THE USE OF SMALL-SCALE EQUIPMENT FOR EVALUATING WATER TREATMENT PLANTS

Final Report to the

Water Research Commission

by

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EXECUTIVE SUMMARY

1 BACKGROUND

Considering the increasing demand for potable water in South Africa, water treatment plants are currently under pressure to increase the production of water. As a consequence, wastewater treatment plants will be faced with an increased flow. In order to respond to this increased demand, water and wastewater treatment plants have to expand. Increase in water production can occur with the construction of new plants, with the expansion of existing water treatment plants, and by the optimisation of current processes. The latter option of optimising existing equipment provides the first step in increasing water production since a more efficient process will lead to higher throughput and better treatment without new construction. Instead of analysing the water treatment process unit by unit, a broad, systems-approach can provide information necessary for increasing the through put.

Water treatment involves the complex interaction of processes such as mixing, coagulation, flocculation, sedimentation, clarification and disinfection. Tools have been developed to simulate the full-scale water treatment plants in order to optimise and monitor the processes in terms of the chemical dosages and physical parameters. These tools include residence time distribution (RTD) studies, the construction of pilot plants, small-scale units, and laboratory tests. Bench-top studies seldom provide sufficient dara for the design of full-scale equipment. Pilot Plants are a useful tool for solving scale-up problems related to dimensional sumilarity and large-scale processing. As the construction and operation of pilot plants can be time consuming and costly, much research has been undertaken to produce smaller, more effective and simpler methods of plant simulation. These include more accurate small-scale systems, more carefully designed jar tests and alternative flocculation systems. In addition to simulation for optimisation, the small-scale systems can be applied in the design and modification of water treatment plants.

One of the important criteria linking the laboratory-scale tests (jar test), small-scale models and the pilot plants to the full-scale plant is the RTD. The RTD is obtained from the computational analysis of tracer test results. The RTD gives the overall flow pattern of the process by studying the distribution of time that fluid elements spend in the process. For the optimisation of a full-scale plant, the RTD can be used to identify non-ideal flow patterns such as dead space, and for linking the full-scale plant to the small-scale and the laboratory-scale tests. It is critical that the fluid elements in the small-scale model and laboratory systems remain in the system for the same duration as those in the full-scale system. In addition, it is essential that the elements are subject to the same hydrodynamic flow conditions.

Residence time distribution studies may be used to simulate the hydrodynamic flow patterns through processes by determining a flow model for the process. One way to improve the performance of treatment works is to improve the flow patterns through the unit processes of the system. If the flow pattern through a unit process is poor, for example, the entire volume of the process is not utilised or flow by-passes part of the process, then the performance will be impaired. A model of flow patterns, or a flow model, will suggest possible improvements to be postulated for more efficient operation and process optimisation. The effect of improvements may also be predicted. A basis for decision-making is thus provided in terms of the time, effort and cost of making the improvements. A flow model can, in addition, be used as a diagnostic tool in the case of process failures which manifest as changes in flow pattern. The length of time necessary to reach steady state after a change in conditions, such as inlet conditions, can be determined. Or, for example, if a pollutant enters a process, the flow patterns will show the quantitative distribution of the pollutant through the process.

Conventionally, laboratory tests are used to simulate the full-scale water treatment plant on a laboratory scale. The jar test is one of the most commonly used tests involving parameters such as the rapid mixing time, and the average velocity gradient (G value), the pH of the raw water and other physical and chemical characteristics which affect the efficiency of flocculation, and hence plant performance. The success of the jar test is dependent on the similarity of the mixing conditions on the plant and in the jar test. Jar tests are often neglected because they are considered to be too time consuming, and do not adequately match the performance of the plant. The jar test, if applied correctly, can closely simulate certain conditions of the plant. In order to achieve this, however, the conditions on each plant must be studied such that the sequence of mixing conditions can be successfully reproduced.

2 *PROJECT AIMS AND OBJECTIVES*

The original aims of the project were :

- The production of a Guide which would enable water authorities to construct small-scale equipment which would behave in a similar fashion to the existing potable water treatment plant.
- The investigation of changes to improve the performance of the existing potable water treatment plant by making physical modifications to the small-scale equipment.
- The design, construction and evaluation of a small-scale plant which would behave in a similar fashion to Umgeni Water's Wiggins Water Works. Both plants to the fed from the same source.

During the project, the Steering Committee targeted specific areas of study, which deviated from the original objectives. The areas targeted involved :

(i) The use of residence time distribution studies for water treatment optimisation.
(ii) A review of literature on the topic of flocculation and plant simulation.
(iii) The use of small-scale and laboratory-scale systems for water treatment optimisation.

3 *SUMMARY OF RESULTS*

While the aim of the work involves the optimisation of water treatment processes, the work was undertaken in three separate areas and is therefore discussed in terms of the comparison between the jar test and the full-scale plant, the design and construction of a small-scale system, and the use of residence time distribution to determine the flow patterns of systems. On a larger scale, however, the three levels of process optimisation are linked by the need for understanding the flow patterns.

3.1. Comparison Between the Jar Test and the Full-Scale Plant

In general the flocculation process consists of a high intensity or rapid mixing process and a low intensity or slow mixing flocculation stage. The rapid mixing stage is the major determinant of the characteristics of the subsequently formed floc. A literature survey was undertaken to determine the important parameters affecting coagulation, flocculation and sedimentation. Attention was paid to the methods of evaluation of flocculation performance, particularly the critiques of the conventional jar test.

The conventional jar test and the modified jar test are commonly used methods for the evaluation of flocculation efficiency, for the evaluation of coagulants and for the establishment of optimum conditions for flocculation. The jar test is limited since it is a batch test and in addition, problems are experienced with scale-up from the jar test to full-scale systems. More importantly, however, it is noted that the settling or sedimentation stage in the jar test is inappropriate for solids-contact clarifiers and it is concluded that the jar test cannot model this type of clarification process. It was concluded that, provided the mixing energy, residence time and particle concentration adequately simulate conditions on the plant, the unit processes of rapid mixing and slow mixing be maintained in the jar test. Comparisons can then be made between the floc characteristics occurring in the beaker and on the plant.

A survey of several Umgeni Water treatment plants was undertaken to determine whether the jar test was being used, whether the conditions used in the jar test were similar to those occuring on the plant, and the perceptions of staff regarding the importance and use of the jar test. Calculations of the mixing energy (G) value and the residence time (T) were undertaken to compare conditions on the plant to those selected in the laboratory. In addition to the estimation of GT values, detail was obtained regarding the flow through the process, taking account of time and mixing energy occurring at weirs. In addition, experimental work was undertaken at Wiggins Water Treatment Plant to optimise the jar test for that plant.

It was found that few of the mixing energy and residence time conditions occurring on the plant were reflected in the jar tests undertaken in the laboratory and that little confidence was held in the use of the jar test for guiding and monitoring plant performance. The cascade test was commonly used by plant personnel for monitoring plant performance. Since there was little correlation between the mixing energy and mixing time on the plant and in the jar test, it is recommended that efforts be made to quantify the mixing energy and mixing time on the plant and to use these values in the jar test. It was also found that the procedure of the jar test did not adequately account for the process of clarification occurring in the pulsator-clarifiers and it is for this reason that it is suggested that research be undertaken to determine the floc characteristics such as floc dimensions and settling rate prior to the settling stage of the jar test. The flocs occurring on the plant can then be compared with the flocs obtained in the jar test.

3.2. *The Use of Small-scale Systems*

Small-scale processes which will operate similarly to a full-size process can be constructed from a flow model. Trials, carried out when planning extensions to processes or modifications on processes, can be performed on a small-scale. In water treatment processes, typical changes investigated include : different blends of raw water ; different flocculants and coagulants ; the effect of pre-ozonation or chlorine dioxide addition ; and the need for and efficiency of granular or powdered activated carbon processes. The advantages of having a small-scale process would be ease of testing, reduced cost where chemicals are needed, as well as the convenience of side-by-side trials.

A 1/8 Perspex model of a full-scale hydraulic jump at Wiggins Water Treatment Plant was constructed according to the principles of similarity. The model was geometrically, kinematically (as determined by the analysis of the hydraulic gradients) and dynamically (as determined by the mean velocity gradients) similar to the full-scale plant. Tracer tests were performed on the model as well as the plant.

Similarity between the plant and the model was successfully achieved on two levels; i.e. geometrically (based on the scaling of the model) and kinematically (based on similar hydraulic gradients in the plant and the model). Dynamic similarity was achieved in the calculation of the mixing energy of the plant as determined by velocity gradient calculations. An acceptable level of similarity was obtained for this model compared to the plant. When the dynamic similarity is determined by residence time distribution methods, it was found that it was possible to simulate the plant operation in terms of the shape of the RTD curve, however the mean residence time was not similar.

In the case of water treatment literature shows that the combination of the effects of mixing energy (G), and time is critical since the flocculation process is determined by the intensity of mixing as well as the duration of mixing. See section 2.2.1 on rapid mixing and flocculation. Thus it would have been preferable to achieve dynamic similarity in terms of (Gt) rather than (G) alone. This insight was only attained after the model had been constructed.

3.3. *Residence Time Distribution*

At a larger scale, a tracer.tests can be used to determine the flow pattern of processes by means of residence time distribution analysis. The effluent stream from a continuous flow process is a mixture of fluid elements that have resided in the process for different lengths of time; the distribution of these residence times is an indicator of flow patterns within a process.

A user-friendly, interactive computer program, IMPULSE, was written for this study. It allows easy modelling of systems using curves obtained from tracer response tests. The user assumes a flow model for the system; the program allows determination of the response curve for the model, and adjustment of the estimated parameters of the model to fit the experimental curve by minimizing the sum of squares error (regression on the data).

IMPULSE manipulates concentration-time data directly in the time domain. Thus no transformation of experimental residence time distribution data (generally determined as concentration-time data) is necessary. The program allows for any input tracer signal and varying flow rates. Generally, constant velocity solutions are used as approximations. It is desirable, however, to have exact solutions where possible.

IMPULSE takes flowrate variation (and therefore velocity) into account.

The following data is entered into the program :

(i) the building blocks of the model (ideal units, dispersion unit, tanks-in-series unit)
(ii) the parameters associated with the building blocks (volume, split fraction, Peclet number)
(iii) whether the parameters must be held constant or can be varied by IMPULSE
(iv) a tracer input curve as a flowrate and concentration history

A tracer response curve (called the REFERENCE curve) is generated against which the model curve is fitted.

Flow modelling of a process and analysis of a process from a flow model needs to be accessible to the staff managing and operating water and waste water treatment works. Staff should be able to use the results to improve the performance of the works. Staff training has been recognised as crucial by the Water Institute of South Africa and the Department of Water Affairs and Forestry. This training should include flow modelling of the works.

Four courses on IMPULSE have been given by staff of the Pollution Research Group.

IMPULSE has been used in several applications to date :

- analysis of the results of a tracer study on an continuous flow anaerobic digester at Umbilo Waste Water Treatment Works. This work coincides with a trial on the anaerobic digestion of textile effluent undertaken in terms of WRC Project No. 456 : *The Regional Treatment of Textile and Industrial Effluents.*
- a tank sizing exercise was undertaken for Tioxide SA using Impulse to determine the optimum size required.
- Umgeni Water undertook a tracer test at Durban Heights Water Treatment Plant, and a dye test and a tracer test on a clarifier at Hazelmere Water Treatment Plant.

4 REALISATION OF OBJECTIVES

This project served to :

- indicate important parameters in the design and construction of small-scale water treatment systems.
- highlight important factors in the planning and execution of the jar test.
- identify and promote the use of tracer test studies and residence time distribution analysis for understanding flow patterns in water treatment plants.

5 *RECOMMENDATIONS*

For the optimisation of plant performance it is recommended that :

(i) The flow patterns of the individual unit processes be determined using tracer test studies and analysed using residence time distribution as the first step. During this phase non-idealities such as dead space and by-passing can be identified and rectified.

(ii) The rapid and slow mixing times and energies on the plant should be quantified and applied to the relevant stages in the jar test along with accurate simulation of the dose of chemicals and the order of the addition of the chemicals. The mixing energy can be calculated based on the dimensions of the plant and the flow to the plant. The residence time can be determined by undertaking tracer tests and analysing the residence time distribution. In addition, care should be taken to identifiy the flow process occurring on the plant in order to account for the presence variations such as mixing occurring at weirs.

(iii)Since the settling stage of the conventional and modified jar tests do not represent the pulsator-clarifier stage occurring on the plant, it is suggested that research be undertaken to analyse and compare the floc characteristics in the jar test and those on the plant.

(iv)Small-scale units or plants should be constructed according to the principles of similarity. Use can be made of tracer studies to make the residence time distribution of the fliud elements similar in the model and in the plant. Similarity of the residence time is indicative of similarity in flow patterns, which is one of the important factors in the design and construction of small-scale systems. In the case of water treatment processes both mixing energy (G) and mixing time (t) should be achieved simulaneously.

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Chapter 1

INTRODUCTION

Due to the increasing demand for potable water in South Africa, water treatment plants are currently under pressure to increase the production of water. As a consequence of increased water production, wastewater treatment plants will be faced with an increased flow. In order to respond to this increased demand, water and wastewater treatment plants have to expand. Increase in water production can occur with the construction of new processes, with the expansion of existing water treatment plants, and by the optimisation of current processes. The latter option of optimising existing equipment provides the first step in expanding water production since an increase in the efficiency of processes will lead to higher throughput and better treatment without new construction.

Water treatment involves the complex interaction of processes such as mixing, coagulation, flocculation, sedimentation, clarification and disinfection. Tools have been developed to simulate the full-scale water treatment plants in order to optimise and monitor the processes in terms of the chemical dosages and physical parameters. These tools include residence time distribution (RTD) studies, laboratory tests, and the construction of pilot plants and small-scale units. Residence time distribution studies may be used to determine the hydrodynamic flow patterns through processes by determining a flow model for the process. One way to improve the performance of treatment works is to improve the flow patterns through the unit processes of the system. If the flow pattern through a unit process is poor, for example, if the entire volume of the process is not utilised or if flow by-passes part of the process, then the performance will be impaired. A model of flow patterns, or a flow model, will suggest possible improvements to be postulated for more efficient operation and process optimisation. The effect of improvements may also be predicted. A basis for decision-making is thus provided in terms of the time, effort and cost of making the improvements. A flow model can, in addition, be used as a diagnostic tool in the case of process failures which manifest as changes in flow pattern. The length of time necessary to reach steady state after a change in conditions, such as inlet conditions, can be determined. Or, for example, if a pollutant enters a process, the flow patterns will show the quantitative distribution of the pollutant through the process.

Conventionally, laboratory tests are used to simulate the full-scale water treatment plant on a laboratory scale. The jar test is one of the most commonly used tests involving parameters such as the rapid mixing time, and the average velocity gradient (G value), the pH of the raw water and other physical and chemical characteristics which affect the efficiency of flocculation, and hence plant performance. The success of the jar test is dependent on the similarity of the mixing conditions on the plant and in the jar test. Jar tests are often neglected because they are considered to be too time consuming, and do not adequately match the performance of the plant. The jar test, if applied correctly, can closely simulate certain conditions of the plant. In order to achieve this, however, the conditions on each plant must be studied such that the sequence of mixing conditions can be successfully reproduced.

The construction of pilot-plant systems is a commonly used tool for simulating the full-scale since problems such as scaling effects and similarity are difficult to achieve in the small-scale systems and in the laboratory. As the construction and operation of pilot plants can be time consuming and costly, much research has been undertaken to produce smaller and simpler methods of plant simulation. These include more accurate small-scale systems, more carefully designed jar tests and alternative flocculation systems. In addition to simulation for optimisation, the small-scale systems can be applied in the design and modification of water treatment plants.

One of the important criteria linking the jar test, small-scale models and pilot plants to the full-scale plant is the RTD. The RTD gives the overall flow pattern of the process by studying the distribution of time that fluid elements spend in the process. It is critical that fluid elements in the small-scale model and laboratory systems remain in the system for the same duration as those in the full-scale system. In addition, it is essential that the elements are subject to the same hydrodynamic flow conditions.

1.1 Work Approach

The original aims of the project were :

- The production of a Guide which would enable water authorities to construct small-scale equipment which would behave in a similar fashion to the existing potable water treatment plant.
- The investigation of changes to improve the performance of the existing potable water treatment plant by making physical modifications to the small-scale equipment.
- The design, construction and evaluation of a small-scale plant which would behave in a similar fashion to Umgeni Water's Wiggins Works. Both plants being fed from the same source.

During the project, the Steering Committee targeted specific areas of study, which deviated from the original objectives. The areas targeted involved :

(i) The use of RTD studies for water treatment optimisation. Tracer tests can be used to determine the flow pattern of processes by means of residence time distribution analysis. Danckwerts developed the residence time distribution concept in 1953 to characterise the overall flow behaviour in a process. The effluent stream from a continuous flow process is a mixture of fluid elements that have resided in the process for different lengths of time ; the distribution of these residence times is an indicator of flow patterns within a process.

(ii) A review of literature on the topic of flocculation and plant simulation. In general, the flocculation process consists of a high intensity or rapid mixing process and a low intensity or slow mixing flocculation stage. The rapid mixing stage is the major determinand of the characteristics of the subsequently formed floc.

(iii) The use of small-scale and laboratory-scale systems for water treatment optimisation. A survey of several Umgeni Water treatment plants was undertaken to determine whether the jar test was being used, whether the conditions used in the jar test were similar to those occurring on the plant, and the perceptions of staff regarding the importance and use of the jar test In addition, experimental work was undertaken at Wiggins Water Works to optimise the jar test for that plant. A small-scale hydraulic jump, based on the mixer at the Wiggins Water Works, was designed, constructed and characterised during the course of this project. This 1/8 scale Perspex model was designed and constructed to be geometrically, kinematically and dynamically similar to the hydraulic jump on the plant.

Chapter 2

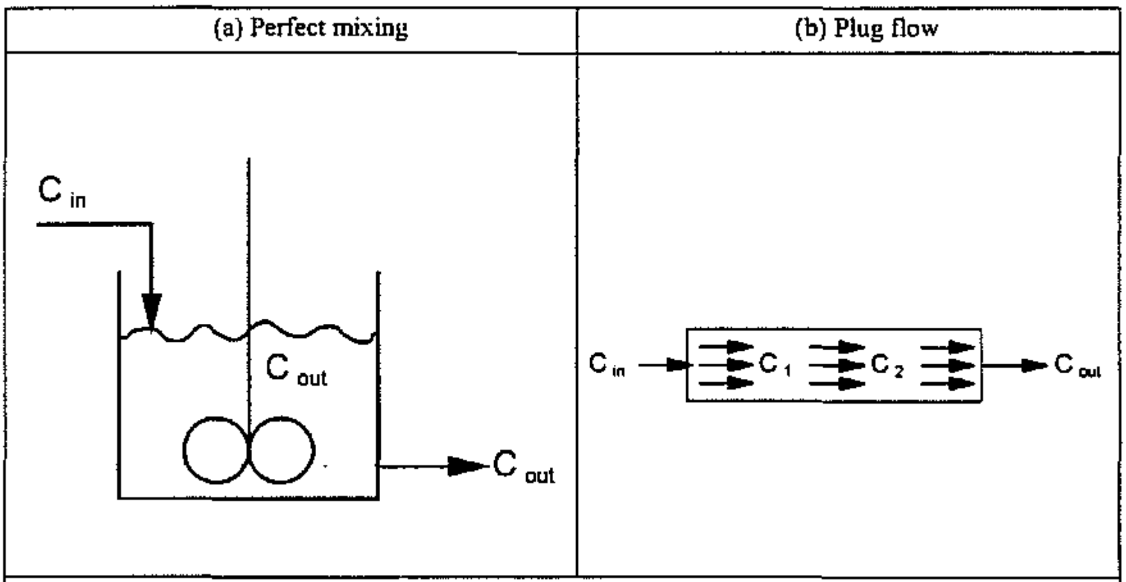
BACKGROUND

As identified in the introduction, three areas of study were recognised as important: the study of flow patterns in unit processes, a review of some of the important aspects of coagulation and flocculation relating to the monitoring and optimisation of water treatment plants, and the use of small-scale models for modelling and optimisation. Some background to these topics is presented in this chapter. Initially the application and analysis of tracer tests is highlighted. Some important criteria for water treatment are described, and, in particular, the use of the jar test is considered. Finally, the similarity criteria such as kinematic, dynamic and geometric similarity for the design of small-scale systems are provided.

2.1 The Determination of Flow Patterns in Unit Processes

In continuous flow processes, such as water treatment processes, the exit stream is a mixture of fluid elements that have resided in the process for different lengths of time. The distribution of these residence times is an indicator of flow patterns within a process.

Previously, in most processes, either perfect mixing or plug flow was assumed (Fig. 2.1).


(a) Perfect mixing

(b) Plug flow

Figure. 2.1 : Description of ideal flow processes

Perfect mixing implies that the fluid in the unit process is completely mixed, so that the properties of the fluid in the unit process and in the exit stream are uniform. **Plug flow** supposes no mixing in the direction of flow, so that elements of fluid entering the process at the same moment move through it on parallel paths, leaving at the same moment.

However, the flow patterns found in real processes usually lie between these two extremes (Smith, 1981), due to by-passing, channelling, dead space, dispersion and recycling. In **by-passing** some elements of fluid bypass the entire process, in contrast to **channelling** where some elements of fluid move through the process significantly faster than others do. **Dead space** refers to a region in the process with extremely poor contacting with the bulk fluid. Fluid in the dead space will generally remain in the process significantly longer than the bulk fluid. **Recycling** occurs when fluid is recirculated to the process inlet or to another region of the process.

The real flow patterns of processes can be determined. The introduction of tracer particles into the inlet stream of a unit process followed by the measurement of the concentration-time relationship of the tracer particles in the exit stream provides an indication of the distribution of the residence times of the tracer particles. A flow model of a process can be determined using the tracer residence time distribution response. The flow pattern of the process can then be inferred from the flow model.

2.1.1 Tracers for Water Treatment Processes

Water and waste water treatment unit processes are aqueous flow processes in which dissolved or suspended solids are removed. In this application, tracers must not adsorb onto, or react with, the solids. Further, the tracer should not be a pollutant or contaminant in the process. Tracers can also be applied to determine the residence time distribution of solids through unit processes, in which case the tracer should have the same flow attributes as the solids, and not leave the system in the aqueous phase. Tracers can generally be classed as radiotracers, dyes, or electrolytes.

Radiotracers, specifically water soluble salts of the solution isotopes, are recommended by Foldiák (1986) for tracing water and aqueous solutions. Many studies (Bergman, 1961; White, 1974) have used radiotracers for water treatment processes. These tracers include tritiated water, sodium-24 (as sodium), bromine-82 (as a bromide ion), and gold-98 (as a complex cation with ammonia).

Tritiated water has flow patterns identical to that of water. No specialised handling equipment is required for tritiated water since the radiation emitted cannot penetrate glass. Disadvantages are that it cannot be monitored on-line (White, 1974), it has a long half-life (12,3 years), so the amount used to trace must not contaminate the process, and it has a low (25 to 60 %) efficiency of detection (Kirk-Othmer, 1982). Sodium-24 has been used in water studies as a radiotracer. As the half-life is 15 h, no lasting contamination is possible. However, it requires specialised handling equipment as it is dangerous to handle. Similarly, bromine-82 and gold-98 require specialised handling and detection equipment. Additionally, the use of any radiotracer requires permission from a controlling board.

Dyes are useful for visual interpretation of the flow patterns in open water treatment and waste water treatment processes. Fluorescein, a readily-available fluorescent dye, is recommended by Smart and Laidlaw (1977) for water studies. Fluorescein has been approved for external human use in the United States, and is detectable at

0,29 µg/ℓ. The U.S. Geological Survey (in Smart and Laidlaw, 1977) recommend, however, where the water is to be used for human consumption, that the final concentration for all dyes used in tracer tests be below 10 µg/ℓ. Where the water from any water or waste water treatment process being traced passes through the system into a reservoir or dam, the tracer will generally be diluted to below 10 µg/ℓ. The concentration of the dye can be measured using a spectrophotometer.

The most common electrolytes used for tracing water processes are lithium chloride and sodium chloride (Agg et al, 1961 ; Nauman and Clark, 1991). Both tracers have been used with success in the sugar industry to trace aqueous flow systems (Rouillard and Smith, 1981). Sodium fluoride has been used as a tracer in waste water treatment processes (Vissers and Williams, 1984), where the anion was detected. Recommended limits in drinking water for these elements are 100 mg/ℓ for sodium, 2,5 mg/ℓ for lithium, 250 mg/ℓ for chloride and 1 mg/ℓ for fluoride (Pieterse, 1989).

Sodium chloride is inexpensive and non-hazardous. Sodium is generally present in higher concentrations in water treatment processes than lithium and is less easily detectable. Thus the concentrations of sodium chloride used are higher. Sodium chloride cannot, therefore, be used if the concentration in the effluent stream will exceed recommended limits, such as for drinking water. The concentration of sodium can be determined by atomic absorption spectrophotometry or by conductivity determinations using a conductivity probe. An assessment should be done prior to the tracer test to determine the amount of tracer adsorption onto the solids present in the process.

Lithium chloride is relatively inexpensive and lithium is detectable at concentrations of nanograms per litre by flame photometric methods (Agg et al, 1961). It is used where it is difficult to add large quantities of tracer to the process, and where the concentration of ions in the effluent stream must be minimised. Lithium is stable in solutions and is not lost by deposition (Agg et al, 1961).

2.1.2 Tracer Input and Monitoring

The introduction of tracer may be as a random signal, as a cyclic signal, a step or jump, as a pulse or as any arbitrary signal. The most common tracer signals, however, are pulse inputs, or changes in the form of step functions.

The experimental residence time distribution response may be influenced by the method of tracer addition and measurement, particularly when the fluid velocity is not uniform across the cross-section of the tracer injection point. Different methods of addition and measurement of the tracer will give different tracer responses. These responses can be related to the true experimental response, but are not identical to it. For determination of the true experimental residence time distribution response, input tracer quantities must be proportional to the velocity of the fluid at each radial position : this may be approximated by a rapid turbulent jet of tracer (a pulse), or by a step change. Measurement must be by means of the *mixing cup reading*, where samples are collected at different times and the average concentration of the samples is measured. Adding the tracer evenly across the injection point or measuring the tracer on-line, without disturbance of the flow, does not give the true tracer response.

2.1.3 Analysis of Tracer Residence Time Distribution Response

In the past residence time distribution theory has not been used to its maximum potential because of computational complexity of the analysis of the tracer test data. These difficulties can be overcome by the accessibility of personal computers and computer software.

A tracer experiment is performed on a process of interest. A flow model is then chosen that reasonably represents the process. It is assumed that the flow model has the same tracer input as in the actual experiment. The two curves are compared and fitted by changing the model parameters. The general modelling approach is shown in Fig. 2.2.

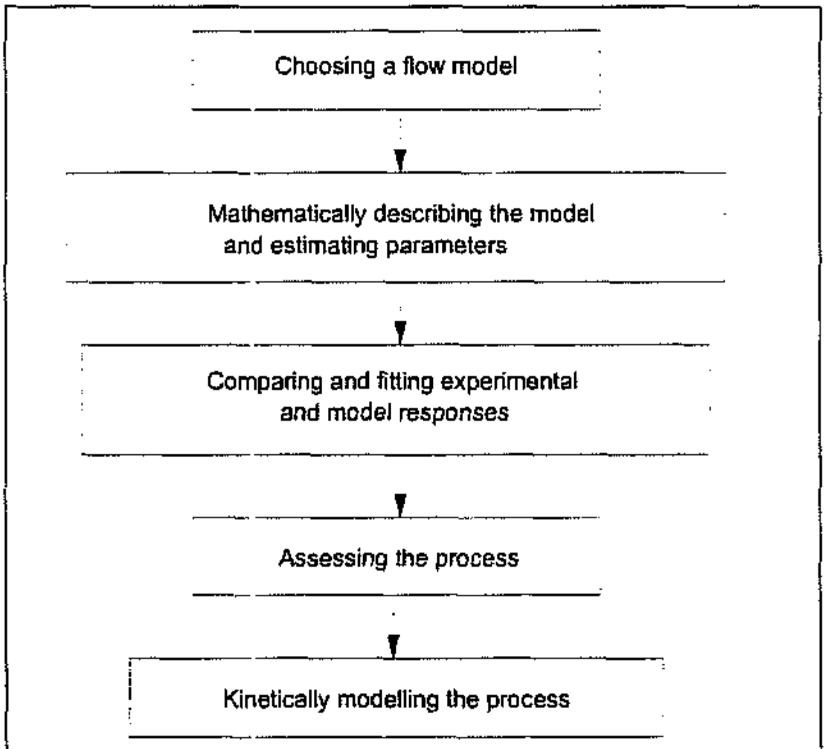


Figure. 2.2 : General modelling approach

The flow model chosen should reasonably approximate real flow. This is done by qualitatively examining the experimental response and comparing it to the residence time distribution responses from known flow models, and by observing the physical process.

The residence time distribution responses for common flow models are documented (Levenspiel and Bischoff, 1963). Matching the shape of the experimental response to the responses for flow models is useful in choosing a flow model. The physical process can be used to choose a model that reasonably approximates real flow. If any visual test was performed (for example, a dye test on an open process), elements of the visible flow pattern can be used. The model must contain the necessary elements to be physically meaningful (Himmelblau and Bischoff, 1968).

It should be noted that to totally account for non-ideal flow, however, knowledge of the complete flow pattern of a process is necessary (Levenspiel and Bischoff, 1963). Tracer experimentation is limited by the fact that the measurement is of residence times and not the particular location of fluid in space (Robinson and Tester, 1986). The residence time distribution indicates the length of time spent in the process by any fluid element, but not the time when fluid elements of different ages mix in the process (Bailey and Ollis, 1986). Information from the residence time distribution is therefore not sufficient to completely define non-ideal flow.

2.1.4 Implications of Non-idealities in Flow

Non-idealities of flow will affect various unit water treatment processes in a variety of ways. Generally, processes are designed to achieve ideal flow, and deviations from this lead to poorer performance and inefficient operation.

In rapid mixing, by-passing can occur which would lead to the poor contact between the dosed polyelectrolyte and the particulates in the water. This would result in poor flocculation since the particle collisions would have been limited. By-passing in a classifier would lead to untreated water overflowing with the treated water stream. In anaerobic digesters, by-passing means that no digestion occurs on a portion of the stream.

If dead space exists in flocculators during the slow mix phase, particle interactions would be minimised and the extent of collision and agglomeration would be reduced. In addition, the active volume would be reduced, therefore the actual residence time allocated for flocculation would not be achieved. This is also true for anaerobic digesters, where dead space would reduce the volume available for digestion, and the residence time would therefore be lower for the same throughput , adversely affecting the anaerobic reaction.

Where plug flow is necessary, such as in some clarifiers, recycling of material can lead to poor clarification due to disturbance of the sludge blanket. In mixed processes, recycling is considered a non-ideality that may enhance the mixing.

2.2 Basic Water Treatment

Conventional water treatment involves the processes of rapid mixing, coagulation and flocculation (or slow mixing), sedimentation, filtration and disinfection. Coagulant chemicals are added during the rapid mix stage to destabilise the colloidal material in the water. During the coagulation/flocculation stage the destabilised particles aggregate to form flocs which subsequently settle during sedimentation or clarification. The performance of each treatment unit affects the efficiency and operation of subsequent units, e.g. larger flocs resulting from an increased detention time, will result in improved sedimentation and longer filter runs (Ramaley et al, 1981). The efficiency of the flocculation, sedimentation and filtration processes depends on many factors including the number, size and mass of particles in the water to be treated.

Coagulation is the process whereby destabilisation of a given suspension or solution is effected (i.e. the function of coagulation is to overcome those factors which promote the stability of a given suspension) (Bratby, 1980). Flocculation occurs when destabilised particles or particles formed as a result of destabilisation make contact, and thereby form larger particles (Bratby, 1980). Natural thermal movement and Brownian motion result in perikinetic flocculation, while turbulence induced by mechanical means results in orthokinetic flocculation.

Colloids and particulates occurring in raw water contribute to the turbidity of the water. These colloidal and suspended particles deflect light waves passing through the sample. Turbidity is dependent on the nature, size, concentration and refractive index of the particles in suspension. Turbidity is not, however, a direct measure of the amount of suspended material in the given water. Turbidity is a commonly used analysis parameter for determining the efficiency of a water treatment process.

The destabilisation of particles and colloids is achieved by the addition of a primary coagulant. Primary coagulants include metal ions as well as polylectrolytes. Other chemical additives such as flocculant aids are added to a destabilised suspension to accelerate the rate of flocculation.

The term polyelectrolyte refers to a variety of water soluble macromolecules with molecular weights ranging from 10^4 to 10^7 daltons, which have the ability to destabilise a colloidal suspension. These macromolecules are generally polymers with readily ionisable functional groups. In solution the macromolecules ionise, creating charged areas on the polymer which can result in adsorption of colloids onto the polymer and hence destabilisation. More than one colloidal particle can be adsorbed onto a macromolecule. Polyelectrolytes that possess both positive as well as negative charges are referred to as ampholytic and those that possess no ionisable functional groups are called non-ionic electrolytes. Some common polyelectrolytes include guar gums, tannins and synthetic polymers. The synthetic polymers are the most popular polyelectrolytes used in industry especially because they can be *tailor made* to suit the raw water to be treated. Properties of polyelectrolytes that affect flocculation include charge density, ionic strength of the solution and molecular weight.

