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Flow Fields in a Mixed-Suspension Mixed-Product-Removal Crystallizer: A Numerical Analysis

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ABSTRACT: Isokinetic withdrawal of solution from research crystallizers is a key factor in obtaining a representative crystal-size distribution and is critical for correct analysis of the crystallization process to accurately determine crystallization kinetic parameters such as crystal growth rates, nucleation rates, breakage rates, and agglomeration rates. Isokinetic withdrawal, in turn, depends upon the characteristics of the flow field in the crystallizer. This research numerically simulated the fluid flow field in a small-scale (experimental) cylindrical round-bottomed, continuous-flow, cooling crystallizer, which was set up for analysis of kinetic parameters in sugar production. The commercial software "CFX 5.5.1" was employed to perform the 3 dimensional simulation with the finite volume method using an unstructured mesh. The impeller used for increasing the gross velocity of the solution inside the crystallizer was modeled using a momentum source. Seven momentum source strength values were used to perform the simulation. The results showed that the momentum source strongly increased the axial flow velocity but only slightly influences the overall flow pattern, except the flow near the outlet tube. The isokinetic withdrawal condition was achieved at a momentum source strength of about 25,000 kg/m^2/s^2. The isokinetic condition for the nuclei crystals was best for the research MSMPR, as it would make the particle size distribution in the product stream most accurate for the analysis of the nucleation and growth rates in the crystallizer. For larger particles, the settling velocity was required for determination of isokinetic withdrawal. The power transmitted by impeller shaft was estimated as 0.6 W, which allowed the impeller rotational speed to be estimated. Furthermore, the velocity at the center of the tank's bottom was lower than that of the surrounding region. This could cause a collection of settled crystals, which was problematic for the operation and analysis of the crystallizer.

Abbreviation: MSMPR, mixed-suspension mixed-product-removal; CFD, computational fluid dynamics.

KEYWORDS: crystallizer modeling, computational fluid dynamics (CFD), fluid flow field.

INTRODUCTION

Crystallization is a transition process of one or more substances from an amorphous solid, liquid or gaseous state to a crystalline state. In solution crystallization, a species crystallizes from a liquid mixture, which will occur only if the solution is super-saturated. There are many methods to obtain supersaturation such as cooling, evaporation, vacuum, pressure, and reaction, or a combination of these processes. Cooling crystallization is widely used to study crystallization in the laboratory. It is a process in which the mixture is cooled down to a super-saturated (or sub-cooled) state, either during the crystallization process, or prior to the addition of seed crystals if the process is to be operated isothermally during the crystallization.

In experimental crystallization processes, isokinetic withdrawal is an important factor to gain a suitable crystal-size distribution of the product. By definition, this condition is achieved when the flow velocities inside and outside the mouth of the outlet tube are equal. When this happens, the path lines of crystals are straight into the mouth of the tube. If the flow in the vessel is sub-isokinetic in relation to the flow in the product tube, the path lines are convergent, causing considerable amount of small crystals to be sucked into the outlet tube while larger ones escape due to their high momenta; the converse is true for the super-isokinetic state. In addition, a typical crystallization process requires fully mixed suspension in the tank and also inside the outlet tube. To attain this condition, the flow pattern at the outlet must be isokinetic. When this condition prevails, the crystal size distribution of an ideal mixed-suspension mixed-product removal crystallizers is exponential, and this results in accurate

experimental data for analysis of growth and nucleation kinetics[1].

Crystal size distribution is one of the key factors that affects the quality of products in several industries, and this is particularly true in the sugar industry. Poor crystal size distributions can also cause many serious technical problems during the production process such as poor filtering and withdrawing. During the crystallization process, small crystal size leads to a collection of crystals at the filter membrane, resulting in a high pressure that could damage the filter membrane. Large crystal size, on the other hand, makes it difficult to withdraw crystals out of the tank, and may also cause a collection of settled crystals at the bottom of the crystallization vessel. From an economic point of view, sugar product of wide crystal size distribution either is a low value product, or requires additional post processing to separate and classify.

There are many factors that affect the kinetic parameters involved in crystallization, such as crystal growth rates, nucleation rates, breakage rates and agglomeration rates, and these kinetic parameters in turn have an effect on the crystal size distribution. Experimental crystallizers are used to determine the kinetic parameters in order to facilitate the design of industrial crystallization units. If the experimental data is poor for any reason, the design of the industrial unit will be not sufficient for production of good quality product. The most significant factors affecting the kinetic parameters are the flow field pattern, mixing, residence time, temperature, concentration, degree of supersaturation, and suspension density. Since the effects of most of the factors (supersaturation, temperature, residence time, suspension density) on the crystallization kinetics are well known already, currently, the knowledge of the flow field pattern is, perhaps, the single most important information toward better design of crystallizers. The geometrical complexity of crystallizers has rendered theoretical flow field study impractical. Experimental study, on the other hand, is difficult, expensive, and time consuming. Currently, high-performance digital computers and simulation software can be employed to obtain very reliable results in an economical manner.

