to the simplest possible gas-liquid regime, bubbly flow, while only a few expressions have been proposed for the other regimes typical of higher void fractions. The preference for bubbly flows among researchers is understandable, since other regimes are more complex and difficult to investigate as the void fraction is increased, when they become unstable and difficult to measure with established experimental techniques.

1.3 PROBLEM STATEMENT

This research project addresses three specific subjects that are closely inter-related, and are presented below:

- **Fluid Flow in Sugar Evaporative Crystallizers**

  The circulation is a key factor for the efficient performance of sugar evaporative crystallizers, and must be as high as practically possible for recovery, quality, capacity, and energy reasons. However, the complexities of the heat transfer involved and fluid flow phenomena and the difficulty in performing experimental measurements have prevented a satisfactory understanding of the process, so the information available about massecuite circulation is limited and often imprecise.
This project investigates the fluid flow in sugar evaporative crystallizers applying state of the art experimental and numerical techniques. The outcome is an enhanced description of the massecuite boiling phenomena and the flow patterns and circulation rate during the crystallization process.

- **Analysis of the Design of Evaporative Crystallizers**

  The lack of information on the fluid flow in evaporative crystallizers has led to an empirical evolution of the technology, where guesses and trial and error have been the main development tools. Although the basic mechanisms that govern the processes of boiling, buoyancy, and two-phase fluid flow have been extensively studied in simpler geometries, their combined behavior in the complex geometry of evaporative-crystallizers makes it difficult to understand the process and optimize the design.

  This project applies CFD in the analysis of the effect of geometric and operation factors connected to the design of evapo-crystallizers on the circulation. The outcome is the identification of desirable design variables for the construction of batch and continuous sugar evapo-crystallizers, based on the established physical principles that are taken into account during the numerical simulation of the flow.

- **Exchange of Momentum in Buoyancy-driven Gas-liquid Vertical Channel Flow**

  The investigation of buoyancy-driven gas-liquid vertical flows is interesting from the fundamental and practical point of view. This type of flow is present in several important industrial applications, like steam generators, gas columns, oil wells, and the calandria tubes in sugar crystallizers. The numerical simulation of this type of flow could be potentially beneficial for engineering analysis and design, but first it is required to understand well the physics and characterize interactions such as the exchange of momentum between the phases.
This project involves the investigation of buoyancy-driven gas-liquid vertical channel flows, looking for a better understanding of the momentum interactions between the phases as the gas flux, circulation, and fluid properties change. The outcome is the characterization of the behavior of the natural circulation produced, and a momentum interaction model suitable for the numerical analysis of gas-liquid vertical channel flows involving highly viscous media and high-void fractions. The model developed is applied to compute the exchange of momentum between the vapor and the massecuite during the CFD simulations of the flow in full-scale sugar evaporative crystallizers.

1.4 OUTLINE OF RESEARCH

Initially a literature review is carried out, looking for information about the circulation and heat transfer in sugar evaporative crystallizers and the theoretical basis available for the analysis of multiphase gas-liquid flows. The review is presented in chapter 2.

The analysis of the fluid flow in sugar crystallizers is performed initially in a lab-scale test rig, involving experimental measurements and numerical simulations. The scaled-down test rig (TR1) was designed and constructed to represent the major features of the flow in sugar evapo-crystallizers, study the effect of several factors on the circulation, and provide data for validation of the CFD solution. The liquid circulation is driven by the density difference between the liquid phase and air injected in vertical channels that represent the calandria tubes. Particle-Image Velocimetry (PIV) is applied to characterize the flow, and the Eulerian-Eulerian model of the commercial CFD code FLUENT is used for the numerical simulations. Results obtained with the lab-scale facility are presented in chapter 3.
For the study of the momentum interaction in buoyancy-driven gas-liquid vertical channel flows a second experimental facility was constructed (TR2). The facility can be described as an instrumented air-lift reactor, provided with a transparent riser that permits visualization of the flow main features and application of high-speed photography. A model for the interfacial exchange of momentum under adiabatic conditions is developed from the experimental data. The model developed is compared with information reported by Rouillard (1985) from experiments performed in a single heated boiling tube facility, observing that a reduction in the momentum interaction with respect to adiabatic conditions is necessary to obtain agreement. Field measurements were performed in a sugar batch crystallizer to verify that a boiling instability occurs within the calandria tubes, causing increased frictional resistance and intermittent vaporization that result in less momentum being transferred to the liquid phase. Results obtained in the study of the interfacial momentum interaction in gas-liquid vertical channel flows and calandria tubes are presented in chapter 4.

Field measurements of the massecuite velocity in a 160 m^3 continuous sugar evaporation crystallizer were performed in a Louisiana sugar mill using commercial insertion flow sensors. The design of the sensor is based in the same principle of the hot-wire anemometers, but the probes are encapsulated within a metal cover that makes them suitable for the rough environment found within the crystallizers. The Eulerian-Eulerian model of the commercial CFD code FLUENT is used for numerical simulation of the flow within the crystallizer studied, as well as in another type of continuous and two batch sugar crystallizers. Results obtained on the flow in full-scale sugar crystallizers are presented in chapter 5.
In the last section the CFD model developed is applied to analyze the design of sugar evaporative crystallizers, performing a systematic evaluation of the effect of different design variables on the circulation rate. The factors that have been studied numerically are:

- Height of the liquid above the top calandria plate (H).
- Length of the heated calandria tubes (L).
- Size of the downtake – circulation ratio (C_R).
- Shape of the downtake (flat vs. bowed calandria wall).
- Angle of the top and bottom calandria plates.
- Geometry of the bottom section.

Results from the application of CFD to improve the hydraulic design of continuous evaporative crystallizers are presented in chapter 6.
CHAPTER 2 - BACKGROUND

2.1 PRACTICAL ASPECTS IN THE DESIGN OF SUGAR EVAPORATIVE CRYSTALLIZERS

Sugar evaporative crystallizers have been developed during the last two centuries\(^2\) using essentially an empirical approach, which has lead to the proven designs of batch and continuous evaporative crystallizers in use today. In the development process it was recognized that a key factor for a successful design is the good circulation, which is often the main comparison point between different designs based on presumed flow and heat transfer characteristics. Some aspects and practical details related to the circulation of massecuite and the design of batch and continuous evaporative crystallizers are presented here.

2.1.1 Batch Evaporative Crystallizers

- **Calandria**

  The calandria is essentially a shell and tube heat exchanger. It consists of two plates at the top and the bottom, and multiple vertical tubes in between that are heated at the outer diameter with steam or vapor and contain the boiling massecuite at the inner diameter surface. The main variations in the design of calandrias are illustrated in Figure 2.1

  - Inclined plate: The plates of the calandria are angled between 10° and 25° to the horizontal. One of the plates might be horizontal in some cases.
  
  - Floating calandria: The downtake is located in the outer annular region. Although it was developed in an attempt to improve the circulation characteristics, measurements involving the use of a radioisotope tracer showed that the conventional central downtake performs better (Wright, 1966). The combination of the annular downtake

\(^2\) Edward C. Howard developed the first sugar batch evaporative crystallizer in 1813.
with a small central downtake has also been tried, resulting in an inefficient design, prone to the formation of false grain (Webre, 1959).

- Horizontal flat plate calandria: This is the most common design. A single central downtake is normally used, with vertical tubes arranged around it, constituting an efficient design of simple construction and maintenance.

![Fig 2.1 Designs of sugar batch evaporative crystallizers.](image)

The ratio between the heat exchange area\(^3\) and the capacity or nominal volume of massecuite is normally around 6 m\(^2\)/m\(^3\), but for high strike grades, where evaporation rates are higher, it can be increased to 9 m\(^2\)/m\(^3\) (van der Poel et al., 1998).

- **Downtake**

  A critical parameter in the design of sugar evapo-crystallizers is the ‘circulation ratio’, devised originally by Smith (1938) for coil crystallizers to describe the relation between the area available for the massecuite up-flow and down-flow. Smith observed that

\[
A_{HT} = \text{No.tubes} \times \pi \times \text{OD} \times L
\]

\(^3\) Heat exchange area is computed normally based on external pipe diameter.
good performance coil evapo-crystallizers were constructed with a circulation ratio not larger than three, and proposed this value as a design constraint, which was applied satisfactorily in many subsequent applications. The concept of the circulation ratio was then extrapolated to the newer calandria crystallizers, and its value was normally around two in the 1960s (Jenkins, 1958). Today it is considered that the circulation ratio should not exceed 2.5, but as Jenkins described years ago, this parameter “is still an empirical figure and simply expresses an area ratio which has been found satisfactory in practice”.

\[
\text{Circulation ratio} = \text{No. tubes} \times \left( \frac{\text{ID}_{\text{TUBES}}}{\text{ID}_{\text{DOWNTAKE}}} \right)^2
\]

The diameter of the downtake is currently designed between 30-50% of the calandria diameter, usually around 40% for natural circulation pans, while smaller downtakes can be used when a stirrer is provided.

Deflectors are sometimes installed above the downtake perimeter in the belief that directing the rising flow of ‘hot massecuite’ apart from the ‘cold’ down-flow at the center is convenient for the circulation (Fig 2.2.b), preventing a mixing that would otherwise interfere (Webre, 1959). Similarly, tapered and rounded downtakes have been introduced in an attempt to improve the flow patterns (Fig 2.2.a). However, the frictional resistance might be actually increased as the cross sectional area is reduced.

- **Calandria Tubes**

The high viscosities of the massecuites demand the use of large diameters to overcome the friction with the available buoyancy. Tubes between 75-125 mm (3-5 in) in diameter are normally used, 100 mm (4 in) being the most common diameter. Tubes above this range exhibit less resistance and promote circulation, but the lower heating area / volume ratio is unfavorable (Webre, 1959) and the graining volume increases (Rouillard, 1985).
Tubes with a length between 900 and 1500 mm (36-60 in) have been traditionally used, but currently they are designed between 700 and 1200 mm. Short tubes are selected for low-grade evapo-crystallizers, while longer tubes are chosen for high-grade evapo-crystallizers or when a stirrer is provided. Short tubes give the best heat transfer coefficients, and probably there is no justification for tubes longer than 900 mm, even if it results in a decrease in the heat exchange area (Webre, 1959).

Fig 2.2 [a-left] Tapered downtake, and [b-right] radial deflector above downtake.

- **Bottom**

The bottom section should promote an even distribution of the massecuite across the calandria, without restricting circulation or providing stagnant areas, and allow the discharge of massecuite by gravity within an acceptable time.

The ideal bottom shape of the sugar evapo-crystallizers has been a major concern for a long time among sugar technologists. For example, Tromp (1936) presented a developed ‘Stream Flow Calandria Pan’\(^4\), where the massecuite was expected to circulate easier as a

---

\(^4\) Netherlands Patent No. 24537, Sep 1929
result of the tapered downtake and a stream-lined bottom, designed with a minimum angle of 20° to facilitate the discharge.

Currently the most frequently encountered bottom shapes are (Fig 2.1):

- Conical bottom: When coil evapo-crystallizers dominated, a conical bottom was the preferred option, ensuring a smooth discharge of the massecuite. In the case of calandria evapo-crystallizers, this shape is believed to lead to a significant stagnant zone. The bottom angle does not need to be greater than 20° (Bosworth, 1959).

- W bottom: While small angles of the bottom reduce the footing volume, they affect the discharge velocity and lengthen the strike time. Two discharge valves are recommended. Usually the angle of the bottom is between 17° and 25° (van der Poel et al., 1998).

- **Strike Height**

  As the strike height increases, the hydrostatic pressure in the calandria tubes increases, rising the boiling temperature. As a result, the available temperature difference becomes smaller, leading to a reduction in evaporation rate and consequently reduced circulation and crystallization rates. The situation is particularly critical at the end of the strike, when the highest level coincides with the maximum density and viscosity of the massecuite, all parameters that impair circulation.

  Tests performed by Austmeyer (1986) suggested that a maximum in the heat transfer coefficient occurs when the liquid level is around $H \sim 0.80 \text{ m}$ for after-product boiling, and around $H \sim 0.15–0.65 \text{ m}$ for white sugar boiling according to the apparent temperature difference available, after which it starts reducing progressively as the massecuite level increases. A maximum height around $H \sim 1.6 \text{ m}$ could give the best balance between quality,
performance and capacity. The strike height may be higher with high grade or refined boiling.

- **Disengagement Height Above the Massecuite**

  Enough space above the massecuite must be provided for disengagement of the liquid from the vapor, minimizing entrainment and allowing for froth explosion, which occurs often on starting a boiling or on cutting over. The disengagement volume should represent between 65-100% of the strike volume, being normally between 1.5–3.0 m (5-10 ft) but preferably as high as possible (Rein et al., 2004).

- **Shell**

  Sugar evaporative crystallizers operate at high vacuum (P ~14 kPa = 0.13 atm) thanks to the use of barometric condensers, and therefore they are catalogued as vessels. The shell of batch crystallizers is cylindrical (with a vertical axis) for mechanical and manufacturing convenience, easing the fabrication process and avoiding undesirable stress concentrations.

  Some designers adopted a conical enlargement of the body above the calandria to increase the capacity of the vessel without increasing the strike height and give a lower graining volume to final strike volume ratio. However, this was largely discredited after a negative effect on circulation was recognized (van der Poel et al., 1998). Numerous comparisons have shown that ‘straight-sided’ evapo-crystallizers perform better than ‘low-head conically-enlarged’ evapo-crystallizers (Chen and Chou, 1993).

- **Condensate and Incondensable Removal**

  Adequate arrangements for the removal of condensate and incondensable gases must be made. It is important that these details are given proper attention, as they can be the cause of under-performance if not properly designed. Condensate is generally removed from the
lowest point of the shell, from more than one outlet. It is good to be generous in sizing these outlets. The size of the condensate drains should be based on liquid outlet velocities of less than 0.45 m/s (1.5 ft/s) at maximum evaporation rate (Rein et al., 2004).

Incondensables need to be purged by the steam flow and are generally located at a point furthest from the steam inlet. This ensures a positive purging of incondensables. The best arrangement is a radial flow of steam from a vapor belt around the calandria. As the steam flow is radially inwards, the incondensables offtake arrangements should consist of two rings around the downtake, one at the top and one at the bottom of the calandria. Care needs to be given to the size and number of holes in the incondensables offtake rings to ensure uniform offtake around the downtake, and that sufficient quantity is vented. The two incondensable offtakes should be separately vented. Incondensable quantity is estimated as about 100 mg/kg of steam. In order to ensure total venting of incondensables, the amount vented should be 100 times the quantity of incondensables. Thus the total quantity to be vented should be about 1% of the steam flow to the calandria (Rein et al., 2004). Thermostatic valves can be used to bleed incondensables if steam economy is important.

Condensate and incondensable venting lines should not be run through the massecuite. Incondensable gases should be vented through the outside of the calandria, and not up through the massecuite. Likewise, condensate outlets should be positioned at the periphery of the evapo-crystallizer, and should not run from the bottom tube plate down through the massecuite.

• **Molasses and Syrup Feeding**

The syrup or molasses feed system should also represent minimum obstruction to the circulating massecuite. The feed should be introduced through the periphery of the evapo-
crystallizer below the calandria or through minimum sized pipes or channels on the floor of the evapo-crystallizer. If the feed is conditioned and at a higher temperature than the boiling massecuite, the feed must be directed under the calandria, so that the flash will aid circulation (Rein et al., 2004).

- **Assisted Circulation - Stirrers**

  Stirrers if properly designed can significantly improve the performance of an evapo-crystallizer. The assisted circulation improves heat transfer and so shortens the duration of the batch boiling, thus improving capacity. It has also been shown that stirrers improve the quality of high-grade sugar produced (Rein, 1988). This is a consequence of the better circulation leading to more homogeneous crystallization conditions within the crystallizer.

  The first stage of a strike is characterized by a high evaporation rate. An intense evaporation occurs, and stirrers do not have much effect in this period, suggesting that the effect of mechanical circulation is small in comparison to the circulation induced by vapor bubbles. As the strike height is increased, the hydrostatic pressure in the calandria increases the massecuite temperature, leading to a reduction in the temperature difference between massecuite and the heating vapor. During the last stages of the strike, the evaporation is minimal, resulting in a low bubble flow. The effect of the forced circulation becomes important at this stage, and is reflected in higher heat transfer coefficients with respect to natural circulation evapo-crystallizers (Austmeyer, 1986). This is shown in Table 2.1

  Mechanical agitators provide the option of achieving an acceptable heat transfer with a temperature difference as low as 20 °C, compared to at least 45-50 °C in the absence of a stirrer. The use of lower pressure vapors becomes possible (e.g. coming from
the 2nd or 3rd evaporator effect), allowing reductions in the factory steam requirements. A stirrer can be used with a smaller diameter downtake, thus enabling a larger heating area to be installed for a given evapo-crystallizer diameter. The tip speed of the stirrers is normally 5-7 m/s.

Table 2.1  Typical heat transfer coefficients in batch sugarcane evapo-crystallizers.

[Source: Bubnik et al., 1995]

<table>
<thead>
<tr>
<th></th>
<th>Natural circulation h (W/m².K)</th>
<th>With stirrer h (W/m².K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Start</td>
<td>570</td>
<td>640</td>
</tr>
<tr>
<td>End</td>
<td>32</td>
<td>224</td>
</tr>
</tbody>
</table>

There is no general agreement about the cost effectiveness of a stirrer. Its installation gives all the advantages associated with good circulation discussed earlier, and promotes circulation at the end of the strike. In contrast, high capital costs, air leakage and high power consumption, particularly at the end of the strike, work against the use of stirrers.

- **Assisted Circulation – Jigger Steam**

Sparging of steam underneath the calandria has often been used as a means of improving circulation. It has the effect of increasing the velocity of the massecuite through the tubes, thus increasing the rate of heat transfer to the massecuite and further improving circulation. The simplicity, low cost and absence of moving parts make this alternative particularly straightforward to put into practice. Venting of incondensables into the jigger arrangement can be done to reduce the consumption of vapor, if steam economy is important.
If the evapo-crystallizer is badly designed and/or the massecuite level gets too high, there comes a time with a highly viscous massecuite when the evapo-crystallizer stops boiling. Heat transfer in the tube is insufficient to cause significant boiling and movement of massecuite slows. This is a vicious cycle. As the heat transfer rate reduces, the rate of movement of massecuite drops, and heat transfer drops even further. Steam assisted circulation can get such an evapo-crystallizer to start boiling again. Thus the effect of steam assistance, often called ‘steam jigger’ is to enable a higher strike height to be achieved as well as a reduced boiling time (Rein et al., 2004).

It should be noted that the vapor admitted through the ring does not condense, but passes straight through the massecuite without causing any superheating or dissolution of crystal. A particular advantage of jigger steam is that it can be shut off at any time, unlike a stirrer, and can be adjusted to give the required degree of circulation.

Air injection: The injection of air instead of vapor has been adopted in some Spanish factories, allowing an increase in circulation with low energy consumption (15 % of that required by stirrers), low costs, and without breakage of crystals (van der Poel et al., 1998). Injection of compressed air at 750 kPa preheated to 65-70 °C beneath the calandria of A-strike crystallizers was reported to be effective in reducing the boiling times by 11 % and to increase recovery without an appreciable effect on the vacuum (Stobie, 1999). However this option requires a much larger vacuum pump or ejector to remove the additional air and is not being recommended here.

2.1.2 Continuous Evaporative Crystallizers

Since the early 1930s several designs of continuous crystallizers have been utilized by the sugar industry, initially without any success due particularly to the poor crystal
distribution size (Moor, 2001). The technology evolved gradually up to a point where many consider the continuous crystallizers as today’s best option.

Crystallization, as any other unit operation, is desirable to be a continuous process from several points of view: The continuous demand and supply of syrup, massecuite, steam, condensate, result in a smoother and more regular operation upstream, downstream, and in the crystallizer itself, easing the control of the plant and increasing the capacity. Other benefits include: Easier automation than batch processes; Lower invest and operation costs; Lower temperature differences are required, making possible the use of low-pressure vapors (e.g. vapor from 2\textsuperscript{nd} or 3\textsuperscript{rd} evaporation effects).

The most commonly used designs of continuous evaporative crystallizers in the sugarcane industry are horizontal, indicating the arrangement of several consecutive compartments or cells in the horizontal plane. The massecuite flows successively through the compartments, where syrup and/or molasses is fed individually as the crystals grow through a continuous process.

A vertical continuous crystallizer is also available (VKT-BMA), where the compartments are placed on top of each other. The vertical design is largely used in the production of sugar from beet, where the massecuites are less viscous, while only two sugarcane mills in the world use it.

Following are presented some of the most well known continuous evapo-crystallizer designs currently available for the production of sugar from cane:

- **Fives Lille-Cail (FCB) Continuous Evaporative Crystallizer**
  - First successful continuous crystallizer, France 1964.
  - Horizontal calandria tubes arranged in the longitudinal direction.
- The volume of the compartments is increased in the direction of the flow. In the case of 13 compartments, compartments 1 to 4 are each one half the volume of the compartments 5 to 12, and compartment 13 has twice the volume of each of the compartments 5 to 12.

- For A-strikes the feed syrup is dispersed in the free surface of the boiling massecuite using slowly rotating distributor arms located in the vapor disengagement space, limiting the incrustation.

- For C-strikes the syrup is feed through the bottom of the crystallizer.

Fig 2.3  FCB continuous evapo-crystallizer.
- **Tongaat-Hulett Continuous Evaporative Crystallizer (Fletcher Smith)**
  - Developed for sugar cane originally, South Africa 1983.
  - Floating vertical-tubed calandria, with vertical plates separating adjacent cells.
  - Horizontal top and bottom calandria plates.
  - 10 to 12 compartments of uniform size.
  - Vessel has an inverted heart-shape cross-section for optimized flow patterns.

[Photo source: Fletcher Smith - http://www.fletchersmith.co.uk/Assets/cvpcane.pdf]

Fig 2.4 Tongaat-Hulett continuous evapo-crystallizer.
• **Sugar Research Institute (SRI) Continuous Evaporative Crystallizer**

- Developed in Australia in the 1980s.
- Floating vertical-tubed calandria.
- Modular construction, each section of compartments has an independent calandria. Attaching new modules can expand the capacity. The volume of the cells is increased progressively.
- Top and bottom calandria plates are both inclined.

[3D Figure source: SRI - [http://www.sri.org.au/prodservices/equipmentdes/continuousva.jsp](http://www.sri.org.au/prodservices/equipmentdes/continuousva.jsp)]

Fig 2.5 SRI continuous evapo-crystallizer.
• **Bosch Continuous Evaporative Crystallizer (Honiron)**

  - The first application was in Westfield, Louisiana, 2001
  - Floating vertical-tubed calandria, with vertical plates separating adjacent compartments.
  - Cylindrical shell, simpler construction.
  - Horizontal top and bottom calandria plates.


  **Fig 2.6** Bosch continuous evapo-crystallizer.

While the floating calandria has proved to be an undesirable feature in batch crystallizers, it is practically the standard practice in modern continuous crystallizers. However, in continuous pans larger downtakes can be used, resulting in less circulation resistance, while in batch pans the peripheral downtake cannot be enlarged significantly when a low footing volume is to be maintained.
2.2  FUNDAMENTALS OF MULTIPHASE BOILING FLOWS

Gas-liquid flows are encountered very often in nature and have very important industrial applications where the buoyancy forces play a significant role, such as in boiling, distillation, flotation, air-lifting, aeration, hydrogenation, fermentation, oil extraction, sugar evaporative-crystallization, and so on. Accurate information about the different interactions in diluted and high void fraction gas-liquid systems is essential for the development of analytical tools to enhance the design and efficiency of multiphase processes. However, the experimental and numerical investigation of multiphase flows is complex in nature, and the traditional techniques applied in the study of single-phase fluid flow and heat transfer problems are limited when applied to multiphase flows, if not impracticable; preventing the development of solid analytic models, and making necessary the use of empirical and semi-empirical approaches in the solution of engineering problems.

Some factors that have been found to be important in the analysis of gas-liquid buoyancy-driven flows are discussed here, giving particular emphasis to the boiling in vertical pipe flows, which corresponds to the phenomena occurring in the calandria tubes and responsible for the circulation of the massecuite in sugar evaporative crystallizers.

2.2.1  Heat Transfer in Vertical Channel Flow

Boiling is a phase-change process from liquid to gas, usually driven by the heat transfer from a solid surface (e.g. walls of calandria tubes) to a fluid in contact (e.g. massecuite near the tube wall), which provides the energy required for the nucleation and growing of vapor bubbles, or evaporation. In most terrestrial applications significant buoyancy forces appear as a result of the density difference between the liquid and the
generated vapor, causing the motion of the fluid and placing the boiling within the convection heat transfer category.

Convection is classified in general as:

- **Natural - free**: Fluid motion due exclusively to buoyancy forces (e.g. conventional sugar evapo-crystallizer).

- **Forced**: Flow induced by external means as a pump, a fan, or a compressor. Buoyancy forces may participate (e.g. stirred sugar crystallizer).

The heat transfer analyses are normally performed for either constant wall temperature (e.g. heating steam condensation) or constant heat flux (e.g. heating electric resistances). It is customary to use the temperature difference between the heating wall \( (T_W) \) and the liquid boiling temperature \( (T_L) \) for the calculation of the heat transfer coefficients, and the external diameter as the characteristic dimension in the case of pipe flows.

\[
\Delta T = (T_W - T_L) \quad \text{Temperature difference, also called excess temperature}
\]

\[
h = \frac{q_w}{\Delta T} = \frac{\dot{Q}_w}{A_{HT} \cdot \Delta T} \quad \text{Boiling heat transfer coefficient}
\]

\[
Nu = \frac{hD}{k} \quad \text{Nusselt number}
\]

Gunther and Kreith (1949) demonstrated that during boiling a superheated film is formed near to the heating surface. It was argued that the vapor bubbles exert a micro-convection effect, and without them the vapor film would impose an enormous thermal resistance. Rohsenow and Clark (1951) demonstrated that the agitation produced by the bubbles generated near the heating surface, and in a minor degree the large amount of energy that a fluid can absorb during vaporization (latent heat), trigger high heat transfer rates during
boiling and make it possible to achieve heat transfer coefficients significantly larger than those found in single-phase convection.

Boiling tests have been carried out for different situations and several correlations have been proposed. Pool boiling (natural circulation) is a popular study case due to its relative simplicity. The internal boiling in heated channels has significant hydrodynamic differences with pool boiling, and has been studied mainly under forced convection conditions. In internal flows the strong interactions between the phases bring several possible flow patterns that are sensible to the flow properties and can change dramatically as more vapor is generated. The particular case of boiling in tubes has received much attention due the extensive industrial applications and its importance in energy conversion processes. However, the complexity and dispersion in the results has prevented the consolidation of generalized expressions.

• **Empirical Correlations**

Convection heat transfer correlations for single phase flows have been developed empirically by recording the heat transfer rate as the temperature difference varies. Normally the results for each fluid can be represented as a straight line on a log-log scale, and consequently algebraic expressions of the following form are used to describe the process (Incropera and Dewitt, 1992):

\[ Nu = C \cdot Re^m \cdot Pr^n \]

In case that the constant values (C, m, and n) are independent of the fluid, a single generalized correlation can be used to describe the convection heat transfer using the ratio \( Nu/Pr^n \) vs. Re. For evaluation of the dimensionless numbers the variations across the boundary layer are considered by evaluating the properties at the film temperature.
• Single-phase Convection

Sieder and Tate (1936) proposed the following heat transfer correlation for single-phase convection of internal laminar flows heated at uniform wall temperature:

\[
Nu = 1.86 \left( \frac{Re*Pr}{L/D} \right)^{1/3} \left( \frac{\mu}{\mu_{FILM}} \right)^{0.14} \quad \text{Sieder and Tate (1936)}
\]

\(0.48 < Pr < 16700\)

Using the Chilton-Colburn analogy expressions for the convection heat transfer in turbulent single-phase flow in circular tubes have been developed, which are valid for small temperature differences, where no significant property variations occur (Incropera 1992):

\[
Nu = 0.023 \ast Re^{4/5} Pr^{1/3} \quad \text{Colburn equation}
\]

\[
Nu = 0.023 \ast Re^{4/5} Pr^n \quad \text{Dittus - Boelter (1937)}
\]

\((n=0.4 \text{ heating}, n=0.3 \text{ cooling})\)

\(Re\geq10000; \ 0.7<Pr<160; \ L/D\geq10\)

In case of large temperature differences the following equation proposed by Sieder and Tate can be used (Incropera, 1992):

\[
Nu = 0.027 Re^{4/5} Pr^{1/3} \left( \frac{\mu}{\mu_{FILM}} \right)^{0.14} \quad \text{Sieder and Tate (1936)}
\]

\(0.7 < Pr < 16700; \ Re\geq10000\)

\(L/D\geq10\)

• Convective Boiling in Pipes

As discussed previously, during boiling processes the vapor bubbles generated agitate their surroundings, mixing the hot liquid close to the exchange surface with the rest of the stream, and triggering high heat transfer coefficients typical of boiling regimes. Therefore, a
fundamental difference with respect to the single-phase convection process exists and the heat transfer correlations should include this component.

Sieder and Tate (1936) proposed a heat transfer correlation for forced convective-boiling that exhibits the same typical empirical form used for single-phase convection:

\[ N_u = 0.027 \operatorname{Re}^{4} \operatorname{Pr}^{1.4} \frac{\mu_{L}}{\mu_{w}} \]  

Sieder-Tate (1936)

Forster and Zuber (1954) based on a similarity rationale argued that the heat transfer in boiling pipe flows can be described as a function of the Reynolds and Prandtl numbers, and therefore an empirical expression with the conventional form could be assumed. Using as reference typical values, the exponents were expected to be \(0.5 < m < 0.8\) and \(n \sim \frac{1}{3}\). Forster and Zuber (1954) compared experimental data reported for boiling of several liquids (n-pentane, benzene, ethanol, water) with their empirical equation, finding that all cases can be represented with confidence using the following expression:

\[ N_u = 0.0015 \operatorname{Re}^{0.62} \operatorname{Pr}^{1.3} \]  

Forster and Zuber (1954) \(^5\)

Chen (1963) proposed an equation for the computation of the heat transfer in forced boiling pipes adding two components: the single-phase liquid \((h_L)\) and the nucleate boiling \((h_{NB})\) heat transfer:

\[ q_w = h_L(T_w - T_L) + h_{NB}(T_w - T_{SAT}) \]  

Heat flux

\[ h = \frac{q_w''}{(T_w - T_{SAT})} \]  

Effective heat transfer coefficient

Multipliers are used to correct the correlations used for each mechanism, this in order to include the interaction between the two heat transfer components.

---

\(^5\)The characteristic length is defined based on bubble dynamic considerations:

\[ L = \frac{C_p \Delta T \rho_l \sqrt{\pi \alpha}}{\delta \rho} \sqrt{\frac{2 \sigma}{\Delta P}} \sqrt[3]{\frac{\rho_l}{\Delta P}} \]
The classic Dittus-Boelter equation (1937) can be applied as if the liquid phase was flowing alone, but making use of a multiplier (F) that is computed as a function of the Martinelli factor (X_{LM}):

\[
h_L = 0.023 \frac{k}{D_h} \left( \frac{G}{
\begin{array}{c}
\text{Liquid phase convection heat transfer}
\end{array}
\right) \]

\[
Re_L = \frac{(1-x)G}{\mu_L}
\]

\[
Pr_L = \frac{Cp_L \mu_L}{k_L}
\]

\[
X_{LM} = \left( \frac{x}{1-x} \right)^{0.9} \left( \frac{\rho_L}{\rho_G} \right)^{0.5} \left( \frac{\mu_G}{\mu_L} \right)^{0.1}
\]

\[
F = 1 \quad 1/X_{LM} \leq 0.1 \quad \text{Multiplier}
\]

\[
F = 2.35 \left( \frac{1}{X_{LM}} - 0.213 \right)^{0.736} \quad 1/X_{LM} > 0.1
\]

A correlation proposed by Forster and Zuber (1955) is used for nucleate boiling, considering as a multiplier (SH) the ratio between the effective superheat of the wall and the total superheat of the wall:

\[
h_{NB} = c \left( T_w - T_{\text{SAT}} \right)^{0.24} \times SH
\]

\[
c = 0.00122 \left( \frac{Cp_{\text{SAT-VAP}} k_{\text{SAT-VAP}}}{\sigma} \right)^{0.5} Pr_L^{-0.29} \rho_{\text{SAT-VAP}}^{0.25} \left( \frac{Cp_{\text{SAT-VAP}} \rho_{\text{SAT-VAP}}}{\rho_{\text{SAT-LIQ}} h_{fr}} \right)^{0.24} \Delta P^{0.75}
\]

\[
\Delta P = P_{\text{SAT}_{\text{MIN}}(T_w, 647)} - P
\]

\[
Re_{MOD} = 0.0001 \times Re_L \times F^{1.25} \quad \text{Modified Reynolds number}
\]

\[
SH = \frac{1}{1 + 0.12 \times Re_{MOD}^{1.14}} \quad 0 < Re_{MOD} \leq 32.5
\]
\[ SH = \frac{1}{1 + 0.42 \text{Re}_{\text{MOD}}^{0.78}} \quad 32.5 < \text{Re}_{\text{MOD}} \leq 70 \]

\[ SH = 0.0797 \exp \left(1 - \frac{\text{Re}_{\text{MOD}}}{70}\right) \quad 70 < \text{Re}_{\text{MOD}} \]

Bejan (1993) presents a similar approach for the solution of forced convection boiling in smooth tubes, combining the Rohsenow correlation for nucleate boiling with the Dittus-Boelter correlation for single-phase convection\(^6\) to estimate the heat flux. This approach is reported to work better for sub-cooled boiling without large generation of vapor near the hot surface:

\[ q'' = q''_{DB} + q''_{NB} \]

\[ \text{Nu}_{DB} = 0.019 \text{Re}^{4/3} \text{Pr}^{0.3} \]  
Dittus-Boelter, single-phase

\[ q''_{NB} = \mu_l h_f \left(\frac{g \Delta \rho}{\sigma}\right)^{1/2} \left[\frac{C_p \left(T_W - T_{SAT}\right)}{\text{Pr}^{1/3} \mu \left(C_{SF} \cdot h_f^{1/3}\right)}\right]^2 \]  
Rohsenow, nucleate boiling

Where S and C\(_{SF}\) are constants dependent on the fluid, the surface material and its roughness (source: Bejan, 1993):

<table>
<thead>
<tr>
<th>Material</th>
<th>Surface Condition</th>
<th>S</th>
<th>C(_{SF})</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water-Copper</td>
<td>Polished</td>
<td>0.013</td>
<td>1.0</td>
</tr>
<tr>
<td></td>
<td>Scored</td>
<td>0.068</td>
<td>1.0</td>
</tr>
<tr>
<td>Water-Stainless steel</td>
<td>Ground and polished</td>
<td>0.008</td>
<td>1.0</td>
</tr>
<tr>
<td></td>
<td>Mechanically polished</td>
<td>0.013</td>
<td>1.0</td>
</tr>
<tr>
<td></td>
<td>Teflon pitted</td>
<td>0.0058</td>
<td>1.0</td>
</tr>
</tbody>
</table>

### 2.2.2 Flow Regimes in Vertical Gas-liquid Channel Flow

In multiphase flows the transfer of mass, momentum, and energy between the phases, as well as between the flow and boundaries, depends considerably on the flow regime. As a

\(^6\) The coefficient C = 0.023 is replaced for C = 0.019
consequence regime dependent correlations must be applied for numerical modeling, and the first step for coupling the differential equations governing the transfer mechanisms is recognizing the dominating flow regime (Kolev, 2002).

Studies on the boiling in vertical tubes have demonstrated that the quality and void fraction increase progressively along the heated pipe, while the flow experiences changes in the regime from the single liquid-phase to bubbly, slug, churn, annular, mist, and finally the single vapor-phase flow (Fig 2.7).

Fig 2.7 Flow regimes and boiling mechanisms in a vertical heated tube.
The boiling in tubes, like evaporators and boilers, is one of the most complicated convective problems found in industrial applications (Kreith et al., 1999) as the flow experiences multiple heat transfer and two-phase regimes along the pipe.

- **Sub-cooled Boiling and Nucleate Boiling**

  At the entrance of the pipe the conditions correspond to conventional single-phase flow, where the standard entry and heat transfer correlations can be applied. The boiling starts soon with the discrete nucleation of small bubbles that are generated at the heating surface.

- **Bubbly Flow**

  This regime is characterized by an approximately homogeneous distribution of the gas-phase in the form of discrete bubbles, which are small in size with respect to the pipe diameter. As the vapor content goes up, the density of bubbles will be higher, increasing the probability of collisions and coalescence. The bubbly regime is kept up to a critical limit where the rate of collisions becomes too high, resulting in the formation of large bubbles known as ‘Taylor bubbles’, and the transition to the slug regime. For adiabatic vertical channel flow the transition limits between the bubbly and slug regimes have been identified in terms of the void fraction:

  Geometrical limit of bubbly flows: The maximum content of bubbles that can coexist together has been determined theoretically assuming that the bubbles can be treated as non-oscillating spherical particles and organized in a rhomboid array, where all the particles are adjacent ‘touching’ each other. Under such conditions the void fraction would be $\alpha_{\text{max}}=0.74$, a limit that is known as ‘Maximum packing density’, and demonstrates that a bubbly flow cannot exist above this void fraction limit (Kolev, 2002). A similar analysis, but assuming a
cubic lattice array of spherical particles would lead to a maximum limit of $\alpha_{\text{max}} = 0.52$ for bubbly flows (Cheremisinoff, 1986).

Practical limit of bubbly flows: For most practical conditions, excepting small bubbles\(^7\), the oscillations, deformation, and randomness in the gas motion make the void fraction limit much lower than the ‘maximum packing density’, since some space is required between the bubbles to prevent significant rates of collision and coalescence to occur.

Radovich and Moissis (1962) proposed a theoretical limit for the transition from bubble to slug regimes occurring when the void fraction reaches $\alpha = 0.25$. The validity of this value would be later confirmed experimentally by Griffith and Snyder (1964), and Hasan and Kabir (1988). In general, the experimental results have indicated that the transition from bubbly to slug flow occurs around $\alpha = 0.25–0.30$. Reported data agrees on the absence of coalescence for void fractions below $\alpha = 0.20$, while normally no bubbly flows are found when the void fraction exceeds $\alpha > 0.35$ (Cheremisinoff, 1986).

Kolev (2002) indicates that turbulent bubbly flows occurring in nature and technical facilities have been observed to have a void fraction up to $\alpha = 0.25 – 0.30$. A similar limit for the transition bubbly-slug has been observed for annular vertical flows (Hasan and Kabir, 1992)

- **Slug Flow (Plug Flow, Piston Flow)**

This regime is characterized by a high coalescence and formation of large bubbles, known as ‘Taylor bubbles’, which have almost the same diameter of the pipe and contain most of the gas phase (Cheremisinoff, 1986). Taylor bubbles are bullet-shaped, and

\(^7\) Bubbles with size $D \leq 6^{*}\lambda_{RT}$ behave as solid particles and display very low coalescence, so the void fraction can reach high values while the bubbly regime is still conserved (e.g. as high as $\alpha=0.54$).

The Raleigh-Taylor wavelength is given by the expression:

$$\lambda_{RT} = \sqrt{\frac{g \cdot \Delta \rho}{\sigma_{L}}}$$
separated by ‘slugs’ of pure liquid (plug) or a mixture of small bubbles and liquid, giving the flow the appearance of a train of bubbles.

- **Churn Turbulent Flow (Froth Flow)**

  Churn flow is similar to slug flow, but more chaotic and frothy. The bullet-shape of the Taylor bubbles becomes narrower and highly distorted, and the slugs are broken by the high gas concentration, resulting in an intermittent fall and rise of the slugs, which causes an oscillatory motion of the liquid phase with a severity that increases with the flow rate (Cheremisinoff, 1986). Churn flow has not been studied as much as the other regimes because of its chaotic nature (Hasan and Kabir, 1992).

  Differentiation between slug and churn regimes is difficult. The oscillatory motion of the liquid phase can be considered as the main difference, and a particular feature of the churn regime, while in slug flow the liquid separating two consecutive Taylor bubbles moves at approximately constant velocity (Cheremisinoff, 1986).

  A criteria for the transition slug-churn in circular channels was proposed by Brauner and Barnea (1986), suggesting that it occurs when the void fraction in the slug that separate the Taylor bubbles is around \( \alpha = 0.52 \). Inclination of the pipe, even by a small amount, reduces the occurrence of the churn flow, and for inclinations with respect to the vertical above 20° the churn regime practically disappears.

- **Annular Flow**

  A further increase in the gas content results in the separation of the two phases in an annular regime, where the liquid phase locates near to the wall(s) of the channel, while the gas phase occupies the central core. This regime is found at very high void fractions.
2.2.3 Hydrodynamic Properties of Rising Bubbles

The intended numerical analysis of the flow in evaporative crystallizers makes use of the Eulerian-Eulerian multiphase approach, treating each phase as a continuum and introducing volumetric fraction terms in the Navier-Stokes equations, which are then solved for each phase individually, but taking into account the interactions with the other phases through exchange terms that couple the transport equations. An exchange coefficient \(K_{ij}\) is used for the coupling of the momentum equations, which is usually calculated as a function of empirical or semi-empirical drag coefficient correlations, allowing the computation of the momentum transfer between the phases at each point.

Unfortunately, the dynamics of fluid particles is far from being well understood, and no generalized correlations for the exchange of momentum have been developed yet, but instead a large number of expressions have been proposed, mainly for the drag of small single bubbles rising under ideal conditions, which can differ significantly from practical applications. In gas-liquid internal flows different flow regimes are possible (eg. bubbly, slug, churn, annular), and it has been recognized that the exchange of mass, momentum, and energy are strongly dependent on the flow regime (Kolev, 2002). Although from the theoretical point of view the Eulerian-Eulerian CFD approach is suitable for any flow regime, the selection of appropriate closure relations, such as the drag coefficient for the momentum equations, is critical for the correct simulation of gas-liquid multiphase flows.

- **Dimensionless Numbers Involved**

The following dimensionless numbers govern the drag coefficient in gas-liquid flows: Bubble Reynolds number (Inertia/Viscous), Eötvös (Gravity/Surface tension), and Morton (Gravity/Molecular fluid acceleration).
\[ \text{Re} = \frac{\rho_L D_G \Delta u}{\mu_L}; \quad Eo = \left[ \frac{D_G}{\lambda_{RT}} \right]^2; \quad Mo = \frac{g \Delta \rho \mu_L^2}{\sigma_L \rho_L \sigma_L} \]

For the calculation of the Eötvös number the Rayleigh-Taylor instability wavelength is given by:
\[ \lambda_{RT} = \sqrt{\frac{\sigma_L}{g \Delta \rho}} \]

Other important parameters for gas-liquid flows are the Froude (Inertia/Gravity), Weber (Inertia/Surface tension), and the gas, liquid, and two-phase Reynolds (Inertia/Viscous) numbers:

\[ Fr = \frac{G}{g D_h \rho_L^2}; \quad We_{\text{BUBBLE}} = \frac{\rho_L \Delta u^2 D_G}{4 \sigma_L}; \quad We_{\text{CHANNEL}} = \frac{G^2 D_h}{\rho_L \sigma_L} \]

\[ Re_G = \frac{\rho_G D_h \mu_G}{\mu_G}; \quad Re_L = \frac{\rho_L D_h \mu_L}{\mu_L}; \quad Re_{TP} = \frac{G D_h}{\mu_L} \]

- If We is small the surface tension effects are important.
- For Re >> 1, the surface tension effects can be neglected if We >> 1
- When \( Fr < 1 \) the flow is sub-critical, and small surface waves can move upstream
- When \( Fr > 1 \) the flow is supercritical, and small surface waves are carried downstream

- **Drag**

A particle or object that travels in a continuum experiences a dissipative force that opposes the movement and is known as ‘drag force’. As a consequence of the movement a non-uniform pressure distribution is developed around the particle and forces such as drag and lift are generated. Additionally, viscous dissipation takes place within the boundary layer and contributes to increase the drag force.
The drag force is usually defined as:

\[ F_D = \frac{1}{2} \rho C_D A_{XS} \Delta u^2 \]

Where \( A_{XS} \) is the frontal area of the body facing the flow, or largest cross-sectional area normal to the flow.

The drag coefficient (\( C_D \)) is a dimensionless representation of the drag, indicating the time averaged drag force (\( F_D \)) that a flow exerts on the moving object or particle. The drag coefficient is expected to be identical for dynamically similar situations (Kundu, 1990), a characteristic that has a very important practical value, since it is possible to generalize experimental results and be applied in multiple conditions, reducing the need for experimental work.

Data and correlations for the drag coefficient have been obtained for a number of geometries, and are typically presented as a function of the Reynolds number in logarithmic scale plots. In multiphase flows the drag plays a fundamental role in the transfer of momentum between the phases. Generally averaged empirical correlations developed for single or multiple particles are used for numerical analysis purposes, allowing computations to be performed with relatively ‘coarse’ grids (Kolev, 2002).

- **Buoyancy**

As a consequence of the earth gravity field, upwards buoyancy forces act on objects or particles that are submerged or floating in a fluid that has a higher density. Conversely, for objects with a density higher than the surrounding media, the gravity force will act downwards, trying to sink the object.

The principles of Archimedes describe the origin of the buoyancy force:
I. A completely submerged body experiences a vertical upward force equal to the weight of the displaced fluid.