2.2.1 Rapid Mixing

Rapid mixing is the process used to disperse coagulants into untreated water. Aspects such as rapid mixing time and speed have been identified as important factors which affect the subsequent coagulation and flocculation processes. Literature on coagulation processes indicates that extremely rapid dispersion is necessary for efficient use of coagulants (Amirtharajah and Mill, 1982; Hudson and Wolfner, 1967). While there is much debate regarding the selection of mixing devices, the mixing energy applied and the duration of mixing, there is no doubt, however, that rapid dispersion of coagulants is advantageous in the rapid mixing phase of water treatment.

The degree of mixing has traditionally been quantified by the root mean square velocity gradient. The velocity gradient was first used by Von Smoluchowski (Polasek and Van Duuren, 1979). In order to account for the duration of the rapid mixing, Camp and Stein (1943 in Polasek and Van Duuren, 1979) combined the root mean square velocity gradient (G) (Eqn 2-1) and rapid mixing time (t) to give a dimensionless number called the Camp number (Ca) (Eqn 2-2). The two parameters which are generally used to quantify the energy associated with flocculation are the velocity gradient and the Camp number.

$$G = \left(\frac{e_d}{\mu}\right)^{\frac{1}{2}} \tag{2-1}$$

$$Ca = Gt \tag{2-2}$$

Where :

μ	viscosity	($N.s.m^{-2}$)
e_d	dissipated power per unit volume	($W.m^3$)
t	space time in a continuous floccolator or time in in a batck reactor	(h)

Both these parameters are derived from the overall energy dissipation. This concept has, however, been questioned by various authors (Han and Lawler, 1992, Polasek, 1979, Cleasby, 1984), and is discussed in **Appendix A.**

2.2.2 Sedimentation

Sedimentation is the process in which the flocs formed during flocculation are allowed to settle by gravity. The efficiency of this separation process is related to loading rate, water quality, temperature, floc size, floc weight and vessel flow conditions. The conventional sedimentation tanks were based on horizontal flow, more recently, however, upflow sedimentation basins have been developed (Hudson, 1981). These basins include clarifiers, clarifier-flocculators, solids contact reactors, tube settlers and plate settlers. It is proposed that floc formation is enhanced by the mass action of floc formation in the presence of previously formed masses of floc in the solids-contact clarifiers (Hudson, 1981).

2.2.3 Monitoring Flocculation Efficiency

Various methods have been devised for monitoring the degree of flocculation as a tool to determine the efficiency of the process. Turbidity is the most commonly measured parameter since the aim of the water treatment process is the clarification of turbid water. Where solids recovery is the objective, the settling rate would be the appropriate parameter to measure. Other methods of evaluation include the monitoring of the settling rate of the flocs, the density of the flocs, the charge of the solution containing the flocculated particulates, the electrophoretic mobility of the solution and the particle size of the flocculated particulates.

Ideally the flocculation test procedure should simulate the operation of the full-scale plant and be capable of predicting plant performance based on laboratory trials. This is difficult to achieve since very often the optimum conditions indicated in the laboratory are not successful when applied to the plant. This can be due to a lack of similarity of conditions on the plant and in the laboratory system. The actual performance of the plant, in terms of clarification, settling rate and other parameters may not be the same as that found in the laboratory test under the same chemical conditions, but it should be possible to establish an empirical relationship which can be used for predictive purposes.

2.3 Jar Test

The jar test is one of the more conventional flocculation evaluation techniques. In the conventional jar test several samples of raw water are placed in a series of vessels, and are stirred in a manner similar to that on the plant. Various dosages of chemicals and coagulants are added. After an appropriate mixing period the stirring is stopped and the floc formed is allowed to settle for a specific time period. The clarity of the supernatant is then used to evaluate the optimum chemical dosages. This procedure can be modified to include a filtration stage. This is often performed with filter paper, and in some instances with the construction of laboratory scale sand filtration systems. The turbidity of the filtered water is then determined. The jar test which includes a filtration stage is known as a *modified jar test.*

Critical factors which should be considered when performing a jar test include the mixing energy provided by the stirring mechanism, the duration of mixing, the manner in which the coagulants and additional chemicals are dosed and the collection of supernatant from the jar.

In addition to determining the optimum dose of coagulant, the jar test can be used for :

- determining the strength of the floc
- predicting filtered water quality
- determining the optimum coagulant pH
- determining the optimum mixing intensity
- evaluating the optimum mixing intensity taper
- evaluating rapid mix duration and intensity
- evaluating the effect of lag time between rapid mix and flocculation
- evaluating optimum coagulant dosages
- evaluating the sequence of coagulant and coagulant aid dosages
- establishing the most effective coagulant
- evaluating the effects of sludge recycling and concentrations of sludge for recycling
- predicting the effects of short circuiting
- predicting sludge deposit configuration
- predicting design criteria for in-plant settling

- establishing design criteria for tube and tray settlers
- evaluating direct filtration possibilities and coagulant dosages

The major disadvantages of the jar test are that it is a batch test which can be very time-consuming, large volumes of water are required (100 *ℓ* or more), and most importantly the results obtained from a jar test may not correspond with the results obtained on a full-scale plant since a small-scale mixing device cannot be assumed to be a true hydraulic model of the plant. In addition, the size of floc particles cannot be scaled down according to the principles of similitude. Many authors including Gregory (1983) indicate the need for a more rapid, convenient and more representative test than the jar test.

The jar test has been modified to model sedimentation basins using a hydraulic comparison equation sensitive to basin flow rates, water residence times and flocculant dosages by Reed and Reece (1984). In this work a hydraulic comparison calculation is used to relate the jar test to the actual operating conditions and is verified by monitoring those conditions in the plant with the same coagulant and settling conditions. From the curves generated the operator could look up the appropriate settling time for the jar test by knowing the operating flow rate for that shift.

The *cascade test* (Fig. 2-3) is a further variation of the jar test. In this test the rapid mixing stage is replaced by the actual mixing process that occurs on the plant. Water is removed from the plant after the rapid mixing stage and the processes of slow mixing and settling occur in the beaker, followed by filtration. This test was devised after it was acknowledged that the rapid mixing conditions on the plant were not adequately represented in the beaker. Final water turbidities in the cascade test can provide an indication of the actual final water turbidities. More often, however, the actual turbidities differ between the plant and the beaker, but operators find that particular ranges of turbidity values occurring from the cascade test indicate particular final water qualities. This technique is often considered more useful than the modified jar test.

Various other systems have been designed to replace the conventional and modified jar tests. These include spiral flocculators and couette flocculators. The processes of rapid mixing, slow mixing and settling are still achieved, but most of the research has dealt with the flocculation process. The reason for this is that the jar test is a batch process, and the spiral flocculator can be used in a continuous flow process, thereby providing more accurate results.

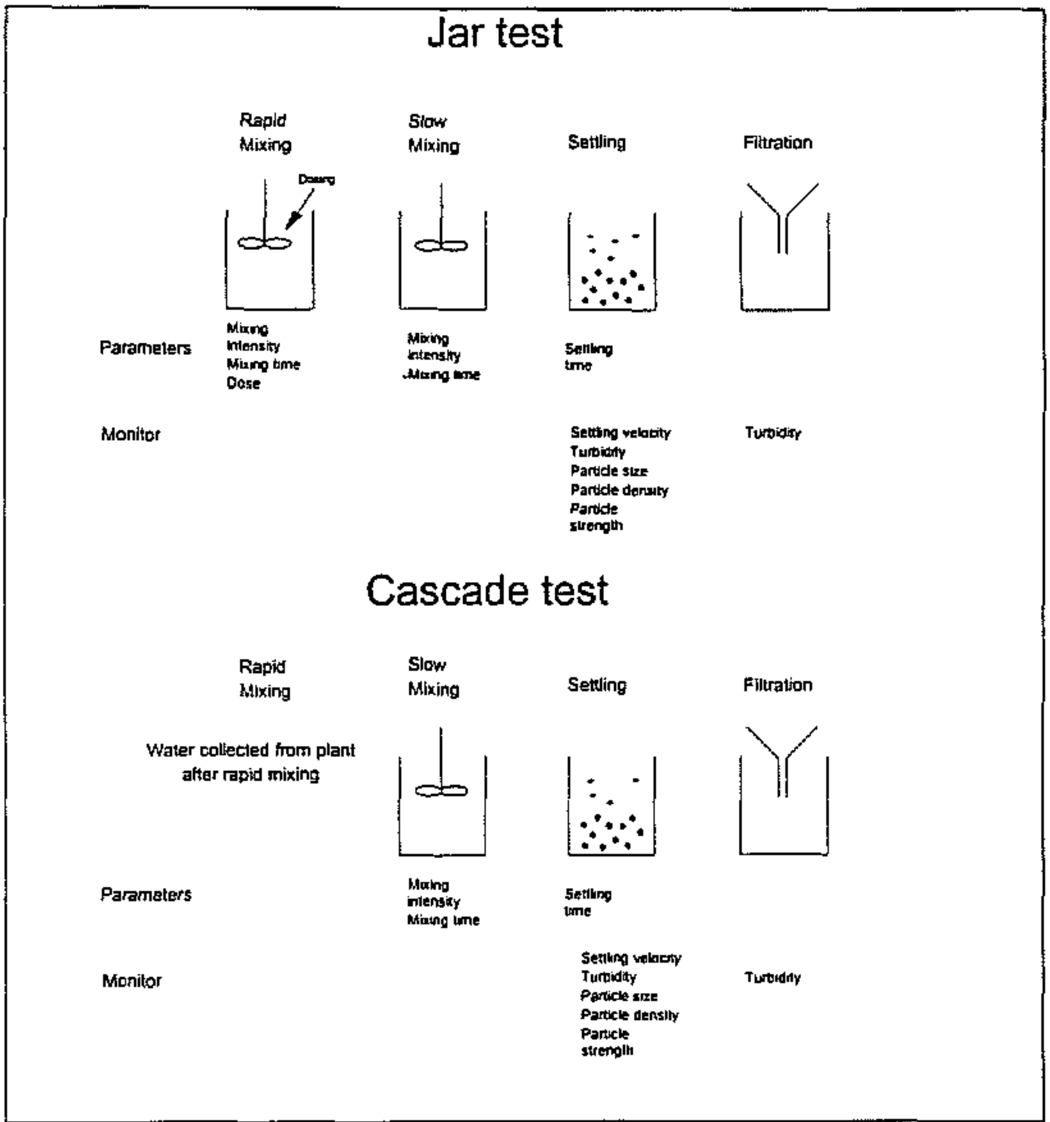


Fig. 2-3 : The modified jar test and the cascade test

One of the major, and most often neglected disadvantages of the jar test is the applicability of the sedimentation stage. The sedimentation of flocs in a beaker is unrepresentative of upflow in solids contact clarifiers since particle interactions are different in the two systems. The jar test sedimentation is considered suitable for conventional sedimentation basins only, and it is unrealistic to consider that the process of clarification occurring in clarifiers can be simulated in a beaker. The jar test is, however, still applied in many inappropriate instances, with poor results.

2.4 Small-scale Equipment

In order to design a scale model of a plant, the laws of physical similarity have to be satisfied. Two systems are said to be physically similar in respect to certain specified quantities when the ratio of corresponding magnitudes of these quantities between the two systems is everywhere the same (Massey 1985).

Generally results from experiments performed on the scale model are used to design as well as predict the behaviour of the prototype. The prototype can be either larger or smaller than the model. The proficient use of the laws of physical similarity together experimental investigation on the behaviour of the model can provide a better understanding of the prototype.

There are three types of physical similarity :

(i) geometric similarity,
(ii) kinematic similarity,
(iii) dynamic similarity.

Geometric similarity is similarity of shape. The characteristic property of geometrically similar systems, is that the ratio of any length in one system to the corresponding length in the other system is everywhere the same (Massey, 1985). This ratio is known as the *scale factor*. Not only must the model be geometrically similar to the prototype, but the roughness of the surfaces must also be geometrically similar. This means that if the model is small and the surfaces are smooth, then the prototype (which is larger than the model) should have surfaces which are rougher. It is not always possible to achieve perfect geometric similarity, however it is essential that geometric similarity be achieved even if it is only in terms of the scale factor. Once this is achieved, the similarity laws can then be applied to achieve kinematic and dynamic similarity.

Kinematic similarity is similarity of motion. This implies that there must be similarity of lengths (geometric similarity) as well as time intervals. If the lengths as well as the corresponding time intervals of the two systems are in a fixed ratio then the velocities of the corresponding elements of fluids must be in a fixed ratio of magnitude at corresponding times.

Dynamic similarity is similarity of forces. Therefore if two systems are dynamically similar, then the magnitudes of forces at similarly located points in each of the two systems, must be in a fixed ratio. For pipe flow, usually the Reynolds number is the same for the model and the prototype. In the case of open channel flow the upper surface of the liquid is exposed to the atmosphere and only the liquid is exposed only to gravitational and inertial forces. The Froude number, a ratio between inertial and gravitational forces is applicable for channel flow.

Chapter 3

MODELLING OF WATER TREATMENT PLANTS FROM RTD DATA

3.1 Residence Time Distribution

Previously, the residence time distribution analysis of tracer tests has not been used to its full potential due to mathematical complexity. Often, tracer test results were merely used as an indication that the flow patterns in the process were non-ideal. No flow model was derived, and the parameters associated with the process were not determined. This research aimed to develop an accessible method to analyse tracer tests to determine the flow model and the parameters of the process.

In the method, the experimental and model responses were compared using the time domain fitting method. The major disadvantage of time domain fitting, numerical complexity, was overcome using a personal computer program, IMPULSE, written for this study. IMPULSE allows relatively simple comparison of a number of models for any particular experimental response. The IMPULSE manual is provided in **Appendix B**.

The user assumes a flow model for the system ; the program allows determination of the response curve for the model, and adjustment of the estimated parameters of the model to fit the experimental curve by minimising the sum of squared errors (regression on the data).

IMPULSE manipulates concentration-time data directly in the time domain. Thus no transformation of experimental tracer test data (generally determined as concentration-time data) is necessary. Bischoff (1964), in his solution of the axial dispersion equation for time variable flow, states that the velocity varies in many applications. Generally, constant velocity solutions are used as approximations. It is desirable, however, to have exact solutions where possible. The program allows for any input tracer signal and varying flow rates.

IMPULSE takes flowrate variation (and therefore velocity) into account.

The following data is entered into the program :

(i) the building blocks of the model (ideal units, dispersion unit, tanks-in-series unit),
(ii) the parameters associated with the building blocks (volume, split fraction, Peclet number),
(iii) whether the parameters must be held constant or can be varied by IMPULSE,
(iv) a tracer input curve as a flowrate and concentration history,
(v) a tracer response curve (called the REFERENCE curve) against which the model curve is fitted,
(vi) the connections between the building blocks.

A flowsheet for the program is given in Fig. 3.1. As IMPULSE is interactive, a flowsheet cannot show all the options available. This is, therefore, only one possible flowsheet.

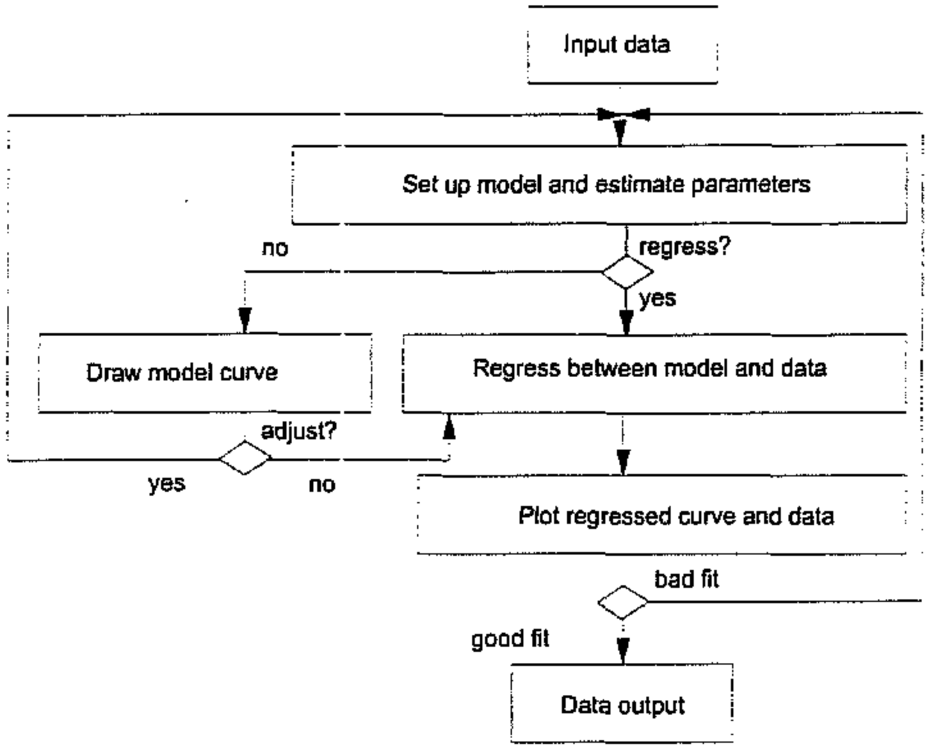


Figure 3-1: A possible flow sheet for IMPULSE

The output from the program after modelling is displayed on the screen and can be saved to a file. The data saved are :

(i) model parameters,
(ii) input curve concentration and flowrate history,
(iii) reference curve concentration history,
(iv) model curve concentration history.

3.1.1. Available Units

IMPULSE has 7 units that can be connected to yield a model of the system. Some units have parameters associated with them. Every unit must have a unique name identification and connection requirements. There is no restriction on the number of units. However, the limitations of personal computers lead to a trade-off between the number of units and the number of time segments. Table 3.1 gives a description of the units and their requirements.

TABLE 3.1 : IMPULSE units

Name	Description	Requirements	
Mixed flow reactor (MFR)	a perfectly mixed region	input output parameters	1 1 volume
Plug flow reactor (PFR)	an ideal plug flow region	input output parameters	1 1 volume
Axial dispersion (PFRDISP)	a region to which the dispersion model can be fitted	input output parameters	1 1 volume Peclet number
Tanks-in-series (MFRSER)	a region to which the fractional tank extension model can be fitted	input output parameters	1 1 volume number of tanks
Input	an input to the process : the tracer input	input output parameters	none 1 flowrate/time concentration/time
Output	the response curve of the model to the input, with either estimated or regressed parameters	input output parameters	1 none concentration/time
Mixer/splitter (Mix Split)	a unit which joins or splits streams	input output parameters	at least 1 at least 1 flow split

Main program features:

(i) The model constructed by the user may be modified at any time by simply adding or removing units and connections.

(ii) Any set of concentration curves generated by the model or entered by the user can be plotted on the screen using a single set of axes for comparative purposes. The plot is regenerated automatically as the user alters modelling parameters and re-runs the model, allowing for almost instantaneous feedback.

(iii) Input flow and concentration profiles can be loaded from an ASCII text file, which can be generated by a spreadsheet program or an ASCII text editor.

(iv) The user provides estimates of the parameters of the model. Any of the parameters estimated may be chosen to be fitted to the experimental tracer response by regression or to be kept at a given value. Thus the effect of changing one or more, but not necessarily all, the parameters may be seen. The regression procedure calculates the sum of squared errors between the model output curve and the experimental curve (Nauman and Clark, 1991) and alters all the parameters chosen for regression to minimise the sum of squared errors.

(v) The data for the response curves can be saved to a file for subsequent plotting using a spreadsheet program.

3.1.2. Limitations to IMPULSE

IMPULSE should only be used to model continuous or near-continuous flow processes since the technique and the program assume a continuous flow through the process. Additionally, the program models the liquid residence time of the process, and not the solids residence time. Tracing of the solids in a process is possible if they can be marked with, for example, a radioactive tracer. The residence time distribution of the solids could then be determined the same way as for liquid residence time given that the same information is known about the solids (for example, the flow of solids).

The determination of the tracer response curve is not trivial and may influence the result obtained from IMPULSE. Thus a user must be aware of how this is obtained, any experimental error that may be associated with obtaining it, and how this influences the curve. Also, the program has been designed for conservative tracers, that is, tracers that are conserved in the process under study and are not consumed in the process.

IMPULSE works with concentration-time responses. Thus, if data is available in another from (for example, radioactivity-time data), this should be converted to concentration. Although the modelling will be possible without this step, the results will be meaningless as the parameters and data will not be in the same units.

A user should have a physical knowledge of the particular unit to be modelled and should be able to propose feasible flow models. Any parameters obtained from the flow model must have physical significance; thus a complex model may fit the tracer curve but may not be feasible. For open processes, a dye test may be used to propose a physically feasible flow model. Further, tracer curves are not unique: they can be modelled accurately by more than one model. Sound engineering judgement is therefore necessary to choose the most feasible model.

IMPULSE works by dividing time into *segments* and then applying the equations describing the system to each segment in turn. A model is sensitive to the number of segments chosen: the larger the number of segments the more accurate the result.

The program is limited by MS DOS memory and there is a trade-off between the number of time segments and the model complexity.

3.1.3. Guidelines

The following guidelines are provided to assist those undertaking tracer tests and using IMPULSE :

(i) Experimental technique : in a continuous flow process the flowrate, as a function of time must be determined accurately. It is essential to record details of the amount of tracer and the concentration of the tracer added, the duration over which the tracer was added to the flow. Sound experimental technique should be observed when taking samples and the analysis method should be appropriate and carefully undertaken.

(ii) Choice of tracers is important : certain tracers are unsuitable for specific purposes, for example, sodium chloride would not be used in a potable water treatment plant. Lithium chloride is more acceptable in this case. Account should be taken of the interaction between the tracer and the liquid, solids and the system. In reacting systems, the tracer must not interfere with the reaction in any way.

(iii) Understand the process : effort should be made to understand the water treatment process in order for the model to be realistic. A model that reasonably represents the process should be chosen, as, given a large number of parameters, a model can be made to fit any response.

(iv) Use of IMPULSE : Impulse can be used to analyse the data from a tracer study. The limitations of the program must be taken into account for useful information to be obtained.

Chapter 4

COMPARISON BETWEEN JAR TESTS AND PLANT PERFORMANCE

In order to determine the important parameters which affect the efficiency of flocculation, and the use of monitoring methods such as the jar test, a survey of several Umgeni Water Works was undertaken. This survey aimed to :

- obtain an appreciation for current practices at the various plants regarding the process, the coagulants used and additional chemicals introduced into the process,
- determine the frequency and accuracy of jar tests and cascade tests,
- determine estimates of the average velocity gradients (G-values) representative of the rapid mixing undertaken at the plants,
- determine the importance of the G value and residence time values in the various stages of flocculation, and
- determine whether the parameters of the jar test undertaken in the laboratory were representative of the conditions on the plant.

The survey criteria included : the source of raw water to the plant; the type of rapid mixing on the plant (where necessary the dimensions were obtained); the flow rate to the plant and the division of flow before and after rapid mixing; the type of clarifier; the clarifier upflow velocity and overflow rate; the fate of the sludge solids; the solids concentration of the sludge; de-sludging practices; the number of filters in operation; the flow to the filters; aspects of filter management; the type and concentration of chemicals added to the raw water; jar test procedure adopted by the plant, particularly the rapid mix time and speed; the frequency of the jar test in the daily laboratory routine; the procedure of the cascade test; the frequency with which the cascade test is performed; the sample point for the cascade test; the opinions on the validity and accuracy of both the jar test and the cascade test; the correlation between the jar and/or the cascade test and the plant performance and, whether the streaming current detector (SCD) is used and how the streaming current detector is set, and choice of set point.

Estimates of the average velocity gradients were calculated based on the dimensions and relevant information for the rapid mixers of the plants. These values and calculated theoretical residence times were compared with the mixing times and speeds employed in the routine jar test and cascade test. This information was also compared with the conditions employed by the Head Office laboratory of Umgeni Water where jar tests are undertaken for the evaluation of new flocculants and assist in trouble-shooting at the plants. The findings of the survey are presented in **Appendix C**.

Eqns 4-1 and 4-2 were used in the calculation of the velocity gradient for weirs and for electric mixers:

WEIR

weir
$$G = \left(\frac{h\,q\,g}{V\,\nu}\right)^{\frac{1}{2}} \tag{4-1}$$

electric mixer
$$G = \left(\frac{P}{V\mu}\right)^{\frac{1}{2}} \tag{4-2}$$

Where :

G = average velocity gradient (s^{-1})
h = height of weir (m)
q = volumetric flow rate (m^3s^{-1})
g = acceleration due to gravity ($m.s^{-2}$)
V = volume of mixing chamber (m^3)
μ = solute viscosity = 10^{-3} at 20°C ($N.s.m^{-2}$)
ν = kinematic viscosity = 10^{6} at 20°C ($m^2.s^{-1}$)
P = electric power drawn by motor (W)
=

The holding time t (or theoretical residence time) was calculated using eqn 4-3.

$$t = \frac{V}{q \times 10^3 \div 86\,400} \tag{4-3}$$

During the survey several factors were noted regarding the jar test :

- On occasion the optimum dosage indicated by the jar test was less than that used on the plant.
- In several cases the actual dose of polyelectrolyte was up to three times the dose indicated in the jar test.
- The plant turbidities were often lower than the filtered water in the cascade test. This could indicate that there is a significant difference in the filtration efficiencies of the plant and the filter paper, and that the condition of the floc is different on the plant and in the laboratory.
- Certain polymer dosages did not *work* on the plant. This means that the dosage of polyelectrolyte obtained in the jar test, when applied to the plant, does not produce similar turbidity values. This can often be explained by the fact that the laboratory perform the jar test to obtain a turbidity of 0,5 NTU, whereas most of the plant operators prefer to work to a limit of 0,2 NTU. Consequently, additional polymer is often

required to achieve the lower turbidity. This makes comparison and evaluation of the accuracy of the jar test difficult.

The purpose of the cascade test is to determine whether the coagulant is being dosed, and whether the dosage on the plant is correct, such that acceptable turbidities are obtained. The purpose of the jar test is to confirm that the dose is appropriate, alter dosages when water conditions change, causing problems on the plant, and to evaluate new products and dosages.

CJB add in a paragraph on GTc is more appropriate....the c's are the same. Tambos comment that not GT but GTC must be similar.

The outcome of the survey indicated that there should be :

(i) similarity between the jar test and the plant in terms of the rapid mixing energy and time (GT),
(ii) similarity between the jar test and the plant in terms of flow processes,
(iii) use of the jar test,
(iv) modification of the jar test to simulate the plant.

4.1 Case Study

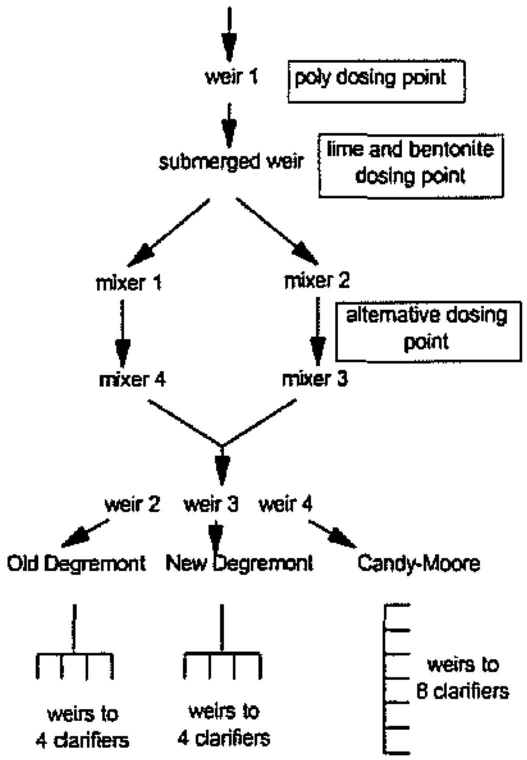
After the calculation of estimates of the velocity gradient at the rapid mixers at the 8 water treatment plants surveyed, discrepancies were was found at 7 of the 8 plants. In general it was found that the jar test G values were lower than the G values occuring in the plants (Table 4.1). The mixing times for the rapid mix stage were also found to differ between the jar tests and the plants. In all cases the time was calculated from the flowrate and the geometry of the system. As a result of the discrepancies between both mixing energy and mixing time, the GT values differed between the plant and the laboratory. In addition to the differences in both time and mixing enegy, it was found that no account was taken of the actual flow processes occurring in the surveyed plants. This aspect is considered important since the behaviour of the floc particles depends on the shearing strength of the water in which the flocs are suspended before reaching the clarifiers.

Taking the example of the Durban Heights Water Works it was found that the G values in the jar tests by both the central laboratories and the plant laboratores were 60 s^{-1} , whereas the calculated G value for rapid mixing weir No. 1 was 175 s^{-1}. The resultant GT values of 22 400 for the plant and 7 200 for the jar test indicate a large discrepancy in mixing conditions. By looking at the flow process (Fig. 4.1a) of the Durban Heights Water Works, it can be seen that the polyelectrolyte is dosed at the weir. The flow then proceeds over a submerged weir (for which the velocity gradient was not calculated) and into chambers with electric mixers. GT values of 22 400 followed by 14 000, followed by 13 000 are experienced by the flow before it is split at a diversion weir. The flow then moves along channels before reaching one of 3 weirs providing GT values of 7 000, 19 000 and 21 000 respectively. While the numbers of GT will vary according to the flow to the plant, it is considered that the GT value of 7 200 as used in the jar test is inappropriate. Sample calculations of the values of G and T are presented in **Appendix C**.

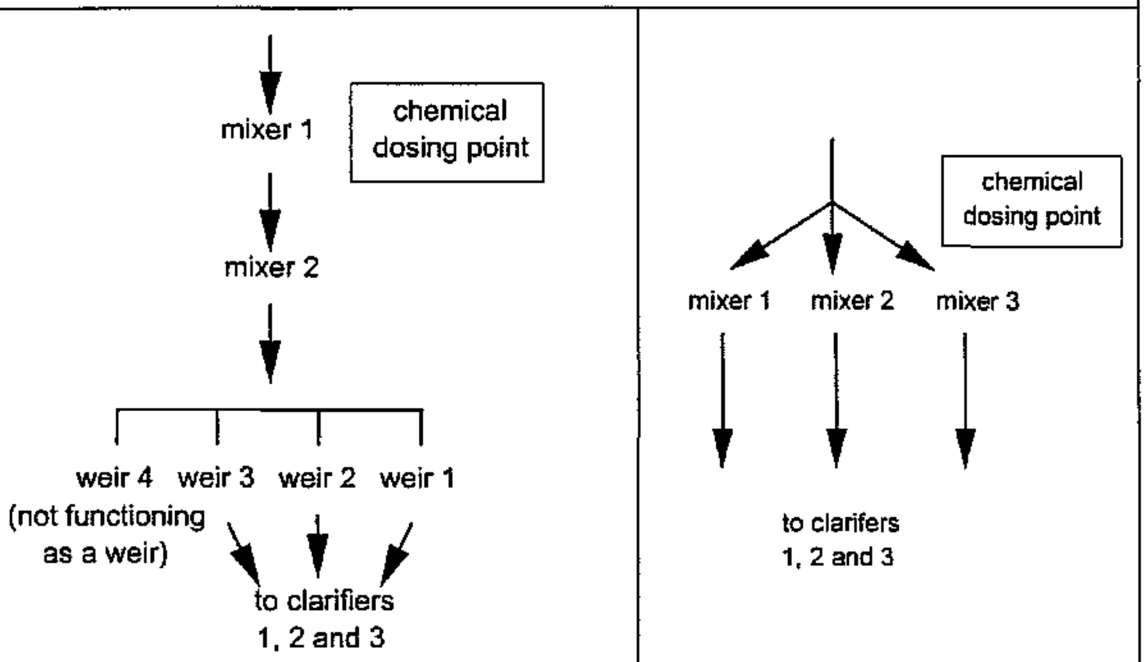
TABLE 4.1 : Summary of findings of 1994 survey

	PLANT (1)				JAR TEST : ON-SITE				JAR TEST : CENTRAL LAB			
PLANT	**Mixing mech.**	**Rapid mix**			**Rapid mix**				**Rapid mix**			
		G-value (s^{-1})	**Time (s)**	**GT (-)**	**Speed (rev^{-1})**	**G (s^{-1})**	**Time (s)**	**GT (-)**	**Speed (rev^{-1})**	**G (s^{-1})**	**Time (s)**	**GT (-)**
DURBAN HEIGHTS (450 Ml/d)	weir 1	175	128	22 400	100	60	120	7 200	100	60	120	7 200
	mixers 1 and 2	211	68	14 263								
	mixers 3 and 4	225	59	13 365								
	weir 2	no volume details available										
	Candy-Moore weir	696	11	7 588								
	New Degremont	498	40	19 928								
	Old Degremont	435	49	21 266								
HAZELMERE (23,5 Ml/d)	mixer 1	218	183	39 776	300	250	180	45 000	300	250	120	30 000
	mixer 2	224	183	41 040								
	weir 1	278	99	27 528								
	weir 2	527	34	17 641								
	weir 3	505	30	14 950								
H D HILL (11 Ml/d)	mixer	937	23	21 543					300	250	120	30 000

(1) Flow diagrams of the plants are given in Fig 4.1.



(a) Durban Heights Water Works



(b) Hazelmere Water Works

(c) DV Harris Water Works

Figure 4.1: Flow schemes at three water works

At the Hazelmere Water Works (Fig 4.1b) the raw water enters at mixer 1 where the polyelectrolytes are dosed. The flow then moves to mixer 2 after which the flow is split over one of three weirs before entering the clarifiers.

Mixer 1 accounts for a GT value of 39 000 followed by 41 000 for mixer 2 and then either 27 000, 14 000 or 17 000 for the weirs. At this plant the jar test procedure allows for a GT value of 45 000 and 30 000 for the central laboratories. While these numbers are similar to the GT value at the inlet, no account is taken of the subsequent mixing conditions.

At the HD Hill Water Works the flow follows a simple process where the water is dosed and split to three mixers with GT values of 21 000. The jar test undertaken for this plant has a GT value of 30 000 which appears to be appropriate since there is no futher significant mixing occurring in the flow. Similar comparisons can be undertaken for the other 7 plants (**Appendix C**).

