ABSTRACT

TANYEL, ZEYNEP. Residence Time Distribution of Multiple Particles in Four Configurations of Holding Tubes (Under the direction of K.P. Sandeep)

Residence time distributions of multiple particles (as affected by process and system parameters) were investigated during non-Newtonian tube flow. Process parameters included flow rate, particle type, carrier fluid viscosity, and particle concentration. The system parameter of interest was the holding tube configuration. Polystyrene and acrylic particles were used as model food particles. Digital imaging analysis was used to obtain residence time data of particles. A novel type of holding tube (chaotic holding tube) was constructed. Comparisons among the straight, single helical, double helical, and chaotic holding tube were performed in terms residence time distribution (RTD) of particles. In addition, the effect of inclination angle (0 º and 45 º) of the chaotic holding tube was investigated. It was found that the narrowest RTD of particles was obtained in the single helical holding tube. RT_{std} in the chaotic holding tube was greater than that in the single helical holding tube, but lower than those in the straight and double helical holding tubes. The most significant process parameter affecting RTD was flow rate. The effect of flow rate was more pronounced for high density (acrylic) particles. Carrier fluid viscosity, particle type, and particle concentration did not have significant effects on the overall RT_{std}. Trends observed in the chaotic holding tube were similar to those observed in the single helical holding tube. However, there were some cases where use of chaotic holding tube resulted in narrower RTD of particles. Changing the inclination angle from 0 º to 45 º in the chaotic holding tube resulted in wider RTD of both types of particles.

RESIDENCE TIME DISTRIBUTION OF MULTIPLE PARTICLES IN FOUR CONFIGURATIONS OF HOLDING TUBES

by

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A thesis submitted to the Graduate Faculty of
North Carolina State University
in partial fulfillment of the
requirements for the Degree of
Master of Science

FOOD SCIENCE

Raleigh

2004

APPROVED BY:

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BIOGRAPHY

Zeynep Tanyel was born on March 12, 1980 in Eskisehir, Turkey. She grew up in Denizli, a small town in the Aegean region of Turkey. In 1993, her family moved to the capital city, Ankara, where she attended high school and college. She graduated from Middle East Technical University in 2002, with a B.S degree in Food Engineering. In the summer of 2002, she was admitted to the graduate school at North Carolina State University where she began her Master of Science degree in the Department of Food Science. She currently works as a research assistant in the same department.

ACKNOWLEDGEMENTS

First of all, I would like to thank my advisor, Dr. K.P. Sandeep, for all of his support and encouragement during my study. I have always felt very lucky to be working with him. I am very thankful for his understanding and guidance.

I would like to thank Dr. Josip Simunovic for all of his assistance during the experimental work of this study and for his suggestions as a member of my committee.

I would also like to thank other members of my committee, Dr. Saad Khan and Dr. Andrey Kuznetsov, for their input.

My special thanks go to Dr. Jason Osborne for helping me with the experimental design and statistical analysis of the data.

I thank Jack Canady for his help with the construction of the experimental setup. He was ready to help whenever I needed and I am grateful for that.

My sincere thanks are extended to the former and current food engineering graduate students, specifically Cristina, Dorin, Pablo, Ediz, Andriana, Junhua, Aswini, and Supriyo for all the good times and friendship we shared.

I wish to thank all my friends both here in the U.S. and in Turkey for always being there for me. Special thanks go to Reha and Burcu for being so supportive during my stay in Raleigh.

I would like to thank my wonderful family for always believing in me. Without their love and encouragement I would never be the person I am now.

Finally, my sincere thanks go to Ilker just for being who he is and for being in my life.

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Chapter 1

INTRODUCTION

Aseptic processing of foods has been a very interesting subject for both researchers and processors worldwide. Aseptic processing involves continuous heating, holding, and cooling of the product followed by aseptic packaging in a sterile environment. The primary objective of this process is to produce high quality, commercially sterile products to meet the growing demand of consumers.

Application of aseptic processing to liquid foods such as milk, fruit juices, wine, salad dressing, and egg has been successful. Foods containing small particles (baby foods, cottage cheese, soups, and rice desserts) have also been aseptically processed. However, aseptic processing of foods with large particulates (larger than 3.2 mm) remains as a major issue in the food industry (Torres and Oliveira, 1998).

The potential of particulate products to be aseptically processed is great and it is known that these products will offer significant benefits over conventionally processed counterparts in terms of quality and consumer friendly packaging (Rice, 1996). If aseptic processing is applied to products containing solid particles, high quality ready-to-eat products such as soups and stews will be available. However, aseptic processing of particulate foods encounters mainly two problems -- particle integrity and assurance of commercial sterility (Richardson *et al*., 1987). Assurance of commercial sterility is related to the heat transfer to the particles and how long they receive the heat treatment. So, in aseptic processing of particulate foods, determination of fluid-to-particle heat transfer coefficient (h_{fp}) and residence time distribution (RTD) of particles are of great importance. Due to the regulation of the Food and Drug Administration (FDA) that gives credit to the lethality accumulated in holding tube section of the system, RTD of particles in the holding tube should be determined for process establishment.

Since it is not yet practically possible to obtain the time - temperature data of cold spots of particles flowing in a continuous system, there is a need to measure the residence times of particles. From product safety point of view, measuring residence time of the critical particle is necessary. The critical particle is the particle that receives the least heat treatment. In a single type particle flow, fastest moving particle receives the least heat treatment. When there is more than one type of particles with different sizes and thermal diffusivities, fastest moving particle may not be the critical one, i.e. faster moving particle with high thermal diffusivity can receive more heat treatment than slower moving particle with low thermal diffusivity (Damiano, 1997). The process conditions should be such that the critical particle receives adequate heat treatment to accumulate the required lethality and overprocessing of the other particles is avoided as much as possible.

While the residence time of the critical particle is important for microbiological safety, residence time distribution of particles is important for product quality. Residence time distribution of particles can be characterized by the standard deviation of residence times. This standard deviation will give an idea of the wideness of the distribution curve, and will hence indicate the degree of uniformity of the heat treatment applied to the product. The narrower the distribution, the product will be treated more uniformly. So, achieving a narrow residence time distribution should result in a uniformly processed and high quality product.

The residence times of particles are affected by various factors. For flow in a holding tube, the configuration of the holding tube plays a very important role in determining the velocity profiles, and thus residence times. A geometric configuration which favors radial mixing of the particles will flatten the axial velocity profile and narrow the residence time distribution. Such a configuration can be obtained by coiling the tubes to yield a helical geometry. In this setup, the direction of flow is forced to change due to curvature of the tube, resulting in a loss of momentum of the fluid and a subsequent increase in pressure at the outside wall of the tube. A flow in the radial direction, secondary flow, is developed because of this pressure gradient across the cross-section of the tube. Thus, more mixing is obtained, which results in a narrow RTD (Sandeep *et al.,* 2000). Other than holding tube geometry, flow rate and rheological properties of the carrier fluid, particle type, size, density, and concentration are the important factors affecting residence times of particles.

Current literature on residence time distribution mostly deals with liquid foods or particulate foods in conventional holding tubes. Limited research has been conducted on particulate foods in helical holding tubes. However, there is still need for more information on system and process parameters affecting the RTD of particles. Specifically, the behavior of mixtures of different types of particles and different types of holding tube geometries need further investigation.

The main objective of this study was to determine the effects of different process parameters such as flow rate, carrier fluid viscosity, particle type, particle concentration, and mixing of particles of different densities on residence time distribution of particles. Another objective was to test different geometric configurations of holding tubes and to perform a comparison study among those. For this purpose, a novel holding tube configuration, which has been named a chaotic holding tube, was developed and compared to single helical, double helical, and straight configurations of holding tubes from the standpoint of RTD. The results of this study will give an insight on the conditions under which a narrow RTD of particles can be obtained, which translates to a uniformly processed product.

Chapter 2

REVIEW OF LITERATURE

2.1. Aseptic Processing and its Application to Particulate Foods

Aseptic processing involves sterilization of the product and the package separately and then bringing them together in a sterile environment. The product is sterilized continuously in a heat-hold-cold process. In aseptic processing of low-acid particulate foods, some major hurdles are present in process establishment. Unlike conventional retort process for which heat penetration studies are easily applied for process evaluation, there is no technique to measure the center temperature of particles that are sterilized during continuous flow.

The Food and Drug Administration (FDA) requires biological validation of sterility and mathematical modeling of the process. Due to the difficulties in ensuring the safety of aseptically processed low-acid foods, not many filings have been received by the FDA. In 1995, a workshop was organized by National Center for Food Safety and Technology (NCFST) and Center for Aseptic Processing and Packaging Studies (CAPPS). This workshop pointed out the main concerns involved in aseptic processing of particulate foods and discussed the possible ways to solve the problems (Larkin, 1997). The issues addressed by this workshop were the critical particle concept, mathematical modeling and physical properties, statistical design and analysis, residence time distribution, and biological validation of the process (CAPPS-NCFST Workshop, 1995, 1996).

2.1.1. History of Aseptic Processing

The history of the aseptic processing goes back to 1920's. Olin Ball was one of the pioneers of aseptic processing who introduced the HCF (heat, cool, fill) process. Although the process itself was not a major success, it was a milestone in the development of aseptic processing. In 1942, the Avoset process was developed by George Grindrod. In this process, steam injection was used to sterilize the product and the product was packaged in containers that were retorted or sterilized by hot air. Another important step in the development of aseptic processing was the Dole aseptic process. This process consisted of four steps: sterilizing the product in tubular heat exchangers, sterilizing the packaging materials with superheated steam, aseptic filling of the product, and sealing the container in an aseptic environment. The use of superheated steam at atmospheric pressure was the biggest advantage the Dole Process brought. The first commercial aseptic plant was built in 1951 in Washington State, USA by Roy Graves and Jack Stambaugh. The process was based on one of the aseptic canning machines used in the Dole process. The next step in the development of the aseptic process was in packaging materials. In 1960s, chemically sterilized packages were introduced by TetraPak. The use of hydrogen peroxide to sterilize packaging materials was approved by the FDA in 1981 (Graves, 1996).

In order to discuss the concerns related to establishment of an aseptic multiphase food process, the National Center for Food Safety and Technology (NCFST) and the Center for Aseptic Processing and Packaging (CAPPS) organized a workshop in 1995. After the workshop, Tetra Pak, Inc. developed the first successful filing for condensed cream of potato soup. The filing was accepted by the FDA in May 1997 (Larkin, 1997).

Today, there are more than 30 companies in the U.S. producing low-acid aseptically packaged food products, such as milk, cheese sauces, creams, infant formulas, puddings, and soups (Graves, 1996).

2.1.2. Advantages of Aseptic Processing

Continuous heat treatment offers many advantages such as minimizing loss of nutrients as compared to traditional canning process. Typical organoleptic properties (color and texture) and heat sensitive nutrients are better retained due to high temperatures and short processing times. Another advantage of aseptic processing has to do with packaging. A wide variety of shapes and materials can be chosen for aseptic packaging, whereas in conventional processing, metal cans and glass jars are the major choices. In addition, use of pre-sterilized packages in aseptic processing avoids recontamination of the sterile food. However, there are some drawbacks of aseptic processing. These are lower speeds of fillers, higher cost of equipment and the automation required for process control. As far as the current status of aseptic packaging is concerned, use of the process for specific product types and destroying heat-resistant enzymes are some other limitations (Ramaswamy *et al*., 1997).

Despite all safety and quality attributes aseptic processing offers, not many commercial filings have been received by the FDA. The reason is attributed to uncertainties involved in continuous thermal processing of particulate low-acid foods (Larkin *et al*., 1997). During thermal treatment, microbiological, chemical and biochemical components undergo thermal destruction, but the rates of these reactions are different. So, an engineering analysis is necessary to determine an optimum thermal treatment for the safest and highest quality product (Jung and Fryer, 1999). However, processors should keep in mind that every situation is unique and there will probably never be instructions for optimum treatment determination applicable to every product (Larkin, 1997).

2.1.3. Aseptic Processing Equipment

The components of an aseptic processing system are a pump, deareator, aseptic surge tank, back pressure valve, heat exchanger and holding tube. The characteristics of the equipment have a big influence on the efficiency of the process so that using different types of heat exchangers results in different heating rates (Heldman, 1992). A pump is critical in terms of maintaining the continuity of the process. Positive displacement pumps (rotary or reciprocating) are used in aseptic processing. Pressure drop, viscosity of the product, and type and size of particles are the factors that will determine the choice of the positive displacement pump. With small particle sizes (< 1/8 ") and low pressure drops (< 150 psi), a rotary positive displacement pump is preferred. When the product contains particulates up to 3 " in size, a reciprocating piston pump is preferred (Lund and Singh, 1993). Dearation is the step where the excessive air is removed before sterilization process. This step is critical especially for products that are under risk of oxidation reactions. In addition, dearation has some advantages such as reducing fouling in heat exchangers and maintaining constant filling conditions by preventing foaming (Carlson, 1996). Another important component is the back pressure valve (located after the cooling unit) which provides a pressure higher than the boiling pressure of the product. The purpose of using an aseptic surge tank is to provide sterile product to the packaging unit. In case a problem is encountered in sterilization process, packaging can be carried on independently using the product stored in the tank.

For heating and cooling of the product, there are two methods that can be applied. One of them is direct heating or cooling and the other is indirect heating or cooling. In direct heating, the product is in contact with steam either by steam injection or steam infusion. Steam is introduced into the product by an injector in the steam injection method. In steam infusion method, the product is dispersed into the steam by spraying or by falling sheets (Lund and Singh, 1993). In direct heating, due to condensation of the steam, which is how the heat transfer occurs, some dilution of the product takes place. This brings the need for an additional operation, flash cooling, to remove the excess water. The advantages of direct heating are very rapid heating of the product and dearation of the product in flash cooling. However, high cost of the equipment, requirement for an additional operation, and the risk of losing desirable volatiles are the disadvantages of direct heating. In indirect heating, plate heat exchangers, tubular heat exchangers, and scraped surface heat exchangers are used. Plate heat exchangers are inexpensive and easier to clean. However, their usage is limited to low viscosity products such as milk or fruit juices. These types of heat exchangers can also be used for preheating or precooling of the product. In tubular heat exchangers, the product and heating or cooling medium are separated by concentric tubes. These heat exchangers can be double tube or triple tube types. Tubular heat exchangers can handle more viscous products than plate exchangers. Another advantage is the lower number of gaskets which reduces the cost and also enhances microbiological safety. However, high pressure drops, low rates of regeneration, and tendency for fouling are the drawbacks of tubular heat exchangers (Bhamidipati and Singh, 1995). For more viscous products such as pastes and fillings scraped surface heat exchangers are preferred. The scraper blades that rotate inside provide uniform heat transfer and avoid burn-on. However, these blades can cause significant mechanical damage and disintegration of particles. Another problem related with the use of scraped surface heat exchangers is the non-uniform particle density distribution in vertical scraped surface heat exchangers.

Coiled heat exchangers are becoming popular due to their ability to provide high heat exchange rates, improved mixing of the product, and thus uniformity of heating. In helical coils, a flow in radial direction, secondary flow, develops due to centrifugal force. This secondary flow provides mixing of fluid elements in the radial direction. However, in helical tubes, more kinetic energy is dissipated due to secondary flow and thus the pressure drop increases. So, the processor should investigate pressure drop and related pumping requirements before using helical heat exchangers (Coronel and Sandeep, 2003).

The most critical section of an aseptic processing system is the holding tube. FDA gives credit for the lethality accumulated only in holding section. In holding tube, commercial sterility of the product is ensured by achieving the adequate time-temperature combination. According to the FDA regulations, the holding tube must be sloped upwards at least 0.25" per foot of the tube in order to eliminate air pockets and help drainage of the product (Lund and Singh, 1993). After holding, the product is cooled to the desired temperature and then sent to the packaging unit. To sterilize packaging materials, mechanical, thermal or chemical methods can be used. Today, the most popular method is the use of hydrogen peroxide (H_2O_2) due to its effective and rapid sterilization. It is the only FDA approved chemical used for sterilization of packaging materials. Generally, 25 - 35 % H_2O_2 is used and the residual of H_2O_2 is limited to 0.5 ppm after packaging (Floros, 1993).

The processing equipment should be sterilized during the clean-in-place (CIP) cycle. For aseptic processing of high-acid foods, CIP cycle consists of hot water, alkali, and hot water. The CIP cycle of low-acid foods is hot water, alkali, hot water, acid, and hot water. The air flow system should also be sterilized. For this purpose, high efficiency particulate air (HEPA) filters are used. Here, air is treated through the filters as it circulates (Carlson, 1996).

2.1.4. Effect of Aseptic Processing on Quality

The primary advantage aseptic processing offers is the high quality attributes resulting from high temperature-short time processing. In aseptic processing, majority of the quality factors such as color, flavor, texture, and nutritional value are preserved while the product is commercially sterilized.

Thermal degradation reactions of vitamins or pigments have higher decimal reduction times (D value: time required for 90% degradation) and higher thermal resistance constants (z value: change in temperature for a ten-fold change in D value) values which correspond to slower rates of reactions compared to the microbial inactivation reactions. Inactivation of *Clostridium botulinum*, which is the target microorganism for low-acid foods, has an activation energy of 70 kcal/mole. Some of the reactions important in terms of food quality have the following activation energies -- Thiamine degradation: 27 kcal/mole, Vitamin A degradation: 14.6 kcal/mole, Maillard browning: 27 kcal/mole. These low activation energies indicate that these reactions are less temperature sensitive, i.e. the higher the temperature, the less damage product will receive (Nielsen *et al*., 1993).

Chemical changes in aseptically processed foods are related to the food components and the reactions they are involved in. For example, proteins may cause changes in color by enzymatic reactions, carbohydrates may affect color due to browning reactions, and lipids may develop textural and flavor changes due to oxidation reactions.

Enzymes are responsible for quality deterioration of foods. They can cause off-flavor development (peroxidase, catalase), texture changes and loss of consistency (amylase, cellulase), color degradation (chlorophyllase, polyphenoloxidase), and nutritional changes (ascorbic acid oxidase and thiaminase). Peroxidase, due to its highly temperature resistant nature, is generally used as an indicator of adequate heat treatment. In aseptic processing, the structure of the protein may only be partially denaturated which results in reactivation of the enzymes during storage of the product. Peroxidase was found to demonstrate such a property during storage of aseptically processed peas in starch slurry (Schwartz, 1992). Amylase, which is responsible for thinning of starch based products, is also one of the heat resistant enzymes. The main source of this enzyme is microbial. Thus, formulation of the product, microbial load, and enzyme structure should be considered together when designing a process. When particulate products are considered, the heat resistant enzymes might be contained within the tissues of the particles, and thus a stronger treatment to deactivate those enzymes may be necessary.

Flavor changes for aseptically processed foods occur either during heat treatment or storage. Although destruction of flavor components is reduced in aseptic processing, some off-flavor formation may result after high temperature treatment. The cooked flavor, mostly encountered in UHT milk, is one of the reasons why consumers choose pasteurized milk over aseptically processed milk (Schwartz, 1992). In aseptic processing of orange juice, some essential volatile flavor is lost during vacuum concentration operation. Lost aroma and essential oils are then added prior to aseptic packaging. It was found that the main factor causing quality loss during storage of aseptic orange juice was not the interaction of the food with the packaging material, but oxidation reaction and chemical degradation of flavoring components (Sizer *et al*., 1988). On the contrary, Schwartz (1992) reported that the loss of D-Limonene flavor compound in orange juice was due to its adsorption onto packaging film.

The main pigments in foods are chlorophylls, carotenoids, anthocynains, and betalaines. Chlorophylls have been studied by many researchers due their importance in green vegetable processing. Decomposition of chlorophyll to pheophytin and further to pyropheophytin results in olive-brown color. In order to retain the original color in retort and aseptic processing of green vegetables, addition of zinc was shown to be promising. Carotenoids, which are relatively heat stabile, can go through trans-cis isomerization upon heat treatment and a subsequent reduction in their vitamin A precursor activity occurs (Schwartz, 1992).