II. A floating or partially submerged body displaces its own weight in the fluid in which it floats.

Then, the buoyancy force acting on a particle of the [G] phase, with volume \([V_B]\), which is submerged in a continuum/primary phase \([L]\), is given by:

\[
F_B = (\rho_L - \rho_G) \cdot g \cdot V_B = \Delta \rho \cdot g \cdot V_B
\]

- **Balance Buoyancy - Drag**

An unbounded object or particle \((G)\) submerged in a media with higher density \((L)\) while exposed to the earth gravitational field will be subjected to buoyancy and drag forces acting in the vertical direction. Initially the buoyancy overcomes the drag, and consequently a vertical positive acceleration \((\hat{\imath})\) is produced. After a short time the drag and buoyancy forces balance each other, the acceleration stops, and the particle is said to reach its terminal velocity.

\[
F_D = F_B \\
\frac{1}{2} \rho_L \cdot C_D \cdot A_{XS} \cdot \Delta u^2 = \Delta \rho \cdot g \cdot V_B \\
\Delta u^2 = \frac{2 \cdot g \cdot \rho_L \cdot V_B}{C_D \cdot \rho_L \cdot A_{XS}}
\]

In most gas-liquid flows the density of the gas phase is negligible with respect to the liquid phase, and the following simplification can be made:

\[
\frac{\Delta \rho}{\rho_L} = \frac{(\rho_L - \rho_G)}{\rho_L} \approx \frac{\rho_L}{\rho_L} = 1
\]

For an ideal spherical particle the volume and cross-sectional area are given by:

\[
A_{XS} = \frac{\pi}{4} D_G^2 \quad ; \quad V_B = \frac{\pi}{6} D_G^3
\]
Then the velocity difference between the phases, or terminal velocity, can be determined as a function of the gravity constant, the bubble size, and the drag coefficient.

\[ \Delta u^2 = \frac{4}{3} g \frac{g^* D_G}{C_D} \]

Fig 2.8 Buoyancy and drag forces acting on an ideal spherical particle.

The developed expression shows that the balance buoyancy-drag is determined in part by the ratio between the volume of the bubble \((V_B)\) and its frontal area facing the flow \((A_{XS})\). For analytical purposes it is customary to assume that bubbles have a constant spherical shape. However, in reality the bubbles can assume different geometries according to their size, the flow conditions, the flow regime, and the properties of the fluid, even when the surface tension forces try to maintain a spherical shape. Data obtained for all liquids has shown that generally after \(Re>100\) the bubbles experience deviation from the ideal spherical shape (Moore, 1965).

- **Solid Spherical Particles**

Stokes (1851) proposed the first solution for the flow around a solid sphere at low Reynolds numbers. Oseen (1910) proposed later an improved approximation.

\[ C_D = \frac{24}{Re} \]

\[ \Delta u = \frac{1}{18} \frac{\Delta \rho g D_G^2}{\mu_L} \]

Stokes solution, solid spheres (1851);
Experimental results have shown that the two previous correlations give reasonable results at \(\text{Re} < 5\) (Kundu and Cohen, 1990). Both expressions imply that the drag force is directly proportional to the velocity, equivalent to saying that the drag coefficient is inversely proportional to the Reynolds number. This concept is usually known as the Stokes law of resistance, and is valid only for low Reynolds numbers, where the inertial forces are small in comparison with the viscous forces.

The drag coefficient becomes independent of the viscosity at \(\text{Re} \sim 1000\), when the inertial forces become dominant and the drag is determined mainly by a wake formation behind the sphere. The drag coefficient takes a constant value \(C_D \sim 0.44\) (Fig 2.9).

Transition to the turbulent regime occurs at a critical Reynolds number \(\text{Re} \sim 500000\), where the drag coefficient drops quickly before a post-critical Reynolds where it start rising again.
• Gas Particles - Bubbles

Deformation of the boundary (gas-liquid interface) and internal circulation differentiate the phenomena occurring in bubbles from solid particles. While solid particles are separated from the fluid by a rigid and non-slip boundary, in the case of fluid particles, such as drops and bubbles, a deformable and zero-tangential-stress boundary is observed, introducing significant differences in the drag and production of vorticity (Bertola, 2004).

A critical parameter in the study of bubbly flows is the bubble size, which has a strong effect on the balance between buoyancy and drag, influencing the transfer of momentum between the phases and determining in part the flow regime. Therefore, an adequate control of the bubble size and dispersion is an important condition for the successful execution of experimental tests in multiphase flows. Researchers have commonly used capillary tubes, membranes with fine pores and spargers for the generation of bubbles, observing that the volume of the generated bubbles is proportional to the surface tension and inversely proportional to the density of the liquid, while other parameters such as temperature, pressure, and viscosity have a small influence.

Bubbles injected in water columns using orifices smaller than D<0.4 mm tend to be spherical, and travel vertically upward. If the orifice is between 0.4<D<4 mm, the bubbles, although spherical at the orifice, become ellipsoidal quickly and travel upwards following a zigzag pattern. If the orifice is larger than D>4 mm, the bubbles are unstable, and can assume different shapes. However, in the case of very viscous liquids the bubbles tend to preserve a symmetrical shape and a vertical path. The following empirical relation gives the size of generated bubbles in air-water systems at 20°C (Soo, 1967):

---

8 It could be argued that the straight rise of small bubbles is responsible for the delay in the transition to the slug/churn regimes, since the collisions are limited to the vertical plane, reducing the probability of coalescence.
\[ R^*_G = 9.05 / R_{ORIFICE} \]
\[ V_B = 0.231 * D_{ORIFICE} \]

More general expressions for calculating the size of bubbles generated blowing gas through a horizontal circular orifice are (Soo, 1967):

\[ R_G = \left[ \frac{3}{2} \frac{R_{ORIFICE} \cdot \sigma_L}{g^* (\rho_L - \rho_G)} \right]^{1/3} \]
\[ V_B = \frac{\pi D_{ORIFICE} \cdot \sigma_L}{g^* (\rho_L - \rho_G)} \]

Another distinctive feature of fluid particles is the ability to merge and form larger particles, or break into smaller ones. Bubbles in media such as tap water, distilled water, and olive oil, are characterized by a high tendency to merge. Conversely, the presence of other substances such as alcohols, organic acids, or salt at a large concentration, tends to reduce the coalescence to a minimum. Because of the reduction caused in the merging, these substances are often added to water during experimental studies, requiring at least the following concentrations:

Table 2.2  Dilution for minimizing the merging of bubbles in water

[Source: Soo, 1967]

<table>
<thead>
<tr>
<th>Solute</th>
<th>Dilution (weight water:solute)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sodium chloride</td>
<td>100</td>
</tr>
<tr>
<td>Methyl-alcohol</td>
<td>600</td>
</tr>
<tr>
<td>Ethyl-alcohol</td>
<td>2000</td>
</tr>
<tr>
<td>Acetic acid</td>
<td>2000</td>
</tr>
</tbody>
</table>

The bubbly flow is the simplest gas-liquid regime, and has been investigated extensively by all experimental and numerical means. A large quantity of information has been presented about bubble dynamics, particularly for single bubbles in low-viscosity
stationary liquid media. However, there are discrepancies between the reported data, and no generalized correlations have been identified (Perry, 1997)

- **Undisturbed Bubbles**

  Bubbles at very low Reynolds numbers tend to adopt generally a spherical shape. In this case the distortion is negligible, and the gas particles are ‘undisturbed’. The drag coefficient of undisturbed bubbles depends on the velocity difference between the phases (slip velocity) and the viscosity of the liquid phase (\(\mu_L\)).

  **Stokes Regime (Re≤16):** Hadamard and Rybczynski (1911) developed an analytical solution for the drag exerted on ideal bubbles (spherical, pure media, mobile interface) that includes the internal circulation effect. The solution was derived coupling the Stokes stream functions for the outer and inner fluids, and indicates that the drag on a spherical bubble rising with straight motion under creeping flow (Re<1) would be:

  \[
  C_D = \frac{8}{Re} \left( \frac{3\chi + 2}{\chi + 1} \right)
  \]

  Where the viscosity ratio is \(\chi = \frac{\mu_G}{\mu_L}\).

  This theoretical solution indicates that the drag of spherical bubbles would be proportional to the velocity difference, like in the solid sphere case, but exhibiting a lower drag coefficient as a result of the internal circulation. Since the viscosity of the gas is normally very small in comparison to the liquid phase viscosity, the drag coefficient for bubbles under ideal conditions can be approximated to (Snabre and Magnifotcham, 1998):

  \[
  C_D = \frac{16}{Re} \quad \text{Mobile interface, pure system (e.g. twice distilled water)}
  \]

  In normal situations it is rare to have a pure system, and the presence of contaminants makes the bubbles ‘stiffer’. As a consequence, for low Reynolds number bubbles the internal
circulation is often negligible, and the dynamic behavior may be approximated by the solid sphere case (Snabre and Magnifotcham, 1998) with the drag coefficient as given by the Stokes law.

$$ C_D = \frac{24}{Re} $$  \hspace{1cm} \text{Rigid interface, non-pure system (e.g. tap water).}

Viscous Regime (Re>16): In this state the drag dependence on the slip velocity becomes non-linear. Numerous expressions covering this regime have been proposed; some of the most well known correlations are:

$$ C_D = \frac{24}{Re} \left( 1 + \frac{3}{16} Re \right) $$  \hspace{1cm} \text{Oseen (1910), originally for solid spheres}

$$ C_D = \frac{24}{Re} \left( 1 + 0.15 * Re^{0.687} \right) $$  \hspace{1cm} \text{Schiller and Naumann (1935)}

The symmetric exchange model available for the Eulerian-Eulerian multiphase solver of the CFD code FLUENT is recommended for bubbly flows, which makes use of the correlation proposed by Schiller and Naumann (1935) to do the coupling between the momentum equations for Re<1000. Durst et al. (1986) studied the application of the Eulerian-Eulerian approach, and concluded that the Schiller and Naumann correlation is satisfactory for the modeling of solid spherical particles between 1<Re<1000. Kolev (2002) recommends the Schiller and Naumann correlation for bubbles as long as Re<500.

The studies reported in the literature on bubbly flows usually involve small bubbles at low void fraction, where the gas particles tend to be approximately spherical and have a negligible internal circulation, behaving similarly to solid spheres. Under such conditions the application of the drag models developed originally for solid particles in the modeling of bubbly flows is reasonable, and leads to acceptable predictions.
Haberman and Morton (1953), and Peebles and Garber (1953) performed two comprehensive studies on the rise of small bubbles, obtaining qualitative agreement, but ambiguous differences from a quantitative point of view (Moore, 1965). Wang et al. (1994) presented the following correlation based on the experimental data obtained by Haberman and Morton (1953) for clean bubbles in distilled water:

\[
\ln(C_D) = \ln 24 - \ln \text{Re} \quad \text{[Same as Stoke’s law]} \quad \text{Re} \leq 1
\]

\[
\ln(C_D) = 2.7 - 0.336 * \ln \text{Re} - 7.14 E - 02 * (\ln \text{Re})^2 \quad 1 < \text{Re} \leq 456
\]

\[
\ln(C_D) = -51.8 + 13.2 * \ln \text{Re} - 0.824 * (\ln \text{Re})^2 \quad 456 < \text{Re} \leq 4000
\]

Experiments have shown that only small bubbles (e.g. \(D_G < 3\) mm in water) are approximately spherical and rise straightly, while a further increase in the size results in ellipsoidal bubbles (3 < \(D_G < 8\) mm) that travel in a zig-zag or helical path (Abou, 1986). In case that the Weber number remains small, the bubbles may keep a spherical shape, even at high Reynolds numbers. Two correlations for ellipsoidal bubbles are:

\[
C_D = \frac{16}{\text{Re}} + 2 + \frac{2}{5} \text{Re} * \ln \frac{\text{Re}}{2} + 1.33 \frac{\text{We}}{\text{Re}} \quad \text{Taylor and Acrivos (1964)}
\]

for \(\text{Re} < 2\) and \(100 < \text{Mo} < 100000\)

\[
C_D = \frac{8}{3} - \frac{16}{3 * \text{We}} \quad \text{Mendelson (1967), for Mo ~ 10^{10}}
\]

- **Distorted Bubbles**

Bubbles can take different shapes according to the pressure distribution along the gas-liquid interface (Durst et al., 1986), where additionally the surface tension, viscosity and inertial forces are interacting continuously. As the bubble size and Reynolds number increase the spherical and ellipsoidal shapes are replaced by distorted geometries, and the
surrounding flow develops a vortex behind the bubble, creating a wake that disturbs the bubble itself and the bubbles that follow.

The strongly distorted or cap bubbles regime is reached when the Weber number is around $\text{We} \sim 20$. Davies and Taylor (1950) studied these cap bubbles, observing a substantial increase in the drag, and reduction of the rise velocity with respect to spherical bubbles of the same volume. The rise velocity was reported to be a function of the dimension of the cap ($R_{\text{CAP}}$) or equivalent volumetric radius ($R_G$), observing for inviscid fluids a constant drag coefficient $C_D = 2.63$; Haberman and Morton (1953) found a similar constant $C_D = 2.60$; In general, the experimental values of the drag of cap bubbles are located between $C_D = 2.6-2.7$; far above the value encountered in the case of solid particles at high Reynolds number ($C_D \sim 0.44$, Fig 2.9).

![Terminal velocity formula](attachment:terminal_velocity.png)

$V = \frac{4}{3} \pi R^3_G$

Terminal velocity (Davies 1950)

$$U_t = \sqrt{\frac{g}{\rho} R_G} = \frac{2}{3} \sqrt{\frac{g}{\rho} R_{\text{CAP}}}$$

Conical angle

$\phi \approx 40-55^\circ$

Fig 2.10 Highly distorted ‘cap’ bubble.

It has been proposed that the drag coefficient of distorted bubbles depends only on the size of the bubble, liquid density and surface tension, and unlike undistorted bubbles, it would be independent of the slip velocity and viscosity (Kolev, 2002).

$$C_D = \frac{2}{3} \frac{D_G}{\lambda_{RT}}$$

Valid for $\frac{24}{\text{Re}} \left(1 + 0.1 \cdot \text{Re}^{0.75}\right) \leq \frac{2}{3} \frac{D_G}{\lambda_{RT}} \leq \frac{8}{3}$
For caps rising in viscous fluids at high Morton numbers, Darton and Harrison (1974) proposed the following correlation:

$$C_D = 2.7 + \frac{16}{\text{Re}}$$

Valid for Re > 1.2 and Eo > 40

Tomiyama (1995) proposed the following correlation for highly distorted bubbles (caps):

$$C_D = \frac{8}{3} \frac{Eo}{Eo + 4}$$

Valid for $$4 \leq \frac{Dg}{\rho RT} ; \text{Re} \leq 100 000 ; 0.01 \leq Eo \leq 1 000$$

At large Reynolds numbers (Re > 5 000) the bubble rise velocity does not depend on liquid viscosity or bubble size. While the rise velocity of small bubbles is controlled by viscosity and surface tension, the turbulence seems to be the controlling factor in the case of large bubbles (Soo, 1967). Harmathy (1960) proposed the following correlation, which implicitly states that for large bubbles the drag is an exclusive function of the Weber number:

$$\Delta u = 1.53 \left( \frac{g \Delta \rho \sigma_L}{\rho_L^2} \right)^{1/4}$$

High Reynolds bubbles (Harmathy, 1960)

Angelino (1966) proposed an empirical correlation for the terminal velocity of large bubbles reported to be useful in practical situations:

$$\Delta u = C * V_B^m$$

Large bubbles (Angelino, 1966)

$$C = \frac{25}{1 + 0.33 \text{Mo}^{0.29}}$$

Where units: $$V_B (\text{cm}^3)$$; $$\Delta u (\text{m/s})$$

$$m = 0.167 \times (1 + 0.34 \text{Mo}^{0.24})$$

- **Effect of Impurities**

Fluid particles in motion, such as undistorted bubbles and drops, develop internal circulation, which has the effect of reducing the drag to values as low as 1/3 with respect to a solid particle of the same size and Reynolds number. The presence of contaminants
however, reduces the mobility of the gas-liquid interface, affecting the flows inside and outside the bubble (Durst et al., 1986), impairing the internal circulation, and increasing the drag coefficient, which under certain conditions can reach the values typical of solid particles (Kolev, 2002). As an example, it is known that the presence of small amounts of surface-active impurities is enough to make the interface of small bubbles ‘stiffer’, which can behave as solid particles, experiencing a substantial increase in the drag coefficient (Moore, 1965).

The terminal velocity of very small bubbles is usually in agreement with the Stokes law, indicating first that ‘contaminated’ conditions are commonly the case, and also that the internal circulation in small gas particles is practically negligible, so they can be treated as spherical rigid particles (Durst et al., 1986) and use the Stokes law for drag modeling.

Lain et al. (1999) proposed a drag expression that combines the correlation developed for solid particles by Schiller and Naumann (1935) for the calculation at low Reynolds numbers with typical drag values of distorted bubbles at high Reynolds number. The correlation seems to match well the data reported for non-pure air-water systems.

\[
C_D = \frac{24}{\text{Re}} \left(1 + 0.15 \text{Re}^{0.687}\right) \quad \text{Re} \leq 500
\]

\[
C_D = 9.5 \times 10^{-5} \times \text{Re}^{1.397} \quad 500 < \text{Re} < 1500 \quad \text{(Lain et al., 1999)}
\]

\[
C_D = 2.61 \quad 1500 \leq \text{Re}
\]

Tomiyama (1998) proposed the following general correlations for pure (two or more times distilled water) and contaminated (tap water) bubbly flows:

\[
C_D = \text{Max} \left[ \text{Min} \left( \frac{16}{\text{Re}} \left(1 + 0.15 \text{Re}^{0.687}\right), \frac{48}{\text{Re}} \frac{8 \text{Eo}}{3 \text{Eo} + 4} \right) \right] \quad \text{Pure liquid system}
\]

\[
C_D = \text{Max} \left[ \frac{24}{\text{Re}} \left(1 + 0.15 \text{Re}^{0.687}\right), \frac{8 \text{Eo}}{3 \text{Eo} + 4} \right] \quad \text{Contaminated system}
\]
Effect of the Void Fraction on Drag

Different designations are used for gas-liquid flows where the gas phase occupies a significant part of the volume, such as: fields of bubbles, families of bubbles, swarms of bubbles, clouds of bubbles, multiple-bubble systems, and in a more general sense high-void fraction flows. In general, it has been observed within the bubbly regime that the momentum interaction between the gas particles and the continuous media is boosted as the concentration or volumetric fraction of the disperse gas phase increases and the proximity between the bubbles reduces.

In the case of multiple solid particles it has been hypothesized that if the volumetric fraction is high enough ($\alpha \geq 2\%$) to make the thickness of the boundary layer exceed the inter-particle distance, a significant increase in the level of interaction will be experienced, and the traditional correlations developed for single particles are no longer applicable. Unfortunately, most experimental and theoretical studies have been performed for single particles and the results can be applied only in the analysis of dilute systems. The scarce data available on the drag in flows with high void fraction have been obtained mainly from fluidization experiments, which are easily conducted for solid particles, but extremely complicated with bubbles due to the coalescence and break-up (Bertola et al., 2004).

For example, settling is a process where the fundamental mechanism is comparable to the rise of bubbles, and it has been observed that the settling velocity of solid particles decreases as the volumetric fraction goes up, and after $\alpha \sim 5\%$ the correlations obtained for single particles can no longer give acceptable predictions (Soo, 1967).

For bubbly flows Behzadi et al. (2004) stated that the use of single-particle based drag correlations should be limited to dilute systems, with a maximum void fraction around
\( \alpha \sim 2-3\% \). From geometrical considerations Michaelides (2003) proposed that bubbly flows can be considered a dilute mixture when the surfaces of the particles are separated by more than one diameter, a limit equivalent to a void fraction around \( \alpha \sim 6\% \). Above this limit, the concentration would have to be accounted for in the calculation of the hydrodynamic forces and heat transfer.

From theoretical considerations the drag force exerted on / by a swarm of bubbles can be calculated as the force acting on a single bubble multiplied by the number of bubbles per unit volume, giving a total drag force that is proportional to the void fraction (Kolev, 2002):

\[
F_D = -\frac{1}{2} C_D \cdot A_{xs} \cdot \rho_l \cdot |\Delta u|^2 \cdot \frac{\alpha}{\pi \cdot D_g^3 / 6} = -\frac{3}{4} \frac{\alpha \cdot \rho_l}{D_g} C_D^* |\Delta u|^2
\]

Attempts to identify expressions for the drag coefficient under high void fraction conditions have been reported usually based on already existing drag correlations for single bubbles. Three approaches are briefly presented:

- Effective continuum viscosity: The viscosity of the continuum used in the computation of the drag is adjusted according to the volumetric concentration of the disperse phase. Using this approach Isshi and Zuber (1979) proposed drag correlations covering a wide range of conditions:

\[
C_D = \frac{24}{\text{Re}_M} \left(1 + 0.1 \text{Re}_M^{0.75}\right) \quad \text{Undistorted [particles, drops, small bubbles]}
\]

\[
C_D = \frac{2}{3} \sqrt{Eo} \left(1 + 17.6 \frac{f(\alpha)}{18.67 \cdot f(\alpha)}\right)^2 \quad \text{Distorted regime [Droplets and bubbles]}
\]

Where: \( \text{Re}_M \): Mixture Reynolds, computed with mixture viscosity

\[
\mu_M = \mu_l \cdot \frac{\left(1 - \alpha\right)^{2.5(\mu_l + 0.4 \cdot \mu_l)}}{\mu_l + \mu_l} \quad \text{Mixture viscosity}
\]
\[ f(\alpha) = \sqrt{1 - \alpha} \cdot \frac{\mu_L}{\mu_G} \]  
Function of the void fraction

- Drag coefficient multiplier: Rusche and Issa (2000) proposed a power-exponential correction for the drag coefficient to take into account the increase in the interaction between the phases as the void fraction becomes larger:

\[ f(\alpha) = \left[ e^{3.64\alpha} + \alpha^{0.864} \right] \]

\[ C_D^H = C_D^{C} \cdot \left[ e^{3.64\alpha} + \alpha^{0.864} \right] \]

The correction function takes the unit value as the void fraction tends to zero, reducing to the drag coefficient for single particles. The authors acknowledge a large scatter in the data for bubbles, while a significantly better fit was obtained for the drag of multiple solid particles.

- Numerical simulation: The flow interactions of fluid and solid particles are difficult to study experimentally, but suitable for computational analysis (Michaelides, 2003). Spelt and Sangani (1998) investigated numerically the effect of the concentration of bubbles on the drag, obtaining the following expression for the viscous drag coefficient of a swarm of bubbles in a viscous fluid, which suggests that the drag increases with void fraction and the velocity fluctuations.

\[ C_D = \frac{1 + 0.15 \cdot \alpha \cdot A}{(1 - \alpha)^2} \]  
Where A is a dimensionless measure of fluid temporary velocity fluctuations

Detailed numerical models of bubbly flows demand extremely powerful computational resources, and their application are limited primarily to low void fractions, so the results unfortunately cannot be used in the analysis of many practical flows (Bertola et al., 2004).
Experimental studies have shown consistently that in bubbly flows the drag coefficient increases with the void fraction. In other words, the drag force on a single particle is lower than the drag that would be experienced by a particle with the same size, relative velocity, and Reynolds number, but under higher concentration or void fraction conditions (Kolev, 2002). On the other hand, as the gas content increases and transition from the bubbly to other regimes associated with high void fractions occur, such as the slug and churn flow regimes, the opposite effect has been observed and the level of drag or momentum interaction tends to decrease as the void fraction increases (Ishii and Chawla, 1979; Ishii and Zuber, 1979). The complex regimes encountered at high void fractions have not been studied as much as the simpler bubbly flow, and as a consequence only a few drag correlations of limited applicability are available.

- **Effect of Non-Newtonian Media**

In spite of the multiple applications involving bubbles in non-Newtonian fluids, and the fundamental relevance of the problem, where important macroscopic differences have been observed with respect to the Newtonian cases, little theoretical and experimental research work has been carried out for bubbly non-Newtonian systems, while most of the attention has been given to Newtonian cases (Kee et al., 1990). The visco-elasticity of fluids is known to govern partially the shape of bubbles, wake characteristics, and coalescence, and therefore important effects on the regime and interfacial forces are naturally expected.

Several attempts to identify the effect of the non-Newtonian character of liquids on the dynamics of bubbles have been reported. The most remarkable observation has been a discontinuity in the bubble rise velocity when a critical size is reached, although some studies have also questioned the existence of such a discontinuity (Kee et al., 1990). The
discontinuity seems to occur when the bubbles reach a volume $V_B \approx 10^{-7}$ m$^3$ (equivalent bubble size $D_G \approx 0.57$ mm), where it has been hypothesized that a change in the regime from rigid to mobile interface could be taking place.

Abou (1986) proposed a generalized correlation for the rise velocity of bubbles, which is reported to be in good agreement with data reported for Newtonian and pseudo-plastic non-Newtonian fluids, and covers the range of spherical, elliptical, and spherical-cap bubbles. Two new dimensionless groups are used, the Flow ($F$) and the Velocity ($V$) numbers, which can be associated with the Reynolds number and the drag coefficient respectively:

$$F = \frac{g \cdot D_G^{2/3} \Delta \rho \cdot \rho_l^{2/3}}{\mu_l^{4/3} \sigma_l^{1/3}} \quad \text{[Valid for } 0.1 < \text{Re} < 10000\text{]}$$

$$V = \frac{\Delta u \cdot D_G^{2/3}}{\mu_l^{1/3} \sigma_l^{1/3}} \quad \text{[Valid for } 0.1 < \text{Re} < 10000\text{]}$$

$$V = 0.75 \cdot (\log F)^2$$

Control of the rheological properties of the primary phase is obviously an important factor in the study of non-Newtonian flows, usually obtained by the controlled addition of specific chemicals. For example, Kee et al. (1990) reported the addition of polyacrylamine (0.5-1.5 % in weight; Separan AP-30) in different concentrations to a mixture 50:50 of glycerin-distilled water to obtain shear-thinning solutions with high viscosities. To prevent biological degradation phenyl-mercuric acetate was added in a small amount (0.01%).

2.2.4 Momentum Interaction in Gas-liquid Vertical Pipe Flow

In gas-liquid vertical pipe upflow the lift force causes a displacement of the small bubbles towards the wall, while large bubbles tend to move towards the center of the tube. This phenomenon has been associated with the interaction between the wakes and the liquid
velocity profile (Tomiyama et al., 2002). In case of high void fractions, usually above \( \alpha \geq 25\% \), the coalescence rate increases quickly, and the flow moves to the slug and churn regimes. In the slug regime large bubbles known as ‘Taylor bubbles’ are developed as a consequence of the merging. Since the internal diameter of the channel restrains the horizontal dimension, the Taylor bubbles adopt a cylindrical geometry with almost the same size of the channel (\( D_{TB} \sim 0.89 \text{ ID} \)) and a bullet shape (Fig 2.11).

![Fig 2.11](image.png)  
Fig 2.11 [a-left] Schematic representing distinctive features of the slug flow regime, and [b-right] image illustrating Taylor bubbles rising in a high-viscosity medium.

As a Taylor bubble rises surrounding liquid is squeezed between the bubble and the walls of the channel and a down-flow film is produced, as well as a wake behind the cylindrical bubble. Liquid slugs are formed between the Taylor bubbles, which may or may not contain smaller bubbles.

Numerous empirical expressions have been proposed to describe the rise velocity of Taylor bubbles and the thickness of the liquid film. While for small bubbles the rise velocity
is normally a function of the equivalent volumetric diameter of the gas particle, for cylindrical bubbles the rise velocity depends on the internal diameter of the tube (Abou, 1986). A correlation for the drag within the slug regime has been proposed as a function of the void fraction, indicating a steep decrease in the momentum interaction as the gas content increases:

\[ C_D = 9.8 \times (1 - \alpha)^3 \]

Slugs, Ishii and Chawla (1979)

\[ D_G = D_{TB} \sim 0.9 \times D_h \]

The previous drag expression for cylindrical Taylor bubbles suggest a decrease in the drag as the void fraction increases, contradicting explicitly the effect of the gas content in bubbly flows. However, fundamental differences exist between both cases as different regimes are encountered: Cylindrical bubbles are developed in internal flows, where the geometry of the bubble and the movement are restricted by the boundaries of the channel. Major wall effects are experienced as the cross section of the bubble facing the flow, associated with the drag, remains practically constant and comparable in size with the channel, while the bubble volume, associated with the buoyancy, increases with the bubble length. Conversely, other types of bubbles (e.g. spherical, elliptical, caps) are smaller in diameter than the channel and almost ‘unbounded’ in the horizontal direction, conditions that permit changes in the bubble shape and cross-sectional area, as well as the typical fluctuations in the path rise.

As the gas content is further increased, coalescence occurs between consecutive Taylor bubbles, breaking the slugs and generating the oscillations typical of the churn regime (Taha and Cui, 2006). The churn flow is highly turbulent, and very difficult to study due to the disordered and chaotic nature of the churns. As a consequence, less information is
available about this regime, and no mechanistic model exists for describing the hydrodynamic properties (Kaya et al., 2001).

- **Air-lift Effect**

  Two important industrial applications involving gas-liquid vertical channel flows are the air-lift reactors and the bubble columns. The air-lift reactor is of particular interest for the present investigation as the configuration and the circulation mechanism are comparable to some extent with the ones found in sugar evaporative crystallizers. Air-lift reactors consist of a riser, a disengagement region, and a downcomer channel: The gas phase is injected in the ‘riser’, where the buoyancy forces due to phase density differences produce an up-flow (as in calandria tubes of sugar evapo-crystallizers); a disengagement volume is available above the riser, where the gas phase is removed (as above the top calandria plate); then the liquid phase flows through the downcomer to enter into the riser again (as in the downtake and bottom). Only few studies have been presented on the hydrodynamics of air-lift reactors, probably due to the satisfactory predictions that can be obtained with 1-D mechanistic models, such as the Drift flux model (Mudde, 2005).

  Merchuk and Stein (1981) reported experimental results obtained in a pilot air-lift reactor for several gas superficial velocities (or flux, $J_G = 0.03-0.3$ m/s) and downtake openings (0-100%). An air-water system was studied, measuring the void fraction with the manometric technique and the liquid recirculation with a flow-meter installed in the downcomer (anubar). It was observed that the induced circulation or flux ($J_L$) increases as the gas flux ($J_G$) is increased, but displaying an asymptotic behavior. It was argued that at low gas rates the bubbles rise straight without merging, and as the gas flow increases the coalescence would become significant, generating larger gas velocities and reducing the air-lift effect as
the slug regime is approached. The reduction in the circulation produced at elevated gas rates was associated with higher frictional losses in the riser, where more of the energy introduced by the gas phase would be dissipated in internal recirculation loops. The flow in industrial air-lift reactors is normally turbulent, and evidence of possible liquid recirculation has been observed as the bubbles near to the pipe wall can display a temporary downward motion (Mudde, 2005).

A large part of the information presented on the air-lift effect is related to the oil industry, as this mechanism is frequently involved in the extraction, transport, and production of hydrocarbons. Oil is encountered between 1-5 km below the surface at pressures up to 200 bar (Nieuwenhuys, 2003), which make possible a pressure-driven ‘natural’ extraction using pipe lines around 0.06-0.09 m in diameter. However, as the oil is removed the pressure logically drops, and so the production rate of the wells. In these cases air can be injected into the production pipe to generate the air-lift effect and reduce the hydrostatic head of the column, increasing the production rate of old wells.

Laboratory experiments with water (Geest and Oliemans, 2001) and oil (Mudde, 2005) in vertical pipes have indicated the convenience of using small bubbles for air-lift wells, as they rise slowly and spread evenly in the cross sectional area, delaying the transition to the slug regime, and maximizing the void fraction and the air-lift effect. Conversely, large bubbles tend to move towards the center and catalyze the transition to the slug flow, which has shown to be unfavorable for the efficiency of air-lift applications. Another reason for preferring the bubbly regime in gas-lift wells is avoiding the pressure fluctuations and instability characteristic of the slug and churn regimes (Nieuwenhuys, 2003). The existence
of liquid slugs traveling at high velocity generates vibrations that can cause damage in the pipelines or other connected elements such as pumps and valves.

- **Bubble Columns**

  The process in bubble columns has been studied extensively, probably because of its widespread application in numerous industries. Krishna et al. (1999) presented experimental and numerical results on the multiphase flow in vertical columns with sizes between 0.051-0.630 m in diameter where water, paraffin oil, and tetradecane were used as the liquid phase. Different multiphase situations were studied, covering single bubbles, pairs of bubbles, and the churn-turbulent regime, where large and small bubbles coexist and rise simultaneously. It is reported that single small bubbles (spherical and elliptical, \(D_G \approx 3-6\) mm) rise with a velocity that can be well predicted with established correlations (e.g. Hamarthy, 1960). On the other hand, the computation of the rise velocity is less clear for large bubbles (spherical caps, \(D_G \approx 20-80\) mm), where no reliable procedure for the estimation of the drag coefficient or bubble size is available. Large bubbles rise faster and are more susceptible to breakage, coalescence, severe deformations, and they represent the main complication in the analysis of many real life flows, as the churn-turbulent regime in bubble columns, where they largely determine the hydrodynamic characteristics of the flow.

  The difficulty in describing the dynamic behavior of large bubbles has been evident in the data reported for a long time, as for example in the drag coefficients presented by Haberman and Morton (1953) for air bubbles in different substances, where the dispersion is clearly worst as the bubble Reynolds number, directly proportional to the bubble size, is increased.
Krishna et al. (1999) reported important wall effects when a large single bubble rises within a stagnant vertical pipe (column). It was observed that bubbles equal in size (volume) would rise faster and adopt a flatter shape in larger tubes. This behavior was attributed to the wall-effects, which would increase the downward liquid velocity as the tube diameter is reduced, and therefore would increase the drag force on the bubble. It was concluded that the rise velocity of large bubbles within a tube is practically independent of liquid properties, but drastically reduced as the column diameter is smaller. In a similar way a test with pairs of bubbles indicated significant wake interactions, observing that the trailing bubble tends to align with the lead bubble, and accelerates progressively until it reaches and merges with the lead bubble. Based on the experimental results it is hypothesized that in the churn-turbulent regime each large bubble will be acting as a trailing bubble, and as a result significantly higher gas velocities are possible than for isolated bubbles.

For the numerical analysis of the churn-turbulent regime Krishna et al. (1999) considered that the Eulerian-Eulerian approach is the only applicable CFD technique, while the Eulerian-Lagrangian and the volume of fluid (VOF) approaches are virtually impracticable. A simulation strategy was developed assuming three phases: the liquid, the small-bubble, and the large-bubble phases. The momentum interaction between small-and-large bubbles was neglected, between small bubbles-and-liquid was calculated with the Harmathy (1960) correlation, and between the large bubbles-and-liquid was computed with a correlation developed from experimental tests. Turbulence was calculated only for the liquid phase using the kE model and standard coefficient values, obtaining numerical results that are reported to be in reasonable agreement with the experimental data. The drag correlations applied by Krishna et al. (1999) in the CFD analysis of churn-turbulent flows in bubble
columns, where SF and AF are the scale and the acceleration factors accounting for the effects of the column diameter and the wake interaction respectively, are as follows:

\[
C_D = \frac{4}{3} \frac{\Delta \rho}{\rho_L} \frac{g D_G}{\Delta u^2} \quad \text{Traditional drag ideal sphere}
\]

\[
\Delta u_{\text{SMALL, BUBBLES}} = 1.53 \left( \frac{\sigma g}{\rho_L} \right)^{0.25} \quad \text{Small bubbles, Harmathy (1960)}
\]

\[
\Delta u_{\text{LARGE, BUBBLES}} = 0.71 \sqrt{g D_G} \cdot SF \cdot AF \quad \text{Large bubbles, Krishna et al. (1999)}
\]

\[
SF = \begin{cases} 
1 & D_G/D_h < 0.125 \\
1.13 \exp(-D_G/D_h) & 0.125 < D_G/D_h < 0.60 \\
0.496 (D_h/D_G)^{0.5} & 0.60 < D_G/D_h
\end{cases}
\]

\[
AF = \alpha + \beta \left[ J_G - U_{\text{transition}} \right] \quad \text{Acceleration factor}
\]

\[
D_{G,\text{LARGE}} = \gamma \left[ J_G - U_{\text{transition}} \right]^{\delta} \quad \text{Large bubble diameter}
\]

\[
\alpha = 2.73; \ \beta = 4.505; \ \gamma = 0.069; \ \delta = 0.376 \quad \text{Fitted coefficients}
\]

### 2.2.5 Algebraic Multiphase Models

Due to the difficulty in characterizing in detail multiphase flows, most of the information that is available is in the form of empirical or semi-empirical 1-D algebraic models, which have been obtained through experiments and statistics, and are presented in the form of mathematical expressions that have proved to be useful in the analysis of engineering problems. The gas-liquid algebraic models that have been proposed could be considered as a valuable data bank if the drag coefficient can be derived from these empirical expressions (Kolev, 2002).

For the algebraic analysis of two-phase flows all the variables considered are averaged along the cross-section, defining a 1D problem that begins with the solution of the momentum equations for the continuous (L = liquid) and disperse (G = gas) phases:
\[
\alpha_L \rho_L \frac{\partial u_L}{\partial \tau} = -\alpha_L \nabla P - \alpha_L \rho_L g + F_{\text{drag}} + F_{\text{VM}} \\
\alpha_G \rho_G \frac{\partial u_G}{\partial \tau} = -\alpha_G \nabla P - \alpha_G \rho_G g - F_{\text{drag}} - F_{\text{VM}}
\]

The drag force and virtual mass force terms represent the exchange of momentum between the two phases and are defined per unit volume as:

\[
F_{\text{DRAG}} = -\frac{3}{4} \frac{\rho_L \alpha_G C_D}{D_g} \Delta u \Delta u \\
F_{\text{VM}} = -\alpha_G \rho_L C_{VM} \frac{\partial (\Delta u)}{\partial \tau}
\]

- **Drift Flux Model**

The semi-empirical algebraic model with more physical basis is probably the drift flux model, which relates the slip velocity to the drift velocity of the gas relative to the liquid. This model was developed mostly by Zuber and Findlay (1965), Wallis (1969), and Ishii (1977), but research is still in progress to refine the drift model, particularly in the search of convincing correlations for the empirical distribution coefficient (Co).

The volumetric flux of the mixture (j) represents the velocity of the center of the volume, and can be computed as the sum of the liquid and gas superficial velocities (J_L + J_G). The gas drift velocity is then defined as the difference between the gas phase velocity (u_G) and the velocity of a reference frame moving with the velocity of the center of volume of the fluid:

\[
J_G = \alpha G + J_L
\]

The void fraction (\(\alpha\)) can be related to the superficial and real velocities of each phase:

\[
J_G = \alpha u_G \\
J_L = (1 - \alpha) u_L
\]
Integrating the local velocity along the channel an expression relating the phase velocities to the void fraction is obtained, which involves an empirical constant that has been denoted as the distribution parameter (Co) and known to be related to the velocity profile. For a given flow the distribution coefficient might depend on the pressure, the geometry of the channel, and the flow rate:

\[ u_G = Co \cdot j + u_{\text{DRIFT}} \]

\[ \alpha = \frac{J_G}{u_G} = \frac{J_G}{Co \cdot j + u_{\text{DRIFT}}} \]

A linear equation describes the gas velocity, where the distribution parameter (Co) and drift velocity (u_{\text{DRIFT}}) can be easily evaluated from experiments as the slope and intercept. For co-current up-flow, Zuber and Findlay (1965) showed that data from various authors could be fitted as:

\[ u_G = Co \cdot j + 1.53 \cdot U_{\text{KU}} \]  

Bubbly and churn turbulent

Where the distribution parameter (Co) varies according to the velocity profile of the liquid phase, and has been characterized according to the dominating flow regime.

\[ Co = 1.2 \]  

Bubbly and churn-turbulent flow (Kolev, 2002)

\[ Co = 0.934 \cdot (1 + 1.42 \cdot \alpha_G) \]  

Bubbly flow vertical tube D=0.1m (Clark, 1985)

And the Kutateladze terminal velocity is given by:

\[ U_{\text{KU}} = \left[ \frac{\sigma \cdot g \cdot \Delta \rho}{\rho^2} \right]^{1/4} \]

For slug flows the velocity of a free rising vertical Taylor bubble is considered the scale. For the slug regime the recommended correlation is:

\[ u_G = Co \cdot j + 0.35 \cdot U_{TB} \]  

Slug flow
Where the velocity for single Taylor bubbles according to Davies and Taylor (1950) is:

\[ U_{TB} = \sqrt{\frac{D_h \cdot g \cdot \Delta \rho}{\rho_L}} \]

Table 2.3 Distribution coefficients and drift velocity.

<table>
<thead>
<tr>
<th>Case</th>
<th>Co</th>
<th>uDRIFT</th>
<th>Source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Co-current upflow</td>
<td>Co = \frac{1}{Ao + (1 - Ao) \cdot \alpha Bo} \quad k = 0.833 \quad Ao = k + (1 - k) \cdot P / 22115000 \quad Bo = \frac{1}{1 - Ao} \left(1 + 1.57 \frac{\rho_G}{\rho_L}\right)</td>
<td>\sqrt{2} \left[ \frac{\sigma g \Delta \rho}{\rho_L^2} \right]^{0.25}</td>
<td>Lellouche, 1974</td>
</tr>
<tr>
<td>Bubbly</td>
<td>Round channel: \quad 1.2 - 0.2 * \sqrt{\rho_G / \rho_L} \quad Rectangular channel: \quad 1.35 - 0.35 * \sqrt{\rho_G / \rho_L}</td>
<td>\sqrt{2} \left(1 - \alpha \right)^{1.75} \left[ \frac{\sigma g \Delta \rho}{\rho_L^2} \right]^{0.25}</td>
<td>Ishii, 1977</td>
</tr>
<tr>
<td>Bubbly: Co-current upflow D=0.1m</td>
<td>0.934(1+1.42*\alpha)</td>
<td>1.53 \left[ \frac{\sigma g \Delta \rho}{\rho_L^2} \right]^{0.25}</td>
<td>Clark and Flemmer, 1985</td>
</tr>
<tr>
<td>Churn-turbulent</td>
<td>1.2 \quad (between 1.0–1.3)</td>
<td>\sqrt{2} \left[ \frac{\sigma g \Delta \rho}{\rho_L^2} \right]^{0.25}</td>
<td>Ishii, 1977</td>
</tr>
<tr>
<td>Slug</td>
<td>1.2</td>
<td>0.35 \left[ \frac{g \Delta \rho D}{\rho_L} \right]^{0.5}</td>
<td>Collier, 1981</td>
</tr>
</tbody>
</table>

Zuber and Findlay (1965) demonstrated that for success in correlating the gas-liquid flow data it is necessary to use regime-dependent correlations. Innumerable studies and applications of the drift flux model have been presented since then in the multiphase literature. Table 2.3 presents some values and expressions that have been recommended for
different situations. In general it has been determined that the distribution coefficient may be lower than one \((Co < 1)\) when wall void picking profiles occur, and can reach values up to \(Co = 1.5\) for strong center line picking of the void fraction (Kolev, 2002). It is important then to emphasize that the distribution coefficient \((Co)\) and the drift velocity \((u_{DRIFT})\) are both regime dependent quantities.

Premoli et al. (1971) presented an empirical model that is reported to give good results over a wide range of conditions. The two-phase correlation is given in terms of the velocity ratio \((S)\), defined as the ratio between the gas and liquid phase average velocities.

\[
S = \frac{u_G}{u_L} = 1 + a \left( \frac{Y}{1 + b.Y} - b.Y \right)^{1/2}
\]

Where:

\[
Y = \frac{x}{1 - x} \left( \frac{\rho_L}{\rho_G} \right)
\]

\[
a = 1.578 \text{Re}_{TP}^{-0.19} \left( \frac{\rho_L}{\rho_G} \right)^{0.22}; \quad b = 0.0273 * \text{We}_{CHANNEL} * \text{Re}_{TP}^{0.51} \left( \frac{\rho_L}{\rho_G} \right)^{-0.08}
\]

\[
\text{We}_{CHANNEL} = \frac{G^2 D_h}{\rho_L \sigma_L} ; \quad \text{Re}_{TP} = \frac{G.D_h}{\mu_L}
\]

2.3 HEAT TRANSFER IN SUGAR EVAPORATIVE CRYSTALLIZERS

The boiling phenomenon is far from being fully understood because of its complex nature and the uncertainty regarding the interaction between the multiple factors involved. Boiling heat transfer or evaporation is an intrinsically unsteady process, with the nucleation and generation of vapor bubbles displaying a random behavior influenced by many factors like size, coalescence, bubble age, densities, viscosities, surface tension, volumetric fraction, regime, hydrostatic head, and so on, that make any analytical study particularly
complex. Evaporation is usually considered a steady state phenomenon, but in reality it is chaotic and difficult to characterize in detail.

The phenomena occurring in the calandria tubes of conventional evapo-crystallizers (natural circulation) could be described as a ‘vertical buoyancy-driven boiling pipe flow’, where the buoyancy forces due to the density difference between the generated gas phase (water vapor) and the surrounding liquid phase is practically the only driving force for the circulation of the massecuite. This type of flow is often referred to as a bubbly-driven flow. Particularly for sugar evapo-crystallizers numerous theories including nucleation, coalescence of bubbles, formation of slugs or columns of vapor, temperature differences, flashing, eruptions, etc. have been proposed to describe the process, but the lack of information has not enabled reasonable conclusions to be drawn. Nonetheless, it is clear that the circulation must be as high as practically possible for quality, recovery and capacity reasons, and that convective boiling plays a vital role, with the density difference between the vapor bubbles and the liquid providing the driving force for massecuite movement.

The boiling heat transfer and fluid flow occurring in sugar evapo-crystallizers have not been clearly identified due to the difficulty in performing field measurements and the numerous operational conditions that are possible. It is believed that a mixture of massecuite and vapor bubbles rises through the calandria tubes, as the vapor generated moves upwards to reach and cross the free surface before going to the barometric condenser. The massecuite would then move in radial direction towards the center, and go back to the bottom through the downtake to enter into the calandria tubes again. The balance between the buoyancy forces and the resistance to flow will determine the circulation rate. The hydraulic resistance
or friction depends essentially on the massecuite viscosity and the geometry of the passages within the evapo-crystallizer, which should be smooth and offer minimum obstruction to flow (Webre, 1959).