At most of the plants it was found that neither the rapid or slow mix stages of the plant were accurately represented in the jar test procedure.

The above findings were confirmed in a study of the comparison of jar test results with the performance of the Wiggins Water Works (Brouckaert, Progress Report, 1993). The standard jar test procedure at the Wiggins Water Works was found to give over-optimistic predictions of plant performance because use is made of a rapid mixing time of 180 s, whereas the actual average holding times on the plant is 43 to 57 s. Experimental work indicated that :

- by using an appropriate rapid mixing time in the jar test, the rapid mixing in the plant can be reasonably matched,
- the filtration stage in the jar test can be matched to the plant provided an appropriate filter paper is selected which represents the plant filtrate quality,
- the results from the study suggest that the jar test is dependent on slow mixing times in the range of 15 to 30 min,
- the pulsator-clarifiers cannot be modelled by the jar test, however it was shown that the final filtrate qualities during the period of the study are roughly equivalent, indicating that the jar test could be used to predict the filtrate quality at the specified operating conditions.

This work concluded that valuable information regarding the floc characteristics can be obtain by a process of simulated rapid mixing and slow mixing followed by analysis of the floc particles. The sedimentation process has been inappropriately applied to water treatment plants where pulsator clarifiers are in operation.

Chapter 5

SMALL-SCALE EQUIPMENT

5.1 Wiggins Water Works

At the time of the study the capacity of the Wiggins Water Works was 175 Mℓ/day and the average flow to the plant was 160 Mℓ/day. Raw water entering the plant was pre-chlorinated and pre-limed before being split equally into four smaller channels which lead to the hydraulic jumps. At each hydraulic jump there was a coagulant dosing point. The hydraulic jumps produce a high degree of turbulence which is used for the rapid mixing of the coagulant. The dosed water leaving the hydraulic jumps then flows into four 4 000 m^3 upward flow pulsator clarifiers. These clarifiers are sludge blanket clarifiers.

Each clarifier has a retention time of approximately 2,4 h with an average up-flow velocity of approximately 2 m/hr when the plant is running at a capacity of 160 Mℓ/day. The sludge blanket formed is approximately 2 m in height. A vacuum pump pulsates the sludge blanket every 40 s by sucking sludge from the sludge blanket into a vacuum chamber and releasing it. The sludge blanket also acts as filter for finely dispersed particles. The fully formed sludge overflows into a hopper while the clarified water flows out through the top of the clarifier.

The hydraulic jump rapid-mixer consists of an upper open channel connected to a lower channel via a downward sloping weir. The flow of water from the upper channel to the lower one causes an increase in velocity. This increase initiates the formation of the hydraulic jump in the lower channel. The turbulence caused by the hydraulic jump is used to mix the coagulants. The coagulant is fed from a feed tank to a pipe distributor which is positioned at point A in Fig 5.1. The coagulant flows out of this distributor and falls evenly onto the water surface.

The level of the water in the lower channel is controlled by the water level in the clarifier. Visual observations suggest that the turbulent mixing zone is in the narrow part of the lower channel. It is expected that the rapid mixing is achieved in this part of the lower channel.

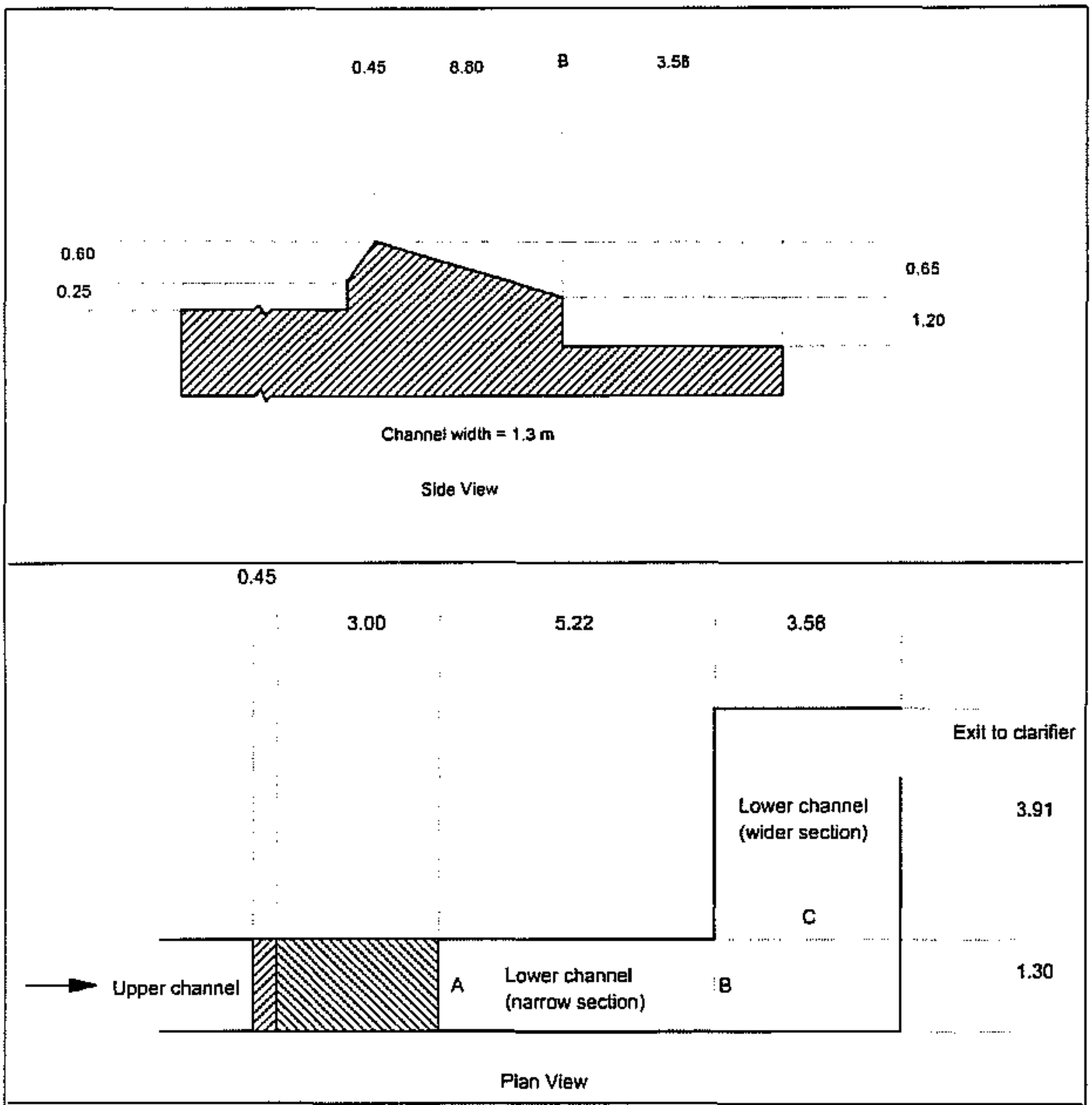


Figure 5.1 : Dimensions of the full-scale hydraulic jump (m)

5.2 Criteria for the Design of a Scale Model

In order to design a scale model of the plant hydraulic jump, the laws of physical similarity have to be satisfied. Two systems are said to be physically similar in respect to certain specified quantities when the ratio of corresponding magnitudes of these quantities between the two systems is everywhere the same (Massey, 1985).

Generally results from experiments performed on the scale model are used to design as well as predict the behaviour of the prototype. The prototype can be either larger or smaller than the model. The appropriate use of the laws of physical similarity together with experimental investigation on the behaviour of the model can provide a better understanding of the prototype.

Three types of similarity were considered :

(i) geometric similarity,
(ii) kinematic similarity,
(iii) dynamic similarity.

Ideally for a model and a prototype to be completely similar, all relevant conditions of similarity must hold. In some instances, due to complexities of the number of variable such as forces and velocities which have to be satisfied simultaneously, this is not always possible. In these cases, the minor variables can be neglected.

5.3 Design of the Model Hydraulic Jump

In open channel flow, kinematic and dynamic similarity between the prototype and the model was achieved by conserving the Froude number (**Appendix D**).

Observations of the mixing in the plant indicate that the most turbulent zone is between sections A and C in Fig. 5.1. The zone between section C (Figure 5.1) and the clarifier is far less turbulent, and it was assumed that only secondary mixing occurs in this region.

The model (Fig. 5.2), constructed of clear Perspex, was scaled down geometrically so that it represented the above mentioned mixing volume. The lower channel was made longer to reduce downstream effects on the mixing upstream.

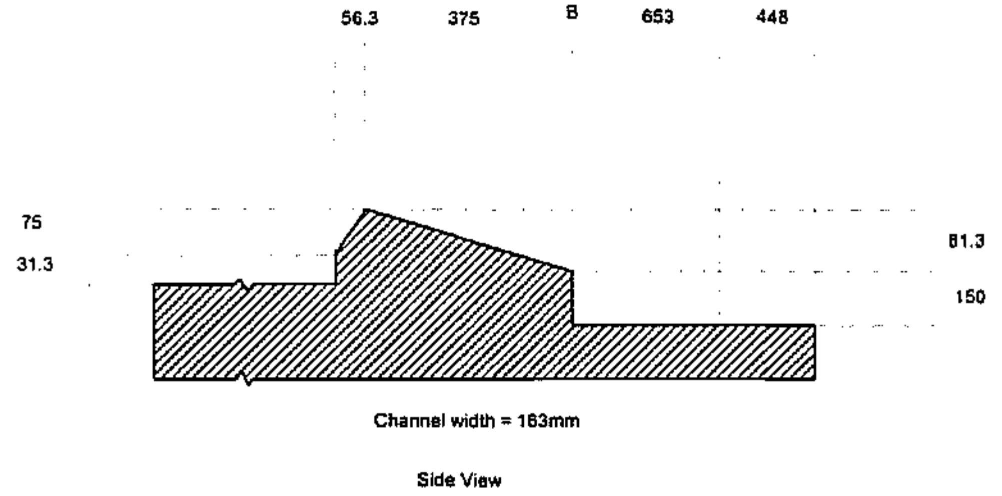


Figure 5.2 : Dimensions of the model hydraulic jump (in mm)

5.4 Characterisation of the Model Jump

For rapid mixing to be similar between the plant hydraulic jump and the model, energy dissipation and flow conditions have to be similar between the two. The geometric similarity was achieved. However this satisfies only one part of physical similarity. It was necessary to verify the kinematic and dynamic similarities between the plant hydraulic jump and the model by characterising the fluid dynamics and the energy dissipation and mixing in the plant and model hydraulic jumps. This was done by comparing the hydraulic gradients and the velocity gradients.

The rapid mixing stage of the flocculation process is one of the most important stages of flocculation. The efficiency of the rapid mixing often depends on the amount of energy that is dissipated during this stage. Sufficient energy must be available for the efficient dispersion of the coagulants. The amount of energy that is dissipated in an open channel can be derived from hydraulic gradients. The energy dissipation obtained can then be used to determine the mean velocity gradients.

For flow conditions in the model to be kinematically similar to conditions on the plant, the hydraulic gradients at kinematic similar flowrates should be similar (i.e. the liquid depth at a point in the plant should be similar to the liquid depth at that equivalent point on the model for kinematic similar flowrates). This was done by measuring the liquid depths in the plant and the model at kinematically similar flow rates to the plant. The depths obtained on the plant were divided by eight and compared to the actual depths obtained on the plant. This comparison showed a reasonable approximation of the plant hydraulic jump with respect to hydraulic gradients with percentage error ranging between 0 and 14 % . Since the hydraulic gradients are similar it is clear that the energies being dissipated in the two systems should be similar at similar points.

The mean velocity gradient developed by Camp (Bratby 1980) is the most widely used term to describe the degree of mixing achieved during any stage of flocculation. In order to determine the mean velocity gradient of the mixing process it is necessary to know the energy that is dissipated. The velocity gradients on the plant as well as the model were calculated and it was found that the model hydraulic jump compare well with the calculated velocity gradients of the plant hydraulic jump with an average percentage error of 4,7 % . This confirmed that the energy dissipation between the model and the plant hydraulic jumps is physically similar.

5.5 Residence Time Distribution Tests on the Model and the Plant.

One method that is used widely to characterise the degree of mixing in unit processes is to determine the residence time distribution. Residence time distributions can also indicate the flow non idealities of a system. Both global pulse and local pulse tracer tests were performed on the model using different concentrations of sodium chloride and monitoring conductance. The global test was done by measuring the residence time distribution of the exit stream of the model hydraulic jump in order to obtain the overall mean residence time of the hydraulic jump. The *local* pulse tracer test was done by measuring the response to a pulse at various locations in the lower channel in order to get a better understanding of the flow patterns in the mixing zone. Residence time distribution tests show that the model hydraulic jump behaves similar to a plug flow reactor in series with a mixed flow reactor. A high degree of backmixing does occur and this backmixing increases the efficiency of the mixing in the lower channel.

A residence time test was undertaken on the Wiggins plant in order to compare the results with those obtained on the Perspex model. The test involved injecting a pulse of lithium chloride at the flocculant addition point. Figure 5.3 shows a direct comparison between the tracer response curves of the perspex model and the plant. Since the flow through the plant on the day of the test was 180 Mℓ/d, which did not correspond to the 160 Mℓ/d that the model experiments were set up to simulate, it was necessary to adjust the time scale to obtain the most meaningful comparison between the plant and the model results. This is shown in Figure 5.4.

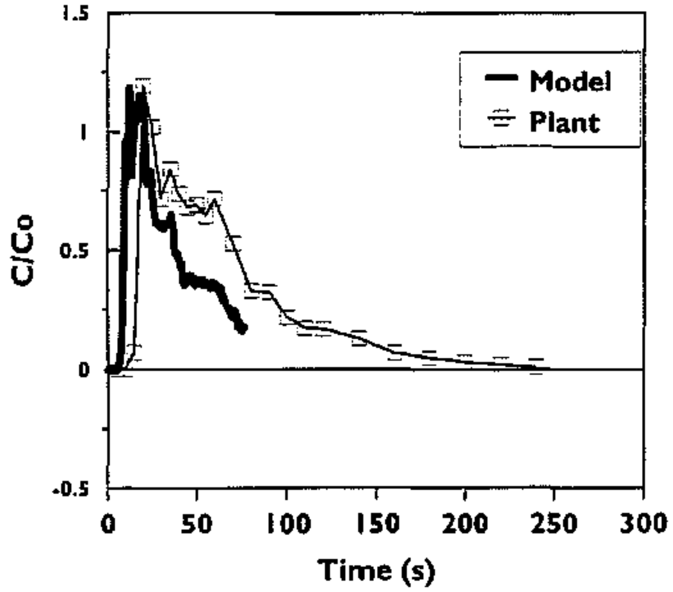


Figure 5.3 : Comparison of the RTD for the Plant and the Model

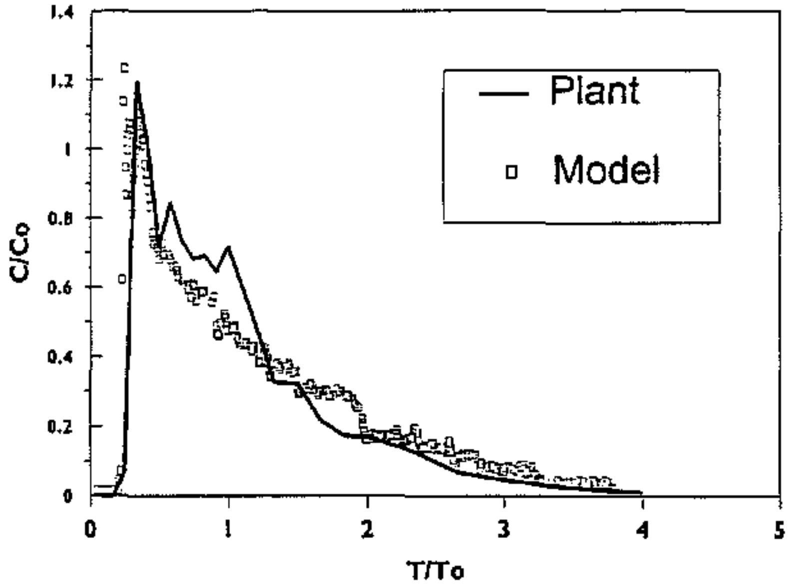


Figure 5.4 : Corrected Comparison of the RTD for the Plant and the Model

From Fig. 5.3 it is clear that while the tests resulted in RTD curves of similar shapes, the mean residence time was shorter in the model than in the plant. Fig. 5.4 shows that the general shape of the response curves are similar, but Fig 5.3 shows that the mean residence time of the plant is considerably longer than that of the model, in spite of the throughflow being relatively too high on the day.

Similarity between the plant and the model was successfully achieved on two levels; i.e. geometrically (based on the scaling of the model) and kinematically (based on similar hydraulic gradients in the plant and the model). Dynamic similarity was achieved in the calculation of the mixing energy of the plant as determined by velocity gradient calculations. An acceptable level of similarity was obtained for this model compared to the plant. When the dynamic similarity is determined by residence time distribution methods, it was found that it was possible to simulate the plant operation in terms of the shape of the RTD curve, however the mean residence time was not similar.

In the case of water treatment literature shows that the combination of the effects of mixing energy (G), and time is critical since the flocculation process is determined by the intensity of mixing as well as the duration of mixing. See section 2.2.1 on rapid mixing and flocculation. Thus it would have been preferable to achieve dynamic similarity in terms of (Gt) rather than (G) alone. This insight was only attained after the model had been constructed.

Chapter 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Residence Time Distribution

Traditionally RTD has not been used to the full potential due to the mathematical complexity. This research aimed to determine an accessible method to analyse RTD's. The experimental and model responses were compared using the time domain fitting method. This method was chosen as the results are easy to interpret as they are available in the time domain. The major disadvantage of time domain fitting, numerical complexity, was overcome using a personal computer program, IMPULSE, written for this study. IMPULSE allows relatively simple comparison of a number of models for any particular experimental response.

A user-friendly, interactive computer program, IMPULSE, was written for this study. It allows easy modelling of systems using curves obtained from tracer response tests. The user assumes a flow model for the system ; the program allows determination of the response curve for the model, and adjustment of the estimated parameters of the model to fit the experimental curve by minimizing the sum of squares error (regression on the data).

IMPULSE works with concentration-time data directly in the time domain. Thus no transformation of experimental residence time distribution data (generally determined as concentration-time data) is necessary. The program allows for any input tracer signal and varying flow rates. Bischoff (1964), in his solution of the axial dispersion equation for time variable flow, states that the velocity varies in many applications. Generally, constant velocity solutions are used as approximations. It is desirable, however, to have exact solutions where possible.

IMPULSE takes flowrate variation (and therefore velocity) into account.
The following data is entered into the program :

(i) the building blocks of the model (ideal units, dispersion unit, tanks-in-series unit),
(ii) the parameters associated with the building blocks (volume, split fraction, Peclet number),
(iii) whether the parameters must be held constant or can be varied by IMPULSE,
(iv) a tracer input curve as a flowrate and concentration history.

A tracer response curve (called the REFERENCE curve) is generated against which the model curve is fitted.

6.2 Jar Test and Plant Comparisons

Following a literature survey of the important factors in flocculation, and particularly on mixing time and energy required for rapid mixing, a survey was undertaken of the current situation at eight Umgeni Water Works regarding the use of the conventional jar test. The survey highlighted the discrepancies between the conditions of the laboratory jar test and the conditions on the plants.

The conceptual procedure of the jar test is sound. It is a procedure where an attempt is made to simulate processes occurring on the plant in a beaker, however the attempted simulation of the sedimentation stage in the normal jar test is inappropriate for sludge blanket clarifiers.

Advances have been made in monitoring equipment such as the dual beam photometic method for measuring floc density and settling velocity. While tests of this nature are applicable and necessary for on-line process control and monitoring a simple procedure is still required to :

- determine the effect on flocculation of changes in dose,
- evaluation coagulants for water treatment plants,
- monitor the effects of changes to the plant on flocculation such as the addition of new mixing systems,
- provide a relatively rapid means of reacting to vast changes in water quality with confidence that subsequent changes on the plant will be successful.

Based on this it is suggested that the rapid and slow mix stages of the jar test procedure are maintained, and efforts be made to improve on the similarities between mixing energy and retention time and flow pattern. This can be achieved by calculation of mixing energy values and the use of tracer studies for residence time values. The the settling process be monitored in terms of floc characteristics such as floc strength, density and settling velocity. If the characteristics for a *good* floc can be identified, it will be possible to predict the success of subsequent processes based on the floc characteristics

While it is acknowledged that advances are being made in the areas of on-line monitoring equipment, the need for a small-scale, relatively rapid procedure is highlighted.

6.3 Small-scale model

A 1/8 Perspex model of a full-scale hydraulic jump at Wiggins Water Works was constructed according to the principles of similarity. The model was geometrically, kinematically (as determined by the analysis of the hydraulic gradients) and dynamically (as determined by the mean velocity gradients) similar.

Tracer tests resulted in RTD curves of similar shapes, the mean residence time was shorter in the model than in the plant. The general shape of the response curves are similar, but the mean residence time of the plant was considerably longer than that of the model, in spite of the throughflow being relatively too high on the day.

Similarity between the plant and the model was successfully achieved on two levels; i.e. geometrically (based on the scaling of the model) and kinematically (based on similar hydraulic gradients in the plant and the model). Dynamic similarity was achieved in the calculation of the mixing energy of the plant as determined by velocity gradient calculations. An acceptable level of similarity was obtained for this model compared to the plant. When the dynamic similarity is determined by residence time distribution methods, it was found that it was possible to simulate the plant operation in terms of the shape of the RTD curve, however the mean residence time was not similar.

In the case of water treatment literature shows that the combination of the effects of mixing energy (G), and time is critical since the flocculation process is determined by the intensity of mixing as well as the duration of mixing. See section 2.2.1 on rapid mixing and flocculation. Thus it would have been preferable to achieve dynamic similarity in terms of (Gt) rather than (G) alone. This insight was only attained after the model had been constructed.

The application of a small-scale model inludes:

- evaluating new polymers for flocculation,
- evaluating the performance of flocculation,
- determining the effect of flow rate on mixing and hence on flocculation.

6.4 *Recommendations*

For the optimisation of plant performance it is recommended that :

(i) The flow patterns of the individual unit processes be determined using tracer test studies and analysed using residence time distribution as the first step. During this phase non-idealities such as dead space and by-passing can be identified and rectified.

(ii) The rapid and slow mixing times and energies on the plant should be quantified and applied to the relevant stages in the jar test along with accurate simulation of the dose of chemicals and the order of the addition of the chemicals. The mixing energy can be calculated based on the dimensions of the plant and the flow to the plant. The residence time can be determined by undertaking tracer tests and analysing the residence time distribution. In addition, care should be taken to identifiy the flow process occurring on the plant in order to account for the presence variations such as mixing occurring at weirs.

(iii)Since the settling stage of the conventional and modified jar tests do not represent the pulsator-clarifier stage occurring on the plant, it is suggested that research be undertaken to analyse and compare the floc characteristics in the jar test and those on the plant.

(iv) Small-scale units or plants should be constructed according to the principles of similarity. Use can be made of tracer studies to make the residence time distribution of the fliud elements similar in the model and in the plant. Similarity of the residence time is indicative of similarity in flow patterns, which is one of the important factors in the design and construction of small-scale systems.

(v) Both mixing energy (G) and mixing time (t) should be achieved simultaneously in applications where processes such as flocculation occur.

6.5 *General*

The optimisation of water treatment processes can be undertaken at various levels. The jar test or laboratory-scale level involves the simulation of the processes of rapid mixing, slow mixing, sedimentation and filtration. The small-scale system often consists of the construction of individual process units which can be used for experimentation. Both of these systems are fraught with difficulties. In the case of the jar test difficulties include the fact that the test is a batch test, that the mixing in a beaker is not similar to that on a plant and that the clarification process cannot be simulated in a beaker. It is necessary for small-scale systems to be designed as scaled-down versions of the full-scale plant. The scaling-down process can result in a lack of kinematic and dynamic similarity. Very often, these factors are neglected.

The broarder approach of residence time distribution studies applied to the full-scale plant can facilitate the design and operation of both the jar test and the small-scale systems. In addition, and of most importance is the view that RTD's can indicate aspects such as non-idealities of flow on the full-scale plant, and thereby provide a first step in the optimisation process. The information from the RTD can be applied into the other processes since aspects such as holding time in vessels, flow rates to areas of the plant, and flow patterns should be taken into account in the design and operation of small-scale and laboratory-scale systems.

The program IMPULSE has provided a user-friendly environment for the analysis of RTD data. With training, this tool can be made available to plant personnel and research facilities in the water treatment field. In addition, RTD's and IMPULSE can be applied to situations other than the water treatment field.

Chapter 7

TECHNOLOGY TRANSFER ACTIVITIES

7.1 IMPULSE Program and Manual :

The program has been upgraded and now includes the following reactor units :
plug flow reactor, mixed flow reactor, mixer/splitter, plug flow with dispersion
The updated manual is presented in **Appendix B.**

7.1.1. IMPULSE Courses

Four courses on IMPULSE have been given by the Pollution Research Group.

- Water Institute of Southern Africa - May 1993 (10 course participants)
- South African Institute of Chemical Engineers - September 1993
- International Association of Water Quality 1994 : Anaerobic Digestion Conference - February 1994 (10 course participants)
- Afriwater Conference - June 1994 (4 course participants) (Table 7.1)

7.1.2. IMPULSE on the Internet

Impulse is now available on Internet through file transfer protocol (FTP). The host computer is aqua.ccwr.ac.za and the directory is /ftp/pub/impulse/.

7.1.3. Applications of IMPULSE

Impulse has been used in the following applications to date :

- analysis of the results of a tracer study on an continuous flow anaerobic digester at Umbilo Waste Water Works,
- a tank sizing exercise was undertaken for Tioxide SA using Impulse to determine the optimum size required,
- tracer studies on a clarifier and a rapid mixer at two Umgeni Water Works.

7.2 Interaction with WRC projects

The application of IMPULSE highlighted the need to calculate flow patterns in unit processes at water treatment plants using Computational Fluid Dynamics (CFD). This has resulted in WRC Project No.648.

The work on the digester at Umbilo Waste Water Treatment Works was part of a trial on the anaerobic digestion of textile effluent undertaken in terms of WRC Project No. 456 : The Regional Treatment of Textile and Industrial Effluents.

7.3 Papers and Conference Proceedings

Conference Proceedings :

Barnett, J.L., Brouckaert, C.J., Baddock, L.A.D., de Benito, B. and Buckley, C.A., **Impulse : Using Personal Computers for Residence Time Modeling,** 5th Biennial PC Symposium (Effective Use of Personnel Computer), SA Institution of Chemical Engineers (N-Tvl Branch), Conference Centre, University of Pretoria, 21 April 1993.

Pillay, V.L., Buckley, C.A., Naidoo, V., Godefory, S. and Carliell, C.M., **Impulse : A PC Program for the Determination of Residence Time Distribution of Biological and Chemical Reactors,** Seventh International Symposium on Anaerobic Digestion, Cape Town, South Africa, 23-27 January 1994.

Brouckaert, C.J., Baddock, L.A.D., Barnett, J.L. and Buckley, C.A., **Impulse : A PC Program for the Determination of Residence Time Distribution of Biological and Chemical Reactors,** Seventh International Symposium on Anaerobic Digestion, Cape Town, South Africa, 23-27 January 1994.

Brouckaert, C.J., Botha, C.J., Baddock, L.A.D. and Buckley, C.A., **Impulse : A PC Program for the Determination of Residence Time Distribution of Biological and Chemical Reactors,** WISA International Specialist Conference on Disinfection of Potable Water, Kruger Park, South Africa, 13-18 March 1994.

The Application of Chemical Engineering Techniques in Environmental Management at Tioxide SA Q E Hurt, B Brouckaert, C J Brouckaert and P Ramlall. Submitted to Water SA, August 1994.

7.4 Interaction with other organisations:

There has been interaction between PRG and Umgeni Water both in G value determination and in undertakiing tracer tests and residence time distribution studiess. The small-scale rapid mixer was construced at the Process Evaluation Facility based at Wiggins Water Treatment Plant.

IMPULSE has been used by the research branch CIRSEE (Centre International de Recherche sur l'Eau et l'Environement) of Lyonnaise des Eaux.

The IMPULSE Program and manual are lodged at the Computing Centre for Water Research (CCWR) and is available for downloading through the Internet.

A meeting was held in September 1995 with Dr Tambo of Hokkaido University. The purpose of the meeting was to exchange views on the current trends in the field of flocculation. The meeting was held at the Process Evaluation Facility at Wiggins Water Works, Durban.

The Atomic Energy Corporation uses Impulse for the interpretation of their isotope tracer tests.

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Water Research Commission Project No. 363

The use of small-scale equipment for evaluating water treatment plants

APPENDIX A

Literature Review on Coagulation, Flocculation and Water Treatment

Compiled by

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University of Natal
Durban

November 1998

1 General Water Treatment

Surface waters usually consist of a stable suspension of colloidal particles. The *stability* relates to the capacity of the colloidal particles to remain in suspension as independent entities over an infinite period of time (Bratby, 1980). Stability occurs as a result of the overriding influence of surface phenomena such as surface charge, surface area and degree of hydration. The performance of each treatment unit affects the efficiency and operation of subsequent units, e.g. larger flocs resulting from an increased flocculation detention time, will result in improved sedimentation and longer filter runs (Ramaley et al, 1981). Conventional water treatment involves the processes of rapid mixing, coagulation and flocculation (or slow mixing), sedimentation, filtration and disinfection. In water treatment it is advantageous to convert the colloidal particles to larger agglomerates so that physical separation can be achieved. This is achieved by modifying the surface properties of the colloidal particles thus destabilising the suspension. Coagulation and flocculation are the two processes involved in the destabilisation of these suspensions.

The interrelationship between the water treatment processes is difficult to analyse quantitatively therefore much research has been undertaken on the individual processes with the aim of the development of equipment or processes which adequately simulate a water treatment plant.

Coagulation is the process whereby the suspension is destabilised with the addition of a chemical coagulant. The function of coagulation is to overcome those factors which promote the stability of a given suspension (Bratby, 1980).

Flocculation occurs when destabilised particles or particles formed as a result of destabilisation make contact and form larger particles (Bratby, 1980). Natural thermal movement and Brownian motion results in perikinetic flocculation, while orthokinetic flocculation is induced by turbulence which is generally caused by some form of mechanical agitation. Flocculation is often enhanced with the addition of a flocculant aid. In this case, a chemical is added to a destabilised suspension to accelerate the rate of flocculation.

Turbidity refers to the lack of clarity of water due to the presence of colloidal and suspended particles. These particles and colloids deflect light waves passing through the sample and produce a representative turbidity. Turbidity is dependent on the nature, size, concentration and refractive index of the particles in suspension, and is therefore not a direct measure of the amount of suspended material in the given water. Turbidity can be measured by

- Turbidimetry, which involves comparing the transmission of light through a standard suspension to that through the water sample, and
- Nephelometry which involves light directed into a water sample at right angles to the direction of response of photoelectric cells within the instrument and converted to a galvanometric reading.

Most colour in water arises from organic compounds leached from the ground and the decaying vegetation. Colour caused by suspended particles in solution is called *apparent* colour while colour derived from the organic matter is called *true* colour. Colour is measured by comparison with an arbitrary standard colour.

The stability of colloids in suspension is largely due to the surface properties of the colloids. The most important surface phenomenon is the surface charge of the particle which can result from :

- **chemical reactions on the surface of the particles :** a colloidal particle which has readily ionisable functional groups such as carbonyl and hydroxyl groups which would readily react with water and result in a charged surface.
- **lattice imperfections**, e.g. if in an array of solid SiO_2 tetrahedra, a Si atom is replaced by an Al atom (Al having one electron less than Si), a negatively charged surface would be formed. Clays are good examples of this type of substitution.
- **adsorption of ions** : Adsorption of a specific ion on a surface can arise from London-van der Waals forces and / or from hydrogen bonding.

The surface charge influences the distribution of nearby ions in solution. Oppositely-charged ions (counter-ions) to the surface charge are attracted towards the surface and ions of like charge (co-ions) are repelled from the surface. This, together with the thermal mixing of the solution and the mutual ionic repulsions and attractions, leads to the formation of an electric double layer around the particle. This electric double layer can be regarded as consisting of an inner region, characterised by a high concentration of counter-ions, which includes adsorbed counter-ions and water molecules and diffuse region in which ions (counter- and co- ions) are distributed according the influence of the electrical forces and random thermal motion.

The adsorption of a molecule from solution onto a surface involves firstly the removal of the molecule from solution and secondly, the removal of the solvent from the surface and the attachment of the molecule onto the surface. Due to Brownian motion and thermal agitation the colloidal particles are randomly being bombarded by molecules in solution. When such contact occurs the molecule could be :

- held to the surface by ionic, covalent, dipolar or hydrogen bonding,
- repelled from the surface by means of electrostatic forces or,
- even if conditions are favourable for adsorption, Brownian motion could move the molecule away from the surface.

When adsorption occurs in solution, it follows that there is a nett release of energy since the energy of attraction between the solid surface and the adsorbate must be greater than the energy of repulsion due to electrostatic effects. Furthermore, since adsorption involves the removal of a certain amount of solvent from the solid surface, then it follows that the energy required to remove the solvent molecules must be less than the energy released by the attachment of the adsorbate. The energy released during adsorption is termed the *overall standard free energy of adsorption* and is equal to the sum of the energy released as a result of chemical bonding, and either the energy released by electrostatic forces of attraction or the energy which has to be overcome because of electrostatic repulsion.