Computational fluid dynamics is a method to obtain solutions for single or multiphase flow analysis, temperature distributions, and chemical composition distributions, through solutions of the equation of continuity, and equations describing momentum transfer, heat transfer, mass transfer, phase change, and chemical reaction. In this research we are dealing with an isothermal system that does not involve chemical reaction or phase change, so that we do not need to model the system with respect to heat transfer, mass transfer, or phase change. We may also assume that the liquid in the crystallizer is incompressible, so that the main equations we require are the equation of continuity

$$\frac{\partial \rho}{\partial t} = -\left(\nabla \cdot \rho \mathbf{v}\right) \tag{1}$$

and the Navier-Stokes equation[2]

$$\rho \frac{D}{Dt}\mathbf{v} = -\nabla p + \mu \nabla^2 \mathbf{v} + \rho \mathbf{g} \tag{2}$$

In turbulent flow systems the equations are more complex, and also can not be exactly modeled at present. The usual method is to use the Reynolds averaged Navier-Stokes equations, which require Reynolds stresses, and therefore (in the simplest approach) eddy viscosities. Eddy viscosities are predicted with a variety of models, the best known of which is the k-ε model. In the current work, the flow is laminar in all areas of the vessel due to the very high viscosity of the supersaturated sugar syrup, and the relatively low flow velocities in the crystallizer. So turbulence models are not required, and will not be further discussed.

In all but the most simple systems, the equations described above do not have analytical solutions, and CFD uses a range of space and time discretization methods in order to enable a numerical solution to be calculated. The most common method to discretize the volume in commercial CFD code is the finite volume technique. Firstly the volume being investigated is discretized into small tetrahedral elements with nodes at each corner of the element. The information relating to the simulation is stored in the nodes of the elements, and equations of continuity can then be applied to each finite volume in the simulation. The last pieces of information that are required by the software are the boundary conditions for each of the boundaries of the total volume (flow rates at the inlets, conditions at the outlets, conditions at solid surfaces, etc…), the physical properties of the fluid(s) and particles in the system, and the physical models that are required for the system.

If the process is multiphase (in this case a fluid containing solid particles) extra relationships are required to model particle movement, momentum, heat, and mass transfer between the phases. As an example of the extra complexity of modeling particle in liquid systems, the forces acting upon a particle in a liquid in a system containing a rotating reference frame include the viscous drag acting on the surface of the particle, the force due to fluid acceleration adjacent to the particle, the virtual mass force (the force required to accelerate the virtual fluid that would be contained in the volume occupied by the particle), the buoyancy force, the Basset force, and the centripetal and Coriolis forces.

The current research was performed using the

commercial software CFX 5.5.1: Exact details of the calculation methods for the software are available in the detailed user notes accompanying the software[3].

In a previous report, the effect of mixing on the crystal size distribution has been investigated by Sha et al[4]. The study concluded that the suspension density in the crystallizer affects the crystallization process; in an imperfect suspension, continuous crystallization depends on both the mixing intensity and the product removal location. This study was experimental, so it would be informative to study the flow field computationally to determine the cause of this effect. Some efforts have been made to simulate the flow field in the crystallizer tank or stirred tank vessel[5,6]. The computed suspension densities were verified with existing experimental data and the predicted volume fraction distribution of the mixed tank agreed reasonably well with the experimental results. Shamlou and Koutsakos[7] determined that each mixing intensity produces a different suspension density distribution, and that the suspension density never reaches total uniformity. These authors also found that at a constant mixing intensity different particle sizes have different suspension density distributions. A similar problem was the mixing or suspension of solids in an agitated vessel, particularly if the vessel contained baffles. An attempt to interpret solid distribution in a fully baffled, mechanically stirred tank affected by flow pattern with three-dimensional simulation and experiment was made by Brucato et al.[8]. The results indicated that the height of the suspension layer is simply proportional to agitation speed. Some other studies have investigated the effect of flow pattern on solid distribution in a stirred tank. Bakker, Fassano and Mayer[9], and Mecale and Montante[10] found that the solid spatial distribution was strongly affected by flow pattern. Impeller diameter, clearance from the bottom of the tank, and the number of impeller blades were all parameters that could affect the flow field in a stirred vessel. None of the reviewed literature incorporates the effect of feeding and removal mechanisms. Therefore, the present work proposes to numerically simulate the flow field in a crystallizer tank with feed and product removal tubes in order to investigate the flow characteristics in that region, as well as in other regions of interest. In particular, this paper aims to numerically study the flow fields inside a small-scale, cooling sugar crystallizer by using the commercial software CFX 5.5.1.

Methods

Crystallizer Configuration

The crystallizer used for numerical simulation in this research is a laboratory model of a continuous mixed-suspension cooling crystallizer at Suranaree University of Technology, Nakhon Ratchasima, Thailand. A schematic diagram of this apparatus is illustrated in Fig. 1.