As normal in convection processes, in a sugar evapo-crystallizer there is a strong interaction between the heat transfer and fluid velocity or circulation. Any mechanism that reduces the evaporation rate will lower the circulation, and conversely, any reduction in the massecuite velocity will be detrimental to heat transfer (Bosworth, 1959).

- **Thermodynamic State**

In many practical applications involving boiling, like sugar evapo-crystallizers, the thermodynamic state lies far below the critical point, and therefore a large density difference between the liquid and gas phases exists. As a consequence, the buoyancy forces associated with the phase density difference between the liquid and the generated bubbles are comparatively high, while the contribution of the buoyancy forces due to temperature differences is virtually negligible.

Boiling processes are normally classified according to the temperature of the liquid phase:

- **Sub-cooled**: Liquid temperature below saturation, bubbles may condense.
- **Saturated**: Liquid temperature slightly exceeds the saturation temperature, bubbles tend to rise.

In a calandria tube the boiling is saturated. It is known that the actual boiling temperature is higher than the water saturation temperature due to the solids content (Brix) and the hydrostatic effect\(^9\), which cause a boiling point elevation (BPE).

---

\(^9\) The saturation state is normally determined from the vacuum measured above the free-surface.
Heat is transferred normally from vapor coming from the first effects of evaporative multi-effect series, or throttled steam / exhaust in the worst cases. The heating vapor is slightly superheated around 85-140 kPa (95-110°C), and condenses on the outer diameter of the calandria tubes. The heat is transferred through the tube wall to the massecuite, which boils inside at around 14 kPa and 65-70°C, and flows upwards driven by the buoyancy forces in a process that can be cataloged as free convective flow boiling within a vertical heated channel.

- **Evaporation Rate**

In industrial practice empirical parameters are frequently used to quantify the heat transfer, such as the evaporation rate in evap-o-crystallizers, which relates the steam consumed or evaporated to the heat exchange area. The evaporation rate tends to be higher as the purity or grade of the massecuite increases, and changes dramatically during each strike, decreasing progressively and finding its minimum value at the last stage of each cycle. Typical values of the evaporation rate in batch and continuous sugar evap-o-crystallizers are presented in Tables 2.4 and 2.5

Table 2.4  Average evaporation rates in batch crystallizers.

[source: Rein et al., 2004]

<table>
<thead>
<tr>
<th>Type of boiling</th>
<th>Minimum evaporation (kg/m².h)</th>
<th>Maximum evaporation (kg/m².h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Grain / Seed</td>
<td>8.2</td>
<td>61</td>
</tr>
<tr>
<td>A</td>
<td>22</td>
<td>38</td>
</tr>
<tr>
<td>B</td>
<td>6.2</td>
<td>25</td>
</tr>
<tr>
<td>C</td>
<td>3.5</td>
<td>18</td>
</tr>
</tbody>
</table>
Table 2.5  Evaporation rate and heat transfer in continuous crystallizers.

[source: Rein and Msimanga, 1999]

<table>
<thead>
<tr>
<th>Type of boiling</th>
<th>Evaporation (kg/m².h)</th>
<th>Heat transfer coefficient h (W/m².K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>16-25</td>
<td>413</td>
</tr>
<tr>
<td>B</td>
<td>8-13</td>
<td>212</td>
</tr>
<tr>
<td>C</td>
<td>3-10</td>
<td>107</td>
</tr>
</tbody>
</table>

- **Heat Transfer Coefficient – Nusselt Correlations**

Austmeyer and Schliephake (1983) studied experimentally the heat transfer in calandria tubes of sugar evaporative-crystallizers. Boiling conditions were replicated using an experimental facility provided with a tube of diameter 0.10 m and length 1.2 m. Tests were run under natural circulation ($J_L \sim 0.29$ m/s)$^{10}$, and forced circulation using an axial-flow pump operated at 600 and 1100 rpm ($J_L \sim 0.43$ and 0.74 m/s respectively). Experimental information was reported on the temperature field, velocity field, supersaturation field, and heat transfer.

The temperature field of the flow was determined using an array of thermocouples aligned in the axial direction and movable in the radial direction, giving information that permitted computation of the local heat transfer. The results reported for natural circulation indicate a rapid increase in the temperature within the first 0.3-0.4 m of the tube. In this region the liquid phase is likely to dominate the flow, and the temperature is understandably higher in the radial direction, reaching a maximum around $T \sim 73 \, ^\circ\text{C}$ near to the wall. This first region of increasing temperature is described by Austmeyer and Schliephake as the ‘pre-

---

$^{10}$ This velocity seems high with respect to real operating conditions.
heating’ zone. After that, a continuous decrease in the flow temperature is observed until it reaches uniformly the saturation temperature at \( L \sim 0.9 \) m.

The reduction in the boiling point as the hydrostatic pressure is reduced would explain the temperature drop measured by Austmeyer and Schliephake (1983) after \( L \sim 0.3-0.7 \) m, where the gas phase fraction might become significant, and consequently a lower temperature would be observed. In the same way, pressure measurements along the tube indicated the presence of gas bubbles, particularly after \( L \sim 0.8 \) m, where a drastic decrease in the slope of the recorded pressure is evident. The results suggested that most of the vaporization takes place at the top of the heated tube.

Austmeyer and Schliephake (1983) proposed a heat transfer correlation based on the experimental measurements in the middle region of the tube, and compared their results with an empirical equation proposed earlier for boiling pipe flow by Forster and Zuber (1954), observing systematically a lower value in the Nusselt number, but practically the same slope and exponent values for the dimensionless groups Reynolds and Prandtl that are involved in the correlation.

Austmeyer and Schliephake (1983) concluded that the liquid inlet velocity, or circulation, did not show any noticeable effect on the heat transfer, as reported by previous works. However, this conclusion openly opposes what has been observed in practice in sugar evaporative crystallizers, and therefore it is likely that the experimental conditions did not represent really what occurs inside a calandria tube. From the experimental data the following heat transfer correlation was proposed:

\[
Nu = 0.0011 \text{Re}^{0.62} \text{Pr}^{0.25}, \\
\text{Re} = \frac{\rho_L \cdot D_h \cdot u_L}{\mu_L},
\]

80
Rouillard (1985) studied experimentally the heat transfer in calandria tubes. The void fraction was measured using Gamma ray adsorption, obtaining the void fraction profile along the tube. The measured void fraction constitutes probably the only experimental information available on the characteristics of the two-phase flow within calandria tubes.

Based on experimental data, Rouillard (1985) concluded that the length of the sub-cooled region is a function of the heat flux and liquid velocity, and can extend up to the tube exit for C-massecuite under certain conditions. The evaporation rates were observed to increase as the heating steam pressure, the vacuum, and the purity are raised. Conversely, longer tubes, higher strike levels and high concentrations reduce the evaporation rate. A regression analysis was performed and the following expression for calculation of the evaporation rate was proposed, which is reported to have a correlation coefficient 0.81 for 254 sets of data:

$$\ln(EC) = 15.92 - 0.165 \cdot Bx - 0.0601 \cdot P_{ABS} + 0.0311 \cdot Pty + 0.00639 \cdot P_{STEAM} - 0.321 \cdot L - 0.298 \cdot H$$

Where:
- $EC$: Evaporation rate ($\text{kg/m}^2\cdot\text{h}$)
- $P_{STEAM}$: Steam pressure (kPa abs)
- $Bx$: Concentration (%Brix)
- $L$: Tube length (m)
- $P_{ABS}$: Absolute pressure (kPa abs)
- $Pty$: Massecuite purity (%)
- $H$: Massecuite head above calandria (m)

The non-dimensional analysis of the data offered a correlation for the internal boiling heat transfer coefficient in a calandria tube in terms of the Nusselt number and the two-phase Reynolds number, and is reported to have an average error of 18% (max 52%):

$$Nu = \frac{h \cdot D}{k} = 3.72 \cdot Re_{{TP}}^{0.387} \left(\frac{\rho_L}{\rho_G}\right)^{0.22}$$

Rouillard (1985)
\[ \text{Re}_{TP} = \frac{\rho_L D Q_L}{\mu_L A_{XS} (1 - \alpha)} \]

Two-phase Reynolds number

Some similarities between the results reported by Austmeyer and Schliephake (1983) and Rouillard (1985) are evident, such as the reduction in the slope of the pressure profile around \( L \sim 0.75 \) m, where the large presence of bubbles is believed to reduce the hydrostatic effect. In the same way, the temperature profiles reported in both cases indicate initially a progressive increase in temperature up to \( L \sim 0.8 \) m, which is followed by a drop in temperature to a value near to the saturation conditions. It is worth noting that in industrial practice the highest heat transfer coefficients have been observed for evapo-crystallizers provided with ‘short’ tubes, whose length is around \( L \sim 0.7-0.8 \) m, just like the critical point reported by the two experimental works discussed above.

Stephens (2002) studied numerically the process in evapo-crystallizers, and after reviewing Rouillard’s data noticed a difference attributed to an alleged \([D/L]^{1/3}\) term that could have been introduced arbitrarily by Rouillard, proposing the following correction:

\[ Nu = \frac{h D}{k} = 4.48 \text{Re}_{TP}^{0.338} \left( \frac{\rho_L}{\rho_G} \right)^{0.202} \]

Modification by Stephens (2002)

2.4 PREVIOUS EXPERIMENTAL STUDIES ON MASSECUITE CIRCULATION

- Conductivity Method - Java Method

Smith (1938) used the conductivity method to study the circulation velocity in coil pans in Mackay, Australia, observing that the circulation speed decreases with increase in concentration (Brix) and crystal content, and therefore viscosity. The conductivity method permits an approximate measurement of the massecuite upward velocity by installing two pairs of electrodes, one above the other separated by 0.9-1.2 m (3-4 ft). The pass of bubbles or irregularities in the massecuite, such as an added pulse of water, would result in changes in
the electrolytic resistance, which is monitored in the two positions, and then from the measured time between the occurrences of two current fluctuations in the two circuits the liquid rising velocity can be computed.

Austmeyer and Schliephake (1983) measured the axial velocity along an experimental calandria tube using an analogous conductivity method. Essentially annular electrodes, referred to as ‘electrolyte detectors’, and an electrolyte tracer solution, were used to determine the velocity from the time required for the tracer to travel between the electrodes. The results indicated a significant increment in the velocity along the tube, particularly near to the walls, which was attributed to the generation and rising of bubbles.

- **Temperature Difference Method**

  This method was used in the past to estimate the circulation in vacuum pans. It was assumed that evaporation takes place near to the free surface, so in the calandria tubes the heat would be transferred essentially to a single-phase flow of massecuite moving upwards. Then, from measurements of the temperature difference below and above the calandria and the latent heat transferred from the steam, it is possible to compute the circulation rate.

  As the temperature difference along the calandria tubes decreases the computed massecuite flow is higher, and it was established that an evapo-crystallizer with good circulation would exhibit a temperature difference equal or below 3 °C measured between the tube entrance and around 0.4 m above the top calandria plate. Massecuites velocities around 0.08 - 0.33 m/s were reported in Java by Alewijn and Honig (1959). Webre (1959) employed the same methodology to estimate the circulation in Cuba, reporting temperature differences typically between 1 - 2 °C, and maximum 7 °C, and estimated velocities starting from 0.46 m/s to 0.003 m/s during a 6-hours C-strike boiling.
Later research showed that significant evaporation takes place within the calandria tubes, and therefore this method is incorrect in principle. However, it constituted an early and relatively easy base for comparison, and the reported increase in the bulk temperature of the massecuite along the tubes in a good crystallizer ($\Delta T \sim 1 - 3 \, ^\circ C$) might remain valid. As presumed in the past, the temperature difference will be larger in poor circulating crystallizers, as lower circulation and evaporation will result in a higher residence time of the massecuite within the tube, and consequently a larger superheating.

- **Radioisotope Capsules**

  Wright (1966) presented a method for measurement of circulation in evaporocrystallizers by inserting a radioisotope tracer contained within a capsule with the same density as the massecuite, and then by following the tracer movements using external gamma rate-meters. Results are reported on the circulation in seven high-grade pans and four low-grade pans, providing information about the velocity, stagnation points, and the time spent on ‘true circulation’, defined as the period required for the capsule to complete one complete circuit within the pan. The results indicated that the floating calandria and the directional louvers have a detrimental effect on circulation, and demonstrated an increase in circulation when steam jiggers are used.

  The measurements performed by Wright (1966) showed an intermittent flow in the calandria tubes, with a severity in the fluctuations that increases with the massecuite viscosity. Downward currents over the top calandria were recorded and associated with the occurrence of eddies above the top calandria plate. Based on the experimental information an ‘eruptive’ boiling mechanism was proposed, where the massecuite remains stagnant within the tubes until enough superheat is reached, and then a large burst of vapor erupts and
expels the fluid upwards. The vapor escapes, while massecuite penetrates the tube from above and below. The measured massecuite velocities were around 0.17 m/s, with a maximum of 0.25 m/s.

The radioisotope capsule, although interesting and potentially applicable in the study of the circulation in crystallizers and the fluid flow in other processes, has not been reported to be used again, probably due to health concerns and the unpopular use of nuclear technology in the food industry.

- **Hot Anemometry**

  Bosworth et al. (1950, 1953 and 1959) reported data on the circulation velocity in one calandria and several coil evapo-crystallizers measured with hot-wire anemometers in Australia. This technique correlates the cross flow velocity with the heat loss to the surrounding media from a standard hot surface (the hot-wire). Bosworth determined the heat loss at different locations of the crystallizers, obtaining an indirect measurement of the local velocities. Initially a sliding single hot-anemometer was used, and then an array of 36 anemometers mounted on a 2 inch steel tubing was constructed to measure simultaneously in a traverse across the body of the crystallizer.

  All the measurements performed by Bosworth were above the heat exchange area, where vapor bubbles are likely to be present and understandably continuous fluctuations in the hot-anemometers signal were reported, which made it necessary to smooth the signal before the readings could be obtained. While the results for coil crystallizers indicated significant variations and peaks across the crystallizer, in the case of calandria crystallizers no significant changes in the velocity were reported, and no clear differentiation between the up-flow and downtake region was appreciable. This would suggest that the distribution of
the massecuite was uniform across the calandria, and similarly the heat transfer coefficient approximately the same in all the tubes. A significant decrease in the circulation as the strike volume increases was evidenced, resulting in the reduction of the measured velocity from $U \sim 0.16 \text{ m/s}$ to $U \sim 0.08 \text{ m/s}$.

Rackemann and Stephens (2002) used commercial insertion flowmeters to determine the massecuite velocity in a B-strike batch evapo-crystallizer in Australia. The sensor employed was a hot-wire anemometer, similar in principle to the sensors used by Bosworth (1950) but more developed and including temperature compensation. The results indicated velocities as high as 1.2 m/s at the beginning of the strike, and a progressive decrease of the circulation as the boiling cycle proceeds.

The thermo-physical properties of the massecuite change continuously during a strike, affecting the convective heat transfer occurring at the probes, and therefore the relationship between the cross-flow velocity and heat loss. A rig for the calibration of the sensors while inserted in a pipe flow was used to determine the effect of different massecuite concentrations and purities on the signal and develop a correlation. It is reported that the orientation of the sensors was not adequately considered, and they were rotated until a maximum in the signal was observed11.

Broadfoot et al. (2004) reported data on the circulation in an A-strike SRI continuous evaporative crystallizer in Australia, where the same procedure described by Rackemann and Stephens (2002) was used to measure the velocity profile in the bottom below the calandria and in the downtake, observing velocities up to 0.4 m/s and 0.2 m/s respectively.

It can be seen that a number of attempts to perform field measurements of the circulation velocity in evaporative crystallizers have been intended since the early 1930s by

11 This is a potential cause of error discussed in section 5.1.2
application of different experimental techniques. Table 2.6 summarizes reported data on the
circulation in sugar evaporative-crystallizers. In general, the information that is available
presents the circulation velocity as the vertical upward velocity in the region where the
calandria tubes or coils are present, and it has been determined to be between 0.02 - 0.25m/s
according to the grade and boiling stage.

Table 2.6  Circulation velocity in sugar evaporative-crystallizers (m/s).

<table>
<thead>
<tr>
<th>Source</th>
<th>Method</th>
<th>Pan type</th>
<th>Heat exchanger</th>
<th>A</th>
<th>B</th>
<th>C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Smith, 1938</td>
<td>Conductivity</td>
<td>Batch</td>
<td>Coil</td>
<td>0.025 -</td>
<td>0.019 -</td>
<td>0.005 -</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0.041</td>
<td>0.040</td>
<td>0.013</td>
</tr>
<tr>
<td>Behne, 1938</td>
<td>Conductivity</td>
<td>Batch</td>
<td>Calandria</td>
<td>0.08</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>max 0.16</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Bosworth, 1959</td>
<td>Hot</td>
<td>Batch</td>
<td>Coil</td>
<td>0.10</td>
<td>0.04</td>
<td>0.02</td>
</tr>
<tr>
<td></td>
<td>anemometers</td>
<td></td>
<td></td>
<td>(refined 0.20)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Wright, 1966</td>
<td>Radioisotope</td>
<td>Batch</td>
<td>Calandria</td>
<td>0.17</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td></td>
<td>capsule</td>
<td></td>
<td></td>
<td>max 0.25</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rackemann, 2002</td>
<td>Hot</td>
<td>Batch</td>
<td>Calandria</td>
<td>---</td>
<td>max 1.2</td>
<td>---</td>
</tr>
<tr>
<td></td>
<td>anemometer</td>
<td></td>
<td></td>
<td></td>
<td>bottom</td>
<td></td>
</tr>
<tr>
<td>Broadfoot, 2004</td>
<td>Hot</td>
<td>Continuous</td>
<td>Calandria</td>
<td>max dwtk 0.2</td>
<td>---</td>
<td>---</td>
</tr>
<tr>
<td></td>
<td>anemometer</td>
<td></td>
<td></td>
<td>max bottom 0.4</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

2.5  PREVIOUS NUMERICAL STUDIES ON MASSECUITE CIRCULATION

Three computational studies on the fluid flow in sugar evaporative crystallizers have
been reported, all involving substantial simplifications and assumptions to deal with the
complexity of the geometry and the different mass, momentum, and energy interactions that
occur between the three participating phases as condensation, evaporation, and solidification
take place simultaneously within the same vessel.
Bunton (1981) in his PhD dissertation reports the first numerical results on the flow in batch evapo-crystallizers. Bunton developed a code for the particular solution of the flow, but in the process he needed to make some strong assumptions such as: homogeneous velocity of the phases (no slip), laminar flow, Newtonian behavior, simplifying the geometry by representing the calandria tubes as a single annular up-flow region and omitting the downtake and bottom regions, where lumped 1D models were used to compute a pressure drop. Convergence problems were reported for the cases corresponding to a high-evaporation rate.

As a part of the numerical analysis Bunton (1981) included the prediction of the crystal growth rate within the computational domain, this based in a mathematical model for sugar growth in industrial crystallization presented by Wright and White (1974). The results indicated that the highest crystallization rates would occur near to the free surface, and that the growth velocity decreases with depth as the hydrostatic pressure increases. At the top end of the tubes zero growth rate is predicted, and surprisingly below this point the opposite behavior is reported, displaying an increase in the growth rate with depth.

Brown et al. (1992) presented results obtained with a commercial CFD code (FLUENT) for a stirred crystallizer. A single-phase model was used, neglecting the effect of the secondary phase and buoyancy forces. The calandria tubes were represented as concentric rings and the massecuite was assumed to be Newtonian. The results indicated massecuite velocities that are suspiciously high (up to 3.8 m/s), and the occurrence of down-flow in the inner tubes.

Stephens (2002) in his PhD dissertation presents a more complete numerical solution of the flow, where the heat transfer is considered to some extent through coupling a 1-D
numerical solution of the boiling in the calandria tubes (FORTRAN routine) with a commercial CFD code (CFX) for the solution of the 2-D flow in the rest of the domain, where isothermal conditions are assumed. The Eulerian-Eulerian approach was applied, and the free surface was treated as a degassing boundary. Stephens results indicated a relatively smooth flow field, except for the development of a vortex above the calandria that grows substantially as the massecuite head increases, and covers the entire section above the top calandria plate for the highest-head case. The maximum liquid velocity is predicted in the free surface region in radial direction towards the center.

The results presented by Stephens constitute probably the most detailed description that is available on the circulation in a sugar evapo-crystallizer. Some observations that can be made are:

- The massecuite is assumed Newtonian, and its properties independent of temperature. Numerous studies have suggested that massecuites behave mostly as pseudo-plastic fluids.

- Ideal spherical bubbles were assumed, and large bubble diameters became necessary to let the gas phase reach the free surface and be removed. While the internal diameter of the calandria tubes is 0.10 m, the assumed diameter of the vapor bubbles was varied arbitrarily between 0.10-0.15 m. The Stokes drag law was used to compute the interfacial momentum interaction in the calandria. However, this law is valid only for undistorted spherical particles at very low Reynolds number, and therefore it is unlikely that the vapor-massecuite interaction was properly represented.
- If the numerically predicted flow field at high massecuite head were correct, a large velocity gradient would be present near to the side walls, and a fast rise of the massecuite would be visible through the side windows of batch evapo-crystallizers. This is contrary to the apparently stagnant flow near to the windows, particularly at the end of the strike, when the viscosity reaches the maximum and the evaporation a minimum.

- The reported pressure fields seem suspicious, as they do not show the hydrostatic effect that can be expected in this highly dense slow moving flow. The right solution of the momentum equations requires a correct calculation of the pressure field.

2.6 FINAL REMARKS

From the information reviewed in this chapter it can be presumed that primarily the buoyancy forces due to phase density differences produced during the boiling in the calandria tubes drives the circulation in sugar evaporative crystallizers. The circulation would be determined by the balance between buoyancy and frictional resistance, and be closely interconnected with the heat transfer in the calandria, as normal in convection processes. Measurements performed in full-scale sugar evaporative crystallizers have indicated an intermittent process within the calandria tubes, which has been attributed to a hypothetic ‘eruptive’ mechanism of boiling. Most of the vaporization seems to take place in the upper section of the tubes in a highly unsteady manner.

Several experiences on the application of experimental and numerical techniques to study the flow in sugar evaporative crystallizers have been reported, but the circulation mechanism and massecuite flow patterns are still unclear due to the complexity of the case and the intrinsic difficulty for performing experimental measurements.
For the numerical analysis of the multiphase flow in sugar evaporative-crystallizers the Eulerian-Eulerian approach is probably the only suitable CFD model. However, several authors have emphasized the importance to provide adequate closure relations for the coupling of the Navier-Stokes equations when gas-liquid flows are simulated applying the Eulerian-Eulerian approach. For the particular case studied the momentum exchange coefficient is the critical factor, which normally is calculated based on drag coefficient correlations and would control how much momentum is transferred from the rising gas phase (water vapor) to the liquid phase (massecuite), determining to a large extend the circulation rate and flow patterns. Therefore, identifying adequate drag correlations able to represent accurately the exchange of momentum between the vapor and massecuite in the calandria tubes is a critical step in obtaining a meaningful CFD solution of the fluid flow in sugar evapo-crystallizers.

A large number of correlations for the calculation of the drag coefficient of bubbles has been proposed from experimental and analytical grounds during the last 150 years. The high number of correlations instead of a wide knowledge could be reflecting a lack of understanding about the physics behind the gas-liquid interaction and all the relevant factors, particularly as the bubble size increases, the bubble shape becomes distorted, and wake interactions with adjacent bubbles take place. Dependable drag correlations have been developed for small single bubbles rising in stagnant media, which can be applied with confidence for the CFD analysis of diluted bubbly flows, but are probably not applicable in more complex situations such as high void fraction bubbly flows, swarms of bubbles, slugs, and churns, where less information has been obtained and no reliable procedure for the calculation of the interfacial interactions is available.
The uncertainty about the drag coefficient of gas particles makes it difficult to simulate the flow in evaporative crystallizers with confidence, since all the regimes starting from the single liquid phase (downtake and bottom region) to the single gas phase (above the free surface) are possible, the liquid phase is non-Newtonian, and large quantities of different substances considered contaminants are present, so it is unclear which would be the best momentum closure relation(s) to be used in the numerical simulation of the flow.
CHAPTER 3 - RESEARCH ON THE TWO-PHASE MOMENTUM INTERACTION AND FLOW PATTERNS IN EVAPORATIVE CRYSTALLIZERS USING A LAB-SCALE EXPERIMENTAL FACILITY

The gas-liquid momentum interaction and the fluid flow patterns in sugar evaporative crystallizers have been studied experimentally using Particle Image Velocimetry (PIV) to determine the flow field in a lab-scale test rig designed to represent the circulation mechanism and the major features of fluid flow in full-scale crystallizers, and to provide data to verify the CFD multiphase simulations. The tests on the scaled test rig provided useful information to understand better the circulation in evaporative-crystallizers and pointed out critical aspects that must be addressed in the numerical simulation of the flow applying the Eulerian-Eulerian CFD approach. In this chapter details and results of the experimental and numerical research in the lab-scale facility are presented.

3.1 MATERIALS AND METHODS

3.1.1 Lab-scale Model

The geometry of the test rig was defined based on an existing batch sugar evapo-crystallizer (90 m³, A, W-bottom, flat side walls) that exhibits satisfactory performance. Industrial batch evapo-crystallizers are cylindrical vessels, displaying an axisymmetric geometry in the vertical direction. However, for convenience the test rig was designed with a rectangular cross section, reducing the volume and giving access for measurements and visualization of the flow along the entire cross section through the front, back, and side walls. The scale factor used is 6:1, and geometric similarity was kept as far as practicably possible.

In full-scale evapo-crystallizers the buoyancy due to density differences between the gas (vapor) and liquid phases (massecuite) during the boiling is the main driving force for
massecuite circulation, while the buoyancy resulting from temperature differences would play a practically negligible role. The buoyancy-driven circulation mechanism has been emulated at lab-scale in a simplified and controllable manner through the injection of air into vertical rectangular channels that represent the calandria tubes (Fig 3.1). The injected air represents the vapor that is generated during the boiling, and the buoyancy resulting from density differences between the air bubbles and the surrounding liquid constitutes the driving force for the circulation, as in the full-scale crystallizers.

Fig 3.1 [a-top] Schematics and [b-bottom] image of the lab-scale experimental facility.
The test rig was entirely constructed in Plexiglas (thickness \(\frac{1}{2}\) inch, refractive index 1.49-1.50, density \(\sim 1190\) kg/m\(^3\)), providing optical access in all the directions and permitting the passage of the laser light that is required for PIV measurements. The air spargers were constructed drilling through sixteen holes [D \(\frac{1}{32}\)” @ \(\frac{1}{4}\)”] on acrylic tubes [D \(\frac{1}{2}\)"], which are inserted in the Plexiglas walls [t \(\sim \frac{1}{2}\)”] separating consecutive channels, for a total of 32 orifices injecting air into each channel. A generously sized manifold [D 4”] located at the top is used to distribute the air between the spargers and procure an even supply. Eight steel L-forms [1.5”x1.5”x1/8”; L \(\sim 1.27\) m.] are situated in the front and back walls to restrict any excessive bowing caused by the hydrostatic pressure.

### 3.1.2 Experimental Conditions and Fluid Properties

The experiments were conducted at sea level, under conditioned room temperature at T \(\sim 25\) °C. Initially tap water was used to represent the liquid phase, which conveniently allows benchmarking the results against experiences reported in the literature on the rise of air bubbles in water. Corn syrup was also used in an attempt to represent the highly viscous massecuities. The corn syrup offers a higher solubility than sugar solutions, and its organic origin makes possible to expect similar properties. The corn syrup has a very low color (convenient for the PIV), and the viscosity can be adjusted by dilution. Table 3.1 presents the three fluid conditions that have been evaluated.

<table>
<thead>
<tr>
<th>Liquid</th>
<th>Density (\rho_L) (kg/m(^3))</th>
<th>Dyn. viscosity (\mu_L) (Pa.s)</th>
<th>Surface tension (\sigma) (N/m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tap water</td>
<td>998.2</td>
<td>0.001 Newtonian</td>
<td>0.073</td>
</tr>
<tr>
<td>Corn syrup 1 - 51 Bx</td>
<td>1243</td>
<td>0.029 Newtonian</td>
<td>0.080</td>
</tr>
<tr>
<td>Corn syrup 2 - 66 Bx</td>
<td>1318</td>
<td>0.281 Newtonian</td>
<td>0.080</td>
</tr>
</tbody>
</table>
As the air is injected in the test rig the free surface is lifted and its exact position becomes difficult to quantify due to the bubble disengagement and fluctuations in the vertical direction. In order to have repeatable experimental conditions the liquid level was set with respect to the top end of the channels (correspondent to the head over the top calandria plate) without any air being injected (gas free). The liquid level was determined using a metallic ruler [0-30 cm] attached to one of the rig sides, and several cases with liquid heads between H ~ 0–30 cm were considered. The rise in the level of the free surface was recorded for each run, giving a rough indication of the total volume occupied by the gas in the rig, which was used to estimate the value of the void fraction.

The airflow rate was determined using a volumetric flow meter [Cole Parmer, Air flow 1-10 CFM, Model 2-44-07-05-4] in combination with a pressure gauge meter [Marsh Mastergauge, K-Monel tip and tube, 0-30 Psig], and assuming that the air obeys the ideal gas law. Although airflow rates between 0–10 CFM could be measured, it was observed that below 6 CFM the air was not distributed evenly between the spargers, and may even become absent in some channels, particularly those located at the sides; while above 8 CFM the airflow is susceptible to oscillations in the compressed air line. Therefore, most of the measurements have been performed with airflow in the range 6–8 CFM. Measured airflow rates under different experimental conditions are presented in Table 3.2. It is calculated that within the vertical channels representing the calandria tubes the gas phase had averaged gas superficial velocities $J_G \sim 6.2 - 7.4 - 8.8$ cm/s under the three airflow conditions considered (6, 7, and 8 CFM respectively).

\[ P = \rho \cdot R \cdot T \]
\[ R = 0.287 \text{ kJ/kg.K} \]
\[ T \sim 298 \text{ K} \]

12 Evaporation rate maybe lower in tubes located farthest from downtake as they are exposed to a longest circulation path and highest flow resistance.
Table 3.2  Airflow rates used during the experiments in the lab-scale model.

<table>
<thead>
<tr>
<th>Liquid head (cm)</th>
<th>Air Flow (CFM)</th>
<th>Air pressure (Psig)</th>
<th>Air pressure (kPa)</th>
<th>Air density (kg/m³)</th>
<th>Air flow (kg/s)</th>
<th>Air flow (LPS)</th>
<th>Air Superficial Velocity Jₖ (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.0</td>
<td>6</td>
<td>1.7</td>
<td>113.0</td>
<td>1.20</td>
<td>0.00374</td>
<td>3.13</td>
<td>0.0621</td>
</tr>
<tr>
<td>0.0</td>
<td>7</td>
<td>2.0</td>
<td>115.1</td>
<td>1.20</td>
<td>0.00444</td>
<td>3.71</td>
<td>0.0738</td>
</tr>
<tr>
<td>0.0</td>
<td>8</td>
<td>2.7</td>
<td>119.9</td>
<td>1.20</td>
<td>0.00529</td>
<td>4.42</td>
<td>0.0879</td>
</tr>
<tr>
<td>2.5</td>
<td>6</td>
<td>1.7</td>
<td>113.0</td>
<td>1.20</td>
<td>0.00374</td>
<td>3.12</td>
<td>0.0620</td>
</tr>
<tr>
<td>2.5</td>
<td>7</td>
<td>2.0</td>
<td>115.1</td>
<td>1.20</td>
<td>0.00444</td>
<td>3.70</td>
<td>0.0736</td>
</tr>
<tr>
<td>2.5</td>
<td>8</td>
<td>2.8</td>
<td>120.6</td>
<td>1.20</td>
<td>0.00532</td>
<td>4.44</td>
<td>0.0882</td>
</tr>
<tr>
<td>5.0</td>
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<td>1.8</td>
<td>113.7</td>
<td>1.20</td>
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<td>0.0622</td>
</tr>
<tr>
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<td>2.0</td>
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<td>1.20</td>
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<td>0.0734</td>
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<td>5.0</td>
<td>8</td>
<td>2.9</td>
<td>121.3</td>
<td>1.20</td>
<td>0.00535</td>
<td>4.45</td>
<td>0.0885</td>
</tr>
<tr>
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<td>113.7</td>
<td>1.21</td>
<td>0.00376</td>
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</tr>
<tr>
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<td>7</td>
<td>2.1</td>
<td>115.8</td>
<td>1.21</td>
<td>0.00447</td>
<td>3.71</td>
<td>0.0737</td>
</tr>
<tr>
<td>7.5</td>
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<td>2.9</td>
<td>121.3</td>
<td>1.21</td>
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<td>4.44</td>
<td>0.0882</td>
</tr>
<tr>
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<td>1.9</td>
<td>114.4</td>
<td>1.21</td>
<td>0.00379</td>
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<tr>
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<td>0.0620</td>
</tr>
<tr>
<td>15.0</td>
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<td>0.00450</td>
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<td>0.0736</td>
</tr>
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<td>3.0</td>
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<td>0.0881</td>
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<td>0.00381</td>
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<td>0.0621</td>
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<td>2.3</td>
<td>117.1</td>
<td>1.22</td>
<td>0.00452</td>
<td>3.71</td>
<td>0.0737</td>
</tr>
<tr>
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<td>3.0</td>
<td>122.0</td>
<td>1.22</td>
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<td>0.0877</td>
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<tr>
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<td>2.0</td>
<td>115.1</td>
<td>1.23</td>
<td>0.00381</td>
<td>3.09</td>
<td>0.0615</td>
</tr>
<tr>
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<td>0.00452</td>
<td>3.67</td>
<td>0.0730</td>
</tr>
<tr>
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<td>3.0</td>
<td>122.0</td>
<td>1.23</td>
<td>0.00538</td>
<td>4.37</td>
<td>0.0869</td>
</tr>
</tbody>
</table>

3.1.3 Flow Measurements Applying PIV

To determine the flow field in the lab-scale test rig Particle Image Velocimetry (PIV) has been applied. This velocimetry technique measures the velocity field in a laser-illuminated plane by following the change of position of tracer particles. A pulsed laser
light sheet synchronized with a high-resolution digital camera (Charged Couple Device – CCD) is used to capture images of the tracer particles. The flow is seeded using microspheres that scatter the laser light, allowing recording of their position in images acquired with the CCD during each laser pulse. The tracer particles must have the right density and size, being small enough to accurately follow the flow, and large enough to be detectable by the CCD. Image pairs are acquired using a very small separation time, which is set to measure accurately the particle displacement and hence the velocity. For the analysis the images are subdivided into small interrogation areas, and cross correlation is used to compare each pair of images using a Fast Fourier Transform, finding the average displacement for each area. PIV measurements give an expected accuracy of 1% of the full-scale velocity (Mercer, 2003).

The PIV system uses a 3 mm thick laser sheet generated by two 15 Hz New Wave Gemini Nd:Yag 532 ηm lasers. A 30 Hz Kodak Megaplus ES 1.0 CCD camera, with a resolution 1008x1012, and an AF Nikkor 50mm f/1.8D lens, acquires images of the plane illuminated by the laser sheet. The seeding used for the PIV measurements consists of 10 μm diameter silver coated glass spheres. The PIV is operated in double exposure mode, and for the analysis a cross correlation with a 42x42 pixel interrogation area is used. During 100 seconds, 400 image pairs are acquired, and the instantaneous velocity results generated from each image pair are averaged to determine the mode, and for the analysis a cross correlation with 33x33 to 42x42 pixel interrogation area is used. During 100 seconds, 400 image pairs are acquired, and instantaneous velocity results generated from each image pair are averaged to determine the velocity field at each position. The analysis is performed using the software pro-VISION™.
The different velocities of the gas and liquid phases impose a challenge in the measurements of this multiphase flow. The bubbles scatter the laser light and become recorded in the PIV images, reducing the signal as it becomes difficult to follow the position of tracer particles, and introducing noise as the PIV processor traces the bubbles. To measure exclusively the liquid phase velocity optical separation was intended by using fluorescent tracer particles (10 µm Coast Yellow, and 20 µm Rhodamine; excited laser 532 ηm; and emitting at 600 and 625 ηm respectively) and an optical orange long pass filter (B+W 041, Schott OG570, cuts off light <570 ηm). The spectral information correspondent to the long-pass filter, the digital camera, and the two fluorophores that were used is presented in Figures 3.2 and 3.3.

Fig 3.2  Spectra for digital camera Kodak Megaplus ES 1.0 and B+W 041 long pass filter.

Fig 3.3  Spectra for [a-left] Suncoast Yellow and [b-right] Rhodamine fluorophores.
Due to the relatively large dimensions of the lab-scale model, for the measurements the experimental domain was divided in ten 20x20cm² zones (Fig 3.4). The post processing included therefore merging the results obtained at each zone for every condition tried into a single velocity vectors plot, which is performed using the software Tecplot.

![Diagram of lab-scale model division](image)

**Fig 3.4** [a-left] Schematic and [b-right] image of the lab-scale model showing the division of the domain in ten zones for the PIV measurements.

### 3.1.4 Numerical Simulation of the Flow Applying CFD

For the numerical simulation of the flow in the lab-scale test-rig the commercial Computational Fluid Dynamics (CFD) code FLUENT was used. To simulate realistically this multiphase flow several critical issues must be simultaneously addressed, including a relatively complex geometry, the multiphase character of the system, the interfacial momentum interaction, the presence of a free surface, and the buoyancy-driven nature of the flow. The numerical models used and the major assumptions are:
- Calculations have been done in 2D and 3D for the lab-scale model. The grids have been generated using quadrangular elements for 2D and octagonal elements for 3D cases. A 2D grid independence study using between 17 000 to 420 000 elements showed no change in the results when more than 90 000 elements were used.

- The Eulerian-Eulerian multiphase model is selected; this is the most appropriate considering the complex interaction between the phases, the high void fraction, and the enormous quantity of bubbles that is present.

- The gas phase is introduced using mass source terms in control volumes that are located in the vicinity of the air spargers.

- The computational domain boundary at the top is placed above the free surface to provide space where the air disengages and ‘leaves’ through an outflow boundary condition.

- The $k-\xi$ turbulence model is used, with a realizibility constraint, and a maximum upper bound of the turbulent viscosity set equal to 500 times the laminar viscosity of the fluid.

- The interaction between the gas-and-liquid phases is incorporated by specifying the momentum exchange coefficient, which is computed based on drag correlations developed for bubbles.

### 3.2 TERMINAL VELOCITY AND DRAG COEFFICIENT OF BUBBLES

The Eulerian-Eulerian approach used for the CFD analysis considers that the secondary phase(s) can be treated as a continuum and solves the Navier-Stokes equations simultaneously for each phase, involving the use of an empirical correlation to couple the momentum equations of the different phases and obtain the closure for the system. The
accuracy of the simulations depends therefore to a great extent on having accurate momentum exchange correlations that are able to represent the interactions under a wide range of practical conditions and for fluids with rheological properties encountered in practice. Tables 3.3 and 3.4 present a compilation of different drag correlations that have been applied in the analysis of bubbles rising in pure and non-pure or ‘contaminated’ media.

Table 3.3 Drag coefficient for bubbles in pure systems.

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Correlation</th>
<th>Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hadamard – Ribezynsky (1911)</td>
<td>( C_D = \frac{16}{Re} )</td>
<td>Re&lt;1; Spherical bubbles with mobile interphase, found in very pure systems</td>
</tr>
<tr>
<td>Clift et al. (1978)</td>
<td>( C_D = \frac{14.9}{Re^{0.78}} )</td>
<td>Low Re; Bubbles with clean interphase</td>
</tr>
<tr>
<td>Lain et al. (1999)</td>
<td>( C_D = 16/Re )</td>
<td>Re \leq 1.5</td>
</tr>
<tr>
<td></td>
<td>( C_D = 14.9/Re^{0.78} )</td>
<td>1.5 &lt; Re \leq 80</td>
</tr>
<tr>
<td></td>
<td>( C_D = 48 \left(1 - \frac{2.21}{\sqrt{Re}}\right) )</td>
<td>80 &lt; Re \leq 700</td>
</tr>
<tr>
<td></td>
<td>( C_D = 1.86E - 15\ast Re^{4.756} )</td>
<td>700 &lt; Re \leq 1530</td>
</tr>
<tr>
<td></td>
<td>( C_D = 2.61 )</td>
<td>1530 &lt; Re</td>
</tr>
<tr>
<td>Wang et al. (1994)</td>
<td>( \ln(C_D) = \ln 24 - \ln Re )</td>
<td>Fit Haberman and Morton (1953) data, distilled water</td>
</tr>
<tr>
<td></td>
<td>( \ln(C_D) = 2.7 - 0.336 \ast \ln Re - 7.14E - 02 \ast (\ln Re)^2 )</td>
<td>Re \leq 1.5</td>
</tr>
<tr>
<td></td>
<td>( \ln(C_D) = -51.8 + 13.2 \ast \ln Re - 0.824 \ast (\ln Re)^2 )</td>
<td>1 \leq Re \leq 456</td>
</tr>
<tr>
<td></td>
<td>( C_D = \max\left{ \min\left[ \frac{16}{Re}, (1 + 0.15 Re^{0.687}) \frac{48}{Re}, \frac{8}{3 Eo + 4} \right] \right} )</td>
<td>456 \leq Re \leq 4 000</td>
</tr>
<tr>
<td>Tomiyama (1998)</td>
<td>( C_D = \frac{16}{Re}, \frac{14.9}{Re^{0.78}} )</td>
<td>Water distilled twice or more times</td>
</tr>
<tr>
<td></td>
<td>( \frac{49.9}{2.61} \left(1 - 2.21 \ast Re^{-0.5}\right) + 1.17E - 8 \ast Re^{2.65} )</td>
<td>Empirical correlation for fluid sphere.</td>
</tr>
<tr>
<td>Sommerfeld (2002)</td>
<td>( C_D = \frac{16}{Re}, \frac{14.9}{Re^{0.78}} )</td>
<td>Re \leq 1.5</td>
</tr>
<tr>
<td></td>
<td>( \frac{49.9}{2.61} \left(1 - 2.21 \ast Re^{-0.5}\right) + 1.17E - 8 \ast Re^{2.65} )</td>
<td>1.5 &lt; Re \leq 80</td>
</tr>
<tr>
<td></td>
<td>( \frac{49.9}{2.61} \left(1 - 2.21 \ast Re^{-0.5}\right) + 1.17E - 8 \ast Re^{2.65} )</td>
<td>80 &lt; Re \leq 1530</td>
</tr>
<tr>
<td></td>
<td>( \frac{49.9}{2.61} \left(1 - 2.21 \ast Re^{-0.5}\right) + 1.17E - 8 \ast Re^{2.65} )</td>
<td>Re &gt; 1530</td>
</tr>
</tbody>
</table>
Table 3.4 Drag coefficient for bubbles in non-pure ‘contaminated’ system.

<table>
<thead>
<tr>
<th>Author(s)</th>
<th>Correlation</th>
<th>Application</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stokes (1851)</td>
<td>( C_D = \frac{24}{Re} )</td>
<td>( \text{Re}&lt;1; \text{Spherical bubbles with ‘rigid’ interphase due to surface impurities} )</td>
</tr>
<tr>
<td>Oseen (1910)</td>
<td>( C_D = \frac{24}{Re} \left(1 + \frac{3}{16} \text{Re} \right) )</td>
<td>( \text{Re}&gt;16 )</td>
</tr>
<tr>
<td>Schiller Naumann (1935)</td>
<td>( C_D = \frac{24}{Re} \left(1 + 0.15 \times \text{Re}^{0.687} \right) )</td>
<td>For solid spheres, no internal circulation.</td>
</tr>
<tr>
<td>Harmathy (1960)</td>
<td>( \Delta u = 1.53 \left( \frac{g \Delta \rho \sigma}{\rho_L} \right)^{\frac{1}{4}} ); ( C_D = \frac{4}{3} \frac{\Delta \rho g \Delta L}{\rho_L \Delta u^2} )</td>
<td>High Reynolds numbers ( 5000&lt;\text{Re} )</td>
</tr>
<tr>
<td>Taylor and Acrivos (1964)</td>
<td>( C_D = \frac{16}{\text{Re}} + 2 + \frac{2}{5} \text{Re} \times \ln \frac{\text{Re}}{2} + 1.33 \frac{\text{We}}{\text{Re}} )</td>
<td>( \text{Re}&lt;2 ) and ( 100&lt;\text{Mo}&lt;100000 ); For ellipsoidal bubbles</td>
</tr>
<tr>
<td>Mendelson (1967)</td>
<td>( C_D = \frac{8}{3} - \frac{16}{3} \times \text{We} )</td>
<td>( \text{Mo} \sim 10^{-10} )</td>
</tr>
<tr>
<td>Clift and Gauvin (1971)</td>
<td>( C_D = \frac{24}{\text{Re}} \left(1 + 0.15 \times \text{Re}^{0.687} \right) + \frac{0.42}{1 + 4.25 \times 4 \times \text{Re}^{-1.16}} )</td>
<td>Contaminated system, extension of Schiller Naumann model ( 0&lt;\text{Re}&lt;300 \ 000 )</td>
</tr>
<tr>
<td>Ishii and Chawla (1979)</td>
<td>( C_D = 9.8 \times (1 - \alpha)^3 )</td>
<td>Slugs</td>
</tr>
<tr>
<td></td>
<td>( D_G = D_{TB} = 0.9 \times D_h )</td>
<td></td>
</tr>
<tr>
<td>Ishii and Zuber (1979)</td>
<td>( C_D = \frac{8}{3} \times (1 - \alpha)^2 )</td>
<td>Churn-turbulent</td>
</tr>
<tr>
<td>Lain et al. (1999)</td>
<td>( C_D = \frac{24}{\text{Re}} \left(1 + 0.15 \times \text{Re}^{0.687} \right) )</td>
<td>( \text{Re} \leq 700 )</td>
</tr>
<tr>
<td></td>
<td>( C_D = 9.5 \times 10^{0.397} \times \text{Re}^{1.397} )</td>
<td>( 700 &lt; \text{Re} \leq 1500 )</td>
</tr>
<tr>
<td></td>
<td>( C_D = 2.61 )</td>
<td>( 1500&lt;\text{Re} )</td>
</tr>
<tr>
<td>Tomiyama (1995)</td>
<td>( C_D = \frac{8}{3} \frac{\text{Eo}}{\text{Eo} + 4} )</td>
<td>Highly distorted– caps ( 4 \leq \frac{D_G}{\lambda_{RT}} ); ( \text{Re} \leq 100 \ 000 ); ( 0.01 \leq \text{Eo} \leq 1000 )</td>
</tr>
<tr>
<td>Tomiyama (1998)</td>
<td>( C_D = \text{Max} \left[ \text{Min} \left{ \frac{24}{\text{Re}} \left(1 + 0.15 \times \text{Re}^{0.687} \right) \frac{72}{\text{Re}} \frac{8}{3} \frac{\text{Eo}}{\text{Eo} + 4} \right} \right] )</td>
<td>For slightly contaminated.</td>
</tr>
<tr>
<td></td>
<td>( C_D = \text{Max} \left[ \frac{24}{\text{Re}} \left(1 + 0.15 \times \text{Re}^{0.687} \right) \frac{8}{3} \frac{\text{Eo}}{\text{Eo} + 4} \right] )</td>
<td>For contaminated.</td>
</tr>
<tr>
<td>Kolev (2002)</td>
<td>( C_D = \frac{2}{3} \frac{D_G}{\lambda_{RT}} )</td>
<td>Distorted bubbles – caps ( \frac{24}{\text{Re}} \left(1 + 0.1 \times \text{Re}^{0.78} \right) \leq \frac{2}{3} \frac{D_G}{\lambda_{RT}} \leq \frac{8}{3} )</td>
</tr>
</tbody>
</table>
A comparison of the different drag correlations was undertaken in terms of the predicted terminal velocity, which is compared against the typical rise velocity of air bubbles in water with size between \( D_G = 0.25-50 \text{ mm} \) in diameter (Fig 3.5), contrasting the different alternatives and identifying drag correlations that represent reasonably the momentum interaction over a wide range of conditions in pure and ‘contaminated’ systems.