In aseptically processed fruit juices, the primary nutrient degradation is the loss of vitamin C due to oxidation reactions. There are different ways oxygen can enter the product -- dissolving in the juice, penetrating through the packaging material, or remaining after hydrogen peroxide treatment. Moreover, besides oxygen, temperature abuse can cause vitamin C degradation during storage of the product (Sizer *et al*., 1988).

David (1992) conducted a review study to summarize the differences in nutrient (ascorbic acid and thiamin) retention between retorted and aseptically processed soups. It was found that 91 % of ascorbic acid was retained in aseptically packed tomato soup, whereas 59 % of ascorbic acid remained in conventionally packed product. Retention of thiamin was also higher in aseptically packed chicken soup as compared to the retorted one. 82 % of thiamin was found in the chicken soup after aseptically processing, whereas only 27 % of thiamin retained in the retorted product.

In aseptic processing of particulate foods, fluid-to-particle heat transfer coefficient (h_{fp}) is a very important factor in determining the extent of microbial destruction and nutrient degradation. In a study by Palazoglu and Sandeep (2002), a conservative mathematical model was developed to determine the effect of h_{fp} on lethality and nutrient retention. The computer program calculated the lethality accumulated and volume average nutrient (thiamine and lysine) retention. For small particles (r = 0.0075 m), even for infinite h_{fp}, thiamine retention was acceptable. Increasing h_{fp} was found to decrease the time required to reach the target lethality. However, increase above a certain value was observed to reduce the overall nutrient content. This was attributed to the fact that near-surface regions of the particles reached higher temperatures and thus, volume average nutrient retention decreased.

Chandarana and Gavin (1989) studied the retention of nutrients for heterogeneous foods which were aseptically processed. They developed a mathematical model to predict the heat transfer into particulate foods. The impact of heat on nutrients and microbial load was assessed. Lower retention of thiamine and peroxidase was reported for large particle sizes. However, it was pointed out that some spoilage enzymes could remain active within smaller particles due to lower surface area available for enzyme inactivation.

Storofos (1992) discussed the effects of various product and processing parameters on process optimization based on maximum nutrient retention. It was found that in small particles (diameter = 0.0127 cm) more thiamin retained after processing. In addition, upon decreasing fluid viscosity, thiamine retention increased. This increase may be due to higher

Reynolds number which results in higher heat transfer coefficients, and shorter processing times to accumulate the required lethality. No effect of particle concentration on thiamine retention was observed.

Gratzek and Toledo (1993) determined the optimum process temperatures for maximum product quality. Enzyme destruction kinetics and volume averaged cook values were modeled by Simpson's numerical integration. The overall cook value was found to decrease with decreasing particle size and increasing temperature. Optimum processing temperature increased as particle size decreased when enzyme inactivation was not considered. When peroxidase inactivation was taken into account, the temperatures for optimum cook values were lower for smaller particles.

2.2. Mathematical Modeling

The need for mathematical modeling in the design of an aseptic process results from the difficulties in determining internal temperature of particles. Mathematical models followed by microbiological verification of the model, are the tools to estimate temperature history of the cold spots in particles (Sastry, 1988). Due to the complex dynamics of particulate flow; influences of particle size, shape, residence time, fluid-to-particle heat transfer coefficient, configurations of the heat exchanger and holding tube, and thermophysical properties are very important on heat transfer rates, and thus there is a need for modeling of heat transfer (Pflug *et al*., 1990). Since non-invasive measurement of the center temperature of moving particles is extremely difficult (practically impossible), the determination of residence times of particles and convective heat transfer coefficients at the liquid-solid interface are the points that mathematical modeling focuses on (Sastry and Cornelius, 2002). The mathematical model developed accounts for the variations in process parameters and conditions. Hence, process control and monitoring will be easier with mathematical modeling (Sastry, 1997).

In case of liquid products, not many problems are encountered in mathematical modeling. Knowing the flow characteristics of the fluid and type of flow, residence time data can be used to determine the required hold tube length to achieve the previously calculated lethality. However, for particulate products, accurate data on residence time and fluid- to-particle heat transfer coefficient is necessary for a reliable mathematical model.

The primary objective of mathematical modeling is to determine the temperature within the critical particle which will determine the particle-center lethality. For this purpose, a partial differential equation governing the heat transfer into the particle is solved with a convective boundary condition (Chandarana *et al*., 1989). For the energy balance in the heat exchanger, it is assumed that the fluid is perfectly mixed in the radial direction of the heat exchanger, and that it is in contact with a population of particles moving at the same average velocity. In the holding tube, the energy balance is performed in a manner similar to that in heating section (Sastry and Cornelius, 2002).

Sastry (1997) summarized the steps to develop a mathematical model for aseptic processing of particulate. These steps involve prediction of fluid temperature from an energy balance which involves energy input from steam and energy loss to walls and particles, prediction of the temperature of the fastest moving particle (critical particle) by solving the heat conduction equation for a convective boundary condition at the particle-fluid interface, and iteration till the temperatures converge.

Sastry (1986) used a finite element analysis to evaluate thermal process schedules for low-acid particulate foods. For this purpose, heat transfers in the heating and holding sections were mathematically modeled. Simulations were performed to study the effects of RTD, particle size, and h_{fp}. It was found that in the heat exchanger little or no lethality was accumulated at the cold spots of large particles (diameter = 0.0356 cm). However, smaller particles were overprocessed even before reaching the holding tube.

Sandeep *et al*. (1999) used a fourth-order Runga-Kutta Method and computed the fluid and particle velocities. Fluid momentum equations were initially solved in the absence of particles. Velocities of particles were then introduced. Momentum transfer resulting from all of the particles was obtained by particle source term equations which took into account the number of particles, drag coefficient, velocities of fluid and particle, and residence times of particles. A finite element program written by the authors calculated the temperature distribution within the particle throughout the whole system. From the energy balances written for heat exchanger, holding tube and cooling section, temperatures of fluid and particle were determined.

Larkin (1989) used the Crank-Nicolson finite difference method in order to compute the center temperature of spherical particles in an aseptic processing system. The author introduced the hyperbolic function and variable power factor into the semi logarithmic plot of Ball's formula which has been widely used for process evaluation of canned foods. Without any modification, Ball's formula would cause overestimation when applied to aseptic processing of particulate foods. In this study, even though cooling section of the system was modeled, it was not considered in lethality calculations due to possible particle breakups which result in rapid cooling.

Bhamidipati and Singh (1994) developed a simultaneous Dufort-Frankel finite difference solution for coupled momentum and energy balances of a Non-Newtonian fluid and particles flowing in a tubular heat exchanger. For the formulation of particle motion, an Eulerian approach was used, in which the particles were treated as a continuum.

2.2.1. Determination of Fluid-to-Particle Heat Transfer Coefficient

In aseptic processing of particulate foods, there are two steps in product heating, first heat transfer occurs from heating medium to the liquid portion of the product, and then heat is transferred from fluid to the particles. Heat transfer from the fluid to the particles is a function of the thermal and rheological properties of the fluid and relative motion between the particle and the fluid. The boundary layer between the particle and the fluid is represented by a convective heat transfer coefficient (h_{fp}) at the interface (Chandarana *et al*., 1989). The value of h_{fp} depends on many factors such as fluid viscosity and temperature, fluid-to-particle relative velocity, particle size, shape and location (Ramaswamy *et al*., 1997).

A lot of research has been conducted on determination of fluid-to-particle heat transfer coefficient. However, if it was possible to measure the time-temperature history of the cold spot of the critical particle, then measurement of h_{fp} would not be of concern.

A recent method to measure h_{fp} is the use of ablation (Tessner *et al*., 2001). In this method, the change in mass of ice was used as an ablation probe. The amount of solid ablated was directly proportional to the energy flux across the fluid-solid interface with constant h_{fp}.

Ramaswamy and Zareifard (2000) used a calorimetric technique to evaluate the effects of flow rate, viscosity, particle size, and temperature on fluid-to-particle heat transfer coefficient in a holding tube. A calorimeter was used to determine the bulk average temperature of a particle. Knowing the medium temperature, initial particle temperature, residence time, and the bulk temperature of the particle, fluid-to-particle heat transfer coefficient was calculated. An interesting result of this study was the effect of temperature on h_{fp}. It was observed that as the temperature increased h_{fp} decreased. The reason behind this was attributed to the fact that at lower temperature ranges (50 - 70 ºC), particle velocity decreased as temperature increased due to a decrease in fluid viscosity and loss in drag force.

A noninvasive method which was developed by Storofos and Merson (1991) is the use of liquid crystals as temperature sensors. The method involves application of a liquid crystal to the surface of a transducer particle and observing the color change as a function of temperature. The experimental system consisted of a cylindrical acrylic container filled with aqueous suspensions of encapsulated liquid crystals. Color changes of liquid crystal coated particles were determined as the container was heated up from an initial temperature of 20 ºC. There are two basic equations to be solved to determine h_{fp}. The first one is an overall energy balance equation on the container using a lumped parameter approach -- energy exchanged between the heating medium and the liquid in the container is equal to the energy accumulated within the fluid and the particle. The second one is the energy balance on the particles -- convective heat transfer from the liquid in the container to the particles is equal to the energy accumulated within the particles.

Zitoun and Sastry (1994) also used the liquid crystal method in order to investigate h_{fp} between carboxymethlycellulose (CMC) solution and aluminum hollow spherical particles. Flow rate, viscosity, and particle size were the experimental parameters. The authors compared the results of the liquid crystal method to those from the particle tracking velocimetry method. It was found that h_{fp} increased with increasing flow rate, decreasing viscosity and decreasing particle size in both of the methods.

Particle tracking velocimetry was used by Balasabrumaniam and Sastry (1994). This method involves introducing a test particle into the fluid medium in a transparent section and videotaping it as it moves through the system. The video is replayed to determine the relative velocity by finding the time required for a selected tracer particle to pass over a transducer particle. Relative velocity data is used to back-calculate h_{fp} by various dimensionless correlations (Sastry and Cornelius, 2002). An example of a correlation for spherical objects is given below:

$$Nu = 2.0 + 0.6\ Re_G^{0.5}\ Pr_G^{0.33} \tag{2.1}$$

Heppell (1985) used a microbiological method to determine h_{fp}. In this method, spores of *Bacillus stearothermophilus* were entrapped in calcium alginate beads and introduced into the system. The number of surviving spores at the end of the process was determined. Reduction in bacterial count is attributed to a time-temperature history, which could be used to determine h_{fp} by a mathematical model. This method is noninvasive, and can be applied to real foods and processing systems. However, only a single end point microbiological determination is used in constructing the entire time-temperature history, which may result in errors (Sastry and Cornelius, 2002).

Fluid-to-particle heat transfer coefficient may also be determined by direct particle temperature measurement. Stationary particle method involves embedding a thermocouple at a specific location of a transducer particle. Time variable temperature measurements of fluid and particle are necessary to estimate h_{fp} (Maesmans *et al*., 1992). Moving thermocouple method involves attaching a thermocouple to a metallic particle which simulates the food particle. This method overcomes the drawbacks of the stationary particle method. However, it still restricts the rotational motion of the particle (Sastry and Cornelius, 2002).

A new approach to determine h_{fp} is the use of magnetic resonance imaging. Although this method has some advantages in terms of being non-invasive and being applicable to multiple particles, there are some disadvantages such as the need to immobilize the particle and lack of residence time information on the particles which are being monitored (Sastry and Cornelius, 2002).

Chemical and enzymatic indicators have been proposed as alternatives for h_{fp} estimation as well as for process validation (Ramaswamy *et al*., 1996a). Kim and Taub (1993) discussed the possible application of compounds that are produced within the food as a marker to validate the sterility. Three carbohydrate compounds formed within carrots, meats, and juice drinks upon thermal treatment were of interest in this study. It was suggested that monitoring the change in the concentration profiles of these compounds can be used to calculate h_{fp}. In spite of being noninvasive, lack of residence time data and uncertainty of the location of the critical particle may restrict the use of this method.

2.2.2. Kinetics Involved in Thermal Processes

Microbial inactivation is the main purpose of all thermal and non-thermal processing of foods. In order to ensure that the food receives adequate treatment to obtain a commercially sterile product, "process F value" should exceed the "required F value". F value of a heat process is the heating time at a specific temperature to achieve a stated reduction in population of microorganisms. It is calculated by the following equation:

$$F = \int_0^t 10^{(T-T_{ref})/z}\, dt \qquad (2.\ 2)$$

Modeling process evaluation is of great importance since process can only be optimized by integrating microorganism and nutrients/enzyme destruction kinetics with residence time distribution and heat transfer modeling (Ramaswamy *et al*., 1997). One of the earliest works in this area is by Manson and Cullen (1974). In this study, a single point F value, integrated F value within the particulate, and total F value of the container were numerically computed for a system of swept surface heat exchangers. Effects of residence time, flow behavior index of the fluid, particle size, and particle concentration were found to be important in determining process lethality.

Sandeep *et al*. (1999) developed a finite difference program to compute the lethality of a process by taking residence times, particle concentration, and kinetics of the microbial reaction into consideration. Overall heat transfer coefficient and temperatures of the fluid and particles were calculated using another computer program as explained in the same study. Within this study, the effects of h_{fp}, residence time distribution (RTD), particle size, and concentration were determined. It was concluded that an increase in particle size decreased the process F_0 value due to the longer time required to accumulate lethality at the center of the particle. Increase in particle concentration also decreased F_0, because higher particle concentrations resulted in less overall thermal diffusivity, and hence lower heating rate of the mixture. Wider RTD of particles resulted in lower F_0 values. It was pointed out that narrow RTD translated into a more uniform process and higher value of lethality. As fluid-to-particle heat transfer coefficient increased, the center of the particle achieved the required time–temperature combination faster and more lethality was accumulated.

Bhamidipati and Singh (1994) studied the role of thermal time distributions on process effectiveness. A mathematical model was developed to obtain the temperatures and lethality values for fluid elements at different locations in a tubular heat exchanger. It was pointed out that although the lowest F_0 value is important in terms of process safety, the entire lethality distribution will give the information on destruction of nutrients. According to the results, at higher temperatures, spread of the thermal time distribution was higher which resulted in low nutrient retention. It was also found that the smaller the radius of the tube, the narrower the distribution and higher the degree of nutrients retained.

Yang *et al*. (1992) modeled the lethal effects of heat accumulated at the center of particles. Temperatures of the fluid and particle and lethality at the coldest point of particles were calculated based on residence time data. However, the analysis was not applied for multiple particle systems. Distribution of F_0 at the center was found to be wider as the h_{fp} decreased, which did not agree with the results of the study by Bhamidipati and Singh (1994).

Chandarana and Gavin (1989) compared three approaches to scheduling a commercial process for aseptically processed heterogeneous foods. These approaches -- total system, F_0 hold, and hold only were considered. By using an explicit finite difference method, a mathematical model was developed to calculate fluid and particle temperatures and the lethality values accumulated at the center of particles. As far as the hold tube length calculations were concerned, the “total” and “F_0 hold” approaches did not result in significant differences. However, “hold only” approach resulted in a requirement of a longer tube length, since the temperature change and lethality accumulated in the heat exchanger were disregarded. In this approach, the F_0 at the center of a particle could go up to 78 minutes for a target value of 6.0 minutes which would result in overprocessing of the particles. However, as particle size increased, this effect was less pronounced.

In another study by Chandarana *et al*. (1989) the "F_0 hold" approach was used for a thermal process establishment. In this approach, it was accepted that the particle does not accumulate lethality in the heat exchanger but there is a temperature rise. By a mathematical model, effects of particle loading, particle shape, and thermal diffusivity of particles on lethality were studied. Two significant factors -- residence time distribution and h_{fp} were taken into consideration. The effects of these factors were modeled. Resulting required length of holding tubes to reach a predetermined target F_0 value at the center of the particle were presented.

David and Shoemaker (1985) designed a transducer, consisting of a thermocouple and an amplifier, to measure and collect lethality rates of a thermal process. A computerized data acquisition system was used to evaluate the transducer. Output from the transducer was recorded with a strip chart recorder and the area under the recorded curve was integrated to determine the F value of the process. Both the hardware and software of this system was easier to modify for different experimental setups than other commercial data loggers. However, its use for continuous systems needs further investigation.

2.3. Biological Validation of an Aseptic Process

In conventional heat treatment of foods, microbiological methods are a part of process evaluation. However, they can be replaced by heat penetration studies. In aseptic processing, biological validation of mathematical model is crucial to support the validity of the mathematical model and to document the lethality delivered to the slowest heating zone of the particle.

In static systems, particulate lethality can be assessed by placing a thermocouple at the center of the particle and monitoring time-temperature data. However, in continuous systems, use of microbiological or chemical indicators is required to assess the adequacy of the process. Several methods have been developed for process evaluation. Dignan *et al*. (1989) summarized some of these methods. One method is the use of encapsulated bacterial spores in a carrier in a particle of food. In this method, bacterial spores do not come into contact with the food, so intrinsic properties of food such as pH, nutrients, or moisture content cannot affect the rate of degradation of spores. A similar internal environment to that in an actual food particle is achieved when an alginate particle is used for encapsulating. Microorganisms are mixed with the food and immobilized with sodium alginate (Ramaswamy *et al*., 1997). The process lethality can be estimated by evaluating the number of spores before and after the process. The second approach is the use of particles of food inoculated with bacterial spores. The slowest-heating part of the particle is inoculated with bacterial spores and then, after the particles have gone through the process, inoculated particles are recovered. After enumerating the surviving spores, sterilization value of the process is determined. This approach has been used for many years for low-acid canned foods. However, when applied to continuous processing, this method may have some disadvantages -- food particles can disintegrate in the cooling unit or spores can leach from the particles (Dignan *et al*., 1989). One other microbiological method is to develop simulated food particles. Bacterial spores are suspended in particles made from materials such as polymethylmethacrylate or alginate. In this approach, it is critical that calibration tests of the spores must be conducted in a solution that is identical to the one in which they are suspended. In addition, spores must have the same chemical and physical properties such as water activity, redox potential, ionic and nutrient conditions, in both the calibration and validation parts of the study (Pflug *et al*., 1990). Sastry (1988) pointed out that a biological indicator which can be processed and tested for sterility is necessary for reliable microbiological validation of aseptic process. Desired properties of a biological indicator are summarized by the author -- large size (representative of the largest particle size in the system), geometry and thermal properties similar to those of real food particles, physical durability (sufficient to withstand process without disintegration), visual distinguishability from real particles, and shelf stability.

In the case study for condensed cream of potato soup, biological verification of an aseptic process was discussed (CAPPS-NCFST Workshop, 1995, 1996). It was pointed out that the test microorganism should have a D value which is not too low (to ensure there are survivors retrieved) and not too high (to eliminate improbable results). Two inoculation methods were discussed -- count reduction and inoculated pack methods. In count reduction method, a known number of microorganisms are implanted into the center of a food particle. After the particle passes through the aseptic processing system, number of surviving spores is determined, and the lethality of the process is established. In this method, additional labor to collect inoculated particles and enumerate the survivors is required. In the inoculated pack method, relatively larger number of particles is inoculated to ensure that an inoculated particle is in the containers to be incubated. The disadvantage of this method is the longer times required for incubation.

Use of chemical indicators to validate aseptic processing has also been studied. Examining chemical compounds in foods have been shown to be a potential alternative for assessment of the lethality accumulated in the particle. The chemical indicator to be used is chosen according to simplicity and effectiveness of the instrumentation, sensitivity of the indicator to changing environmental conditions, and thermal stability of the indicator at aseptic processing conditions (Ramaswamy *et al*., 1997). Although microbiological verification will finally be required to establish the process, using chemical indicators can save time and serve as a starting point (Bhamidipati and Singh, 1996).