The coagulation of colloids and particles may be achieved with the addition of metal coagulants or polyelectrolytes. The two most commonly used metal coagulants are the aluminium-based coagulants and the iron-based coagulants. Common metal coagulants are shown in Table 1:

Table 1 : Common metal coagulants	
aluminium sulphate (alum)	$[Al_2(SO_4)_3 \cdot 14H_2O]$ or $Al_2(SO_4)_3 \cdot 18\ H_2O$
aluminium chloride	$[AlCl_3 \cdot 6H_2O]$
polyaluminium chloride	$[Al(OH)_{1.5}(SO_4)_{0.125}(Cl_{1.25})_n]$
sodium aluminate	$[NaAlO_2]$
ferric sulphate	$[Fe_2(SO_4)_3 \cdot 3H_20]$
ferric chloride	$[FeCl_3]$
ferrous sulphate	$[FeSO_4 \cdot 7H_2O]$

The aluminium and ferric salts form hydrated reaction products in solution. These metal ions normally exist as hydrated ions of the form :

$[Al(H_2O)_6]^{3+}$ and $[Fe(H_2O)_6]^{3+}$

In solution, a step-wise substitution of ligand molecules for water molecules occurs. The extent of substitution will depend on the concentration of the substituted ligand. For destabilisation of hydrophobic colloids with aluminium and iron coagulants, the ligands of interest are those of H_2O and OH^-, OH^- ligands arise from dissociation of bound H_2O ligands or from replacement of H_2O by OH^- ligands. The extent to which OH^- ions are bound to the metal complex is dependent on pH or concentration of the OH^- ions in solution.

In general there are four mechanisms of destabilisation observed when metal coagulants are used. These mechanisms are summarised in the Table 2 :

Table 2 : Influence of Indicated Parameter According to Mechanism (after Ramthol, 1993)				
Parameter	**Physical double layer**	**Adsorption Destabilisation**	**Bridging**	**Precipitation**
Electrostatic interactions	Predominant	Important	Subordinate	Subordinate
Chemical interactions and adsorption	Absent	Important	Predominant	May occur but not essential for removal
Zeta potential for optimum destabilisation	Near zero	Not necessarily zero	Usually not zero	Not necessarily zero
Addition of excess coagulant	No detrimental effect	Restabilisation usually accompanied by charge reversal	Restabilisation due to complete surface coverage	No detrimental effect
Relationship between optimum floc formation	Optimum dosage virtually independent of colloid conc.	Stoichiometry possible but does not always occur	Stoichiometry between dosage and particle conc.	Optimum dosage virtually independent of colloid conc.
Physical properties of floc produced	Dense, great shear strength but poor filterability in cake filtration	Flocs of widely varying shear strength and density	Flocs of 3-dimensional structure, low shear strength but good filtration	Flocs of widely varying shear strength and density

Metal (i.e. Al and Fe) complexes act as weak acids in water. A ligand from one metal ion may form a co-ordinative bond with another metal ion, thus the ligand acts as a bridge between the two metal ions. Thus the bridging of ligands of metal ions leads to metal ion complex formation. There is the ability to form multi-charged polynuclear complexes in solution with enhanced adsorption characteristics, metal complex coagulants usually work well in reducing the electric double layer effect.

Metal coagulants during the transition from free aquo metal ions to insoluble metal hydroxide precipitates undergo a series of hydrolytic reactions. The hydrolysis products of the metal coagulants in aqueous solution are adsorbed more readily than the free aquo metal ions. It has been noted that the greater the degree of hydrolysis the more extensive the adsorption. The reason for this enhanced adsorption is not understood completely. However it is evident that the presence of hydroxyl ions in the co-ordination sheath is largely responsible for the adsorption. Therefore the greater the degree of hydrolysis, the higher is the concentration of hydroxyl ions and thus the adsorption will be enhanced.

It has been postulated that the replacement of water molecules by hydroxyl ions in the co-ordination sphere of the metal ion, imparts a degree of hydrophobicity to the complex. Dispersions destabilised by hydrolysed metal coagulants have been observed to restabilise when the solution is made sufficiently acidic. Such destabilisation is a result of desorption of the metal ion species. This is due to the de-hydrolysation of the metal complexes, when the acid is added. The coagulant dosage affects the mechanism of destabilisation occurring. Where repression of the double layer by increasing the ionic strength is the predominant mechanism, the suspension passes from stability to destabilisation over a narrow range of coagulant concentration known as the critical coagulant concentration (CCC). The CCC is independent of particle concentration since destabilisation relies only on the concentration of the counter-ions in solution. Increasing the coagulant concentration beyond the CCC has no effect since once the ionic strength of the solution is sufficient to compress the diffuse part of the double layer destabilisation occurs spontaneously.

The pH of the colloidal suspension is critical since the predominance of certain hydrolysis species of the metal coagulant is largely dependent on the pH of the suspension. For a particular suspension, a particular metal hydrolysis species is required for effective destabilisation. To obtain an optimum concentration of this metal hydrolysis species a particular suspension pH needs to be maintained.

The term **polyelectrolyte** refers to a variety of water soluble macro molecules which have the ability to destabilise a colloidal suspension. These macro molecules are generally polymers with readily ionisable functional groups. In solution, these macro molecules ionise thus creating charged areas on the polymer which results in destabilisation. More than one colloidal particle can be adsorbed onto a macro molecule. Polyelectrolytes that possess both positive as well as negative charges are referred to as ampholytic polyelectrolytes. Those that do not possess ionisable functional groups are called non-ionic electrolytes. Polyelectrolytes range in molecular weights from 10^4 to 10^7 daltons. Activated silica, guar gums, tannins and synthetic polymers are common polyelectrolytes .

The charge density, molecular weight and the ionic strength of the solution are factors which affect flocculation. By altering the charge density, or the percent hydrolysis of polyelectrolytes, the configuration of the molecule in solution is changed. Increasing the charge density of the polymer, increases the viscosity of the polymer solution. Since viscosity increases with the length of the molecule, the implication is that with increasing charge density, the polyelectrolyte chains are increasingly stretched by the increasing electrostatic repulsions between charged units. At low charge densities, the polyelectrolyte chains are tightly coiled into spherical configurations. At medium charge densities, the polyelectrolyte chains are randomly linked or flexed coils while at high charge densities, the polyelectrolyte chains are fully extended filamentous rods. The longer the polyelectrolyte molecule the higher the molecular weight. Therefore the higher the molecular weight of the polyelectrolyte the higher the viscosity. Increasing the ionic strength of a polyelectrolyte solution decreases the range of influence of the charged sites on the polyelectrolyte chain and an increasing coil formation is achieved.

A fully-extended filamentous polyelectrolyte will expose a greater number of charged sites on the chain therefore it is most desirable to have a polyelectrolyte solution which has a high charge density, a high molecular weight and a low ionic strength for good flocculation to occur.

Two basic mechanisms have been proposed for the destabilisation of colloidal suspensions using polyelectrolytes. These are the bridging mechanism and the electrostatic patch model. The bridging mechanism is based on the ability of the polyelectrolyte to destabilise particles bearing the same charge. Furthermore the polyelectrolyte has the ability to form bridges between the particles. The first stage in bridging is the dispersion of the polyelectrolyte in the suspension. Polyelectrolytes have high molecular weights, they therefore exhibit high viscosities and low diffusion rates. It is essential that the polyelectrolyte be dispersed quickly and evenly in the suspension, because the rate of adsorption of the polyelectrolyte onto the particle is much faster than the rate of diffusion. Therefore quick vigorous rapid mixing is required. The second stage is the adsorption at the solid liquid interface. Adsorption depends on the chemical characteristics of the polyelectrolyte and the adsorbent surface. Adsorption could be due to cation exchange, electrostatic linkages, hydrogen bonding or ionic bonding. Adsorption is influenced by ionic strength which may reduce repulsion between similarly charged particle surfaces and polyelectrolytes, permitting adsorption or reduce the size of the polyelectrolyte coil, allowing more polyelectrolyte chains to be accommodated on the particle. Adsorption is effectively irreversible.

The third stage in bridging is the compression of the adsorbed chains. The occurrence and degree of bridging between two adjacent particles depends on the configuration of the polyelectrolyte on the particle. As progressively larger numbers of polyelectrolyte molecules adsorb onto the surface and consequently become compressed onto the particle surface the chance of bridging becomes smaller. Bridging between two particles occurs most effectively initially when there are a few "long looped" polyelectrolyte molecules adsorbed onto one particle. The other end of the "long loop" then has a greater chance of attaching itself onto another particle. Therefore for effective bridge formation it is important that extensive compression of adsorbed chains has not taken place. The strength of flocs depends on the number of bridges formed.

The bridging model describes destabilisation or non-ionic and anionic polyelectrolytes applied to negatively-charged suspensions. However, for charged polyelectrolytes applied to suspensions with particles carrying charges of opposite sign, the bridging model is inadequate. In this case the electrostatic patch model can be used to describe the destabilisation.

In the patch model the polyelectrolyte completely adsorbs itself onto the particle surface, forming regions of alternating positive and negative charges. Destabilisation occurs when these adsorbed particles align themselves to other adsorbed particles to provide electrostatic attraction. More particles are attracted to each other, forming agglomerates, which leads to the formation of flocs.

The dispersion of coagulants is dependent on rapid mixing at the time of dosing. The destabilisation reactions occur during rapid mixing and floc formation begins. The adsorption and destabilisation reactions occur very much faster than the diffusion rates. Therefore quick, very vigorous mixing is required to distribute the coagulant evenly. In order to appreciate the importance of rapid mixing one can observe the time intervals involved in the destabilisation reactions when compared to the diffusion reactions. Destabilisation reactions for polynuclear complexes requires 10^{-2} to 1 s and of the order 10^{-10} s for adsorption. The time needed for adjusting the structure of the double layer is of the order 10^{-8} s and for Brownian collisions to occur the order of time is 10^{-7} s to 10^{-3} s. Overbreek (1977) notes that the longest step in the destabilisation process (*rate limiting step*) is the time required for charge adjustment at the particle surface. This time could be as short as 10^{-6} s and as long as 10^{4} s. The above reaction times emphasises the importance of rapid mixing. Rapid mixing in excess of the optimum rapid mix times will cause the break up of the newly formed flocs, leading to less efficient flocculation in the clarifier.

Once destabilisation is sufficiently achieved, flocculation begins. Perikinetic flocculation arises from thermal agitation and it is a natural random process. It commences immediately after destabilisation and is completed within seconds, since there is a limiting floc size beyond which Brownian motion has no effect. That is, as particles coalesce, the magnitude of the energy barrier increases, so that eventually Brownian motion is insufficient to overcome this potential energy barrier. After perikinetic flocculation the only way to increase the collision rate between the particles is by inducing shear motion in the liquid. This induced shear motion is usually achieved by some sort of mechanical agitation. The above process is termed orthokinetic flocculation. The greater the velocity gradients induced in the liquid the more particle contacts there will be in a given time. However, the greater the velocity gradient, the smaller will be the ultimate floc size. This occurs because of floc break-up due to the high shear rates. Therefore for a given velocity gradient there will be a specific flocculation time beyond which floc particles will not grow further. If larger floc particles are required, a small velocity gradient with a large flocculation time is necessary.

The processes of sedimentation/clarification, filtration and disinfection follow flocculation in conventional water treatment processes. In this literature review the topics of mixing, coagulation, flocculation and sedimentation are considered.

2 Rapid Mixing

Rapid mixing is the process used to disperse coagulants into untreated water. Aspects such as rapid mixing time and speed have been identified as important factors which affect the subsequent coagulation and flocculation processes. Literature on coagulation processes indicates that extremely rapid dispersion is necessary for efficient use of coagulants (Amirtharajah and Mill, 1982; Hudson and Wolfner, 1967). While there is much debate regarding the appropriate mixing devices, the mixing energy applied and the duration of mixing, there is no doubt, however, that rapid dispersion of coagulants is advantageous in the rapid mixing phase of water treatment. The main reason for this is the speed of the destabilisation reaction.

The reaction of the coagulant with the colloidal particles (i.e. the dispersion of the coagulant, hydrolytic reaction of the metal cation, formation of colloidal hydroxo-metal polymers and the adsorption of polymer onto the colloid particles) occurs within one second if adequate coagulant is added at the correct pH and if sufficient mixing is provided (Kawamura, 1973). Amirtharajah (1981) found that destabilisation with alum occurs by adsorption destabilisation and by sweep floc destabilisation. The reactions for adsorption destabilisation were extremely fast, within micro seconds without the formation of Al hydrolysis polymers and within 1 second if polymers form. Sweep flocculation was found to occur between 1 and 7 seconds.

2.1. The Velocity Gradient

Bachman in Kawamura (1976) notes that mixing is required to

- provide an even distribution of chemicals in the medium, thereby producing a more uniform reaction
- provide energy to the Brownian diffusion to enable orthokinetic flocculation.

The degree of mixing has traditionally been quantified by the root mean square velocity gradient. The velocity gradient was first used by Von Smoluchowski (Polasek and Van Duuren, 1979). In order to account for the duration of the rapid mixing, Camp and Stein (1943, in Polasek and Van Duuren, 1979) combined the root mean square velocity gradient (G) and rapid mixing time (t) to give a dimensionless number called the Camp number. The two parameters which are generally used to quantify the energy associated with flocculation are the velocity gradient and the Camp number (Ca).

$$G = \left(\frac{e_d}{\mu}\right)^{\frac{1}{2}}$$

Where e_d is the dissipated power per unit volume, μ is the viscosity

$$Ca = Gt\theta$$

Where t is space time in a continuous flocculator or time in a batch reactor.

Both these parameters are derived from the overall energy dissipation. This concept has, however, been questioned by various authors (Han and Lawler, 1992, Polasek, 1979, Cleasby, 1984), and this will be discussed later in this review.

Polasek (1979) calculated the root mean square velocity gradient for a variety of devices used in water treatment for coagulation in his review. In gravitational systems the intensity of agitation depends only on the total head loss produced by the device. It is directly proportional to the rate of flow, and cannot be varied by the operator. He describes flash mixers, baffled mixers, orifice mixers, jet mixers, water jump mixers and baffled channels. Agitation by mechanical means is caused by the motion of a solid body, the impeller, submerged in the water. Revolving agitators and reciprocating agitators are considered.

The need for the development of small-scale equipment for the evaluation of full-scale plants has been highlighted, many authors have sought to quantify the G value occurring on the plant and apply it in a laboratory scale test such as the conventional jar test. Lai et al (1975) note that the mixing intensity is related to the rotational speed of the mixer, the configuration of the agitator and the geometry of the mixing vessel. They found that similar G values can be obtained for impellers with different shapes provided that the projected areas were the same. In this work they describe four different jar test configurations and monitor the mean velocity gradient, drag coefficient, Reynolds and Power numbers. Cornwell and Bishop (1983) also considered the importance of the Power number and state that the power number is a function of the mixer Reynolds number and the system geometry.

Francois and Van Haute (1984) investigated the effect of rapid mixing time on the flocculation process and highlight that a critical rapid mix time exists and mixing for a period longer than this critical time leads to a disturbance in floc growth during the flocculation period.

Various authors recommend G and Gt values for optimum mixing and subsequent flocculation. Fettig and co-workers (van der Merwe, 1990) suggest a G value of 700 s^{-1} at a Gt value of 2,1 x 10^5, for rapid mixing and 23 s^{-1} at a Gt value of 2,1 x 10^4 for orthokinetic flocculation. Morrow and Rausch (1974) suggested G values between 400 to 1000 s^{-1} for efficient flocculation and for metal coagulants it was found that G values from 1200 s^{-1} to 2500 s^{-1} were advantageous for cationic polyelectrolytes. Velocity gradients of 300 to 500 s^{-1} and detention times of 10 to 30 seconds have been recommended (Monk and Willis, 1987).

2.2. Types of Mixers

Rapid mixers may be radial turbines, weirs or hydraulic flumes and in-line static mixers. Conventional rapid mix systems are based on back mixing in a mixing chamber with detention times of seconds or even minutes, not fractions of seconds (Monk and Willis, 1987).

Amirtharajah (1981) undertook a comparative study on a flat blade, a propeller and a blender mixer to evaluate both mixing energy and mixing time. The Gt values ranged from 16 000 to 20 000. It was found that the high intensity rapid mixer (G = 16000) is significantly better than the other two backmix devices for adsorption destabilisation conditions. However when sweep flocculation is dominant, results indicated that there was no significant difference between the three types of mixers.

Studies by Kawamura, Griffith and Williams (in Kawamura, 1976) have indicated that the back mix flash mixer with a mechanical stirrer is inefficient and the flash mixing time should be shorter than the time recommended by the design guidelines. The low efficiency is due to short circuiting of flow and mass rotation of water instead of turbulence and eddy formation. Stators or baffles could improve the mixing conditions in such instances. Other rapid mixers include plug-flow devices with chemical injection nozzles where the mixing time is limited to a fraction of a second and in-line blender-type mixers.

Vrale and Jorden (1971) determined the mixing efficiency of mechanical mixers, backmix reactors and in-line blenders by monitoring the aggregation rate when using aluminium and iron as flocculants. The backmix reactor was found to be inefficient for rapid mixing, while the tubular reactor was the most efficient. They concluded that the average mixing velocity gradient is inadequate for characterising rapid mix units efficiency in terms of achieving the maximum aggregation rate for a given chemical dosage.

2.3. The Significance of the Velocity Gradient

The kinetics of flocculation of heterodisperse suspensions are usually described by the Smoluchowski equation which incorporates collision frequency functions by Brownian motion, fluid shear and differential sedimentation. These functions are based on a rectilinear approach of collisions. By this they mean that the short-range forces and changes in fluid motion as particles approach one another are ignored. Within this rectilinear approach the G value is the principal design parameter for flocculation units. Han and Lawler (1992), however, suggest a curvilinear approach, one which accounts for short-range effects in particle collisions. One of the main findings was that, if the curvilinear approach is adopted, the significance of G decreases. Many studies including Camp's have been undertaken where the assumption has been made that all particles are the same size (monodisperse), and short-range forces were ignored. In this study the effects of hydrodynamic interactions and van der Waals forces were accounted for in the analysis of collisions which resulted in a more realistic curvilinear approach. They concluded that the importance of G has been over emphasised in that they consider the primary importance of mixing to be keeping particles in suspension so that collisions can occur and that the velocity gradient (G) is secondary for the collisions of particles > 1μm. Letterman et al (Janssens and Beukens,1987) showed that the mean velocity gradient did not provide complete characterisation of mixing, although it is the most widely used parameter.

Elmaleh and Jabbouri (1991) criticise the approach of using the Camp number and the velocity gradient since both of these parameters are derived from the overall energy dissipation. They note, however, that only a small amount of the dissipated energy is required to promote interparticle contact and floc growth, while the remaining energy is required for the transport of the fluid. They therefore sought to quantify the flocculation energy requirement, as opposed to the total energy dissipated. For this purpose they used a flocculation device which was comprised of a PVC coiled circular pipe around a vertical cylinder, and monitored the head loss. They concluded that the number of units of energy was related to primary particle concentration and Reynolds number.

Cleasby (1984) also questioned whether the velocity gradient is a valid turbulent flocculation parameter. He found that G is only a valid parameter for the flocculation of particles smaller than the Kolmogoroff microscale of turbulence. He suggests that mean power input per unit mass to the two thirds power is a more appropriate flocculation parameter than G for common water treatment because the turbulent eddies are larger than the Kolmogoroff microscale.

Ives (Janssens and Beukens ,1987) when comparing various paddle geometries used in jar tests on the basis of mean velocity gradients that were imparted to water in one litre beakers concluded that the mean velocity gradient was not an adequate criterion to choose from among several paddle blade designs.

2.4. Models of Rapid Mixing

Amirtharajah and Trusler (1986) developed a theoretical model of rapid mixing using accepted models of particles collisions in turbulent fields. Amirtharajah and Trusler used the intensity of turbulence, the scale of turbulence and the Kolmogoroff microscale to quantify the degree of mixing. Their theory predicts that a minimum destabilisation rate occurs at specific ratios of microscale size to particle diameter d. It was also found from theory and experiments that particle destabilisation seem to be controlled by the maximum turbulence in the zone around the impeller of a back mix device. This is attributed to the presence of microscale eddies present in this region.

The kinetics of aggregate formation during rapid mixing was studied by Wiesner (1992) using fractal geometry. He notes that mixing conditions in most full scale plants are sufficiently large for breakup and aggregate restructuring to control the size distribution of materials leaving the rapid mix basin.

3 Coagulation and Flocculation

Coagulation refers to the chemical destabilisation of particles and the formation of small aggregates under the influence of Brownian motion. Flocculation is the subsequent process in which small aggregates form large flocs under the influence of hydrodynamic shear. Flocs form as result of the collision of particles which may occur as a result of Brownian motion (perikinetic flocculation) and by induced velocity gradients (orthokinetic flocculation). In each case the flocculation rate depends on the collision rate. In the case of perikinetic flocculation collisions occur entirely as a result of random Brownian motion of the particles and the collision rate can be calculated from standard diffusion theory.

3.1. Coagulation

Coagulation is affected by the intensity and duration of rapid mixing, the intensity of turbulence (i.e. velocity gradients), the coagulant concentration and dosage and the retention time in flocculator. The total time required for the chemical reaction of the coagulant to destabilise colloids (i.e. the dispersion of the coagulant, the hydrolytic reaction of the metal cation, the formation of colloidal hydroxo-metal polymers and the adsorption of the polymer to the colloidal particles) occurs within 1 second (Kawamura, 1973). Therefore the dispersion of the coagulant has to be achieved in a time of similar order to that of the overall time required to destabilise the colloids.

The diffuse part of the electrical double layer proposed by Gouy and Chapman is based on the assumption of point charges in solution. However, Stern recognised that the finite size of the ions will limit the boundary of the diffuse part of the double layer, since the centre of the ion can only approach the surface to within its hydrated radius without becoming specifically adsorbed. Thus Stern proposed a model in which the double layer has two parts, separated by a plane called the Stern plane. The Stern plane would be located at one hydrated ion radius away from the surface. Ions located beyond the Stern plane are part of the diffuse layer, therefore the Gouy Chapman model is used to describe this layer. The potential at the particle surface is given by the Nernst potential and that of the Stern plane is given by the Stern potential. When adsorption takes place arising from electrostatic forces, counter-ion adsorption generally predominates over co-ion adsorption. Verwey and Overbreek, and Derjaguin and Landau developed a model to describe the stability of hydrophobic colloids in solution, using the Gouy-Chapman model. This theory involves the estimation of energies of attraction and energies of repulsion in terms of distance between particles.

The point of addition of coagulant was identified by Moffett (Kawamura,1973) where it was indicated that 27 % more alum was require to achieve zero zeta potential when the alum was dosed onto the surface rather than at the stirrer blade.

Regarding the selection of coagulants, van der Merwe (1990) reported that high molecular weight substances (M.W. $> 10^4$) are readily removed by aluminium and ferric coagulants at optimum pH while low molecular weight particles (MW < 1000 - 1500) are not.

Colloid stability has been recognised as an important factor in the efficiency of coagulation. When two particles collide, contact may be prevented by inter-particle repulsion which can occur as a result of hydrophilic and hydrophobic conditions. Dissolved macromolecules such as proteins, starches and humic acids repel each other hydrophobically when they occur in contact with water molecules. Separation of these colloids can only occur by altering the solubility of the substance in water. This may be achieved by changing the temperature, adding large quantities of salts or small amounts of ions which form insoluble complexes with the macromolecules. Hydrophobic colloids are particles such as clays which are insoluble in water and are stable by virtue of an electric charge carried by the particles. When two such particles approach, electrical repulsion may be sufficient to prevent contact thus making the particles stable. Theories of the stability of hydrophobic colloids were developed independently by Deryagin and Landau in Gregory (1977) and Verwey and Ovebeek in Gregory (1977). This work is known collectively as the DLVO theory.

The DLVO theory regards colloid stability in terms of a balance between Van der Waals forces and electrical repulsion. Electrical repulsion arises from the charge carried by most colloidal particles and van der Waals attraction is based on fundamental intermolecular forces. When repulsion exceeds attraction at most separation distances the total energy curve produces a pronounced potential energy barrier which must be overcome before the colliding particles can contact. If the collision energy is insufficient to overcome the barrier, aggregates cannot form and the colloidal particles are stable. When the energy barrier height is exceeded, a certain portion of collisions are successful and slow flocculation occurs. If the barrier is removed altogether, there is no further repulsion between the particles and rapid flocculation takes place at a rate determined by the collision frequency.

Electrical repulsion can be reduced by increasing the ionic strength with any soluble salt (or indifferent electrolytes). The second way in which electrical repulsion can be reduced is by the addition of counter-ions which specifically adsorb on the particles and hence reduce the surface charge. This is known as adsorption flocculation.

Certain macromolecules such as polymers adsorb around the particles, preventing contact. Small amounts of polymer, however, can sensitise colloidal particles so that lower salt concentrations are required. This effect is recognised as *bridging* between particles by polymer chains. Bridging occurs when different segments of a polymer chain adsorb onto different particles, enabling the formation of aggregates.

There is often confusion regarding the selection of polyelectrolyte or metal coagulants. The type and size of turbidity-causing material determines the choice of coagulant due to the difference in ion exchange capacity of the colloid. Langlier et al. (in Kawamura, 1976) notes that particles of 1 to 5 μm in diameter provide a suitable structural unit for the development of a dense, rapidly settling floc. Particles of less than 1 μm provide a binder action for flocculant growth resulting in a porous, bulky floc with very slow settling velocity when alum is used as a coagulant. If both fractions of particle size exist in the water, a good floc will be provided. When raw water lacks either the colloidal or the coarse fraction it is recommended that the particle-size distribution be adjusted by adding a colloidal clay or other suitable material. Proper consideration of the particle-size distribution on a seasonal basis may affect the arrangement of process in the water treatment plant.

Coagulation by cationic polyelectrolytes can produce sludge of greater density, more rapid settling, improved dewatering capabilities and a lower volume. Because of this, and the increased ease of handling, cationic polyelectrolytes provide the advantage of lower overall treatment costs than metal coagulants (Morrow and Rausch, 1974). They concluded that cationic polyelectrolytes are able to completely replace inorganic coagulants as primary coagulants for both low and high turbidity surface waters when G values greater than 400 s^{-1} are applied. With the addition of excess polymer, however, restabilisation of particles can occur, either because of saturation adsorption, or by charge reversal (Gregory, 1977).

Clay, activated silica and organic polymers have been used as coagulant aids (Kawamura, 1976). The use of polyelectrolytes as flocculation aids has several advantages including a reduction in sludge volume, effectivity in small amounts, improvement in the sludge dewatering process, polyelectrolytes are more biodegradable than inorganic alum or iron salts, ease of handling, non-toxic in nature, and a minimal alkali dosage for final pH adjustment is required.

The type of floc formed is an important consideration for subsequent processes. Coagulation and flocculation prior to sedimentation requires different characteristics than coagulation and flocculation prior to dissolved air filtration. Klute et al. (1995) note that coagulation with Fe^{3+} alone resulted in the formation of flocs with good sedimentation not flotation characteristics and that dosing with cationic polyelectrolyte after coagulation with

Fe^{3-} resulted in improved separation. They also note that reaction time and energy input after polyelectrolyte dosing influences separation efficiency. Increased reaction time and energy input results in the formation of compact flocs with a smaller diameter, which offer more adhesion sites for air bubbles. In addition low turbidity, soft and highly coloured waters often produce very light flocs which settle slowly. In these cases, dissolved air flotation is seen as a more appropriate technology (Janssens, 1992)

Enhanced coagulation can be achieved with the addition of excess coagulant for improved removal of disinfection by-product precursors by conventional water treatment. This concept was developed in response to the Disinfectants/Disinfection By-products (DBP) Rule which stipulates the reduction of trihalomethane maximum contaminant level to 0,080 mg/l (Crozes et al, 1995). Natural organic matter (NOM) is perceived as the source of DBP precursors, and removal of NOM through coagulation has been widely documented in literature (Crozes et al, 1995). NOM is generally removed by charge neutralisation of colloidal NOM, by precipitation as humates or fulvates and co-precipitation by adsorption on the metal hydroxide (Randtke in Crozes et al, 1995). The degree of NOM removal through coagulation is affected by the type and dosage of coagulants as well as the pH. Synthetic organic cationic polymers can achieve colloidal NOM charge neutralisation and possibly participate in the precipitation of humic and fulvic acids. They do not, however, provide a substrate for the adsorption of the organic matter. Bench scale studies evaluating enhanced coagulation showed that inorganic coagulants were superior to the synthetic organic polymers for NOM coagulation (Crozes et al, 1995).

3.2. Flocculation

The aim of slow mixing is to transform the destabilised smaller particles into larger aggregates. The transport processes that govern the overall growth of aggregates is called flocculation (AWWA, 1991). Flocculation is due to collisions between particles which may occur by three mechanisms :

- Brownian or perikinetic flocculation due to the thermal energy of the fluid.
- velocity gradient, or othokinetic flocculation due to bulk fluid motion
- differential settling due to larger particles overtaking and colliding with a slower settling particle

According to Kawamura (1973) the most important factors that affect flocculation are: the velocity gradient, the coagulant concentrations and the location of the point of coagulant addition. Experimental results indicate that the optimum economical alum dosage should be selected on the basis of the most efficient coagulant concentration for turbidity reduction rather than a simple correlation between coagulant dosage and turbidity reduction. Kawamura suggests that for flocculation, a maximum energy input for most waters should not exceed a G value of 100 s^{-1} and an optimum G value of 50 s^{-1} is sufficient for most raw waters. The minimum detention time in a flocculator should not be less than 10 minutes.

In AWWA (1991) three modes of flocculation, occurring during different mixing flow patterns are described. **Random flocculation** is the most common floc-growth pattern which occurs in conventional flocculators. The process is dependant on the size of the primary particles introduced into the mixing stream. This form of flocculation results in a range of floc sizes and the density generally decreases with an increase in the size of the floc or the degree of agglomeration.

Contact flocculation is characterised by the action of flocculation in which larger, well-grown particles adsorb incoming minute flocs onto the surface. This mode of flocculation occurs in solids contact clarifiers and can be seen in conventional flocculators where backmixing occurs. Larger flocs of high concentration circulate through a turbulent flow field or are suspended in an upflow stream. The flow fields cause collisons and agglomeration to occur between the grown-flocs and the incoming microflocs. As a result of the difference in size between the two colliding flocs, an adsorption-like flocculaion process occurs. Due to the presence of grown-flocs with high concentration *C*, high *GCT* values result. These high *GCT* values enable the duration of the flocculation process to be shortened. Contact flocculation occurring in an upflow clarifier occurs when microflocs enter the floc blanket from the bottom and contact and agglomerate with the suspended grown-flocs. This flocculation process was formulated by Tambo and Hozumi (1979 in AWWA, 1991). It was shown that the process of contact flocculation in a floc blanket is characterised as a first-order reaction with respect to microfloc concentration. The rate constant is proportional to the product of the volumetric concentration of the suspended grown-floc, and the inverse of the floc diameter.

The third mode of flocculation is **pellet flocculation** which occurs in a conventional flocculation process. The charge-neutralised elementary particles collide randomly to form bulky agglomerates known as random floc. During the course of floc growth, the density of the floc decreases with the increase in floc size, since more void water is incorporated into the agglomerates. If the water could be excluded, particles of higher density and a larger size would be formed. Devices such as a horizontal-drum dewatering system, and a fluidised-bed pellet contact clarifier have been developed (Yusa and Gaudin, 1964 in AWWA, 1991). In the case of the clarifier, slow mixing blades in the floc blanket control the irregular growth of the pellet floc.

Andreu-Villegas and Letterman (1976) investigated the influence of alum concentration and the duration of flocculation on the optimum velocity gradient (G) values. The following results were obtained :

- the optimum G value decreases as the duration of the flocculation period increases,
- the residual turbidity at optimum G for each flocculation period decreases as the flocculation period increases, however the rate of decrease in turbidity diminishes after 20 minutes,
- the optimum G is apparently a result of the increasing influence of the erosion mechanism on the flocculation reaction at G values greater than 25 s^{-1}.

Gregory (1977) notes that an important feature of orthokinetic flocculation is that, for a given shear rate, flocs do not increase in size indefinitely, but a certain limiting floc size is found. Under turbulent conditions, the flocculation rate depends on the average shear rate, but the limiting floc size depends on the maximum shear encountered by the growing flocs. The strength of the aggregated floc was considered an important consideration for the design of separation processes in both water and wastewater treatment by Leentvaar and Rebhun (1983). Ferric hydroxide floc break-up mechanisms depend on the hydrodynamic conditions. In dilute agitated suspensions, floc break-up is governed by the interaction of the flocs with the fluid forces. A floc can be viewed as an aggregate of primary microparticles that are bound together to form a matrix incorporating fluid. The size and compactness of the matrix and the size and shape of the microparticles, the number and strength of the bonds at the microparticle contacts all contribute to the floc structure and the ability to withstand disruptive fluid forces.

Regarding the choice of flocculant, Treweek et al (1979) carried out flocculation tests on *E. coli* suspensions using a cationic polymer called polyethyleneimine (PEI)of various molecular weights (MW). . The following results were obtained :

- PEI polymers of molecular weights greater than 10000 achieved more than 90 percent reduction in number of single cells for doses of 0,5 mg/*ℓ* whereas PEI polymers of molecular weights less than 1000 recorded less than 70 percent reduction in single cell. Dosages in excess of 0,5 mg/*ℓ* caused restabilisation.
- The degree of flocculation with low molecular weight PEI polymers is sensitive to the duration (time) of the stirring during flocculation, whereas with higher molecular weight PEI polymers it was found that the degree of flocculation was unaffected by the duration of stirring.
- Flocculation was carried out at velocities gradients of 20 to 60 s^{-1}. Experimental results showed that there was a negligible difference in particle count as G increased from 20 to 60 s^{-1}.