![Graph showing terminal velocity vs. bubble size](image)

Fig 3.5 Terminal velocity of air bubbles rising in water computed using different drag correlations, and typical values for pure and contaminated systems.

It has been observed that bubbles tend to rise faster in pure systems, while the presence of contaminants causes a ‘stiffening’ of the inter-phase that results in a slower rise. The typical range of terminal velocity of air bubbles in water is contained in Fig 3.5 between...
the two thick green lines representing ‘pure’ and ‘contaminated’ air-water systems. Several aspects about the drag correlations can be noted from the comparison between computed rise velocities and typical terminal velocities:

- The cleanliness of the system has a significant effect on the rise of bubbles with diameter in between $0.5 \text{ mm} < D_G < 7.0 \text{ mm}$, where a lower terminal velocity is displayed for the contaminated cases.

- The Schiller and Naumann (1935) drag correlation can be used with confidence only for small bubbles ($D_G < 2.5\text{mm}$), and is therefore limited to low Reynolds numbers. The models that are available in Fluent for the exchange of momentum in bubbly flows are based on this correlation, which would under-predict the drag in cases where the bubble size exceeds 2.5 mm in diameter, and would introduce errors in the momentum transfer between the phases and the predicted flow field.

- Several correlations represent closely the rise of bubbles in pure systems, including the ones proposed by Wang et al. (1994), Tomiyama et al. (1995), and Sommerfeld et al. (2002) (all presented in red in Fig 3.5). Experimental studies in fluid mechanics are normally conducted in laboratories under ideal conditions, and therefore it is not surprising that several correlations would describe the behavior of bubbles in pure media. The Sommerfeld et al. (2002) correlation is chosen to represent pure air-water systems.

- The prediction using the Lain et al. (1999) correlation gives rise velocities that fall within the typical range and seem to represent better than others the hydrodynamic phenomena occurring in bubbles that rise in ‘contaminated’ systems.
Fig 3.6 presents the terminal velocity computed with the drag correlation selected to represent contaminated systems, which is super-imposed on the range of rise velocity of air bubbles in water under diverse clean and contaminated conditions measured by Patro et al. (2001). As normal in studies of gas-liquid flows, the range of dispersion in the reported data is significant, but still agreement can be observed between the calculated and reported terminal velocities, corroborating the capacity of the drag correlation to represent approximately the behavior of bubbles in practical ‘non-pure’ situations.

The drag coefficient correlations have been developed usually for single bubbles rising in stagnant media, and are exclusive functions of the bubble Reynolds number, like in the case of solid spherical particles. It is therefore interesting to visualize the different drag
correlations in the plane $C_D$ vs. Re (Fig 3.7), observing how the drag coefficient varies with Reynolds and compares with the drag of solid and gas particles:

![Graph showing drag coefficients for solid spheres and bubbles vs. Reynolds number.](image)

Fig 3.7 Drag coefficient of solid spheres and bubbles with respect to the Reynolds number.

The Schiller and Naumann (1935) drag correlation was developed originally for solid particles, and the predictions are in effect close to the drag coefficients measured for solid spheres (Schlichting, 1960). Small bubbles in contaminated systems are thought to behave as solid particles as a result of the stiffening of the gas-liquid interface, and therefore the use of this equation seems reasonable under such conditions, which probably would extend up to $Re < 30$.

The main differences between the drag in gas and solid particles seem to occur after $Re > 100$, where it is observed that the drag coefficient for bubbles increases up to $C_D \sim 2.61$ and reaches a constant value (drag coefficient of distorted ‘cap’ bubbles), while for solid particles the drag coefficient reduces monotonically with the Reynolds number until it reaches a constant value around $C_D \sim 0.44$. 


From the previous comparison of the drag coefficient vs. Reynolds number it can be said that in the numerical analysis of bubbly flows with small bubbles the selection of the drag model does not represent a major challenge, and using any drag correlation, including those developed originally for solid particles, would give satisfactory predictions of the exchange of momentum and the flow field. On the other hand, a less convenient situation is experienced as the bubble size and the Reynolds number are increased, observing significant differences between the correlations that suggest larger drag forces acting on gas-particles than on solid particles under equivalent hydrodynamic conditions. Therefore, the right selection of the drag model seems to become critical for analytical purposes as the bubbles and Reynolds number get large ($D_G > 0.25 \text{ mm} \ ; \ Re > 30$).

For the simulation of the flow in the lab-scale facility the momentum exchange correlations are introduced in the CFD code through the use of User Defined Functions (UDFs) programmed in C++. Special attention is given to the following three drag correlations, which represent the drag of solid spherical particles, bubbles in contaminated systems, and the effect of increasing the void fraction in bubbly flows:

i. Based on Schiller & Naumann (1935), this standard correlation is commonly used, and is the default model available to use with the commercial CFD code employed.

$$C_D = \frac{24}{Re} \left(1 + 0.15 Re^{0.687}\right) \quad ; \ Re \leq 1000$$

$$C_D = 0.44 \quad ; \ Re > 1000$$

ii. Based on work developed by Lain et al. (1999), the following drag correlation is applied to compute the exchange of momentum between air bubbles and tap
water, considered contaminated. This correlation incorporates the effect of contaminants.

\[
C_D = \frac{24}{Re} \left(1 + 0.15 Re^{0.687}\right) \quad ; \quad Re \leq 700
\]

\[
C_D = 9.5E - 0.5 \cdot Re^{1.397} \quad ; \quad 700 < Re \leq 1500
\]

\[
C_D = 2.61 \quad ; \quad 1500 < Re
\]

iii. Based on results presented by Rusche and Issa (2000), the following drag correction is applied to the drag coefficient to take into account the postulated increase of the interaction between the phases as the void fraction becomes larger:

\[
C_D^M = C_D^\alpha \left[ e^{3.64\alpha} + \alpha^{0.864}\right]
\]

3.3 
EXPERIMENTAL RESULTS

3.3.1 Two-phase Flow

The size of the bubbles has been determined by image analysis, calculating an equivalent diameter in the case of non-spherical bubbles. It was found that for the tap water case the smaller bubbles (\(D_G < 3\) mm) are approximately spherical, while larger bubbles tend to have an elliptical shape (Fig 3.8). An averaged volume equivalent to spheres of diameter \(D_G \sim 4.44\) mm was found; rounded to \(D_G \sim 4.50\) mm for analysis purposes.

The coalescence has been observed to be increasingly significant as the viscosity of the liquid is higher. When tap water was used a regular bubbly regime was obtained, without evidence of any significant merging or breaking of the injected bubbles. However, as corn syrup was added and the viscosity increased, it became noticeable that the bubbles would move towards the mid-plane of the vertical channels representing the calandria tubes and merge, forming larger bubbles and generating 3D recirculation effects (Fig 3.9).
Fig 3.8  Optical determination of the bubble size.

Water  Corn syrup 1  Corn syrup 2

Fig 3.9  Schematic representing coalescence of the air bubbles within the vertical channels of the lab-scale model at high viscosity.
It was difficult to determine a representative bubble size for the corn syrup cases due to the merging of the bubbles, which results in a large dispersion in the size and highly distorted shapes as the bubbles grow larger (Fig 3.10). Besides, it was found that the large bubbles produced severely interfere with the passage of laser light (Fig 3.11), making the application of Particle Image Velocimetry to measure the velocity of the two-phase flow region virtually impossible.

An additional difficulty related to the presence of the gas-phase was experienced only in the corn syrup cases, observing that very small bubbles ($D_g \ll 1\text{mm}$) were generated at the top, probably as a consequence of the collapse of departing bubbles as they cross the free surface, which were quickly entrained by the circulating flow to the entire domain of the test-rig. In spite of being small, the large quantity of these fine bubbles turned the liquid completely opaque (Fig 3.11), blocking the passage of light that is required for PIV measurements and making the detection of the tracer particles unfeasible.

![Tap water, regular bubbles](0.001Pa.s) ![Corn syrup, large & distorted bubbles](0.028Pas)

**Fig 3.10** Images illustrating that the coalescence increases with liquid viscosity, resulting in large highly distorted bubbles.
Fig 3.11  Images illustrating that with corn syrup the coalescence produced [a-left] large bubbles that cut the passage of laser light; and that [b-right] fine bubbles generated at the free surface were entrained and turned the liquid opaque.

The intended use of fluorescent tracer particles in combination with an optical long pass filter to measure the liquid velocity of the two-phase flow was unsuccessful. The bubbles cause severe light scattering that increases dramatically with the void fraction, diminishing the excitation of the fluorescent pigments and making it difficult to capture the position of the particles with the CCD.

The Coast yellow fluorophore was found inappropriate for PIV due to the low relative fluorescence (Fig 3.3), which impeded the detection of the particles with the CCD unless a very high laser power is used, which causes the bubble contours to appear in the images, failing the optical separation of the phases. Rhodamine has superior relative fluorescence (Fig 3.3), and the particles were clearly captured in the digital images, but as long as the airflow rate was null or small. Under the evaluated range of conditions in the test rig the void fraction is relatively high (estimated $\alpha \sim 18$-35%), causing severe scattering and obstruction of the laser and fluorescing light, and impeding therefore the detection of the fluorescent particles. Previous studies have referred to this difficulty, indicating that the refraction and reflection of the light occurring at the gas-liquid interface restrains the use of
optical techniques in gas-liquid flows. The use of the PIV technique is limited for bubbly flows with void fraction not higher than 5% (Larue de Tournemine et al., 2001)

3.3.2 Flow Patterns

The flow patterns observed when tap water was used as the liquid-phase can be broadly classified into two regions:

- Single-phase region: In the region below the channels and in the lower portions of the downtake the flow is mostly bubble-free and is classified here as essentially a single-phase region. Only very small bubbles are present in this section (small size bubbles are easily dragged by the downflow), leading to a low-void fraction region, where good PIV results were obtained without any major complication. The velocity measurements indicated a smooth flow pattern in this single-phase region at low circulation rates, with no flow separation as the flow negotiates the sharp corner at the bottom of the inner calandria channels (Fig 3.12.a). For higher circulation rates a flow separation is obtained at the bottom immediately below the inner calandria channels that are closer to the downtake (Fig. 3.12.b).

- Two-phase region: A significant quantity of gas bubbles is present in the channels, in the top section, and in the downtake, where many bubbles are dragged down, reducing the effective area and opposing the circulation (see bubbles in Figs 3.1 or 3.4). In this section the regime can be described as a high void-fraction flow, where the bubbles scattered light that was captured by the CCD, making it difficult to distinguish the seeding particles from the numerous bubble contours recorded. Measuring in this region was difficult, and although the flow patterns obtained with PIV seem to represent the gas-phase flow, the
accuracy of these results could not be verified. In this two-phase region vortices are consistently developed near the sidewalls above the calandria and in the upper half of the downtake region (Fig 3.13). Since the bubbles dominate the images in the two-phase region, the velocity vectors obtained by the PIV measurements in this region are interpreted here as bubble velocities.

![Image](image_url)

**Fig 3.12** PIV velocity vectors determined in the single-phase region indicating [a-top] a smooth flow in the bottom at low circulation rates, and [b-bottom] a flow separation developed below the channels at high circulation rates.

To identify possible three-dimensional effects, measurements were performed in three different planes: near to the front wall, at one quarter, and at half (mid-plane) of the depth of the test rig (Fig 3.14). For the single-phase section, water case, the results showed similar flow patterns and velocities that are close in magnitude between the three evaluated planes,
but slight differences are present indicating that 3D effects are likely to exist but are small.

In the bottom region only liquid is present, and therefore it is ideal for the integration of the velocity results and quantification of the circulation rate, which was obtained averaging the horizontal velocity in the space between the downtake wall and the bottom W shape.

![PIV velocity vectors determined in the lab-scale model.](image)

[Tap water - $H = 12$ cm - $J_G = 8.92$ cm/s]

Fig 3.13  PIV velocity vectors determined in the lab-scale model.

For all the experimental conditions evaluated a large-scale separation was detected in the two-phase section above the calandria near the side-walls (Fig 3.15), whose size varies according to the liquid head.

115
Fig 3.14  PIV velocity vectors determined in three different planes.

Fig 3.15  Vortex developed above the calandria near the side walls.
The circulating liquid entrains some of the injected bubbles into the downtake channel, where two adjacent vortices are generated and keep a recirculation of the secondary phase, which was illustrated by the PIV velocity vectors determined in that region (Fig 3.16). As the liquid head and circulation increase, more air is entrained into the downtake and the recirculation of the bubbles is higher, as can be seen illustrated in Figure 3.4.

![Tap water]

Fig 3.16 Large scale vortices developed in the downtake of the lab-scale model.

Different flow patterns were obtained as the liquid viscosity was increased by addition of corn syrup. PIV measurements were obtained for the less viscous corn syrup (1), but only for the bottom region and are not considered highly accurate due to the difficulties imposed by the coalescence and the entrainment of fine bubbles already discussed. Nonetheless, the determined PIV velocity vectors represent well some visible features of the flow, as lower velocity magnitudes in comparison with the tap water case and the development of a flow separation in the lowest part of the bottom (Fig 3.17). Note that the measured peak velocity with the corn syrup 1 was around $U_{\text{MAX}} \approx 0.3 \text{ m/s}$, while with tap water it reached $U_{\text{MAX}} \approx 0.5 \text{ m/s}$ (as seen in Figure 3.12).
In the tests with highest liquid viscosity [corn syrup 2] the entrainment of fine air bubbles was severe, and as a consequence the PIV measurements were impracticable. The flow patterns were outlined qualitatively by marking at several locations of the front wall the velocity direction that was observed (Fig 3.18). With respect to the previous cases a slower flow was evident, displaying smoother changes in direction and exhibiting a separation at the lowest part of the W shape towards the sides (marked in red in Fig 3.18).
Paradoxically the entrainment of large-scale bubbles (not the fine micro-bubbles) into the downtake was less pronounced for the highest viscosity liquid. This is attributed to the slower circulation and increased coalescence, which result in large bubbles that are harder to drag into the downtake and can escape more easily.

Important differences were noticed in the flow patterns and circulation as the liquid height and viscosity were changed, while the airflow did not show an important effect, probably due to the narrow range that was evaluated ($J_G = 6.2 - 7.4 - 8.8 \text{ cm/s}$). In general, the larger circulation rates are found at high liquid heads and cause the development of a large-scale flow separation in the bottom, whose size and location depends largely on the liquid viscosity (Fig 3.19). At low liquid heads above the channels it was observed that the circulation is progressively reduced as the head is lowered, and the flow separation at the
bottom shrinks and eventually vanishes as a result of lower liquid velocities. Similarly, vortices that are developed in the two-phase region above the calandria against the sidewalls and in the downtake get smaller as the viscosity is increased.

3.3.3 Circulation and Exchange of Momentum between Gas-liquid Phases

The circulation rate is a key performance metric of interest. Considering that the conditions for the PIV measurements were more favorable when tap-water was used and the results more accurate, and that most of the information reported in the literature on gas-liquid flows corresponds to the rise of air bubbles in water, the analysis of the circulation and the exchange of momentum will be focused largely on this case, which conveniently permits benchmarking with data from previous investigations. The velocity measurements in the bottom region, where reliable PIV results were obtained, have been used to quantify the circulation rate.

The circulation is perhaps the most palpable manifestation of the drag forces occurring within the test-rig, reflecting the transfer of momentum from the injected gas bubbles to the liquid phase, although the frictional resistance has also an effect by reducing the circulation. Figure 3.20 presents the circulation in the lab-scale model that has been determined from the PIV measurements at different liquid heads ($H \sim 0-30$ cm) and airflow rates ($J_G \sim 6.2 - 7.4 - 8.8$ cm/s), and Figure 3.21 presents the void fraction estimated from the displacement of the free surface. It can be appreciated that initially as the liquid head rises, there is a steep increase in the circulation and a steep decrease in the void fraction. This behavior persists until a critical head is achieved ($H \sim 10$ cm for the test rig, corresponds to 60 cm in a full-scale crystallizer), after which further increases in the liquid level have a small effect on the circulation and void fraction.
Fig 3.20 Effect of the liquid head over the calandria on the circulation rate. The circulation is presented in terms of [a-left] the total liquid volumetric flow and [b-right] the liquid superficial velocity within the vertical channels.

Fig 3.21 Effect of the liquid head over the calandria on the void fraction.

The measurements in the test-rig at low liquid heads have indicated that the void fraction tends to increase as the superficial liquid velocity, or circulation, is lower. This trend is in agreement with two-phase algebraic models and experimental data obtained in the study of air-lift reactors.

The results presented in figures 3.20 and 3.21 suggest that at low liquid heads the drag exerted on the bubbles might show important variations as the circulation increases and the void fraction goes down, but it becomes practically constant after the ‘critical’ head is
exceeded, when the drag coefficient reaches an almost constant value. The poor circulation observed for low liquid heads is attributed to the smaller sectional area above the channels or calandria, and as a consequence, an increase in the frictional resistance of the circuit.

In full-scale sugar evaporative crystallizers, for well-established reasons, the evaporation rate decreases during each strike, causing the circulation to decrease correspondingly as the massecuite level increases. The maximum evaporation occurs at the start of the strike, but the perception that the maximum circulation takes place simultaneously is not supported by the results obtained, which show that for low heads a ‘bottle-neck’ effect above the calandria is obtained, restricting the liquid from flowing toward the downtake. This behavior was inferred years ago by Allan (1962), who proposed that a lower head means a smaller sectional area above the calandria, and therefore an increase in the flow resistance, and consequently lower circulation. This result suggests the existence in the design of continuous sugar evapo-crystallizers of an optimum massecuite height.

Since the liquid circulation rate is related to the heat transfer coefficient, the present results are consistent with the reported data on the effect of the liquid head on full scale evapo-crystallizers by Austmeyer (1986), who observed that initially increasing the liquid level increases the heat transfer up to a maximum at H ~ 0.8 m (for batch after-product boiling), beyond which a decrease in the heat transfer coefficient occurs due to a reduction in the temperature difference caused by the higher hydrostatic pressure.

It is acknowledged that in full-scale evaporative crystallizers the evaporation rate and massecuite circulation are interrelated. The test rig data obtained with tap-water did not show a significantly distinguishable effect of the injected airflow on the circulation rate, probably due to the narrow range of airflow conditions evaluated.
3.3.4 Slip Velocity and Drag Coefficient

The experiments with tap water in the lab-scale rig have indicated slip velocities within the vertical channels in between $\Delta u = 0.12$–0.23 m/s, which are comparable with the typical range of terminal velocities of air bubbles equal in size ($D_G \sim 4.5$ mm). Figure 3.22 presents a graphic comparison of the determined slip velocity with terminal velocities that were reported by Patro et al. (2001), where agreement can be noticed between the high slip values recorded at high head/circulation and the reference data, but slip velocities far below the typical range are also observed. It should be kept in mind that the experimental results correspond to the averaged slip velocity in the vertical channels of the test-rig, where a bubbly flow exists presenting complex momentum interactions, high void fraction, distorted bubbles, and mutual bubble interactions, while the reference data correspond to the terminal velocity of single bubbles rising in a stagnant media under ideal conditions.

![Fig 3.22 Comparison of the slip velocity determined in the vertical channels of the lab-scale model against data on the terminal velocity of air bubbles in water reported by Patro et al. (2001)]
Table 3.5 presents the hydrodynamic parameters of the bubbles rising in the channels of the test-rig and for single air bubbles in water predicted with the drag correlations selected to represent pure (Sommerfeld et al., 2002) and contaminated systems (Lain et al., 1999). It can be appreciated how the lower values of the slip velocity result in lower Reynolds numbers and higher drag coefficients, which can reach values that exceed several times the typical drag coefficient for single bubbles (4.08/1.53 = 2.66 times!).

Table 3.5  Hydrodynamic parameters for air bubbles with $D_G = 4.5$ mm in diameter rising in water computed from drag correlations and determined in the lab-scale model.

<table>
<thead>
<tr>
<th></th>
<th>Pure system</th>
<th>Contaminated system</th>
<th>Test rig Tap-water</th>
</tr>
</thead>
<tbody>
<tr>
<td>Void fraction - $\alpha$</td>
<td>0%</td>
<td>0%</td>
<td>18-35%</td>
</tr>
<tr>
<td>Re</td>
<td>1150</td>
<td>1027</td>
<td>540 – 1010</td>
</tr>
<tr>
<td>$C_D$</td>
<td>1.22</td>
<td>1.53</td>
<td>1.15 – 4.08</td>
</tr>
<tr>
<td>$\Delta u$ (m/s)</td>
<td>0.231</td>
<td>0.206</td>
<td>0.12 – 0.23</td>
</tr>
</tbody>
</table>

Based on previous experiences reported on bubbly flows, the lower values of the slip velocity with respect to the terminal velocity seem reasonable as they would be the natural consequence of the alleged increase in the drag or momentum interaction as the void fraction becomes higher, or also a possible consequence of the loss of buoyancy of the bubbles as the density of the surrounding mixture is reduced.

Figure 3.23 presents the slip velocity determined in the channels of the test-rig with respect to the liquid level over the calandria and the void fraction. Significant variations in the slip are obtained for the low liquid heads, or low circulation rates, while a practically constant velocity is observed after the ‘critical’ head is exceeded, which displays some proportionality with the injected airflow and remains relatively close to the terminal velocity predicted with the drag correlation for contaminated systems (Lain et al., 1999). The effect
of the void fraction on the terminal velocity is introduced with the drag correction multiplier proposed by Rusche and Issa (2000) and is presented in Figure 3.23, predicting a monotonic decrease in the theoretical velocity as the void fraction is higher.

The predicted effect of the void fraction reducing the rise velocity of the bubbles agrees with the experimental data obtained in the lab-scale facility for liquid heads above $H \geq 5$ cm, while below this level the opposite tendency is found and the slip increases with void fraction (Fig 3.24). At low liquid heads ($H < 5$ cm) the circulation is minimal and the vertical channels could be compared with bubble columns, where strong liquid recirculation takes place. On the other hand, the results obtained with high heads ($H > 5$ cm) are comparable with air-lift reactors, observing significant circulation rates and high liquid superficial velocities. Therefore, major differences in the hydrodynamic characteristics of the flow within the channels of the test-rig potentially exist between the low and high liquid heads, which could be related to the ambiguous effect of the void fraction on the slip that has been observed. It is noticed that the critical values where the slip minimums are located
occur at void fractions between $\alpha \sim 0.24-0.28$, in the same range where the departure from the bubbly regime is reported to occur ($\alpha \sim 0.25-0.30$), although no visible evidence of such a transition was noticeable.

A more fundamental appreciation of the experimental results is obtained from the plot of the drag coefficient with respect to the Reynolds number (Fig 3.24), which permits a comparison of the drag coefficient determined from the tests with the drag correlations that were discussed previously and will be applied in the numerical simulation of the flow. It is observed that the drag coefficients determined at high liquid heads are closer to the curve representing the drag in contaminated system, while all the other considered correlations would under-predict the drag forces. The drag coefficient determined at low heads exceed the values predicted with the correlations, presumably as a consequence of the high void fraction and/or liquid recirculation effects.

![Fig 3.24 Drag coefficient vs. Reynolds number determined from the experimental data obtained in the lab-scale model, and computed with drag correlations for bubbles.](image-url)
Figure 3.25 presents the effect of the void fraction on the drag coefficient, which is compared with the drag coefficient for bubbles in contaminated media predicted with the Lain et al. (1999) correlation and corrected with the expression proposed by Rusche and Issa (2000) to include the effect of the void fraction. It is observed that the experimental results obtained at high liquid heads, or high circulation rates, display an increase in drag with void fraction that is comparable in slope with the curve representing the theoretical correction proposed by Rusche and Issa (2000), although the almost constant value reached in all cases after the liquid head exceeded the ‘critical’ head (H > 10 cm) remains relatively close to the drag for single bubbles (C_D ~ 1.53). Conversely, at low liquid heads the drag exhibits a tendency to decrease with void fraction, particularly after α ~ 0.24–0.28, which openly contradicts the scarce information that is available on the effect of void fraction in bubbly flows.
The application of high-speed photography has illustrated that in the test rig the bubbles suffer continuous changes in shape or deformation, and that the bubble geometry differs significantly from the ideal spherical shape that is customarily assumed for numerical analysis (Fig 3.26). It could be hypothesized that the continuous variations in geometry of the gas particles at high void fractions are a palpable consequence of the multiple interactions with neighboring bubbles and the fluctuations in the continuous phase associated with the pseudo-turbulence, which are likely to be connected with the observed increase in the drag coefficient at moderately high liquid heads. The opposite behavior has been detected after the void fraction exceeds $\alpha \sim 0.24–0.28$, where reductions in the drag coefficient with void fraction have been recorded. This change in the trend could be attributed to larger wake interactions and coalescence occurring as the void fraction gets higher and the space between the gas-particles is reduced\textsuperscript{14}, and is likely to be associated with the departure from the bubble regime.

Fig 3.26  Image of a distorted air bubble rising in tap water within the lab-scale model.

Bubbly channel flows have been extensively investigated, and several models are available for the prediction of different parameters and the flow regime. Figure 3.27 presents

\textsuperscript{14} Krishna et al. (1999) indicates that the gas rise velocity in churn-turbulent flows increases as a result of higher wake interactions (trailing bubbles rise faster) and the larger size of the bubbles due to coalescence.
a comparison of the experimental results focusing in the channels of the test rig against an empirical model developed by Premoli et al. (1971), which is reported to cover a wide range of conditions. Reasonable agreement is observed between the experimental data and the algebraic multiphase model, particularly at high liquid heads, where vigorous circulation occurs and the lowest values of the void fraction were recorded.

![Comparison of the void fraction measured in the lab-scale model and predicted using the algebraic multiphase correlation developed by Premoli et al. (1971)](image)

**Fig 3.27** Comparison of the void fraction measured in the lab-scale model and predicted using the algebraic multiphase correlation developed by Premoli et al. (1971)

### 3.4 NUMERICAL RESULTS

The CFD simulation results showed strong sensitivity to the model used for the exchange of momentum between the gas and liquid phases. In general, the results obtained with the exchange model based on the drag correlation developed by Schiller and Naumann (1935), which might represent well solid spherical particles and fine bubbles, displayed significant over-prediction in the velocity field, and consequently in the circulation, with respect to the PIV data. On the other hand, the results obtained computing the phase exchange of momentum with the drag correlations proposed by Lain et al. (1999) and the
correction for void fraction proposed by Rusche and Issa (2000) display a flow field and a circulation rate that are closer to the experimental results.

Figure 3.28 presents the velocity vectors in the single-phase region (below channels) that were measured with PIV and the liquid velocity vectors that were predicted using the drag correlation chosen to represent bubbles rising in non-pure systems (Lain et al., 1999). In this single-phase region both measurements and simulations show velocity vectors that are comparable in magnitude and direction, depicting the flow separation in the bottom, and indicating a reasonable agreement between the measured and predicted flow fields.

![Fig 3.28 - Velocity vectors measured using PIV, and liquid velocity vectors predicted numerically for the single-phase region of the lab-scale model.](image)

Fig 3.28 [a-top] Velocity vectors measured using PIV, and [b-bottom] liquid velocity vectors predicted numerically for the single-phase region of the lab-scale model.

A more exhaustive comparison of the liquid flow field is shown in Fig. 3.29, where the predicted and measured velocity profiles below the downtake wall are presented. The profiles correspond to measurements effectuated for two different liquid head levels ($H = 5,$
12 cm) and the three airflow conditions evaluated ($J_G = 6.2$-7.4-8.8 cm/s). For the lower head ($H = 5$ cm), where there is no flow separation below the calandria channels, velocities increase from the lower wall reaching a peak of around 0.3 m/s, and then remain relatively flat till the calandria channel wall is reached. There is a slight increase in the velocity with increasing airflow ($J_G$). For liquid head $H=12$ cm, as seen in Fig. 3.29, flow separates below the calandria channels, and the peak velocities are skewed closer to the lower wall of the test-facility. The predicted peak velocities are about 25% lower, and the predicted profile is flatter compared to the measurements.

![Velocity profile measured underneath the downtake wall of the lab-scale model applying PIV, and predicted with CFD.](image)

Figure 3.30 presents velocity vectors in the two-phase region that were measured with PIV, and gas velocity vectors predicted numerically using the drag correlation proposed by Lain et al. (1999) and the void fraction correction proposed by Rusche and Issa (2000). It was argued earlier that the PIV measurements in the two-phase region could be interpreted as the gas-phase velocities, but with some uncertainty regards to accuracy. Some similarity in
the flow patterns is evident, and both measurements and computations indicate vectors that are similar in direction, and exhibit two large vortices, one in the downtake region and another along the sidewall immediately above the channels. Despite the qualitative similarity, and the same range of velocities in the predictions and measurements, the two sets of results differ in magnitude.

Fig 3.30  Velocity vectors [a-left] measured using PIV, and [b-right] gas velocity vectors predicted numerically for the two-phase region of the lab-scale model.

Figure 3.31 presents the contours of gas volumetric fraction predicted numerically. The results reflect the concentrated injection of the air in the channels at the spargers, the rise of the mixture gas-liquid towards the free surface, where most of the gas crosses and
separates from the liquid, while some air is entrained by the circulation into the downtake region. The void fraction predicted within the channels is in average $\alpha \sim 0.2108$, a value that is relatively close to the void fraction estimated for the same case ($\alpha \sim 0.23$) from the rise of the free surface. The difference between the computed and estimated void fraction cannot be judged as a bad sign considering the uncertainty in the procedure that was followed to determine the void fraction.

![CFD predicted contours of void fraction for the lab-scale model.](image)

[Tap water - H = 12 cm - J_G = 8.8 cm/s]

**Fig 3.31** CFD predicted contours of void fraction for the lab-scale model.

Figure 3.32 presents the contours of static pressure obtained with the CFD computations, displaying a field that is essentially dominated by the hydrostatic effects. This pressure field may be considered normal due to the relatively low velocity and high density of the liquid phase. The pressure results are roughly validated comparing the maximum predicted pressure with the maximum hydrostatic head:
- The maximum static pressure predicted (CFD) at the lowest point is \( P = 5720 \text{ Pa} \).

- The maximum hydrostatic head is 0.583 m, and the correspondent hydrostatic pressure can be computed as: \( P = \gamma g h = 998.2 \times 9.8 \times 0.583 = 5708 \text{ Pa} \)

- Therefore a relatively small difference (0.2%) exists between the computations and the estimated maximum pressure, indicating that the pressure field is being predicted acceptably and the magnitude is physically reasonable.

![CFD predicted contours of static pressure for the lab-scale model.](image)

[Tap water - \( H = 12 \text{ cm} \) - \( J_G = 8.8 \text{ cm/s} \) – unit: Pa]

Fig 3.32  CFD predicted contours of static pressure for the lab-scale model.

The pressure predicted just below and at the entrance of the vertical channels representing the calandria tubes is lower than in the downtake region. This pressure difference has been discussed by sugar technologists for a long time, and would be the consequence of the lower density of the gas-liquid mixture present within the channels.
Figure 3.33 presents a comparison of the circulation rate obtained experimentally from the PIV measurements and predicted numerically by integrating the liquid flow across the downtake. The simulations using the drag model based on the Schiller and Naumann (1935) correlation (Fluent default) display a significant overprediction of the circulation rate with respect to the measured values (nearly twice). The inclusion of the drag correlation proposed by Lain et al. (1999) in the computations resulted in less momentum transferred to the liquid phase, and predictions of the velocity field and circulation that are closer to the PIV results. Further increase in the level of momentum interaction through the use of the correction for high void fractions proposed by Rusche and Issa (2000) resulted in an even lower circulation rate, which falls slightly below the experimental results.
From the comparison presented in Fig. 3.33 it is evident that the use of the drag correlation proposed by Lain et al. (1999), which was found to represent reasonably well the rise of air bubbles in non-pure water (contaminated, as tap water), was a critical step for obtaining correct predictions of the flow velocities and the circulation rate in the lab-scale test rig. Therefore, it can be said that the use of momentum exchange models based on suitable drag correlations for the particular conditions of the flow is fundamental in the numerical analysis of this multiphase case. A particular characteristic of this buoyancy-driven process is that the liquid flow depends largely on the momentum interaction, which must be precisely computed in order to attain reliable predictions, this even within the ‘simplest’ gas-liquid regime, the homogeneous bubbly flow that was obtained in the lab-scale facility when tap water was the primary phase.

3.5 FINAL REMARKS

The circulation and flow patterns in sugar evaporative-crystallizers have been studied experimentally in a lab-scale model using PIV, and numerically using CFD. Several important aspects about the flow and the necessary conditions for a correct application of the numerical techniques in the analysis of the process have been identified:

- The buoyancy-driven mechanism of massecuite circulation has been represented in the lab-scale facility injecting air bubbles in vertical channels that represent the calandria tubes. The experimental results have been compared with the typical rise velocity of single bubbles in water and different drag correlations, observing agreement with an expression selected to represent the rise of bubbles in contaminated media (Lain et al., 1999). Similarly, agreement was observed between the experimental data and an empirical two-phase model (Premoli et al., 1971).
- Significant differences in the flow patterns and circulation have been observed to occur in the lab-scale model as the liquid head above the calandria and the liquid viscosity change, developing flow separations and vortices that vary in position and size. The circulation results indicated that too low liquid heads above the calandria would be detrimental for the circulation as they increase the frictional resistance. This result is in agreement with information presented by Austmeyer (1986) on the effect of the massecuite head on the heat transfer coefficient in full-scale batch sugar evapo-crystallizers. The compromise between frictional resistance above the calandria and boiling point elevation must be carefully considered during the design and operation of evaporative-crystallizers. Although a decrease in the liquid level reduces the hydrostatic pressure in the boiling region, increasing the effective temperature difference, below certain critical limit it increases drastically the frictional resistance, and might cause the opposite effect since a lower circulation will reduce the heat transfer.

- An ambiguous effect of void fraction on the drag coefficient has been observed at low circulation rates. At high liquid heads and high circulation rates the drag increases with void fraction, as has been reported in the literature (Rusche and Issa, 2000). Conversely, at low liquid heads, when the circulation was small and the void fraction reached the highest values, the opposite trend was evidenced and the computed drag decreased with void fraction. The change in the effect of the void fraction is attributed here to the liquid recirculation and the increase in wake interaction and coalescence that would take place as the void fraction goes up and the distance between the gas-particles is reduced. This phenomena could be related to the
departure from the bubbly regime, which has been reported to occur when the void fraction is around \( \alpha \approx 0.25-0.30 \), while in the lab-scale tests the critical points where the minimum slip velocities (maximum drag coefficients) were recorded correspond to void fractions in the same range, estimated to be around \( \alpha \approx 0.24-0.28 \).

The numerical prediction of the flow in the lab-scale model using the Eulerian-Eulerian CFD approach with different drag equations has shown the importance of selecting an adequate model to compute the phase momentum interaction, which plays a particularly critical role in this buoyancy-driven flow, largely determining the magnitude of the velocities and circulation rates that are predicted, and consequently the flow patterns. It is anticipated from the experience with the lab-scale test rig that if the exchange of momentum between the gas and liquid phases is solved properly, an acceptably accurate solution of the entire flow-field within sugar evaporative crystallizers will become feasible.
CHAPTER 4 - FUNDAMENTAL STUDIES OF THE MOMENTUM INTERACTION IN BUOYANCY-DRIVEN GAS-LIQUID VERTICAL CHANNEL FLOW

The experimental and numerical results obtained with the lab-scale facility (chapter 3) indicated that the flow in sugar evaporative crystallizers may be well represented by applying the Eulerian-Eulerian CFD multiphase model, provided that the momentum interaction or drag coefficient between the gas and liquid phases is specified properly. However, most drag correlations have been developed for single small bubbles rising in stagnant media, usually water, or within the bubbly regime at low void fractions. No established correlations exist for the more complex regimes that occur at high void fractions and are potentially present in the calandria boiling tubes.

Considering the lack of understanding on the drag or momentum interaction between the gas and liquid phases at high void fraction, a fundamental study of the momentum exchange in buoyancy-driven gas-liquid vertical channel flows has been carried out as a part of this project. The goal of the study is to provide a closer insight into the phenomena occurring in vertical multiphase and boiling channel flows, and a correlation for the interfacial exchange of momentum that can be applied in the simulation of the fluid flow in sugar evaporative crystallizers. To address these goals, an experimental facility was constructed to study the interfacial transfer of momentum in gas-liquid buoyancy-driven vertical channel flows under adiabatic conditions. Results from this experimental study are presented, observing that the circulation increases with the gas flow rate asymptotically and displays significant oscillations, particularly with highly viscous liquids, which are related to the unsteadiness and intermittency of the slug flow regime. The results show that the
interfacial transfer of momentum is reduced progressively as the gas content and void fraction are higher.

The complex phenomena occurring in calandria tubes are discussed based on information reported from experimental work on mass and massecuite boiling (Austmeyer and Schliephake, 1983; Rouillard, 1985; Austmeyer, 1986; Bruhns, 1996) and relevant information reported from more fundamental studies of boiling in vertical heated tubes (Griffith, 1962; Jeglic and Grace, 1965; Shoukri et al., 1981; Subki et al., 2004; Wienecke, 2006). It is proposed that flow boiling instability is developed in the calandria tubes of vacuum pans, generating oscillations that affect the heat transfer and interfacial momentum interaction and cause intermittent circulation. Results from field measurements in a batch evaporative crystallizer showing that the vaporization is a discontinuous and periodic process are presented, confirming that a boiling instability takes place.

4.1 MATERIALS AND METHODS

An experimental facility was developed for the evaluation of the interfacial transfer of momentum in buoyancy-driven gas-liquid vertical channel flows, which essentially comprises a circulation loop, similar in concept to air-lift reactors, where air is injected at the bottom of a ‘riser’ section and the buoyancy force due to density differences between the gas and liquid phases is the only driving force for liquid circulation. The experimental facility is shown in Figures 4.1-4.3, illustrating the main components and dimensions. The main components of the experimental facility are:

- Riser: 3” PVC transparent pipe with internal diameter ID ~ 0.0779 m and length L ~ 2.43 m (Figures 4.1 & 4.2.i). The transparent wall of the pipe permits using image analysis to study the flow and visualization to identify the dominant flow regime.
Fig 4.1 Schematic of the experimental facility for evaluation of transfer of momentum in buoyancy-driven gas-liquid vertical channel flow showing main components and dimensions.
Fig 4.2 Images of the experimental facility for evaluation of transfer of momentum in buoyancy-driven gas-liquid vertical channel flow, illustrating [i] the main components of the test rig, [ii] the high speed camera and lamps, [iii] the air pressure and flow meters, and [iv] the absolute pressure transmitter.
Fig 4.3 Schematic of arrangement of needles used at the bottom of the riser to inject bubbles from orifices with different size.

- A 3” stainless steel pipe and 3” clear tubing are used as down-comer (Figures 4.1 & 4.2.i), permitting the return of the liquid phase to the bottom of the riser.

- A sparger was constructed drilling 110 holes of 3.175 mm (1/8”) in diameter around a 3” PVC pipe, which is used to inject the gas phase at the bottom of the riser section (Figure 4.1). Additionally, needles with different orifice sizes (ID = 0.254, 0.508, 1.066, 2.159, 2.99, 3.81, 4.8, and 10.74 mm) were installed at the bottom of the rig to inject single bubbles (Figures 4.1 & 4.3). The buoyancy force generated by the density difference between the injected air and the liquid phase is the only driving force for the circulation, as in full-scale sugar evaporative crystallizers.

- A plastic tank is installed above the riser, which has the top open, permitting the separation and release of the air to the atmosphere (Figures 4.1 & 4.2.i).
- A 1½” globe valve with full port connected between the tank and the Coriolis meter is used to increase the frictional resistance of the loop, controlling to some extent the circulation rate or superficial velocity of the liquid in the riser (Figures 4.1 & 4.2.i).

The instruments used for the measurements in the experimental facility are:

- High-speed camera: A Fastec Troubleshooter HR digital camera (Figure 4.2.ii) has been employed to record images of the gas-liquid flow in the riser, giving information about the shape, dimensions, and velocity of the gas-particles. The camera is provided with an optical MV-13 sensor, 1” C-mount format, maximum resolution of 1280 x 1024, pixel size 12 microns, and active area of 15.36 x 12.29 mm. To determine the size of bubbles and the respective velocity the software Motion Measure Suite, provided by Fastec Imaging, has been used.

- For the measurement of the gas flow rate two Cole Parmer flow meters are used, covering a low (4 - 50 SCFH) and a large (60 - 6000 SCFH) range correspondingly (Figure 4.2.iii). The mass flow of air is obtained in combination with a Marsh Master Gauge pressure meter (K-Monel tip and tube, 0-30 Psig), assuming that the air obeys the ideal gas law

- A Coriolis mass flow meter Honeywell SCM 3000 PLUS F-Series (Figures 4.1 & 4.2.i) with a 1.5” sensor (CMFS 40 AAWØØA25 B1A; 0-45 t/h; Accuracy ±0.005% of full scale; Repeatability ±0.0025% of full scale), has been used to quantify the circulation rate. Additionally, it gives information on the temperature and density of the circulating liquid phase.

- A critical variable for the intended analysis is the void fraction, which has been determined using the manometric method, measuring the pressure at the bottom of the
riser with an absolute pressure transmitter Rosemount 3051 CA (range 0-207 mm Hg, calibrated 0-760 mm Hg, accuracy ±0.1% full scale, damping 0.74 sec) (Figure 4.2.iv).

The tests were designed to determine the transfer of momentum between the gas and liquid phases, focusing attention on the regimes at high viscosity and high void fraction. Initially tap water was used as the medium, and then corn syrup diluted with water at several proportions to cover a range of viscosities. Table 4.1 presents the physical properties of the different liquid media conditions that were evaluated during the tests.

Table 4.1  Physical properties of the liquid media used during the experiments.

<table>
<thead>
<tr>
<th>Liquid condition</th>
<th>Concentration (° Brix)</th>
<th>Temperature T_L (° C)</th>
<th>Viscosity (\mu_L) (Pa.s)</th>
<th>Density (\rho_L) (kg/m³)</th>
<th>Inverse viscosity number - (N_\mu)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 – tap water</td>
<td>0.00</td>
<td>29.7</td>
<td>0.0008</td>
<td>996</td>
<td>84 783.25</td>
</tr>
<tr>
<td>2</td>
<td>57.70</td>
<td>31.9</td>
<td>0.054</td>
<td>1248</td>
<td>1 582.84</td>
</tr>
<tr>
<td>3</td>
<td>60.59</td>
<td>31.0</td>
<td>0.08</td>
<td>1265</td>
<td>1 076.11</td>
</tr>
<tr>
<td>4</td>
<td>64.23</td>
<td>29.4</td>
<td>0.18</td>
<td>1286</td>
<td>495.88</td>
</tr>
<tr>
<td>5</td>
<td>67.65</td>
<td>32.1</td>
<td>0.30</td>
<td>1306</td>
<td>293.44</td>
</tr>
<tr>
<td>6</td>
<td>68.62</td>
<td>31.4</td>
<td>0.38</td>
<td>1312</td>
<td>232.68</td>
</tr>
<tr>
<td>7</td>
<td>71.16</td>
<td>31.5</td>
<td>0.71</td>
<td>1328</td>
<td>128.07</td>
</tr>
<tr>
<td>8</td>
<td>74.33</td>
<td>31.9</td>
<td>1.91</td>
<td>1348</td>
<td>48.07</td>
</tr>
<tr>
<td>9</td>
<td>79.06</td>
<td>26.6</td>
<td>10.05</td>
<td>1379</td>
<td>9.34</td>
</tr>
<tr>
<td>10 – corn syrup</td>
<td>83.30</td>
<td>29.5</td>
<td>111.33</td>
<td>1407</td>
<td>0.86</td>
</tr>
</tbody>
</table>

NOTES: The viscosity was determined with a cone and plate viscometer Brookfield model DV-II using the spindles cP-42 and cP-52 and a thermostatic bath at the average temperature displayed by the coriolis temperature sensor during the tests (~30 °C).