Bernard *et al*. (1990) explained the steps a processor should follow when filing an aseptic process with the appropriate regulatory agency. Once the sterilization process has been established, inoculated packs of product are processed for confirmation of the system. Then, lethality delivered to the system is calculated and compared to the results of the inoculated pack study. Microorganisms which can be used in verification of the process are chosen according to the sterilizing medium -- for superheated steam *Bacillus stearothermophilus* is chosen as the test microorganism.

Bhamidipati and Singh (1996) used a bioindicator system in which food particles were simulated by alginate and potato spheres and the bioindicator was an immobilized *Horseradish peroxidase*. The thermal characteristics of the enzyme were determined by the TDT can method, where cans with special thermocouple attachments were used to measure the temperature of the liquid. Model food particles and immobilized enzyme were subjected to continuous thermal processing. Final peroxidase analyses of this study showed that there was agreement between numerical simulation of heat transfer and destruction kinetics of peroxidase at 110 and 120 °C. However, significant difference was reported at 130 °C. In general, simulations underestimated peroxidase destruction, which may result in safer process conditions. As a result, use of this enzyme was shown to be a good method to verify thermal processing of particles providing reductions in both time and effort compared to microbiological indicators.

Sadeghi (1987) proposed to use encapsulated and immobilized calibration materials to establish the equivalent point method developed by Swartzel (1982). The equivalent point technique describes thermal systems with one time and temperature, independent of an activation energy. In Sadeghi’s study, sucrose solution was immobilized first in calcium alginate spheres and then in polymethylmethacrylate. Immobilized materials were encapsulated in stainless steel and aluminum cylinders. Thermal characteristics of these particles were determined after processing them in retort and obtaining concentration data. Residence time distribution was established based on concentration changes. Further application of this procedure to continuous systems has potential for determination of slowest and fastest moving particles as well as the effects of residence time, shear stress, Reynolds and Nusselt numbers on concentration changes, and hence time-temperature equivalent points and lethality.

Ramaswamy *et al*. (1996) used immobilized bovine pancreas trypsin (type III) to verify a finite difference model developed to determine the fluid-to-particle heat transfer coefficient in a holding tube. The enzyme was immobilized in a steel capsule and embedded in cylindrical potato particles. The retention of enzyme estimated from the finite difference program was compared to the results of actual enzyme assays. Predicted and measured values were in good agreement. As a result, using encapsulated enzyme as a method of verification of heat transfer coefficients was shown to be promising.

Pflug *et al*. (1990) discussed thermal process establishment for both conventional and aseptic processing of low acid foods. Process conditions verified for any in-container sterilization system can be used for other similar systems. However, in continuous treatment of particulate foods, all critical factors are specific for that particular system. In this study, it was emphasized that biological validation methods were not yet complete tools to measure the actual lethality and mathematical models could only be as accurate as the input parameters.

2.4. Residence Time Distribution (RTD)

Residence time distribution (RTD) of particles is one of the two critical factors that play a role in aseptic processing. In a continuous flow system, some of the particles remain in the system longer than others, i.e. particles close to wall or in a dead space move much slower than those traveling along the centerline. Hence, the particles experience a distribution of residence times within the system. Residence time of particles must be as narrow as possible in order to reduce the degree of overprocessing and obtain a uniformly heated product (Ramaswamy *et al*., 1995).

Residence time distributions can be described by E (t) and F (t) functions. The E (t) function gives the RTD of the fluid or particles for any non-ideal flow. A typical E (t) curve will describe the fraction of material in the outlet stream that has been in the system between times t and t+dt. The F (t) function represents the accumulation of particles at the exit with a residence time t or less. F (t) can be obtained by integrating E (t) over time t (Ramaswamy *et* *al*., 1995). Standard deviation of residence times has been used by researchers to characterize RTD and to determine the spread of the distribution.

2.4.1. Techniques to Determine Residence Time

There are several studies which involve determination of residence time distribution of liquid or multiphase media. In most of the multiphase studies, model food particles such as polystyrene spheres, alginate beads, and rubber cubes have been used (Sandeep and Zuritz, 1995; Palazoglu and Sandeep, 2002, 2004; Fairhurst *et al*., 1999; Dutta and Sastry, 1990a, b). There are some other studies performed by using real food particles such as carrot and potato cubes (Ramaswamy *et al*., 1992; Alhamdan and Sastry, 1997; Abdelrahim *et al*., 1997).

There are various types of methods used in determination of RTD. These include optical methods, use of magnetic particles, magnetic resonance imaging, chemical tracers, fractional collection, metal detectors, radioactive tracers, and ultrasonic Doppler velocimetry methods.

Introducing a tracer solution has been widely used in determining RTD of liquids. In a study by Bateson (1971), age distribution of fluid particles (water and starch solution) in a heat exchanger was determined by introducing 20 % saline solution. Milton and Zahradnik (1973) performed their research for a heat-hold-cool system. A pulse type of 60 % sucrose solution was injected as a tracer material in a polyethylene glycol solution which was the carrier fluid. The concentration of sucrose in the fluid leaving the system was analyzed by a polarimeter. Sancho and Rao (1992) used a pulse input of sodium chloride (NaCl) solution to determine RTD of water, sucrose solution, and guar gum solutions flowing in a straight holding tube. The method involved injecting the salt solution with a syringe before the inlet of the holding tube and detecting the conductivity at the outlet. The signals (in mA) obtained from the conductivity transmitter were converted to mV and recorded. Concentration versus time data was obtained. Mean residence time, and variance of residence times were calculated using a FORTRAN program.

Abichandini and Sarma (1987) performed a similar analysis to determine the RTD of water in a scraped surface heat exchanger. A pulse input of saturated NaCl solution was used as the disturbance at the inlet and the response was measured (at periodic intervals of 5 seconds) at the outlet by a conductivity bridge. Similarly, KCl and NaCl solutions and conductivity meter were used by Zhang *et al*. (1990) and Lee *et al*. (1995) to determine the RTD of fluids.

Dickerson *et al.* (1968) used radioactive tracers to determine the residence time of milk products such as cream, ice cream mix, condensed skim milk, chocolate milk, and raw milk. Radioactive iodine, silver iodide crystals, and bacteria stained with iodine were different types of tracers used in this study. Passage of these tracers was detected by two gamma-sensitive scintillation detectors located outside the stainless steel holding tube. Although consistent results were obtained from the three tracers used, this method was not recommended by the authors due to the high cost of instrumentation and loss of the product due to radioactive contamination.

Torres *et al*. (1998) used a dye, methlylene blue, as a tracer in residence time experiments. The flow of water through a double tube heat exchanger and a conventional holding tube was investigated. A pulse of tracer was injected by a syringe and continuous on-line sampling was done at the outlet. A spectrophotometer connected to a sampler recorded the absorbance every second, and from this residence time distribution was obtained.

In an early study by Roig *et al*. (1976), red dye (no. 2) was used to determine the RTD of water in the holding section of a plate heat exchanger. After introducing a step change in the concentration of the dye, samples at the outlet stream were collected at 2 s intervals and analyzed by a colorimeter at 625 and 430 nm. Taeymans *et al*. (1986) also used dyes to determine RTD of liquids containing particles. For this purpose, tracer calcium alginate beads were colored in blue dye before entrapping spores in them.

Hall effect sensors have been used by many researchers to investigate the RTD of particles in carrier fluids. Hall effect sensors consist of a semiconductor device whose ability to pass electrical current is influenced by the presence of an external magnetic device. So, when a magnet moves through the sensing field of the sensor, the output current is detected. Furthermore, with a stop watch, residence time can be calculated if the distance between the two sensors is known (Tucker and Heydon, 1998). This approach was used by Tucker and Withers (1994), Fairhurst *et al.* (1999), and Eliot-Godereaux *et al.* (2001). Although this method was shown to be promising, the primary drawback was its inapplicability to multiple particle systems.

Yang and Swartzel (1991) investigated the use of photo electronic sensors to determine the residence time distributions of particles in a continuous thermal processing system. Photo electronic sensors are capable of detecting objects in air, hence it was also proposed that their application in monitoring particle movements could be successful. Their approach involved using optical beams creating optical grids which divided the cross sectional area of a glass holding tube into several regions. An array of photo-sensors was set up around the transparent glass tube to generate this optical grid. When an opaque particle passed through this optical-grid, light beams were blocked and corresponding receiver was turned off. By analyzing signals turning on and off, particle positions within this cross section were determined. When the distance between two optical grids are known, the velocity or residence time of the particle can be calculated. Their study validated the use of photo-sensors to determine the residence times of single particles. However, the limitations of this method are the need for a transparent pipe and carrier fluid, as well as particles with similar size and shape for easy data processing.

Chandarana and Unverferth (1996) used magnetic tracer particles to determine RTD. The basics of this approach is related to Faraday's law of electromotive induction, which states that if a magnet moves towards or away from an electrical coil, an electric current is induced in the coil. The electric current induces an electromotive force (EMF). This phenomenon was used to detect potato and chicken/alginate particles in which small flexible magnets were embedded. Copper coils were wrapped around stainless steel tubing at the inlet and outlet of heater, holding tube, and cooler. Passage of a magnet through the coils caused a sudden change in EMF signal which was received by the data acquisition system. Presence of a tracer particle was detected by a spike in the EMF recorded. Elapsed time between the spikes in the EMF at the beginning and the end of a section of the system was the residence time of the particle. Primary limitation to this approach is its inapplicability to multiple particle systems since it is unable to distinguish between multiple particles passing through the coils at the same time.

Another technique used by researchers to determine RTD of particles is fractional collection. In a study by Abdelrahim *et al.* (1997), the experimental system consisted of a feed tank to store the working fluid, a particle tank connected to a three way valve, holding tube, and a screen to collect the particles. After introducing the particles in the system through the three way valve, they were collected in the strainer placed at the exit of the holding tube. The time between introduction of particles and the arrival of the first particle at the screen was accepted as the fastest particle residence time. Fluids and particles at the exit were collected at every 20 s and weighed. The concentration of particles at each time interval was used to obtain the residence time distribution.

Visual observation of colored tracer particles was applied by many researchers such as Fan and Fu (1996), Baptista *et al*. (1996), and Sandeep and Zuritz (1994).

Alhamdan and Sastry (1998) used videotaping in order to detect tracer particles flowing through a transparent SSHE and a holding tube. A mirror was used to monitor the particles that were blocked by the shaft of the SSHE. Residence times of the particles were obtained from the readings of the built-in digital stopwatch. Palazoglu and Sandeep (2002) used a digital imaging analysis technique. After videotaping of experiments, images captured from the entrance and exit of the holding tube were digitized by a media converter and downloaded on a computer. Residence time data were obtained by analyzing these images frame by frame.

Some recent studies reveal that use of magnetic resonance imaging or ultrasonic Doppler velocimetry have potential in determining RTD since they provide a full velocity field information (Sastry and Cornelius, 2002). However, more research is required to apply these methods to multiparticle food systems.

2.4.2. Research on RTD

Residence time distribution of liquids and particles depend on many system and process variables. The system variables are diameter, length, and geometry of the test equipment (holding tube or heat exchanger). Rheological properties, flow rate, temperature, and density of the carrier fluid, shape, size, density and concentration of the solid particles are the process variables (Ramaswamy *et al*., 1995). A review of the effects of some of these parameters on RTD is presented below.

A study on RTD of water in a scraped surface heat exchanger (SSHE) was conducted by Abichandini and Sarma (1988). Saturated NaCl solution was injected as a tracer at the inlet and response (electrical conductivity) was measured at the outlet. Dimensionless concentration and time values were plotted on E (t) curves. It was found that as mass flow rate increased, RTD improved. In a SSHE, due to the centrifugal effect of blades, the fluid spreads over the heat transfer surface in the form of a thin film. As the mass flow rate increased, higher turbulence was present for this film of fluid. Another parameter studied was the number of blades. Increase in number of blades resulted in a narrowing of RTD due to the turbulence and subsequent radial mixing. However, since this would increase the power requirements, it was pointed out that an optimum design of the rotor would be required.

Dutta and Sastry (1990a, b) investigated velocity distributions of food particle suspensions in a straight holding tube. Both mathematical modeling and experimental studies were conducted to determine the average and fastest particle velocities. Model polystyrene particles (ρ = 1044 kg/m^3) were used. Low particle concentrations (ranging from 0.2 to 0.8 %) were used. Flow of particles in CMC solution was continuously videotaped. A

FORTRAN program was written to simulate particle motion as affected by buoyancy, Magnus lift force and Saffman lift force. These studies revealed that the most critical factor affecting the particle velocity was the viscosity of the carrier fluid. In the second study two interesting observations were made. Channeling occurred for high particle concentrations, when a large number of slow-moving particles traveling at the bottom caused an increase in the velocity of the remaining fluid and particles. The other observation was the leap frog effect which was described as the mechanism where a fast particle rolling along the bottom approached a slower one and passed over it. The authors concluded that pressure changes in the inter-particle gap caused attraction-repulsion interactions of particles. Depending on the radial position and interaction with other particles, some particles experienced a change in velocity. In some cases, acceleration effects were significant as particles leaped over each other, and decelerations occurred due to settling and collisions with slow moving particles.

Lee and Singh (1991) studied particle residence time distributions of particles in a horizontal scraped surface heat exchanger. Model food particles were potato cubes suspended in a CMC solution. Tracer particles dyed with food colors were introduced to the system. Individual residence times of tracer particles between the inlet and outlet ports of SSHE were recorded. Normalized particle residence times were analyzed as affected by particle concentration, flow rate, mutator speed, carrier viscosity, and particle size. The results showed that increasing carrier fluid viscosity decreased the mean normalized residence times, which was explained by the fact that higher CMC concentrations provided more lifting force, allowing particles to move faster. Mutator speed, particle size, flow rate, and two-way interactions significantly affected mean normalized residence times, but no significant affect was observed for standard deviations, minimum, and maximum residence times.

In another study by the same authors (Lee and Singh, 1992), RTD of potato particles passing through vertical and horizontal scraped surface heat exchangers were mathematically modeled. RTD was correlated to generalized axial Reynolds number, generalized rotational Reynolds number, particle Froude number, viscosity and particle size. In both vertical and horizontal heat exchangers, particle size was the most influential factor. Normalized residence time was more affected by particle size in horizontal SSHE. In vertical SSHE, normalized residence time was a stronger function of particle Froude number.

Ramaswamy *et al.* (1992) developed an apparatus for particle-to-fluid relative velocity measurement. A transparent glass tube was used to monitor the flow of a particle. Experiments were conducted with water and 3 % starch solution at temperatures of 60, 80, and 90 °C. Potato, carrot, and polypropylene particles were used. It was found that the larger particles traveled faster than smaller particles due to a larger cross sectional area being available for high velocity streamlines. Starch solution was found to have a lifting effect on food particles, thereby yielding higher velocities. Velocities of food particles showed a greater variability than that of polypropylene particles. This was attributed to the smooth rolling of inert polypropylene spheres. When multiple particles of different diameters were used, the leapfrog phenomenon that was described by Dutta and Sastry (1990a, b) was observed. Smaller particles traveled as a string of particles, being pushed by the larger particles. Larger particles that were ahead of the smaller particles traveled faster and increased their distance from the smaller ones.

Abdelrahim *et al.* (1993) used a fractional collection technique to collect residence time data of carrot cubes (6 and 13 mm diameter) in a starch solution. The consistency coefficient of the solution ranged from 0.052 to 0.657 $Pa.s^n$ for a temperature range of 80 and

100 °C. Larger particles were found to have higher residence times -- contrary to what Ramaswamy *et al*. (1992) found. Ramaswamy *et al.* (1992) found that larger sized particles traveled faster than their smaller counter parts. This discrepancy could be due to range of parameters such as flow rate or concentration of particles (5 % w/w versus single particle) used. As the temperature increased, the residence time of particles decreased due to increased turbulence as a result of the reduced fluid viscosity. The spread of residence times was characterized by the variance. It was found that variances increased with particle size and decreased with flow rate (15 and 20 kg/min), carrier fluid viscosity, temperature, and holding tube length (Ramaswamy *et al*., 1992).

Sandeep and Zuritz (1994, 1995) studied the effects of several process parameters on residence times of multiple particles in a straight holding tube. Polystyrene spheres were used at 4, 7, and 10 % (v/v). Particles were close to neutrally buoyant with densities of 1,018 and 1,021 kg/m^3. The carrier fluid was CMC solution with consistency coefficient ranging from 0.6 to 1.4 $Pa.s^n$. Equations for dimensionless mean and minimum residence times were developed as functions of particle Reynolds number, particle concentration, and flow behavior index. Correlations for standard deviation of residence times involved particle size to tube diameter ratio. The results revealed that the standard deviation of residence times decreased with increasing flow rate, decreasing viscosity, increasing particle size, and increasing particle concentration. No channeling effects were observed which was opposite to what Dutta and Sastry (1990 a, b) found. This difference was attributed to very homogeneous suspensions due to the high particle concentrations used and the neutrally buoyant characteristics of the particles.

Effects of particle size and flow rate on residence time distribution in a curved section of holding tube were investigated by Salengke and Sastry (1995). Particles (11 and 20 mm in diameter) were used at 20 % (v/v) concentration in CMC solution. RTD was characterized by E (θ) and F (θ) curves. As the particle size increased, E (θ) became narrower, indicating a uniform distribution of particles. This was attributed to the fact that when smaller particles were used, these particles were channeled at or near the bend axis, while others lagged behind as they moved with low velocity streamlines close to the wall at the bend. Normalized residence time distribution curves revealed that as the flow rate increased, the flow profile approached the plug flow characteristics. At this particle concentration (20 %), radial mixing became very effective at high flow rate, and the motion of particles was restricted.

In another study, effects of particle concentration and the radius of curvature of the bend were analyzed (Salengke and Sastry, 1996). Cylindrical particles with a density of 1,130 kg/m^3 were used at particle concentrations of 20, 30, and 40 %. No obvious effect of particle concentration was obtained for standard deviation of residence times. However, normalized residence times were found to increase with increasing particle concentration. Experiments conducted at the tube with a 0.28 m of radius of curvature showed that increasing particle concentration to the highest level caused residence times to decrease. Changing the bend radius had more clear effects on the spread of residence times. Increasing the radius from 0.089 to 0.22 m tended to decrease the standard deviation of residence times. However, further increase in the radius (to 0.28 m) did not have a pronounced effect on standard deviations of residence times.

Baptista *et al.* (1996) investigated the effect of mixing particles with different characteristics (size and density) on RTD. Both single particle (ρ = 1,065 kg/m^3) and multiple particles at a concentration of 1 % v/v (ρ = 1,185 kg/m^3) experiments were performed. At this low particle concentration, it was observed that the standard deviation of residence times increased due to the presence of other particles, especially for smaller particles. This was due to higher dispersions for smaller particles when compared to larger particles at the same concentration. Interactions between high numbers of smaller particles increased the dispersion and consequently resulted in a wider RTD. At low flow rates, mixing different particles increased the dispersion, while no significant effects were observed at high flow rates. This study revealed that single particle studies could not be applied to every situation. Even at low particle concentrations, particle-particle interactions existed and affected the residence times.

Chandarana and Unverferth (1996) pointed out that if a thermal process was designed according to residence time estimates in the holding tube, excessive thermal treatment and loss of texture and nutritional qualities would result. Their study concentrated on an aseptic processing system consisting of a horizontal SSHE heater and cooler, and a stainless steel holding tube. Tracer particles were potato and chicken/alginate cubes with flexible magnets inserted in them. The residence time data was obtained at temperatures as high as 135 -140 ºC. Although a high particle concentration (15 % w/w) was used, no channeling or leapfrog effects were observed. The residence time data fit both Gamma and Lognormal probability distributions. From these distributions, one is able to estimate the residence time of the fastest moving particle.