Flocculators may involve uniform or tapered mixing. Kawamura (1976) notes that the best mixing pattern for flocculation is continuous mixing with tapered energy input. A problem is often encountered when mixing is not continuous between rapid mixing and the flocculation tanks. It is common to find plants where, after the coagulant has been added, there are unmixed distribution channels before the flocculation tanks. An improvement in turbidity and colour removal was noted by Kawamura (1976) when tapered mixing was implemented after rapid mixing. Continuous agitation can be done in distribution channels and pipes with the installation of water jets or compressed air.

Flocculation in laminar flow conditions has been studied. Gregory (1981) studied the extent of orthokinetic flocculation of suspensions as a result of laminar tube flow and found that flocculation depends primarily on the tube dimensions and not the flow rate. It was also found that coiling the tube provides increased mixing and as a result, a more uniform shear rate and enhanced flocculation.

3.3. *Modelling of Coagulation and Flocculation*

Janssens and Beukens (1987) developed an equation that describes flocculation in a clarifier in terms of a series of completely mixed flow reactors. The efficiency of each reactor was then be determined by measuring the inlet and outlet concentrations of each reactor. The above compartment model was tested and validated by Argaman and Kaufman (1970).

Von Smoluchowski (in Polasek and van Duuren, 1979) developed a model for perikinetic flocculation, where the frequency of collisions was obtained from the diffusional flux of particles towards a single stationary particle (Polasek and van Duuren, 1979). Diffusion towards the stationary particle was assumed to be in a radial direction. The number of particles diffusing radially inwards is proportional to the Brownian diffusion coefficient of the particles, the surface area of the sphere, and the radial particle concentration gradient. It was also assumed that the central particle also experiences diffusion. Von Smoluchwski also developed a model for orthokinetic flocculation under laminar flow (Polasek and van Duuren ,1979).

A diffusion model similar to Von Smoluchowski's was proposed by Argaman and Kaufman (1970). In this model it was hypothesised that particles in turbulent flow experience random motion similar to that of gas molecules. A bimodal floc size distribution comprising primary particles and large flocs was assumed.

Lee and Gregory (van der Merwe, 1990) found that for cationic polymer flocculation kinetics, the state of the charge density of the polymer is important in determining optimum dosage. The rates of adsorption and the rates of flocculation may, however be affected by the charge density and molecular mass of the polymer.

Tambo and Watanabe (1979) studied the properties of floc density using a clay suspension. A floc density function was mathematically derived and was modelled on computer. Experimental results were compared to the solution obtained by computer. It was found that the floc density decreases as floc size increases and that the aluminium ion concentration divided by the suspended particle concentration (i.e. ALT) ratio greatly affects clay-aluminium floc density. As the ALT ratio decreases, the floc density at the fixed size increases.

Tambo and Hozumi (1979) used a kaolin suspension and a special paddle flocculator to study floc strengths. An equation relating maximum floc diameter to effective energy dissipation was derived. They developed theoretical and experimental methods to evaluate maximum floc size during agitation and to provide a function of floc strength for the design of the flocculation and sedimentation process including the process of contact flocculation in clarifiers.

Amirtharajah and Mills (1982) developed the alum coagulation diagram as a predictive tool to define coagulant dosage and pH conditions where each of these mechanisms would predominate.

Wilson (1978) studied the kinetics of flocculation within the frame work of the Gouy-Chapman model. It was found that the rate of the achievement of flocculation equilibrium increased markedly with an increase in the ionic strength of the system.

Tambo and Watanabe (1979) developed a dimensionless equation for floc growth based on the rate of collisions between particles during flocculation. This dimensionless equation is a function of three dimensionless variables. The three dimensionless parameters are m (the dimensionless time), S (dimensionless maximum floc volume) and K_p (parameter describing variations in floc density). By evaluating these three dimensionless variables, the rate of flocculation can be determined.

3.4. Mixing in flocculation

The differences in design criteria for flocculators, both sedimentation and flotation systems were presented by Odegaard (1995). It was found that flocculators with G values of 20 to 40 s^{-1} were appropriate for flocculation/sedimentation systems, while values of 60 to 80 s^{-1} were found to be suited to flocculation/flotation systems. The reason for this was that small flocs, achieved by more intense mixing were considered preferable in a flotation system. It was also found that tapered flocculation units were more suitable for flocculation/sedimentation systems, than for flotation systems.

Tomi and Bagster (1978 in AWWA, 1991) showed that at the initial stages of flocculation with polymers aggregates grow rapidly in size, reaching a critical maximum size (d_{max}). This maximum size is inversely related to the energy input during mixing. For any condition of mixing there is an optimum concentration of polymer flocculant, which decreases with impeller speed. Greater speed facilitate increased compression of polymer which maximises particle bridging. At the critical size, rupture occurs followed by degradation (AWWA, 1991, Janssens, 1992).

Regarding the design of flocculators, it is considered that flocculators should be designed to allow for flexibility and to minimise short-circuiting. This flexibility should include alternative flocculant aid application points and the capability to vary the velocity gradient, tapered flocculation and an ability to vary the retention time (Monk and Willis, 1987). The retention time required for effective flocculation varies with the characteristics of the raw water.

4 Sedimentation and filtration

Sedimentation is the process whereby the floc formed in flocculation is allowed to settle by gravity. The efficiency of this separation process is related to loading rate, water quality, temperature, floc size, floc weight and tank currents. Conventional rectangular horizontal flow basins contain an inlet zone, the settling zone and the outlet zone (Monk and Willis, 1987).

In contrast to earlier developed horizontal flow sedimentation basins, upflow sedimentation basins are commonly used in water treatment (Hudson, 1981). Hudson (1981) lists 5 types of upflow sedimentation tanks including clarifiers; clarifier-flocculators; solids contact reactors; tube settlers and plate settlers. Solids-contact clarifiers provide the advantage of enhanced by the mass action of floc formation in the presence of previously formed masses of floc (Hudson, 1981).

Takacs et al (1991) presented a multi-layer model of the clarification process. This model was based on the solids flux concept and mass balances around each layer of a one-dimensional settler. The model was designed to predict a solids profile along a settling column, including effluent and underflow suspended solids. The model was tested on a pilot scale settler, with reasonable results.

Clarke et al (1978a) studied the operation of quiescent clarifiers operating under the hindered settling model to produce a mathematical model based on an equation for coalescing and disintegrating particles. A set of non-linear partial differential equations are obtained. These equations were solved numerically with the aid of a computer package. The dependence of settling characteristics on flocculation parameters were studied using the model.

Clarke et al (1978b) also studied the theory of operation of sludge blanket and up-flow reactor clarifiers. The operation of axially symmetric continuous up-flow clarifiers were modelled using continuity equations. The partial differential equations were again solved numerically using a computer.

Conventional water clarification equipment operates at surface loading in the range of 0,5 to 5,0 $m^3/m^2/h$ with retention times of 1 to 4 hours, while the High Rate (HR) clarifier is shown to operate at surface loading rates in excess of 15 $m^3/m^2/h$ with retention times below 20 minutes (Polasek, 1981). The HR Clarifier is a vertical flow type of clarifier designed to facilitate two stage flocculation, separation of a suspension in a sludge blanket maintained in a fully fluidised state by only the hydraulic means and being capable of sludge thickening. This rapid clarification process involves the initial orthokinetic flocculation, followed by post orthokinetic agglomeration with the addition of an organic flocculant aid.

Filtration through granular media is designed to remove the residual floc in the settled water, or in the case of direct filtration, the pin-point floc from coagulation (Monk and Willis, 1987). Filter performance is a function of the quantity of floc remaining in the settled water, the quality (size and strength) of the floc, the type, depth and size of the filter media and the hydraulic loading rate on the filter. The performance is measured by filtered water quality and by the length of filter runs between backwashes (Monk and Willis, 1987).

5 Monitoring and simulation

Proper operation of a flocculation process requires that the correct dosages of reagents are used with the required mixing and that the destabilised particles have sufficient opportunity to form aggregates of a specific size and having specific properties (Gregory, 1983). Monitoring and simulation must be designed to suit the application. If the aim of the process implemented is to clarify turbid water then the optimum conditions may be assessed by monitoring turbidity after the sedimentation of flocs. Where solids recovery is the objective, the settling rate may be the parameter to model. Ideally the flocculation test procedure should simulate the operation of the full scale plant and be capable of predicting plant performance based on laboratory trials. Often this is difficult to achieve and many water treatment plant personnel rely on achieving optimum conditions in the laboratory which correspond to those giving optimum plant performance. *The actual performance of the plant, in terms of clarification, settling rate and other parameters may not be the* *same as that found in the laboratory test under the same chemical conditions, but it should be possible to establish an empirical relationship which can be used for predictive purposes.*

TeKippe and Ham (1970) evaluated various methods including the conventional jar test, the modified jar test, the speed of floc formation, visual floc size comparisons, floc density, settled floc volume, floc volume concentration, residual coagulant concentration, silting index, filterability number, membrane refiltration, inverted gauze filtration, cation exchange capacity, surface area concentation, conductivity, zeta potential, streaming current detection, colloid titration, pilot column, filtration, filtration parameters, cotton-plug filtration and electronic particle counting to determine different properties of a flocculated suspension .

Other methods include turbidity, chlorine residual, bacterial counts in the finished water (Ramaley et al, 1981). A model for integral water treatment plant design based on particle size distributions was undertaken by Lawler et al in Ramaley et al (1981) to analyse the sensitivities of selected plant performance criteria to changes in design variables. This model can be used to examine and predict how particle sizes and concentrations are altered through flocculation and sedimentation.

5.1. The Jar Test

Many authors inluding Kawamura (1976), Dentel (in AWWA, 1989) and Gregory (1983) have identified the jar test as the most practical tool to determine the chemical application sequence. In the conventional jar test several samples of raw water is placed in a series of vessels, and is stirred in a manner similar to that on the plant. Various dosages of chemicals and coagulants are added. After an appropriate mixing period the stirring is stopped and the floc formed is allowed to settle for a specific time period. The clarity of the supernatant is then used to evaluate the optimum chemical dosages.

The jar test is a batch process experiment which is used to evaluate the various parameters of flocculation. Some of the parameters which are usually evaluated are:

- flocculant effectiveness,
- intensity of agitation during rapid mixing as well as during slow mixing,
- optimum pH,
- duration of the flocculation process,
- size and nature of flocs.

Regarding the conduct and use of jar tests Hudson and Wagner (1981) list the many used of the jar test other than for coagulant dose control :

- determining the strength of the floc
- predicting filtered water quality
- determining the optimum coagulant pH
- determining the optimum mixing intensity

- evaluating the optimum mixing intensity taper
- evaluating rapid mix duration and intensity
- evaluating the effect of lag time between rapid mix and flocculation
- evaluating optimum coagulant dosages
- evaluating the sequence of coagulant and coagulant aid dosages
- establishing the most effective coagulant
- evaluating the effects of sludge recycling and concentrations of sludge for recycling
- predicting the effects of short circuiting
- predicting sludge deposit configuration
- predicting design criteria for in-plant settling
- establishing design criteria for tube and tray settlers
- evaluating direct filtration possibilities and coagulant dosages

In order to calculate the velocity gradient, the power input into the liquid needs to be measured by connecting a torque meter to the stirrer and to the motor. Leentvar and Ywema (1980) found that the power number is dependent on the geometry of the system under study. An electrical method for the measurement of power input to an impeller mixer was devised by Mhaisalkar et al (1986).

Lee (1986) recognises that many factors affect the performance of a clarification process and stresses that efforts should be made to reproduce the plant conditions in the beaker. Often flash mixing intensity is neglected and Lee (1986) refers to equations for the calculation of G value for a weir and an electric mixer

Often the correlation between the jar test data and the plant is satisfactory, however, several problems have been identified (AWWA, 1989). These include:

- problems with dilution, during which the coagulant may not be sufficiently mixed
- proper dispersion must occur during addition to avoid adhesion to container walls
- there may be an interdependence between the primary coagulant and the coagulant aid.

The major disadvantages of the jar test are:

- that such a small scale mixing device cannot be assumed to be a true hydraulic model of the plant,
- the size of floc particles cannot be scaled down according to the principles of similitude,
- since the jar test is a batch process, short-circuiting does not occur

The conventional jar test, however, is ineffective for determining optimal coagulant dosages for waters of turbidities less than 1 NTU since settleable flocs do not form. Brink et al (1988) developed a bench-scale procedure for this application. The test is similar to the conventional jar jest except that a bench-scale filter is added to the process and the slow mixing and settling stages are omitted.

The jar test has been modified to include a filtration stage consisting of a small granular-bed filter. This method of separation does not simulate the sedimentation period of the conventional procedure, but requires less time to perform. Where there is correlation between the settled turbidity and the filtrate turbidity, this modified technique may provide equivalent results.

5.2. Laboratory-scale Flocculators

As a result of the porblems with the conventional jar test, varius authors have investigated alternative systems. Vigneswaran et al (1984) used a **spiral flocculator** to investigate the effects of velocity gradient, flocculation time and coagulant dose on direct filtration performance. The spiral flocculator consisted of a thin tube wound around a cylindrical core. The authors found that the velocity gradient in the tube does not depend on the length of the spiral tube, but on the velocity of the suspension passing through the tube. The flocs formed at specified flocculation conditions were found to be uniform in size, suggesting that uniform velocity gradients are achieved in the spiral flocculator. The spiral flocculator provides a means of obtaining and controlling uniform velocity gradients. Al-Hashimi and Ashjyan (1989) used a system of helical pipes for flocculation and compared this system to the jar test system. A similar system was used by Hameed et al (1995) for flocculation of river water and comparing the flocculation with that achieved in the conventional jar test. The increase in the degree of mixing achieved by the coiling of tubes was noted by Gregory (1981) in a study of flocculation in laminar tube flow.

Grohman et al (1981) devised a flocculation unit consisting of a tubular reactor for coagulation, followed by a cylindrical stirrer and a tubular reactor for flocculation. They found that a lower amount of energy is required for flocculation in these units than in a conventional flocculator.

Stahl and Kleine (1986) studied the physical properties of flocs using a continuous flow flocculation unit with a **couette flocculator**. The couette flocculator provides a uniform shear field (i.e. uniform velocity gradient). Studies have shown that when flocculation is carried out under high velocity gradients, the flocs formed become compact and more resistant to external forces, as verified by Stahl and Kleine (1986). Ives and Bhole (1977) used both the tapered and uniform gap Couette flocculators in a study of flocculation kinetics. They also compared this type of flocculator with the paddle flocculator used in the jar test.

5.3. Particle Characteristics

Particle counting involves the direct measurement of particulate material in water. Colloidal particles fall within the size range that particle counters are capable of measuring. Particle counters use the properties of light blockage and conductivity, to directly measure the number of particles in a given volume of water. The particle size distribution can then be calculated. Petrusevski et al (1995), in their investigation of coagulation for direct filtration found that, that in addition to turbidity, particle count is an essential parameter for the determination of optimal coagulation.

Azad and Kleine (1989) used an on-line **particle size analyser** to study the sizes and particle size distributions of flocs formed during flocculation using a commercially available automatic image analysis system.

Gregory (1983) proposed that the most direct method of monitoring a flocculation process can be achieved by particle counting using a **coulter counter.** Certain particle counters can analyse suspensions continuously. This is an advantage for on-line raw water analysis.

Gregory and Nelson (in Gregory and Chung ,1995) used a technique based on measurements of fluctuations of light transmission though flowing suspensions to determine the state of **aggregation** of particles. With improvements to the optical arrangement it is possible to obtain quantitative information on floc size and density. A flocculating suspension is sampled continuously and monitored by a narrow light beam. From the transmitted light intensity and fluctuations the average size and fractal dimension of the flocs can be obtained.

Floc density has been identified by Tambo and Watanabe (1979) as an important factor to monitor in order to design and operate the chemical coagulation process. If the density characteristics are illuminated, the settling velocity distribution of flocs can be derived from an output floc size distribution of a flocculator for the design of a sedimentation tank. The density and size of the flocs influence the amount and thickening characteristics of the sludge and the behaviour of filters. The floc settling velocity and floc size were measured and the floc density calculated. After using a modified version of the Stokes equation, they concluded that floc density decreases as floc size increases. They noticed a linear relationship between the floc diameter and the floc effective density (the buoyant density = floc density - water density). A floc density function was developed from this work.

Gregory (1983) refers to work done on sedimentation and settling where sedimentation of flocs is often measured in zone **settling** and reference is made to the importance of the settling characteristics of discrete flocs.

Tambo and Watanabe (in Gregory, 1983) developed methods for measuring the **settling** rates of individual aluminium hydroxide flocs and determining their size photographically. Measurement of supernatant turbidity if often used as an indicator of flocculation performance. The Silting Density Index (SDI) has been suggested as an alternative to turbidity measurements for low turbidities. This technique depends on the blockage of 0,45μm membrane filter and a consequent reduction in flow rate through the filter.

Bache et al (1995) describe how optimum coagulation conditions can be identified for both coloured waters and turbid suspensions in terms of the floc properties of size, effective density and strength. A general correspondence was found between the optimal coagulation conditions identified in the jar test with the optimal characteristics of the floc. Much of this work concentrated on the techniques for determining the floc-size density distributions and floc strength. Floc density was determined by the observation of individual flocs in a settling column. The settling column should be large enough to negate wall effects and the temperature should be controlled since temperature differences as small as 0.01°C can result in convection. Floc strength was determined in single grid oscillatory mixers.

5.4. Electrophoretic Measurements

Most suspended particles are charged and electrical repulsion prevents aggregation Gregory (1983). After destabilisation, a measurement of the particle charge as a function of the reagent dose can give an indication of the optimum dose. The **zeta potential** is defined as the potential at the surface of shear between the particle and the surrounding fluid (Hunter in Gregory, 1983). In this case the velocity of the particles is measured when an electrical field is applied. The particle velocity per unit field strength is known as the electrophoretic mobility, from which the zeta potential is calculated. Zeta potentials (ZP) of particles in water are commonly -50mV and complete destabilisation generally occurs when the zeta potential has been reduced to about +- 10mV.

Dentel (in AWWA, 1989) described a simple model for predicting the appropriate dosage of coagulant. The electrophoretic mobility $(EM)_{1,2}$ of the contaminant-coagulant complex is the weighted average of the mobilities of the contaminant and the coagulant :

$$(EM)_{1,2}\,(C_1 + C_2) = (EM)_1\,C_1 + (EM)_2\,C_2$$

Where the subscript 1 is for the contaminant and 2 is for the coagulant and C represents concentration. The value C_2 is determined by subtracting the theoretical soluble alum or iron concentration from the total that has been added. The $(EM)_1$ is measured for the raw water, $(EM)_2$ is measured for C_2 . These values are used to predict $(EM)_{1,2}$ at other coagulant dosages for pH values. Changes in the TOC or turbidity of the raw water may be translated into new values for C_1. This model is simplistic because it assumes that the EM of the coagulant is independent of time and mixing conditions. Dentel, however suggests that empirical means are still preferable in determination of optimum coagulant dose.

The absorption-destabilisation model for ferric ion flocculation indicates that coagulation will occur when repulsive electrical forces between particles are at a minimum. The ZP of particles can be measured in a particular system from the velocity at which these particles traverse a measured path in a DC current at a specific voltage (Bean et al in Johnson and Amirtharajah, 1983). This velocity is known as the electrophoretic mobility which is proportional to the particle charge or ZP. At zero charge, or at neutral conditions, maximum coagulation would occur. In practice however, maximum coagulation can occur at values less than or greater than zero electromotive force or ZP (Singley and Black in Johnson and Amirtharajah, 1983).

Swope (1977) deals with zeta potential in both polyelectrolyte and alum dosing schemes in some detail. It is noted that polyeletrolytes cause large changes in the zeta potential without affecting either the alkalinity or pH. Swope (1977) maintains that a jar test should be undertaken, varying coagulant dose and the zeta potential measured.

Electrophoresis refers to the movement of a charged particle suspended in fluid induced by an applied electric field. When a direct current (d.c.) is applied to a suspension of colloidal particles with a nett double layer charge, the charged particles will migrate to the pole of opposite charge. The counter-ions surrounding the particles will move in the reverse direction of the particles. The results from the electrophoresis measurements can then be converted to zeta potentials.

The technique used in electrophoresis measurements usually involves the visual observation of the movement of particles. The movement of the particles are timed over a specified distance for a given electric field. The equipment used for these measurements comprises of a stereoscopic microscope with an ocular micrometer, an illuminator and a variable power supply, a clear plastic electrophoresis cell with electrodes and a cell holder and timer. A disadvantage of this experimental technique is that it is a time consuming process.

Water-soluble polyelectrolytes of opposite charge to the suspension particles will adsorb strongly until the particle charge is neutralised. **Colloid titrations** can be undertaken by back titrating with excess polyelectrolyte to determine the amount adsorbed onto the particles. The original surface charge of the particles can then be calculated.

Coagulation diagrams have been developed by Johnson and Amirtharajah (1983) for Fe(III) and aluminium coagulants.

The **streaming current technique** involves the placing of a sample of a suspension in a special cylinder which has electrodes at the top and bottom. A loose fitting piston is fitted into the cylinder and submerged. The piston is then reciprocated along the cylinder's axis by a motor at a set speed. The movement of the liquid through the annulus (i.e. the space between the piston and the cylinder walls) creates an alternating current. This current is called the streaming current. This streaming current is used to control the dosage of coagulants in water treatment plants.

The Streaming Current Detector (SCD) consists of three main components:
- sample flow line
- streaming current sensor
- electrical components

The sample flow line is taken from the coagulated stream. The streaming current sensor consists of a reciprocating piston in a closed cylinder. This piston receives the sample flow and causes a portion of the flow to move back and forth in the gap, or annulus, between the piston and the cylinder due to the piston movement. Particulate, colloidal or otherwise absorbable materials attach temporarily to the piston and cylinder walls. The fluid flow relative to these surfaces causes sinusoidal movement of the counter-ions outside the shear planes, and this movement is equivalent to an electrical current. The magnitude of the current is proportional to the amount of net charge in the counter-ion layer, which is related to the electrical potential at the shear plane, known as the zeta potential. The electrical components of the SCD include the electrodes at the opposite ends of the sensor gap that pick up the streaming current through appropriate half-reactions, a filter, an amplifier and a rectifier. The output may be analog, digital, or graphical, or may be used as the input to a controlling device. A proportional integral differential (PID) controller is used to adjust the pumping rate of the coagulant feed (Dentel, 1995).

Dentel et al (1989a,b) evaluated the use of the Streaming Current Detector (SCD) for use in jar tests and continuous systems. Many researchers believe that charge neutralisation accompanies good coagulation in water (Kawamura et al, 1967 in Dentel et al, 1989). Traditional measurements such as electrophoretic mobility and colloid titrations are tedious to perform on a continuous basis, and could be superseded by the SCD for coagulant dose control. The SCD can provide meaningful information regarding the degree of destabilisation of a particulate suspension. The use of an SCD in conjunction with jar tests was found to be convenient and provided reproducible results. In continuous flow systems the SCD responded rapidly and reversibly to changes in the coagulant dose.

Dentel (1995) notes that the SCD has been accepted in the water treatment industry for on-line monitoring of coagulant dosage. In some cases, however the SCD does not correlate with the observed trends in particle destabilisation due to the effects of the attachment of non-representative materials to the internal sensor surfaces, the incomplete coverage of the sensor surfaces by the coagulated colloids and possible electrical artefacts in the electrical current from the electrodes of the SCD. Considering the above limitations Dentel concludes that SCD is feasible for controlling coagulant feed rates when the variations in water characteristics are moderate.

5.5. Turbulence Intensity

McConnachie (1991) assessed the fluid turbulence intensities and local velocities in relation to flocculation efficiency. Laser-Doppler anemometry was used for three types of stirrers finding that turbulence intensity is an alternative measure of flocculation efficiency to velocity gradient or power input.

5.6. Small-scale Tests and the Design of Water Treatment Plants

Barthelme (1994) modelled a continuous package lamella clarifier using batch jar test data. In this work the jar test procedure was used in the design and process description of the rapid mix and flocculation processes of water clarification and was adapted to describe the operation of a fill-scale continuous shallow depth sedimentation unit process. It was found from analysing the jar test flocculation kinetics that the floc break-up and floc aggregation rate constants are statistically correlated to the floc settling velocities. The constants provide a link between the batch jar test and the continuous sedimentation process. In this work it was also found that the jar test is useful in evaluating the effect of clarifier hydraulic characteristics such as short-circuiting and back mixing. One of the main reasons for the success of this work was that the settling depth in the jar test was similar to that in the lamella clarifier (due to the plate settlers) and therefore it was possible to apply the settling data directly to the continuous clarifier.

Bratby et al (1977) studied the validity of designing a continuous flocculation system based on data from batch experiments. The aggregation and break-up constants are obtained from jar tests. Normal jar test apparatus should *not* be used to determine the flocculation characteristics of a particular water since flocculation performance will be over predicted. They recommend that for the design of square flocculation basins, square batch reactors should be used.

Bratby (1981) studied the scope and limitations for the use of batch test results for the design of rapid mixing and flocculation systems. He notes several advantages of jar tests including cost and complexity when compared with pilot plant tests. Bratby (1981) refers to work done on correlations between batch test and continuous systems. He also stresses the importance of scale-up effects and refers to work by Argaman and Kaufman where they concluded that laboratory-scale batch experiments can be extrapolated successfully to full-scale design. He notes the batch tests are not appropriate for contact flocculation which occurs in solids contact blanket clarifiers as a continuous form of experimentation would be more appropriate.

A jar test was modified to model sedimentation basins using a hydraulic comparison equation sensitive to basin flow rates, water residence times and flocculant dosages by Reed and Reece (1984). In this work a hydraulic comparison calculation is used to model the jar test to the actual operating conditions and is verified by monitoring those conditions in the plant with the same coagulant and settling conditions. From the curves generated the operator could look up the appropriate settling time for the jar test by knowing the operating flow rate for that shift.

Reed and Robinson (1984) when developing the method to predict sedimentation basin performance noted that settling data suggested that a measurement of the floc size distribution might be a more reliable method for coagulant dose control. They note that the best settling characteristics occurred at the optimum dose, suggesting that floc size distribution may be a reliable method for dosage control.

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Water Research Commission Project No. 363

The use of small-scale equipment for evaluating water treatment plants

APPENDIX B

IMPULSE Manual

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Durban

November 1998

IMPULSE MANUAL

a computer program for
Residence Time Modelling

For use with IMPULSE beta version

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This manual is for use with **IMPULSE : a computer program for residence time modelling** (beta version). An introduction to the concept of residence time distribution and residence time modelling is given. **IMPULSE** structure and limitations are examined. A user manual then follows.

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LIST OF FIGURES

1 Introduction

This computer program was developed in response to a Water Research Commission-sponsored project entitled : *The use of small-scale equipment for evaluating water treatment plants.* Previously, the residence time distribution analysis of tracer tests has not been used to its full potential due to mathematical complexity. This research aimed to develop an accessible method to analyse tracer tests to determine the flow model and the parameters of the process.

1.1 Introduction to the residence time distribution concept

To optimise continuous flow processes it is necessary to understand and assess the physical process. Possible improvements for more efficient operation and process intensification must be postulated and the effect of improvements must be predicted, providing the basis for decision-making in terms of the time, effort and cost of making the improvements.

In systems where a flow model does not have significant meaning alone, a knowledge of the flow patterns in the system can be used as a diagnostic tool analyse the cause of process failure, as condition changes will be indicated by change in flow patterns. Further, the flow patterns will indicate the length of time required to reach steady state after a change in conditions in the system.

After determining the flow model of a system and the flow parameters, the overall kinetic rate equation of the reaction occurring in the system can be combined with a mass balance performed over the flow model. This yield the outlet concentration with respect to the flow parameters of the flow model. Thus the effect on the outlet concentration of a change in the flow parameters can be quantified.

To understand and assess a system it is necessary to model the system. For the modelling of a system a knowledge of the flow patterns in the system is important (Rabbitts, 1982). Danckwerts (1953) introduced residence time distribution methods to enable a quantitative description of the flow patterns in a system. Previously, it was usual to assume either *perfect mixing* or *plug flow* (Figure 1.1).

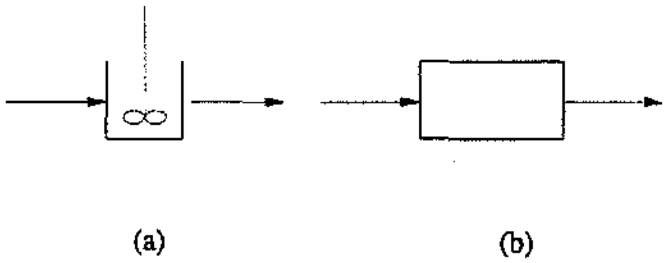

Figure 1.1 : *Description of ideal systems :*
(a) perfect mixing
(b) plug flow

Ideal mixing implies the fluid in the system is completely mixed, so that the properties of the fluid in the system and in the effluent stream are the same ; **plug flow** implies no mixing in the direction of flow, so that elements of fluid entering the system at a particular time flow through the vessel together and leave at the same time in the future. The flow patterns found in real processes usually lie between these two extremes (Smith, 1981), due to *bypassing*, *channelling*, *dead space* and *recycling* (Figure 1.2).

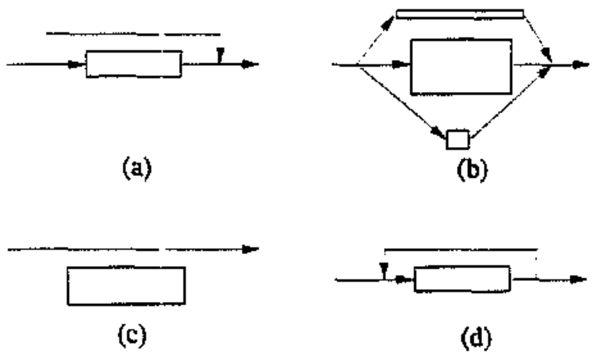

Figure 1.2 : *Deviations from ideal systems :*
(a) bypassing ; *(b) channelling* ;
(c) dead space ; *(d) recycling*

In *bypassing* and *channelling*, some elements of fluid move through the system significantly faster than others do. *Dead space* refers to a region in the system with extremely poor contacting with the bulk fluid ; fluid in the dead space will generally remain in the system significantly longer than the bulk fluid. *Recycling* occurs when fluid is recirculated to the system inlet or to another region of the system.

Thus, the effluent stream of a continuous flow system is a mixture of fluid elements that have resided in the system for different lengths of time ; the distribution of these residence times is an indicator of flow patterns within a system.

Introducing tracer into the inlet stream of a system and measuring the concentration-time relationship of the tracer in the effluent stream provides an observation of the distribution of residence times for the tracer. If the tracer particles have the same flow attributes as the fluid, their residence time distribution can be said to approximate the residence time distribution of the fluid (Naumann and Buffham, 1983).

The flow pattern of the system is determined by establishing a flow model of the system from the tracer residence time distribution curve.

1.2 Flow model determination

Flow models are constructed from combinations of ideal mixed and plug flow **blocks**, joined by recycles and bypasses. A flow model of this type, and its parameters, is guessed from the tracer residence time distribution curve. The model residence time distribution curve is determined from known equations associated with the ideal blocks. This curve is compared to the tracer curve and the parameters of the model adjusted until the closest fit is obtained.

In the past, this procedure was not easily accomplished due to mathematical complexity associated with finding the closest fit between the tracer curve and the model curve. Therefore, finding the best flow model and parameters of the model was a time consuming and inaccurate procedure, particularly if different models were attempted.

1.3 Computer programs for determining flow models

A procedure to model systems from the tracer residence time distribution curve using a Turbo Pascal computer program was reported by Barnett et al. (1992). This program found the closest fit between a model guessed from the tracer residence time distribution curve and the tracer curve itself to determine the flow parameters of the system. However, the program was not interactive nor user-friendly.

A user-friendly, interactive computer program, **IMPULSE**, was written by Baddock (1992). It allows easy modelling of systems using curves obtained from tracer response tests. The user assumes a flow model for the system ; the program determines the theoretical response curve for the model, and optimises a chosen set of parameters of the model to fit the experimental curve.

The program is currently in its beta, or testing, phase. It has been designed for conservative tracers, that is, tracers that are not consumed in the system. It is suitable for a wide range of applications of continuous flow processes.

These include :

- continuous flow anaerobic digesters
- clarifiers
- mixing systems
- reaction vessels
- pipelines

2Program structure

2.1 Introduction

The program was written in Turbo C++ and is executable on all IBM and IBM compatible personal computers. with Hercules, EGA or VGA graphics cards.

IMPULSE was developed with a menu system, consisting of a main menu and submenus, which allow the user to select a desired program option from a list of possible options displayed on the screen. A number of user-friendly features have been included in the program, for example :

- on certain options the program guards against meaningless data input by evaluating data entered into the program as the keys are pressed
- the user can choose whether to display the model curve with guessed parameters before curve fitting, to reduce the time spent by the program fitting the curve
- automatic updates of the screen curves are done as data is entered into the program

Input to the program is via the ADD and EDIT options on the main menu. Also on the main menu is a FILE option, allowing file handling ; a PLOT option, to add or remove data sets from the plot displayed on the screen ; a CONSTANTS option, setting the mathematical constants for the curve fitting ; and a RUN option, which starts calculations and curve fitting. The program exits to DOS by pressing the *Escape* key.

A complete description of the menu system and the running of the program is presented in Chapter 4.

2.2 Data input

Data is entered into the program via the ADD and EDIT options on the main menu (see Chapter 4). The following data may be entered into the program :

- the building blocks of the model (mixed flow, plug flow, mixer/splitter, input, output)
- the parameters associated with the building blocks (volume, split fraction)
- whether the parameters must be held constant or can be varied by **IMPULSE**
- an tracer input curve as a flowrate and concentration history
- a tracer response curve (called the REFERENCE curve) against which the model curve can be fitted

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* C++ is a registered trademark of Borland International Inc.