The Brix, or concentration of soluble solids, was determined using a refractometer Bellingham + Stanley Ltd. model RFM 340 and samples at room temperature (~25 °C).

The dimensionless inverse viscosity number, originally proposed by Wallis (1962) to account for the effect of viscosity in the slug regime, is computed according to the following formula using the internal diameter of the riser of the test rig (0.0779 m):

\[
N_\mu = \frac{1}{\mu_L} \sqrt{D^3 \cdot g \cdot \rho_L \cdot \Delta \rho}
\]
4.2 RISE OF SINGLE GAS PARTICLES - LOW VOID FRACTION

4.2.1 Flow Visualization

The size and velocity of single bubbles have been determined from images captured with the high-speed digital camera operating at 1280 x 512 pixels, 125 fps, shutter 30x; rate that is high enough to register the rising velocity at low-void fraction.

In the experiments it was observed that bubbles rising in highly viscous media tend to preserve their sphericity even at large sizes, until they approach dimensions corresponding to the diameter of the channel. At this stage, growth is only possible in the axial (vertical) direction and the typical shape of Taylor bubbles begins to develop. Ishii and Hibiki (2006) indicated that when the bubble size reaches 60% of the channel the wall effects become important and start affecting the shape of the gas particle. Figure 4.4 illustrates the described behavior with bubbles of different size rising in highly viscous corn syrup.

![Fig 4.4 Images of air bubbles rising in a high-viscosity medium (μL ~ 10.1 Pa.s), displaying sphericity even at large size. The image on the right illustrates how as the bubble size approaches the internal diameter of the channel an elongated Taylor shape is developed.]

When the viscosity of the media is low, the small bubbles display an approximately spherical geometry; while ellipsoidal and distorted (cap) shapes are found as the size of the bubble is increased. Figure 4.5 presents images of bubbles captured during the tests with
water, showing how the geometry changes continuously and differs in all the cases from the ideal spherical geometry. The distortion and variations in shape are more severe as the size of the bubble increases, making it difficult to characterize the dimensions and velocity from 2-D images. A highly distorted bubble (cap) observed when air was injected at low rates through the sparger is also presented, which corresponds to the churn-turbulent regime. The pictures presented in Figure 4.5 illustrate how the geometry of bubbles passes from roughly spherical to elliptical and distorted ‘mushroom’ or ‘cap’ as the size is increased.

Fig 4.5 Images of [a-top] non-spherical bubbles \( \{\text{Re} \sim 1\, 170 - 2\, 140; \text{We} \sim 0.86 - 1.65\} \), and [b-bottom] a highly distorted cap bubble \( \{\text{Re} \sim 19\, 400; \text{We} \sim 22\} \) rising in tap water.
The changes in geometry of bubbles observed during the tests are consistent with observations reported by Habermann and Morton (1953) and Pebles and Garber (1953), who performed extensive studies on the rise of bubbles in different liquid media. They observed that small bubbles tend to be spherical, but as the size increases they become ellipsoids (Fig 4.5.a), and then they adopt a spherical or elliptical cap shape, like mushrooms (Fig 4.5.b). The small spherical and large cap bubbles rise straight and display consistent drag coefficient values, approximating the drag trend for solid spheres in the case of small bubbles, and around $C_D \sim 2.6$ for large cap bubbles (Moore, 1965). On the other hand, the elliptical bubbles observed in between rise with oscillatory motion and display different drag behavior according to particular properties of the media (Brennen, 2005). This dissimilarity in the drag has prevented developing a universal drag expression to describe the hydrodynamic behavior of single bubbles.

In the case of ‘small’ bubbles, the surface tension forces are dominant and lead to spherical shapes. Surface active impurities make the interface of small bubbles stiffer, resulting in fluid particles that behave as solid spheres in most practical situations. The Weber\textsuperscript{15} number gives the ratio between inertial and surface tension forces, and has been used to define the limit where the surface tension effect is overcome, and departure from sphericity or deformation initiates. Brennen (2005) suggests the following limits for spherical bubbles:

\begin{align*}
\text{Low Reynolds (Re }<<1): & \quad \text{bubble remain spherical if } \quad \frac{\text{We}}{\text{Re}} \leq 1 \\
\text{High Reynolds (Re }>>1): & \quad \text{bubble remain spherical if } \quad \text{We} \leq 1
\end{align*}

\textsuperscript{15} We ver number: $\text{We} = \frac{\text{Inertia}}{\text{Surface tension}} = \frac{\rho_L \cdot \Delta u^2 \cdot D_g}{4 \cdot \sigma}$
A small Weber number indicates that the surface tension force is large, and consequently the corresponding bubbles would tend to be spherical. It has been observed that even large-scale bubbles can be spherical as long as the Weber number remains small (Soo, 1967).

As the size of the bubbles increases an intermediate region is reached, where the bubbles become flatter, adopting the shape of an oblate spheroid or ellipsoid that undertakes continuous deformation (Fig 4.5.a). Haberman and Morton (1953) observed for bubbles in different liquids that departure from sphericity occurs when the Reynolds number reaches the order of Re ~100. Visualization experiments have indicated that the wake behind the bubbles becomes turbulent when Re ~ 360 (Brennen, 2005).

Further increases in the size of bubbles results in transition to a region where turbulence becomes the controlling factor (Soo, 1967), and the resultant bubbles adopt a highly distorted asymptotic cap geometry (Fig 4.5.b). The transition to this turbulent region occurs at large Weber numbers (We ~ 20), where flow separation from the bubbles (Davies and Taylor, 1950) and instability at the interface (Moore, 1965) cause significant changes in the flow. Cap bubbles rise with a straight motion defined by a universal drag coefficient ($C_D \sim 2.6$), although they display continuous deformations.

When the size of the bubbles approaches the size of the channel, Taylor bubbles are generated, which are normally described as elongated bullet-shaped bubbles that cover almost the entire cross sectional area of the channel ($D_{TB} \sim 90\%$ ID), and rise generating a liquid film that flows downwards between the bubble and the wall. Figure 4.6 presents images corresponding to Taylor bubbles rising in water and a highly viscous corn syrup. A significant difference between the two cases is observed at the tail of the bubble, which is flat.
and unstable with low-viscosity liquids, while a bullet shape tail is obtained with high-viscosity liquids. This behavior is attributed to the reduction in the strength of the wake behind the bubbles as the viscosity increases. Note that surface tensions are not substantially different in the two cases.

Fig 4.6 Images of Taylor bubbles rising in [a-top] tap water and [b-bottom] high-viscosity corn syrup, illustrating a difference in the shape of the bubble, whose tail is flat and unstable at low viscosity, while it adopts a bullet shape with the high viscosity medium

4.2.2 Drag Coefficient

The drag coefficient of the bubbles is evaluated using the formula that results from the balance between buoyancy and drag for ideal spherical spheres that was presented in chapter 2. The drag is computed from the slip velocity ($\Delta u$) and the equivalent volumetric
size of the bubbles (Dg) determined from analysis of the images acquired with the high speed camera. As usual in the study of hydrodynamic forces exerted by flows around bodies, the results are presented according to the corresponding Reynolds number.

\[ C_D = \frac{4}{3} \frac{\Delta \rho g D_g}{\rho_L \Delta u^2} \quad \text{Drag coefficient for ideal spheres} \]

\[ \text{Re} = \frac{\rho_L D_g \Delta u}{\mu_L} \quad \text{Reynolds number for bubble} \]

Figure 4.7 presents the determined drag coefficients with respect to the Reynolds number for disperse bubbles rising in highly viscous corn syrup and water, and a comparison of the results with drag correlations selected to represent solid spheres (Schiller and Naumann, 1935)\textsuperscript{16} and bubbles in non-pure water (Lain et al., 1999)\textsuperscript{17}. In the high-viscosity case (Fig 4.7.a), the drag determined for the smallest bubbles at lowest Reynolds numbers \((10^{-7} \text{ to } 10^{-5})\) showed good agreement with the Schiller and Naumann (1935) correlation. On the other hand, the drag coefficients obtained for larger bubbles tend to exceed the values predicted with the Schiller and Naumann (1935) correlation, even when the Reynolds number remained low \((0.01-1.0)\). The higher drag obtained for these large bubbles is explained by the increasingly important effect of the wall friction as the size of the bubble approaches the internal diameter of the channel.

\textsuperscript{16} Schiller and Naumann (1935) drag correlation: \( C_D = \frac{24}{\text{Re} \left(1 + 0.1 \ast \text{Re}^{0.687}\right)} \)

\textsuperscript{17} Lain et al. (1999) drag correlation: \( C_D = \frac{24}{\text{Re} \left(1 + 0.15 \ast \text{Re}^{0.687}\right)} \quad \text{Re} \leq 500 \)

\( C_D = 9.5 \ast 10^{-5} \ast \text{Re}^{1.397} \quad 500 < \text{Re} < 1500 \)

\( C_D = 2.61 \quad 1500 \leq \text{Re} \)
Fig 4.7 Drag coefficients vs. Reynolds number determined for bubbles at low void fraction in [a-top] highly viscous corn syrup, and in [b-bottom] tap water, showing reasonable agreement with the drag correlations proposed by Schiller and Naumann (1935) for solid spheres and Lain et al. (1999) for bubbles in non-pure systems respectively.

For the low-viscosity case (water) a significant dispersion in the data is noticeable, although it can be said that the measurements are consistent with the correlation representing the rise of bubbles in non-pure media (Lain et al., 1999). The scatter in the plot of drag coefficient observed in Figure 4.7.b is attributed to the distorted geometry of the bubbles,
which makes it difficult to obtain an accurate assessment of the size and displacement of the center of mass from the 2-D images captured with the high-speed camera.

4.2.3 Drag Reduction due to Wake Interaction

Wake effects are known to increase the velocity of trailing particles, either gas or solids. This effect has been observed in the experimental facility during the tests at low-void fraction, where small bubbles trapped in the wake of larger leading bubbles have been observed to travel with practically the same rise velocity as illustrated in Figure 4.8. Krishna et al. (1999) suggested that in the churn-turbulent regime the bubbles rise faster than if they were alone due to wake interactions. Large bubbles would act as leading particles, and a part of the population of small bubbles would constitute instantaneously the trailer particles, resulting in a system where the gas phase rises faster than for isolated bubbles, and therefore the overall drag interaction is reduced.

![Images illustrating how the wake interaction causes a small bubble to travel together with a larger leading bubble. It is also observed that small bubbles adopt an approximately spherical geometry, while larger bubbles tend to assume distorted shapes. In the two bubbles continuous deformations are evident.](image-url)
A comparable behavior has been observed for Taylor bubbles as illustrated in Figure 4.9, where it can be appreciated that the leading bubble travels slower, making it possible for the trailing bubble to reach it and merge, forming a larger slug. Moissis and Griffith (1962) observed that there is a critical separation of about six tube diameters between two consecutive Taylor bubbles, and if the distance is lower the wake interaction will cause the trailing bubble to travel faster than the leading bubble and merge. This interaction has large implications, making the fully developed slug flow nearly impossible in practice and difficult to study numerically (Levy, 1999).

Fig 4.9 Images illustrating how the wake interaction produces a trailer bubble to accelerate and merge with the leading Taylor bubble ($U_{\text{LEADING}} \sim 0.37 \text{ m/s}$)
The results presented in Figures 4.8 and 4.9 illustrate how wake interactions can reduce the drag experienced by trailing bubbles. The effect of turbulence on multiphase flows is not well understood yet, but it is believed that bubbles tend to collect in regions of low pressure, such as wakes and centers of vortices (Brennen, 2005), and it is apparent from the discussed result that the wakes may play a significant role in high void fraction flows reducing the drag interaction. The behavior discussed is opposite to the increase in the drag coefficient with void fraction that has been reported for the bubbly regime by several researchers. The interfacial momentum exchange in gas-liquid systems is affected by numerous parameters not fully understood, including mutual interactions between bubbles.

### 4.3 HIGH-VOID FRACTION VERTICAL CHANNEL FLOW

When higher gas flow rates were injected at the bottom of the riser using the sparger, a more complicated situation was observed, where the flow regime obtained and circulation produced varied significantly according to the viscosity of the liquid and gas flux.

#### 4.3.1 Regimes

In the runs where highly viscous corn syrup was used it was noticed that the slug regime tends to develop and dominate under the entire range of conditions evaluated. Figure 4.10 illustrates the flow patterns observed in the experimental facility with highly viscous corn syrup at different gas fluxes. In general, ‘chains of bubbles’ or ‘slug trains’ were developed with high viscosities, where the size of the liquid slugs tends to reduce as the gas flow is higher, while the Taylor bubbles become larger.

The slug flow observed with high viscosity media was noticeably unstable, where trailing bubbles merged with their leading bubbles, forming larger Taylor bubbles that can travel faster and reach easily the subsequent leading bubble. Because of this gas merging the
resulting bubbles at the upper part of the tube are larger than the bubbles originally produced by the sparger. A liquid film flowing downwards between the Taylor bubbles and the tube wall was permanently observed.

Fig 4.10 Images illustrating slug trains observed with highly viscous liquid ($\mu_L \sim 10.1$ Pa.s) and how the length of the Taylor bubbles increases and the length of the liquid slugs reduces as the gas flux is higher.
For the high-viscosity runs, the large Taylor bubbles formed at high gas rates were accompanied by strong fluctuations and vibrations of the test rig. This represented a limit to the maximum gas flux rate. The severity of the fluctuations was lower as the corn syrup was diluted and its viscosity reduced. The characteristics of the flow observed in the test rig suggest that flow oscillations can potentially occur within the calandria tubes of sugar evaporative crystallizers, and they are likely to be more severe as the viscosity is higher.

\[ J_G = 1.9 \text{ cm/s} \quad \text{Bubbly} \]

\[ J_G = 5.9 \text{ cm/s} \quad \text{Churn turbulent} \]

\[ J_G = 24.5 \text{ cm/s} \quad \text{Slug} \]

\[ J_G = 156 \text{ cm/s} \quad \text{Churn} \]

Fig 4.11 Images illustrating disordered flow observed with low-viscosity media (0.05 Pa.s)

It was observed that the coalescence and breakup mechanisms become more important as the viscosity of the liquid media is lowered. With low-viscosity media the flow in the facility remained within the bubbly regime only at very low gas rates; as the airflow
rate was increased, the bubbles tend to merge at the center of the riser and form larger and
distorted gas particles (caps) as the one presented in Figure 4.5.b, experiencing rapidly
transition to the churn-turbulent, slug and churn regimes. Figure 4.11 presents images
obtained with a low viscosity medium and different gas flux rates, showing high dispersion
in the size and shape of the bubbles difficult to characterize in detail. For this reason the
internal diameter of the tubes has been often assumed as characteristic dimension in the
analysis of high-void fraction channel flows.

4.3.2 Frictional Pressure Drop

An important parameter to be considered is the pressure drop due to the frictional
shear stress at the wall, since it affects the void fraction and drag coefficient computations.
For the calculation of wall shear stress in gas-liquid channel flows it is normally required to
use empirical models due to the complexity associated with multiphase flows and the
differences between multiple possible patterns. For the analysis of the wall frictional
resistance in the test rig used to study the momentum interaction in vertical gas-liquid flows,
several correlations reported in the literature have been considered, comparing the results
with corresponding numerical predictions (CFD) of the friction at the wall. Following are
briefly described the two models that showed better agreement with the CFD simulations.

- Lockhart-Martinelli correlation (1949)

This assumes that a gas-liquid flow comprises two co-current separated streams, with
equal pressure drop, which can be calculated from the pressure drop that each stream would
have as if they were flowing alone in the channel. The calculations begin using conventional
models to compute wall frictional losses of single-phase flows, applied for each phase as if it
was flowing alone:
\[
\frac{dP}{dx}_{fr} = C_{fr} \cdot \frac{1}{D_H} \left( \frac{1}{2} \cdot \rho \cdot U^2 \right)
\]
\[
\Delta P_{fr} = C_{fr} \cdot \frac{4 \cdot L}{D_H} \left( \frac{1}{2} \cdot \rho \cdot U^2 \right)
\]

Where: \( D_H = \frac{4 \cdot A}{P} \)

\[
Re = \frac{\rho \cdot D_H \cdot U}{\mu}
\]

For smooth channels the analytical approximations developed by Hagen-Poiseuille (1840) and Altshul (1975) for the friction pressure loss coefficient can be applied:

\[
C_{fr} = \frac{64}{Re} \quad \text{Re} < 2300 \quad \text{Laminar.}
\]
\[
C_{fr} = 0.028 \left( \frac{Re}{2300} \right)^{2.667} \quad 2300 \leq Re < 2818 \quad \text{Transition}
\]
\[
C_{fr} = \left( 1.8 \log Re - 1.644 \right)^{-2} \quad 2818 \leq Re \quad \text{Turbulent}
\]

Once the friction per each phase is known, the Martinelli parameter can be computed as the ratio between the liquid and gas pressure drop, and a pressure multiplier can be obtained from empirical functions to compute the pressure drop in the multiphase flow. Table 4.2 presents the main equations involved in the Lockhart-Martinelli model. The constant ‘c’ depends upon the regimes determined from the calculated Reynolds numbers for each phase:

- Liquid laminar – gas laminar: \( c = 5 \)
- Liquid laminar – gas turbulent: \( c = 12 \)
- Liquid turbulent – laminar gas: \( c = 10 \)
- Liquid turbulent – gas turbulent: \( c = 20 \)

The Lockhart-Martinelli model was developed from air-water tests near atmospheric pressure. It is known to be applicable satisfactorily to steam-water flows at low pressure (eg. 14.7 psi), but at high pressure tends to give underestimated values of frictional pressure drop.
Table 4.2  Lockhart-Martinelli multipliers for calculation of frictional pressure drop.

<table>
<thead>
<tr>
<th></th>
<th><strong>Liquid phase</strong></th>
<th><strong>Gas phase</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td>Reynolds</td>
<td>$Re_L = \frac{\rho_L D_H J_L}{\mu_L}$</td>
<td>$Re_G = \frac{\rho_G D_H J_G}{\mu_G}$</td>
</tr>
<tr>
<td>Pressure drop</td>
<td>$\left(\frac{dP}{dx}\right)<em>{f-r-L} = C</em>{f-r-L} \cdot \frac{1}{2} \rho_L J_L^2$</td>
<td>$\left(\frac{dP}{dx}\right)<em>{f-r-L} = C</em>{f-r-G} \cdot \frac{1}{2} \rho_G J_G^2$</td>
</tr>
<tr>
<td>Phase multiplier</td>
<td>$\Phi_L^2 = \frac{\left(\frac{dP}{dx}\right)<em>{f-r-TP}}{\left(\frac{dP}{dx}\right)</em>{f-r-L}}$</td>
<td>$\Phi_G^2 = \frac{\left(\frac{dP}{dx}\right)<em>{f-r-TP}}{\left(\frac{dP}{dx}\right)</em>{f-r-G}}$</td>
</tr>
<tr>
<td>Martinelli factor</td>
<td>$X_{LM} = \frac{\Phi_G^2}{\Phi_L^2} = \frac{\left(\frac{dP}{dx}\right)<em>{f-r-L}}{\left(\frac{dP}{dx}\right)</em>{f-r-G}}$</td>
<td>$X_{LM} = \frac{\Phi_G^2}{\Phi_L^2} = \frac{\left(\frac{dP}{dx}\right)<em>{f-r-L}}{\left(\frac{dP}{dx}\right)</em>{f-r-G}}$</td>
</tr>
<tr>
<td>Phase multiplier</td>
<td>$\Phi_L^2 = 1 + \frac{c}{X_{LM}} + \frac{1}{X_{LM}^2}$</td>
<td>$\Phi_L^2 = 1 + X_{LM} + X_{LM}^2$</td>
</tr>
</tbody>
</table>

- Martinelli-Nelson correlation (1964)

This is an extension of the Lockhart-Martinelli model to account for the phase change effect in steam-water flows, particularly at high pressures. An additional multiplier corresponding to an only liquid flow (LO) with the total mass flux of the two-phases is introduced to replace the Martinelli factor.

$Re_{MN} = \frac{D_H \cdot G}{\mu_L}$  

Reynolds number, liquid only flow

$\left(\frac{dP}{dx}\right)_{f-r-LO} = C_{f-r-L} \cdot \frac{1}{D_H} \cdot \frac{1}{2} \rho_L \cdot \frac{G^2}{\rho_L}$. 

Pressure drop for liquid only flow

$\Phi_{LO}^2 = \frac{\left(\frac{dP}{dx}\right)_{f-r-TP}}{\left(\frac{dP}{dx}\right)_{f-r-LO}}$. 

Martinelli-Nelson multiplier
More accurate expressions have been developed recently by comparison with experimental data. For the case of $\mu_L/\mu_G \geq 1000$ (e.g. $\mu_{\text{MASSECUTE}}/\mu_{\text{VAPOR}} \sim 2 \text{ Pa.s}/1.1 \times 10^{-5}$ Pa.s = 200 000) the following correlations are recommended by Kolev (2002).

- Martinelli-Nelson: Analytical solution for the Blasius regime, use for $G \leq 100 \text{ kg / s.m}^2$

$$\Phi_{LO}^2 = (1 - x)^{7.5} \Phi_{L}^2$$

Where: $x =$ gas mass fraction or quality.

- Baroczy (1965) correlation with Chisholm (1983) modification, use for $G > 100 \text{ kg / s.m}^2$

$$\Phi_{LO}^2 = 1 + \left( Y^2 - 1 \right) \left( B \cdot \left[ x \cdot (1 - x) \right]^{0.875} + x^{1.75} \right)$$

$$Y = \sqrt{\frac{\rho_L}{\rho_G} \left( \frac{\mu_G}{\mu_L} \right)^{0.25}}$$

Where $B$ is defined according to $G$ and $Y$:

<table>
<thead>
<tr>
<th>$G \leq 500$</th>
<th>$Y \leq 9.5$</th>
<th>$G \leq 600$</th>
<th>$9.5 \leq Y \leq 28$</th>
<th>$28 \leq Y$</th>
</tr>
</thead>
<tbody>
<tr>
<td>$500 \leq G \leq 1900$</td>
<td>$B = 4.8$</td>
<td>$B = 520/(Y \cdot G^{0.5})$</td>
<td>$B = 2400/G$</td>
<td>$600 &lt; G$</td>
</tr>
<tr>
<td>$1900 \leq G$</td>
<td>$B = 55 \cdot G^{-0.5}$</td>
<td>$B = 15000/(Y^2 \cdot G^{0.5})$</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Figure 4.12 presents the frictional pressure drop calculated for the gas-liquid flow in the riser of the test rig with the correlations developed by Lockhart-Martinelli (1949), and Martinelli-Nelson (1964) combined with Baroczy (1965). These friction correlations are compared with the friction predicted numerically with the Eulerian-Eulerian CFD model for corresponding two-phase flows. Considering the consistency of the Lockhart-Martinelli results with the numerical simulations, this model was chosen for the calculation of the frictional pressure drop in the analysis of the experimental results. The Lockhart-Martinelli is known to be a sturdy correlation, presented repeatedly in multiphase literature and frequently applied for the analysis of practical flows.
During the analysis of the tests corresponding to low viscosity it was noticed that the effect of friction is small, particularly at low gas flow rates, where the pressure field is determined essentially by the void fraction. This seems to be a normal behavior, which makes it possible to neglect the frictional component in the analysis of some multiphase flows, as for example is often the case in vertical bubbly and slug flows (Levy, 1999). On the other hand, as the viscosity increases it was noticed that the contribution of the friction becomes significant, and its effect has to be considered in order to obtain meaningful results.

### 4.3.3 Circulation

During the tests involving highly viscous liquids it was observed that the circulation displayed a pulsating behavior, exhibiting unsteadiness with some periodicity. Figure 4.13 presents signals of the circulation recorded with the Coriolis meter for high-viscosity corn syrup and water with a similar gas flow rate ($J_G \sim 0.5$ m/s). For the corn syrup case a lower
circulation rate is observed, which is explained by the increase in the frictional resistance due to higher viscosity, this even when the void fraction was higher, and therefore more pressure driving force was available. Significant oscillations in circulation were recorded in both cases, observing that the relative amplitude of fluctuation is larger for the high-viscosity medium. The slug regime, normally encountered at high void fractions \((\alpha \geq 0.25)\), is intermittent and unsteady by nature, and can be described as a kinematic or concentration wave (Brennen, 2005), explaining the origin of the oscillations observed during the tests at high viscosity.

The circulation is the clearest indicator of the exchange of momentum between the gas and liquid phases in the test rig. Figure 4.14 presents the circulation determined in the rig in terms of the liquid superficial velocity in the riser \((J_L)\) with respect to the air flux rate or superficial gas velocity \((J_G)\). In general, it is observed that as the gas flux increases the

![Graph showing liquid circulation](image)

**Corn syrup -** \(\mu_L \approx 1.91 \text{ Pa.s}\)  
\(J_G \approx 0.56 \text{ m/s}, \alpha \approx 81\%\)  

**Water -** \(\mu_L \approx 0.0008 \text{ Pa.s}\)  
\(J_G \approx 0.51 \text{ m/s}, \alpha \approx 36\%\)  

Fig 4.13  Liquid circulation determined in the experimental facility with [a-left] high-viscosity and [b-right] low-viscosity media

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circulation induced is higher, but it displays an asymptotic behavior. This result suggests that the capacity of the gas to transfer momentum to the liquid phase is lost progressively and/or the increase in frictional resistance is severe. As a consequence the gain in circulation tends to be lower as the gas flux increases.

Fig 4.14 Circulation determined in the test rig vs. gas flux rate, showing an asymptotic behavior of the circulation as the gas flow becomes higher
The asymptotic behavior of the circulation with respect to the gas flow observed during the experiments is in agreement with data reported from the study of air-lift reactors (Merchuk and Stein, 1981; Yun and Shen, 2003), gas-lifted oil wells (Mudde, 2005), and natural circulation boiling tubes (Subki et al., 2005). This result points to the existence of an optimal maximum length for the calandria tubes of vacuum pans, since the transfer of momentum responsible for the circulation would be progressively reduced as the tubes are longer and more vapor is generated.

The decay in the transfer of momentum with increase of gas flux can be associated with the wake effect previously discussed (Figs 4.8–4.9). This behavior is consistent with drag correlations proposed by Ishii and Chawla\textsuperscript{18} (1979) for the slug regime and Ishii and Zuber\textsuperscript{19} (1979) for the churn-turbulent regime, where significant reductions in drag are predicted as the void fraction increases.

### 4.3.4 Void Fraction

Figure 4.15 presents the circulation and void fraction determined with water at different openings of the valve. As expected, a reduction in circulation as the valve is closed occurs, indicating that an increase in frictional resistance of the loop hinders circulation. Correspondingly, in a sugar crystallizer using a smaller downtake or reducing the separation between the calandria plates and the bottom increases the frictional resistance and will reduce the circulation.

\textsuperscript{18} For slug flow, from Ishii and Chawla (1979):

\[ C_D = 9.8 \times (1 - \alpha)^3 \]

\textsuperscript{19} For churn-turbulent, from Ishii and Zuber (1979):

\[ C_D = \frac{8}{3} (1 - \alpha)^2 \]
Fig 4.15 Effect of the gas flux on the [a-left] induced circulation and [b-right] void fraction determined in the experimental facility with tap water, displaying asymptotic behaviors.

The experimental results showed that the void fraction displays a similar trend as the circulation, and shows a steep increase at low gas rates followed by an asymptotic behavior. The void fraction determined is compared with data reported by Jia et al. (1984) from the study of air-water flow in a vertical channel, and displays similarity. The void fraction tends to reach higher values as the valve opening is reduced, this in agreement with observations in the lab-scale test rig, and with algebraic multiphase models that predict a higher void fraction as the superficial velocity of the liquid phase decreases.

Figure 4.16 presents the void fraction determined with different liquid media conditions, illustrating the discussed asymptotic behavior as the gas flux increases. The void fraction tends to reach higher values as the liquid viscosity is higher, this as a consequence of the reduction in the gas rise velocity.
4.3.5 Drag Coefficient – A New Correlation

To examine the experimental information from a more fundamental perspective the drag coefficient has been evaluated assuming a 1-D averaged situation representing the flow in the entire cross section. While the traditional plot of drag vs. Reynolds number was used previously with reasonable success for the analysis of dispersed bubbles at low void fraction, it proved to be incapable to provide a distinguishable trend when dealing with the more complex flows associated with high void fractions. Identifying the drag forces is relatively easy for dilute systems and has been covered in numerous studies, but as the void fraction and the wall effect become larger, the problem becomes progressively more complex (Ishii and Hibiki, 2006) and the information available is scarce.

The drag correlations normally used to obtain closure of the flow transport equations when applying the Eulerian-Eulerian CFD approach have been developed originally for...
single solid spheres or small bubbles in stagnant media, usually water, or within the bubbly regime at low void fraction. Consequently, the use of these drag models is formally restricted for simple low-void fraction bubbly flows, since the correlations cannot account for effects such as mutual inter-particle interactions, wakes, and regime transitions, which would be openly neglected when dealing with complex multiphase flows. No established correlations exist for the more complex regimes that occur at high void fraction, such as the slug regime, which according to the visual observations illustrated in Figures 4.10 and 4.11 is likely to develop in vertical channels with high-viscosity media.

In gas-liquid buoyancy-driven flows, the momentum transferred from the gas to the liquid phase is responsible for overcoming the frictional resistance, and as a result the drag forces exerted on the bubbles tend to be higher than in stagnant flows. Levy (1999) presents an analysis where the wall frictional force ($f_w$) is introduced in the two-phase momentum equations, leading to an expression for the drag coefficient of gas-liquid channel flows that takes into account the effect of the void fraction and the friction at the wall. For the analysis of the experimental data the expression presented by Levy (1999) for calculation of the drag coefficient in multi-particle systems is applied, showing that the drag interaction is lower as the gas content is higher, and increases with friction.

$$C^M_D = \frac{4}{3} \rho_l \times \Delta u^2 \left\{ \Delta \rho \cdot g \cdot (1 - \alpha) + f_w \right\}$$

Levy (1999)

$$f_w = \left. \frac{\Delta P}{dx} \right|_{\text{FRICCTION}} = \frac{4}{D_h} \tau_w$$

$$f_w = \text{wall frictional force.}$$

Based on correlations proposed for high-void fraction regimes by Wallis (1969) and Ishii and Hibiki (2006), two terms including the inverse viscosity dimensionless number ($N_\mu$)
and the void fraction ($\alpha$) were introduced in the analysis, affected by exponents ($a,b,c$) and a constant ($k$) that were determined by iteration until a good correlation in the log-log plane was obtained. Figure 4.17 presents the multi-particle drag coefficient determined in the experimental facility (TR2) in terms of two auxiliary functions $f(N_\mu)$ and $g(\alpha)$ with respect to the Reynolds number, illustrating a distinguishable trend.

$$f(N_\mu) = N_\mu$$
$$g(\alpha) = (1 - \alpha)$$

$$C_D^M = f^a g^b \frac{k}{Re^c}$$

Fig 4.17 Trend of the drag coefficient vs. Reynolds number determined from measurements in the experimental facility (TR2). The identified trend indicates a decrease in the momentum interaction as the void fraction and the Reynolds number are higher.

The correlation obtained has shown reasonable agreement with results obtained during the experiments conducted in the lab-scale unit described in chapter three (TR1), and
also with experimental data reported by Jia et al. (1984) and Merchuk and Stein (1981) from comparable experiments, suggesting that the identified relationship could be applied with confidence in the analysis of adiabatic gas-liquid vertical flows at low liquid flux within a wide range of viscosities ($10^{-3}$ to $10^2$ Pa.s). This correlation appears useful for the analysis of natural circulation systems, particularly at high-void fractions ($\alpha > 0.2$) and low-Reynolds numbers, as expected in calandria tubes due to the elevated viscosity of the massecuite.

The developed drag correlation shows a decrease in the drag interaction with void fraction, which reflects the already discussed asymptotic behavior of the natural circulation as the gas flux increases (Figures 4.14 - 4.15). The observed trend is opposite to the increase in the drag within the bubbly regime with void fraction that has been reported by several researchers for bubbly flows (e.g. Rusche and Issa, 2000; Behzadi et al., 2004), but agrees with results that suggest that within the churn-turbulent and slug regimes the drag decreases with increasing void fraction (Ishii and Chawla, 1979; Ishii and Zuber, 1979), this probably as a consequence of the wake interaction between large and small bubbles (as illustrated in Figure 4.8), which would produce an easy way for the gas phase to rise faster. Therefore, the interfacial momentum interaction in dilute systems increases with void fraction up to certain point, probably where departure from the bubbly regime occurs. Then, within the regimes associated with high-void fraction, the effect is opposite and the drag interaction is progressively reduced as the gas content increases.

4.4 EXCHANGE OF MOMENTUM IN VERTICAL BOILING TUBES

The analysis of the flow in boiling vertical channels is highly complicated due to the wide range of conditions and regimes that can be developed, which affect dramatically the different mass, momentum, and energy interactions between the participating phases. Most
studies of flow boiling have been dedicated to the fundamental water-steam flow, while not much information exists on boiling of highly viscous liquids, as in the case of massecuites boiling in calandria tubes.

The present study makes use of the momentum interaction correlation obtained from the tests in the experimental facility, under adiabatic conditions, for the analysis of the flow in the non-adiabatic calandria tubes of sugar evaporative crystallizers. To adjust the numerical model for the more complex conditions observed when boiling takes place, the predictions have been compared with experimental data presented by Rouillard (1985), who measured the pressure and void fraction along a forced-circulation boiling tube facility designed to represent the process in calandria tubes. The results obtained are discussed and compared with experimental information reported by Austmeyer and Schliephake (1983), Rouillard (1985), and Bruhns (1996) from tests performed in single-tube facilities, observing that the interfacial momentum interaction has to be reduced with respect to corresponding adiabatic conditions in order to obtain agreement with the experimental data. This reduction in the momentum interaction is attributed here to the increase in frictional pressure drop caused by vapor bubbles generated near to the walls and the development of flow boiling instability.

4.4.1 Comparison with Experimental Data Reported by Rouillard

Rouillard (1985) performed a series of tests within a wide range of conditions using syrup, molasses, and C-massecuite seed as liquid media. It is considered here that the viscosity of the molasses (0.2-3.9 Pa.s) and syrup (0.03-0.08 Pa.s) were too low, and consequently the heat transfer too high (~ 900-1300 W/m².K), with respect to the actual conditions in sugar evaporative-crystallizers. On the contrary, the C-massecuite seed is
closer to real operational conditions (1-13 Pa.s; 200-500 W/m².K), and therefore only those runs using this medium are considered for comparison in this study.

The information presented by Rouillard (1985) from the tests using C-massecuite seed permits identifying two types of characteristic void fraction profiles, which are distinguishable because one type shows a steep increase at the upper section of the tube, as presented in Figure 4.18, while the other type illustrates a flat profile at the upper section of the tube as shown in Figure 4.19.

Table 4.3 presents the operational conditions for the runs considered classified according to the type of void fraction profile and the liquid viscosity, where the following common characteristics can be observed for each case:

- **Steep void fraction at upper section (Fig 4.18):** This profile is found for a wide range of viscosities (1-11 Pa.s) at high heat transfer rates (15-37 kg/h.m²; 300-530 W/m².K), probably representing what occurs in the calandria tubes at the beginning of a boiling for batch sugar evaporative crystallizers, or in the first cells for continuous sugar crystallizers. The profile suggests that most of the evaporation takes place at the top of the tube.

- **Flat void fraction at upper section (Fig 4.19):** This profile is observed at relatively high liquid viscosity (~ 7-13 Pa.s) and low heat transfer (8-15 kg/h.m²; 165-290 W/m².K), probably representing what occurs in the calandria tubes towards the end of a boiling for batch sugar evaporative crystallizers, or in the lasts cells for continuous sugar crystallizers. The profile suggests that a maximum evaporation is reached and no significant vaporization takes place at the top of the tube.
Fig 4.18 Void fraction profiles measured along the experimental tube by Rouillard (1985) displaying a steep increase in the upper section, suggesting that significant evaporation would occur at the top of the calandria tubes. Data corresponding to C-massecuite seed at relatively ‘high’ heat transfer; 1-11 Pa.s; 15-37 kg/h.m²; 300-530 W/m².K
Fig 4.19 Void fraction profiles measured along the experimental tube by Rouillard (1985) displaying flat behavior in the upper section, which suggests that no significant evaporation would occur at the top of the calandria tubes. Data corresponding to C-seed massecuite at relatively ‘low’ heat transfer; 7-13 Pa.s; 8-15 kg/h.m²; 165-290 W/m².K
Table 4.3 Operation parameters during Rouillard (1985) tests with C-massecuite seed.

<table>
<thead>
<tr>
<th>Run</th>
<th>( \mu_L ) (Pa.s)</th>
<th>( J_L ) (m/s)</th>
<th>Evaporation (kg/h.m²)</th>
<th>( \Delta T ) (°C)</th>
<th>( h ) (W/m².K)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>(High)</td>
<td>(High)</td>
<td></td>
<td>(High)</td>
<td>(High)</td>
</tr>
<tr>
<td>I - Steep top</td>
<td>10 11.4 0.063 17.4 37.7 299</td>
<td>11 11.3 0.046 23.5 39.4 386</td>
<td>12 7.7 0.079 19.1 29.7 413</td>
<td>39 4.6 0.046 32.7 42.7 497</td>
<td>40 4.9 0.071 36.8 45.5 527</td>
</tr>
<tr>
<td></td>
<td>(Low)</td>
<td>(High)</td>
<td></td>
<td>(Low)</td>
<td>(Low)</td>
</tr>
<tr>
<td></td>
<td>43 1.1 0.046 16.3 31.0 335</td>
<td>44 0.9 0.071 14.8 28.4 332</td>
<td>45 2.1 0.096 28.0 36.9 489</td>
<td>46 2.1 0.046 19.3 38.9 320</td>
<td>47 1.9 0.071 25.2 39.9 408</td>
</tr>
<tr>
<td>II - Flat top</td>
<td>9 11.4 0.079 11.9 38.4 201</td>
<td>13 7.3 0.063 11.5 32.6 226</td>
<td>14 7.6 0.079 12.1 32.7 238</td>
<td>15 7.4 0.063 12.9 33.2 250</td>
<td>16 7.5 0.046 15.2 33.3 292</td>
</tr>
</tbody>
</table>

- **Evaporation Profile**

  Using the information reported by Rouillard (1985) on condensation for each C-massecuite seed run and its respective void fraction profile, the evaporation occurring along the tube has been estimated. Figure 4.20 presents the evaporation profiles normalized with respect to the maximum evaporation. The results suggest that at high heat transfer rates, corresponding to steep void fraction profiles at the top, most of the evaporation takes place in the upper section of the tubes. On the other hand, at low heat transfer rates a maximum evaporation would be achieved at \( L \sim 0.6-0.9 \) m.
Fig 4.20 Evaporation profile along a boiling heated tube estimated from Rouillard (1985) experimental data illustrating that [a-top] most of the evaporation occurs in the upper section at high heat transfer rates ($\sim$ 300-500 W/m$^2$.K), while [b-bottom] a maximum evaporation around $L \sim 0.6-0.9$ m is obtained at low heat transfer ($\sim$ 150-300 W/m$^2$.K)
Particularly interesting in Figure 4.20 is the presence of a maximum in the estimated local evaporation around $L \sim 0.6-0.9\ m$ at low heat transfer rates, which could have important implications in the design of the length of calandria tubes. In practice it has been found that crystallizers provided with calandria tubes within this range display higher heat transfer coefficients than with longer tubes.

Averaged evaporation profiles have been identified using statistics, which are displayed with thick curves in Figure 4.20. The expressions that showed best agreement with the estimated evaporation profiles were:

- High heat transfer, most evaporation occurring at the top (Fig 4.20.a)
\[
Evaporation(\hat{x}) = \left( \frac{\text{Total \ evaporation}}{0.15146} \right) \times \left( \frac{1}{94.6780 - 187.78^*\hat{x} + 94.135^*\hat{x}^2} \right)
\]
\[
\hat{x} = x/L
\]

- Low heat transfer, maximum evaporation occurring at $L \sim 0.6-0.9\ m$. (Fig 4.20.b)
\[
f(L) = \left[ 1 + \exp(7.3255 - 8.4524^*L) \right]^{0.4226}
\]
\[
Evaporation(\hat{x}) = \left( \frac{\text{Total \ evaporation}}{f(L)} \right) \times \left( \frac{0.2011 - 0.1357^*\hat{x}}{1 - 2.1935^*\hat{x} + 1.3321^*\hat{x}^2} \right)
\]
\[
\hat{x} = x/L
\]
\[
L \leq 1.5m
\]

4.4.2 Numerical Simulation of the Flow in Boiling Vertical Tubes

The evaporation profiles identified from Rouillard (1985) experimental data have been used to simulate the flow in calandria tubes using the Eulerian-Eulerian multiphase model of the CFD commercial code Fluent. The energy interaction is approximated using a mass source function (UDF DEFINE_SOURCE) to inject vapor in the computational domain representing the calandria tubes. The UDF is programmed based on the evaporation profiles identified, which are evaluated locally at each cell according to the axial position along the
tube. For the momentum interaction, the (adiabatic) correlation developed from the experiments in the test rig is applied, multiplied by a factor that was adjusted iteratively until reasonable agreement with the (non-adiabatic) pressure measurements reported by Rouillard was obtained. Figure 4.21 presents the grid and boundary conditions used for the numerical analysis, and some results illustrating a pressure field largely determined by the hydrostatic effect, and a void fraction field indicating that most of the vaporization occurs at the top section of the tube.

![Schematic illustrating grid and boundary conditions used to simulate the flow in Rouillard’s test rig, and the predicted contours of absolute pressure and void fraction](image)

Fig 4.21 Schematic illustrating [a-top] the grid and boundary conditions used to simulate the flow in Rouillard’s test rig, and the predicted contours of [b-middle] absolute pressure and [c-bottom] void fraction

[Data corresponding to C-massecuite seed run No. 10]
Figure 4.22.a presents numerical results obtained assuming uniform evaporation along the heated tubes, which are compared with corresponding void fraction and pressure measurements reported by Rouillard (1985 – Run No. 10). The Schiller and Naumann (1935) drag correlation and the drag model developed in this study (Figure 4.17) are applied for the computations, observing poor agreement in both cases from the qualitative and quantitative points of view. The difference between computations and measurements indicates that considering a uniform evaporation in the calandria tubes would be a poor assumption, leading to overprediction of the void fraction, pressure driving force, and consequently of the circulation and massecuite velocity.

Figure 4.22.b presents numerical results obtained using a UDF programmed with the evaporation profile identified from Rouillard (1985) experimental data to specify the generation of vapor along the heated tube. Improved agreement between the numerical solution and the measurements is evident with respect to the previous solution assuming uniform evaporation. The use of the drag correlation developed in this study (Figure 4.17) to model the exchange of momentum between the gas (vapor) and liquid phases resulted in a prediction that shows reasonable agreement with the void fraction and pressure reported by Rouillard (1985), although some differences from the quantitative point of view are obtained.

The numerical results presented in Figure 4.22 indicate that specifying correctly the generation of vapor along the calandria tubes is a critical step to obtain a valid prediction of the two-phase flow studied. Using a realistic evaporation profile to represent the boiling in numerical flow simulations is important for a correct representation of the buoyancy driving forces in the calandria tubes, and therefore of the natural circulation of massecuite in sugar evaporative crystallizers.
Fig 4.22 Void fraction and absolute pressure along a heated boiling tube measured by Rouillard (1985), and computed using CFD assuming [a-top] uniform vapor generation and [b-bottom] an evaporation profile along the calandria tube.

Important reductions in the drag interaction with respect to corresponding adiabatic conditions have been required to obtain agreement with the experimental data presented by
Rouillard (1985), suggesting that the transfer of momentum in the heated boiling channels studied is lower than for corresponding continuous adiabatic gas-liquid flows. Figure 4.23 presents a comparison of numerical results with data reported by Rouillard (1985) for two cases where reasonable agreement was obtained, illustrating the two types of evaporation and void fraction profiles that have been identified.

Fig 4.23 Comparison of the void fraction and absolute pressure along a heated boiling tube measured by Rouillard (1985) and computed using CFD, showing reasonable agreement for two cases displaying [a-top, run 10] a steep void fraction profile at the top, and [b-bottom, run 16] a flat void fraction profile in the upper part of the tube.
Figure 4.24 presents the adjustment factor that has been applied to the adiabatic drag correlation developed from tests in the experimental facility in order to attain agreement with the non-adiabatic experimental data reported by Rouillard (1985). The drag correction tends to be more significant as the heat transfer increases and the system moves farther from the adiabatic case, where ideally no correction would be necessary.

Based on the result displayed in Figure 4.24 it can be reasoned that the momentum interaction in certain gas-liquid flows such as calandria tubes is considerably affected by boiling, which seems to reduce the quantity of momentum being transferred from the gas to the liquid phase with respect to corresponding adiabatic conditions. It can also be observed in Figure 4.24 that the required adjustment of the drag tends to be greater as the viscosity is lower. This could be related to the turbulent nature of boiling and the consequent development of fluctuations and shear stresses near to the wall, which would be logically more significant in a low-viscosity medium.
4.4.3 Pressure and Evaporation along a Calandria Tube

The difference between the hydrostatic and static pressures presented in Figure 4.23 indicates that a significant part of the circulation driving force is generated in the upper section of the tube, where the gas content is logically higher. The same trend is also observed in experimental data reported by Austmeyer and Schliephake (1983) under natural circulation, showing a departure from the hydrostatic pressure profile after \( L \sim 0.8-0.9 \) m from the bottom of the heating tube, which was attributed to the presence of vapor bubbles exclusively in the upper section. Although this interpretation is largely true, it ignores the effect of the frictional pressure drop, which is significant for large viscosity media such as massecuite. On the other hand, the void fraction measured by Rouillard (1985) and experiments performed by Skyring and Beale (1967) evidenced the presence of vapor bubbles not far from the entrance of the calandria tubes, which according to the numerical simulations cause only a small difference between the hydrostatic and static pressures at the lower section of the tube (see Figure 4.23), and therefore are likely to be difficult to recognize from pressure readings under unsteady boiling conditions.