Fan and Fu (1996) studied the residence time of particles during vertical tubular flow. The radial migration effect caused by gravitational force was eliminated by using vertical holding tubes. Both single and multiple particle (up to 20 %) experiments were conducted with water as the carrier fluid. Reynolds numbers varied between 6,300 and 26,000. Dimensionless mean residence times increased with Reynolds number. This was attributed to using particles that were heavier than the carrier fluid. Dimensionless standard deviations were found to decrease with increasing flow rate. It was observed that in vertical flow, particles moved at different speeds in the upward and downward parts of the tube. They reported that the standard deviation of residence times obtained in the vertical tube arrangement were lower (0.01 to 0.02) when compared to the results obtained from horizontal tubes (0.03 to 0.07) studied by Sandeep and Zuritz (1995). However, more data would be required to verify the benefits of this vertical tube configuration.

Another study in a vertical pipe flow system was reported by Fairhurst *et al.* (1999). Residence times of neutrally buoyant alginate particles in carboxymethylcellulose solution were determined by Hall effect sensors and visual tracers. Particle concentrations ranged from 16 to 55 % (v/v). Other experimental parameters were particle size (5 and 10 mm diameter) and carrier fluid viscosity (apparent viscosity ranged from 29 to 422 mPas). Qualitative observations during the experiments revealed that an annular region of slow moving particles existed near the wall of the tube. For flows where the annular region existed, some particles were observed to stop on the tube wall for a short time before moving downstream. Particle concentration was the strongest factor affecting the mean and minimum residence times. The effects of parameters on maximum residence times were more complicated. Maximum residence time depended mostly on the existence of the annular region. In the absence of the annular region ($Re_p > 0.08$), particle concentration did not have a significant effect on maximum residence time. However, increasing particle concentration increased maximum residence times when there was an annulus ($Re_p < 0.08$).

Sandeep *et al.* (1997) compared the residence time distribution of particles in conventional and helical holding tubes. Transparent conventional and helical tubes of the same length and diameter were used to assess the effects of viscosity, flow rate, particle size, and concentration on RTD. Aqueous solutions of CMC and neutrally buoyant polystyrene particles were used in the experiments. Flow rates were 0.40 and 0.77 kg/s, particle diameters were 7 and 10 mm, particle concentrations were 3 and 10 %, and carrier fluid consistency coefficients were 0.3 and 1.3 $Pa.s^n$. After obtaining the residence time data for 30 particles for each combination of parameters, statistical analysis showed that residence times were normally distributed in every case. ANOVA tests showed that viscosity, flow rate, and particle concentration had significant effects on RT_{mean}. RT_{min} was found to be affected significantly by all parameters. RTD was significantly influenced by flow rate, viscosity, and particle size. By applying multiple regressions, dimensionless equations were developed to predict RT_{mean}, RT_{min}, and RT_{std}. For comparison to conventional holding tube results, non-dimensional mean, minimum, and standard deviation of residence times were used. Increasing particle concentration increased the mean and minimum residence times, but decreased the standard deviation. Increasing viscosity resulted in an increase in standard deviation and a decrease in mean and minimum residence times. Increasing particle size decreased minimum and standard deviation of residence times. Mean residence times in the conventional holding tube were found to be shorter than that in the helical tube due to radial momentum transfer in the helical tube. Standard deviations of residence times were also found to be similar in the two types of tubes, which could be attributed to radial mixing developed in the three 180 ° bends of the straight tube.

Another study by Sandeep *et al*. (2000) was undertaken to model non-Newtonian two-phase flow in conventional and helical holding tubes. For this purpose, a finite difference FORTRAN program using fourth order, four-stage, explicit Runge-Kutta method was written, fluid velocities and particle trajectories were determined. The unique point of this study was that, change in fluid velocities due to presence of particles and collisions between particles and the walls were all accounted for. It was seen that residence times of particles were higher in helical tubes, but RTD was narrower. Increasing particle diameter decreased both residence times and standard deviation of residence times. When the inclination of the tube was increased from 2.0 ° to 4.0 °, more radial mixing was observed which resulted in narrower RTD. As particle concentration was doubled from 10 % to 20 %, a decrease in net cross-sectional area for the flow resulted in channeling of particles and a decrease in their residence times. Another effect of increasing the particle concentration was the obstruction of particle motion which increased the residence time of particles. However, the overall effect was a narrowing of the RTD.

Palazoglu and Sandeep (2004) studied the RTD of particles in a single helical tube. Experimental parameters were flow rate and rheological properties of the carrier fluid, curvature ratio, particle concentration, and type of particle. Carrier fluid consistency coefficient (ranged from 0.05 to 0.20 $Pa.s^n$) did not have a significant effect on RTD. Mean and minimum residence times were found to increase as particle concentration increased from 4 to 12 %. Moreover, RTD was wider for higher particle concentration. This is in contradiction with what Sandeep *et al*. (1997) concluded by increasing particle concentration from 3 to 10 %. Another previously reviewed study revealed no obvious effect of particle concentration within the range of 20 – 40 % (Salengke and Sastry, 1996). RTD of particles was strongly dependent on the curvature ratio of the helical tube. When the tube with a small curvature ratio (0.077) was used, the acrylic particles accumulated near the inner wall at the bottom of the tube. This indicated that the secondary flow was not strong enough to lift these particles up. When the tube with a large curvature ratio (0.143) was used, acrylic particles were found to be distributed uniformly due to the effect of stronger secondary flow.

The effect of holding tube configuration on RTD was studied by Palazoglu and Sandeep (2002). A novel configuration, a double helical holding tube, was used and compared to the previously studied single helical tube. The double helical tube, formed by coiling the tube in shape of a figure-8, was hypothesized to improve mixing by changing the direction of secondary flow. Effect of curvature ratio, flow rate, carrier fluid consistency, and particle concentration were the other parameters studied. Polystyrene, acrylic, and a mixture of the two were used at concentrations of 4 and 12 % (v/v). A digital image analysis technique, which was developed by Simunovic (1998) was used to obtain residence time data. It was found that upon mixing, acrylic particles were lifted up and caught in secondary flow, which decreased the standard deviation of residence times. Increasing flow rate from 1.26 kg/s to 1.51 kg/s was more effective in reducing the standard deviation of residence times of acrylic particles due to the fact that neutrally buoyant polystyrene particles were already more uniformly distributed. RTD became narrower as particle concentration increased. This was contrary to the findings of Sandeep and Zuritz (1995), and Sandeep *et al*. (1997); but in agreement with what Sandeep *et al*. (2000) found. Curvature ratio, which is the ratio of the tube diameter to the coil diameter, was found to be a significant factor that affected RTD. As curvature increased from 0.077 to 0.143, the RTD of acrylic particles became narrower since a stronger secondary flow was able to mix the particles radially.

When the results were compared to those of the single helical holding tube experiments, it was observed that for the smaller curvature ratio the standard deviations of residence times were similar. However, at large curvature ratio, the single helical tube was more effective in improving RTD since the secondary flow was distorted in the double helical tube as the direction of the flow changed from one loop to the next. In this case, the effect of asymmetric velocity profile in the double helical holding tube overcame the radial mixing effect of secondary flow, and resulted in wider RTD.

2.5. Secondary Flow in Helical Tubes

Helical tubes have been widely used in the food industry due to the higher heat transfer rates in them and their compact design. Helical heat exchangers offer high efficiency and uniformity in product heating. However, special care is necessary for the design of this equipment. For proper design of helical holding tubes and heat exchangers, the fluid and particulate flow in these should be investigated.

During flow in curved pipes, the pressure increases monotonically from the inner convex wall of tube to the outer concave wall across any cross-section due to centrifugal force. This centrifugal force exists because of the curvature of the tube (Sandeep and Palazoglu, 1999). A maximum value of the centrifugal force is present along a line parallel to the horizontal axis of the tube. The fluid in the central region, which is moving with a higher velocity, experiences this centrifugal force, and moves towards the outer wall. However, the fluid particles near the top and bottom parts of the tube move towards the inner wall (Palazoglu, 2001). So, two symmetric, counter-rotating vortices are formed. As a result, radial mixing of fluid elements takes place. The flow in the radial direction is known as the secondary flow. Reynolds number is used to characterize the type of flow and rate of heat transfer in straight tubes. In helical tubes, Dean number, given by the following equation, is used to characterize secondary flow:

$$\mathrm{Dn} = \mathrm{Re}\sqrt{d / D} \tag{2.3}$$

In this equation, d/D is the curvature ratio of the tube, where d is the inside diameter of the tube, and D is the coil diameter between the centers of the tube.

Dean and Hurst (1959) defined secondary flow as the motion that accompanies the primary motion along the line of a curved pipe. This secondary motion decreases the rate of flow produced by a given pressure gradient and causes an outward movement of the region where the primary motion is greatest. The fluid moving at the center of the pipe requires a smaller pressure gradient to balance its centrifugal force. So, the fluid at the center moves outwards, whereas the fluid near the top and bottom moves inwards. As a result of this secondary flow, the region of maximum velocity in the main stream is shifted towards the outer wall of pipe. In addition, frictional energy loss near the wall increases considerably when compared to friction loss in straight pipes (Barua, 1963).

Koutsky and Adler (1964) investigated the effect of secondary flow in minimizing axial dispersion in helical tubes. Three mechanisms affecting axial dispersion were discussed. The first mechanism, non-uniform velocity profile, causes fluid elements to visit different locations and to experience different axial velocities. Therefore, axial dispersion is promoted. However, this mechanism is more effective in the laminar regime, since flattening of the profile reduces this effect in turbulent flow regime. In the second mechanism, random mixing due to diffusion or turbulent eddies is considered. This effect reduces axial dispersion by causing each fluid element to spend almost the same amount of time in the tube. The last mechanism is the secondary flow which inhibits the axial dispersion by continuously changing the position of fluid elements. At low flow rates (Re < 300), axial dispersion in the helical tube was similar to that in the straight tube. Up to a Reynolds number of 3,000, axial dispersion in the helical tube was considerably less. Between Reynolds numbers of 3,000 and 7,000, transition from laminar to turbulent flow regime occurred in the helical tube. The transition to turbulent regime was found to be more gradual for small radius of curvatures. As the Reynolds number increased to 40,000, the axial dispersion in the helical tube decreased greatly in comparison to that for the laminar regime. In addition, axial dispersion in the helical tube in this Reynolds number range was much smaller than that in the straight tube.

Dravid *et al*. (1971) studied the effect of secondary flow on heat transfer in helical tubes. In numerical and experimental studies, only laminar flow regime and Dean numbers above 100 were considered. The development of the thermal entrance regime in the temperature profile was studied. It was concluded that the thermal boundary layer was not affected by the secondary flow at very short distances from the entrance. The ratio of heat transfer coefficient in the coiled tube to that in the straight tube varied as $De^{1/6}$ near the entrance region. As the distance from the tube inlet increased, this ratio varied as $De^{1/2}$. However, increase in heat transfer coefficient in the helical tube flow was not continuous -- cyclic oscillations were observed. This was attributed to the fact that even at high Dean numbers and far from the entrance, not all the fluid elements were well-mixed.

Austin and Seader (1973) developed a numerical solution to the equations of motion in order to compute the velocity and pressure profiles over a wide range of Dean numbers. Their study involved Dean numbers from 1 to 1,000, and Reynolds numbers from 10 to

4,000. The axial velocity component of the equation of motion and stream function of secondary flow were used in the numerical method. At low Dean numbers, secondary flow streamlines were symmetrical and the axial velocity profile in the helical tube was found to be similar to that in a straight tube. As the Dean number increased, non-convex regions were developed in the secondary flow profile due to distortion of symmetric streamlines.

Secondary flow and natural convection effects on heat transfer coefficient were studied in helically-coiled tubes under laminar flow conditions (Abul-Hamayel and Bell, 1979). Natural convection effect becomes an important factor in heat transfer at low Reynolds numbers. In this case, the gravity force due to density difference causes vertical movement of the fluid. If the fluid near the wall is heated, the colder or heavier fluid at the center will move towards the bottom and the fluid at the bottom will move upwards. This study revealed that at high Reynolds numbers (Re = 937, Re = 2,761), heat transfer coefficient was highest near the outer wall, and lowest at the inner wall. This was attributed to the effect of symmetric secondary flow patterns and lack of natural convection. However, at a lower Reynolds number (Re = 73), natural convection was found to dominate the secondary flow effect. This study suggested that high heat transfer rates in helical tubes result from two mechanisms -- secondary flow or natural convection, depending on Reynolds number.

Flow visualization studies were conducted by Cheng and Yuen (1987a) to observe secondary flow patterns at the exit of a 180 ° bend and in the downstream region of a straight tube. Flow visualization was done by injecting smoke into air, and recording the experiments using a video camera. Reynolds numbers in their study ranged from 283 to 1,807. It was concluded that at the exit of the 180 ° bend and for Dean number ~ 100, onset of centrifugal instability occurred. This instability was observed in the form of an additional pair of vortices near the central outer wall region.

Another study by Cheng and Yuen (1987b) dealt with secondary flow patterns in an isothermally heated curved pipe. The combined effects of centrifugal and buoyancy forces in the hydrodynamically and thermally developing entrance regions were investigated. Vertical curved pipe with an upward flow, horizontal curved pipe flow, and vertical curved pipe with a downward flow were of interest. Dean vortices were observed only for downward flow in the vertical curved pipe. Secondary flow patterns in the horizontal pipe were found to be skewed, since buoyancy and centrifugal forces were perpendicular to each other. The distortion of secondary flow in the curved pipes was due to the buoyancy effects. It was pointed out that the symmetry of secondary flow pattern could only be maintained if the viscosity effect was greater than the body force effect.

Austen and Soliman (1988) investigated the effect of pitch on heat transfer in helically coiled tubes. As pitch, h, increases for the same coil diameter, D, centrifugal force decreases and secondary flow weakens, thus the flow approaches the straight tube flow. The results of this study showed that the effect of pitch appeared at low Reynolds numbers and diminished as Reynolds number increased. This was attributed to free convection, which was the dominating effect since secondary flow was weaker at low Reynolds numbers.

Single and two-phase (gas-liquid) flows through helically coiled tubes were studied by Hart *et al*. (1988). Friction loss and pressure drops in the axial and radial directions were investigated. The secondary flow has a stabilizing effect on laminar fluid flow which results in higher critical Reynolds number--at which transition from laminar to turbulent flow regime occurs. This critical Reynolds number in helical tubes can be obtained from the following equation (Srinivasan *et al*., 1968):

$$Re_{crit} = 2100\ [1+12\ (d / D)^{1/2}] \tag{2.4}$$

The results of the study by Hart *et al*. (1988) suggested that in the helically coiled tube, the increase in pressure gradient is much lower for two-phase flow, as compared to that for single-phase flow.

2.5.1. Chaotic Advection Flow

Chaotic advection is defined as the production of chaotic particle paths in the laminar regime. Chaotic advection, or Lagrangian chaos, is a flow regime in which chaos is generated in the physical space. In this regime, fluid elements follow irregular trajectories different from the Eulerian streamlines (Chagny *et al.*, 2000).

Chaotic advection flow has been of interest to many researchers. Raynal and Gence (1997) compared chaotic mixing in laminar flow and turbulent flow from an energy saving point of view. Mixing time in chaotic laminar flow was defined as the time taken for the concentration gradient created by advection to be smothered by molecular diffusion. Product of mixing time and energy dissipated was used as a criterion to compare chaotic mixing to turbulent mixing. At high Schmidt number (~10^4 at least), which is the ratio of momentum diffusivity to mass diffusivity, chaotic mixing in laminar regime resulted in lower power requirements than turbulent mixing.

Chagny *et al*. (2000) compared helical and chaotic flow in heat exchangers having the same heat transfer area and same number of bends. For this purpose, they used a criterion, which involved thermal performance and energy expenditure. A curved duct, which was formed by rotating each bend 90° with respect to the neighboring bend, was used in the study. The rotation of the bends resulted in reorientation of centrifugal force and thus, in chaotic advection. Temperature and pressure losses in the helical and chaotic heat exchangers were measured. Reynolds numbers ranged from 30 to 30,000. It was found that at low Reynolds numbers, heating was more homogeneous and heat transfer was intensified in the chaotic advection regime, with no increase in energy expenditure. At high Reynolds numbers, there was no difference in thermal performance between the helical and chaotic configurations. Flow visualization studies done by laser-induced fluorescence showed that at low Reynolds numbers, mixing was better in the chaotic configuration, which explained the improvement of the homogeneity and heat transfer in this configuration. This study showed that using chaotic configuration in heat exchangers would be more economical regardless of Reynolds number.

Acharya *et al*. (2001) numerically analyzed the enhancement of heat transfer in two different coiled tube heat exchangers, one with regular mixing and the other with chaotic mixing. Chaotic mixing was achieved by periodic rotation of the coil axis, with no additional flow obstructions. It was pointed out that in straight tubes, physical obstructions located periodically in the flow path can generate repeating developing flow patterns. However, in coiled tubes, a transverse flow is already present in the fully developed flow regime and repeating flow patterns are easily generated by modifying the direction of centrifugal force. Alternating the geometry of the axis of the coil resulted in a 7 - 20 % enhancement in the heat transfer coefficient as compared to the constant axis coil geometry over a Reynolds number range of 50 - 1,200. The enhancement in the heat transfer coefficient was found to increase uniformly with Prandtl number (ratio of momentum diffusivity to thermal diffusivity). In addition, the angle of the rotation of the coil axis had an optimum value of about 120°.

Lemenand and Peerhossaini (2002) developed a mathematical model to predict Nusselt number as affected by number of bends (3 – 33), Reynolds number (100 – 300), and Prandtl number (30 – 100) in chaotic flow. Helical and chaotic configurations of heat exchangers were compared. The chaotic configuration was composed of a succession of quarter–circle bent tubes with a 90 ° angle of rotation of the curvature plane. In chaotic configuration, Nusselt number increased with the number of bends at low Reynolds numbers (Re < 150). However, Nusselt number did not evolve beyond seven bends. After six bends, a homogeneous mixing was obtained due to a destruction and reorganization of the secondary flow structure at each bend rotation. Dependence of Nusselt number on Reynolds number in chaotic tube was lower than that in the helical tube. This was explained by the fact that mixing in the chaotic geometry was more dependent on chaotic trajectories, rather than secondary flow.

Lefevre *et al.* (2003) reviewed the studies performed on chaotic advection in real Stokes flow, i.e., flows governed by viscous forces. In real Stokes flow regime, although the flow is laminar and dominated by viscous forces, fluid particle trajectories are chaotic due to a change in geometry. It was pointed out that two-dimensional periodic and three-dimensional real flows can be used to induce chaotic trajectories in industrial processes in batch and continuous operations.

2.5.2. RTD in Helical Tubes

For two phase flow in bends, one would expect to see a highly complicated flow behavior. Velocity profiles and residence time distribution of particles are different in helical tubes than those in straight tubes. Toda *et al*. (1972) investigated the flow of suspensions of polystyrene beads in through horizontal and vertical pipe bends. In a horizontal pipe bend, at low velocities particles were lifted up along inside the wall, whereas in straight pipe flow, particles were observed to move along the centerline of the pipe. At higher velocities, particles were transported towards the outside wall of the pipe bend due to a strong centrifugal force (Toda *et al*., 1972). It is expected that the presence of secondary flow in helical tubes or in bends of straight tubes will enhance mixing of particles and result in narrow residence time distributions in holding tubes.

Ahmad *et al*. (1993) studied concentration profiles of solid particles in multisized particulate slurry flowing through a 90 ° pipe bend. They observed that at a lower velocity and concentration level, concentration profile in the bend of the vertical plane was asymmetric but less asymmetric than that in straight pipe. In the horizontal plane, the concentration distribution was uniform. However, it was slightly distorted at the inner wall of the bend due to separation of particles resulting from strong secondary flow. In the vertical plane, asymmetry of the distribution was found to reduce with decreasing particle size. In the horizontal plane, all particles except the largest ones appeared to be uniformly distributed. Largest particles were observed to migrate very close to the inner wall of the bend. This study gave an insight of particle motion and redistribution through bends.