** IBM is a registered trademark of International Business Machines Corporation

2.3 IMPULSE flow diagram

As **IMPULSE** is interactive, a flowsheet cannot show all the options and the paths between the options. A flowsheet showing possible user interaction with the program is given in Figure 2.1.

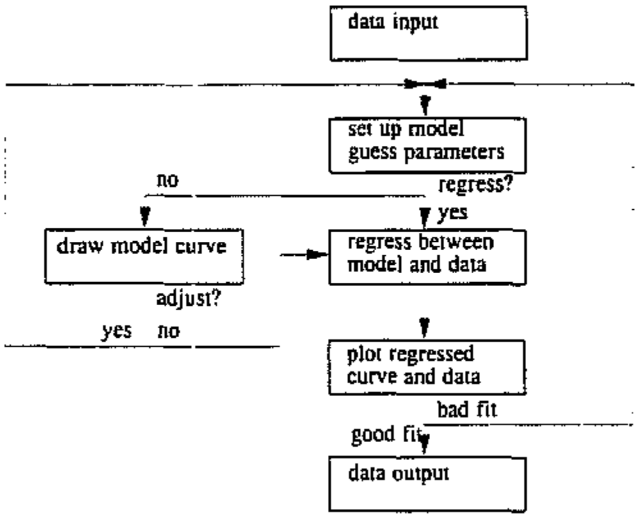


Figure 2.1 : *Typical suggested flowsheet for using IMPULSE to investigate a residence time problem for residence time modelling*

2.4 Output from the program

The output from the program after modelling is displayed on the screen and can be saved to a file, the filename of which is entered by the user. The following is saved in the file :

- modelling units and associated parameter
- unit connections
- any curve that is currently in the set that is to be plotted (Section 4.2.7).

3 Program limitations

Certain limitations exist for the use of **IMPULSE**. These are listed below :

- **IMPULSE** should only be used to model continuous or near-continuous flow systems.
- The program models the liquid residence time of the system, and not the solid residence time.
- The experimental determination of the tracer response curve is not trivial and may influence the result. Thus a user must be aware of the techniques for tracer tests, and be able to estimate experimental error, and how this influences the curve.
- A user should have a physical knowledge of the system to be modelled and should be able to guess feasible flow models. Any parameters obtained from the flow model must have physical significance ; thus a complex model may fit the tracer curve but may not be feasible. For open systems, a dye test may indicate the most feasible flow model.
- Tracer curves are not unique : they can be modelled accurately by more than one model. Sound engineering judgement is therefore necessary to choose the most feasible model.
- **IMPULSE** works by dividing up time into **'segments'**, and then applying the equations describing the system to each segment in turn. Any model constructed is thus sensitive to the number of segments. The larger the number of segments, where $n_{segments} = \frac{StopTime - StartTime}{deltaT}$, the more accurate the results, but the more time required to reach a result (Section 4.2.8).

4User manual

4.1 Installing the program

4.1.1 System requirements

IMPULSE is executable on all IBM PC, XT, AT or true IBM compatibles. It requires DOS 3.00 or higher and at least 512K of RAM to run. **IMPULSE** includes routines that allows the program to make use of an 8087, 80287 or 80387 numeric co-processor if available. It can only be used on a system with Hercules, EGA or VGA compatible graphic adapters.

It is recommended that **IMPULSE** be run on a 386 or, preferably, a 486 computer with a numeric co-processor and a colour monitor for optimum performance.

4.1.2 Files on the distribution disk

The distribution disk contains the following files :

- IMPULSE.EXE : the executable program file
- EX?.REF : the reference time-concentration data for the examples described in Section 4.3
- EX?.FLO : The time-flowrate data for the examples described in section 4.3
- EX?.CON : the time-concentration data for the examples described in Section 4.3

4.1.3 Installing and running IMPULSE

The most convenient method for running **IMPULSE** is on a hard disk system. The procedure outlined below can be used to set up **IMPULSE** on a hard disk :

Assuming that the hard disk is designated as drive C, make a subdirectory to work in, and change to that subdirectory by typing the following commands (pressing *Enter* at the end of each line) :

- C:
- CD\
- MD IMPULSE
- CD IMPULSE

Place the distribution disk in the disk drive (presumably drive A)

Copy the files on the distribution disk by typing the following command and pressing *Enter* :

- COPY A:*.* C:

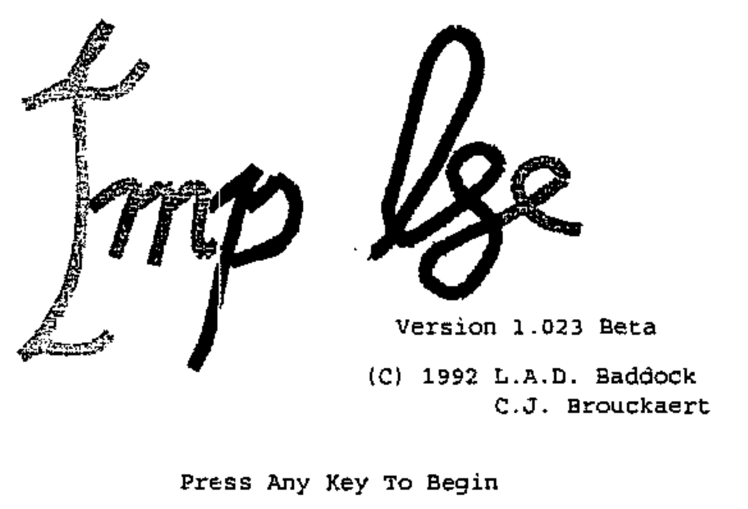
To run the program from anywhere on the hard disk, type the following commands (pressing *Enter* at the end of each line) :

- C:
- CD \IMPULSE
- IMPULSE

4.2 Using the program

4.2.1 Starting Program Execution

Once the program has been run the title page will gradually be displayed, as shown in Figure 4.1. Press any key to advance execution as indicated on the screen.

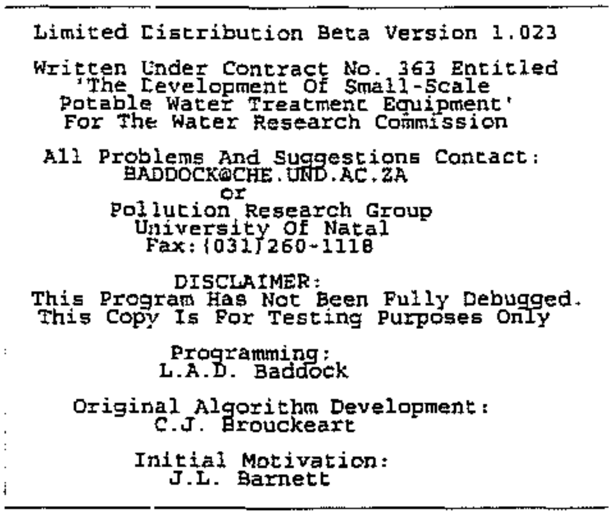


```
Version 1.023 Beta
(C) 1992 L.A.D. Baddock
         C.J. Brouckaert

Press Any Key To Begin
```

Figure 4.1 : *Title page of the IMPULSE program*

Information about the program will then be displayed, as shown in Figure 4.2. Press any key to advance execution as indicated on the screen.



```
        Limited Cistribution Beta Version 1.023

        Written Under Contract No. 363 Entitled
          'The Cevelopment Of Small-Scale
         Potable Water Treatment Equipment'
          For The Water Research Commission

         All Problems And Suggestions Contact:
               BADDOCK@CHE.UND.AC.ZA
                         or
               Pollution Research Group
                University Of Natal
                 Fax:(031)260-1118

                    DISCLAIMER:
       This Program Has Not Been Fully Debugged.
        This Copy Is For Testing Purposes Only

                    Programming:
                   L.A.D. Baddock

           Original Algorithm Development:
                   C.J. Brouckeart

                 Initial Motivation:
                    J.L. Barnett
```

Figure 4.2 : *Second title page of the IMPULSE program : about the program*

4.2.2 The main menu

After continuation from the title page, the main menu of the program appears (Figure 4.1)

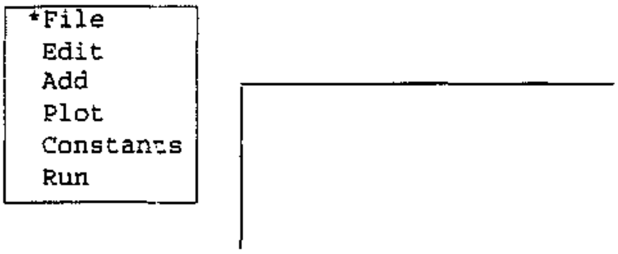


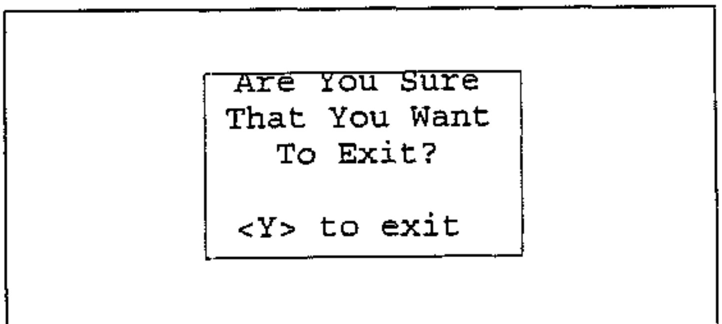
Figure 4.3 : *The main menu*

When the menu first appears, the cursor (shown as an asterisk) is positioned next to the FILE option. A menu option (or submenu option) can be selected by using the *Up* and *Down* arrow keys on the keyboard to move the cursor through the list of options, and pressing *Enter* when the cursor is positioned next to the desired option.

Pressing *Escape* when the cursor is in any of the submenus return the cursor to the main menu and erases the submenu from the screen.

4.2.3 To leave IMPULSE

When the cursor is in the main menu, pressing *Escape* prompts a confirmation query (Figure 4.4). Pressing *'y'* or *'Y'* exits to DOS.



Are You Sure That You Want To Exit?

<Y> to exit

Figure 4.4 : *The exit screen*

4.2.4 The FILE option

By selecting the FILE option, a submenu appears on the screen as shown in Figure 4.5.

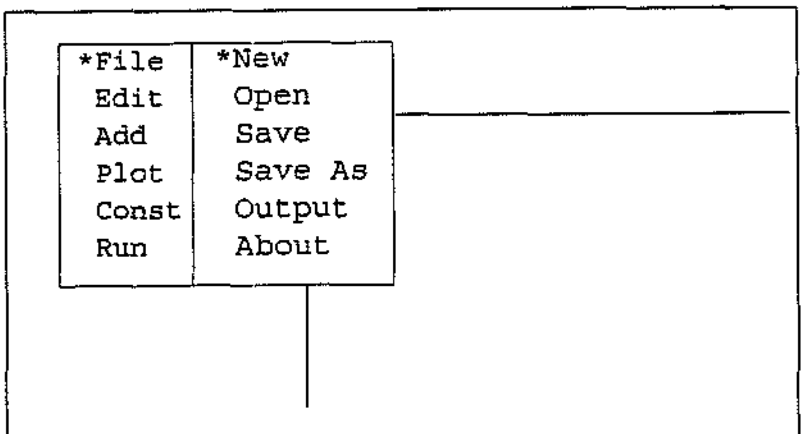


Figure 4.5 : *The FILE submenu*

Moving the cursor to an option and pressing *Enter* selects that option.

- **New** clears the screen and starts a blank file.
- **Open** opens a file specified. The user is prompted for a filename, as shown in Figure 4.6. A filename can either be entered using the keyboard, and the asterisk deleted using the *Delete* or *Backspace key*, or a list of possible files can be called by pressing *Enter*. Positioning the cursor next to one of the filenames and pressing *Enter* loads the file. The default extension is *rtd* : if the filename does not have this extension, the default can be deleted using the *Delete* or *Backspace* key, and replaced with an asterisk, or the actual extension can be entered from the keyboard.

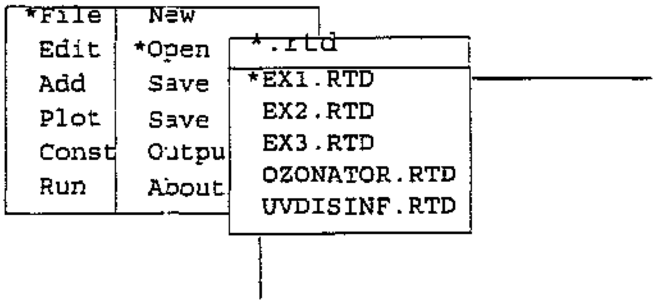


Figure 4.6 : *OPEN on the FILE submenu*

- **Save** allows the user to save a new file under a name of their choice, or saves an old file under its name. It is recommended that the extension *.rtd* be added to any filename.
  If changes have been made to an existing file, and **Save** is chosen, the old settings will be overwritten with the new ones.
- **Save As** allows a user to save a new or old file under a filename of their choice. This allows changes to be made to a file and both new and old versions to be retained with different names. It is recommended that the extension *.rtd* be added to any filename. A note of filenames should be kept by the user, as the program will overwrite any other file given the same filename and extension.
- **Output** allows the user to save all the data associated with the file, including the parameters of the model, the input history, the reference curve and the model curve. The file is saved in a format that can be imported into any spreadsheet program. It is recommended that the file be given the extension *.prn*, as this is the default extension for most spreadsheet programs.
- **About** gives information about the program, as shown in Figure 4.2.

4.2.5 The EDIT option

By selecting the EDIT option, a submenu appears on the screen as shown in Figure 4.7.

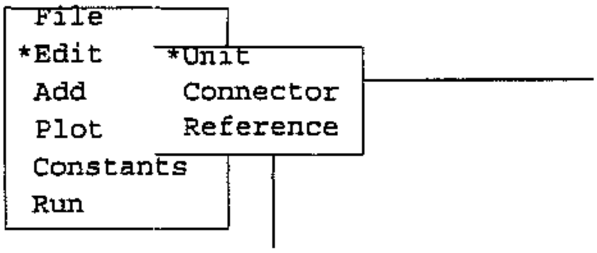


Figure 4.7 : *The EDIT submenu*

Moving the cursor to an option and pressing *Enter* selects that option.

- **Unit** allows the user to add edit the parameters for any unit on an existing model.
- **Connector** allows the user to edit any connector linking objects in an existing model, and to connect objects in a different order.
- **Reference** is where a reference curve (that is, the tracer response curve) is linked to the program. This is the curve against which the program fits the model curve for the model chosen. It is an editable curve, as shown in Figure 4.8. A scaling factor can be added to the curve to compensate for units. Data can be entered from the keyboard as time and concentration, with each data set on a separate line and time and concentration separated by a space.

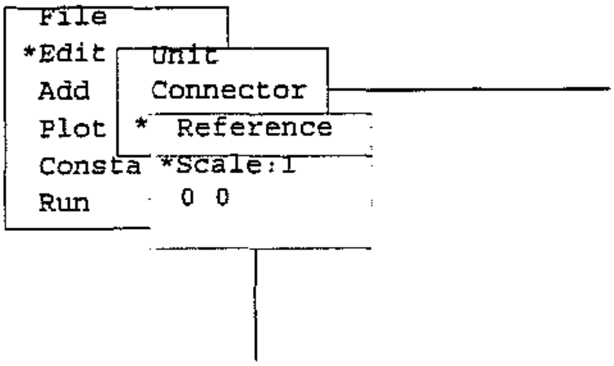


Figure 4.8 : *REFERENCE on the EDIT submenu*

Data can also be imported from a file made in a spreadsheet program. The file should contain only time and concentration, with each data pair on a separate line. The elements in each data set should be separated by spaces. To start importing processes, *F1* is pressed. The filename is then entered in the space provided (Figure 4.9).

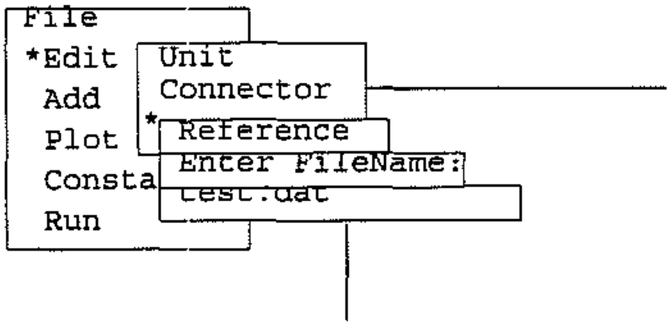


Figure 4.9 : *Showing a list of possible files for the REFERENCE curve*

4.2.6 The ADD option

By selecting the ADD option, a submenu appears on the screen as shown in Figure 4.10.

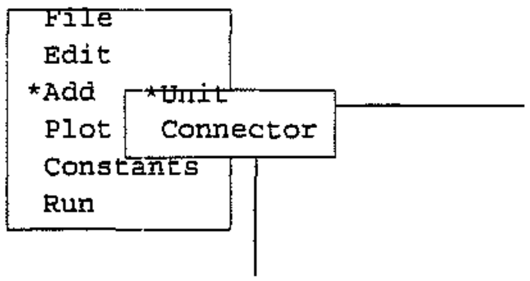


Figure 4.10 : *The ADD submenu*

Moving the cursor to an option and pressing *Enter* selects that option.

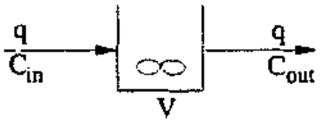
- **Unit** allows the user to add a unit to the model. The user is asked to name the unit, guess parameters for the unit and decide whether the parameters must be held constant or whether they can be varied during curve fitting. It is required that the user give a different name to each unit and that they keep a record of the units and where they fit into the model the user wishes to construct. The units are : **input, mixed flow (mfr), plug flow (pfr), mixer/splitter (ms)** and **output.**

  input : an input unit is used to model a stream entering the process being modelled. Its parameter set consists of a flowrate history, a concentration history and associated scaling factor.

  The user must assign a name to each input. The input is an editable curve. A scaling factor can be added to the curve. Data can be entered from the keyboard as time and concentration, with each data set on a separate line and time and concentration separated by a space.

  Data can also be imported from a file made in a spreadsheet program. The file should contain only time and concentration, or time and flowrate, with each data set on a separate line, with spaces between data elements in each. This filename is typed into the space given for a name, after *F1* has been pressed.

  mixed flow (mfr) : the mixed flow (mfr) unit is a perfectly mixed vessel, as defined in Chapter 1, with an equation associated with it (Figure 4.11).



$$qC_{in} - qC_{out} = V\frac{dC_{out}}{dt}$$

Figure 4.11 : *IMPULSE modelling block for mixed flow (mfr)*

where **q** is the volumetric flow through the mixed flow and **V** is the volume of the mixed flow.

The user must assign a unique name to each mfr, and guess a volume. The user must decide whether the volume should be held constant by **IMPULSE**, or whether it can be varied, as shown in Figure 4.12.

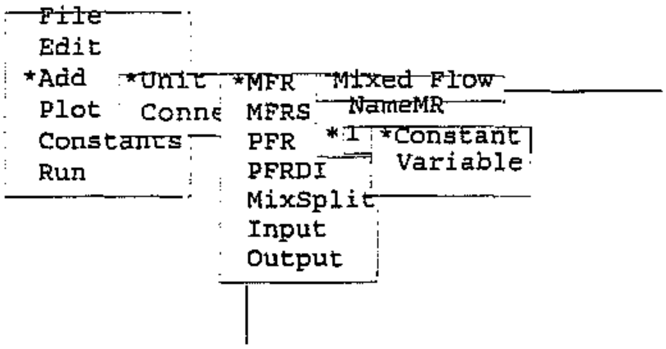
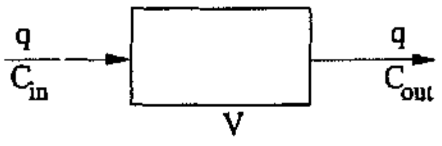


Figure 4.12 : *Screen display for mfr*

plug flow (pfr) : the plug flow (pfr) unit is a perfectly mixed vessel, as defined in Chapter 1, with an equation associated with it (Figure 4.13).



$$C_{out}(t)=C_{in}(t-\frac{V}{q})$$

Figure 4.13 : *IMPULSE modelling block for plug flow (pfr)*

where **q** is the volumetric flow and **V** is the dead volume.

The user must assign a name to each pfr, and guess a dead volume. The user must decide whether the volume should be held constant by **IMPULSE**, or whether it can be varied, as shown in Figure 4.14.

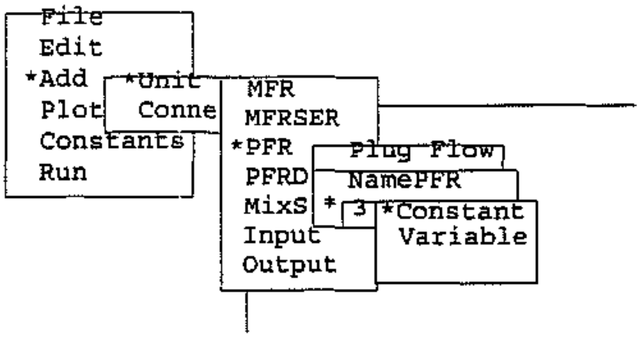


Figure 4.14 : *Screen display for pfr*

mixer/splitter (ms) : a mixer/splitter is added to the model whenever a stream needs to be joined or split (Figure 4.15).

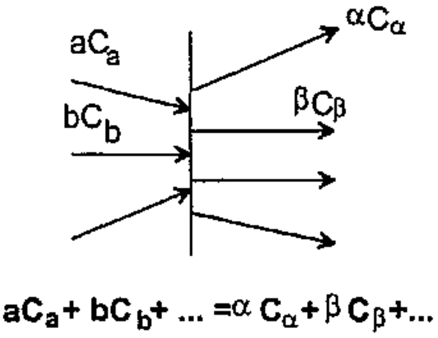


Figure 4.15 : *IMPULSE modelling block for mixer/splitter (ms)*

where **a** , **b** , etc. are volumetric flows.

The user must assign a name and split fraction to each ms.

output : an output records the history of any flowrate and concentration. A name needs to be associated with this output for plotting purposes.

- **Connector** : allows the user to connect the units together in any way. A list of the units to be connected *FROM* is shown and the cursor can be moved to the desired unit using the arrow keys (Figure 4.16). A list of the units to be connected *TO* is shown and the cursor can be used to select the desired unit.

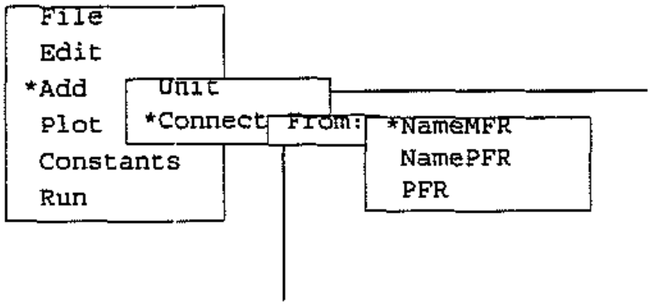


Figure 4.16 : *A list of possible units for connections*

Only the units possible are shown so, for example, the input will only be shown on the *FROM* submenu, as it has only one connection. An unit will be erased from the lists of possible options once the unit's maximum connection requirements are met. A mixer/splitter has no connection limitations, and will remain on both list. An output will only appear on the *TO* submenu as it has only one possible connection.

4.2.7 The PLOT option

By selecting the PLOT option, a submenu appears on the screen as shown in Figure 4.17.

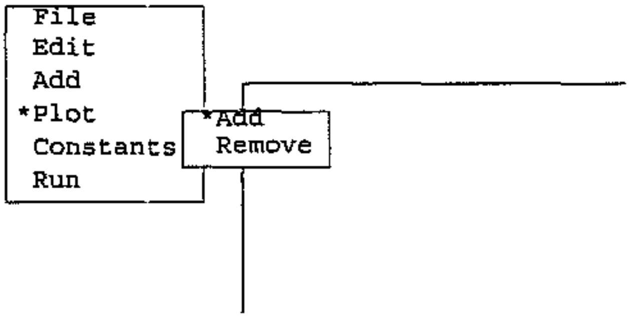


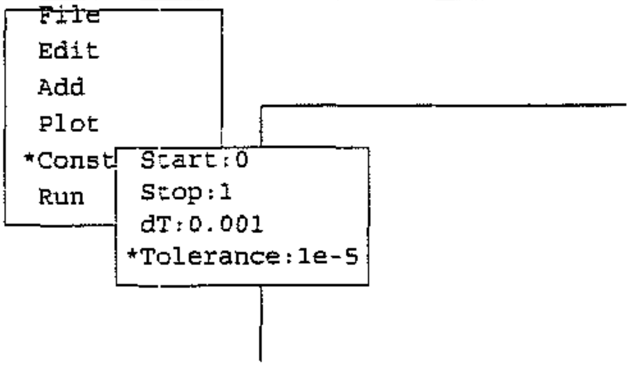
Figure 4.17 : *The PLOT submenu*

Moving the cursor to an option and pressing *Enter* selects that option. The PLOT option allows the user to determine what data sets are plotted on the screen. This is useful if there are many data sets, or if the data sets have very different scales, so that it is difficult to see any one set.

- **Add** allows the user to add a data set to the plot that appears on the screen. A list of possible data sets will appear, and the cursor can be used to select a set.
- **Remove** allows the user to remove a data set from the plot that appears on the screen. A list of possible data sets will appear, and the cursor can be used to select a set.

4.2.8 *The CONSTANTS option*

By selecting the CONSTANTS option, a submenu appears on the screen as shown in Figure 4.18.



```
Start:0
Stop:1
dT:0.001
*Tolerance:1e-5
```

Figure 4.18 : *The CONSTANTS submenu*

Moving the cursor to an option and pressing *Enter* selects that option. The CONSTANTS option allows the user to determine the constants in the curve fitting.

- **Start** determines the relative time at which the curve fitting begins.
- **Stop** determines the relative time at which the curve fitting ends.
- **dT** determines the time interval over which the program iterates. Reasonably, this should be less than the minimum interval between the time values of any adjacent pair of data points on any date curve. With in these confines, this value should initially be chosen as large as possible to reduce computing time. Once the model curve fits the reference curve moderately, this can be reduced to give a more accurate result.
- **Tolerance** determines the maximum allowable error for flowrate convergence where there are non-constant flowrates in the model. The recommended setting is 1.0e-5.

4.2.9 *The RUN option*

By selecting the RUN option, a submenu appears on the screen as shown in Figure 4.19.

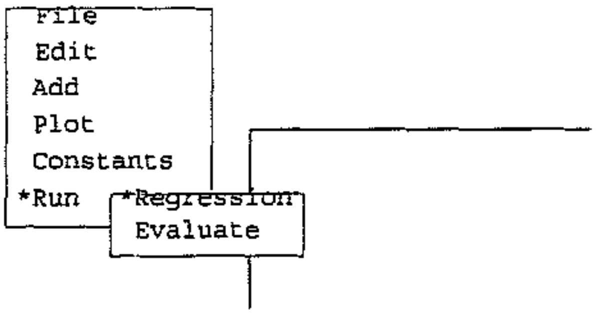


Figure 4.19 : *The RUN submenu*

Moving the cursor to an option and pressing *Enter* selects that option.

- **Regression** has two options :

  Regress starts the regression procedure on all model parameters specified as variable.
  Choose chooses the output curve to compare with the reference curve. There may be more than one output curve on some models.

- **Evaluate** starts the processing and the curve fitting procedure.

4.3 Examples

A number of typical residence time problems are presented to illustrate the program's capability.

4.3.1 *Example 1*

Figure 4.20 shows the tracer response curve for a reactor which is meant to behave as an ideal mixer. A spike of tracer was injected into the inlet line leading into the reactor, and the concentration was measured at the outlet of the reactor. The flow through the reactor was held constant during the experiment. Using **IMPULSE**, determine the model for the reactor and assess its performance.

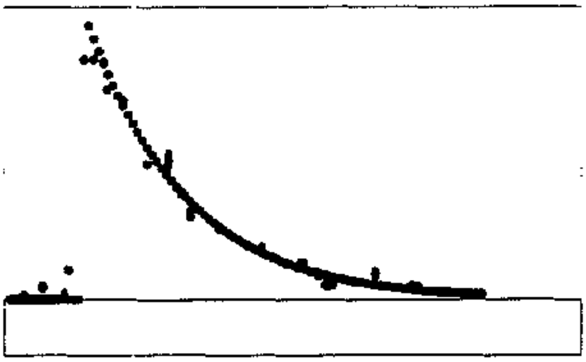


Figure 4.20 : *Raw data for example 1*

The delay before the tracer appears indicates an ideal plug flow region. The perfect exponential curve indicates an ideal mixed flow region. This is in series with the plug flow region. There appears to be no other flow characteristics. The quantity of dead space cannot be determined as there are no quantities given. The model is shown in Figure 4.21.

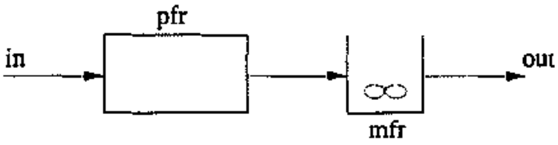


Figure 4.21: *Model for example 1*

The plug flow region in the model is explained by the inlet line to the rector. This delayed the tracer. The reactor operates as an ideal mixer. The modelling results are shown in Figure 4.22.

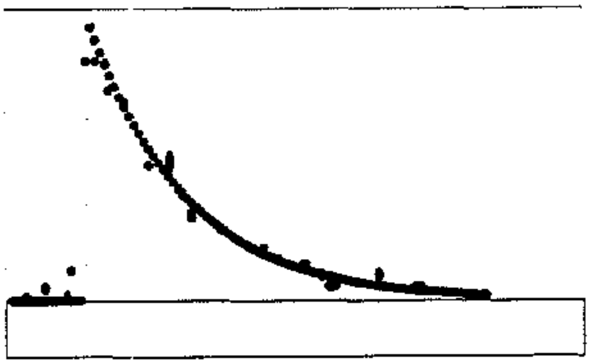


Figure 4.22 : *Result for example 1*

4.3.2 Example 2

A tracer was injected into a mixing vessel as there appeared to be a bypass over the vessel. The tracer response curve, from a spike of tracer injected into inlet line to the vessel, is given in Figure 4.23. Determine whether there is a bypass over the vessel, assuming flow was held constant during the experiment.

Figure 4.23 : *Raw data for example 2*

The delay before the tracer appears indicates a plug flow region. This is in series with a mixed flow region is indicated by the exponential curve. There is a bypass over the mixed flow region, but there is some mixing in the bypass, as the spike has an exponential character. The model is shown in Figure 4.24.

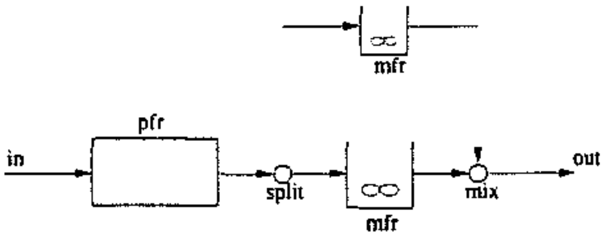


Figure 4.24 : *Model for example 2*

The plug flow region may be explained by the inlet line to the vessel. The vessel has a large ideal mixing zone by there is, however, a bypass on the vessel. The modelling results are shown in Figure 4.25.

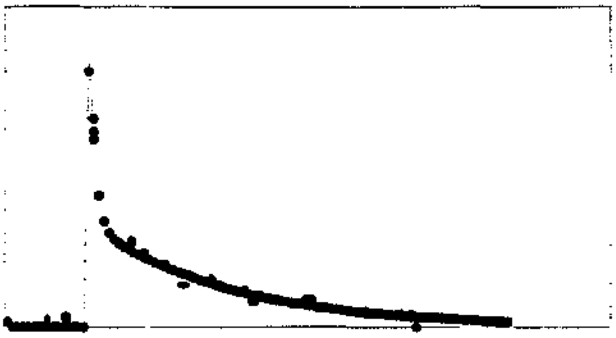


Figure 4.25 : *Results for example 2*

4.3.3 *Example 3*

A tracer was injected into the inlet of a system through to act as a plug flow vessel. The tracer response curve determined is shown in Figure 4.26. Determine the flow pattern for the system and whether the assumption of plug flow is correct.

Figure 4. 26 : *Data for example 3*

The data indicated that there may be more than one plug flow region, from the delay before the tracer and from the slight 'bumpiness' of the curve. The regular 'bumps' indicate that there is a recycle, possibly with a plug flow delaying the 'bumps' regularly. The overall shape is an exponential curve, indicating an ideal mixer. The model is shown in Figure 4.27.

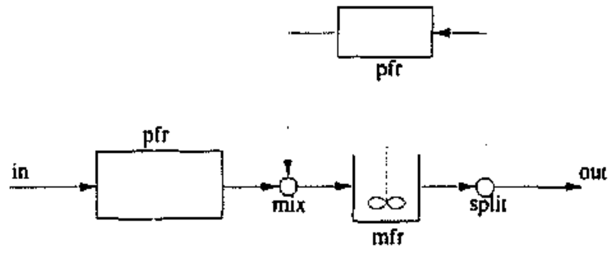


Figure 4.27 : *Model for example 3*

There is a large mixing region that is recycling in the system and the system has very little plug flow character. The modelling results are shown in Figure 4.28.