The static pressure measured by Austmeyer and Schliephake (1983) along the radial direction displays minimum variations, this in agreement with the numerical results presented in Figure 4.21.b

4.4.4 Flow Regime and Frictional Pressure Drop in Calandria Tubes

Based on the tests conducted in the experimental facility that showed that the slug regime tends to dominate with high-viscosity liquids, such as is the case of massecuites, it can be proposed that the slug flow is likely to be present in calandria tubes and responsible for flow boiling oscillations. This hypothesis is in agreement with a comparable case
reported by Liu et al. (1995), who studied experimentally the boiling of highly viscous pseudo-plastic down-flow using a transparent film heater, observing that the bubbly regime occurred rarely and only at the entrance, while the slug and annular regimes were dominant.

Kolev (2002) states that correlations developed for frictional pressure drop in adiabatic gas-liquid channel flows tend to under-predict by several times the friction observed at sub-cooled boiling at the same mass flow rate. Experiments performed by Miropolski et al. (1967) indicated that the frictional pressure drop in flow boiling increases significantly with respect to adiabatic flows as the quality increases, this up to a critical point where it was argued that transition from nucleate to film boiling occurs. In the same way, Shoukri et al. (1981) studied experimentally the effect of heat flux on flow boiling, concluding that the application of heat flux results in a significant increase in the frictional pressure drop, and that under boiling conditions the frictional resistance tends to be higher as the gas content increases and transition to the slug regime occurs. It was postulated that the friction increases with quality up to a maximum point where transition to the annular regime occurs.

The increased frictional resistance of flow boiling has been attributed to the turbulent effect of vapor bubbles generated near to the hot surface, which would induce changes in the velocity and shear stress profiles. It is noticed that the presence of micro-bubbles near to the hot surface is also believed to be the reason for the large boiling heat transfer coefficients, as they exert a micro-convection effect that enhances the heat transfer.

Considering that friction is a determining factor in gas-liquid multiphase flows when a high-viscosity medium is involved, and that experimental evidence has demonstrated that the frictional pressure drop in heated boiling channels is higher than in adiabatic gas-liquid
channels under corresponding conditions, it can be anticipated that a significant frictional pressure drop occurs in calandria tubes that cannot be accurately predicted with conventional two-phase flow models. Since more momentum would have to be dedicated to overcome the frictional resistance in a calandria tube than in a corresponding adiabatic channel, the circulation produced would be correspondingly lower. The discussed increase in frictional resistance under boiling conditions partially explains the necessity to reduce the drag interaction during the CFD simulations of the flow in non-adiabatic calandria tubes, where the boiling flow is modeled applying a drag correlation developed under adiabatic conditions in the experimental facility to compute the interfacial momentum interaction.

4.4.5 Temperature and Saturation along a Calandria Tube

Measurements performed by Austmeyer and Schliephake (1983) and Bruhns (1996) in experimental single-tube facilities showed an increase in the flow temperature along the tubes according to the radial position, where the largest temperatures are logically observed near to the walls, while the temperature of the core does not change significantly. A contradictory behavior is observed at the upper section of the tubes, where the data corresponding to natural circulation indicates that the flow reaches a uniform saturation temperature around $H \approx 0.9$ m, while the information corresponding to forced circulation displays significant temperature differences between the fluid at the wall and the center. Rouillard (1985) measured the temperature only at the center of the tube, observing a relatively small gain that cannot give representative information on temperature variations of the flow.

Figure 4.25 summarizes characteristics of the two-phase flow in a calandria tube suggested by Rouillard (1985) from his experiments. It was proposed that at the entrance
subcooled boiling occurs with moderate evaporation and a progressive increase in temperature. When the bulk temperature reaches the saturation conditions, transition to saturated boiling would occur, accompanied by a rapid increase in the slope of the void fraction profile and departure of the static pressure from the hydrostatic profile. At this location the temperature reaches a maximum value, and begins decreasing as the pressure reduces and makes the rising liquid superheated, resulting in vaporization. This description of the phenomena in calandria tubes implies a continuous process, similar to the classic description of boiling in vertical heated channels, where well defined changes in the regime of the flow occur along the tubes as more vapor is generated and the void fraction increases.

Fig 4.25 Schematic representing two-phase flow properties along a calandria tube, indicating that as transition from subcooled to saturated boiling takes place, a rapid increase in the void fraction occurs, accompanied with separation of the static pressure from the hydrostatic profile and a maximum in flow temperature.
4.5 BOILING INSTABILITY IN VERTICAL HEATED TUBES

The most known description of boiling in vertical heated tubes considers that the flow and heat transfer phenomena take place in an unsteady but relatively continuous manner, displaying progressive transitions in the flow regime as evaporation takes place and the gas content increases along the tube. However, a less familiar boiling mechanism is also encountered in vertical heated tubes, which are prone to develop instabilities particularly in low-velocity systems (e.g. natural circulation) with low-heat transfer, low-thermal conductivity (e.g. organic fluids), and large liquid-to-vapor density ratios (e.g. water vapor at low pressure), where severe fluctuations in the evaporation, pressure, circulation, and temperature can occur and make the boiling process distinctively discontinuous.

Griffith (1962) demonstrated that the initial boiling in boiler tubes can be unstable and periodic. This phenomenon has been associated with thermo-hydraulic instabilities such as density waves, the Ledinegg instability, and geyser instability (Brennen, 2005), which are typically encountered under low-flow low-pressure conditions, and can cause significant oscillations, for example during the startup of old low-pressure water boilers and in coffee percolators. These boiling instabilities show resemblance to the discontinuous process that takes place in geo-thermal geysers.

Measurements conducted by Jeglic and Grace (1965) indicated that the onset of boiling oscillations coincides with an abrupt transition to the slug regime, which would be largely responsible for the oscillations. They concluded that as the pressure reduces and/or sub-cooling increases, the system is destabilized and boiling flow oscillations are easier to develop. It has been proposed that the high occurrence of geysering at low pressures is associated with the lower density of the vapor, as this results in larger vapor bubbles that can
occupy easily the entire cross section of the heated channel, probably catalyzing the transition to the slug regime. In practice geysering is avoided in boilers by increasing the operating pressure, suppressing the development of large bubbles and making it easier to attain stable natural circulation (Subki et al., 2004). Some steam generators are for example required to initiate the boiling above a minimum pressure (e.g. $P>500$ kPa) to avoid boiling instability.

### 4.5.1 Boiling Instability in Calandria Tubes of Vacuum Pans

Considering the characteristics of flow boiling instabilities discussed above, it seems highly probable that this can take place in the vertical heating tubes used in sugar evaporative crystallizers, particularly taking into account the low velocity of the highly viscous massecuite, the relatively low heat transfer, and the large liquid-to-vapor density ratio due to vacuum conditions.

Subki et al. (2004) determined for geysering at sub-atmospheric conditions that the fluctuations in circulation and pressure are synchronous. Based on this behavior and in an attempt to confirm whether this type of phenomenon takes place in calandria tubes, field measurements were performed installing two absolute pressure transmitters\(^{20}\) at the bottom and the top of a calandria tube in a 51 m\(^3\) C-boiling batch pan at Lula-Westfield sugar mill, Louisiana. Figure 4.26 presents details on the installation of the pressure meters and an image illustrating the two sensors installed at the bottom and the top of the calandria. A small flow of water (~2 drops/s) was purged continuously through the sensors in order to prevent blockage of the lines with massecuite. The water flow was dosed using two rotameters and needle valves. The analogue signals from the transmitters (4-20 mA) were recorded using a data logger operating at 100 Hz.

\(^{20}\) Rosemount absolute pressure transmitter 3051 CA, 0-30 Psia, calibrated 0-101.3 kPa (0-14.69 Psia), Dead time ~45 ms, total response time: ~100 ms, accuracy: ± 0.15% full scale
Fig 4.26  [a-top] Schematic illustrating details of the installation of absolute pressure transmitters for measuring at the top and the bottom of a calandria tube in a batch pan, and [b-bottom] image showing the two transmitters attached to the pan.
Figure 4.27 presents the averaged pressure and fluctuations (together with ±1 standard deviation ranges) recorded during two boilings, indicating a progressive reduction in the amplitude of fluctuation with time, which can be explained by the decrease in evaporation as the level increases. In general, the first part of the cycle is characterized by a progressive increase in pressure as a consequence of the higher hydrostatic head resulting from feeding molasses and increasing the liquid level. Then, during the ‘tightening’, the pressure decreases at the top of the tube, and this is attributed here to the lower hydrostatic head resulting from evaporating water. On the other hand, no significant changes in pressure are evident at the bottom during this final stage.

![Figure 4.27: Pressure trends determined at the bottom and the top of a calandria tube, displaying decreasing amplitude of fluctuations with time as the massecuite level increases.](image)

Pressure oscillations such as those illustrated in Figure 4.28 were observed during the first ~1½ hour. The fluctuations were particularly noticeable at the top of the tube, and are
attributed here to the discontinuous pass of vapor bubbles at this location as a consequence of boiling instability and intermittent vaporization within the calandria tubes. The fluctuations became smaller when the pressure reached its highest values (t ~ 100-130 min), and after that a radical change was appreciated, with fluctuations evident now at the bottom of the tube and highly periodic, while the oscillations at the top of the tube remained small. A signal analysis applying Fast Fourier Transform indicated that the oscillation period varied during the cycle between 1.5–5.0 s, and was on average 2.7 s.

![Fig 4.28](image)

**Boiling time 20 min**

**Boiling time 80 min**

**Boiling time 120 min**

**Boiling time 160 min**

---

Fig 4.28  Pressure oscillations recorded at the bottom and the top of a calandria tube at different boiling times, displaying discontinuity and some periodicity that indicates discontinuous evaporation and flow boiling instability in calandria tubes.

The pressure oscillations presented in Figure 4.28 correspond to ‘normal’ batch boilings, when the pan is filled up to ~40 m^3. However, a different response was also recorded during the end of a shortened boiling, where the massecuite was discharged in a receiver when the level was around ~31 m^3, and the oscillations recorded at the bottom and
the top of the tube displayed synchrony and strong periodicity as can be seen in Figure 4.29. Signal processing of the pressure recorded indicated strong peaks around $\tau \sim 3.4$ sec at the bottom and the top of the tube, indicating that the oscillations recorded at the two locations are connected and evidencing a highly periodic boiling instability.

![Figure 4.29](image)

Fig 4.29  [a-top] Pressure oscillations recorded at the bottom and the top of a calandria tube at the end of a boiling displaying high periodicity and synchrony, and [b-bottom] spectra obtained applying Fast Fourier Transform to signals showing strong peaks around 3.4 sec

Wienecke (2006) studied experimentally the boiling of high-viscosity liquids in a vertical annulus, and presented images acquired simultaneously with pressure measurements
showing that during boiling oscillations the pressure reaches peak values while the content of gas is minimum, with only a few single bubbles being present. Conversely, the minima in pressure coincide with a high content of bubbles within the vertical channel described as ‘foam’. The result seems rational in view of the large effect of the gas volumetric fraction on hydrostatic pressure in vertical gas-liquid flows.

The pressure oscillations recorded in the full-scale sugar evaporative crystallizer can be interpreted based on the observation reported by Wienecke (2006), indicating minimum presence of bubbles within the tube when the pressure reaches peak values that are marked as (i) in Figure 4.30. In the same way, the highest gas content within the calandria tubes would coincide with minimums in pressure (iv). Therefore, it can be said that most of the evaporation should take place between the peaks and the minimums (i ⇒ iv), and that during the rest of the cycle (iv ⇒ i) sub-cooled liquid would enter the tube and replace the vapor evacuated by the buoyancy forces, and be heated until the vaporization starts again.

Fig 4.30 Pressure oscillations recorded at the bottom and the top of a calandria tube displaying periodicity and synchrony
The oscillations in pressure that have been recorded indicate that there are radical differences between the boiling in calandria tubes and the typical depiction of boiling in vertical tubes that shows a continuous process with successive regime transitions as the vapor content increases. Although it is difficult to deduce in detail what occurs within a calandria tube from the information available, a hypothetical description of the process based on reported characteristics of flow boiling instability (Subki et al., 2004; Barbosa, 2005; Duffey and Rohatgi, 1996) is presented in Figure 4.31, suggesting a discontinuous but periodic process characterized by intermittent vaporization.

(i) Only sub-cooled liquid is present inside the calandria tube, being heated at approximately uniform wall temperature.

(ii) The heated massecuite approaches boiling conditions, displaying some superheat near the hot tube wall.

(iii) Nucleation and rapid growth of vapor bubbles take place particularly in the upper section of the tube, where the boiling temperature is comparatively low due to lower hydrostatic pressure. Quick transition to the slug regime is likely to occur.

(iv) The generation of vapor at the top reduces the hydrostatic pressure, promoting flash vaporization progressively downwards along the tube, which results in ‘explosive’ or ‘eruptive’ boiling.

(v) As the vapor leaves, ‘fresh’ massecuite enters the tube to re-fill the evacuated space and start the cycle again. Some backflow might occur at the end.

Fig 4.31  [a-left] Hypothetical description of the boiling in calandria tubes of sugar evaporative crystallizers based on known features of flow boiling instabilities, and [b-right] typical description of boiling in vertical tubes displaying progressive changes in the flow regime as evaporation takes place in a continuous manner
The boiling instability identified in calandria tubes explains fluctuations and intermittency in circulation observed in full-scale vacuum pans from field measurements which followed radioisotope capsules within batch crystallizers using gamma ray rate-meters to study the flow (Anon., 1964-5; Wright, 1966), observing that the circulation displays intermittency that becomes more and more significant as the viscosity increases. In the same way, the boiling instability agrees with observations by Skyring and Beale (1967) after performing tests with an external boiling tube provided with sight glasses that was attached to a batch C-boiling pan, observing intermittent flow and a ‘spasmodic’ rise of the vapor bubbles. The visualization of the boiling massecuite showed that in the lower part of the tube single vapor bubbles are nucleated and grow rapidly, and at the upper part of the tube merging takes place and large bubbles are generated, this probably as a quick transition to the slug regime occurs. These observations from experiences in full-scale evaporative crystallizers reinforce the idea that boiling instability takes place within calandria tubes.

Another factor that could relate boiling instability with the process in full-scale sugar evapo-crystallizers is the effect of the tube length. The tests performed by Jeglic and Grace (1965) included the evaluation of several heating lengths, observing that as the tube is longer the heat flux required for the onset of flow boiling oscillations reduces, and therefore short tubes are less prone to develop instabilities. Analogously, it has been determined that sugar crystallizers provided with short calandria tubes ($L \sim 0.7$ m) perform better and display higher heat transfer than those with long tubes ($L > 1$ m). This effect of tube length appears rational, since long tubes can potentially generate more vapor, and therefore would be more susceptible to experience transition to the slug regime, where the energy and momentum interaction would be affected and result in less circulation and lower heat transfer.
The boiling instability in the calandria tubes helps to justify the reduction in momentum interaction during CFD simulations of the flow sugar evaporative crystallizers, where the flow equations are solved in an idealized continuous steady process, while in reality it is highly unsteady and intermittent.

4.6 FINAL REMARKS

The results obtained in an experimental facility constructed for the evaluation of the interfacial transfer of momentum in buoyancy-driven gas-liquid vertical channel flows have indicated that the bubbles tend to move to the middle of the tube and merge, particularly at high viscosities, displaying a rapid transition to the slug and churn regimes as the gas flux is increased. The slug flow regime is unsteady and accompanied by significant oscillations that lead to intermittent or ‘pulsating’ circulation.

A progressive decay in the transfer of momentum to the liquid phase as the gas flow rate increases has been determined experimentally, suggesting that long calandria tubes can lead to poor circulation, this in agreement with practical experience that pans with short calandria tubes are more efficient. The transfer of momentum from the gas to the liquid phase appears to be more efficient within the bubbly regime, while transition to the slug and churn results in reduced drag interaction.

A model for gas-liquid momentum exchange in vertical tubes has been developed from experimental results, which is applied to compute the interaction between vapor bubbles and massecuite in CFD simulations of the flow in calandria tubes of vacuum pans. However, comparison with experiments reported by Rouillard (1985) has shown that a reduction in the drag interaction is required to obtain realistic predictions of the flow in vacuum pans. The reduced interaction has been attributed to the effect of convective boiling
based on experimental evidence that shows that the generation of vapor increases the frictional resistance and can lead to severe flow and heat transfer instabilities.

The process in calandria tubes of vacuum pans appears to be dominated by a thermo-hydraulic boiling instability typically found in low-circulation low-heat transfer low-pressure systems, which results in intermittent evaporation and corresponding oscillations in circulation and pressure. Field measurements of the pressure at the top and the bottom of a calandria tube performed in a batch pan indicated significant oscillations, suggesting that flow boiling instability occurs in calandria tubes. A possible description of the process in the calandria tubes based on reported features of flow boiling instabilities is proposed, which explains the fluctuations and intermittency in circulation found in sugar evaporative crystallizers.
CHAPTER 5 - RESEARCH ON THE FLOW IN FULL-SCALE SUGAR EVAPORATIVE CRYSTALLIZERS

The continuous sugar evaporative crystallizers, normally known as continuous vacuum pans (CVP), constitute a convenient alternative for the experimental study of the circulation at full-scale for several reasons associated with the relatively steady operation compared to the batch process, with less variation in the liquid properties, in the massecuite level, in the evaporation rate and heat transfer, and consequently a more regular circulation rate and flow patterns are expected to occur. Taking these factors into consideration, the experimental study at full-scale has been focused primarily on continuous evaporative crystallizers, and field measurements of the massecuite velocity within a CVP have been performed using insertion flow sensors that are similar in principle to the conventional ‘hot-wire’ anemometers. Numerical simulations of the flow in full-scale sugar crystallizers have been carried out applying a strategy developed based on the previous experiences in the simulation of the flow in the lab-scale test rig. The interfacial momentum interaction is modeled applying the adiabatic drag correlation obtained from the tests in the experimental facility presented in chapter 4, expecting a decrease in this interaction due to higher frictional pressure drop and flow boiling instability.

Numerical predictions of the flow in two continuous (Tongaat-Hulett and SRI) and two discontinuous or batch (straight side and flared) sugar evaporative crystallizers are presented and compared with performed and reported field measurements, observing reasonable agreement. It is believed that the CFD predicted flow field realistically represents the process within vacuum pans, and provides an enhanced description of the circulation of massecuite.
5.1 MATERIALS AND METHODS

5.1.1 Study Cases

The first study case corresponds to a Tongaat-Hulett CVP, supplied by Fletcher Smith, which was installed recently at the Enterprise Sugar Factory, Louisiana, USA, where flow measurements were performed in the 2nd and 11th cells during the 2004 and 2005 seasons to study the circulation.

The second study case is a Sugar Research Institute CVP that is installed at Tully Sugar Mill, Queensland, Australia, which has been analyzed using CFD as a part of this study, comparing the numerical results with velocity measurements reported by Broadfoot et al. (2004).

Finally, predictions of the flow field in discontinuous or batch crystallizers with straight sides and with a conical enlargement above the calandria or ‘flared’ are presented and compared with circulation flow patterns identified in full-scale pans by following the movement of a radioactive capsule with Gamma ray scintillation counters (Anon., 1964-1965).

Table 5.1 presents some of the main parameters in the design and nominal operating conditions of the continuous evapo-crystallizers studied. Several similarities between the two designs that can be appreciated are:

- The capacity in terms of volume of the crystallizer and the massecuite flow rate is practically the same in both cases, as well as the main dimensions of the shell.
- The ratio between heating area and volume is similar (~ $10 \text{ m}^2/\text{m}^3$).
- The number of cells is relatively close. The Tongaat-Hulett (12 cells) has only two additional cells with respect to the SRI design (10 cells).
Table 5.1 Nominal parameters of the continuous evaporative crystallizers studied.

<table>
<thead>
<tr>
<th></th>
<th><strong>CASE 1</strong></th>
<th><strong>CASE 2</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td>Design</td>
<td>Tongaat-Hulett</td>
<td>SRI</td>
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<tr>
<td></td>
<td>(Sup. Fletcher Smith)</td>
<td></td>
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<tr>
<td>Sugar mill</td>
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<td>Tully Sugar Limited</td>
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<td>Massecuite type</td>
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<td>Massecuite volume (m³)</td>
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<td>Nominal heating surface (m²)</td>
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<td>Ratio heating area/volume (m²/m³)</td>
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<td>Number of cells</td>
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<tr>
<td>Massecuite concentration</td>
<td>92.5 % Brix</td>
<td>91 % Dry substance</td>
</tr>
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</table>

Figures 5.1 and 5.2 present the cross section (elevation view) of the Tongaat-Hulett and SRI continuous crystallizers respectively, illustrating some significant differences in the design. The locations where velocity measurements have been performed are illustrated at
the bottom and upper-downtake regions, representing the sensors and the maximum insertion lengths from the shell.

Fig 5.1 Cross section of the Enterprise Tongaat-Hulett CVP - Case 1
The measurement locations and maximum insertion lengths are indicated at the bottom and in the upper-downtake regions.
Some design features that can have important effects on the circulation and flow patterns and are noticed from Figures 5.1 and 5.2 are:

- A vertical-tubed floating calandria is used in both cases, but the Tongaat-Hulett design has horizontal calandria plates, while the SRI design has both plates inclined.
- The massecuite head above the top calandria plate is lower in the Tongaat-Hulett design (H ~ 300 mm), while a higher liquid average head is observed in the SRI case (H ~ 600 mm). It can be assumed that the first case would exhibit a higher frictional
resistance above the calandria, while in the second case the effective temperature difference would be lowered as a consequence of the higher hydrostatic pressure.

- In both cases a generous downtake path is provided, resulting in circulation ratios ($C_R \sim 0.9$) well below the maximum limit recommended for batch crystallizers ($C_R \leq 2.5$). The downtake in the SRI design is narrower ($\leq 770$ mm) in comparison with the Tongaat-Hulett design ($\leq 970$ mm).

- An intermediate vertical plate is used in the Tongaat-Hulett design, separating two adjacent cells at each side of the pan. In the SRI crystallizer this plate does not exist, and the entire cross section corresponds to the same cell.

- Differences in the shape of the shell are observed, which would affect to some extent the flow patterns and circulation rate.

5.1.2 Measuring Technique

- Insertion Flow Sensors

The velocity of the massecuite has been measured in the two CPVs studied using commercial insertion flow meters based on the hot-wire anemometry technique, and suitable for the rough environment inside the vessels. This insertion flow sensor was used originally by Rackemann and Stephens (2002) to measure the massecuite velocity in batch evapor-crystallizers, and then by Broadfoot et al. (2004) as a part of an evaluation of the Tully SRI CVP that is being considered here as case two. The same type of sensor has been used in the present study for the field measurements in the Enterprise Tongaat-Hulett CVP (Case 1), where a ‘Turck Insertion Flow Monitor FCS-G1/4 A4-NA/D100’ was utilized to measure the velocity of the massecuite in the locations indicated in Figure 5.1 at the second and eleventh cells.
The flow sensors used in the field measurements operate based on the same principle of the traditional ‘hot-wire’ and ‘hot-film’ anemometers, measuring indirectly the fluid velocity from the convective heat loss. However, in this case the measuring elements are encapsulated within a metallic shield, allowing the operation under harsh conditions where traditional hot anemometers cannot function.

The insertion flow sensors utilized have a two-resistor configuration for temperature compensation: A first ‘reference’ resistor gives the temperature of the media, and a second ‘sensing’ resistor is heated a few degrees above that temperature. The temperature difference results in heat transferred by convection to the medium proportional to the velocity of the cross-flow, and therefore the velocity can be determined from the heat loss if the relationship between the two is known. Figure 5.3 presents the internal two-resistors and the main dimensions of the flow sensor used to measure the massecuite velocity at Enterprise.

![Schematics of the flow sensor](image)

**FCS-G1/4 A4-NA/D100**

Fig 5.3  Schematics of the flow sensor used to measure the massecuite velocity in CVPs illustrating [a-left] the internal two-resistors configuration for temperature compensation, and a third variable resistor for calibration of a Wheatstone bridge used to generate the voltage output signal, and [b-right] the main dimensions of the sensor (source: Turck Inc 2004).

The difference in resistance between the two-resistors located at the tip of the sensor is proportional to the heat transfer, and increases with the velocity of the flow. A Wheatstone
bridge is used to amplify this difference in resistance, obtaining a voltage signal proportional to the flow velocity. A variable resistance connected to the Wheatstone bridge serves for calibration purposes, giving the reference value. The relationship between resistance and flow velocity is given by King’s law, and determined in practice through a calibration process.

According to the manufacturer datasheet the Turck insertion flow sensors utilized can operate around 3-300 cm/s (oil) and have a repeatability ± 5%. Commercial signal processors ‘Turck MK96-LI01’ were used to read the signal and generate an analog current output 4-20 mA proportional to the liquid velocity, which was recorded with a Data Acquisition System ‘OMEGA ReadyDAQ AD-128’ operating at 4 Hz.

- **Calibration**

  The technique applied to measure the massecuite velocity is indirect, and consequently the validity and accuracy of the results depends to a large extend on the calibration process, which in the case of sugar evaporative-crystallizers is particularly challenging due to the complex composition and fluctuating characteristics of the massecuites, which determine physical properties that affect the relationship between the liquid velocity and the heat loss from the hot-anemometers.

  For the measurements in the Tongaat-Hulet CVP (case 1) the anemometers were calibrated on site using a rotating rig with variable speed and liquid samples taken from the same cells of the vessel where the sensors would be inserted, ensuring as far as possible representative properties of the massecuite during the calibration. The calibration was performed immediately after removing the liquid from the CVP and before a significant temperature and viscosity change could occur. Figure 5.4.a presents the calibration rig
illustrating the sensors submerged in the liquid at a known radial distance from the center of rotation. Different rotational speeds were used, obtaining the typical response of this type of sensors that is illustrated in Fig 5.4.b, and enabling a correlation between the linear speed at the tip of the sensor and the signal from the anemometer.

During the calibration it was noticed that the orientation of the flow sensor with respect to the direction of the flow is an important factor, and the recommendation of the manufacturer to install it on a particular direction (reading the ‘A4’ printed on one face as shown in Fig 5.5.a) must be carefully followed. It was observed that at 90° clockwise from the recommended orientation, as seen from the backside of the sensor, a maximum signal is always obtained; at 180° clockwise a closer but still different value is obtained; and at 270° clockwise a minimum signal is registered. This behavior is attributed here to the two-resistor configuration of the sensors and is illustrated in Figure 5.5, where the heat transfer from the hot-resistor would vary according to its position around the perimeter of the sensor and with respect to the stream-flow direction, affecting the output signal.
Fig 5.5 [a-left] Orientation for the installation of the insertion flow sensors that is specified by the manufacturer, and [b-right] effect of the orientation of the sensor with respect to the stream flow direction on the output signal.

The existence of a minimum and a maximum in the response according to the orientation of the flow sensors with respect to the stream direction was manipulated to determine the direction of the massecuite flow in the CVPs. The sensors were rotated slowly until the maximum and minimum signals were found, indicating approximately the direction of the flow. This strategy was particularly useful at the upper-downtake, suggesting the presence of a vorticity similar to the one that was observed in the downtake of the lab-scale model, while at the bottom it was consistently determined that all the massecuite flows in the same direction towards the entrance of the tubes.

For the calibration of the flow sensors used in the SRI CVP (Case 2) Broadfoot et al. (2004) report the utilization of a circulation rig that was described by Rackemann and Stephens (2002), which includes a pump and a variable speed drive that permit the control of the flow through a pipe where the sensor is inserted, and the output signal is correlated with the reading from a magnetic flow-meter installed in series and used for calibration. Solutions within the range of concentration and purity in vacuum pans were mixed and used as the
liquid phase during the calibration. The sensors were reported to be oriented in the position that gives the maximum signal²¹.

- **Mounting of the Sensors**

  For the measurements in the Tongaat-Hulett CVP (Case 1) the flow sensors were installed simultaneously in the second and eleventh cells as indicated in Figure 5.1. The maximum insertion lengths measured from the wall were 720 mm at the bottom and 1225 mm at the upper-downdraft region, covering a small section above the top calandria plate. The field measurements in the SRI CVP (Case 2) reported by Broadfoot et al. (2004) were performed in the fourth cell, and displayed maximum insertion lengths around 600 mm at the bottom and 770 mm in the downdraft as indicated in Figure 5.2.

  The measurements at the bottom were taken below the corner of the calandria plate, where all the circulating liquid passes before entering the tubes and the velocity is expected to be approximately horizontal (shown in the CFD predictions in Figure 5.19). Therefore, the integration of the velocity profile along the measurement plane is expected to provide an accurate value for the circulation rate. The measurements were taken at 15 intermediate locations recording data during 5 minutes at each one, for a total duration per measured profile around two hours.

  The insertion flow sensors were attached to the end of a tube and introduced into the vessel using a relatively simple sealing system that is presented in Figure 5.6. With this arrangement it was possible to install and remove the sensors on the run, and measure velocity at several positions along a linear path without breaking the vacuum or disturbing the process in a perceptible manner.

²¹ Probable error source, since the maximum signal is obtained at 90° from the orientation recommended by the manufacturer as illustrated in Figure 5.5.b
Fig 5.6 Arrangement for the mounting of the insertion flow sensors showing [i-left] the main components, [ii-middle] the assembly, and [iii-right] the mounting attached at the bottom of the 2nd cell of the Tongaat-Hulett CVP.

a. Steel threaded cap NPT 1”, perforated at the end to attach the sensor.
b. Steel threaded coupling NPT 1½” welded to Socket-End Compression for 1” pipe.
c. Seal, gasket is loosen to slide the sensor and then tighten to keep the vacuum.
d. Steel pipe threaded at one end NPT 1”, where the cap and the sensor are hold.
e. Cable connecting the flow sensor with the signal processor.
f. Steel nipple threaded at one end NPT 1 ½”. Welded to the wall of the crystallizer, connects with the NPT 1½” coupling (b), permitting the connection and removal of the measuring array on the run.

5.1.3 Numerical Simulation of the Flow Applying CFD

The fluid flow in full-scale crystallizers has been simulated using the commercial CFD code FLUENT. The Eulerian-Eulerian multiphase approach has been applied using a strategy similar to the one developed for the simulation of the flow in the lab-scale test rig.
The analysis has been performed using 2D grids, and assuming a vertical symmetry plane in the middle of the SRI CVP.

Although strictly three phases are present during sugar crystallization (mother liquid + vapor + sugar crystals), the high viscosity of massecuites and relatively small size of sugar crystals make it reasonable to assume that the liquid and solid phases move together, so a two-phase flow model is used (vapor and massecuite).

Vertical channels were used to represent the calandria tubes of the continuous evapo-crystallizers, which are analogous to the rings employed by Brown et al. (1992) and Stephens (2002) to simulate the calandria tubes in batch sugar evapo-crystallizers. To represent approximately the complex mass and energy interactions occurring between the vapor and massecuite in the calandria tubes, vapor has been injected along the computational domain representing the heated tubes using mass source functions that reproduce the evaporation profiles identified from the analysis of Rouillard (1985) experimental data and were presented in chapter 4 (Fig 4.20).

It is assumed that a uniform evaporation takes place across the calandria. However, knowing that tubes with less circulation evaporate less, a flow distribution across the calandria smoother than in real evapo-crystallizers is expected. In the model the vapor is injected without any momentum, and rises due to its density difference with the surrounding massecuite.

The primary buoyancy is generated by the difference between the liquid and vapor densities. The buoyancy forces caused by liquid temperature differences are assumed negligible with respect to those caused by density difference between the phases; thus the system can be treated as isothermal, avoiding the solution of the energy equation.
No information exists on the size of the bubbles in sugar evapo-crystallizers, or a drag coefficient correlation that could be applied in the numerical simulation of this flow. Stephens (2002) developed a numerical analysis assuming bubbles with size $D_G \sim 10$-15 cm in diameter, and used the Stokes drag law for the analysis ($C_D = 24/\text{Re}$). In the computations presented here the size of the bubbles has been defined assuming spherical gas particles with the same size as the calandria tubes ($D_G \sim 10$ cm), this considering that in high-void fraction gas-liquid regimes, such as the slug and churn, the cross sectional area of the bubbles approaches the internal diameter of the channel, which consequently has been assumed as characteristic dimension in previous studies. However, in reality a large dispersion in size and shape of the bubbles that would be nearly impossible to consider in detail is expected to occur as the boiling takes place and different regimes are experienced.

The correlation developed in the experimental facility under adiabatic conditions\footnote{Correlation developed for gas-liquid adiabatic channel flow: $C_D^M = N^1.6286 \alpha^{1.3951} \frac{3.6351}{\text{Re}^{1.7}}$} (chapter 4) is applied here to model the interfacial momentum interaction in the non-adiabatic calandria tubes. The exchange of momentum has been adjusted iteratively to obtain agreement between the predicted liquid velocity and the field measurements, observing that a significant reduction in drag is necessary ($M \sim 49$-52%), as it was the case in the comparison with Rouillard (1985) experimental data. The reduced interaction is a consequence of the higher frictional resistance (Fig 4.25) and severe flow and heat transfer instabilities (Figs 4.32 - 4.34) caused by the unsteady generation of vapor within the calandria tubes.

### 5.1.4 Fluid Properties and Heat Transfer

For the Tongaat-Hulett CVP (case 1) the information on the concentration and purity of the A-massecurites, syrup, and molasses was obtained from the Manufacturing Report that...
the Enterprise Laboratory issues daily. Measurements of viscosity and crystal content in samples of massecuite taken from the second and eleventh cells were performed on site using a spindle viscometer and a Nutsch pressure filter. Table 5.2 presents results from analysis of the massecuites and respective mother liquors carried by the Audubon Sugar Institute Laboratory. The viscosity measurements indicated values around 2.2-2.5 and 8.9-13.3 Pa.s at operating temperature (~ 65 °C) for the second and eleventh cells respectively, which have been averaged to 6.5 Pa.s. The densities have been obtained from tables for sucrose solutions (Bubnik et al., 1995) for the corresponding temperature, brix, and purity of the massecuites.

Table 5.2  Analysis of the composition and crystal content in massecuites at the second and eleventh cells of the Tongaat-Hulett CVP (Case 1).

<table>
<thead>
<tr>
<th>Cell 2</th>
<th>Season</th>
<th>Material</th>
<th>Brix</th>
<th>Sucrose</th>
<th>Glucose</th>
<th>Fructose</th>
<th>Fru/Glu</th>
<th>Suc/Bx</th>
<th>Cryst cont</th>
</tr>
</thead>
<tbody>
<tr>
<td>2004</td>
<td>Massecuite</td>
<td>88.40</td>
<td>81.807</td>
<td>1.058</td>
<td>1.062</td>
<td>1.00</td>
<td>92.54</td>
<td>50.57</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Mother liquid</td>
<td>79.91</td>
<td>65.983</td>
<td>1.970</td>
<td>2.044</td>
<td>1.04</td>
<td>82.57</td>
<td></td>
<td></td>
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<tr>
<td>2004</td>
<td>Massecuite</td>
<td>86.43</td>
<td>78.508</td>
<td>1.178</td>
<td>1.191</td>
<td>1.01</td>
<td>90.83</td>
<td>42.44</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Mother liquid</td>
<td>80.81</td>
<td>66.256</td>
<td>2.049</td>
<td>2.226</td>
<td>1.09</td>
<td>81.99</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2005</td>
<td>Massecuite</td>
<td>85.50</td>
<td>76.100</td>
<td>2.700</td>
<td>2.700</td>
<td>1.00</td>
<td>89.01</td>
<td>37.96</td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>79.90</td>
<td>64.100</td>
<td>4.400</td>
<td>4.300</td>
<td>0.98</td>
<td>80.23</td>
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<tr>
<td>2005</td>
<td>Massecuite</td>
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<td>80.800</td>
<td>2.000</td>
<td>2.000</td>
<td>1.00</td>
<td>92.45</td>
<td>46.11</td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>80.70</td>
<td>67.800</td>
<td>3.500</td>
<td>3.500</td>
<td>0.97</td>
<td>84.01</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2005</td>
<td>Massecuite</td>
<td>88.30</td>
<td>80.500</td>
<td>1.800</td>
<td>1.900</td>
<td>1.06</td>
<td>91.17</td>
<td>47.18</td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>83.30</td>
<td>67.500</td>
<td>3.200</td>
<td>3.500</td>
<td>1.09</td>
<td>81.03</td>
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<tr>
<td>Average</td>
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<td>87.21</td>
<td>79.543</td>
<td>1.747</td>
<td>1.771</td>
<td>1.01</td>
<td>91.20</td>
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<table>
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<tr>
<th>Cell 11</th>
<th>Season</th>
<th>Material</th>
<th>Brix</th>
<th>Sucrose</th>
<th>Glucose</th>
<th>Fructose</th>
<th>Fru/Glu</th>
<th>Suc/Bx</th>
<th>Cryst cont</th>
</tr>
</thead>
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<tr>
<td>2004</td>
<td>Massecuite</td>
<td>92.38</td>
<td>85.394</td>
<td>1.174</td>
<td>1.299</td>
<td>1.11</td>
<td>92.44</td>
<td>65.06</td>
<td></td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>76.81</td>
<td>57.171</td>
<td>2.782</td>
<td>2.720</td>
<td>0.98</td>
<td>74.43</td>
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<tr>
<td>2005</td>
<td>Massecuite</td>
<td>91.50</td>
<td>81.700</td>
<td>2.600</td>
<td>2.600</td>
<td>1.00</td>
<td>89.29</td>
<td>55.89</td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>81.40</td>
<td>59.000</td>
<td>5.900</td>
<td>5.900</td>
<td>1.00</td>
<td>72.48</td>
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<tr>
<td>2005</td>
<td>Massecuite</td>
<td>91.70</td>
<td>83.100</td>
<td>2.100</td>
<td>2.300</td>
<td>1.10</td>
<td>90.62</td>
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<tr>
<td></td>
<td>Mother liquid</td>
<td>87.60</td>
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<td>1.10</td>
<td>73.74</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2005</td>
<td>Massecuite</td>
<td>92.30</td>
<td>83.500</td>
<td>1.800</td>
<td>1.900</td>
<td>1.06</td>
<td>90.47</td>
<td>60.40</td>
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<td>Mother liquid</td>
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<td>63.000</td>
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<td>1.07</td>
<td>72.41</td>
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<tr>
<td>Average</td>
<td>Massecuite</td>
<td>91.97</td>
<td>83.424</td>
<td>1.919</td>
<td>2.025</td>
<td>1.06</td>
<td>90.70</td>
<td>60.07</td>
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<table>
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<tr>
<th>Cell 12</th>
<th>Season</th>
<th>Material</th>
<th>Brix</th>
<th>Sucrose</th>
<th>Glucose</th>
<th>Fructose</th>
<th>Fru/Glu</th>
<th>Suc/Bx</th>
</tr>
</thead>
<tbody>
<tr>
<td>2004</td>
<td>Final Mcte</td>
<td>92.17</td>
<td>85.478</td>
<td>1.140</td>
<td>1.224</td>
<td>1.07</td>
<td>92.74</td>
<td></td>
</tr>
</tbody>
</table>

Several relevant variables such as the calandria pressure, vacuum, massecuite boiling temperature, and total heat supplied to the pan (computed from the condensate flow), were
obtained from corresponding sensors (pressure transducers, thermocouples, and flow meter) mounted at appropriate locations in the crystallizer. The crystallizer overall heat transfer coefficient is computed as:

\[
h = \frac{\dot{Q}_w}{A_{HT} \Delta T}
\]

Where: Heat transfer rate: \( \dot{Q}_w = \dot{m}_{\text{condensate}} \cdot h_{fg} \)

Area: \( A_{HT} = \pi \cdot OD_{\text{TUBE}} \cdot L_{\text{TUBE}} \cdot \text{No. tubes} \)

Temp difference: \( \Delta T = T_{\text{HEATING VAPOR}} - T_{\text{BOILING}} \)

The heat transfer coefficient is expressed in a non-dimensional form as a Nusselt number:

\[
Nu = \frac{h \cdot OD_{\text{TUBE}}}{k}
\]

For case 2 the massecuite properties and operation conditions reported by Broadfoot et al. (2004) are considered in the numerical analysis, and unknown parameters are assumed based on typical values.

5.2 FIELD MEASUREMENTS

5.2.1 Massecuite Velocity

Figure 5.7 presents the velocity profiles determined with the flow sensors inserted at the bottom of the Tongaat-Hulett CVP beneath the corner of the calandria. A significant dispersion in the data is appreciated, particularly for the second cell, which is attributed to the scaling of the tubes and corresponding decay in heat transfer, to the inherent unsteadiness of the boiling process, and to the fluctuations in the operational variables of the CVP during the relatively long periods required to complete each run (\(~2\) h). The averaged trends for the two cells are represented in Figure 5.7 with thicker black curves, which have been integrated to quantify representative values of the circulation rate in each cell.
The velocities recorded in the second cell are higher than in the eleventh, reaching peak values up to 0.280 and 0.09 m/s respectively. This result indicates clearly that the circulation rate is more vigorous in the second cell, and this is the natural consequence of the increase in viscosity of the massecuite with the number of the cell as the sugar crystals grow, increasing the frictional resistance and lowering the heat transfer rate, both factors that are detrimental to circulation.

![Velocity profiles](image)

Fig 5.7 Velocity profiles measured at the bottom of the Tongaat-Hulett CVP, case 1, below the corner of the calandria in [a-left] the second cell and [b-right] the eleventh cell. The position is defined respect to the width between the bottom wall and the calandria L= 758mm

The averaged velocity profiles at the bottom of the second and eleventh cells are compared in Figure 5.8 in terms of the magnitude and the ratio with respect to the maximum velocity recorded at each cell. Non-centered velocity profiles, skewed towards the calandria
plate are observed in both cases, which show some similarity and indicate that the velocity of the liquid phase is faster near the bottom calandria tube plate, and slower towards the bottom wall of the vessel.

![Diagram](image)

**Fig 5.8** Averaged velocity profiles from measurements at the bottom of the Tongaat-Hulett CVP beneath the corner of the calandria in terms of [a-left] the velocity magnitude and [b-right] the ratio with respect to the maximum velocity at each cell. The position is defined respect to the width between the bottom wall and the calandria L = 758 mm.

Since the Tongaat-Hulett CVP has a single calandria, it is not possible to determine the specific evaporation rate at each one of the cells from the total condensate flow measurement that is available. To deal with this situation in the numerical analysis an averaged velocity profile has been computed from the results obtained in the second and eleventh cells, which represents an averaged flow situation where the global evaporation rate and averaged properties of the massecuite will be assumed applicable. The averaged profile is illustrated in Figure 5.8.a, displaying a peak velocity around $U_{\text{MAX}} \approx 0.11$ m/s.
The measurements in the upper-downtake region proved to be more difficult due to the non-perpendicular orientation of the flow with respect to the inserted flow sensors. From observation through the sight-glass windows of the pan it is appreciated that a practically horizontal stream of massecuite flows rapidly from the region above the calandria towards the side wall, probably as a consequence of the relatively low massecuite head above the top calandria plate. This horizontal stream is aligned with the axis of the sensor, and normal to the correct orientation for the measurements. As a result, the data obtained in this region cannot be considered an accurate representation of the liquid velocity.

Figure 5.9 presents the data obtained in the upper-downtake region, illustrating similar trends in the two cells studied. The maximum values were obtained when the hot-anemometers were located above the calandria plate (illustrated in Figure 5.9 as x/L>1), where the high signal is attributed to the elevated turbulence occurring in this region as the bubbles that exit from the heating tubes rise and induce severe flow disturbances, which would result in the increase in heat transfer from the hot-anemometer and were manifested in severe vibration of the array used to insert the sensors. Just as in the bottom, the velocity determined in the downtake of the second cell is higher than in the eleventh cell (0.14 / 0.06 = 2.3), this as a consequence of the increase in viscosity with cell number.

The effect of the orientation of the flow sensors with respect to the direction of the stream-flow was discussed previously in this chapter when the calibration procedure was presented, indicating that a maximum and minimum were detected at 90° and 270° with respect to the orientation specified by the manufacturer. This response has been exploited here to determine the direction of the flow component normal to the inserted sensors, which
were rotated slowly to identify the maximum and minimum positions, indicating approximately the orientation of the flow.

![Graph showing velocity vs. insertion length](image)

Fig 5.9 Velocity determined in the upper-downtake region of the Tongaat-Hulett CVP. These results are considered inaccurate. The position is defined respect to the width between the shell and the calandria $L \sim 975$ mm.

The use of the described procedure rotating the sensors at the bottom of the CVP indicated consistently that the massecuite flows uniformly from the downtake towards the entrance of the tubes. On the other hand, in the upper-downtake it was found that the massecuite flows upwards above the calandria ($x/L > 1$) and in a large part of the downtake adjacent to the calandria, covering around $\sim 40\%$ of the downtake area in the second cell, and around $\sim 20\%$ in the eleventh cell. In the rest of the downtake ($x/L \leq 0.60-0.80$), next to the side wall, the readings indicated that the massecuite flows downwards. The detected direction of the flow is represented in Fig 5.10 indicating with positive unit values the flow upwards, and negative unit values the flow downwards.
Fig 5.10 Orientation of the flow determined in the upper-downtake of the Tongaat-Hulett CVP by rotating the sensors. The flow directed upwards is represented as +1, and downwards as -1. The position is defined with respect to the width between the shell and the corner of the calandria L ~ 975 mm (0=side wall).