The shell of the tank is of cylindrical shape with an internal diameter of 137-mm; its bottom is of a hemi-

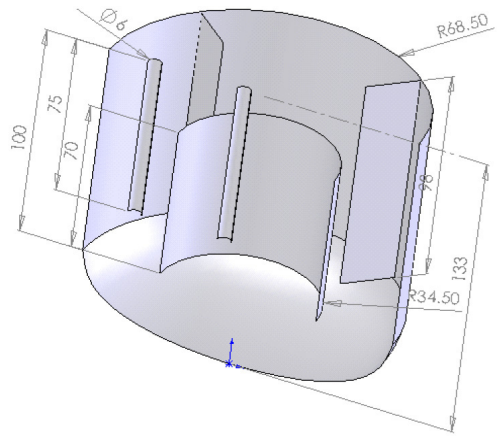

Fig 1. Schematic model of the experimental crystallizer.

spherical dome shape. Attached to the inner wall are four 24-mm wide (in the radial direction), 98-mm high vertical baffle plates which are symmetrically placed 90 degrees apart. A 69-mm-diameter vertical draft tube is located at the center of the shell, within which an impeller to force a downward flow of the solution is installed. The inlet tube (6 mm diameter) is fixed at the center of the draft tube and above the impeller, whereas the outlet tube (6 mm diameter) is placed outside the draft tube and mid-way between two baffle plates. The solution feed rate is kept constant at 2.5 liter per hour, in accordance with a common experimental condition.

Flow Simulation

The flow fields of the crystallizer in this study are intrinsically a three-dimensional flow, and thus it is appropriate to employ commercial software (CFX 5.5.1) to simulate the flow field. Due to symmetry, it is adequate to simulate only one half of the flow field. The plane partitioning the vessel into the two halves must lie through the centre of both the inflow and outflow tubes, and hence it is not possible to reduce the fraction of the vessel modeled further. The draft tube, the vertical baffles, the inlet and the outlet tubes are sufficiently thin to safely assume two-dimensional geometrical models.

Since the flow field at the inlet of the draft tube is quite far away from the rotor, the effect of swirl from the rotor would have already died out there due to the effect of the baffle plates as well as background dissipation. Therefore it should be sufficient that the

rotor be simulated by a momentum source without having to resort to unnecessarily more expensive alternatives, such as by use of moving grid technology, which can capture swirl at the rotor. To model the impeller, a momentum source term has been added at the base of the draft tube, at the position and size of the true impeller. Numerical experiments with both axial and radial components for the momentum source have been used to compare with a momentum source of pure axial momentum. The results show that radial components in the momentum addition do not create a significant difference to the fluid flow profile, particularly at the important section near the product outflow. The radial velocity components of the flow field have been eliminated due to the major role of the vertical baffles. This indicated that the impeller could be adequately modelled by an axial momentum source in order to reduce the computation time and model complexity.

The crystal suspension density used in the laboratory scale crystallizer is low (typically less than 5%) and the crystals are small (less than 0.5 mm), which means that the particles have no significant effect on the flow field of the liquid phase. Although the research aims to determine conditions relating to isokinetic removal of suspension from the crystallizer, it is not possible for one condition to be applicable to all sizes of particles, because the settling velocity of a particle is a strong function of its size. To determine the most appropriate criteria for isokinetic sampling of particles the particle size distribution in the crystallizer must be considered. In an ideal unseeded mixed suspension, mixed product removal crystallizer it is well known that the crystal population density, as predicted by the population balance, is given by

$$n = n^0 \exp\left(-\frac{L}{G\tau}\right) \tag{3}$$

where L is the crystal size; G is the growth rate of individual crystals, which is uniform across the entire population; τ is the drawdown time of the crystallizer, which is equal to the crystallizer volume divided by the flow of mother liquor in the product; and n^0 is the population density of infinitely small particles which can be calculated from the nuclei birth rate, B^0, via the relation $B^0 = n^0 G$. The population density as a function of size is represented by the function n, which is not a function of time (because the crystallizer operates at a steady-state) nor a function of spatial position because the suspension is assumed to be "well mixed".

The distribution can be put in a non-dimensional form to make it a generic function by introduction of the variables

$$\bar{n} = \frac{n}{n^0} \qquad \bar{L} = \frac{L}{G\tau} \tag{4}$$

thus producing the model

$$\bar{n} = \exp\left(-\bar{L}\right) \tag{5}$$

This function is plotted in Figure 2.