Salengke and Sastry (1996) investigated the effect of bend curvature on residence time distribution of cylindrical particles at concentrations of 20, 30, and 40 %. Normalized residence times of particles slightly increased as the radius of curvature increased from one level to another (0.089, 0.22, and 0.28 m). Visual observations and video analyses revealed that the outward migration, expected to push and accelerate the particles radially, was restricted by high solid concentrations. So, it was presumed that the acceleration due to centrifugal force caused the particles to speed up in the axial direction. This presumption was justified by the presence of Magnus effect and particle interactions which tend to keep particles away from the outer wall. Standard deviation of normalized residence times decreased as the radius of curvature increased from 0.089 to 0.22 m. A further increase to 0.28 m did not result in a significant change in the standard deviation of residence times. As the radius of curvature changed, the length of the paths traversed by the particles also changed. As the length of paths decreased, the amount of time particles spent in the bend, decreased. Smaller bends resulted in higher residence time differences between the particles which moved with high velocity near the tube axis and those which moved with low velocity near the tube wall.

Liu and Zuritz (1995a) modeled solid-liquid two phase flow in the bend section of a straight tube. Residence time of spherical neutrally buoyant particles (10 % v/v) was also taken into consideration. In the first straight section before the bend, axial velocity contours of the fluid were symmetric and the particles had a slight influence on the axial velocity profile. However, radial positions of particles started to change upon entrance at the bend. At the bend section, secondary flow carried the fluid from the inner to the outer wall and axial dispersion was suppressed. Velocity profiles of the fluid were affected by the presence of particles due to a large momentum transfer between the two phases. Secondary flow shifted the maximum fluid axial velocity region towards the outer wall, and particles disturbed this asymmetric profile even more. After the bend section, during flow in the second straight section, the effect of the bend section was still observed. The radial pressure gradient set up to balance the centrifugal force propagated downstream away from the bend. However, after the effect of bend was lost in the straight section, axial dispersion started to develop again, which resulted in wider residence time distribution of particles.

Another study by Liu and Zuritz (1995b) showed that the residence time distribution of particles in helical tube flow was narrower than that in a conventional holding tube which was used in their previous study. The simulations performed showed that secondary flow existed in the entire flow due to the continuous curvature of the tube. Secondary flow reduced the axial dispersion of particles by varying their radial locations. Hence, residence time distributions in helical tube were narrower than that in the straight tube. The ratio of mean residence time to minimum residence time was 1.04 in the helical tube, whereas it was 1.26 in the conventional tube. It was also concluded that particles spent more time in the helical tube (mean residence time in helical tube was 21.3 s, in conventional tube this was 17.3 s).

Grabowski and Ramaswamy (1998) found that the linear velocity of particles in the curved section (180°) of a straight holding tube was reduced by 8 % for a larger radius of curvature (0.2 m) and 15 % for a smaller radius of curvature (0.1 m) compared to velocities in a straight section. The effects of particle size, carrier fluid viscosity, and radius of curvature of the bend on particle velocity were investigated. Centrifugal force in the curved section was believed to increase particle-wall and particle-particle interactions in addition to inducing some turbulence to the flow. The resulting force between the particle and the tube surface in the bend was given by:

$$F_s = [(F_g - F_b)^2 + F_c]^{0.5} \tag{2.5}$$

where F_c is the centrifugal force ($F_c = m_p u_p^2 / R$), F_g is gravitational force, and F_b is buoyancy force. Experimental results indicated that the effect of curvature was more pronounced when carrier fluids of low viscosity were used. The linear velocity of the particle at the bend was more affected at lower viscosity than higher viscosity. The centrifugal force, which was expected to reduce the linear velocity of particles, was found to have a greater effect on the particles of larger sizes (15 mm compared to 10 mm of diameter). In a straight tube, increasing particle size would increase the particle velocity since there would be more surface area of the particle available for high velocity streamlines. This increase in particle velocity is expected to be higher in curved section due to the drag action of the centrifugal force. This study gives insight on the effect of curvature ratio on the particle velocity. However, the results may not be applicable to all situations since experiments were conducted with a single particle.

Sandeep and Zuritz (1999) investigated the residence time distribution of neutrally buoyant polystyrene particles in a holding tube system which consisted of four straight sections connected by 180° bends. The effect of parameters such as flow rate, viscosity, and particle concentration were determined. At higher flow rate (0.8 kg/s), particles traveled closer to the center of the tube than the tube walls. This was attributed to the higher degree of radial mixing at the high flow rate which caused particles to travel in locations not corresponding to their neutrally buoyant characteristics. An increase in flow rate, particle size, or particle concentration resulted in a decrease in the standard deviation of particle residence times. Viscosity was the only factor which showed the opposite effect. It was found that RTD was always narrower in the 4th tube section. As the particles passed through the first 3 bends, they experienced radial mixing, so RTD became narrower after each bend section. If more bends were used, the final RTD would be much narrower due to stronger secondary flow.

A study was undertaken by Castelain *et al*. (1997) to compare residence time distributions of a fluid in a helically coiled system and a chaotic system. The objective of this study was to investigate axial dispersions in the helically coiled and chaotic pipes for Reynolds numbers between 800 and 13,500. Axial dispersion plug-flow model developed by Trivedi and Vasudeva (1975) was used. This model was based on two different flow zones -- predominant regular flow at the inlet and fully developed chaotic zone which prevails after a few bends. Residence time distributions were determined experimentally by a conductimetric method with two measurement points. Pure water was used as a carrier fluid. The concentration of an injected tracer (NaOH solution) was measured, as a function of time, at the inlet and outlet of the pipes. In the straight pipe flow, in the laminar regime, mixing is induced by molecular diffusion which is a slow mechanism and causes wide residence time distributions due to high velocity gradients. The flow in a helically coiled system provides regular mixing. However, if chaotic mixing is present, irregular mixing will exist due to chaotic trajectories induced. In chaotic tube, there are two mechanisms that result in lower axial dispersion -- development of secondary flow and modification of the velocity field by reorientation of the direction of centrifugal force. Both the helical and chaotic systems in this study were made up of 33 bends. In the chaotic system, the plane of curvature of each bend had a 90° angle with that of its adjacent bend. In the helical system, an increase in Reynolds number resulted in an increase in the Peclet number (ratio of convectional mass transfer to diffusion mass transfer). Experimental tests were conducted with 3, 9, 15, 27, and 33 bends. As the number of bends increased, a slight increase in axial dispersion coefficient resulted, which was attributed to the effect of the straight sections between the bends. In the helical tube, transition from laminar to turbulent regime occurred at Reynolds number values of 8,000 - 10,000. After this transition, the turbulent regime dominated the effect of secondary flow and dispersion increased with further increase in Reynolds number. The chaotic system was found to be similar to the helical system in terms of critical Reynolds number at which the transition occurs. For Re < 2,500, axial dispersions in the two systems were similar. However, as Reynolds number increased further, the residence time distribution narrowed in the chaotic configuration due to the chaotic advection effects. To compare the efficiency of these two systems as mixing devices, Peclet number was plotted against Reynolds number. Pressure drop or energy consumption were not taken into account. In addition, for Re < 2,500, the agreement with the plug flow axial dispersion model and residence time data was less reliable for the chaotic system (tails of the residence time distribution curves were wider). This study revealed that the use of the chaotic system would result in narrow residence time distributions and more uniform processing and not cause problems related to cleaning, mechanical damage or extra pressure drop as in the case of static mixers, blades, vans or fins.

Castelain *et al*. (2000) investigated mixing in a spatially chaotic flow. Experimental characterization was done by means of residence time distribution determination of fluid elements. The residence times were obtained by using the two-measurement conductrimetric method as explained in the previous study. Instead of water, saccharose solutions were used as working fluids. Mixing in the chaotic system was determined for Reynolds number range of 30 – 1,700. In addition, pressure drop was taken into consideration to characterize the efficiency of the chaotic system. For this purpose, a mixing criterion was defined as follows: $C_r = \nu / D_{ax} C_f$, where C_f is the friction factor. For the same pressure drop, the criterion increased when the axial dispersion decreased. For the same axial dispersion, the criterion increased when the pressure drop decreased. In general, Peclet number was found to be larger in the chaotic system than that in helically coiled system, which indicated a smaller axial dispersion in the chaotic system. This was in agreement with the results of a previous study by Castelain *et al.* (1997) which was conducted for larger Reynolds numbers ($Re > 800$).

SYMBOLS

C_f	friction factor	
C_r	mixing criterion	
D	diameter of coil	m
d	diameter of tube	m
D_{ax}	axial dispersion coefficient	m^2/s
Dn	Dean number ($\mathrm{Re}\sqrt{d/D}$)	
D-value	decimal reduction time	min.
F	sterilization value	min.
F_b	buoyancy force	N
F_c	centrifugal force	N
F_g	gravitational force	N
F_s	resultant force	N
h	pitch of helical tube	m
h_{fp}	fluid-to-particle heat transfer coefficient	W/m^2-K
Nu	Nusselt number ($h_{fp}\, L_c / k_f$)	
m	mass of particle	kg
Pr	Prandtl number ($c_p\, \mu / k_f$)	
r	radius of particle	m
R	radius of tube	m
Re	Reynolds number ($\rho_f\, u\, d / \mu$)	
t	time	min.
u	velocity	m/s
z-value	temperature change required to change D-value by a factor of ten	K

Greek Letters

ρ	density	kg/m^3
ν	kinematic viscosity	m^2/s

Subscripts

crit	critical
g	generalized
min	minimum
p	particle
ref	reference
std	standard deviation

Abbreviations

CMC	carboxymethlycellulose
RT	residence time
RTD	residence time distribution
SSHE	scraped surface heat exchanger
TDT	thermal death time

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Chapter 3

MANUSCRIPT I

Residence Time Distribution of Multiple Particles in Straight, Single Helical, and Double Helical Holding Tubes

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Paper No. FSR-04-18 of the Journal Series of the Department of Food Science, North Carolina State University, Raleigh, NC 27695-7624.

ABSTRACT

RTD of polystyrene and acrylic particles in three different holding tube configurations were studied. Straight, single helical, and double helical holding tubes were used. The effects of process parameters on RTD of particles were determined. These process parameters included flow rate, particle type (mixing of particles), carrier fluid viscosity, and particle concentration. Holding tube configuration and flow rate were found to be the significant parameters affecting RTD of particles. Particle type, carrier fluid viscosity, and particle concentration had significant effects on RTD for individual holding tube configurations and flow rates. The narrowest RTD (the lowest RT_{std}) of particles was obtained in the single helical holding tube. Increasing flow rate resulted in narrower RTD of particles, the effect being more pronounced for acrylic particles. Upon mixing polystyrene and acrylic particles, RT_{std} of polystyrene and acrylic particles increased in the straight and single helical holding tubes, whereas an opposite effect was found in the double helical holding tube.

Keywords: residence time distribution, holding tube configuration, multiphase flow

INTRODUCTION

Determination of residence time distribution (RTD) of particles is one of the key factors in the establishment of aseptic processing for low-acid particulate foods. Aseptic processing offers many advantages such as improved quality of the product and less labor and time required for the process. Although liquid foods have been easily adapted to continuous thermal processing, there are some hurdles in the application of aseptic processing to foods containing particles. Since accurate temperature measurements within freely flowing particles in the heat-hold-cool system is practically impossible, mathematical modeling of heat transfer to the center of the particles is necessary for process design (Sastry, 1986). Moreover, simulated results should be supported by experimental verification of commercial sterility (Bhamidipati and Singh, 1996). The most notable uncertainties during establishment of particulate food processing are residence time distribution of particles and the convective heat transfer coefficient at the solid-liquid interface (Chandarana *et al*., 1989).

Introduction of particles into the fluid causes relatively complex flow conditions, which will in turn result in residence time distributions different from the fluid portion of the product. Since the primary goal of any thermal treatment is ensuring the safety of the product, residence time of the fastest heating particle should be known to determine process lethality. However, in order to maximize the quality advantages of continuous high temperature short time processing, special attention must be paid to obtain uniform heat treatment, thus minimizing overprocessing of slower moving particles.

Uniform heat treatment is possible by making sure that the RTD of particles in the system is as narrow as possible. The residence time (RT) of particles is affected by many factors and can be grouped as carrier fluid and food particle factors. The carrier fluid factors include type, density, flow rate, rheological properties; and food particle factors include type, size, shape, density, and concentration. Furthermore, system parameters play a very important role in determining the fastest residence time and RTD of particles. These parameters are holding tube length, diameter, configuration, and type and orientation of holding tubes and heat exchangers (Ramaswamy *et al*., 1995).

Researchers have investigated the RTD of particles in heat exchangers as affected by the processing and system parameters. RTD of fluid elements in scraped surface heat exchangers (SSHE) have been studied by Abichandini and Sarma (1988) and Lee *et al*. (1995). RTD of food and simulated particles were found to be narrower when high flow rate, high shaft rotational speed, and low particle concentrations were used in a horizontal SSHE (Alhamdan and Sastry, 1998). Abdelrahim *et al*., (1993) pointed out that in a pilot scale aseptic processing system the distribution of particle residence times widened with an increase in particle size and narrowed with an increase in flow rate, holding tube length, and carrier fluid concentration.

Although heat exchangers contribute to the lethality at the center of particles, FDA gives credit to the lethality accumulated only in the holding tube. So, many studies have been performed on RTD in holding tubes. In straight holding tubes, Sandeep and Zuritz (1994) concluded that an increase in flow rate or particle concentration decreased mean and minimum residence times and standard deviation of residence times. Sandeep and Zuritz (1995) also found that for non-Newtonian flow, an increase in carrier fluid viscosity increased the standard deviation of particle residence times. Alhamdan and Sastry (1997) investigated the RTD of actual and simulated food particles and found out that RTD curves became narrower with decreasing particle concentration, increasing flow rate, and using more viscous fluids. Baptista *et al*. (1996) determined the effect of mixing particles with different characteristics on RTD of particles. In single type particle flows, the critical particle, on which the lethality calculations are based, is the fastest moving particle. However, for multiple types of particles, the critical particle, which receives the least heat treatment, may not be necessarily the fastest moving particle. Large particles of low density were mixed with small particles of high density. It was found that when different types of particles were mixed, a reduction in mean and minimum residence times was observed. However, standard deviation of residence times increased upon mixing. This study showed that if different types of particles are present in a fluid, RTD results obtained from a single type particle study cannot be applied (Baptista *et al*., 1996).

Helical tubes have gained attention due to their high heat transfer rates and improved mixing of the product. For a flow in a curved pipe, the primary motion along the line of the pipe is accompanied by a secondary motion (Dean and Hurst, 1959). This secondary motion in the radial direction occurs due to a pressure gradient across the tube, which results from the centrifugal force due to the curvature (Dean, 1927, 1928). Secondary flow manifests itself as two counter rotating vortices and is characterized by Dean number which is the ratio of centrifugal forces to viscous forces. Secondary flow is known to decrease the axial dispersion of fluid elements by continuously changing their radial positions (Koutsky and Adler, 1964). Hence, it is expected that RTD of particles will be narrower if helical holding tubes are used. Studies on RTD in the bend sections of straight holding tubes have been conducted by several researchers (Ahmad *et al*., 1993; Liu and Zuritz, 1995a; Salengke and Sastry, 1995, 1996; Grabowski and Ramaswamy, 1998; Sandeep and Zuritz, 1999). Numerical and experimental studies on RTD of particles in helical tubes reveal that axial dispersion of particles was greatly reduced and RTD of particles became narrower as compared to that in conventional straight tubes (Liu and Zuritz, 1995b; Sandeep *et al*., 1997). Recent studies in the area focus on the effects of holding tube configuration and curvature ratio on RTD of particles. Palazoglu and Sandeep (2002) compared RTD of particles in single and double helical holding tubes (double helical holding tube was built by coiling the tube in shape of a figure-8 around two cylindrical mandrels). It was found that the single helical tube configuration was more effective in reducing residence times when small tube curvature ratios were used. In the double helical tube configuration, secondary flow streamlines were distorted at the straight regions between any two loops and the effect of secondary flow on narrowing RTD was lessened. Increasing the curvature ratio of the helical tube resulted in narrower RTD of particles (Palazoglu and Sandeep, 2004). In the same study, mixing two types of particles of different densities (polystyrene and acrylic particles) was found to decrease the residence times of high density (acrylic) particles.

Current literature reveals that there are still many uncertainties about RTD of particles that need further research. Particle-particle interactions in multiphase flow, mixing different types of particles, and testing the effects of parameters in various types of holding tubes have not yet been investigated in detail. So, the objective of this study is to determine the effects of process parameters (flow rate, carrier fluid viscosity, particle type, and particle concentration) and a system parameter (holding tube configuration) on the RTD of particles.

MATERIALS AND METHODS

Experimental setup

The experimental setup consisted of a holding tube, a positive displacement pump, and a reservoir (Figure 3.1). Holding tubes (straight, single helical, and double helical) used in this study were 15.24m long, 0.0508m i.d. transparent tubes (Cleargard, CSC200, M.G. Newell Corp., Greensboro, NC). The straight holding tube assembly consisted of two straight sections connected by a 180° bend. The tube was inclined upwards of a quarter inches per foot length of the tube according to FDA regulations. The single helical tube was built by coiling the transparent tube around a cylindrical mandrel (diameter = 0.355 m). The resulting curvature ratio (the ratio of the tube diameter to coil diameter) was 0.143. The double helical holding tube was coiled in shape of figure-8 around two cylindrical mandrels of the same diameter. The pitch of the helical tubes, defined as the vertical rise per unit length, was 0.076 m.

A positive displacement pump (Model 45U2, Waukesha Cherry-Burrell, Delavan, WI) was used to circulate the carrier fluid between the reservoir and the holding tube. Particles were introduced to the system through a perforated cone assembly inside the reservoir. Particles did not pass through the pump, instead they were bypassed from the pump by the perforated cone assembly to a point at the exit of the pump.

Carrier Fluid

The carrier fluid used in this study was carboxymethylcellulose (CMC) solution (CMC is a typical viscosifier used in food industry). CMC powder (6000 C, TIC Gums, Belcamp, MD) was dissolved in tap water, and after hand-mixing the CMC –water mixture for some time, it was pumped through the system to make a more homogeneous solution. The solution was stored for 6 -10 hours to obtain a clear solution free of air bubbles. The final CMC solution had density of 1,001 kg/m³. Particles were introduced to the system after the flow of solution reached steady-state conditions.

Rheological characterization of CMC solutions was done using a Haake Viscotester (Haake VT 550, sensor MV-DIN, Haake Inc., Paramus, NJ) before and after each run. A bob and cup geometry was used with this instrument. Rheological properties such as shear stress, shear rate, and apparent viscosity were obtained at room temperature. The data was fitted to the Oswald De Waale model ($\sigma = K\ \dot{\gamma}^{n}$). Flow behavior index (n) and consistency coefficient (K) values were used to control viscosity range of the solution at which the experiments were conducted. These values are shown in Table 3.1 with other range of experimental parameters. Apparent Viscosity of CMC solution ranged between 0.025 – 0.108 Pa.s at shear rates of 0 -100 1/s. Due to a slight decrease in viscosity with time, addition of CMC powder was required between each run.

All experiments in this study were conducted at room temperature and under laminar flow conditions. Particle Reynolds numbers ranged from 117 to 835. Dean numbers in the single and the double helical holding tubes ranged from 44 to 290.