The results on the orientation of the flow suggest the existence of a large-scale flow separation in the upper-downtake region towards the corner of the calandria, which would imitate the vortex that was detected at the same position in the lab-scale model. According to the information registered the center of the vortex might be located around x/L ~ 0.8 in the second cell, and x/L ~ 0.9 in the eleventh cell. In the same way, it can be seen that the vortex would be smaller in the eleventh cell, a logical consequence of the higher viscosity and lower circulation velocity. The existence of a vortex in the downtake is reasonable considering the sharp change in the direction of the liquid flow that is forced by the geometry of the crystallizer, expecting a 90° turn within a relatively small section, and explains the wall-picking profile obtained with the flow sensors at this location (Fig 5.9). It could be argued that the generation of this vortex is detrimental to the circulation as the effective area of the downtake is reduced and the massecuite velocity increases, and consequently the frictional resistance would be higher. However, the generation of this vortex seems difficult to prevent.
The massecuite velocities reported by Broadfoot et al. (2004) for the SRI CVP case are reproduced in Figs 5.11 and 5.12, indicating the velocity profiles measured in the bottom and downtake respectively.

Despite that a direct comparison of the velocities measured in the two continuous crystallizers studied is inappropriate due to differences in the design, operation conditions, massecuite, and the location and procedure followed for the measurements, it can be seen that some similarities are evident, such as:

- The velocities determined below the calandria of the SRI CVP (Fig 5.11) are comparable in magnitude with the results obtained at the bottom of the Tongaat-Hulett CVP (Fig 5.7). Besides, in both cases a profile skewed towards the bottom calandria plate is noticeable.

![Fig 5.11 Velocity profile measured by Broadfoot et al. (2004) at the bottom of the SRI CVP below the calandria in the fourth cell.](source: Broadfoot et al., 2004)
The velocity profile in the downtake region of the SRI CVP presented by Broadfoot et al. (2004) displays a minimum around the middle of the downtake (Fig 5.12), supporting the idea that a vortex is developed there, as was observed in the lab-scale model and detected through rotation of the flow sensors in the downtake of the Tongaat-Hulett CVP.

It is noticed that in the downtake of the SRI CVP the highest velocity was determined near to the side wall ($U \sim 0.20\, \text{m/s}$), and in the Tongaat-Hulett CVP relatively high velocities were also registered at this location ($U \sim 0.08-0.14\, \text{m/s}$). These high velocities are potential consequences of the vortex at the downtake that has been hypothesized, and would indicate the presence of high velocity gradients, and therefore high frictional resistance in this part of the crystallizers.
5.2.2 Massecuite Circulation and Heat Transfer

The steam consumption and heat transfer of the Tongaat-Hulett CVP (Case 1) are monitored using a magnetic flow-meter to measure the flow of the condensate leaving the calandria, which is a good indicator of the total amount of vapor condensed by the crystallizer. However, the readings from this meter were found to be suspiciously low for the duty of the CVP. To deal with this situation the evaporation rate has been calculated from a mass balance involving the main streams of the pan, using the information that is available in the control room and in the daily report produced by the Laboratory of the sugar mill. For the analysis successive days were chosen covering periods between consecutive cleanings, so high and low evaporation rates would be included.

Figure 5.13 presents the calculated steam consumption of the CVP for several different days with respect to the apparent temperature difference. The trend observed in the results indicates lower steam consumption as the apparent temperature difference increases, this due to the scaling of the heating tubes. As incrustation takes place over the heat exchange surface the conduction resistance goes up, and the heat transfer progressively deteriorates. To compensate for this effect, the pressure in the calandria is increased, resulting in a higher apparent temperature difference, but the capacity and heat transfer have been affected by the scaling, and consequently a reduction in the demand of heating vapor is experienced. The calculations indicated a condensate flow rate around 50-60% higher than displayed by the condensate flow meter (1.56:1), suggesting steam consumptions between 13-26 kg/h.m², with an averaged value around 19.4 kg/h.m² at $\Delta T \sim 32 ^\circ$ C.

---

23 Condensate flow-meter indicated steam consumption between 10-15 kg/h.m², with average ~12 kg/h.m² vs. typical values around 16-25 kg/h.m²
Table 5.3 presents operational variables of the Tongaat-Hulett CVP that were determined and collected from the control room of the pan floor and the fabrication reports. This information will be assumed to be representative of the conditions within the CVP for the numerical simulation of the flow.

The circulation rate in the second and eleventh cells of the Tongaat-Hulett CVP has been obtained by integration of the velocity profiles measured at the bottom along the line beneath the corner of the calandria. In spite of the large dispersion in the field measurements, it was noticeable that the velocity and circulation rate increase with the evaporation rate. Figure 5.14 presents the circulation measured, in terms of the superficial velocity of the liquid phase in the calandria tubes ($J_L$) or liquid flux, with respect to the evaporation rate, illustrating an increase in circulation as the evaporation is higher. A
significantly higher circulation rate was obtained for the second cell with respect to the eleventh (2.6:1) as a consequence of the increase in viscosity with the number of the cell.

Table 5.3 Operation parameters considered for the analysis of Case 1

<table>
<thead>
<tr>
<th>HEATING VAPOR</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure (bar abs)</td>
<td>1.03</td>
</tr>
<tr>
<td>Temperature (C)</td>
<td>100</td>
</tr>
<tr>
<td>Flow (t/h)</td>
<td>32.1</td>
</tr>
<tr>
<td>Flow (kg/h.m²)</td>
<td>19.4</td>
</tr>
<tr>
<td>ΔT (C)</td>
<td>32.2</td>
</tr>
<tr>
<td>h (W/m².K)</td>
<td>377</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>EVAPORATION</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure (bar abs)</td>
<td>0.16</td>
</tr>
<tr>
<td>Saturation temp (C)</td>
<td>55</td>
</tr>
<tr>
<td>Flow (t/h)</td>
<td>30.7</td>
</tr>
<tr>
<td>Flow (kg/h.m²)</td>
<td>18.5</td>
</tr>
<tr>
<td>Density (kg/m³)</td>
<td>0.20</td>
</tr>
<tr>
<td>Viscosity (Pa.s)</td>
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</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>MASSECUITE</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Concentration (Brix)</td>
<td>91%</td>
</tr>
<tr>
<td>Purity</td>
<td>85%</td>
</tr>
<tr>
<td>Boiling temp (C)</td>
<td>68</td>
</tr>
<tr>
<td>Density (kg/m³)</td>
<td>1450</td>
</tr>
<tr>
<td>Viscosity (Pa.s)</td>
<td>6.5</td>
</tr>
</tbody>
</table>
Fig 5.14 Effect of the average pan evaporation on circulation measured in the Tongaat-Hulett CVP (case 1). The circulation rate is given in terms of the superficial liquid velocity in the calandria tubes ($J_L$).

The circulation data presented in Fig 5.14 shows a non-linear behavior that suggests a progressive loss in the capacity to induce circulation or effective transfer of momentum to the liquid phase as the evaporation is higher. The circulation results display the same asymptotic behavior registered previously in the test rig as the gas flux increases, and are consistent with experiences in the study of slugs and churns, where a reduction in the drag coefficient has been observed to occur as the void fraction is higher; and also with experimental results obtained in a natural circulation boiling heated tube (Subki et al., 2004), where the circulation reported displays an asymptotic behavior as the heat transfer increases.

The circulation period ($\tau$) is computed as the ratio between the nominal volume of massecuite of the evaporative crystallizer and the volumetric circulation rate, obtaining the averaged time that would be required for the fluid particles to complete a circuit along the path within the crystallizer. The field measurements indicated that the circulation produced
is higher as the evaporation rate increases, increasing the superficial velocity of the liquid in
the tubes and reducing the circulation period. Table 5.4 presents the circulations for different
evaporation rates obtained from the trends of the measurements displayed in Figure 5.14. An
averaged condition between the circulations in the second and eleventh cells is calculated,
giving reference values for comparison with the numerical simulations.

Table 5.4 Effect of the evaporation rate on the circulation determined from
velocity measurements in the Tongaat-Hulett A-CVP (Case 1).

<table>
<thead>
<tr>
<th>EVAPORATION (kg/h.m²)</th>
<th>CIRCULATION - J_L (m/s)</th>
<th>CIRCULATION PERIOD – τ (s)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Cell 2</td>
<td>Cell 11</td>
</tr>
<tr>
<td>10</td>
<td>0.065</td>
<td>0.026</td>
</tr>
<tr>
<td>15</td>
<td>0.082</td>
<td>0.031</td>
</tr>
<tr>
<td>18.5 (Average)</td>
<td>0.093</td>
<td>0.034</td>
</tr>
<tr>
<td>20</td>
<td>0.098</td>
<td>0.035</td>
</tr>
</tbody>
</table>

Figure 5.15 presents the effect of the circulation measured at the bottom of the
Tongaat-Hulett CVP on the apparent heat transfer coefficient (h) and the Nusselt number
(Nu). It is clearly appreciated that the heat transfer is higher as the circulation increases, and
vice-versa, showing the strong interaction that exists between the two phenomena in sugar
evaporative-crystallizers. This result agrees with practical experiences in the application of
assisted circulation to full-scale sugar evaporative-crystallizers, where the increase in the
velocity of the massecuite has resulted in enhanced heat transfer and capacity.

For the SRI-CVP (case 2) Broadfoot et al. (2004) reported several operational
parameters that are summarized in Table 5.5, including also some properties that will be
assumed here for the numerical simulation of the flow.
Fig 5.15 Effect of the circulation on the heat transferred in terms of [a-left] the apparent heat transfer coefficient, and [b-right] the Nusselt number. The circulation rate is presented as the superficial velocity of the liquid phase within the calandria tubes ($J_L$).

Table 5.5 Operation parameters considered for the analysis of Case 2.

<table>
<thead>
<tr>
<th><strong>HEATING VAPOR</strong></th>
<th>Pressure (bar abs)</th>
<th>~1.25</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Temperature (C)</td>
<td>106</td>
</tr>
<tr>
<td></td>
<td>Flow (t/h)</td>
<td>29</td>
</tr>
<tr>
<td></td>
<td>Flow (kg/h.m$^2$)</td>
<td>19.4</td>
</tr>
<tr>
<td></td>
<td>$h$ (W/m$^2$.K)</td>
<td>350</td>
</tr>
<tr>
<td><strong>EVAPORATION</strong></td>
<td>Pressure (bar abs)</td>
<td>0.16</td>
</tr>
<tr>
<td></td>
<td>Saturation temp. (C)</td>
<td>55</td>
</tr>
<tr>
<td></td>
<td>Flow (kg/h.m$^2$)</td>
<td>18.5 – Assumed</td>
</tr>
<tr>
<td></td>
<td>Density (kg/m$^3$)</td>
<td>0.20</td>
</tr>
<tr>
<td></td>
<td>Viscosity (Pa.s)</td>
<td>1.1x10$^{-5}$</td>
</tr>
<tr>
<td><strong>MASSECUITE</strong></td>
<td>Concentration (DS)</td>
<td>91%</td>
</tr>
<tr>
<td></td>
<td>Purity</td>
<td>89%</td>
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<td></td>
<td>Boiling temp (C)</td>
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</tr>
<tr>
<td></td>
<td>Density (kg/m$^3$)</td>
<td>1450 – Assumed</td>
</tr>
<tr>
<td></td>
<td>Viscosity (Pa.s)</td>
<td>6.5 – Assumed</td>
</tr>
</tbody>
</table>
5.3 NUMERICAL RESULTS

5.3.1 Interfacial Exchange of Momentum

During the simulations of the flow in the Tongaat-Hulett CVP (case 1) it was observed that the momentum interaction had to be reduced with respect to the drag correlation developed in the experimental facility under adiabatic conditions (chapter 4) in order to obtain a flow prediction within the range that was determined in the field measurements, this consistently with the reduction that was required previously to obtain agreement with Rouillard (1985) experimental data. In general, the computations have shown that as the drag interaction is decreased the gas phase rises faster, less momentum is transferred to the liquid phase, and as a consequence the slip velocity is higher and the void fraction goes down.

Figure 5.16 presents the measured and computed circulation rate within a range of evaporation rates, where agreement between the predicted and measured circulation was obtained reducing the drag interaction by $M \sim 49\text{-}52\%$ with respect to corresponding adiabatic conditions. It can be appreciated that the drag correction is more important as the evaporation and the heat transfer rates increase, moving the system farther from adiabatic conditions.

It is observed that the momentum exchange has been adjusted within a relative narrow range ($\pm 0.015/0.505 = \pm 2.9\%$), which is below the repeatability of the flow sensors employed for the field measurements ($\pm 5\%$). It can be said therefore that the drag correlation developed is effective predicting the effect of the gas flux and void fraction on the circulation produced, although a significant reduction associated with the non-adiabatic discontinuous boiling has been necessary.
Fig 5.16  Circulation in the Tongaat-Hulett CVP measured and computed with CFD for different evaporation rates, illustrating that important reductions in the momentum interaction (M ~ 49-52%) with respect to corresponding adiabatic conditions are required and that this reduction is more significant as the evaporation and heat transfer are higher.

5.3.2  Grid Independence

A grid-independence study was performed to determine a minimum acceptable grid-size, observing a significant increase in the solution time as the grid is larger. This is particularly critical for the studied case since the Eulerian-Eulerian model solves the transport equations for each phase, resulting in a highly expensive technique from a computational point of view. Table 5.6 presents details on the different grids that have been considered.

Table 5.6  Grid-sizes considered for grid independence analysis

<table>
<thead>
<tr>
<th>Case</th>
<th>Grid size (mm)</th>
<th>No. faces</th>
<th>No. nodes</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>20</td>
<td>57 178</td>
<td>29 986</td>
</tr>
<tr>
<td>2</td>
<td>15</td>
<td>77 966</td>
<td>40 433</td>
</tr>
<tr>
<td>3</td>
<td>10</td>
<td>134 006</td>
<td>68 553</td>
</tr>
<tr>
<td>4</td>
<td>5</td>
<td>531 693</td>
<td>268 970</td>
</tr>
</tbody>
</table>
In general, similar flow fields were predicted with all the grid sizes, except for the first case that corresponds to the coarsest grid, which showed difficult convergence and slight differences in the void fraction predicted. Figure 5.17 presents the contours of void fraction obtained with different grid-sizes, indicating for cases three and four the development of ‘gas columns’ above the channels representing the calandria tubes. This feature is still appreciated in case three, but cannot be clearly distinguished in case one, and therefore it is considered that corresponding grid-sizes (>15 mm) are inappropriate for the intended analysis.

Figure 5.18 presents a comparison of the liquid velocity profile predicted with different grid sizes below the corner of the calandria in the location where the field measurements were conducted using the insertion flow sensors. No significant differences between the profiles predicted are appreciated, and this seems rational considering that in this section of the crystallizer the flow is single-phase, with only highly viscous massecuite being present, and therefore the flow should be comparatively easy to predict, even with coarse grids representing the geometry of the crystallizer. The results presented in Fig 5.18 indicate that the circulation predicted is not affected significantly by the grid-size, and that the last two cases (5 and 10 mm) display practically the same velocity profile.

Numerical flow predictions are known to be increasingly accurate as the grid is finer, and it would be desirable to use an extremely fine grid for this study, but the high computational cost of the Eulerian-Eulerian model and the complexity of the studied case oblige being pragmatic about this aspect. Considering the previous grid independence analysis, it is considered that a grid-size around ~10 mm is enough to obtain correct predictions of the fluid flow in sugar crystallizers while keeping the required time for
solution within reasonable limits (~ 1 day). Further increases in the grid size result in higher computational expense (e.g. 3 days with 5 mm) while yielding similar solutions.

Fig 5.17 Contours of void fraction obtained with different grid-size, suggesting that the grid-size should not exceed ~ 10-15 mm.
5.3.3 Fluid Flow in the Tongaat-Hulett CVP - Case 1

Figure 5.19 presents the velocity vectors that have been predicted for the Tongaat-Hulett CVP with an evaporation of 18.5 kg/h.m$^2$, indicating a relatively smooth flow in the bottom and downtake regions. A small vortex is predicted to develop in the upper-downtake towards to the corner of the calandria. In the two-phase section above the calandria a complex situation is observed as the liquid moves across the rising gas phase. Apart from a small portion at the bottom next to the middle wall, no significant stagnant areas are predicted.
Figure 5.20 presents liquid flow streamlines predicted for different evaporation rates (10-20 kg/h.m²), illustrating that within the range evaluated the main features of the flow patterns predicted are the same. According to this result, the decay in heat transfer due to scaling of the calandria tubes does not induce significant changes in the flow patterns within continuous vacuum pans, although the circulation and massecuite velocity are decreased.
Fig 5.20 Streamlines of the liquid-phase flow predicted for the Tongaat-Hulett CVP indicating similar flow patterns within a range of evaporation rates.

Figure 5.21 presents a comparison of the averaged velocity profiles measured at the bottom of the Tongaat-Hulett CVP with the insertion flow sensors, and predicted with CFD for an evaporation rate of 18.5 kg/h.m², showing a reasonable agreement, particularly in the shape of the profiles, which are skewed towards the bottom calandria plate in both cases.
Fig 5.21 Velocity profile underneath the corner of the calandria of the Tongaat-Hulett CVP measured and numerically predicted. The position is defined respect to the width between the bottom wall and the calandria L = 758 mm

It was discussed earlier that the field measurements at the upper-downtake location are not accurate due to the non-normal orientation of the flow with respect to the sensors and the violent disturbances caused by the gas bubbles erupting from the calandria tubes. Fig 5.22 presents a comparison of the averaged velocity profiles measured in this region of the Tongaat-Hulett CVP with the insertion flow sensors and the velocity predicted with CFD for an evaporation rate of 18.5 kg/h.m², showing a poor agreement between computations and measurements. Some similarity is noticed between the trends from the qualitative point of view, indicating a large velocity gradient near to the side wall and a steep increase in the velocity in the part of the downtake close to the calandria (x/L ~ 0.8–1.0), which would be caused by the vortex that has been recorded in the lab-scale model and recognized during the
field measurements. Measurements and computations indicate a minimum in velocity towards to the center of the downtake channel.

![Velocity profile graph](image)

**Fig 5.22** Velocity profile at the upper-downtake region of the Tongaat-Hulett CVP measured and numerically predicted. The position is defined with respect to the width between the shell and the corner of the calandria $L = 975$ mm

Figure 5.23 presents the numerically predicted contours of void fraction, indicating an almost vertical rise of the vapor bubbles, whose presence seems limited to the heating tubes and the region above the top calandria plate. The predicted flow of the gas phase indicates that the vapor would leave the tubes and rise forming distinguishable ‘vapor columns’. However, in reality boiling instability would cause severe fluctuations and deviation from the predicted flow field. It must be kept in mind that the numerical results presented here
constitute an averaged description of an unsteady and discontinuous process, and the predicted vapor columns represent bubbles rising intermittently.

Figure 5.23 Contours of void fraction numerically predicted illustrating a vertical rise of the gas phase, whose presence is restricted to the calandria tubes and above the top calandria plate.

Figure 5.24 presents the contours of absolute pressure predicted numerically, showing that the hydrostatic head determines largely the pressure field. This result is similar to the pressure field computed for the lab-scale test rig, and seems physically reasonable in view of the high density and low velocity of the massecuite. The maximum pressure is predicted at the lowest grid point (50 121 Pa abs), and is close in magnitude with the pressure obtained from the maximum possible hydrostatic head.
\[ P = P_0 + \rho \cdot g \cdot h = 14000 + 1450 \text{ kg/m}^3 \cdot 9.8 \text{ m/s}^2 \cdot (0.8+1.4+0.3) \text{ m} = 49500 \text{ Pa} \]

Fig 5.24 Contours of absolute pressure numerically predicted for the Tongaat-Hulett CVP, illustrating a pressure field determined largely by the hydrostatic effect.

5.3.4 Fluid Flow in the Sugar Research Institute CVP - Case 2

For the simulation of the flow in the SRI CVP the same approach used previously for the Tongaat-Hulett CVP has been applied, assuming identical material properties and equivalent boundary conditions. The effect of having tubes with different length and hydrostatic head has been considered by injecting vapor with mass source functions defined individually for each one of the channels representing a set of tubes of certain length, this assuming that the effect of the tube length and liquid head on the vaporization can be
represented approximately with the correlation developed from Rouillard (1985). Due to the uncertainty on the different interactions that take place within sugar evapo-crystallizers and the assumptions made for the numerical simulations, a direct comparison between the two continuous evaporative crystallizers studied must be careful with respect to quantitative values such as the circulation rate, although the features of the flow patterns are expected to be correct in both cases.

Figure 5.25 presents the liquid velocity vectors that have been predicted for the SRI CVP, indicating smooth flow patterns in the low-downtake and the region below the calandria tubes, while a vortex would be developed in the upper section of the downtake, which extends above part of the calandria. The region above the top calandria plate is exposed to severe disturbances caused by intermittently rising bubbles as they ‘erupt’ out of the tubes and disengage from the liquid phase. However, on average the circulating flow could resemble the predicted field, and it can be hypothesized that the vortex generated would reduce the effective area of the cross section, and therefore increase the flow resistance above the calandria and in the upper-downtake region.

Figure 5.26 presents a comparison of the massecuite velocity measured at the bottom of the SRI CVP by Broadfoot et al. (2004) with insertion flow sensors and the velocity predicted in this work with CFD. The predicted and measured velocities are within the same order of magnitude, although the simulations tend to indicate lower values. Both profiles resemble the characteristic velocity profile of laminar channel flows, although some difference is noticeable, indicating the computations a centered profile (peak at x/L ~ 0.53), while the measurements show a profile slightly skewed towards the bottom calandria plate (peak at x/L ~ 0.64).
Fig 5.25 Liquid-phase velocity vectors predicted using CFD for the SRI CVP

Fig 5.26 Velocity below the calandria of the SRI CVP measured by Broadfoot et al. (2004) and numerically predicted in this study. The position is defined respect to the width between the bottom wall and the calandria \( L = 600 \text{ mm} \)
Figure 5.27 presents a comparison of the velocity measured by Broadfoot et al. (2004) in the upper-downtake of the SRI CVP with insertion flow sensors and the velocity predicted with CFD in this study, showing a poor agreement between computations and measurements. As in the first case, some similarity is noticed between the trends from a qualitative point of view, indicating a large velocity gradient near to the side wall, where the largest measured velocity is registered, and a steep increase in the velocity in the part of the downtake near to the calandria. The measurements indicated a minimum velocity towards the middle of the downtake channel, and this is attributed here to the vortex that has been predicted in the upper section of the downtake, which would cause a non-perpendicular orientation of the flow with respect to the sensor at this location, just as was observed in the analysis of the Tongaat-Hulett CVP.

Fig 5.27  Vertical velocity at the downtake of the SRI CVP measured by Broadfoot et al. (2004) and numerically predicted in this study. The three measurements located closest to the calandria have been inverted to up flow based on the predicted vortex. The position is defined with respect to the width between the shell and the corner of the calandria L~770 mm
Figure 5.28 presents the numerically predicted contours of void fraction, indicating a practically vertical rise of the vapor phase, whose presence is limited to the heating tubes and the region above the calandria plate. The quantity of evaporated water per tube increases logically with the tube length, and this is reflected in the flow simulations in void fractions at the top of the tubes that are higher as the length increases.

![Contours of void fraction numerically predicted for the SRI CVP](image)

Figure 5.29 presents the predicted contours of absolute pressure, showing that the hydrostatic effect largely determines the pressure field. The maximum pressure is predicted at the lowest grid point (53 903 Pa abs), and is close in magnitude with the pressure obtained from the maximum hydrostatic head:
$P = P_0 + \rho g h = 14\,000 \text{ Pa} + 1450 \text{ kg/m}^3 \times 9.81 \text{ m/s}^2 \times 2.88 \text{ m} = 54\,900 \text{ Pa}$

Fig 5.29 Contours of absolute pressure numerically predicted for the SRI CVP.

### 5.3.5 Fluid Flow in Batch Sugar Crystallizers

Anon. (1964-5) studied the circulation in full-scale batch sugar evapo-crystallizers by following the movement of a radioactive capsule incorporated in the massecuite with gamma ray scintillation counters. The field measurements illustrated possible flow patterns in batch crystallizers provided with floating calandria and with central-downtake, which have been considered in this study for comparison with corresponding numerical simulations.

For the analysis it has been assumed that the massecuite has a viscosity of 11 Pa.s, density 1450 kg/m$^3$, that the evaporation rate is 18.5 kg/h.m$^2$, and that the volume of the pans is 45 m$^3$. Tubes with 0.10 m in diameter and 1.0-1.4 m in length are assumed. The design of the crystallizers has been defined based on the pan geometries presented by Anon. (1964-5).
Floating Calandria - Case 3

In this design the downtake is located in the outer annular region. Floating calandrias produce poor circulation and promote false grain formation, and because of this they are obsolete for batch crystallizers. The case studied includes also an inclined bottom calandria plate, a smooth conical enlargement of the shell above the calandria, and a conical-bottom.

Figure 5.30 presents the flow patterns determined experimentally by Anon. (1964-5) and liquid velocity vectors predicted in this study, showing qualitative agreement in indicating that a large-scale vortex is developed towards the sides above the downtake and the top calandria plate. The simulations indicated low circulation (τ ~ 420 sec), evidenced in low velocities at the downtake and the bottom. The poor circulation is attributed to the increased frictional resistance in the downtake channel and the strong vortex at the sides, which seems to make it difficult for the massecuite to penetrate into the downtake.

Figure 5.30  Flow patterns in batch sugar crystallizers provided with floating calandria that were [a-left] identified experimentally by Anon. (1964-5), and [b-right] numerically predicted in this study. The simulations confirm the development of a large vortex above the downtake and part of the calandria, and indicate poor circulation.
Most continuous sugar crystallizers in use today incorporate floating calandrias, but provided with generous downtake channels. The main feature observed in the flow patterns of the batch floating-calandria case is the presence of a large vortex above the downtake, which shows some similarity with the experimental and numerical results obtained for the Tongaat-Hulett and the SRI continuous crystallizers. It can be hypothesized that the large downtake area used in continuous crystallizers makes it possible for this vortex to develop and penetrate into the upper-section of the downtake, potentially reducing its effect on the circulation rate with respect to a batch crystallizer provided with floating calandria.

- **Central-downtake Calandria - Case 4**

  In this design a single central downtake is used, with vertical tubes arranged around it. This is the most common design for batch crystallizers, probably due to the simple construction and maintenance, and good circulation results. The case studied incorporates straight-side walls and a stream flow-bottom.

  Figure 5.31 presents the flow patterns determined experimentally by Anon. (1964-5) and the liquid velocity vectors predicted in this study for the central-downtake case, showing reasonable agreement with the main characteristics of the flow field. The computed liquid velocity vectors indicate smooth flow patterns in the downtake and the region below the calandria tubes, while a vortex is developed above the downtake and part of the calandria. Another vortex, smaller and opposite in orientation, is predicted to occur above the calandria towards the sides. The presence and orientation of the two vortices shows qualitative agreement between measurements and computations, indicating that the numerical solution might be representing the massecuite flow reasonably well.
Most of the batch evapo-crystallizers in use today have design similarity with the central-downtake case studied, and it is highly probable that they would display comparable flow patterns to the ones that were delineated roughly by Anon. (1964-5) and that are described with more detail in the present study.

5.4 FINAL REMARKS

The fluid flow in two widely adopted designs of continuous sugar evapo-crystallizers has been studied. Field measurements performed in a CVP indicated that a strong interrelation exists between circulation and heat transfer, as is normal in natural convection processes. A loss in the capacity of the vapor to transfer momentum to the liquid phase and produce circulation as the gas flux increases has been evidenced from the field measurements, and this behavior is consistent with the results obtained in the
experimental facility that are presented in chapter 4, as well as with experiences in natural circulation water-boiling, in air-lift processes, and in fundamental studies of slugs and churns, where it has been established for high void fraction flows that a lower momentum interaction occurs as the gas flux and the void fraction are increased. For the simulation of the flow in full-scale crystallizers a reduction in the interfacial momentum interaction ($M \sim 0.49-0.52$) with respect to corresponding adiabatic conditions has been required to obtain agreement with field measurements, and as previously in the comparison with Rouillard (1985) experimental data (chapter 3), this is attributed to the increase in frictional resistance associated with non-adiabatic conditions, and to the boiling instability associated with boiling under low-flow low-heat transfer low-pressure conditions.

Vortices have been predicted to occur in continuous vacuum pans in the upper-downtake region towards the corner of the calandria. The predicted vortices are considered physically reasonable due to the sharp turn that the geometry of the crystallizers imposes in the circulating massecuite at this location, and agree with results obtained in field measurements and at lab-scale. The vortices predicted could be detrimental for the circulation of massecuite and possibly related to the unsuccessful use of floating calandrias for batch crystallizers, where a generous downtake cannot be afforded as in the continuous units to keep the graining volume within a reasonable limit.

Numerical simulations of the flow in two batch crystallizers provided with a floating calandria and with a central-downtake have been performed, obtaining reasonable agreement from a qualitative point of view with flow patterns identified by Anon. (1964-5) experimentally by following a radioactive capsule immersed in the circulating massecuite.
The experiences gained from this study of the circulation in full-scale crystallizers suggest that physically sound predictions of the flow can be obtained with the developed gas-liquid momentum exchange model and CFD modeling strategy. The approach developed for analysis of massecuite circulation is valid for engineering purposes and can be applied in the study of the effect of the geometry on sugar evaporative crystallizers, keeping in mind that the heat transfer process within the calandria tubes is critical for the circulation and important interactions can be overlooked.
CHAPTER 6 - APPLICATION OF CFD TO IMPROVE THE DESIGN OF SUGAR EVAPORATIVE CRYSTALLIZERS

The strategy that has been developed in previous chapters for the numerical simulation of the fluid flow in sugar evaporative crystallizers is applied here to investigate the effect of different changes in the design of continuous units on the flow field and natural circulation, looking for potential alternatives for improvement. Based on the flow simulations recommendations for the design of vacuum pans are presented at the end, which might enhance the circulation, and therefore the efficiency and capacity, of the evaporative crystallizers used by the sugar industry.

6.1 CRITERIA FOR COMPARISON OF DESIGN ALTERNATIVES

The circulation of massecuite, defined here as the liquid downflow through the downtake, is chosen as the main criterion for evaluation in the intended analysis. As normal in convection processes, the heat transfer in sugar evapo-crystallizers improves as the circulation increases, and practical experience has demonstrated that a good circulation favors the crystallization rate, the capacity of the crystallizers, and the quality of the sugar crystals produced. The user of stirrers in batch crystallizers has illustrated that improvement in circulation, resulting from increased speed of agitation, leads to enhanced heat transfer (Bruhns et al., 2006). Similar trends have been reported from diverse experiences using mechanical agitators and steam jiggers.

The circulation period (τ) is defined as the ratio between the volume of massecuite in the pan and the circulation rate through the downtake, and is used here for comparison of different designs and situations involving different massecuite volumes.
The experimental results obtained in this study have indicated that to obtain a good circulation the transfer of momentum from the gas (vapor) to the liquid phase (massecuite) in the heated calandria tubes must be maximized, and the frictional resistance of the circuit should be minimized. However, there are constrains that are also important from the technical and practical point of view and are taken into account in the present analysis:

- Stagnation: It is undesirable to have stagnant or low-velocity regions, since this can result in crystal settling and differential growing rates of the grain. A minimum velocity of 10 mm/s is defined as critical value for comparison purposes.

- Accessibility: Enough space for inspection and maintenance is required, particularly at the bottom. Based on existing designs of sugar evaporative crystallizers it is assumed that a minimum distance of 100 mm (4 in) from the bottom tube plate is required for accessing and maintenance of the calandria.

- Discharge time: Particularly for batch crystallizers, the time that is required for discharging the massecuite at the end of the strike has some importance, since it contributes to the total duration of the cycle, and therefore the capacity of the crystallizers. Based on recommendations presented in the literature it is considered that the use of low slopes at the bottom must be avoided (e.g. must be $\theta \geq 17^\circ$).

- Practicability: Simplicity in the design and manufacturability are valuable features, since they can affect significantly the fabrication and maintenance costs. It is assumed here that complicated configurations or geometries should be avoided unless they result in a substantial improvement of performance.

The numerical analysis of the design of sugar evaporative-crystallizers is performed studying systematically the effect of diverse parameters on the circulation in an attempt to
identify optimum conditions that would lead to improved circulation. The design parameters considered are:

- Massecuite head – H: Liquid level above the top calandria plate.
- Length of the heated calandria tubes - L.
- Downtake size: The circulation ratio (C_R) relates the nominal upflow (calandria tubes) and downflow (downtake) areas, indicating the relative size of the downtake.
- Geometry of the calandria-downtake wall: Flat or rounded (bowed calandria).
- Inclination of the top tube calandria plate
- Inclination of the bottom tube calandria plate
- Bottom: Geometry and separation below the bottom calandria tube plate.

6.2 ANALYSIS OF THE DESIGN OF CONTINUOUS CRYSTALLIZERS

6.2.1 Effect of the Liquid Level above the Top Tube Plate in Continuous Crystallizers

Table 6.1 presents the nominal liquid head used in three horizontal continuous crystallizers currently available in the market, indicating that relatively low massecuite heads are normally used with respect to batch crystallizers, where at the end of the strike the massecuite level can reach values that are higher by several times.

Table 6.1 Liquid head in continuous sugar evaporative crystallizers.

<table>
<thead>
<tr>
<th>Design</th>
<th>H (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tongaat-Hulett</td>
<td>0.30</td>
</tr>
<tr>
<td>SRI</td>
<td>0.50-0.70</td>
</tr>
<tr>
<td>Honiron</td>
<td>0.30</td>
</tr>
</tbody>
</table>

As the liquid level above the top calandria plate rises, the hydrostatic pressure in the heated tubes increases, resulting in higher saturation temperature and reduction of the driving
temperature difference responsible for the heat transfer. This well known hydrostatic and thermodynamic effect makes very low liquid heads apparently attractive. However, the experiments performed at lab-scale (chapter 3) indicated that below certain critical height the flow resistance above the calandria becomes significant and causes poor circulation. It was observed that particularly for continuous crystallizers an optimum head must exist.

Rouillard (1985) studied the effect of several variables on the evaporation in calandria tubes, and based on experimental results presented a correlation that indicates an exponential reduction in evaporation as the liquid head increases:

\[
\ln(EC) = 15.92 - 0.165 \times Bx - 0.0601 \times P_{ABS} + 0.0311 \times Pty + 0.00639 \times P_{STEAM} - 0.321 \times L - 0.298 \times H
\]

Where:
- \( EC \): Evaporation rate (kg/h.m\(^2\))
- \( Bx \): Concentration (%Brix)
- \( P_{ABS} \): Absolute pressure (kPa abs)
- \( Pty \): Massecuite purity (%)
- \( P_{STEAM} \): Steam pressure (kPa abs)
- \( L \): Tube length (m)
- \( H \): Massecuite head above calandria (m)

For the flow simulations it is assumed that the correlation developed by Rouillard (1985) can be applied to predict the evaporation with the different liquid heads to be considered. The analysis is based on the geometry of the Tongaat-Hulett CVP studied previously and assuming the same operational conditions presented in chapter 5. During the simulations the liquid head was varied within rational limits, exhibiting a free surface located between \( H = 0.3-1.1 \) m above the top tube calandria plate\(^{24}\). Figure 6.1 presents the geometry of the crystallizer studied, illustrating the range of liquid levels that has been evaluated.

\(^{24}\) The correlation developed by Rouillard (1985) is based on experimental data corresponding to liquid heads \( H \sim 0.20-0.95 \) m
Table 6.2 presents the evaporation that has been estimated for different liquid heads based on the correlation developed by Rouillard (1985), and the corresponding circulation that has been predicted applying numerical flow simulation for each condition.

Figures 6.2 and 6.3 present numerical results on the flow field in the continuous evaporative crystallizer studied, showing an increasingly complex two-phase situation above the calandria as the liquid level increases. The velocity vectors presented in Fig 6.3 indicate comparatively low massecuite velocities and circulation for the cases representing the lowest liquid heads, this in spite of the higher evaporation that is considered (Table 6.2). This behavior is in agreement with the results obtained in the lab-scale model, and is attributed to the ‘bottle-neck’ effect or higher frictional resistance above the top tube plate with low liquid heads as a consequence of the reduced cross sectional area above the calandria.
Table 6.2  Effect of the level above the top tube plate on the circulation predicted numerically

<table>
<thead>
<tr>
<th>Liquid head H (m)</th>
<th>Evaporation (kg/h.m²)</th>
<th>Massecuite volume (m³)</th>
<th>Circulation JL (m/s)</th>
<th>Ratio JL / JL_MAX</th>
<th>Circulation period τ (s)</th>
<th>Ratio τ / τ_MIN</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.30</td>
<td>18.5</td>
<td>153</td>
<td>0.063</td>
<td>0.91</td>
<td>82</td>
<td>1.00</td>
</tr>
<tr>
<td>0.50</td>
<td>17.4</td>
<td>170</td>
<td>0.066</td>
<td>0.95</td>
<td>88</td>
<td>1.07</td>
</tr>
<tr>
<td>0.70</td>
<td>16.4</td>
<td>185</td>
<td>0.068</td>
<td>0.98</td>
<td>93</td>
<td>1.13</td>
</tr>
<tr>
<td>0.90</td>
<td>15.5</td>
<td>201</td>
<td>0.069</td>
<td>1.00</td>
<td>99</td>
<td>1.20</td>
</tr>
<tr>
<td>1.10</td>
<td>14.6</td>
<td>215</td>
<td>0.068</td>
<td>0.98</td>
<td>108</td>
<td>1.31</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes (J_L) and the circulation period (τ).

The numerical simulations indicate that the vortex generated in the upper downtake region grows as the liquid level increases and more room is available to extend its field. The increase of the liquid level results in a progressive displacement of the eye of the vortex upwards and increase of its strength, resulting in larger liquid velocities and undesirable massecuite recirculation as shown in Fig 6.3.

Fig 6.4 illustrates the predicted effect of the massecuite level on the circulation, indicating an initial increase in circulation in terms of the superficial liquid velocity (J_L) as the liquid head is higher, this in spite of the reduction in evaporation that has been estimated. The result suggests that the gain in circulation is significant up to H ~ 0.70 m, where the optimum head for the considered case is probably located. According to the numerical results an increase in circulation of +7.6 % can be expected from raising the liquid head above the top tube plate from H = 0.30 m to 0.70 m.
Fig 6.2 Contours of void fraction numerically predicted for a continuous sugar crystallizer with different liquid levels above the top tube plate.
Fig 6.3 Liquid velocity vectors numerically predicted for a continuous sugar evaporative crystallizer with different liquid levels above the top tube plate.
A larger volume of massecuite is logically obtained as the liquid level increases, and this results in higher circulation periods as shown in Figure 6.4, which indicates that the average time required for each fluid particle, and sugar crystal, to complete a circulation loop within the crystallizer is extended as the liquid level rises. Based on the flow simulations it can be proposed that the massecuite level above the top tube calandria plate should be designed between $H \approx 0.60-0.80$ m for A-boiling continuous evaporative crystallizers.

The predicted effect of the massecuite level on the circulation displays similarity with the trend that was determined from the experiments performed at lab-scale (chapter 3), which suggested an optimum liquid head around $H \approx 0.10$ m, corresponding to $H \approx 0.60$ m for full-scale crystallizers. In the same way, the optimum level predicted is comparable with results reported by Austmeyer (1986) from tests performed in full-scale stirred batch crystallizers, where it was determined that a maximum in the heat transfer coefficient occurs when the liquid level is around $H \approx 0.80$ m for after-product boiling, and $H \approx 0.15–0.65$ m for white sugar boiling.
6.2.2 Effect of the Geometry of Calandria Tubes in Continuous Crystallizers

The heated calandria tubes used in sugar continuous crystallizers are normally between \( D \approx 0.09-0.13 \text{ m} \) (3 ½"-5") in diameter, a range that has shown in practice a good compromise between the frictional resistance and the heat transfer. Smaller diameters display high frictional resistance and impair the circulation, while larger tubes present a low area per volume ratio, affecting the heat transfer effectiveness of the calandria and increasing the footing volume in batch applications.

Rouillard (1985) and Austmeyer and Schliephake (1983) used experimental facilities consisting of single heated tubes to study the heat transfer phenomena in calandria tubes, but considered only one pipe size in both cases (\( D \approx 0.10 \text{ m} \)), and therefore the effect of the tube diameter on the boiling heat transfer was not established. Bruhns (1986) presented experimental results on the heat transfer in calandria tubes with different diameters, but without boiling, reporting data that shows similarity with correlations for single-phase convective heat transfer. It can be said then that the effect of the diameter of the calandria tubes on the boiling heat transfer and circulation has not been studied experimentally, this in spite of being a parameter of paramount importance for the design of vacuum pans.

Considering the complexity of the interactions involved in the convective boiling phenomena occurring in calandria tubes and the relatively narrow range of tube diameters that are used in practice, it is assumed for this analysis that tubes with 0.10 m (4 in) in diameter are adequate. However, this cannot be considered conclusive and a different optimum size can exist, which probably would range from smaller diameters for low-viscosity massecuities and the first cells of continuous crystallizers, to larger diameters for high-viscosity massecuities and the last cells of continuous crystallizers.
• **Length of the Calandria Tubes**

It has been established that short calandria tubes display higher heat transfer than long tubes, although no rational explanation for this effect has been proposed. Results obtained by Tongaat-Hulett in tests performed with a test rig provided with tubes of different length (L = 0.6 - 1.0 - 1.4 - 1.8 m) demonstrated the benefit of using short tubes.

Modern batch crystallizers are provided with calandria tubes around L ~ 0.6-1.0 m in length, while in the past longer tubes were normally used. Table 6.3 presents the main dimensions of the calandria tubes currently used for continuous crystallizers, illustrating that relatively long tubes are normally used with respect to batch applications.

<table>
<thead>
<tr>
<th>Design</th>
<th>Tube length L (m)</th>
<th>Tube size OD (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Tongaat-Hulett</td>
<td>1.20 – 1.50</td>
<td>0.10</td>
</tr>
<tr>
<td>SRI</td>
<td>1.20 – 1.80</td>
<td>0.10 – 0.13</td>
</tr>
<tr>
<td>Honiron</td>
<td>1.40 – 1.70</td>
<td>0.09 – 0.10</td>
</tr>
</tbody>
</table>

The correlation proposed by Rouillard (1985) indicates an exponential reduction in evaporation as the length of the calandria tubes increases, which has permitted an estimation of the evaporation rate with different tube lengths for the present study. Table 6.4 presents the evaporation estimated with tubes of different length and the corresponding circulation that has been numerically predicted. The number of tubes required to obtain the same total evaporation and the corresponding length of the calandria are also presented, showing that for the same duty the foot print of the continuous crystallizer would need to be increased as the tubes are shortened.
Table 6.4  Evaporation and circulation numerically predicted for a continuous sugar evaporative crystallizer provided with calandria tubes of different length

<table>
<thead>
<tr>
<th>Tube length L (m)</th>
<th>Evaporation (kg/h.m²)</th>
<th>No. tubes</th>
<th>Calandria length (m)</th>
<th>Circulation JL (m/s)</th>
<th>Ratio JL / JL_MAX</th>
<th>Massecuite volume V (m³)</th>
<th>Circulation period τ (s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.6</td>
<td>23.9</td>
<td>6952</td>
<td>29.3</td>
<td>0.0504</td>
<td>0.70</td>
<td>194.2</td>
<td>73</td>
</tr>
<tr>
<td>0.8</td>
<td>22.4</td>
<td>5560</td>
<td>23.4</td>
<td>0.0562</td>
<td>0.78</td>
<td>172.1</td>
<td>72</td>
</tr>
<tr>
<td>1.0</td>
<td>21.0</td>
<td>4743</td>
<td>20.0</td>
<td>0.0605</td>
<td>0.84</td>
<td>160.0</td>
<td>73</td>
</tr>
<tr>
<td>1.2</td>
<td>19.7</td>
<td>4215</td>
<td>17.7</td>
<td>0.0639</td>
<td>0.88</td>
<td>155.5</td>
<td>76</td>
</tr>
<tr>
<td>1.4</td>
<td>18.5</td>
<td>3852</td>
<td>16.2</td>
<td>0.0667</td>
<td>0.92</td>
<td>153.1</td>
<td>78</td>
</tr>
<tr>
<td>1.8</td>
<td>16.3</td>
<td>3406</td>
<td>14.3</td>
<td>0.0723</td>
<td>1.00</td>
<td>153.3</td>
<td>82</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes (JL) and the circulation period (τ).

Figures 6.5 and 6.6 present numerical results on the two-phase flow in the continuous crystallizer studied for tubes of different length, showing higher liquid velocities and higher void fractions at the upper part of the tubes as the length increases, this as a natural consequence of the larger evaporation per tube.

Figure 6.7 illustrates the predicted effect of the length of the calandria tubes on the circulation, showing an increase in circulation in terms of the superficial liquid velocity (JL) as the tubes are longer. However, the trend displays a non-linear behavior that indicates that the gains in circulation are progressively reduced as the length of the tubes increases, suggesting that they should be ideally as short as practically possible. The circulation period predicted indicates an optimum in tube length around L ~ 1.0 m as can be appreciated in Fig 6.7. Further reductions in the length of the tubes would not offer a significant advantage and would lead to a considerably larger vessel.
Fig 6.5 Contours of void fraction numerically predicted for a continuous sugar evaporative crystallizer provided with calandria tubes of different length.
Fig 6.6 Liquid velocity vectors numerically predicted for a continuous sugar evaporative crystallizer provided with calandria tubes of different length.
Fig 6.7 Effect of the length of the calandria tubes on the circulation in a continuous sugar evaporative crystallizer predicted numerically.

According to the numerical results presented, reducing the length of the tubes from \( L = 1.4 \) to 1.0 m would make necessary to extend the length of the calandria by +23 % (3.75 m), still remaining within reasonable limits (20 m), and would increase the evaporation per unit area in +13.7 % and improve in +6 % the circulation period. This reduction in length of the tubes would result in 891 additional tubes being required, but the total heat exchange area would be actually reduced.
Fig 6.8 Estimated effect of the length of the calandria tubes on the length of the calandria of a continuous evaporative crystallizer.

The experimental results on the transfer of momentum in gas-liquid vertical channel flows that were presented in chapter 4 indicated that the interfacial interaction reduces as the gas content and the void fraction increase, resulting in a progressive loss in the capacity of the gas to transfer momentum to the liquid and produce circulation as the gas flux is higher. This experimental result indicates that the high gas content that is predicted at the top of long calandria tubes (Fig 6.5) affects the interfacial momentum interaction, causing poor circulation and performance in pans provided with long tubes. In the same way, the higher void fraction associated with longer tubes would reduce the heat transfer efficiency, increasing the severity of boiling instabilities and frictional resistance.