This analysis clearly shows that the particle size that has the highest population density in the crystallizer, and is therefore most significant, is the particles which

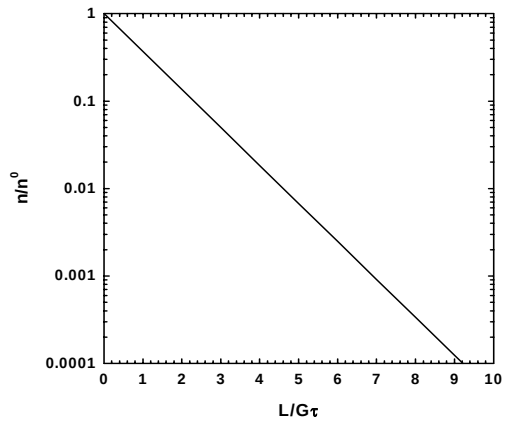

Fig 2. Non-dimensional population density of particles in an ideal MSMPR.

have a size approaching zero, i.e. the nuclei crystals. For this reason the optimum condition to be considered isokinetic with respect to the particles is the condition which is isokinetic with respect to the nuclei. Isokinetic sampling of nuclei is identical to isokinetic sampling of the mother liquor, as the nuclei will move with exactly the same flow field as the liquid. When this is combined with the low suspension density in the crystallizer, the isokinetic sampling can be determined without analysis of the flow of the crystals. As such a two phase simulation is not required, and was not attempted in this study.

Crystals larger than nuclei will be sampled at fractions approaching the isokinetic case, but the difference will become larger as the particle size becomes larger. The maximum significant particle size in the crystallizer is of the order of 100 μm, and crystals of this size will still be sampled reasonably accurately when nuclei are sampled isokinetically. Particles of this size are orders of magnitude lower in number concentration than nuclei, and therefore small errors in the analysis of these particles will not greatly affect the accuracy of the final analysis of the crystal growth kinetics. The calculation of the nucleation rate is based upon the concentration of particles of zero size in the crystallization vessel, and therefore the nucleation kinetics will be measured with the highest accuracy if nuclei are sampled isokinetically.

The fluid is assumed to be a single-phase, sugar solution, whose density and viscosity are 1342.7 kg/m^3

and 236.77 mPa.s, respectively. Steady, laminar flow is assumed throughout because the Reynolds numbers of all the test cases are rather low, of the order of 1.

Because of the low mass flow rate fed into the tank, the temperatures of the incoming solution and the solution inside the tank are only slightly different (not over 2 degrees Celsius). Therefore, the flow can be assumed to be an isothermal process without significant error. The inlet boundary condition was a mass flowrate, and the outlet boundary condition was a 1 atm back pressure at the exit point: these are optimum types of boundary conditions for a robust solution to the problem. A zero slip boundary condition was used at all solid surfaces.

The finite volume methodology with an unstructured grid is built into the software and employed in this study. A grid structure of the studied domain is portrayed in Figure 3. In this configuration, the grid has 73735 nodes, 376481 elements and 40716 faces.

The solution was first obtained using a coarse grid. The calculation is performed iteratively until the residuals reduce to an acceptable level. To achieve grid independent results, the grids were refined step by step until changes in the numerical solution were unnoticeable.

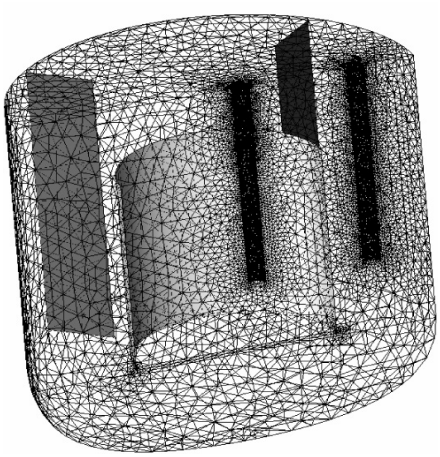

Fig 3. Generated volume mesh for the crystallizer model.

Results and Discussion

Seven test cases were performed, using momentum source values of 0, 1000, 10000, 15000, 18000, 30000 and 50000 kg/m^2/s^2. Figures 4, 5 and 6 show the overall velocity vectors, the enlarged velocity vectors near the mouth of the outlet tube, and the streamlines near the mouth of the outlet tube, respectively, for 0, 1000, 18000 and 50000 kg/m^2/s^2. These figures indicate that for the baseline case (no momentum addition at the impeller location) the overall magnitude of the velocity of the flow field was rather low (Figure4a), except for those portions near and within the inlet and outlet tubes. As such, the overall flow feature was non-uniform. Flow acceleration and a convergent flow pattern near the mouth of the outlet tube (Figures 5a and 6a) were quite obvious. This, as mentioned earlier, was undesirable because it would lead to a poor size distribution of the crystals in the product tube, with a greatly exaggerated amount of fine particles in the product.