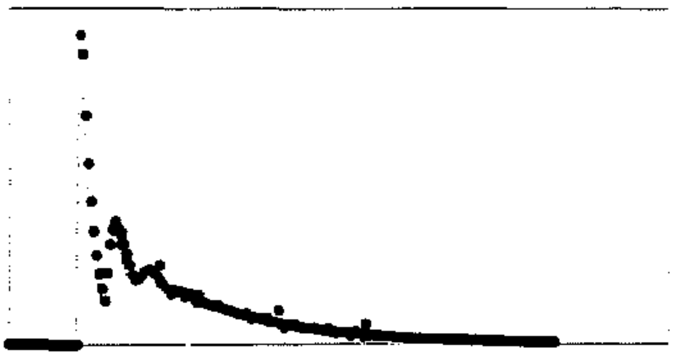


Figure 4.28 : *Results for example 3*

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Third Southern African Anaerobic Digestion Symposium, University of Natal, Pietermaritzburg, South Africa, 13-16 July 1992.

DANCKWERTS, V P (1953) *Continuous flow systems : distribution of residence times.*
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SMITH, T N (1988) *Chemical engineering kinetics.*
John Wiley & Sons, Inc., New York.

INFORMATION AND ERROR MESSAGES

- **Are You Sure That You Want To Exit? <Y> to exit :**

  This is displayed when the user presses the escape key from the main menu. When this message is displayed, pressing 'y' or 'Y' will exit the application, and return the user to the dos prompt, or any menu system that was used to invoke the program. Pressing of any other key will return the user to the main menu. Care should be taken to ensure that any work that is needed at a later stage is saved before exiting, as automatic saving of work on exit is not implemented.

- **Cannot Open File *<FILENAME>* :**

  There has been an error during the file open process. There may be a number of causes for this. The only cause specific to the package is if the file that is specified is not in the current working directory. All files MUST be in the directory that the package was invoked from. For any other causes, the user should consult a DOS manual or a network manual if the files are stored on a network drive.

- **Enter File Name :**

  This message appears during the file import process (for importing a set of points into an *'Editable Curve'* by pressing the 'F' key), during the 'File -> Save' process if the working file was not originally loaded from disk (i.e. created from scratch), or during the 'File -> Save As' process. Note that no path should be specified for the file... the file will be saved to the current working directory. (The directory that the package was invoked from).

- **Error Opening File : *<FILENAME>* :**

  See '*Cannot Open File <FILENAME>*'.

- **File Already Exists! Overwrite? <y/n> :**

  This message is displayed when the user requests a save for the model under development, and the specified save filename already exists. This message will not be seen if the current model was initially loaded from disk, altered and then saved using the 'save' option, rather than the 'save as' option.

- **No Valid Point Could Be Found In File *<FILENAME>* :**

  This occurs when the user is in an 'Editable Curve' and has selected a file from which to import data, where the specific file does not contain at least one numeric pair on a single line of ASCII text, separated by anything non-numerical.

- **Not Implemented Yet!! :**

  This message should never be seen by the user. If it is, the user is requested to contact the Pollution Research Group, to report it.

- **Press Any Key to Stop The Regression :**

  While in an regression procedure, this box is continually displayed. The regression information box will continuously update during this procedure, display the model iteration number. If the user presses a key, it is unlikely that the regression procedure will terminate immediately. This is often disturbing, but nevertheless necessary, in order to ensure that the set of parameters that remain after the user has requested a regression termination are better than the starting estimates.

- **Reduction Tolerance % :**

  This message is displayed, along with a numerical entry box when the user has selected the file import facility in an 'Editable Curve'. This figure is used for point filtering, and only really has to be used if the input file has more than about 100 points in it.

- **Running.. :**

  This message is displayed while the current model is being evaluated. As soon as it is no longer visible, the model calculations are complete. Users running simple models on fast machines may not see this message at all.

- **Saving : *<FILENAME>* :**

  This message is displayed whenever a file save has been requested and stays on the user display as long as the file writing process is in action. The file saving is complete as soon as the message disappears.

- **There Are Units That Still Need Input Connections :**

  Each unit in the package that is used for modelling has certain connection requirements, that must be satisfied. It does not, for example, make sense for a mixed flow reactor to have more than one input, but it does have to have the input. (It would be useless for modelling flow processes if it did not have a flow input!) If, for example, a mixed flow reactor was created, and did not have any connection going to it from another unit, this message will be displayed when the user tries to run the model. The connection requirements are presented in the table below.

Table A.1 : Unit Connection Requirements

UNIT	INPUTS	OUTPUTS
MFR	1	1
MFRSER	1	1
PFR	1	1
PFRDISP	1	1
MIXSPLIT	At least 1	At least 1
INPUT	0	1
OUTPUT	1	0

- **There Are Units That Still Need Output Connections :**
  See the above message for details.

- **The Following Units Need Modification :**
  This message appears whenever one of the parameters specified in any of the units used in the current model is invalid, and a model run is requested. A list of the offending units will be displayed, and the user may alter them at this point, (a unit disappears off the list once all its parameters are valid) or the user can press the escape key to return to the menu system, and alter them using the *EDIT->UNIT* option. The most common causes of this message are listed in the table below, by unit type.

Table A.2 : Causes of Incorrect Unit Configuration

UNIT	CHIEF CAUSE
MFR	The volume has been set to a value of zero or less.
MFRSER	The volume has been set to a value of zero or less. The number of series mixed flow reactors is less than or equal to zero.
PFR	The dead volume has been set to zero or less.
PFRDISP	The dead volume has been set to zero or less. The Peclet number is less than or equal to zero. The Peclet number is greater than 1500
MIXSPLIT	A split proportion has been specified with a negative value.
INPUT	One of the input profiles (flowrate or concentration) has a negative value of flowrate or concentration/
OUTPUT	-

- **The Unit Name (The First Line In The Box) Must Consist Of Non-Blank Characters :**

  The name of each unit MUST consist of a non-empty set of ASCII characters. Spaces and tabs are not counted, as they are blank, and cannot be readily seen by the user.

- **This Is Not A Valid 1.01 Beta Analyser File :**

  This message appears when the user is trying to load an invalid file. Note that the same file format has been used from version 1.01 to the present, 1.022. If the file is indeed an impulse file and will not load, then the file has probably been corrupted.

- **Unit Name Must Be Unique :**

  Each unit that is specified in the working set MUST have a unique name.

- **You Have Not Specified Any Regressable Parameters! :**

  This is displayed when a regression has been requested, but no model parameters (split proportions, volumes, peclet numbers, scaling factor etc.) have been set to *'variable'*.

- **You Haven't Chosen a Curve To Regress Against Yet! :**

  As there is a single reference curve (normally containing experimental data), and no restriction on the number of output units that may be specified, the user is required to select the output unit that is to be compared with the experimental data that is contained in the reference.

- **Your Delta Needs Adjustment :**

  Unfortunately, due to the silly memory restriction that were placed on DOS applications, no solution may have more than about 3000 integration intervals in it. The User must ensure that $\frac{Stop\ Time - Start\ Time}{Delta\ T} \leq 3000$. (These settings can be found in the 'Constants" menu.

- **Your Delta T Is Too Small**

  See above.

- **Your Flowrate Convergence Tolerance Is Too Small :**

  The flowrate convergence tolerance must be greater than or equal to 1.0e-6. This is necessary, to ensure that convergence is possible. (Round-off errors which are inherent in floating point calculations can accumulate in the system, and may stop the solution from converging if this number could be set to a smaller value. A recommended setting is 1.0e-5.

- **Your Start Time and Stop Time Need Adjusting :**

  This appears if the user has specified the stop time to be smaller than or equal to the start time. (This obviously does not make sense!)

How Do I?

- **Make A Curve?**

  Firstly, get into an *'Editable Curve'*. (Examples are : The reference [Edit ->Reference], an input unit concentration curve and an input flowrate curve). If no editing has taken place, the user will see a scale factor on the first line (initially set to 1), and a point on the second line (initially set to 0 0). (The first value is the time, and the second value is the property value at that time [concentration or flowrate, depending on which curve is being edited)].

  To **edit** any point that is already in the list, simply move the cursor to the point for alteration, and press enter. An editing box will appear, with the values in it. Simply type in the new pair of values, separated by spaces. When finished, simply press enter, and the point will be updated in the list.

  To **add** a new point, set the cursor next to any point that already exists, or to the blank line immediately below the past point in the list, and press *'INSERT'*. An editing box will appear, this initial abscissa and ordinate values of 0. Alter, if necessary, and press enter when complete. The new point will appear in the list. Points do NOT have to be placed in the list in any particular order. They will be sorted as soon as the user exits the *'Editable Curve'* edit box.

  To **remove** a point from the set, simply place the cursor next to the point for deletion, and press the 'DELETE' key.

- **Import Data Into An 'Editable Curve' From a Spreadsheet?**

  From the spreadsheet, use the ***'Print Range To File'*** option, or some equivalent (extract semantics vary from spreadsheet to spreadsheet). Make sure that you have only two columns of data selected for output, and that the time fields are in the FIRST column. Give the print file a name, and on exiting the spreadsheet, ensure that the file is in the Working directory when invoking IMPULSE.

  Once in IMPULSE, go into the *'Editable Curve'* where the exported data is to be put, and press the 'F1' key. You will be prompted for a file name. Simply enter the file name that you created from the spreadsheet. (This file MUST be in the current working director!!!). You will now be asked for a reduction tolerance percentage. If you do not want any points in the input file to be filtered out (removed), simply press the 'ESCAPE' key. If a percentage value of *X* percent deviation from a straight line between their immediate left right neighbours will be removed.

FREQUENTLY ASKED QUESTIONS

- **What Unit System Does IMPULSE Use?**

  There is no prescribed system of units in IMPULSE. The user may use any set of units that makes sense, as long as all the data entered by the user are in the same set of consistent units. (If volumes in cubic angstroms, masses in megarams and times in microseconds seem fit for an application for some reason or other, then all flowrates entered should be entered in megagrams per cubic angstrom). It is up to the user to remember what the unit set is that they have chosen to work in.

- **What Are The Limitations?**

  The number of integration intervals for any run of a model is limited to about 3000 or so. This should be more than sufficient for any model, but is an artificial restriction imposed due to the small amount of memory available to DOS applications.

- **Why Do My Results Look Odd?**

  There are a number of causes for this :

  1) As there is only a single viewpoint on the screen, all the curves that have been included in the list of curves to be displayed are automatically scaled, in order to display all of them simultaneously in their entirety. If one curve in the set has upper and low values of 100 and 0, respectively, and all other curves have a cumulative maximum and minimum of 1 and 0, respectively, the first curve can 'overpower' the other curves on the display. (They will virtually appear as straight lines on the bottom of the screen, as their upper and lower values pale in comparison with [100,0].
  2) The user has not specified flowrate and concentration curves over the specified simulation interval. ([start time, stop time]). Every point outside the time interval range that is given in an *'Editable Curve'* is considered zero.
  3) The DeltaT specified for the simulation is too small. (If, for example, the start time and stop time are set to 0 and 2, respectively, and the DeltaT is set to 1, only 3 integration intervals will be used!. This is not suitable for any half-descent simulation!).

- **How Do I Represent An Impulse (Spike) Tracer Input?**

  When doing this, the user must remember that concentrations are used for input, NOT masses. (The input unit should be thought of conceptually as a pipe entering the system, into which the tracer is injected). Therefore, the concentration which is attributed to the addition of a certain specified mass is dependent on the flowrate in the 'pipe'. The user should also realise that the addition of a specified mass has a certain, finite time associated with it. (The dirac delta so often used in mathematical modelling is an idealised situation, and cannot be physically realised in practice).

EXAMPLE :

Let us say, for example, that the steady addition of 5 kg of tracer to an input stream flowing at 10 m^3/s takes the experimenter 4 seconds. This means that the actual concentration of tracer in the input stream during this period is :

$conc = \frac{5kg}{10\frac{m^3}{s} * 4s} = 0.125\frac{kg}{m^3}$. Note that this concentration exists for the entire 4 second interval!

As linear interpolation is used internally between any two points in the *'Editable Curves'*, a point list looking like the following for the flowrate and concentration curves can be used (assuming that the simulation start time and stop time are 0 and 100 seconds, respectively) :

FLOWRATE		**CONCENTRATION**	
Time	Value	Time	Value
0	10	0	0
100	10	0.001	0.125
		4.001	0.0125
		4.002	0

- **Who Do I Contact If I Have a Query?**

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Water Research Commission Project No. 363

The use of small-scale equipment for evaluating water treatment plants

APPENDIX C

Summary of Findings of a Comparison Between the Conditions for Jar Tests and Actual Plant Performance

Compiled by

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Department of Chemical Engineering
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Durban

November 1998

Calculation of GT values

In the calculation of G for a weir:

$$G = \left(\frac{h \times q \times g}{V \times \nu}\right)^{\frac{1}{2}}$$

$$= \frac{h \times \left[q \times \left(\frac{10^3}{3600 \times 24}\right) \times 9{,}81\right]}{V \times 10^{-6}}^{\frac{1}{2}}$$

$$= \left(\frac{h \times q}{V} \times 1.1354 \times 10^5\right)^{\frac{1}{2}}$$

The terms h,q and V must be entered in the units m, Mℓ/d and m³ respectively.

In the case of an electric mixer

$$G = \left(\frac{P}{V \times \mu}\right)^{\frac{1}{2}}$$

$$= \left(\frac{P}{V \times 10^{-3}}\right)^{\frac{1}{2}}$$

where :

G	=	average velocity gradient	s^{-1}
h	=	height of weir	m
q	=	volumentric flow rate	m^3/s
g	=	acceleration due to gravity	$9.81\ m.s^{-2}$
V	=	volume of mixing chamber	m^3
ν	=	kinematic viscosity = 10^{-6} at 20°C	$m^2.s^{-1}$
μ	=	solute viscosity = 10^{-3} at 20°C	$N.s.m^{-2}$
P	=	electric power drawn by motor	W

TABLE 3 : Summary of findings of a comparison between the conditions of jar tests and actual plant performance

	PLANT				**JAR TEST : ON-SITE**				**JAR TEST : CENTRAL LAB**			
PLANT	**Mixing mech.**	**Rapid mix**			**Rapid mix**				**Rapid mix**			
		G-value (s^{-1})	**Time (s)**	**GT (-)**	**Speed (rev^{-1})**	**G (s^{-1})**	**Time (s)**	**GT (-)**	**Speed (rev^{-1})**	**G (s^{-1})**	**Time (s)**	**GT (-)**
DURBAN HEIGHTS (450 Ml/d)	weir 1	175	128	22 400	100	60	120	7 200	100	60	120	7 200
	mixers 1 and 2	211	68	14 263								
	mixers 3 and 4	225	59	13 365								
	weir 2	no volume details otained										
	Candy-Moore weir	696	11	7 588								
	New Degremont	498	40	19 928								
	Old Degremont	435	49	21 266								
D V HARRIS (90 Ml/d)	in-line mixing	not available			100	60	120	7 200	100	60	180	10 800
	weirs	351	48	16 726								
HAZELMERE (23,5 Ml/d)	mixer 1	218	183	39 776	300	250	180	45 000	300	250	120	30 000
	mixer 2	224	183	41 040								
	weir 1	278	99	27 528								
	weir 2	527	34	17 641								
	weir 3	505	30	14 950								

TABLE 3 - Summary of findings of 1994 survey (continued)

	PLANT				JAR TEST : ON-SITE				JAR TEST : CENTRAL LAB			
PLANT	Mixing mech.	Rapid mix			Rapid mix				Rapid mix			
		G-value (s^{-1})	Time (s)	GT (-)	Speed (rev^{-1})	G (s^{-1})	Time (s)	GT (-)	Speed (rev^{-1})	G (s^{-1})	Time (s)	GT (-)
H D HILL (11 Ml/d)	mixer	937	23	21 543					300	250	120	30 000
MILL FALLS (5,5 ML/d)	weir 1	236	134	34 009	200	150	120	18 000	100	60	120	7 200
	weir 2	442	20	8 876								
AMANZIMTOTI (7,5 Ml/d)	Information not available to date											
UMLAAS ROAD (38,5 Ml/d)	weir 1	271	89	24 208					100	60	120	7 200
	weir 2	557	21	11 575								
WIGGINS (140 Ml/d)	weirs 1 and 3	451	43	19 483	100	60	120	7 200	300	250	180	45 000
	weirs 2 and 4	393	57	22 440								

Water Research Commission Project No. 363

The use of small-scale equipment for evaluating water treatment plants

APPENDIX D

Construction and Characterisation of a Small-Scale Rapid Mixer

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November 1998

1 DESCRIPTION OF WIGGINS WATER TREATMENT PLANT RAPID MIXER

Wiggins water treatment plant is capable of handling a maximum capacity of 175 Mℓ/day. At present the plant is running at 160 Mℓ/day.

Raw water entering the plant is pre-chlorinated and pre-limed before being split equally into four smaller channels which lead to the hydraulic jumps. At each hydraulic jump there is a coagulant dosing point. The hydraulic jumps produce a high degree of turbulence which is used for the rapid mixing of the coagulant.

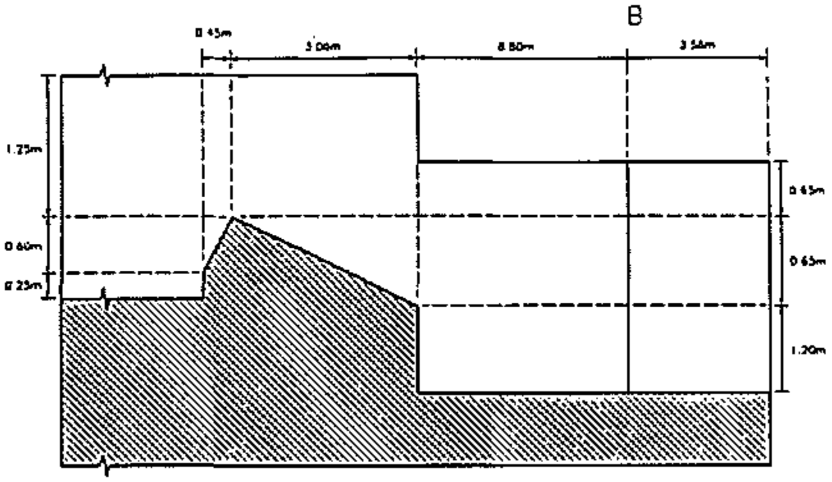
The dosed water leaving the hydraulic jumps then flows into four 4 000 m^3 upward flow pulsator clarifiers. These clarifiers are sludge blanket clarifiers.

Each clarifier has a retention time of approximately 2,4 hours with an average up-flow velocity of approximately 2 m/hr when the plant is running at a capacity of 160 Mℓ/day. The sludge blanket formed is approximately 2 m in height. A vacuum pump pulsates the sludge blanket every 40 seconds by sucking sludge from the sludge blanket and releasing it. The sludge blanket also acts as filter. The fully formed sludge overflows into a hopper while the clarified water flows out through the top of the clarifier.

The hydraulic jump rapid-mixer consists of an upper open channel connected to a lower channel via a downward sloping weir. Figure 1.1 illustrates the configuration of the hydraulic jump. The flow of water from the upper channel to the lower one causes an increase in velocity. This increase initiates the formation of the hydraulic jump in the lower channel. The turbulence caused by the hydraulic jump is used to mix the coagulants. The coagulant is fed from a feed tank to a pipe distributor which is positioned at point A in Figure 1.1. The coagulant flows out of this distributor and drops evenly onto the water surface.

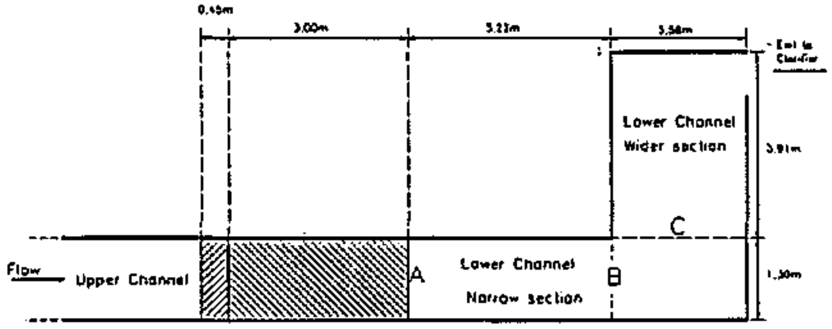
The level of the water in the lower channel is controlled by the water level in the clarifier. Visual observations suggest that the turbulent mixing zone is in the narrow part of the lower channel. It is expected that the rapid mixing is achieved in this part of the lower channel.

The objective of this study is to construct a small scale hydraulic jump that would would simulate the large scale jump.



Width of Channel = 1.30m

Side View



Plan View

FIGURE 1.1 : Dimensions of Plant Hydraulic Jump.

2 CRITERIA FOR THE DESIGN OF A SCALE MODEL

In order to design a scale model of the plant hydraulic jump, the laws of physical similarity have to be satisfied. The following section briefly describes the various laws of physical similarity.

In the design of scale models from prototypes, the laws of physical similarity have to be satisfied. Two systems are said to be physically similar in respect to certain specified quantities when the ratio of corresponding magnitudes of these quantities between the two systems is everywhere the same (Massey).

Generally results from experiments performed on the scale model are used to design as well as predict the behaviour of the prototype. The prototype can be either larger or smaller than the model. The proficient use of the laws of physical similarity together experimental investigation on the behaviour of the model can provide a better understanding of the prototype.

There are three types of physical similarity :-

(i) Geometric similarity,

(ii) Kinematic similarity,

(iii) Dynamic similarity.

2.1 Geometric Similarity

Geometric similarity is similarity of shape. The characteristic property of geometrically similar systems, is that the ratio of any length in one system to the corresponding length in the other system is everywhere the same (Massey). This ratio is known as the *scale factor*. Not only must the model be geometrically similar to the prototype, but the roughness of the surfaces must also be geometrically similar. This means that if the model is small and the surfaces are smooth, then the prototype (which is larger than the model) should have surfaces which are rougher. It is not always possible to achieve perfect geometric similarity, however it is essential that geometric similarity be achieved even if it is only in terms of the scale factor. Once this is achieved, the similarity laws can then be applied to achieve kinematic and dynamic similarity.

2.2 Kinematic Similarity

Kinematic similarity is similarity of motion. This implies that there must be similarity of lengths (geometric similarity) as well as time intervals. If the lengths as well as their corresponding time intervals of the two systems are in a fixed ratio then the velocities of the corresponding elements of fluids must be in a fixed ratio of magnitude at corresponding times.

2.3 Dynamic Similarity

Dynamic similarity is similarity of forces. Therefore if two systems are dynamically similar, then the magnitudes of forces at similarly located points in each of the two systems, must be in a fixed ratio.

For pipe flow usually the Reynolds number conserved between the model and the prototype. In this case flow is through an open channel. It is therefore important to determine which parameters are the most significant in open channel flow occur.

The liquid flowing through an open channel or vessel usually has the upper surface exposed to the atmosphere and since atmospheric pressure is relatively constant, open channel flow is usually caused by the weight of the liquid.

In uniform pipe flow the piezometric pressure drop across the pipe is defined by $p + \rho \times g \times z$, but in open channel flow only the second term, $\rho \times g \times z$, is significant. Open channel flow is always accompanied by a fall in the level of the surface.

Some examples of open channel flow are streams, rivers, canals and flumes. In general open channel flow is more complex than pipe flow.

Laminar flow in open channels very seldom occurs and in terms of engineering interest open channel flow can be assumed to be completely turbulent (Massey). The significant forces in open channel flow are gravity forces, inertial forces and pressure forces. Since open channel flow is usually turbulent, the inertial forces far outweigh the viscous forces. Consequently Reynolds number is seldom used in open channel flow to characterise flow conditions. Instead another dimensionless number called the Froude number (Fr) is used. The Froude number is the ratio of inertial forces to gravity forces, and is defined as :-

$$Fr = \frac{u}{(l \times g)^{\frac{1}{2}}}$$

where Fr = Froude number

u = characteristic velocity (ms^{-1})

l = characteristic length (m)

g = gravitational constant = 9.81 ms^{-2}

Ideally for a model and a prototype to be completely similar, all three of the above conditions of similarity must hold (i.e. there must be geometric, kinematic and dynamic similarity). However this is not always possible, due to the great number of variables (such as forces and velocities) which have to be satisfied simultaneously. In such cases the variables which have a negligible effect on the systems should be neglected.

3 DESIGN OF THE MODEL HYDRAULIC JUMP.

In open channel flow, kinematic and dynamic similarity between the prototype and the model is achieved by conserving the Froude number. Therefore :-

$$\left[\frac{u}{(l \times g)^{\frac{1}{2}}}\right]_m = \left[\frac{u}{(l \times g)^{\frac{1}{2}}}\right]_p \qquad 3.1$$

where the subscripts m and p refer to the model and the prototype respectively.

The model hydraulic jump was built, based on the geometric similarity of lengths and the conservation of equation 2.1.

The length scale ratio L_r is defined as :-

$$\frac{l_p}{l_m} = L_r$$

where the subscript r refers to the ratio of prototype to model, and the subscripts p and m refer to the prototype and model respectively.

From equation 3.1,

$$\left[\frac{u}{(l \times g)^{\frac{1}{2}}}\right]_m = \left[\frac{u}{(l \times g)^{\frac{1}{2}}}\right]_p$$

$$\frac{u_p}{u_m} = \left[\frac{l_p}{l_m}\right]^{\frac{1}{2}} \qquad 3.2$$

Let the velocity ratio be $U_r = u_p/u_m$ then equation 3.2 becomes :-

$$U_r = L_r^{\frac{1}{2}} \qquad 3.3$$

The time scale ratio T_r which is equal to t_p/t_m can be obtained as follows :-

$$U_r = \frac{u_p}{u_m} = \frac{\left[\frac{l_p}{t_p}\right]}{\left[\frac{l_m}{t_m}\right]}$$

$$= \left[\frac{l_p}{l_m}\right] \times \left[\frac{l_p}{l_m}\right]^{-1}$$

$$= \left[\frac{L_r}{T_r}\right]$$

where T_r = time scale ratio,

l = characteristic length (m),

u = characteristic velocity (ms^{-1}),

t = characteristic time interval (s).

p, m = subscripts which denote the prototype and the model respectively.

Therefore

$$T_r = \frac{t_p}{t_m} = \frac{L_r}{U_r}$$

Substituting for U_r using equation 3.3 gives :-

$$T_r = \frac{L_r}{L_r^{\frac{1}{2}}}$$

$$T_r = L_r^{\frac{1}{2}} \qquad 3.4$$

Q_r which is equal to $Q_p \;/\; Q_m$ is obtained similarly :-

$$Q_r = \frac{Q_p}{Q_m} = L_r^{2\frac{1}{2}} \qquad 3.5$$

where Q_r = Volumetric flow scale ratio,

Q = Volumetric flow (m^3s^{-1}),

p, m = subscripts which denote the prototype and the model respectively.

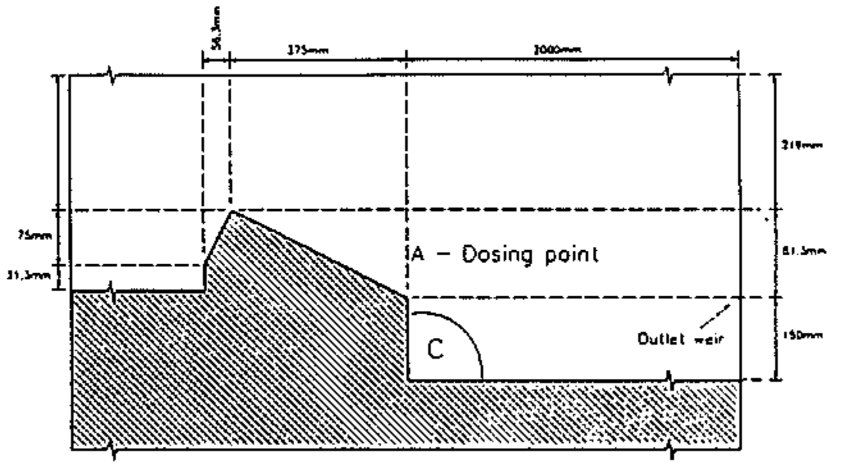
Equations 3.1 to 3.5 were used to determine flow conditions for the model hydraulic jump.

Earlier the dimensions of the plant jump were shown in Figure 1.1. These lengths were scaled down by eight to give the model dimensions. Therefore :-

$$L_r = \frac{l_p}{l_m} = 8$$

Observations of the mixing in the plant indicate that the most turbulent zone is between sections A and C in Figure 1.1. The zone between section C and the clarifier is far less turbulent, and it was assumed that only secondary mixing occurs in this region.

Considering the above assumption, the mixing volume was assumed to be between section A and section C (in Figure 1.1). The model was scaled down geometrically so that it represented the above mentioned mixing volume. The lower channel was made longer to reduce downstrem effects on the mixing upstream. Clear perspex was used to make the model. Figure 3.1 shows the dimensions of the model.



Width of Channel = 163mm

Side View

FIGURE 3.1 : Dimensions of Model Hydraulic Jump

In order to get a visual idea of the mixing and flow patterns in the model, dye was added at point A in Figure 3.1. It was found that the dye dispersed very rapidly in the turbulent zone. However a fairly stagnant recirculating zone (region C) existed directly under the hydraulic jump. Dye that entered this recirculating zone took a long time to pass through the system.

Another fact that was established by the dye tests was that large backflow currents existed throughout the lower channel. The cause of these backflow currents was the weir wall. These backflow currents were affected the mixing upstream.

To determine the effect of the backmixing caused by the weir wall, 20 mℓ of an 80 g/ℓ NaCl solution was added 50 mm before the weir. The conductance of the water was then measured 1 200 mm upstream. After 5 s the conductance began to increase from the initial 100 μ S. The conductance peaked at 250 μ S, 9 s after the NaCl was added. This implied an increase of 0,07 g/ℓin NaCl concentration at the peak. This showed that the backmixing caused by the weir significantly affected the mixing in the mixing volume.

In order to eliminate the back mixing, the lower channel was modified. The lower channel was then redesigned to represent the section from the hydraulic jump to the section leading into the clarifier on the plant. The lower channel now consists of a narrow turbulent region as well as a wider less turbulent region. Figure 3.2 shows the dimensions of the modified model hydraulic jump.

The volumetric flowrate through the model jump was obtained using equation 3.5. For a plant flow of 160 Mℓ/day, the flow through one full scale jump will be 40 Mℓ/day. Therefore :-

$$Q_p = 40 \ Ml/day$$

$$= \left[\frac{40 \times 1000000}{24 \times 60}\right] \ l/\mathrm{min}$$

$$= 27\ 778 \ l/\mathrm{min}$$

Since $L_r = 8$, equation 3.5 becomes,

$$Q_r = L_r^{2\frac{1}{2}} = 8^{2\frac{1}{2}}$$

$$= 181,02$$

Now,

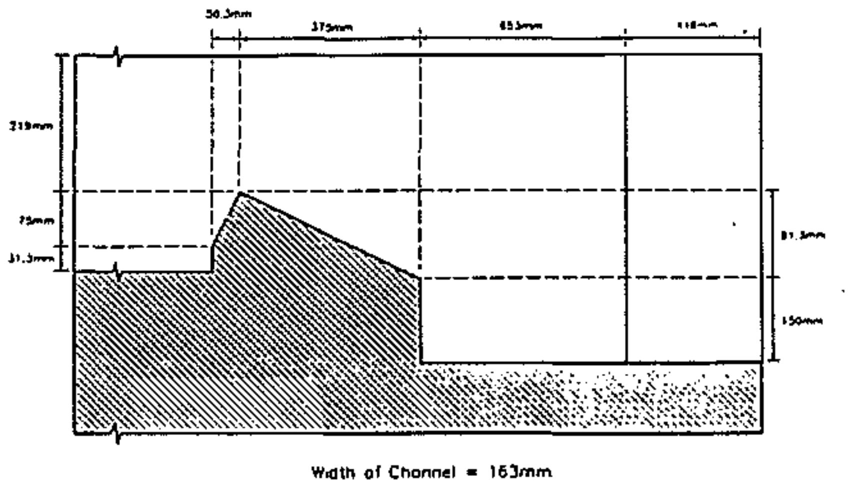
$$Q_r = \frac{Q_p}{Q_m} = 181,02$$

$$\therefore \ Q_m = \frac{Q_p}{181,02}$$

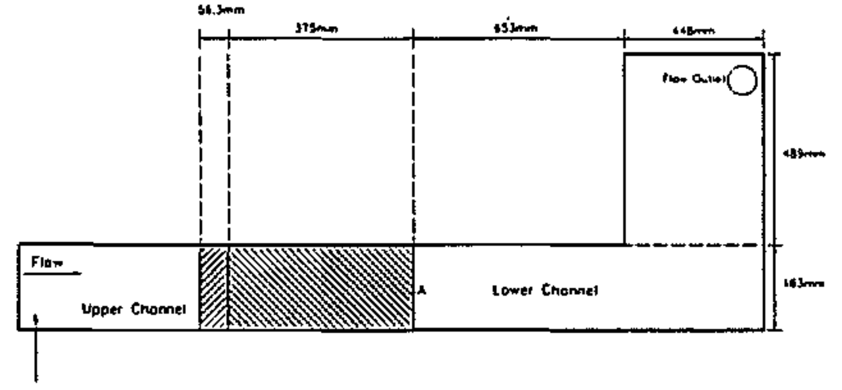
Q_p = 27778 ℓ/min, therefore

$$Q_m = \frac{27778}{181,02}$$

$$= 153,45 \ l/\mathrm{min}$$



Side View



Plan View

FIGURE 3.2 : Dimensions of the Modified Hydraulic Jump

By measuring average velocities at specific points in the plant hydraulic jump, the expected water velocities in the model can be calculated using equations 3.2 and 3.3. These calculated flowrates and velocities can then be compared to experimentally measured quantities to check whether the model has been built correctly.

4 TESTS AND CHARACTERISATION OF THE MODEL JUMP.

For rapid mixing to be similar between the plant hydraulic jump and the model, energy dissipation and flow conditions have to be similar between the two. This section concerns the characterisation of energy dissipation and flow conditions in the model jump.