Model Food Particles

Model food particles used in this study were polystyrene and acrylic spherical particles (Precision Plastic Ball Co., Franklin Park, IL). All particles had a diameter of 0.014 m. Some of these particles were used as tracer particles by painting them with different colors and patterns for easy identification. Density measurements were performed on these tracer particles by using a gas pcynometer (Model Accupcy 1330, Micrometrics Instrument Corporation, Norcross, GA). Polystyrene particles were closer to neutrally buoyant ($\rho = 1{,}012 \pm 9$ kg/m^3) than acrylic particles ($\rho = 1{,}169 \pm 18$ kg/m^3). Either single or mixed types of particles (both acrylic and polystyrene particles) were used at concentrations of 4 % and 12 % (v/v) in the experiments.

Digital Image Analysis

Residence time data of tracer particles were obtained by the digital image analysis method developed by Simunovic (1998). In this method, the experiments are continuously recorded using a camcorder (DCR-TRV103, Sony Corporation, New York, NY) and two industrial color CCD cameras (Toshiba, IK-M43, New York, NY) located at the entrance and the exit of the holding tube. A character generator (Panasonic, WJ-KB30, Matsushita Communication Industrial Co., Ltd., Japan) with a built-in stop watch function was used to display time (resolution: 0.01 s) on the recordings. A digital video mixer (Videonics, MX-1, Campbell, CA) was used to combine the images from the entrance and the exit of the holding tube. The recordings from the camcorder were downloaded onto a computer using a video capturing software (DV Gate Motion, Sony Corporation, New York, NY). These images were analyzed frame by frame by a video editing software (Adobe premiere 5.0, Adobe systems, Inc., San Jose, CA). For every run, residence times of 100 particles were obtained from the images. These were saved in a spreadsheet for further analysis.

Experimental Design

A fractional factorial experimental design was used to determine the combination of parameter levels to be used in the experiments. With this design, it was possible to statistically analyze the effects of holding tube configuration (three levels), flow rate (two levels), carrier fluid viscosity (two levels), particle concentration (two levels), and particle type (three levels).

RESULTS AND DISCUSSION

Experimental results of this study were statistically analyzed to determine the effects of process and system parameters on mean residence time (RT_{mean}), minimum residence time (RT_{min}), and standard deviation of residence times (RT_{std}). For this purpose, an analysis of variance (ANOVA) was conducted using the General Linear Model of SAS (Release 8.2, SAS Institute Inc., Cary, NC). Before conducting the analysis, validity of ANOVA assumptions (normality of distribution and homogeneity of variances) was checked based on residual plots by using univariate procedure of SAS. Residual diagnostics did not reveal any non-normality or in homogeneity of variance.

Results of the ANOVA tests showed that RT_{mean} and RT_{min} were significantly affected by holding tube configuration, particle type, flow rate, carrier fluid viscosity, and particle concentration ($p < 0.05$) while RT_{std} was significantly affected by holding tube configuration and flow rate. However, significant effects of particle type, carrier fluid viscosity, and particle concentration on RT_{std} were found when the data was analyzed separately for each holding tube configuration and flow rate. Only the effects which significantly affected RT_{mean}, RT_{min}, and RT_{std} are discussed and the related data tables are presented.

One can get a better understanding of the mechanisms affecting RT_{mean}, RT_{min}, and RT_{std} by considering the average values obtained from all of the experiments (Table

3.2). Since the primary focus of this study is to compare the performance of straight, single helical, and double helical holding tube configurations, it is important to first consider the effect of holding tube configuration on RT_{mean}, RT_{min}, and RT_{std}. Holding tube configuration had significant effects on RT_{mean} ($p < 0.0001$), RT_{min} ($p = 0.0268$), and RT_{std} ($p < 0.0001$). From Table 3.2 it can be seen that the narrowest RTD (the smallest RT_{std}) was obtained in the single helical holding tube, whereas the distribution was the widest for the straight holding tube. RT_{std} in the double helical holding tube was smaller than that in the straight holding tube, but greater than that in the single helical holding tube. Similar trends were observed for RT_{mean} and RT_{min}. These results are in accordance with what Palazoglu and Sandeep (2002) found when comparing a double helical holding tube having a curvature ratio of 0.143 with a single helical holding tube of the same curvature ratio. However, the results of RT_{std} in the straight and single helical holding tubes are contradictory to those of Sandeep *et al*. (1997) who found that RT_{std} in a conventional holding tube was similar to that in a helical holding tube. This discrepancy may be due to three sharp 180° bends in their straight holding tube assembly compared to only one bend in the straight holding tube used in our study.

Flow rate was one of the most significant parameters affecting the overall RT_{mean}, RT_{min}, and RT_{std}. Increasing the flow rate from 1.25 kg/s to 1.50 kg/s resulted in a significant decrease in RT_{mean} ($p < 0.0001$), RT_{min} ($p < 0.0001$), and RT_{std} ($p = 0.0223$) as seen in Table 3.2. In order to better understand the effect of flow rate on RT_{std}, it is important to understand how this effect is influenced by holding tube configuration. The effect of flow rate on RT_{std} was significant ($p = 0.0075$) in the straight holding tube but not in the single and double helical holding tubes (Table 3.3). The reason for not observing an effect in the single and double helical holding tubes could be because a secondary flow is present in these holding tubes, and increasing the flow rate from 1.25 kg/s to 1.50 kg/s did not further increase the strength of the secondary flow and improve RT_{std}. Effect of flow rate was significant for RT_{std} of all particle types ($p < 0.0001$). However, the effect was more pronounced for acrylic particles, indicating that at high flow rates, sedimentation of high density acrylic particles was prevented and they were forced to travel at regions close to the tube center (Table 3.4). This result is in agreement with the findings of Sandeep and Zuritz (1994), Alhamdan and Sastry (1997), and Palazoglu and Sandeep (2002).

Particle type had a significant effect on RT_{mean} ($p < 0.0001$) and RT_{min} ($p < 0.0001$), but not on RT_{std}. Though RT_{std} was not significantly affected (overall) by particle type, it was apparent that RT_{std} was affected by particle type in varying degrees in different configurations of holding tubes. Thus, further analysis was conducted on the effect of particle type on RT_{std} by grouping data based on the holding tube configuration. When the effect of mixing acrylic and polystyrene particles was analyzed, it was seen that the effect of particle type was significant for RT_{mean} in the single helical ($p = 0.0011$), double helical ($p = 0.0012$), and straight ($p < 0.0001$) holding tubes. For experiments involving mixing acrylic and polystyrene particles, it was found that RT_{mean} and RT_{min} decreased in the single helical configuration, but opposite effects were found in the straight and double helical holding tubes (Table 3.5). In the straight and single helical holding tubes, mixing polystyrene particles with acrylic particles resulted in a significant increase ($p = 0.0101$) in RT_{std} of polystyrenes particles (in agreement with the findings of Palazoglu and Sandeep, 2004). However, opposite result was obtained in the double helical holding tube. Decrease in the RT_{std} of acrylic particles was found in the double helical holding tube, whereas increases were found in the single helical and straight configurations. This is contradictory to the results of a study by Palazoglu and Sandeep (2004) who found that RT_{std} of acrylic particles decreased as a result of mixing with polystyrene particles in the single and double helical holding tubes.

Carrier fluid viscosity significantly affected the overall RT_{mean} ($p = 0.0009$) and RT_{min} ($p = 0.002$), but not RT_{std}. Though RT_{std} was not significantly affected (overall) by carrier fluid viscosity, it was apparent that RT_{std} was affected by carrier fluid viscosity in varying degrees in different configurations of holding tubes. Thus, further analysis was conducted on the effect of carrier fluid viscosity on RT_{std} by grouping data based on the holding tube configuration. When the effect of carrier fluid viscosity was analyzed for each holding tube configuration, significant increases are found for RT_{mean} and RT_{min} in the double helical ($p = 0.0044$ and $p = 0.0045$, respectively) and straight holding tube configurations ($p = 0.0002$ and $p = 0.0010$, respectively) upon increasing the viscosity of the CMC solution (Table 3.6). Increasing carrier fluid viscosity resulted in a significant decrease (from 0.44 to 0.39 s) in RT_{std} in the single helical holding tube ($p = 0.0486$). No significant effects of carrier fluid viscosity (on RT_{std}) were found in the double helical and straight holding tubes. Observing no significant effect of the carrier fluid viscosity on the overall RT_{std} is in agreement with the findings of Palazoglu and Sandeep (2002, 2004). However, increases in RT_{mean} and RT_{min} in the straight holding tube are in contradiction with the results of the studies by Sandeep *et al*. (1994, 2000), Alhamdan and Sastry (1997), and Tucker and Heydon (1998) who reported that mean residence times decreased upon increasing consistency coefficient of the carrier fluid. Sandeep *et* *al*. (1994, 2000) attributed the decrease in RT_{mean} to the increase in the drag force on the particles. Alhamdan and Sastry (1997) observed that the particles were lifted up (close to the tube axis), resulting in a significant decrease in mean normalized residence times of particles when 1.0 % CMC solution was used instead of water. The contradiction in the results of these studies and our study may be because of the different viscosity ranges used. The consistency coefficients used in our study were much lower than those in the other studies.

Increasing particle concentration from 4 % to 12 % resulted in a significant increase in the overall RT_{mean} ($p = 0.0004$) and RT_{min} ($p = 0.0033$). However, no significant effect was found for the overall RT_{std}. When the effect of particle concentration was analyzed for each holding tube configuration, it was found that RT_{mean} and RT_{min} significantly increased with an increase in particle concentration in the straight ($p = 0.0013$ and $p = 0.0264$, respectively) and double helical holding tubes ($p = 0.0039$ and $p = 0.0093$, respectively). RT_{std} of particles increased significantly in the straight holding tube ($p = 0.0087$) as seen in Table 3.7. In addition, it was found that the effect of particle concentration on RT_{std} was significant at low flow rates ($p = 0.0224$). At high flow rate (1.50 kg/s), the effect of particle concentration was masked by the effect of high flow rate and no significant change was observed upon increasing particle concentration from 4 % to 12 % (Table 3.8). Our result on the average RT_{mean} (Table 3.2) is in agreement with the findings of Salengke and Sastry (1996), Sandeep *et al*. (2000), and Palazoglu and Sandeep (2002, 2004). Observing an increase in the RT_{std} (with an increase in the particle concentration) in the straight holding tube is in agreement with what Alhamdan and Sastry (1997) and Palazoglu and Sandeep (2002, 2004) found, but in contradiction to the findings of Salengke and Sastry (1996) and Sandeep *et al*. (1997, 2000). Alhamdan and Sastry (1997) attributed the widening of RTD with an increase in the particle concentration to the formation of two particle layers (at the top and bottom of the tube) which moved with different velocities. Salengke and Sastry (1996) did not observe any clear effects of particle concentration (20 % to 40 %) on RT_{std} in the curved section of the holding tube. Sandeep *et al*. (2000) reported a decrease in RT_{std} of particles in helical tubes as the particle concentration increased from 10 % to 20 %. Contradictory results found in the literature indicate that the effect of particle concentration on RTD is dependent on the range of concentrations used as well as the other experimental parameters such as particle type, size, and holding tube configuration.

CONCLUSIONS

The results of this study showed that RTD of particles were significantly affected by holding tube configuration and flow rate. The lowest RT_{std} was obtained in the single helical holding tube. RT_{std} in the double helical holding tube was lower than that obtained in the straight holding tube, but higher than that in the single helical holding tube.

The effect of flow rate was more pronounced in the straight holding tube. Increasing flow rate from 1.25 kg/s to 1.50 kg/s did not further improve RT_{std} in the single and double helical holding tubes. In addition, acrylic particles were more affected by the flow rate than the polystyrene particles.

Mixing polystyrene and acrylic particles increased the RT_{std} of polystyrene and acrylic particles in the single helical and straight holding tubes, whereas opposite results were obtained in the double helical holding tube. Polystyrene particles always had lower RT_{mean} and RT_{std} due to their neutrally buoyant characteristics. The effect of mixing two types of particles was more pronounced for acrylic (higher density) particles.

Increasing carrier fluid consistency coefficient and particle concentration did not have any significant effects on the overall RT_{std} of particles. However, RT_{mean} and RT_{min} in the straight and double helical holding tubes increased significantly upon increasing the viscosity of the carrier fluid or particle concentration. Results of this study are valuable in terms of deciding the type of holding tube configuration and processing conditions to obtain the narrowest RTD for the highest product quality.

ACKNOWLEDGEMENTS

Support for the research study undertaken here, resulting in the publication of Paper No. FSR-04-18 of the Journal Series of the Dept. of Food Science, NCSU, Raleigh, NC 27695-7624, from USDA-NRICGP and North Carolina Agricultural Research Service is greatly acknowledged.

The use of trade names in this publication does not imply endorsement by the North Carolina Agricultural Research Service of the products named nor criticism of similar ones not mentioned.

SYMBOLS

Dn	Dean number Dean number (Re $\sqrt{d/D}$)	
h_{fp}	fluid-to-particle heat transfer coefficient	W/m²-K
K	consistency coefficient	Pa-s^n
n	flow behavior index	
p	p-value of a statistical test	
Re	Reynolds number (ρ_f u d / μ)	

Greek Letters

$\dot{\gamma}$	shear rate	1/s
ρ	density	kg/m^3
σ	shear stress	Pa

Subscripts

min	minimum
std	standard

Abbreviations

CCD	charge-coupled device
CMC	carboxymethlycellulose
RT	residence time
RTD	residence time distribution

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Table 3.1. Range of process and system parameters

Particle type	Polystyrene Acrylic Polystyrene-mixture Acrylic-mixture
Flow rate (kg/s)	1.25 1.50
Viscosity of the carrier fluid	K ($Pa.s^n$): 0.05 – 0.29 n: 0.67 – 0.84
Particle concentration (v/v)	4 % 12 %
Configuration of the holding tube	Single-helical Double-helical Straight

Table 3.2. Effect of process and system parameters on RT_{mean}, RT_{min}, and RT_{std}

Parameter	**Parameter level**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Configuration	Straight	28.78	24.06	1.71
	Single helical	22.71	21.67	0.42
	Double helical	24.27	22.23	0.64
Flow rate	Low (1.25 kg/s)	29.94	26.80	1.10
	High (1.50 kg/s)	20.56	18.50	0.74
Particle type	Polystyrene	21.07	18.95	0.76
	Acrylic	28.88	25.92	1.05
	Polystyrene-mixture	22.30	20.38	0.79
	Acrylic-mixture	28.76	25.36	1.08
Carrier fluid	Low (0.05 $Pa.s^n$)	23.98	21.38	0.88
consistency coefficient	High (0.29 $Pa.s^n$)	26.53	23.93	0.97
Particle concentration	Low (4 %)	23.84	21.46	0.79
	High (12 %)	26.98	23.85	1.05

Table 3.3. Effect of flow rate (for different configurations of holding tubes) on RT_{mean}, RT_{min}, and RT_{std}

Flow rate	**Configuration**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Low	Double helical	28.44	25.91	0.79
High	Double helical	20.11	18.55	0.48
Low	Single helical	26.86	25.82	0.42
High	Single helical	18.56	17.51	0.42
Low	Straight	34.54	28.68	2.09
High	Straight	23.02	19.45	1.33

Table 3.4. Effect of flow rate (for different particle types) on average RT_{mean}, RT_{min}, and RT_{std}

Flow rate	Particle type	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Low	Acrylic	35.89	32.16	1.33
High	Acrylic	21.87	19.67	0.77
Low	Acrylic-mixture	34.67	30.27	1.32
High	Acrylic-mixture	22.82	20.45	0.83
Low	Polystyrene	23.18	20.90	0.86
High	Polystyrene	18.95	17.00	0.67
Low	Polystyrene-mixture	26.00	23.88	0.88
High	Polystyrene-mixture	18.60	16.89	0.71

Table 3.5. Effect of particle type (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Particle type	Configuration	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Acrylic	Double helical	26.50	24.05	0.77
Acrylic-mixture	Double helical	27.45	25.28	0.62
Polystyrene	Double helical	21.22	20.45	0.63
Polystyrene-mixture	Double helical	21.93	19.13	0.53
Acrylic	Single helical	26.93	25.53	0.46
Acrylic-mixture	Single helical	23.37	22.33	0.58
Polystyrene	Single helical	20.13	19.36	0.24
Polystyrene-mixture	Single helical	20.40	19.46	0.40
Acrylic	Straight	33.22	28.17	1.93
Acrylic-mixture	Straight	35.46	28.48	2.04
Polystyrene	Straight	21.85	21.35	1.42
Polystyrene-mixture	Straight	24.58	18.26	1.45

Table 3.6. Effect of carrier fluid viscosity (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Fluid viscosity	Configuration	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Low	Double helical	22.41	20.16	0.66
High	Double helical	26.14	24.30	0.61
Low	Single helical	23.46	22.42	0.44
High	Single helical	21.95	20.91	0.39
Low	Straight	26.07	21.54	1.53
High	Straight	31.49	26.58	1.89

Table 3.7. Effect of particle concentration (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Particle conc.	**Configuration**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Low	Double helical	22.37	20.39	0.60
High	Double helical	26.17	24.07	0.68
Low	Single helical	22.58	21.44	0.44
High	Single helical	22.84	21.89	0.40
Low	Straight	26.59	22.55	1.34
High	Straight	30.97	25.58	2.08

Table 3.8. Effect of particle concentration (for different flow rates) on RT_{mean}, RT_{min}, and RT_{std}

Particle conc.	**Flow Rate**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Low	Low	27.67	25.04	0.83
High	Low	32.21	28.57	1.37
Low	High	20.02	17.89	0.75
High	High	21.10	19.12	0.74

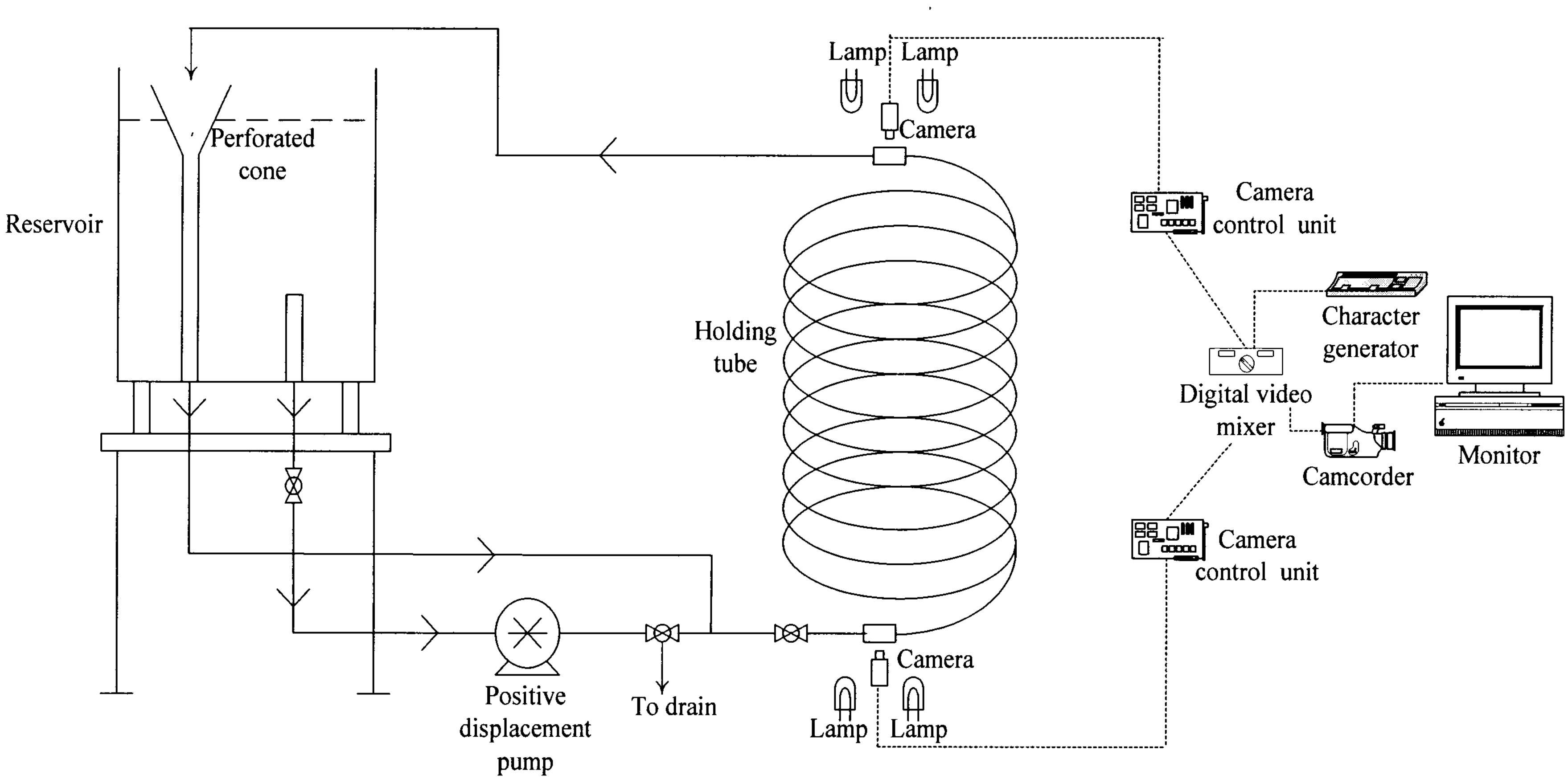

Figure 3.1: Schematic representation of the experimental setup for helical holding tubes

Chapter 4

MANUSCRIPT II

Residence Time Distribution of Multiple Particles in Different Configurations of Holding Tubes

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Paper No. FSR-04-19 of the Journal Series of the Department of Food Science, North Carolina State University, Raleigh, NC 27695-7624.