As the momentum added through the impeller increased the velocity vectors became larger, both in the downflow section in the draft tube, and in the upflow area in the annular space. This resulted in a stronger flow at the bottom of the draft tube, and also a higher velocity flow up in the annular space, which would assist in suspension of particles. Since the flow in the circular product tube was laminar, it was possible to determine the average flow velocity in the tube as half of the maximum flow, which occurred at the central point of the tube (assuming this was performed far enough from the tube end so that entrance effects were not significant). Thus it was evident from Figure 5 that the flow velocity in the annular space approached, but was slightly less than, the average flow velocity in the product tube when the momentum added was 18,000 kg/m^2/s^2, while at 50,000 kg/m^2/s^2 the flow velocity outside the product tube was significantly higher than the isokinetic condition.

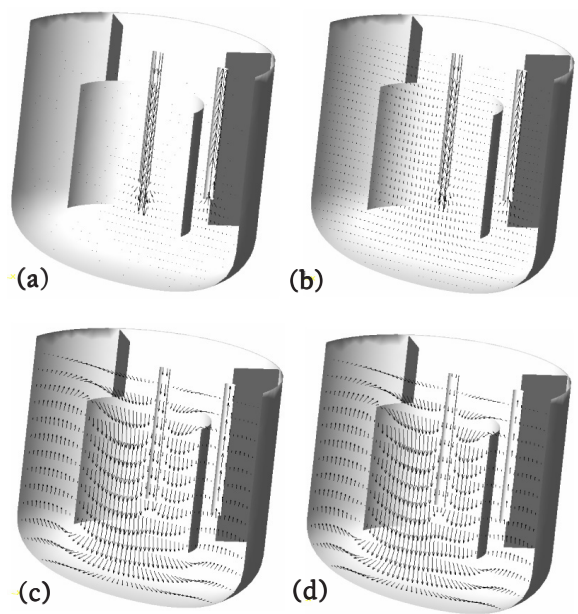

Fig 4. Velocity vectors for one half of the tank for momentum source additions of (a) 0 kg/m^2/s^2; (b) 1000 kg/m^2/s^2; (c) 18000 kg/m^2/s^2; (d) 50000 kg/m^2/s^2.

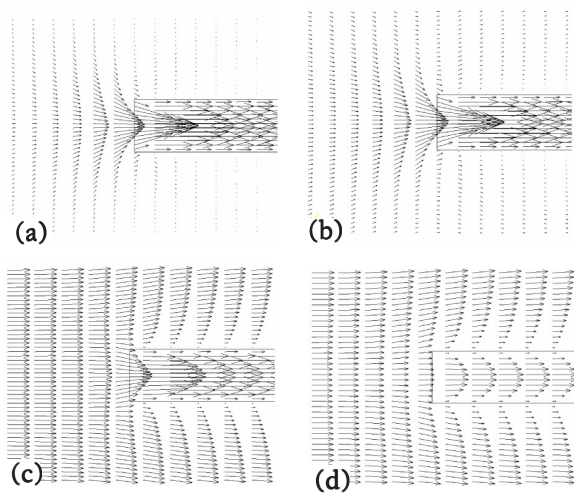

Fig 5. Velocity vectors around the outlet tube for momentum source additions of (a) 0 kg/m²/s²; (b) 1000 kg/m²/s²; (c) 18000 kg/m²/s²; (d) 50000 kg/m²/s².

Overall velocity magnitude contours were depicted in Figure 7, wherein the same general features were observed without any drastic change in structure, such as the formation of the recirculation zones inside the draft tube sometimes observed within a flow with an impeller.

As described previously, isokinetic withdrawal leads to the desirable crystal size distribution. It was useful to estimate the appropriate value of the momentum source that gives the isokinetic condition. Plane A in Figure 8 was used to obtain the average velocity, which was compared with the average velocity inside the outlet tube. With the results obtained from the seven test cases, the fitted curve for the velocity ratio (the current velocity divided by the velocity at no source condition) versus the percent of momentum source added is shown in Figure 9 (the percent increase was calculated based on the momentum flux of the baseline value.) The appropriate value of the momentum added that would give the isokinetic withdrawal was estimated as 6.43 % of the baseline condition; this corresponded to the absolute momentum value of approximately 25400 kg/m²/s². Then the power transmitted by impeller shaft was 0.58 Watt. The impeller rotational speed could be determined iteratively from the impeller power correlation curve[11]. In the laboratory crystallizer, which is normally used with an impeller of diameter : height ratio of 8:1, with a 45° pitched blade turbine, it requires approximate 2.2 rev/s to obtain the isokinetic condition.