Figure 6.8 presents the length of the pan or the calandria that would be required to achieve the same total evaporation rate with tubes of different length, illustrating a rapid increase in the size of the vessel as the tubes are shortened. While the numerical analysis has suggested that the tubes should be as short as possible to optimize the heat transfer and circulation rate, from the practical point of view the dimensions of the pan must be kept
within reasonable limits, and the use of very short tubes would result in a vessel unreasonably large (e.g. 29.3 m for $L = 0.6$ m). Tubes with length between $L = 1.0$-$1.2$ m seem to offer a reasonable compromise between the circulation (Fig 6.7) and the size of the pan (Fig 6.8); the required length of the calandria grows rapidly when shorter tubes are considered, while longer tubes would reduce only by a small amount the size of the calandria and display lower heat transfer.

6.2.3 Effect of the Downtake Size or Circulation Ratio\(^{25}\) in Continuous Crystallizers

Sugar continuous evaporative crystallizers are normally provided with a relatively large down-comer or downtake channel, resulting in low circulation ratios (e.g. $C_R \sim 0.9$), well below the minimum value recommended for the design of batch evaporative crystallizers ($C_R \leq 2.5$). This is possible because of the continuous character of the process, where having a small footing volume is not required as in batch crystallizers.

Figures 6.9 and 6.10 present numerical results on the two-phase flow in a continuous sugar crystallizer provided with a downtake channel varying between 0.2-2.0 m in width. The downtake of the commercial case studied is $\sim 1.0$ m, corresponding to a circulation ratio $C_R \sim 0.90$

The velocity vectors presented in Figure 6.10 show higher velocities within the downtake channel as its size is decreased, this as a logical consequence of the corresponding reduction in cross sectional area. On the other hand, the vortex that is developed at the top of the downtake is predicted to become larger as the size of the downtake increases, causing undesirable massecuite recirculation and a low velocity region towards the calandria-downtake wall.

---

\(^{25}\) Circulation ratio ($C_R$) = Ratio between the total cross sectional area of the calandria tubes and the downtake, indicating the proportion between the upflow / downflow areas
Fig 6.9 Contours of void fraction numerically predicted for a continuous sugar evaporative crystallizer with different downtake size - circulation ratios
Fig 6.10 Liquid velocity vectors numerically predicted for a continuous sugar evaporative crystallizer with different downtake size - circulation ratios
Table 6.5 presents the circulation that has been numerically predicted for the continuous crystallizer studied considering different sizes of the downtake channel or circulation ratio. The corresponding increase in massecuite volume is also presented.

Table 6.5  Circulation numerically predicted for a continuous sugar evaporative crystallizer with different downtake size - circulation ratios.

<table>
<thead>
<tr>
<th>Downtake size (m)</th>
<th>Circulation ratio CR</th>
<th>Massecuite volume V (m³)</th>
<th>Circulation JL (m/s)</th>
<th>Ratio JL / JL_MAX</th>
<th>Circulation period τ (s)</th>
<th>Ratio τ / τ_MIN</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>--</td>
<td>--</td>
<td>0.0000</td>
<td>--</td>
<td>--</td>
<td>--</td>
</tr>
<tr>
<td>0.20</td>
<td>4.516</td>
<td>106.29</td>
<td>0.0520</td>
<td>0.76</td>
<td>70</td>
<td>1.14</td>
</tr>
<tr>
<td>0.25</td>
<td>3.612</td>
<td>110.07</td>
<td>0.0606</td>
<td>0.89</td>
<td>62</td>
<td>1.01</td>
</tr>
<tr>
<td>0.35</td>
<td>2.580</td>
<td>117.62</td>
<td>0.0655</td>
<td>0.96</td>
<td>61</td>
<td>1.00</td>
</tr>
<tr>
<td>0.50</td>
<td>1.806</td>
<td>127.42</td>
<td>0.0675</td>
<td>0.99</td>
<td>64</td>
<td>1.05</td>
</tr>
<tr>
<td>0.75</td>
<td>1.204</td>
<td>146.30</td>
<td>0.0684</td>
<td>1.00</td>
<td>73</td>
<td>1.19</td>
</tr>
<tr>
<td>1.00</td>
<td>0.903</td>
<td>165.21</td>
<td>0.0682</td>
<td>1.00</td>
<td>83</td>
<td>1.35</td>
</tr>
<tr>
<td>1.25</td>
<td>0.722</td>
<td>184.06</td>
<td>0.0677</td>
<td>0.99</td>
<td>93</td>
<td>1.51</td>
</tr>
<tr>
<td>1.50</td>
<td>0.602</td>
<td>203.05</td>
<td>0.0669</td>
<td>0.98</td>
<td>104</td>
<td>1.69</td>
</tr>
<tr>
<td>2.00</td>
<td>0.452</td>
<td>240.70</td>
<td>0.0662</td>
<td>0.97</td>
<td>124</td>
<td>2.02</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes (JL) and the circulation period (τ).

Figure 6.11 presents the circulation that has been predicted with respect to the size of the downtake, showing an asymptotic behavior that suggests a critical value around ~ 0.50 m. The use of a downtake channel below this critical value, corresponding to a circulation ratio above CR ≥ 1.8, is predicted to be detrimental for the circulation. On the other hand, the use of a downtake larger than ~ 1.25 m seems to be unnecessary and probably inconvenient as no further gain in circulation is observed, while undesirable recirculation and low velocity regions are generated as illustrated in Figure 6.10.
Fig 6.11 Numerically predicted effect of [a-top] the downtake size and [b-bottom] the circulation ratio on the circulation in a continuous sugar evaporative crystallizer

Figure 6.12 presents the circulation period that has been predicted with respect to the circulation ratio, showing an optimum value around $C_R \sim 2.58$. The minimum circulation period predicted corresponds to a downtake of $\sim 0.35$ m in width, considerably narrower than the values used in practice for sugar continuous crystallizers. However, it is interesting to notice that the predicted optimum size of the downtake is close to the design
limit for batch sugar crystallizers, where it is recommended to keep the circulation ratio below \( C_R \leq 2.50 \).

![Fig 6.12 Numerically predicted effect of the circulation ratio on the circulation period in a continuous sugar evaporative crystallizer.](image)

Although the optimum circulation ratio observed in Figure 6.12 is predicted to optimize circulation period \((C_R \sim 2.5 - 0.35 \text{ m})\), it also would reduce the circulation rate by \(\sim 4\%\) as indicated in Table 6.5. Taking into account that for continuous crystallizers having a small downtake is not a constraint, it is considered here that reducing the downtake size to the predicted optimum value is unnecessary and inconvenient.

Based on the numerical results that have been presented it is proposed that the continuous crystallizer studied should be constructed with a downtake around \(\sim 0.75-1.0 \text{ m}\), corresponding to circulation ratios between \(C_R \sim 0.9-1.2\), which would result in maximum circulation and prevent an undesirable expansion of the vortex developed at the top of the downtake.
6.2.4 Effect of Rounding the Calandria-downtake Wall in Continuous Crystallizers

In sugar evaporative crystallizers a vertical metallic wall is used to separate the massecuite within the downtake and the heating steam within the calandria. This wall is normally rounded (bowed calandria) in continuous crystallizers in an attempt to favor the circulation by making smoother the changes in direction of the flowing massecuite.

Figure 6.13 presents numerical results on the flow in a continuous sugar crystallizer provided with a flat and a rounded calandria-downtake wall, showing minor differences between the flow fields predicted. The use of the flat wall is predicted to give a circulation slightly higher (+0.2 %), and this is explained by the larger downtake cross sectional area obtained in this case. However, this also results in a larger volume of massecuite that affects the circulation period.

Table 6.6  Circulation numerically predicted for a continuous sugar evaporative crystallizer provided with a flat and a rounded calandria-downtake wall.

<table>
<thead>
<tr>
<th>Calandria - downtake wall</th>
<th>Circulation $J_L$ (m/s)</th>
<th>Ratio $J_L / J_{L_{MAX}}$</th>
<th>Pan volume (m$^3$)</th>
<th>Circulation period - $\tau$ (s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flat</td>
<td>0.0644</td>
<td>1.000</td>
<td>161.87</td>
<td>85.8</td>
</tr>
<tr>
<td>Rounded (R 1.2m)</td>
<td>0.0643</td>
<td>0.998</td>
<td>153.16</td>
<td>81.4</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes ($J_L$) and the circulation period ($\tau$).

The numerically predicted liquid velocity vectors presented in Figure 6.13 show a low velocity region within the downtake towards the calandria wall, which is predicted to be less pronounced when a rounded calandria-downtake wall is used. Because of this control of the size of the low velocity region, and considering that the effect on the circulation rate is almost negligible, it can be said that a rounded calandria-downtake wall is a desirable feature for the design of continuous sugar evaporative crystallizers.
<table>
<thead>
<tr>
<th>a. Flat Calandria-downtake Wall</th>
<th>b. Rounded Calandria-downtake Wall</th>
</tr>
</thead>
<tbody>
<tr>
<td><img src="image1" alt="Flat Calandria-downtake Wall" /></td>
<td><img src="image2" alt="Rounded Calandria-downtake Wall" /></td>
</tr>
<tr>
<td><img src="image3" alt="Flat Calandria-downtake Wall" /></td>
<td><img src="image4" alt="Rounded Calandria-downtake Wall" /></td>
</tr>
<tr>
<td><img src="image5" alt="Flat Calandria-downtake Wall" /></td>
<td><img src="image6" alt="Rounded Calandria-downtake Wall" /></td>
</tr>
</tbody>
</table>

Fig 6.13  Contours of void fraction, liquid velocity vectors, and low velocity areas numerically predicted for a continuous sugar evaporative crystallizer provided with [a-left] flat and [b-right] rounded calandria-downtake wall.
6.2.5 Effect of Inclining the Top Tube Calandria Plate in Continuous Crystallizers

Inclined or sloped calandria tube plates are used sometimes in batch and continuous sugar evaporative crystallizers in the belief that it improves the circulation. In these cases the tube plates are angled usually between $\alpha \sim 10$-25 °. There is no agreement with respect to the convenience of using inclined tube plates, and today most of the pans are built using horizontal plates, probably because of the simpler construction and maintenance, and therefore lower costs.

Figures 6.14 and 6.15 present numerical results on the two-phase flow in a continuous crystallizer with different inclinations of the top calandria tube plate, covering a range of slopes within rational limits ($\alpha = 0$-30 °). The velocity vectors presented in Figure 6.15 indicate that the vortex developed at the upper downtake tends to grow in size and strength as the inclination of the top tube plate increases, resulting in significant liquid recirculation.

Table 6.7 presents the circulation that has been numerically predicted for different inclinations of the top tube plate, suggesting that inclining the plates up to $\alpha \sim 10$ ° has a minimum effect on the circulation, while further increases would become rapidly detrimental as illustrated in Figure 6.16.

Table 6.7 Circulation predicted numerically for a continuous sugar evaporative crystallizer with different inclinations of the top calandria plate.

<table>
<thead>
<tr>
<th>Inclination top calandria plate (deg)</th>
<th>Circulation $J_L$ (m/s)</th>
<th>Ratio $J_L / J_{L,\text{MAX}}$</th>
<th>Circulation period $\tau$ (s)</th>
<th>Ratio $\tau / \tau_{\text{MIN}}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0.0780</td>
<td>1.00</td>
<td>81.8</td>
<td>1.00</td>
</tr>
<tr>
<td>5</td>
<td>0.0781</td>
<td>1.00</td>
<td>81.7</td>
<td>1.00</td>
</tr>
<tr>
<td>10</td>
<td>0.0780</td>
<td>1.00</td>
<td>81.8</td>
<td>1.00</td>
</tr>
<tr>
<td>15</td>
<td>0.0757</td>
<td>0.97</td>
<td>84.4</td>
<td>1.03</td>
</tr>
<tr>
<td>20</td>
<td>0.0702</td>
<td>0.90</td>
<td>90.9</td>
<td>1.11</td>
</tr>
<tr>
<td>30</td>
<td>0.0667</td>
<td>0.85</td>
<td>95.7</td>
<td>1.17</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes ($J_L$) and the circulation period ($\tau$).
Fig 6.14 Contours of void fraction numerically predicted for a continuous sugar evaporative crystallizer with different inclinations of the top calandria plate.
Fig 6.15 Liquid velocity vectors numerically predicted for a continuous sugar evaporative crystallizer with different inclinations of the top calandria plate.
Fig 6.16 Effect of the inclination of the top calandria tube plate on the circulation in a continuous sugar evaporative crystallizer predicted numerically

The numerical results have suggested that inclining the top tube plate in continuous sugar evaporative crystallizers is unnecessary and does not lead to any significant gain in circulation. The use of slopes above $\theta \sim 10^\circ$ is predicted to be unfavorable, this probably due to an increase in recirculation above the calandria (Fig 6.15) and to the implicit use of longer tubes, where the transfer of momentum between the vapor and massecuite tends to be less effective. Based on the flow simulations it is proposed that the top tube plates of continuous sugar evaporative crystallizers should be horizontal.

6.2.6 Effect of Inclining the Bottom Tube Calandria Plate in Continuous Crystallizers

Figures 6.17 and 6.18 present numerical results on the flow in a continuous sugar crystallizer with different inclinations of the bottom calandria tube plate ($\theta = 0-30^\circ$). In this case the geometry of the shell had to be adjusted for the different inclinations, keeping constant the volume of the bottom section.
Fig 6.17 Contours of void fraction numerically predicted for a continuous sugar crystallizer with different inclinations of the bottom calandria tube plate.
Fig 6.18 Liquid velocity vectors numerically predicted for a continuous sugar crystallizer with different inclinations of the bottom calandria plate.
Table 6.8 presents the circulation rate that has been numerically predicted for the different inclinations of the bottom tube plate evaluated, suggesting that inclining the plates up to $\theta \sim 15^\circ$ does not affect the circulation significantly, and that further increases would be slightly detrimental as illustrated in Figure 6.19. Based on the flow simulations it is considered that the bottom tube plates of continuous sugar crystallizers should be horizontal.

Table 6.8  Circulation predicted numerically for a continuous sugar evaporative crystallizer with different inclinations of the bottom calandria plate.

<table>
<thead>
<tr>
<th>Inclination bottom plate (deg)</th>
<th>Circulation $J_L$ (m/s)</th>
<th>Ratio $J_L / J_{L,\text{MAX}}$</th>
<th>Circulation period $\tau$ (s)</th>
<th>Ratio $\tau / \tau_{\text{MIN}}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0.0691</td>
<td>1.00</td>
<td>79.1</td>
<td>1.00</td>
</tr>
<tr>
<td>10</td>
<td>0.0690</td>
<td>1.00</td>
<td>79.2</td>
<td>1.00</td>
</tr>
<tr>
<td>15</td>
<td>0.0689</td>
<td>1.00</td>
<td>79.3</td>
<td>1.00</td>
</tr>
<tr>
<td>20</td>
<td>0.0683</td>
<td>0.99</td>
<td>79.9</td>
<td>1.01</td>
</tr>
<tr>
<td>25</td>
<td>0.0678</td>
<td>0.98</td>
<td>80.6</td>
<td>1.02</td>
</tr>
<tr>
<td>30</td>
<td>0.0674</td>
<td>0.98</td>
<td>81.1</td>
<td>1.02</td>
</tr>
</tbody>
</table>

NOTE - The circulation is presented in terms of the liquid superficial velocity or flux in the calandria tubes ($J_L$) and the circulation period ($\tau$).

Fig 6.19  Effect of the inclination of the bottom calandria plate on the circulation in a continuous sugar evaporative crystallizer predicted numerically.
6.2.7 Effect of the Geometry of the Bottom in Continuous Crystallizers

Considering that only massecuite is present at the bottom of sugar evaporative crystallizers and the elevated computational expense of multiphase CFD modeling, it was decided to study the effect of the geometry of the bottom applying single-phase CFD modeling. For the numerical analysis the geometry of the continuous crystallizer has been simplified restricting the computational domain to represent only the bottom and the downtake. Figure 6.20 presents the section of the crystallizer studied and the boundary conditions assumed. A mass inlet boundary condition is used to introduce the circulating massecuite above the top tube calandria, while outflow boundary conditions are set in multiple locations representing the entrance of the calandria tubes. A frictionless wall is used at the top to represent the free surface. The boundary conditions are set to represent the circulation measured in a full-scale CVP that is presented in chapter 5 ($J_L \sim 0.065$ m/s).

![Figure 6.20 Geometric simplification of continuous sugar evaporative crystallizers for analysis of the effect of the bottom geometry applying single-phase CFD modeling](image)

To compare different alternatives of the bottom, the pressure drop between the inlet and outflow boundary conditions is computed from the numerical flow simulations, which corresponds to the frictional resistance of the path conformed by the downtake and the
bottom. Attention is also given to areas of low velocity or stagnation, which are undesirable for the crystallization process.

Different alternatives for the design of the bottom of continuous crystallizers have been evaluated numerically, changing systematically the openings or separations between the bottom tube plate and the bottom wall below the corner (L_{CB}) and the middle (S) of the calandria, parameters that are shown in Fig 6.20

- **Separation between the Bottom Tube Plate and the Bottom Wall - L_{CB}**

A simple rectangular geometry of the bottom has been considered initially, where the separation between the bottom tube plate and the bottom wall was defined based on the optimum size of the downtake determined earlier (Fig 6.11 – L_{CB} = 0.75 m). Figure 6.21 presents numerical results on the flow within the rectangular bottom, showing large stagnation areas at the corners that indicate that using this geometry would be inappropriate for the crystallization process. However, the streamlines obtained for this case delineate a geometry that might help to identify a convenient shape for the bottom.

![Fig 6.21 Streamlines and low velocity areas numerically predicted for a continuous sugar crystallizer provided with a rectangular bottom.](image)

In view of the flow patterns observed in Figure 6.21, it was decided to consider an ‘elliptical’ bottom, which has been defined based on two concentric ellipses centered in the outer corner of the calandria, one below the downtake and one below the bottom tube plate. The elliptical geometry guarantees tangency automatically at the intersections on the edges.
Different openings between the bottom tube plate and the bottom wall have been evaluated, covering a range between $L_{CB} = 0.25-1.00$ m. Figure 6.22 presents numerical results on the flow within the elliptical bottoms considered, showing smooth changes in the flow direction and no significant stagnation points. The streamlines indicate accelerations and higher mass sucuie velocity as the size of the bottom is reduced, resulting in larger velocity gradients towards the bottom of the vessel.

Fig 6.22 Streamlines and low velocity areas numerically predicted for a continuous crystallizer with elliptical bottom and different separations between the bottom tube plate and the bottom wall - $L_{CB}$
Table 6.9 presents the frictional pressure drop that has been numerically predicted for different openings between the bottom tube plate and the elliptical bottom. The simulations indicate lower frictional resistance as the size of the bottom, and therefore the cross sectional area, are increased. Based on the numerical results it can be proposed that the separation of the bottom must be $L_{CB} \geq 0.65$. Reducing this opening below this value is predicted to be detrimental for the circulation due to a rapid rise in frictional resistance that is illustrated in Figure 6.23.

Table 6.9 Frictional pressure drop predicted numerically for a continuous sugar crystallizer with different separations between the bottom tube plate and the bottom wall - $L_{CB}$

<table>
<thead>
<tr>
<th>$L_{CB}$ (m)</th>
<th>$\Delta P_r$ (kPa)</th>
<th>$\Delta P_r / \Delta P_{r \text{ MIN}}$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.250</td>
<td>1.486</td>
<td>1.224</td>
</tr>
<tr>
<td>0.350</td>
<td>1.316</td>
<td>1.084</td>
</tr>
<tr>
<td>0.500</td>
<td>1.248</td>
<td>1.028</td>
</tr>
<tr>
<td>0.600</td>
<td>1.232</td>
<td>1.015</td>
</tr>
<tr>
<td>0.675</td>
<td>1.226</td>
<td>1.009</td>
</tr>
<tr>
<td>0.750</td>
<td>1.221</td>
<td>1.006</td>
</tr>
<tr>
<td>1.000</td>
<td>1.214</td>
<td>1.000</td>
</tr>
</tbody>
</table>

Fig 6.23 Effect of the separation between the bottom tube plate and the bottom wall - $L_{CB}$ - on the frictional pressure drop in a continuous sugar crystallizer predicted numerically
• Separation between the Bottom Tube Plate and the Bottom Wall below the Middle of the Calandria – S

The separation between the bottom tube plate and the bottom wall below the middle of the calandria (S) is the narrowest opening. The effect of the separation S is studied here defining a geometry based on an elliptical shape below the downtake and a straight line below the bottom tube plate. Based on existing sugar evaporative crystallizers it is assumed that a minimum distance of $S \sim 0.10$ m (4 in) from the bottom tube plate is required for access and maintenance of the calandria. Figure 6.24 presents some of the geometries that have been considered, where the separation $S$ has been adjusted between $S = 0.10$-$0.40$ m, while a constant value of the separation $L_{CB} = 0.65$ m is maintained.

![Fig 6.24 Schematic illustrating bottom geometries with different separation between the bottom tube plate and the bottom wall below the middle of the calandria - S](image)

Figure 6.25 presents numerical results on the flow within the bottom shapes evaluated, showing acceleration and less stagnation as the opening between the bottom tube plate and the bottom wall (S) reduces. The velocity gradients near the bottom wall increase as the separation S is narrower, suggesting higher frictional resistance.
Table 6.10 presents the frictional pressure drop that has been numerically predicted for different separations between the bottom tube plate and the bottom wall, indicating a relatively small effect of this parameter on the circulation. Based on the numerical results it is proposed that the separation S should be designed around S ~ 0.20-0.25 m, where some inflection in the frictional resistance has been predicted as can be appreciated in Figure 6.26. Increasing the opening above this value might be slightly beneficial for the circulation, but is inconvenient due to the consequent enlargement of the low velocity region (Fig 6.25).
Table 6.10  Frictional pressure drop numerically predicted for a continuous sugar evaporative crystallizer with different separations between the bottom tube plate and the bottom wall - S

<table>
<thead>
<tr>
<th>S (m)</th>
<th>ΔPfr (kPa)</th>
<th>ΔPfr / ΔPfr MIN</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.10</td>
<td>1.249</td>
<td>1.016</td>
</tr>
<tr>
<td>0.15</td>
<td>1.242</td>
<td>1.011</td>
</tr>
<tr>
<td>0.20</td>
<td>1.238</td>
<td>1.007</td>
</tr>
<tr>
<td>0.25</td>
<td>1.235</td>
<td>1.005</td>
</tr>
<tr>
<td>0.30</td>
<td>1.233</td>
<td>1.003</td>
</tr>
<tr>
<td>0.40</td>
<td>1.229</td>
<td>1.000</td>
</tr>
</tbody>
</table>

Fig 6.26  Effect of the separation between the bottom calandria plate and the bottom wall S on the frictional pressure drop in a continuous sugar crystallizer predicted numerically

- **Practical Aspects Related with the Bottom of Continuous Crystallizers**

  The Fletcher Smith continuous sugar crystallizer used as the study case has a bottom geometry based on three circles as illustrated in Figure 6.27 (Case 1). The preference of a circular shape is associated with the manufacturing process, where it is easier to determine the mechanical procedure required to bend the steel and achieve the desired geometry than with other geometrical shapes. The smallest circle, in the middle of the bottom, defines a rounding that eases the fabrication of the crystallizer, but causing undesirable stagnation as illustrated in Fig 6.27.a
Figure 6.27  Geometry and numerical results on the flow in the bottom of [a-left] a commercial sugar continuous crystallizer and [b-right] a similar continuous crystallizer with simpler bottom geometry

Figure 6.27.b presents numerical results on the flow in a continuous crystallizer with a simplified geometry of the bottom (case 2), which has been defined using the same openings found in the Fletcher Smith continuous crystallizer (S ~ 0.30 m; L_{CB} ~ 0.80 m), but based on two circles and a straight geometry below the calandria, eliminating the rounding of massecuite velocity below 10 mm/s.

<table>
<thead>
<tr>
<th>Case 1</th>
<th>Case 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>S = 0.32 m</td>
<td>S = 0.30 m</td>
</tr>
<tr>
<td>L_{CB} = 0.80 m</td>
<td>L_{CB} = 0.80 m</td>
</tr>
</tbody>
</table>

ΔP_{fr} = 1.2200 kPa  
ΔP_{fr} / ΔP_{fr-1} = 1.0004

ΔP_{fr-1} = 1.2195 kPa  
ΔP_{fr} / ΔP_{fr-1} = 1.0000
the bottom at the middle ($R_C \sim 0$). An improved feature with respect to Case 1 is the decrease of the cross sectional area of the bottom across the calandria, which is assumed to be convenient due to the progressive removal of the massecuite from the bottom section as it enters to the calandria tubes.

The flow simulations indicate practically the same frictional pressure drop for the two cases presented in Figure 6.27. Therefore, the simplification of the geometry (case 2) does not have any detrimental effect on the circulation, which would be the same for a given pressure driving force. The numerical results presented in Figure 6.27 suggest that eliminating the rounding in the middle of the bottom reduces the size of the low velocity region at this point, but probably not enough.

A comparison of the different cases numerically studied shows that the frictional pressure drop predicted with different bottom geometries and the same opening ($L_{CB} \sim 0.75$ m) is similar\(^{26}\), suggesting that the shape of the bottom has small effect on the circulation rate as long as it is not undersized ($L_{CB} < 0.65$ m). On the other hand, the numerical results show that correctly designing the geometry of the bottom is important to prevent low velocity or stagnant regions.

Figure 6.28 presents numerical results on the flow in a continuous crystallizer with different bottom geometries, where the separations have been varied around the values that were determined as favorable in previous analyses ($S = 0.10-0.25$ m; $L_{CB} = 0.65-0.80$ m). The frictional pressure drop numerically predicted for the different bottom geometries is presented in Table 6.11, showing a logical increase in friction as the size of the bottom section decreases.

\(^{26}\) Rectangular bottom $\Delta P_f = 1.218$ kPa (Fig 6.24)
Elliptical bottom $\Delta P_f = 1.221$ kPa (Table 6.9)
Ellip + Straight bottom $\Delta P_f = 1.220$ kPa (Fig 6.30)
<table>
<thead>
<tr>
<th>Case 3</th>
<th>S = 0.25 m</th>
<th>L_{CB} = 0.75 m</th>
</tr>
</thead>
<tbody>
<tr>
<td>Case 4</td>
<td>S = 0.25 m</td>
<td>L_{CB} = 0.65 m</td>
</tr>
<tr>
<td>Case 5</td>
<td>S = 0.20 m</td>
<td>L_{CB} = 0.80 m</td>
</tr>
<tr>
<td>Case 6</td>
<td>S = 0.20 m</td>
<td>L_{CB} = 0.75 m</td>
</tr>
<tr>
<td>Case 7</td>
<td>S = 0.20 m</td>
<td>L_{CB} = 0.65 m</td>
</tr>
<tr>
<td>Case 8</td>
<td>S = 0.10 m</td>
<td>L_{CB} = 0.80 m</td>
</tr>
</tbody>
</table>

NOTE – Red area indicates velocity below 10 mm/s

Fig 6.28 Streamlines and low velocity areas numerically predicted for a continuous sugar crystallizer with different bottom geometries.
Table 6.11  Frictional pressure drop numerically predicted for a continuous sugar evaporative crystallizer with different bottom geometries.

<table>
<thead>
<tr>
<th>Case</th>
<th>S (m)</th>
<th>L_CB (m)</th>
<th>ΔP_{fr} (kPa)</th>
<th>ΔP_{fr} / ΔP_{fr-min}</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.32</td>
<td>0.80</td>
<td>1.2195</td>
<td>1.0000</td>
</tr>
<tr>
<td>2</td>
<td>0.30</td>
<td>0.80</td>
<td>1.2200</td>
<td>1.0004</td>
</tr>
<tr>
<td>3</td>
<td>0.25</td>
<td>0.75</td>
<td>1.2243</td>
<td>1.0039</td>
</tr>
<tr>
<td>4</td>
<td>0.25</td>
<td>0.65</td>
<td>1.2318</td>
<td>1.0101</td>
</tr>
<tr>
<td>5</td>
<td>0.20</td>
<td>0.80</td>
<td>1.2220</td>
<td>1.0020</td>
</tr>
<tr>
<td>6</td>
<td>0.20</td>
<td>0.75</td>
<td>1.2263</td>
<td>1.0056</td>
</tr>
<tr>
<td>7</td>
<td>0.20</td>
<td>0.65</td>
<td>1.2348</td>
<td>1.0125</td>
</tr>
<tr>
<td>8</td>
<td>0.10</td>
<td>0.80</td>
<td>1.2260</td>
<td>1.0053</td>
</tr>
<tr>
<td>9</td>
<td>0.10</td>
<td>0.75</td>
<td>1.2303</td>
<td>1.0088</td>
</tr>
<tr>
<td>10</td>
<td>0.10</td>
<td>0.65</td>
<td>1.2419</td>
<td>1.0184</td>
</tr>
</tbody>
</table>

The bottom geometry denoted as case 5 (S \sim 0.20 m; L_CB \sim 0.80 m) is predicted to give similar frictional resistance and less stagnation than the commercial continuous crystallizer used as study case (case 1). The stagnation region predicted at the middle of the bottom could be controlled further by rounding the corner as indicated in Figure 6.29.b, modification that would add minor complexity to the construction of the crystallizer.
Figure 6.29 Geometry and numerical results on the flow field in a continuous crystallizer with [a-left] a square and [b-right] a rounded corner of the bottom at the middle.

### 6.3 Alternatives for the Design of Continuous Sugar Evaporative Crystallizers

Considering the numerical results obtained on the effect of different design parameters of continuous sugar crystallizers on the circulation, two design alternatives that
are expected to display enhanced hydraulic characteristics are proposed here and evaluated numerically applying the developed multiphase CFD model.

- **Continuous Sugar Crystallizer with Long Calandria Tubes - L ~ 1.4 m**

A first alternative is considered assuming that the current length of the calandria tubes is to be maintained, corresponding to a conservative solution that involves low risk and potential of improving the circulation. Figure 6.30 presents the proposed design and the main geometrical parameters. Two bottom options using a square and a rounded corner at the middle are considered.

![Figure 6.30 Schematic representing a design alternative for continuous sugar crystallizers using long calandria tubes and [a-left] a square and [b-right] a rounded corner of the bottom at the middle](image)

Figure 6.31 presents the proposed geometry (red) and the commercial case studied (dotted black), illustrating two main design variations consisting in rising the liquid level from $H = 0.30$ m to 0.60 m to reduce the frictional resistance above the top tube calandria plate, and in redesigning the shape of the bottom to reduce stagnation. For comparison purposes, the same massecuite properties and operational conditions that were used in the analysis of the commercial sugar continuous crystallizer studied in chapter 5 are assumed.
Fig 6.31  Schematic comparing the proposed geometry using long calandria tubes with the current design of a commercial continuous sugar evaporative crystallizer.

Figure 6.32 presents numerical results on the two-phase flow in the proposed continuous sugar crystallizer, showing similarity from the qualitative point of view with the results obtained at lab-scale and full-scale presented in chapters 3 and 5 respectively.

Fig 6.32  Contours of void fraction and liquid velocity vectors numerically predicted for the proposed continuous sugar evaporative crystallizer using long tubes.
Figure 6.33 presents the flow predicted for the continuous crystallizer when the corner of the bottom below the middle of the calandria is square ($R_C \sim 0$ m) and rounded ($R_C \sim 0.15$ m), showing practically the same flow field in the two cases, except for a reduction of the stagnation area.

NOTE – The red area indicates massecuite velocity below 10 mm/s

Fig 6.33 Liquid velocity vectors and stagnation regions numerically predicted for the proposed continuous sugar evaporative crystallizer using long calandria tubes and [a-left] a square and [b-right] a rounded corner of the bottom at the middle.
Continuous Sugar Crystallizer with Shorter Calandria Tubes - L ~ 1.0 m

A second design alternative has been considered assuming that the length of the calandria tubes can be reduced from L = 1.4 m to 1.0 m to increase the heat transfer efficiency. This option involves more radical changes in the design, and can be considered risky, although potentially useful for enhancing the hydraulic and heat transfer performance of continuous sugar crystallizers.

Figure 6.34 presents the geometry and the main design parameters of the proposed continuous crystallizer with shorter tubes (red) and a comparison with the commercial design used as study case (dotted black). The main design variations are a reduction in length of the tubes from L ~ 1.4 m to 1.0 m to increase the heat transfer; an increase of the liquid level above the top tube plate from H = 0.30 m to 0.60 m to reduce the frictional resistance; an expansion of the calandria towards the sides (~ 0.18 m) and corresponding reduction of the downtake; and the redesign of the shape of the bottom to reduce stagnation. A longer calandria would be required for the projected crystallizer, increasing the length of the vessel from ~ 16.1 m to ~ 18 m.

Fig 6.34 Schematic representing [a-left] a design alternative for continuous crystallizers using short calandria tubes and [b-right] a comparison with a commercial design.
Figure 6.35 presents numerical results on the two-phase flow in the proposed continuous sugar crystallizer with shorter calandria tubes, showing well defined ‘vapor columns’ that suggest a simpler two-phase interaction above the top tube plate in comparison with the crystallizer provided with longer tubes (Fig 6.32). The liquid velocity vectors indicate smooth flow patterns and less massecuite recirculation above the calandria plate, suggesting that the variations in the design might be advantageous.
Table 6.12 presents the circulation that has been numerically predicted for the two design alternatives considered and the commercial continuous sugar crystallizer taken as study case. For the first alternative, with long calandria tubes, the flow simulations indicate an improvement in circulation around +10%. For the case with shorter tubes a direct comparison is difficult due to the significant changes in design, but the predictions indicate in general improvement; with the circulation period enhanced by -15%, and the circulation rate increased by +1% in terms of the liquid superficial velocity in the calandria tubes, and by +24% in terms of the total circulation flow.

<table>
<thead>
<tr>
<th>Study case</th>
<th>Design 1 long tubes</th>
<th>Design 2 short tubes</th>
</tr>
</thead>
<tbody>
<tr>
<td>Massecuite volume – V (m³)</td>
<td>160</td>
<td>174.4</td>
</tr>
<tr>
<td>Calandria / pan length (m)</td>
<td>16.14</td>
<td>16.14</td>
</tr>
<tr>
<td>Tube diameter - ID (m)</td>
<td>0.0984</td>
<td>0.0984</td>
</tr>
<tr>
<td>Tube length - L (m)</td>
<td>1.40</td>
<td>1.40</td>
</tr>
<tr>
<td>Number of tubes</td>
<td>3852</td>
<td>3852</td>
</tr>
<tr>
<td>Cross section - A_NS (m²)</td>
<td>29.3</td>
<td>29.3</td>
</tr>
<tr>
<td>Heat transfer area - A_HT (m²)</td>
<td>1667</td>
<td>1667</td>
</tr>
<tr>
<td>Ratio A_HT / V (m²/m³)</td>
<td>10.3</td>
<td>9.6</td>
</tr>
<tr>
<td>Width (m)</td>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>Circulation ratio - C_R</td>
<td>0.9</td>
<td>0.9</td>
</tr>
<tr>
<td>Separation - L_CB (m)</td>
<td>0.80</td>
<td>0.81</td>
</tr>
<tr>
<td>Separation - S (m)</td>
<td>0.32</td>
<td>0.2</td>
</tr>
<tr>
<td>Minimum angle (deg)</td>
<td>28</td>
<td>22</td>
</tr>
<tr>
<td>Shell base radius - R_S (m)</td>
<td>2.69</td>
<td>2.69</td>
</tr>
<tr>
<td>Bottom downtake radius - R_D (m)</td>
<td>1.28</td>
<td>0.81</td>
</tr>
<tr>
<td>Bottom corner radius - R_C (m)</td>
<td>0.45</td>
<td>0 / -0.15</td>
</tr>
<tr>
<td>Evaporation (kg/h.m²)</td>
<td>18.5</td>
<td>18.5</td>
</tr>
<tr>
<td>Total evaporation (kg/h)</td>
<td>30841</td>
<td>30841</td>
</tr>
<tr>
<td>Liquid flux heated tubes-J_L (m/s)</td>
<td>0.0667</td>
<td>0.0732</td>
</tr>
<tr>
<td>Ratio J_L / J_L - STUDY_CASE</td>
<td>1.000</td>
<td>1.098</td>
</tr>
<tr>
<td>Circulation period - T (s)</td>
<td>81.9</td>
<td>81.3</td>
</tr>
<tr>
<td>Ratio T / T - STUDY_CASE</td>
<td>1.000</td>
<td>0.993</td>
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<tr>
<td>Circulation rate (m³/s)</td>
<td>1.95</td>
<td>2.14</td>
</tr>
<tr>
<td>Ratio Circ / Circ - STUDY_CASE</td>
<td>1.000</td>
<td>1.098</td>
</tr>
</tbody>
</table>
CHAPTER 7 - CONCLUSIONS

A study of the circulation of massecuite in sugar evaporative crystallizers has been performed, involving experimental measurements in lab-scale models and full-scale crystallizers, and numerical simulations of the flow applying Computational Fluid Dynamics (CFD). The conclusions of the study are:

- **Circulation Studies Using a Lab-scale Model**

  The circulation mechanism and flow patterns in sugar evaporative-crystallizers have been studied experimentally in a lab-scale model using Particle Image Velocimetry (PIV), and numerically using CFD. The experiments showed development of vortices in the upper corner of the calandria and above the top tube plate towards the side wall. The circulation results indicated that too low liquid heads above the calandria would be detrimental for the circulation as they increase the frictional resistance. This suggests the existence in the design of continuous sugar evaporative crystallizers of an optimum massecuite height. The gas-liquid interaction determined in the lab-scale model have been compared with reported information on bubbly flows, observing agreement with data and drag correlations that represent the rise of bubbles in non-pure media.

  The flow in the lab-scale model has been simulated with the Eulerian-Eulerian CFD approach applying different drag correlations to model the interfacial momentum interaction. Agreement between measurements and computations was obtained when a correlation that represents correctly the exchange of momentum between the two participating phases was introduced (Lain et al., 1999). This result has shown the importance of selecting an adequate model to compute the exchange of momentum between the gas and liquid phase, which plays a particularly critical role in the buoyancy-driven flow studied, largely determining the
magnitude of the velocity field and circulation rate predicted. The experience with the lab-scale model suggested that if the exchange of momentum between the gas and liquid phases is solved properly, an acceptably accurate solution of the entire flow-field within sugar evaporative crystallizers will be feasible.

- **Momentum Interaction in Gas-liquid Buoyancy-driven Vertical Channel Flow**

  A large number of correlations for the calculation of the drag coefficient of bubbles has been proposed from experimental and analytical studies during the last ~150 years. The high number of correlations instead of a wide knowledge could be reflecting a lack of understanding about the physics behind the gas-liquid interaction and all the relevant factors, particularly as the properties of the media change, the bubble size increases, the bubble shape becomes distorted, and wake interactions with adjacent bubbles take place. Dependable drag correlations have been developed for small single bubbles rising in stagnant media, which can be applied with confidence for the CFD analysis of dilute bubbly flows, but are not applicable in more complex situations such as high void fraction bubbly flows, swarms of bubbles, slugs, and churns, where less information has been obtained and no reliable procedure for the calculation of the interfacial interactions is available.

  The interfacial transfer of momentum in buoyancy-driven gas-liquid vertical channel flows has been studied in an experimental facility, which essentially comprises a circulation loop, similar in concept to an air-lift reactor, where air is injected at the bottom of a riser section and the buoyancy force due to density differences between the gas and liquid phases is the only driving force for liquid circulation.

  Visualization of the two-phase flow in the experimental facility has shown that the bubbles tend to move to the middle of the tube and merge, particularly at high viscosities,
displaying a rapid transition to the slug and churn regimes as the gas flux increases. The slug flow regime has shown to be highly unsteady, and is accompanied by significant oscillations that lead to intermittent or ‘pulsating’ circulation.

A progressive decay in the transfer of momentum to the liquid phase as the gas flow rate increases has been determined from the experiments, suggesting that long calandria tubes can lead to poor circulation, this in agreement with practical experience that pans with short calandria tubes are more efficient. The transfer of momentum from the gas to the liquid phase appears to be more efficient within the bubbly regime, while transition to the slug and churn results in reduced drag interaction.

A model for the drag or exchange of momentum in gas-liquid vertical channel flows has been developed from the experiments, which is applied to compute the interaction between vapor bubbles and massecuite in CFD simulations of the flow in calandria tubes of sugar crystallizers. The correlation developed represents the momentum interaction under adiabatic conditions as a function of the Reynolds number (Re), the void fraction (α), and the dimensionless viscosity number (Nμ) developed for slug flows by Wallis (1969). The correlation developed has shown reasonable agreement with comparable experiments, suggesting that it can be applied with confidence in the analysis of adiabatic gas-liquid vertical flows at low liquid flux within a wide range of viscosities (10^{-3} to 10^2 Pa.s).

Comparison with experimental data reported by Rouillard (1985) on the flow in calandria tubes has shown that a reduction in the drag interaction with respect to the correlation developed is required to obtain realistic predictions of the flow. The reduced interaction is attributed to the effect of convective boiling based on experimental evidence

\[ C_D^M = N_{\mu}^{1.6286} (1-\alpha)^{3.951} \frac{3.6351}{Re^{1.7}} \]

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27 Correlation developed for gas-liquid adiabatic channel flow: \[ C_D^M = N_{\mu}^{1.6286} (1-\alpha)^{3.951} \frac{3.6351}{Re^{1.7}} \]
that shows that the generation of vapor increases the frictional resistance and can lead to severe flow and heat transfer instability.

The process in calandria tubes of vacuum pans appears to be dominated by a thermo-hydraulic boiling instability typically found in low-circulation low-heat transfer low-pressure systems, which results in intermittent evaporation and corresponding oscillations in circulation and pressure. Field measurements of the pressure at the top and the bottom of a calandria tube performed in a batch evaporative crystallizer indicated significant oscillations, suggesting that flow boiling instability occurs in calandria tubes. A possible description of the process in the calandria tubes based on reported features of flow boiling instabilities has been proposed, which explains the fluctuations and intermittency in circulation found in sugar evaporative crystallizers.

- **Circulation in Full-scale Sugar Evaporative Crystallizers**

  The buoyancy forces due to phase density differences produced during the boiling in the calandria tubes drives the circulation in sugar evaporative crystallizers. The balance between buoyancy and frictional resistance determines the circulation. To optimize the circulation, the transfer of momentum between the vapor and massecuite should be maximized, and the friction of the circuit should be minimized as far as practically and economically possible.

  The fluid flow in two widely adopted designs of continuous sugar evaporative crystallizers, otherwise known as continuous vacuum pans (CVP), has been studied. Field measurements performed in a CVP indicated that a strong interrelation exists between circulation and heat transfer, as is normal in natural convection processes. A loss in the capacity of the vapor to transfer momentum to the liquid phase and produce circulation as the
gas flux increases has been shown from the field measurements, and this behavior is consistent with the experimental results obtained in the experimental facility in the present study, as well as with experiences in natural circulation water-boiling, in air-lift processes, and in fundamental studies of slugs and churns, where it has been established for high-void fraction flows that a lower momentum interaction occurs as the gas flux and the void fraction are increased.

The fluid flow in full-scale sugar evaporative crystallizers has been simulated applying the Eulerian-Eulerian model of the commercial CFD code Fluent. The (adiabatic) drag correlation developed from the tests in the experimental facility has been applied to model the (non-adiabatic) interfacial momentum interaction between vapor and massecuite in full-scale crystallizers, observing that a reduction in the exchange of momentum (~50%) with respect to corresponding adiabatic conditions is required to obtain agreement with field measurements. The reduced momentum interaction is attributed to the increase in frictional resistance due to non-adiabatic conditions and development of flow boiling instability.

Vortices have been predicted to occur in continuous vacuum pans in the upper-downtake region towards the corner of the calandria. The vortices predicted are considered physically reasonable due to the sharp turn that the geometry of the crystallizers imposes in the circulating massecuite at this location, and agree with results obtained in field measurements and at lab-scale. The vortices predicted could be detrimental for the circulation of massecuite and possibly are responsible for the unsuccessful use of floating calandrias for batch crystallizers, where a generous downtake cannot be afforded as in the continuous units to keep the graining volume within a reasonable limit.
The strategy developed for the numerical simulation of the fluid flow in sugar evaporative crystallizers has been applied to investigate the effect of changes in the design of continuous sugar crystallizers on the flow field and natural circulation, looking for potential alternatives for improvement. The numerical results have indicated that continuous sugar evaporative crystallizers should be designed with a liquid level above the top tube plate around $H \sim 0.6-0.8$ m, a downtake corresponding to circulation ratios between $C_R \sim 0.9-1.2$, and that the use of shorter tubes (e.g. $L \sim 1.0$ m) could bring potential improvements in circulation and performance. No significant benefits in circulation are predicted to result from the use of complicated geometries, such as inclined tube plates, or from the use of elaborate bottom shapes. Based on the flow simulations prospective designs of vacuum pans are presented and analyzed numerically, which might enhance the circulation (+10-24 %), and therefore the efficiency and capacity of the evaporative crystallizers used by the sugar industry.
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VITA

Luis Fernando Echeverri Davila was born and grew up in Cali, Colombia. He received a bachelor’s degree in mechanical engineering from the Universidad del Valle, Colombia, in 1997. Upon graduation, Luis joined the sugar research center CENICAÑA, where he was involved in projects dedicated to improve the sucrose extraction and energy efficiency in Colombian sugarcane mills. In 2002 Luis entered Louisiana State University (LSU) and began his graduate work at the Audubon Sugar Institute and the LSU Mechanical Engineering Department, focusing on the study of the multiphase flow in sugar evaporative crystallizers. Luis will receive a doctoral degree in mechanical engineering in the Spring of 2007.