ABSTRACT

Residence time distribution of particles in a novel holding tube configuration (chaotic holding tube) was studied. The results obtained from the chaotic holding tube were compared with those obtained from the straight, single helical, and double helical holding tubes. Polystyrene and acrylic particles were used as model food particles. The effects of flow rate, carrier fluid viscosity, particle type, and particle concentration on RTD of particles were also studied. The effect of inclination angle of the horizontal section of the chaotic holding tube on RTD of particles was determined by testing two different angles (0 º and 45 º). The performance of the chaotic holding tube was better than the straight and double helical holding tubes in terms of narrowing RTD, but the RTD in the single helical holding tube was narrower than that in the chaotic holding tube. However, at high flow rate (1.50 kg/s) the lowest RT_{std} was obtained in the chaotic holding tube. Changing the inclination angle of the chaotic holding tube from 0 º to 45 º resulted in wider RTD of both polystyrene and acrylic particles.

Keywords: residence time distribution, chaotic holding tube, particulate flow

INTRODUCTION

Liquid foods such as milk and fruit juices are commonly subjected to continuous aseptic processing. However, extension of aseptic processing to low-acid particulate foods suffers from the lack of data on two critical factors -- convective heat transfer coefficient at the liquid-particle interface (h_{fp}) and residence time distribution (RTD) of particles in the holding tube of the aseptic processing system (Ramaswamy *et al*., 1995).

Minimum residence time of the particle in the holding tube is related to product safety, whereas residence time distribution relates to product quality. Determination of residence time of the critical particle (fastest moving particle) is required to ensure that a target lethality value has been accumulated inside the particle. Ideal type of flow (to obtain a narrow RTD of particles) is the plug flow where all the particles experience the same residence time in the tube (Sandeep *et al*., 1997). However, under laminar conditions in a conventional holding tube particles which travel close to the walls of the tube will experience longer residence times and be overprocessed (due to a parabolic velocity flow profile). To achieve uniformity of heat treatment in a process, and hence minimize overprocessing of a product, mixing of the product and flattening of the velocity profile is required. This can be obtained by orifice plates which induce turbulence or by static mixers which redistribute the fluid elements along the radial direction (Sandeep *et al*., 1997). However, using these may damage the product due to the high shear stresses involved (Castelain *et al*., 1997). In addition, cleaning of orifice plates and static mixers may be difficult.

Changing tube geometry is another way to obtain mixing of particles. It is well known that flow in a helical tube is different from that in a straight tube due to the centrifugal force exerted by the curvature of the tube. During flow through a curved pipe, a pressure gradient is required to balance the centrifugal force resulting from the curvature of the tube (Barua, 1963). As a consequence, a flow (secondary flow) is set up in radial direction. In the secondary flow the fluid near the top and the bottom moves inwards whereas the fluid at the center moves outwards (Barua, 1963). Axial dispersion in helical tubes is smaller than that in conventional tube which results in a profile close to plug flow. Koutsky and Adler (1964) discussed the three mechanisms which affect axial dispersion in helical tubes. Secondary flow results in shifting of the maximum velocity region towards the outer wall of the tube (Barua, 1963). This non-uniform velocity profile which promotes axial dispersion in laminar flow is the first mechanism. The second mechanism, random mixing due to diffusion or turbulent eddies, inhibits axial dispersion. The third mechanism, secondary flow, inhibits axial dispersion by continuously changing the radial positions of particles (Koutsky and Adler, 1964).

In coiled tubes, developing flow patterns can be generated in a repeated manner by changing the direction of centrifugal force. As a result, chaotic mixing is achieved (Acharya *et al*., 2001). Chaotic mixing has been shown to enhance heat transfer in heat exchangers (Chagny *et al*., 2000; Acharya *et al*., 2001; Lefevre *et al*., 2003). A spatially chaotic system was found to decrease axial dispersion of fluid particles in laminar flow due to the generation of chaotic trajectories. Chaotic trajectories can be introduced to the helically coiled pipe by shifting the plane of curvature by a ± 90 ° angle from one bend to the next (Castelain *et al*., 1997). Castelain *et al*. (1997) found that the transition from laminar to turbulent regime occurred at Reynolds numbers of the same order of magnitude for both helical and chaotic systems. However, for Re > 2,500, axial dispersion was found to be 25 % smaller in the chaotic system than that in the helical system. This indicated that in this range of Reynolds numbers, the residence time distribution became narrower by the generation of chaotic trajectories.

Castelain *et al.* (2000) conducted RTD experiments in a spatially chaotic flow for Reynolds numbers between 30 and 13,000. One result of this study was an increase in Peclet number with Reynolds number in the chaotic system, which indicated smaller axial dispersion in the chaotic configuration compared to the helical configuration. This trend was observed regardless of the number of bends in the chaotic system. Reduction in the axial dispersion was attributed to the presence of regular and chaotic zones in the flow regime. At the inlet of the spatially chaotic system, flow was similar to that in a helically coiled tube (regular zone). After a few bends, fully developed chaotic zone began to dominate due to inducement of chaotic trajectories of fluid particles.

The effects of pitch in helically coiled tubes and the inclination between the planes of curvature between the bends on fluid flow and heat transfer have been studied by several researchers. Mishra and Gupta (1979) found that as the pitch of a helical tube increased (for the same diameter of coil) centrifugal force decreased and secondary flow lost its strength. Austen and Soliman (1988) pointed out that for a helical tube of infinite pitch, the flow approached that in a straight tube. Cheng and Takuma (1991) studied the development of secondary flow in curved pipes with bend angles of 45, 90, and 180 º. They observed less effect of secondary flow and stronger effect of inertial force for the entrance region of the 45 º bend. At the 180 º bend, flow was close to fully developed and the effect of centrifugal force dominated the flow field.

The effect of secondary flow on RTD of particles has been investigated by Sandeep and Zuritz, 1994; Salengke and Sastry, 1996; Sandeep *et al*., 1997; Grabowski and Ramaswamy, 1998; Sandeep and Zuritz, 1999; Sandeep *et al*., 2000; Palazoglu and Sandeep, 2002, 2004. However, none of these studies deal with the chaotic tube configuration. Thus, the primary objective of this study was to investigate the effect of a chaotic holding tube configuration on RTD of particles and to compare the results obtained from the chaotic holding tube to the results from the other configurations studied -- single helical, double helical, and straight holding tubes. The second objective of the study was to determine the effects of various processing parameters -- flow rate, carrier fluid viscosity, particle concentration, and mixing particles with different densities on the RTD of particles. The third objective was to determine the effect of inclination angle of the horizontal section of the chaotic holding tube on RTD.

MATERIALS AND METHODS

Experimental setup

The experimental setup consisted of a holding tube, a positive displacement pump, and a reservoir (Figure 4.1). Holding tubes (straight, single helical, double helical, and chaotic) used in this study were 15.24m long, 0.0508m i.d. transparent tubes (Cleargard, CSC200, M.G. Newell Corp., Greensboro, NC). The straight holding tube assembly consisted of two straight sections connected by a 180° bend. The tube was inclined upwards of a quarter inches per foot length of the tube according to FDA regulations. The single helical tube was built by coiling the transparent tube around a cylindrical mandrel (diameter = 0.355 m). The resulting curvature ratio (the ratio of the tube diameter to coil diameter) was 0.143. The double helical holding tube was coiled in shape of figure-8 around two cylindrical mandrels of the same diameter. The chaotic holding tube was built by connecting vertical and horizontal coiled tubes (of equal length) by an angle of 90 º. The pitch of the helical tubes, defined as the vertical rise per unit length, was 0.076 m.

A positive displacement pump (Model 45U2, Waukesha Cherry-Burrell, Delavan, WI) was used to circulate the carrier fluid between the reservoir and the holding tube. Particles were introduced to the system through a perforated cone assembly inside the reservoir. Particles did not pass through the pump, instead they were bypassed from the pump by the perforated cone assembly to a point at the exit of the pump.

Carrier Fluid

The carrier fluid used in this study was carboxymethylcellulose (CMC) solution (CMC is a typical viscosifier used in food industry). CMC powder (6000 C, TIC Gums, Belcamp, MD) was dissolved in tap water, and after hand-mixing the CMC – water mixture for some time, it was pumped through the system to make a more homogeneous solution. The solution was stored for 6 -10 hours to obtain a clear solution free of air bubbles. The final CMC solution had density of 1,001 kg/m^3. Particles were introduced to the system after the flow of solution reached steady-state conditions.

Rheological characterization of CMC solutions was done using a Haake Viscotester (Haake VT 550, sensor MV-DIN, Haake Inc., Paramus, NJ) before and after each run. A bob and cup geometry was used with this instrument. Rheological properties such as shear stress, shear rate, and apparent viscosity were obtained at room temperature. The data was fitted to the Oswald De Waale model ($\sigma = K\ \dot{\gamma}^n$). Flow behavior index (n) and consistency coefficient (K) values were used to control viscosity range of the solution at which the experiments were conducted. These values are shown in Table 4.1 with other range of experimental parameters. Apparent Viscosity of CMC solution ranged between 0.025 – 0.108 Pa.s at shear rates of 0 -100 1/s. Due to a slight decrease in viscosity with time, addition of CMC powder was required between each run.

All experiments in this study were conducted at room temperature and under laminar flow conditions. Particle Reynolds numbers ranged from 117 to 835. Dean numbers in the single helical, double helical, and chaotic holding tubes ranged from 44 to 290.

Model Food Particles

Model food particles used in this study were polystyrene and acrylic spherical particles (Precision Plastic Ball Co., Franklin Park, IL). All particles had a diameter of 0.014 m. Some of these particles were used as tracer particles by painting them with different colors and patterns for easy identification. Density measurements were performed on these tracer particles by using a gas pcynometer (Model Accupcy 1330, Micrometrics Instrument Corporation, Norcross, GA). Polystyrene particles were closer to neutrally buoyant (ρ = 1,012 ± 9 kg/m^3) than acrylic particles (ρ = 1,169 ± 18 kg/m^3). Either single or mixed types of particles (both acrylic and polystyrene particles) were used at concentrations of 4 % and 12 % (v/v) in the experiments.

Digital Image Analysis

Residence time data of tracer particles were obtained by the digital image analysis method developed by Simunovic (1998). In this method, the experiments are continuously recorded using a camcorder (DCR-TRV103, Sony Corporation, New York, NY) and two industrial color CCD cameras (Toshiba, IK-M43, New York, NY) located at the entrance and the exit of the holding tube. A character generator (Panasonic, WJ-KB30, Matsushita Communication Industrial Co., Ltd., Japan) with a built-in stop watch function was used to display time (resolution: 0.01 s) on the recordings. A digital video mixer (Videonics, MX-1, Campbell, CA) was used to combine the images from the entrance and the exit of the holding tube. The recordings from the camcorder were downloaded onto a computer using a video capturing software (DV Gate Motion, Sony Corporation, New York, NY). These images were analyzed frame by frame by a video editing software (Adobe premiere 5.0, Adobe systems, Inc., San Jose, CA). For every run, residence times of 100 particles were obtained from the images. These were saved in a spreadsheet for further analysis.

Experimental Design

A half factorial $2^{3-1} \times 3 \times 4$ experimental design was used to determine the combination of parameter levels to be used in the experiments. This design covered 48 experiments. The experiments were statistically designed to determine the effects of particle type (three levels), flow rate (two levels), viscosity of the carrier fluid (two levels), particle concentration (two levels), and holding tube configuration (four levels).

Additional experiments

Additional experiments were performed on the chaotic holding tube configuration to determine the effect of inclination angle of the horizontal section on RTD of particles. For this purpose, two sets of experiments were conducted on the holding tube with 45 ° and 0 ° of inclination angle of the horizontal section. 0 ° of inclination angle refers to the original chaotic holding tube assembly where a 90 ° angle was present between the horizontal and vertical sections. 45 ° inclination angle between the sections was obtained by tilting the horizontal section of the chaotic holding tube from its original position of 0 ° inclination angle. Single type (polystyrene or acrylic) particles were used. Experiments were conducted for single levels of flow rate, carrier fluid viscosity, and particle concentration (Table 4.2). For each experiment, two separate runs were performed -- one to record the flow through the whole holding tube (with cameras located at the entrance of the vertical section and at the exit of the horizontal section), and the second one to record the flow only through the vertical section of the chaotic holding tube (with cameras located at the entrance and the exit of the vertical section). After analyzing the recordings and obtaining residence times of particles from the whole length and from the vertical section of the holding tube, residence times of particles in the horizontal section of the holding tube were backcalculated. These residence times from the horizontal section of the holding tube were statistically analyzed.

RESULTS AND DISCUSSION

The effects of system and process parameters on the mean, minimum, and standard deviation of residence times were statistically analyzed by an analysis of variance (ANOVA) using the General Linear Model of SAS (Release 8.2, SAS Institute Inc., Cary, NC). Before conducting the analysis, validity of ANOVA assumptions (normality of distribution and homogeneity of variances) was checked based on residual plots by using univariate procedure of SAS. Residual diagnostics did not reveal any non-normality or in homogeneity of variance. Data obtained from the experiments conducted with different inclination angles of the horizontal part of the chaotic holding tube configuration was analyzed by using a homogeneity of variance test (Levene's test).

Statistical analysis showed that all system and process parameters had a significant effect ($p < 0.05$) on RT_{mean} and RT_{min}. Standard deviation of residence times were significantly affected by the holding tube configuration ($p < 0.0001$) and flow rate ($p = 0.0079$). When the statistical analysis was conducted for each configuration and flow rate additional parameters (particle type, particle concentration, and carrier fluid consistency coefficient) were found to have an effect on RT_{std} and these results are discussed in detail.

The lowest value for RT_{std}, and thus the narrowest RTD, was obtained in the single helical holding tube (Table 4.3). RT_{std} in the chaotic holding tube (0.45 s) was higher than that in the single helical tube (0.42 s) but lower than those in the double helical (0.64 s) and straight (1.71 s) holding tubes. This may indicate that even though the direction of flow changed upon tilting the tube by 90 º (to shift from single helical to chaotic configuration), recovery of this change in the horizontal section of the chaotic tube was fast, which dampened the beneficial effect of changing the direction of flow. Observing a higher RT_{std} in the chaotic holding tube (compared to that in the single helical holding tube) is in contradiction to what Castelain *et al.* (2000) found. They reported that a smaller axial dispersion (larger Peclet number) was obtained in the chaotic system, and that the RTD of the fluid elements was narrower in the chaotic holding tube as compared to that in the helical tube. The disagreement between this study and our study may be due to the differences in the chaotic holding tube assemblies. In their study, the chaotic geometry was generated by continuously shifting each bend by a ±90º angle with respect to the previous one. So, a reduced axial dispersion in the chaotic system was attributed to the continuous shift in the plane of curvature from one bend to the next.

Flow rate had significant effects on RT_{mean} ($p < 0.0001$), RT_{min} ($p < 0.0001$), and RT_{std} ($p = 0.0079$). When the flow rate increased from 1.25 to 1.50 kg/s, RT_{std} in the chaotic holding tube decreased from 0.52 to 0.38 s (Table 4.4). The decrease in RT_{std} in the chaotic holding tube was greater than that observed in the single helical tube, but less than those in the double helical and straight tubes. This may be due to the already strong secondary flows in the single helical and chaotic holding tubes. At high flow rate, the narrowest RTD was obtained in the chaotic holding tube, whereas at low flow rate, the

RTD in the single helical holding tube was narrower than that in the chaotic holding tube (in agreement with the findings of Castelain *et al*., 2000). When a statistical analysis was conducted separately for each particle type, RT_{std} of acrylic particles was found to significantly decrease ($p = 0.0116$) with an increase in flow rate (Table 4.5). Upon increasing flow rate, acrylic particles experienced more mixing and started to travel off the bottom of the tube (in agreement with observations of Palazoglu and Sandeep, 2004). Polystyrene particles were already well distributed along the holding tube due to their neutrally buoyant densities, and hence their RT_{std} did not significantly decrease.

Particle type significantly affected the overall RT_{mean} ($p = 0.0003$), but not overall RT_{min} and RT_{std}. However, significant effects of particle type on RT_{std} were found in the single helical holding tube ($p = 0.0297$) and at low flow rate ($p = 0.086$) as seen in Tables 4.6 and 4.7, respectively. When the effect of particle type, and thus the effect of mixing two types of particles, was investigated for each configuration, it was found that upon mixing polystyrene and acrylic particles, RT_{mean} of acrylic particles decreased, whereas no change was found for polystyrene particles in the chaotic holding tube (Table 4.6). Decreases in RT_{mean} of acrylic particles were also found in the single helical holding tube, but not in the straight or double helical holding tubes. Mixing with acrylic particles did not affect the flow of polystyrenes in the single helical, double helical, and chaotic holding tubes, whereas acrylic particles may have started to travel at the regions closer to the tube center upon mixing with polystyrene particles in the single helical and chaotic holding tubes. Since the decrease in RT_{mean} of acrylic particles was observed in the chaotic and single helical holding tubes, it can be said that strong secondary flows in these holding tube configurations enhanced the effect of mixing two types of particles.

Carrier fluid consistency coefficient had a significant effect on overall RT_{mean} ($p = 0.0018$) and RT_{min} ($p = 0.0328$), but not on the overall RT_{std}. However, RT_{std} in the chaotic holding tube significantly decreased (from 0.60 s to 0.30 s) when the carrier fluid viscosity increased ($p = 0.0328$). Increase in viscosity resulted in a decrease in RT_{mean} and an increase in RT_{min} of particles in the chaotic holding tube (Table 4.8). Upon increasing the carrier fluid viscosity, the drag force on the particle, which resists sedimentation, increased and resulted in a decrease in the residence times of particles. This was observed in the chaotic and single helical holding tubes. However, in the straight and double helical holding tubes, opposite effects were found. This may be due to the presence of strong secondary flow in the single helical and chaotic holding tubes. Our results on the effect of carrier fluid viscosity on RT_{std} are in agreement with the result of the study conducted by Abdelrahim *et al.* (1993) who found that the RTD of carrot cubes narrowed with an increase in the carrier fluid (starch) viscosity. However, there are contradictions to the findings of Sandeep and Zuritz (1999) and Abdelrahim *et al.* (1997) who found that an increase in viscosity resulted in an increase in RT_{std} in the straight holding tubes with bends. This may be due to the higher carrier fluid consistency coefficients used in their studies which resulted in lower Reynolds numbers, and thus wider RTD of particles.