The difficulty in the design of mixed-suspension crystallizers is that isokinetic flow at the product point

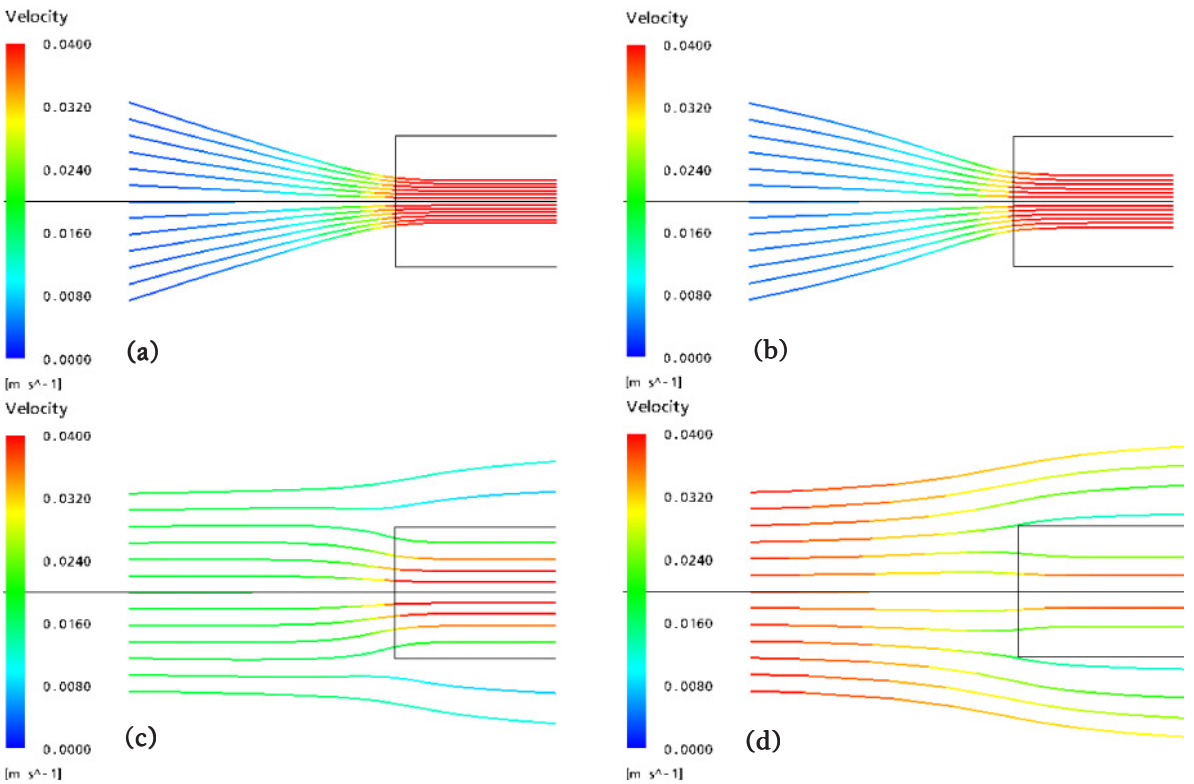

Fig 6. Streamlines around the outlet tube for momentum source additions of (a) 0 kg/m²/s²; (b) 1000 kg/m²/s²; (c) 18000 kg/m²/s²; (d) 50000 kg/m²/s².

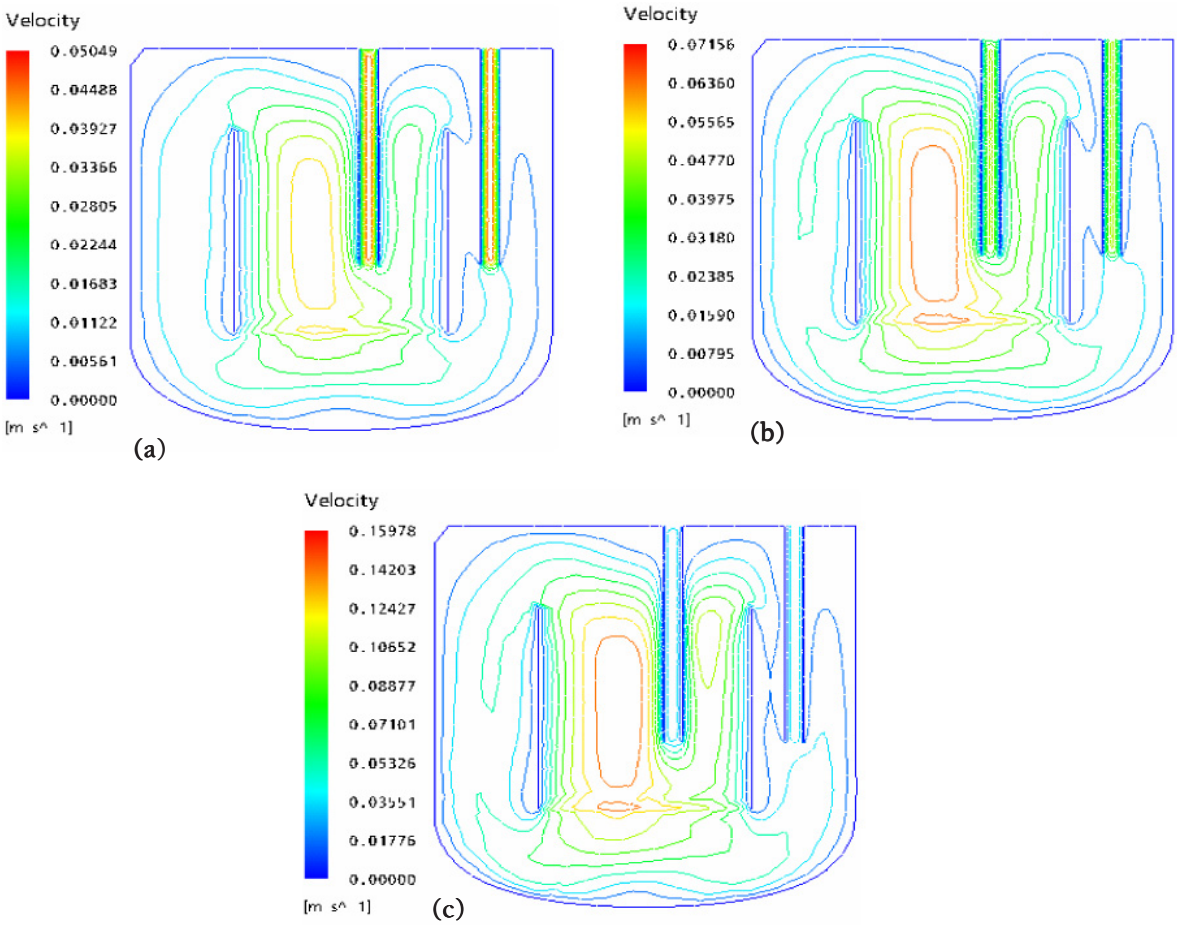

Fig 7. Velocity contour plots at the tank centerline for momentum source additions of (a) 10000 kg/m^2/s^2; (b) 18000 kg/m^2/s^2; (c) 50000 kg/m^2/s^2.

is not a sufficient criterion to ensure the product size distribution. It is also essential that the crystals in the vessel are sufficiently well suspended in the mother liquor. Suspension of crystals only occurs if the upflow velocity of the fluid produces sufficient drag to overcome the effect of the weight of the individual particles. The suspension quality in a vessel is an extremely difficult parameter to determine experimentally, as suspensions are typically opaque, such that the common light sensing methods cannot detect suspension quality. Hence, a typical parameter used to describe impeller speeds suitable for sufficient suspension is the "just suspended" criterion: this criterion states that no particle should spend more

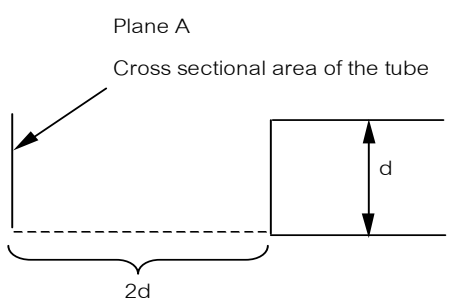

Fig 8. Location of the plane used to check for the isokinetic flow condition.

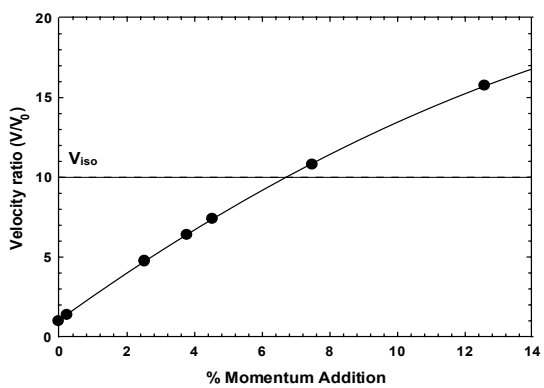

Fig 9. Velocity at a plane near the outlet of the crystallizer as a function of the amount of momentum added by the impeller.

than 1-2 s on the base of the vessel at any time if the solids in the vessel are to be considered well suspended.

The most well known correlation to determine the impeller speed required in a stirred vessel to achieve a suspension at the just suspended condition is then Zwietering equation[12]

$$N_{js} = S\frac{\nu^{0.1}D_p^{0.2}\left(g\Delta\rho/\rho_l\right)^{0.45}X^{0.13}}{D^{0.85}} \qquad (6)$$

where S is some function of the ratio of the tank diameter to the impeller diameter (T/D) and the ratio of the tank diameter to the impeller clearance (T/C_B). The impeller clearance is the distance from the base of the impeller to the bottom of the vessel. The most commonly used equation for the parameter S was developed by Nienow[13]

$$S = 2\left(\frac{T}{D}\right)^{1.33} \qquad (7)$$

A more recent correlation[14] also accounts for variations in the impeller clearance. For a pitched blade impeller similar to the type used in the laboratory scale crystallizer, the correlation is

$$N_{js} = 2.32\left(\frac{T}{D}\right)^{0.84} \times\left(0.72\frac{C_B}{T}+1\right)D^{-0.82}X^{0.11}D_p^{0.23}\nu^{0.09}\left(\frac{g\Delta\rho}{\rho_l}\right)^{0.51} \qquad (8)$$