Particle concentration was found to have a significant effect on RT_{mean} ($p = 0.0002$) and RT_{min} ($p = 0.0017$), but not on overall RT_{std}. RT_{std} was significantly affected by particle concentration at low flow rate ($p = 0.0154$) and in straight holding tube ($p = 0.0019$). RT_{mean} and RT_{min} in the chaotic holding tube increased with an increase in the particle concentration (Table 4.9). Greater increases were observed in the straight and double helical holding tubes, whereas no effect was seen in the single helical tube. These results indicate that the effect of particle concentration is dependent on the holding tube geometry, and thus the strength of the secondary flow. In the chaotic and single helical tubes, an increase in particle concentration resulted in a narrower RTD, which indicated that the flow of particles in these holding tube configurations may have approached plug flow at high particle concentration. In the double helical holding tube, where the beneficial effect of secondary flow was lost in straight region (between two loops) of the tube, and in straight tube (where no secondary flow was present) a higher particle concentration resulted in higher residence times. When the effect of increasing particle concentration was analyzed for the two levels of flow rate, it was seen (Table 4.10) that the effect of particle concentration on RT_{std} was significant for the low flow rate ($p = 0.0154$). This may be due to the fact that at high flow rate, the effect of flow rate overcame the effect of particle concentration and resulted in a lower RT_{std}. Our results (in double helical holding tube) are in agreement with those of Palazoglu and Sandeep (2002) who found that RT_{mean}, RT_{min}, and RT_{std} increased as particle concentration increased from 4 % to 12 % in single and double helical holding tubes. In addition, the results obtained from the single helical and chaotic holding tubes are in agreement with the results of a study by Sandeep and Zuritz (1999) who found that an increase in particle concentration (from 3 % to 10 %) decreased RT_{std} in a straight holding tube with four bends.

Data obtained from 0 ° and 45 ° inclined horizontal sections of the chaotic holding tube are presented in Table 4.11. Levene's test was conducted to determine the significance of changing the inclination angle of the horizontal section of the chaotic holding tube on the residence times and RTD. The ANOVA procedure in Levene's test showed that changing the angle of the horizontal section of the chaotic holding tube from 0 ° to 45 ° had a significant effect on RT_{mean} ($p < 0.0001$) and RT_{std} ($p = 0.0004$) for both polystyrene and acrylic particles. RT_{mean} of both types of particles decreased upon changing the inclination angle to 45 °. However, RT_{std} of particles in the 45 ° inclined tube was greater than that in the 0 ° inclined holding tube. This may be attributed to the fact that in the 45 ° inclined tube, buoyancy was a factor in determining the radial position of particles due to the inclination of the horizontal section, whereas for the vertical flow in the 0 ° inclined tube, buoyancy was not a direct radial force (it only had an accelerating or decelerating effect on the velocity of particles). Since the density difference between the carrier fluid and acrylic particles was higher than that between carrier fluid and polystyrene particles, the resultant force of buoyancy and centrifugal forces on acrylic particles was higher than that on polystyrene particles. Thus, acrylic particles experienced buoyancy as a dominant radial migration force during the flow in vertical direction. The results on wider RTD for both types of particles in the 45 ° inclined tube (Dean number range: 127 – 143) are in agreement with the findings of Cheng and Yuen (1987) and Cheng and Takuma (1991) who observed that the buoyancy forces dominated over centrifugal forces for Dean numbers smaller than 182 and that there was less effect of centrifugal force at the entrance of 45 ° bend compared to those in 90 ° and 180 ° bends. Although these findings support our observations of wider RTD for both types of particles in the 45 ° tilted holding tube, more research is required to investigate the effect of inclination angle on RTD of particles.

CONCLUSIONS

In this study, the RTD of particles in a novel holding tube configuration (chaotic holding tube) is discussed. The results obtained from the chaotic holding tube were compared to those obtained from the straight, single helical, and double helical holding tubes. The best holding tube configuration in terms of narrowing RTD was the single helical holding tube. RT_{std} in the chaotic holding tube was greater than that in the single helical holding tube. However, it was seen that at high flow rate, the use of chaotic holding tube resulted in a narrower RTD compared to those in the other holding tubes.

Significant effects of particle type, carrier fluid consistency coefficient, and particle type on RT_{std} were found for individual holding tube configurations and flow rates. Upon mixing polystyrene and acrylic particles in the chaotic holding tube, RT_{mean} of acrylic particles decreased whereas no change was seen for polystyrene particles. Increasing carrier fluid viscosity resulted in a decrease in RT_{mean} and RT_{std} in the chaotic holding tube. RT_{std} in the single helical and chaotic holding tube decreased with an increase in particle concentration.

This study gives valuable insight into the use of different types of holding tube configurations in aseptic processing of multiphase food products. The results of this study will be helpful to obtain the maximum product quality by determining the optimum range of process and system parameters. However, it should be kept in mind that the experimental results are specific to this study and may not be applicable for other situations.

ACKNOWLEDGEMENTS

Support for the research study undertaken here, resulting in the publication of Paper No. FSR-04-19 of the Journal Series of the Dept. of Food Science, NCSU, Raleigh, NC 27695-7624, from USDA-NRICGP and North Carolina Agricultural Research Service is greatly acknowledged.

The use of trade names in this publication does not imply endorsement by the North Carolina Agricultural Research Service of the products named nor criticism of similar ones not mentioned.

SYMBOLS

Dn	Dean number Dean number (Re $\sqrt{d/D}$)	
h_{fp}	fluid-to-particle heat transfer coefficient	W/m^2-K
K	consistency coefficient	Pa-s^n
n	flow behavior index	
p	p-value of a statistical test	
Re	Reynolds number (ρ_f u d / μ)	

Greek Letters

$\dot{\gamma}$	shear rate	1/s
ρ	density	kg/m^3
σ	shear stress	Pa

Subscripts

min	minimum
std	standard

Abbreviations

CCD	charge-coupled device
CMC	carboxymethlycellulose
RT	residence time
RTD	residence time distribution

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Table 4.1. Range of process and system parameters

Particle type	Polystyrene Acrylic Mixture
Flow rate (kg/s)	1.25 1.50
Viscosity of the carrier fluid	K (Pa.s^n): 0.05 – 0.25 n: 0.67 – 0.84
Particle concentration (v/v)	4 % 12 %
Configuration of the holding tube	Single-helical Double-helical Chaotic Straight

Table 4.2. Experimental conditions for additional experiments

Flow rate (kg/s)	1.38
Viscosity of carrier fluid	K (Pa.s^n): 0.11 – 0.12 n: 0.73 – 0.78
Particle concentration	8 %
Reynolds number	270 – 421
Dean number	106 – 160

Table 4.3. Effect of process and system parameters on RT_{mean}, RT_{min}, and RT_{std}

Parameter	**Parameter Level**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Configuration	Straight	28.78	24.06	1.71
	Single Helical	22.71	21.67	0.42
	Double Helical	24.27	22.23	0.64
	Chaotic	22.93	21.93	0.45
Flow rate	Low (1.25 kg/s)	28.84	26.19	0.95
	High (1.50 kg/s)	20.50	18.76	0.66
Particle type	Polystyrene	21.10	19.27	0.70
	Acrylic	27.87	25.27	0.94
	Polystyrene-mixture	22.18	20.59	0.68
	Acrylic-mixture	27.53	24.77	0.90
Carrier fluid consistency	Low (0.05 Pa.s^n)	23.77	21.47	0.81
coefficient	High (0.29 Pa.s^n)	25.58	23.48	0.80
Particle concentration	Low (4 %)	23.55	21.50	0.72
	High (12 %)	25.80	23.45	0.89

Table 4.4. Effect of flow rate (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Flow rate	**Configuration**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Low	Straight	34.46	28.60	2.09
High	Straight	23.09	19.53	1.33
Low	Single helical	26.78	25.74	0.42
High	Single helical	18.63	17.60	0.42
Low	Double helical	28.51	26.00	0.79
High	Double helical	20.03	18.47	0.47
Low	Chaotic	25.63	24.41	0.52
High	Chaotic	20.23	19.45	0.38

Table 4.5. Effect of flow rate (for different particle types) on RT_{mean}, RT_{min}, and RT_{std}

Flow rate	**Particle type**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Low	Acrylic	34.59	31.31	1.22
High	Acrylic	21.16	19.22	0.66
Low	Acrylic-mixture	22.52	28.94	1.08
High	Acrylic-mixture	22.54	20.60	0.73
Low	Polystyrene	22.93	20.94	0.80
High	Polystyrene	19.23	17.60	0.60
Low	Polystyrene-mixture	25.30	23.55	0.72
High	Polystyrene-mixture	19.07	17.62	0.64

Table 4.6. Effect of particle type (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Particle type	**Configuration**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Acrylic	Straight	33.23	28.18	1.93
Acrylic-mixture	Straight	35.46	28.48	2.04
Polystyrene	Straight	21.85	18.26	1.42
Polystyrene-mixture	Straight	24.58	21.35	1.45
Acrylic	Single helical	26.93	25.53	0.46
Acrylic-mixture	Single helical	23.37	22.33	0.58
Polystyrene	Single helical	20.13	19.46	0.24
Polystyrene-mixture	Single helical	20.40	19.36	0.40
Acrylic	Double helical	26.50	24.05	0.77
Acrylic-mixture	Double helical	27.45	25.28	0.62
Polystyrene	Double helical	21.22	19.13	0.63
Polystyrene-mixture	Double helical	21.93	20.45	0.53
Acrylic	Chaotic	24.85	23.31	0.59
Acrylic-mixture	Chaotic	23.83	23.00	0.38
Polystyrene	Chaotic	21.22	20.22	0.51
Polystyrene-mixture	Chaotic	21.83	21.20	0.33

Table 4.7. Effect of particle type (for different flow rates) on RT_{mean}, RT_{min}, and RT_{std}

Particle type	**Flow rate**	**RT_{mean} (s)**	**RT_{min} (s)**	**RT_{std} (s)**
Acrylic	Low	34.59	31.31	1.22
Acrylic	High	21.16	19.22	0.66
Acrylic-mixture	Low	32.52	28.64	1.08
Acrylic-mixture	High	22.54	20.60	0.73
Polystyrene	Low	22.98	20.94	0.80
Polystyrene	High	19.23	17.60	0.60
Polystyrene-mixture	Low	25.30	23.55	0.72
Polystyrene-mixture	High	19.07	17.62	0.64

Table 4.8. Effect of carrier fluid viscosity (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Viscosity	Configuration	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Low	Straight	26.16	21.62	1.53
High	Straight	31.40	26.46	1.89
Low	Single helical	23.55	22.54	0.44
High	Single helical	21.86	20.79	0.39
Low	Double helical	22.32	20.04	0.66
High	Double helical	26.23	24.42	0.61
Low	Chaotic	23.04	21.62	0.60
High	Chaotic	22.83	22.24	0.30

Table 4.9. Effect of particle concentration (for different holding tube configurations) on RT_{mean}, RT_{min}, and RT_{std}

Particle conc.	Configuration	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Low	Straight	26.74	22.66	1.34
High	Straight	30.82	25.47	2.08
Low	Single helical	22.73	21.56	0.44
High	Single helical	22.68	21.78	0.40
Low	Double helical	22.22	20.28	0.60
High	Double helical	26.32	24.18	0.68
Low	Chaotic	22.50	21.49	0.51
High	Chaotic	23.37	22.37	0.39

Table 4.10. Effect of particle concentration (for different flow rates) on RT_{mean}, RT_{min}, and RT_{std}

Particle conc.	Flow rate	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
Low	High	20.06	18.18	0.69
High	High	20.94	19.40	0.62
Low	Low	27.04	24.87	0.75
High	Low	30.66	27.50	1.16

Table 4.11. Effect of inclination angle (of the chaotic holding tube) on RT_{mean}, RT_{min}, and RT_{std}

Inclination	Particle type	RT_{mean} (s)	RT_{min} (s)	RT_{std} (s)
0 º	Polystyrene	11.99	11.35	0.17
45 º	Polystyrene	8.97	6.91	0.41
0 º	Acrylic	11.62	11.27	0.20
45 º	Acrylic	8.97	8.78	0.36

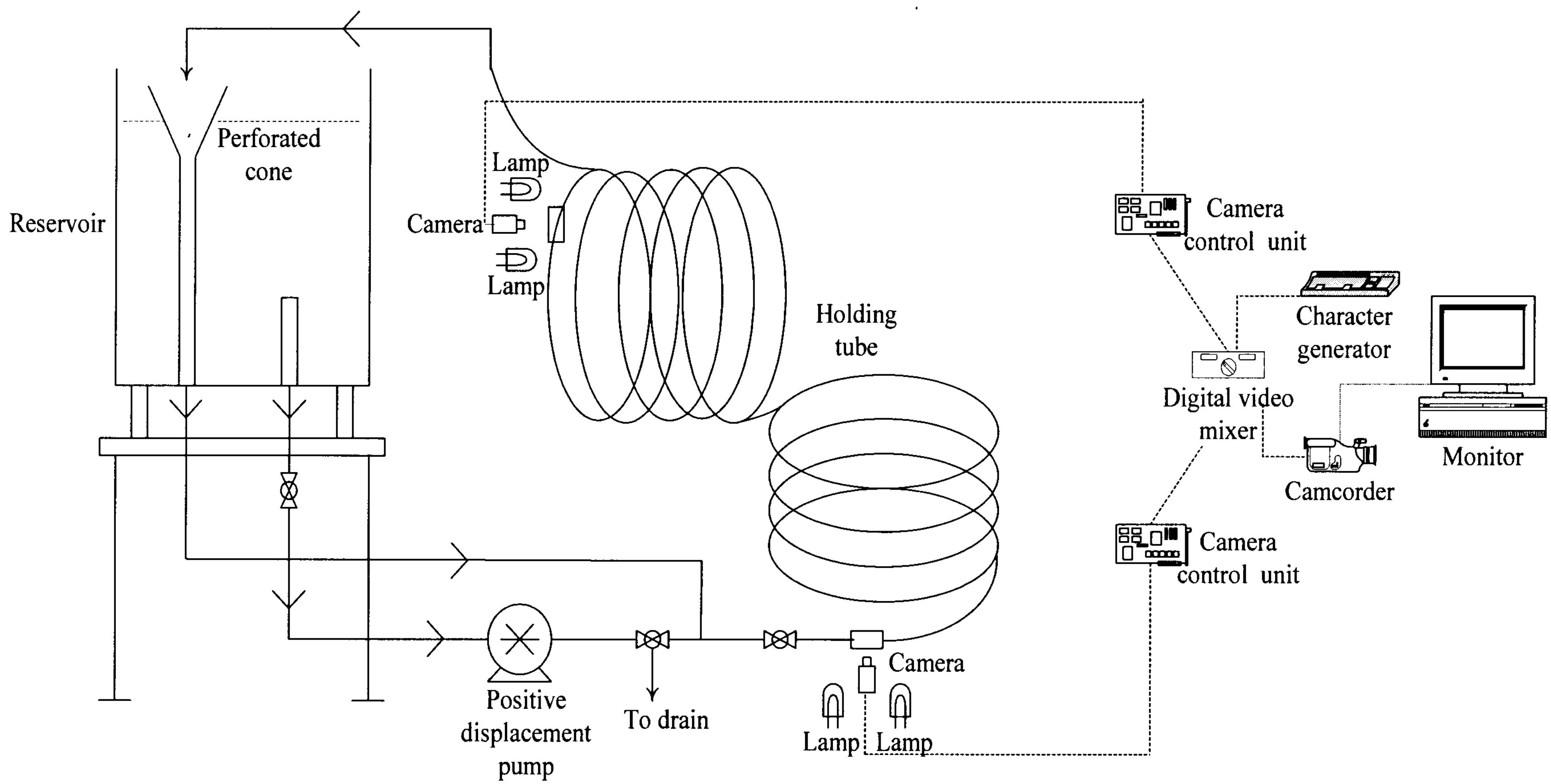

Figure 4.1: Schematic representation of the experimental setup for the chaotic holding tube

Chapter 5

CONCLUSIONS

The results of this study showed that the lowest RT_{std} was obtained in the single helical holding tube. RT_{std} in the novel (chaotic) holding tube was higher than that in the single helical holding tube, but lower than those in the double helical and straight holding tubes. The widest RTD was in the straight holding tube. Based on these results, the single helical holding tube was found to be the best in terms of obtaining the narrowest RTD, and thus providing the best product quality.

Increasing flow rate significantly decreased RT_{mean}, RT_{min}, and RT_{std} of particles. The decrease in RT_{std} of particles was more pronounced in the straight and double helical holding tubes. However, at high flow rate, the lowest value of RT_{std} was obtained in the chaotic holding tube. Hence, use of chaotic holding tube at the high flow rate was shown to be more beneficial in terms of narrowing RTD.

Polystyrene particles always had lower RT_{mean}, RT_{min}, and RT_{std} than acrylic particles. Upon mixing polystyrene and acrylic particles, RT_{mean} of polystyrene particles increased in the straight holding tube, whereas no changes were observed in the other configurations. RT_{mean} of acrylic particles increased in the straight and double helical holding tubes, whereas decreases were seen in the single helical and chaotic holding tubes. RT_{std} of polystyrene and acrylic particles increased as a result of mixing the two types of particles in the straight and single helical holding tubes. However, RT_{std} of both types of particles decreased in the double helical and chaotic holding tubes.

Increasing the carrier fluid consistency coefficient significantly decreased the RT_{std} in the chaotic and single helical holding tubes. However, overall RT_{mean} and RT_{min} increased upon increasing the carrier fluid consistency coefficient. The analysis conducted by grouping data based on the holding tube configuration showed that RT_{mean} of particles increased in the straight and double helical holding tubes, whereas an opposite effect was seen in the single helical and chaotic holding tubes.

As a result of increasing particle concentration, RT_{std} in the straight holding tube significantly increased. An analysis based on flow rates showed that the effect of particle concentration on RT_{std} was significant at low flow rate, but not at high flow rate. This was because at high flow rate the effect of particle concentration was masked.

Changing the inclination angle (from 0 ° to 45 °) of the horizontal section of the chaotic holding tube resulted in a significant increase in the RT_{std} of polystyrene and acrylic particles. Hence, the use of non-inclined chaotic holding tube was shown to provide better mixing and narrower RTD of particles for the levels of parameters studied.

This study will help processors to determine the system and processing conditions under which RTD of particles is narrower and highest product quality is obtained. However, the results are specific to the experimental conditions used in this study, and thus cannot be applicable to every situation.

Chapter 6

RECOMMENDATIONS FOR FUTURE WORK

This study gives an insight into the effects of process and system parameters on RTD of particles during two-phase non-Newtonian flow in holding tubes. In order for processors to apply the results of the experimental works to a commercial process, more research is required. Some recommendations for future work to better understand the effects of process and system parameters on RTD of particles in a holding tube of an aseptic processing system are presented below:

Experimental work should be conducted for a wider range of parameters and by obtaining data for more levels of parameters. For instance, within a range of particle concentration of 4 – 20 %, experiments should be conducted with particle concentrations of 4, 8, 12, 16, and 20 %. This type of experimental design will provide more information on how particle concentration affects RTD and at which value the effect becomes more pronounced or when an opposite effect is obtained.

Different forms of chaotic holding tube configurations need to be investigated to determine the effect of flow inversion on RTD. In addition, different levels of inclination angles (30 ° and 60 °) and pitches should be studied in existing or other possible helical holding tube configurations to determine the effect of inclination on RTD.

Flow visualization experiments (by dye-injections) are suggested in order to observe secondary flow profiles, and thus investigate the degrees of mixing in different holding tube configurations.