Unfortunately, these correlations (and other similar correlations) were based on results in flat bottomed vessels without draft tubes, whereas the crystallizer being modeled has a round bottom and a draft tube, for which no correlation for suspension conditions is available in the literature. Draft tubes, in particular, greatly increase the ability of an agitator to suspend particles in a vessel because they create a condition where the upward flow and downward flow parts of the flow loop are clearly segregated, allowing a high upflow velocity in the annular space of the vessel. This means that results of these correlations will predict impeller speeds that are higher than necessary to suspend a particle in a vessel such as we were using, but it is still instructive to use them to obtain an idea of the general magnitude of the particles suspended at the isokinetic condition.

The Zwietering correlation (equations 6 and 7) and the Armenante correlation (equation 8) were used to estimate a value for the size of particles that would be suspended if the stirrer was operated at 2.2 rev/s in the experimental crystallizer. Geometrical parameters for the crystallizer were the same as those used in the CFD model (see Figure 1), as were the fluid and particle properties. Using the calculated impeller speed to achieve isokinetic flow at the product point (2.2 rev/s), it was possible to estimate the maximum size of particle that would be fully suspended in the crystallizer. At a solid suspension of one percent (a typical value for the laboratory crystallizer studied here), the Zwietering correlation predicted that only particles smaller than 10 micrometer would be suspended adequately, while the Adamante correlation predicted that particles smaller than 20 micrometer would be suspended adequately. These correlations may be inadequate for the crystallizer being investigated because they were for flat bottomed vessels not containing a draft tube. As mentioned above, the draft tube increases the agitator's ability to suspend particles by more completely segregating the upflow and downflow parts of the flow loop in the vessel. In addition, the kinematic viscosity of the fluid in the crystallizer (176×10^{-6} m^2/s) is higher than that of the most viscous liquid used to predict the correlation (148×10^{-6} m^2/s). This suggests that the crystallizer may suspend particles larger than those suggested by the correlations (and introductory two-phase simulations suggest that this is true). However the correlations did indicate that the larger particles in the crystallizer (which may be in the order of 100 μm) would not be well suspended, particularly if they approached the dead zone at the base of the crystallizer, where the fluid velocity was very low.

This leads to one of the apparent contradictions in the design of crystallizers: the agitator speed required for complete suspension may be quite different to the speed required for isokinetic flow at the product outlet. These two conditions (suspension of particles and isokinetic flow) may appear to be related, but in fact they are quite different. Isokinetic flow can be achieved at any product flow rate either by varying the agitator speed, or by adjusting the diameter of the product tube. Smaller diameter tubes create higher velocities at the sampling point for a particular product flowrate. The typical method used to correct this in laboratory crystallizers is to have a periodic outflow in the product tube, which allows for a significantly higher flow velocity over a shorter period (typically 10 – 20 percent of the total operating time). This allows for much higher agitation speeds to be used, allowing for larger particles to be completely suspended, while maintaining an approximately isokinetic flow condition at the outlet so that the sampling of particles is ideal. This approach is very difficult to model with CFD because the simulation becomes transient (over a significant period of time to see the entire flow period), and requires a free surface as the crystallizer empties slightly during the sampling period, and refills during the period the outflow is stopped.

Conclusions

The proposed computational study to obtain flow fields within the model sugar crystallizer appeared to be a successful endeavour and the results were reasonable. The overall magnitude of the velocity within the crystallizer could be increased by increasing the axial momentum source. Isokinetic withdrawal condition could be achieved at approximately 25,400 $kg/m^2/s^2$ of the momentum source value, in accordance with the 2.2 rev/s for the crystallizer with the impeller of 8:1 diameter : height ratio and a 45° pitched blade turbine. In all the seven test cases studied, the overall flow patterns remained generally the same, which was desirable because it would not complicate the design for optimization process. Even with higher values of the momentum source, the velocity at the center of the tank's bottom was still low; this was undesirable because sugar crystals may settle, causing a lump of settled crystals there. The velocity magnitude inside and outside the draft tube was rather different due to the effect of the flow area; this problem could be eliminated by designing the two regions to have equal areas.

Nomenclature

C_B Impeller clearance from the vessel bottom, m
D Diameter of the impeller, m
D_P Particle diameter, m
g Gravitational constant, 9.81 m/s^2
Njs Impeller speed required for particles being "just suspended", rev/s
T Tank diameter, m
X Solids concentration, g solids/g suspension × 100 %

Greek symbols

ν Kinematic viscosity, m^2/s
ρ_l Density of the liquid phase, kg/m^3
ρ_s Density of the solid phase, kg/m^3
$\Delta\rho$ Solid-liquid density difference, $(\rho_s - \rho_l)$, kg/m^3

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