4.1 Objectives

Geometric similarity has been achieved. However this satisfies only one part of physical similarity. It is now necessary to verify the kinematic and dynamic similarities between the plant hydraulic jump and the model.

The objectives are:-

(i) characterisation of the fluid dynamics and the energy dissipation in the plant and model hydraulic jumps,

(ii) characterisation of the mixing on the model and the plant hydraulic jumps.

4.2 Characterisation of the Fluid Dynamics and the Energy Dissipation in the Model and the Plant Hydraulic Jumps.

The rapid mixing stage of the flocculation process is one of the most important stages of flocculation. The efficiency of the rapid mixing often depends on the amount of energy that is dissipated during this stage. Sufficient energy must be available for the efficient dispersion of the coagulants. The amount of energy that is dissipated in an open channel can be derived from hydraulic gradients. The energy dissipation obtained can then be used to determine the mean velocity gradients.

4.2.1 Comparison of Hydraulic Gradients

For flow conditions in the model to be kinematically similar to conditions on the plant, the hydraulic gradients at kinematic similar flowrates should be similar (i.e. the liquid depth at a point in the plant should be similar to the liquid depth at that equivalent point on the model for kinematic similar flowrates.)

To determine the similarity of hydraulic gradients between the plant jump and the model, liquid depths were measured on both the plant and the model. Figures 4.1 and 4.2 show the points at which the liquid depths were measured.

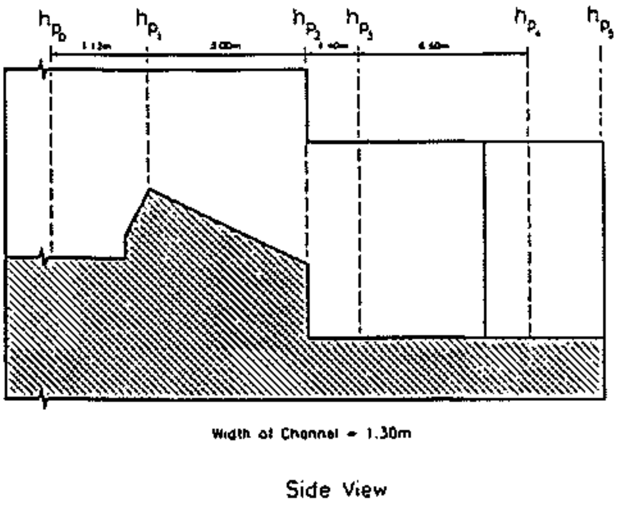


FIGURE 4.1 : Points at which Liquid Depths were taken on the Plant.

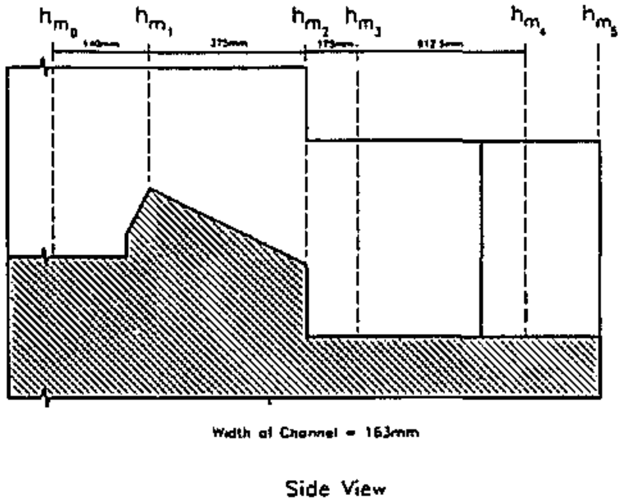


FIGURE 4.2 : Equivalent Points at which Liquid Depths were taken on the Model.

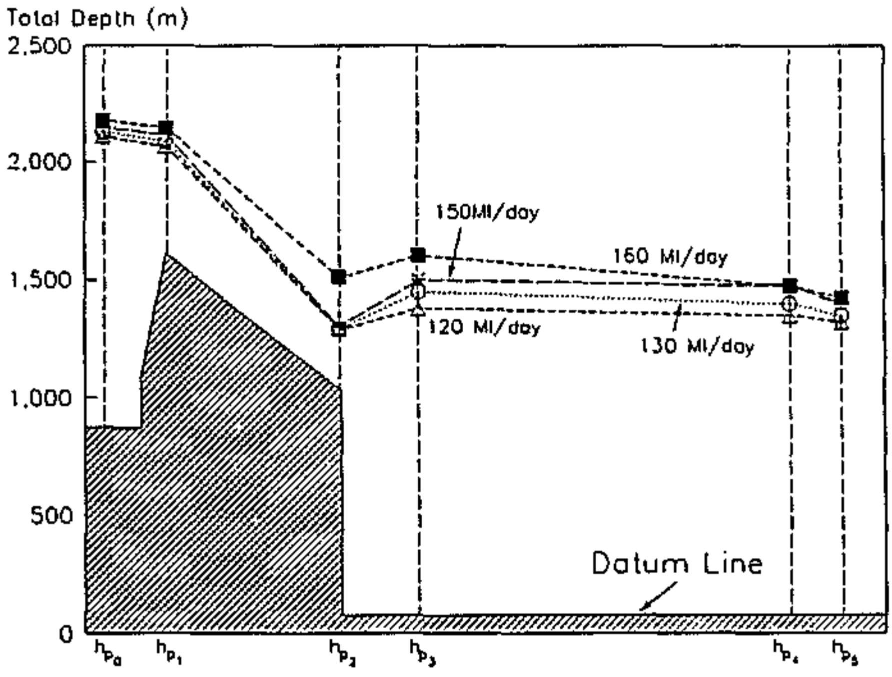
Points h_{p5} and h_{m5} are situated at the exit of the lower channel of the plant jump and the model jump respectively.

Table 4.1 shows the water depth measurements obtained from the plant hydraulic jump. It is important to note that flow conditions above point h_{p2} are independent of the flow conditions in the lower channel. The water depth in the lower channel however, is controlled by the level of the water in the adjoining clarifier.

TABLE 4.1
Water Depths Measured on the Plant

Feed Flow Mℓ/day	h_{p0} mm	h_{p1} mm	h_{p2} mm	h_{p3} mm	h_{p4} mm	h_{p5} mm
120	1160	270	90	1380	1350	1300
130	1180	290	95	1450	1400	1325
150	1200	320	105	1500	1475	1375
160	1230	360	110	1600	1475	1380

Figure 4.3 shows the above data in terms of hydraulic gradients. The bed of the lower channel is taken as the datum line.



FGURE 4.3 : Hydraulic Gradients at Various Flowrates.

The results from Table 4.1 were divided by eight to give the equivalent liquid depths that are expected in the model jump at similar flowrates. These expected depths are shown in Table 4.2.

$h_{pe} = h_p/8$ where h_{pe} is the expected liquid depth for the model derived from h_p, and h_p is the liquid depth measured on the plant hydraulic jump.

Water depths were then measured on the model at kinetically similar flowrates to those flowing through the plant hydraulic jumps. These measured depths are shown in Table 4.3. h_m is the water depth measured on the model hydraulic jump.

TABLE 4.2
Expected Depths in the Model Hydraulic Jump

Equivalent Model Flowrate ℓ/min	h_{pe0} mm	h_{pe1} mm	h_{pe2} mm	h_{pe3} mm	h_{pe4} mm	h_{pe5} mm
115	145	33,75	11,25	172,5	168,75	162,5
125	147,5	36,25	11,88	181,25	175	165,63
144	150	40	13,13	187,5	184,4	171,88
154	153,75	45	13,75	200	184,4	172,5

TABLE 4.3
Actual Water Depths Measured on the Model

Equivalent Model Flowrate ℓ/min	h_{m0} mm	h_{m1} mm	h_{m2} mm	h_{m3} mm	h_{m4} mm	h_{m5} mm
115	145	33	10	167,5	165	162,5
125	146	35	10,5	170	170	165
144	150	39	11,5	177,5	175	165
154	152	44	12	185	175	165

A comparison of model water depths to water depths derived from the plant data shows that the model is a fairly good approximation of the plant hydraulic jump with respect to hydraulic gradients. Table 4.4 shows the % error between h_{pe} and h_m at the various flows.

TABLE 4.4
% Error Between Plant and Model Water Depths.

Equivalent Model Flow ℓ/min	% error between h_{pe0} and h_{m0}	% error between h_{pe1} and h_{m1}	% error between h_{pe2} and h_{m2}	% error between h_{pe3} and h_{m3}	% error between h_{pe4} and h_{m4}	% error between h_{pe5} and h_{m5}
115	0	2,2	11	3	2,2	0
125	1,03	3,6	12	6	2,9	0,3
144	0	5	14	5,6	5	4
154	1,15	2,2	14	8	5	4,5

The largest percentage error exists at point h_2. Apart from the discrepancies at this point the hydraulic depths at the other points compared very well. A plot of the model hydraulic gradients against the expected hydraulic gradients derived from plant data is shown in Figure 4.4.

From the above it is clear that the model and the plant jumps behave similarly with respect to hydraulic gradients. Since the hydraulic gradients are similar it is obvious that the energies being dissipated in the two systems should be similar at similar points. This is verified in the next section.

4.2.2 Comparison of Velocity Gradients

The mean velocity gradient $\overline{G}$ was developed by Camp (Bratby 1980)is the most widely used term to describe the degree of mixing achieved during any stage of flocculation. In order to determine the mean velocity gradient of the mixing process it is necessary to know the energy that is dissipated.

The following method was used to calculate the energy dissipation in the lower channel of the hydraulic jump. The mean velocity gradient $\overline{G}$ was calculated based on this method. Consider Figure 4.5.

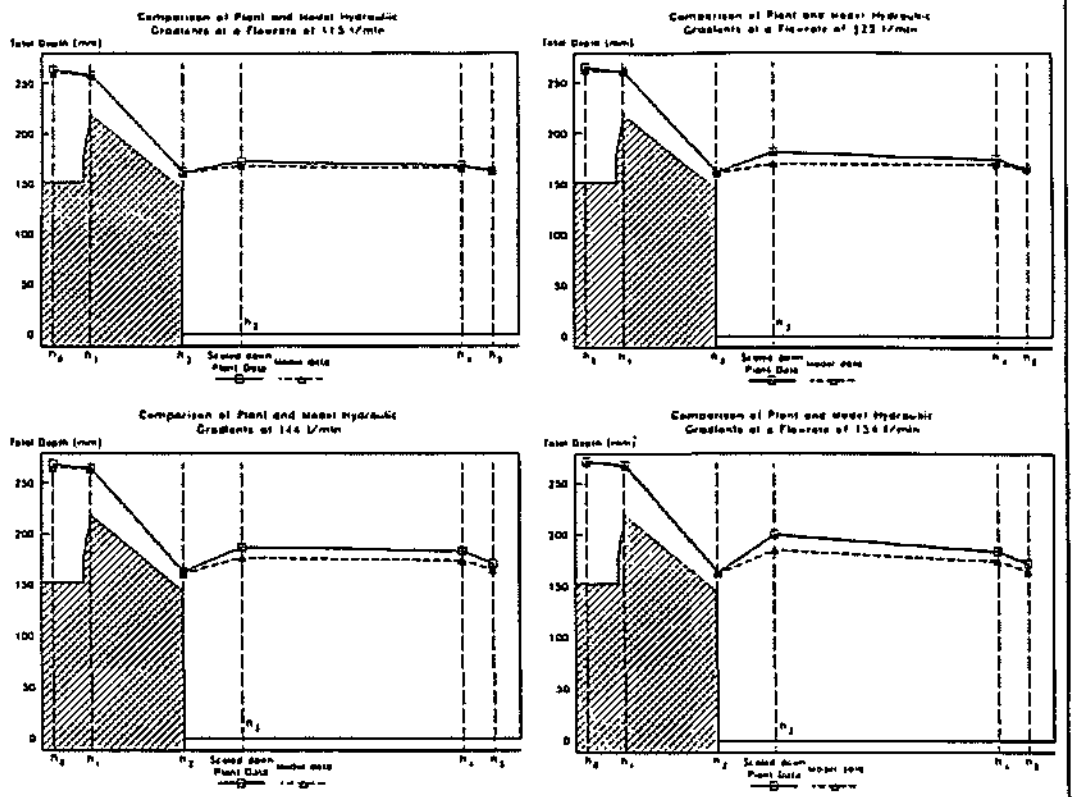


FIGURE 4.4: Comparison of Plant and Model Hydraulic Gradients

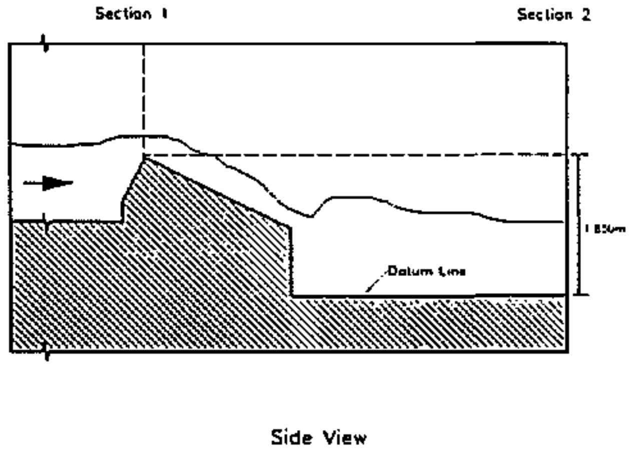


FIGURE 4.5 : Hydraulic Jump

Let the bed of the lower channel be the datum line. The loss of energy between points 1 and 2 is then given by :

$$\Delta H = \left[z_1 + h_1 + \frac{(u_1^2)}{(2 \times g)} \right] - \left[z_2 + h_2 + \frac{(u_2^2)}{(2 \times g)} \right]$$

where ΔH = head loss (m),

z = height of the channel bed above the datum line (m),

h = vertical height from the channel bed to the water surface (m),

u = average velocity (m/s)

1, 2 = indices representing sections 1 and 2 on Figure 3.5

ΔH is converted to power as follows:-

$$P = \rho \times g \times \Delta H \times Q \qquad 4.2$$

where P = Power or Energy dissipation (W),

ρ = density of water = 1000 kg/m^3,

g = 9,81 m/s^2,

Q = volumetric flowrate (m^3/s).

The mean velocity gradient is given by:-

$$G = \sqrt{\frac{P}{\mu \times V}} \qquad 4.3$$

P = Power or Energy dissipation (W),

G = mean velocity gradient (s^{-1}),

V = mixing volume (m^3)

μ = viscosity (kg/ms)

$$u = \frac{Q}{A} \qquad 4.4$$

where Q is the volumetric flowrate and A is the cross sectional area.

The relationship between the plant mean velocity gradient G_p and the model mean velocity gradient G_m is obtained as follows. Subscripts p and m refer to the plant and the model hydraulic jumps respectively.

$$G_p = \sqrt{\frac{P_p}{\mu \times V_p}} \qquad 4.5$$

and

$$G_m = \sqrt{\frac{P_m}{\mu \times V_m}} \qquad 4.6$$

From equation 3.5

$$Q_r = \frac{Q_p}{Q_m} = L_r^{2\frac{1}{2}} \qquad 4.7$$

Also

$$V_r = \frac{V_p}{V_m} = L_r^3 \qquad 4.8$$

$L_r = 8$ equations 4.5 and 4.6 become :-

$$Q_m = \frac{Q_p}{8^{2\frac{1}{2}}}$$

$$= 0,00552 \times Q_p \qquad 4.9$$

and

$$V_m = \frac{V_p}{8^3} = 0,001953 \times V_p \qquad 4.10$$

The relationship between the plant and the model head loss ΔH is given by :-

$$\Delta H_m = \frac{\Delta H_p}{8} = 0,125 \times \Delta H_p \qquad 4.11$$

The head loss on the plant is given by :-

$$P_p = \rho \times g \times \Delta H_p \times Q_p$$

and the head loss on the model is :-

$$P_m = \rho \times g \times \Delta H_m \times Q_m \qquad 4.12$$

Substituting equations 4.9 and 4.11 in equation 4.12 gives :-

$$P_m = \rho \times g \times 0{,}125\ \Delta\ H_p \times 0{,}00552\ Q_p$$

$$= 0{,}0006905\ \rho \times g \times \Delta H_p \times Q_p$$

$$= 0{,}0006905\ \Delta\ P_p \qquad 4.13$$

G_m can be found in terms of G_p by substituting equations 4.10 and 4.13 into equation 4.6 gives:-

$$G_m = \sqrt{\frac{0{,}0006905\ P_p}{\mu \times 0{,}001953\ V_p}}$$

$$= 0{,}595\ \sqrt{\frac{P_p}{\mu \times V_p}}$$

$$= 0{,}595\ G_p \qquad 4.14$$

For a plant feed flowrate of 160 Mℓ/day, the flow through one channel is 160/4 Mℓ/day. The water depths at points 1 and 2 are obtained from Table 4.1. The vertical water depth at section is 0,36 m and at section 2 is 1,38 m.

$$Q = \frac{40 \times 10^6}{(1000 \times 24 \times 60 \times 60)}$$

$$= 0{,}463\, m^3/s$$

$$u_1 = \frac{0{,}463}{(0{,}36 \times 1{,}3)}$$

$$= 0{,}989\, m/s$$

$$u_2 = \frac{0{,}463}{(1{,}38 \times 1{,}3)}$$

$$= 0{,}258\, m/s$$

Therefore from equation 4.1 the head loss is given by :-

$$\Delta H = \left[1{,}850 + 0{,}36 + \left(\frac{0{,}989^2}{2 \times 9{,}81}\right)\right] - \left[0 + 1{,}38 + \left(\frac{0{,}258^2}{2 \times 9{,}81}\right)\right]$$

$$= 0{,}877\, m$$

The energy dissipation in the lower channel is then given by :-

$$P = 1000 \times 9{,}81 \times 0{,}877 \times 0{,}463$$

$$= 3980\,W$$

$$V = vol \ in \ lower \ channel$$

$$= (8{,}8 \times 1{,}3 + 3{,}58 \times 3{,}91) \times 1{,}40$$

$$= 35{,}61\,m^3$$

Therefore the mean velocity gradient at 160 Mℓ/day is :-

$$G = \sqrt{\frac{3980}{9{,}5 \times 10^{-4} \times 35{,}6}}$$

$$= 343 s^{-1}$$

The expected mean velocity gradient of the model $G_{m\,e}$ is obtained using equation 4.14

$$G_{m\,e} = 0{,}595 \times G_p$$

$$= 0{,}595 \times 343$$

$$= 204{,}1 s^{-1}$$

The above method was used to calculate the velocity gradients on the plant as well as the model, Table 4.5 shows the G values obtained for the plant and Table 4.6 shows the expected G for the model (derived from the plant G values).

TABLE 4.5
Plant Velocity Gradients

Plant Feed Flowrate	Mean Velocity Gradient, G
Mℓ/day	s^{-1}
120	304
130	313
150	326
160	343

TABLE 4.6
Expected Mean Velocity Gradient for Model

Equivalent Model Flowrate	Mean Velocity Gradient, G
ℓ/min	s^{-1}
115	181
125	186
144	194
154	204

The actual G values for the model were calculated from Table 4.3 and the results are shown in Table 4.7.

TABLE 4.7
Model Velocity Gradients Calculated from Model Data

Feed Flowrate	Mean Velocity Gradient, G
Mℓ/day	s^{-1}
120	186
130	191
150	207
160	218

The expected G values for the model (derived from the plant plant G values) as well as the actual G values for the model were plotted against the equivalent model flowrate (Figure 4.6).

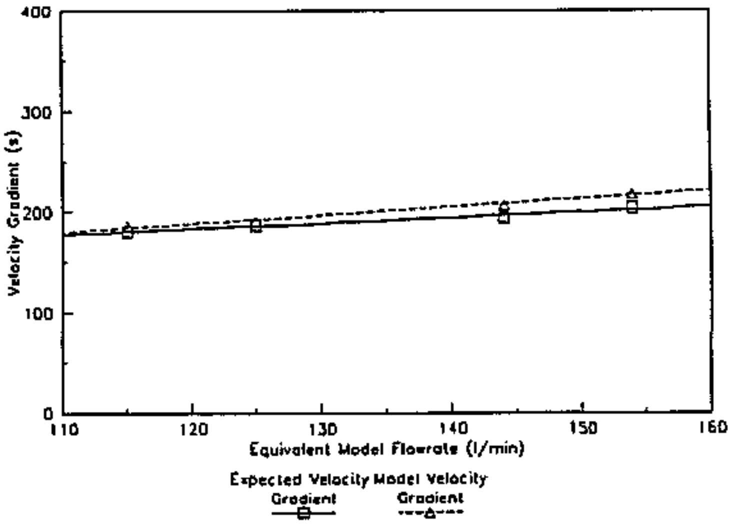


FIGURE 4.6 : Comparison of Plant and Model Velocity Gradients

The computer and conductance meter were then calibrated using standard NaCl solutions. Temperature has a significant effect on conductance therefore the calibrations were done at a number of temperatures. Figure 4.7 shows the effect of temperature on conductance.

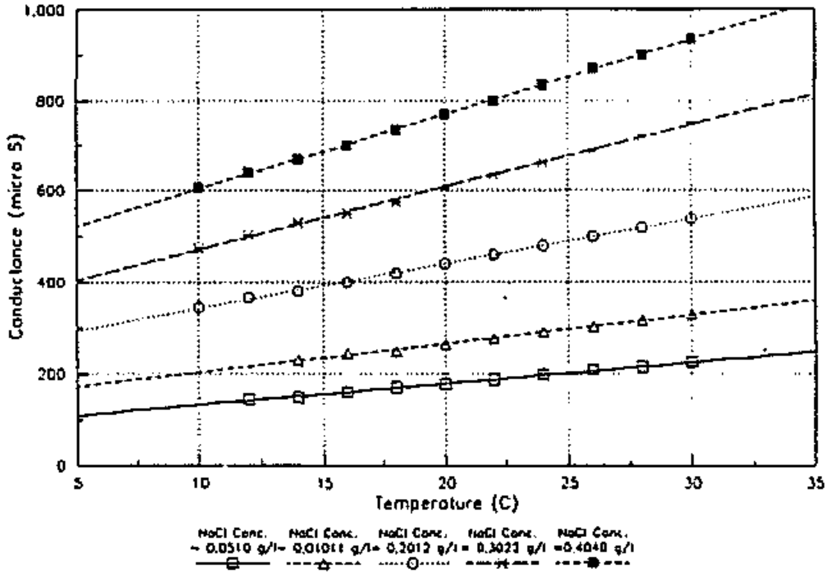


FIGURE 4.7 : *Effect of Temperature on Conductance*

Therefore for different temperatures, different calibration curves are available.

After many preliminary pulse tests done with different concentrations of NaCl, it was found that a 20 m*ℓ* pulse of an 80 g/*ℓ* NaCl solution provided the most suitable response. The following experimental procedure is used to carry out pulse tracer tests on the model :-

(i) A feed flowrate is set and the temperature of the water is recorded.

(ii) The computer and the conductance meter are connected and switched on. The conductance probe is placed into the channel at the required position and the conductance meter is zeroed. In the case of the global tracer test the probe is place in the exit stream.

(iii) The program is started, the sampling time is set to 0,2 seconds and a blank run is performed in order to get the conductance of the water.

(iv) 20mℓ of an 80 g/ℓ NaCl solution is placed into the doser and the program is started when the NaCl is dosed into the water.

(v) Once the conductance has decreased back to its original value the run is stopped and the data is saved on disc.

4.3.2 Results and Analysis

As discussed earlier two types of pulse tracer tests were done on the model hydraulic jump. The first was a "global" pulse tracer test and the second was a "local" tracer test.

4.3.2.1 "Global" Tracer Tests

In a process it is evident that not all elements of a fluid spend the same time in the unit. That is flow conditions alter the lengths of time that fluid elements spend in the unit.

The distribution of these times for the liquid leaving the unit is called the exit age distribution $E(t)$ or the residence time distribution of the fluid. $E(t)$ is usually defined in terms of the following normalized equation :-

$$\int_0^\infty E(t) \times dt = 1 \qquad 4.15$$

where t is time.

From the above equation the fraction of the exit stream of age between t and $t+dt$ can be represented by $E(t)dt$. Residence time distributions are usually described in terms of $E(t)$ and t

For the pulse test on the hydraulic jump, a mass M (g) of NaCl tracer is added to the system at $t=0$. The conductance was monitored and concentration profiles were obtained.

The mean residence time, $\bar{t}$ of a unit is defined as :-

$$\bar{t} = \int_0^\infty t \times E(t) \times dt \qquad 4.16$$

It is therefore convenient to convert the concentrations $C(t)$ to $E(t)$ for the mean residence time to be calculated.

By definition the fraction of material leaving a unit between t and $t+dt$ is $E(t) \times dt$. This is equivalent to $C(t)$ divided by the total area under the $C(t)$ vs t curve.

But from a unit balance on the mass of tracer added, the total area under the $C(t)$ vs t curve is also given by :-

$$A = \frac{M}{Q} \qquad 4.17$$

where A = total area under the $C(t)$ vs t curve,

M = mass of tracer added (g),

Q = volumetric flowrate (m^3/s).

Therefore

$$E(t) = \frac{Q}{M} \times C(t) \qquad 4.18$$

Therefore from equation 4.16

$$\bar{t} = \int_0^\infty t \times \frac{Q}{M} \times C(t) \times dt \qquad 3.19$$

But

$$\frac{M}{Q} = A = \int_0^\infty C(t) \times dt \qquad 4.20$$

Therefore equation 4.19 becomes

$$\bar{t} = \frac{\int_0^\infty t \times C(t) \times dt}{\int_0^\infty C(t) \times dt} \qquad 4.21$$

The above equation can be approximated by :-

$$\bar{t} = \frac{\sum_{0}^{\infty} t \times C(t) \times dt}{\sum_{0}^{\infty} C(t) \times dt} \qquad 4.22$$

The ideal mean residence time τ , is :-

$$\tau = \frac{V}{Q} \qquad 4.23$$

where τ = volumetric flowrate (m^3/s),

Q = volumetric flowrate (m^3/s),

V = volume of mixing (m^3).

Equation 4.22 was used to calculate the mean residence time of the water passing through the hydraulic jump rapid-mixer.

The feed flowrate was set at 115 ℓ/min and a pulse test was done with the probe located in the exit stream. Figure 4.8 shows the plot of NaCl concentration vs time.

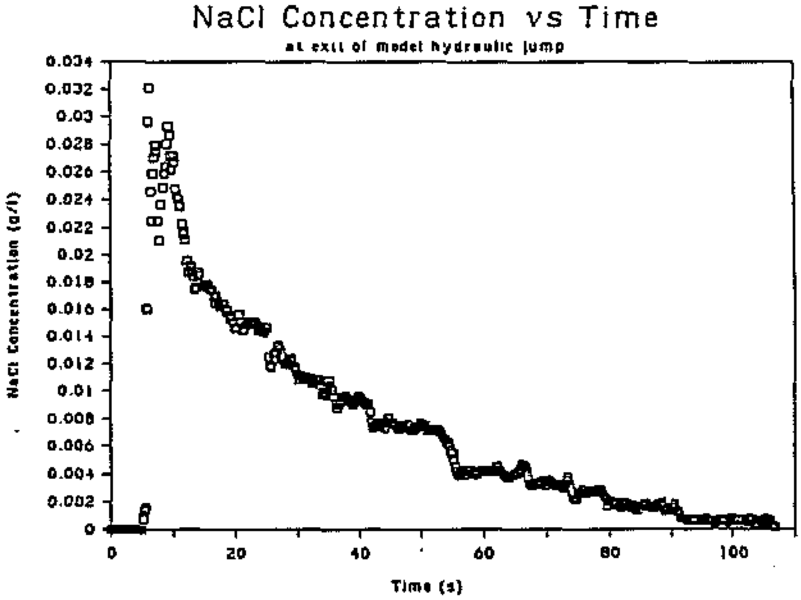


FIGURE 4.8 : Concentration vs Time Curve for the Hydraulic Jump

The shape of the curve clearly suggests that the hydraulic jump behaves like a mixed flow reactor (MFR). This type of curve is typical of unit processes that provide good mixing. The residence time curve can be modelled as a plug flow reactor in series with a mixed flow reactor as shown in Figure 4.9

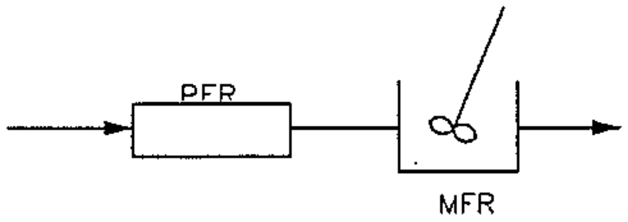


FIGURE 4.9 : Description of Mixing in the Model Jump

$$Q = 115 \ l/\text{min}$$

$$= 0{,}00192 \ m^3/s$$

The volume of water in the lower channel is the product of the area of the lower channel and the average height of the water above the bed of the channel.

$$V = area \times average \ height$$

$$= [(1{,}1 \times 0{,}163)+(0{,}448 \times 0{,}489)] \times 0{,}169$$

$$= 0{,}06732 m^3$$

The ideal mean residence time τ is

$$\tau = \frac{V}{Q}$$

$$= 0,\frac{06732}{0},00192$$

$$= 35{,}12 s$$

The mean residence time $\bar{t}$ was calculated using equation 4.23 and was found to be 31,66 s.

Therefore the active volume in the lower channel is :-

$$V_{act} = Q \times \bar{t}$$
$$= 0,00192 \times 31,66$$
$$= 0,0607 \ m^3$$

That is the percentage dead space in the lower channel is :-

$$\% \ dead \ space = \frac{(0,06732-0,0607)}{0},06732 \times \frac{100}{1}$$
$$= 9,8\%$$

One method to confirm the validity of the experimental procedure is to perform a mass balance on the tracer. Initially 20 mℓ of an 80 g/ℓ NaCl tracer was added. Therefore the mass of tracer into the system was 1,6 g.

The mass of NaCl leaving the system is given by :-

$$M_{out} = [\sum C_i \times \Delta t_i] \times Q$$

For the above experiment $\sum C_i \times \Delta t_i =$ 7605 (g/m^3) *s. Therefore

$$M_{out} = 760,5 \times 0,00192$$
$$= 1,46g$$

The percent error in the mass balance is

$$\% \ error = \left[\frac{1,60-1,46}{1},6\right] \times \frac{100}{1}$$
$$= 8,75\%$$

Hence, this confirms the validity of the global RTD experimental technique.

4.3.2.2 "Local" Pulse Tracer Tests

Pulse tests were performed at a series of grid points on the model jump. The objective here was to characterise the local flow patterns in the model jump. At each grid point tests were done at 50, 100 and 150 mm above the bed of the channel. Figure 4.10 shows the grid points at which the tests were done. Each dark node represents the point at which the test was done. Tests were repeated many times at each point.

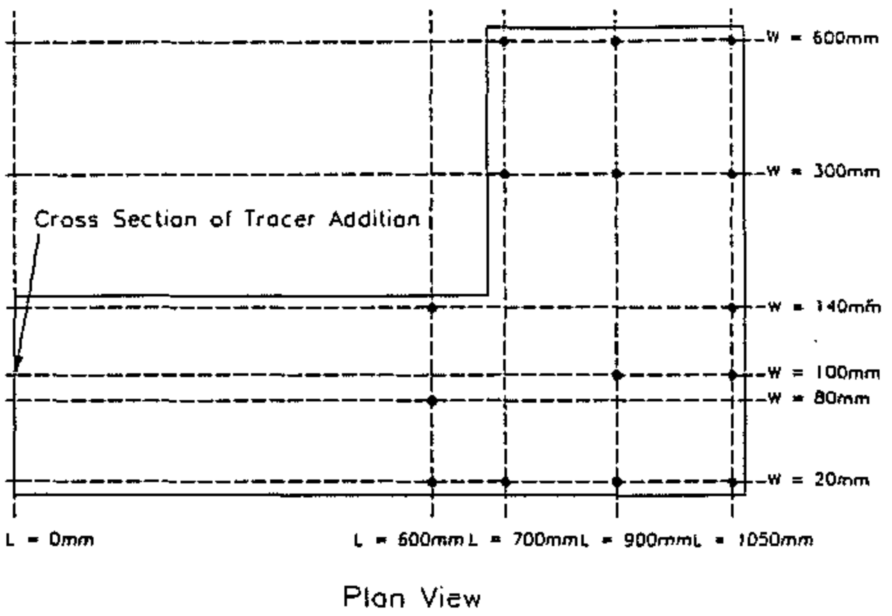


FIGURE 4.10 : *Grid Points at which Tests were Performed.*

Figures 4.11 to 4.14 shows a set of results of these tests run at a feed flowrate of 115 ℓ/min. Each set of graphs represents a cross section of the channel.

From Figure 4.11 it is clear that at L = 600 mm a large degree of bypassing occurs. This is most pronounced close to the surface of the water and least pronounced near the bed of the channel. This suggests that the tracer is flowing in a band (or section) close to the water surface. Once the plug of high concentration fluid passes it takes a very long time before the NaCl concentration drops to back to zero. One reason for this long tail on the graph is the backmixing. It is also evident that the bypassing occurs only in the region around the narrow section of the lower channel. Dye tests confirmed the backmixing.

As can be seen in Figures 4.12 to 4.14 the the mixing in the wider section of the lower channel behaves more like the mixing expected in a MFR. One possible reason for this is that there is more volume available in the wider section for backmixing to occur more efficiently.

This indicates that the mixing is not completed in the narrow turbulent section, as was expected earlier. Since it has been shown that the model jump is a fairly good representation of the plant hydraulic jump, it is highly possible that the mixing is not completed in the turbulent zone of the plant jump.

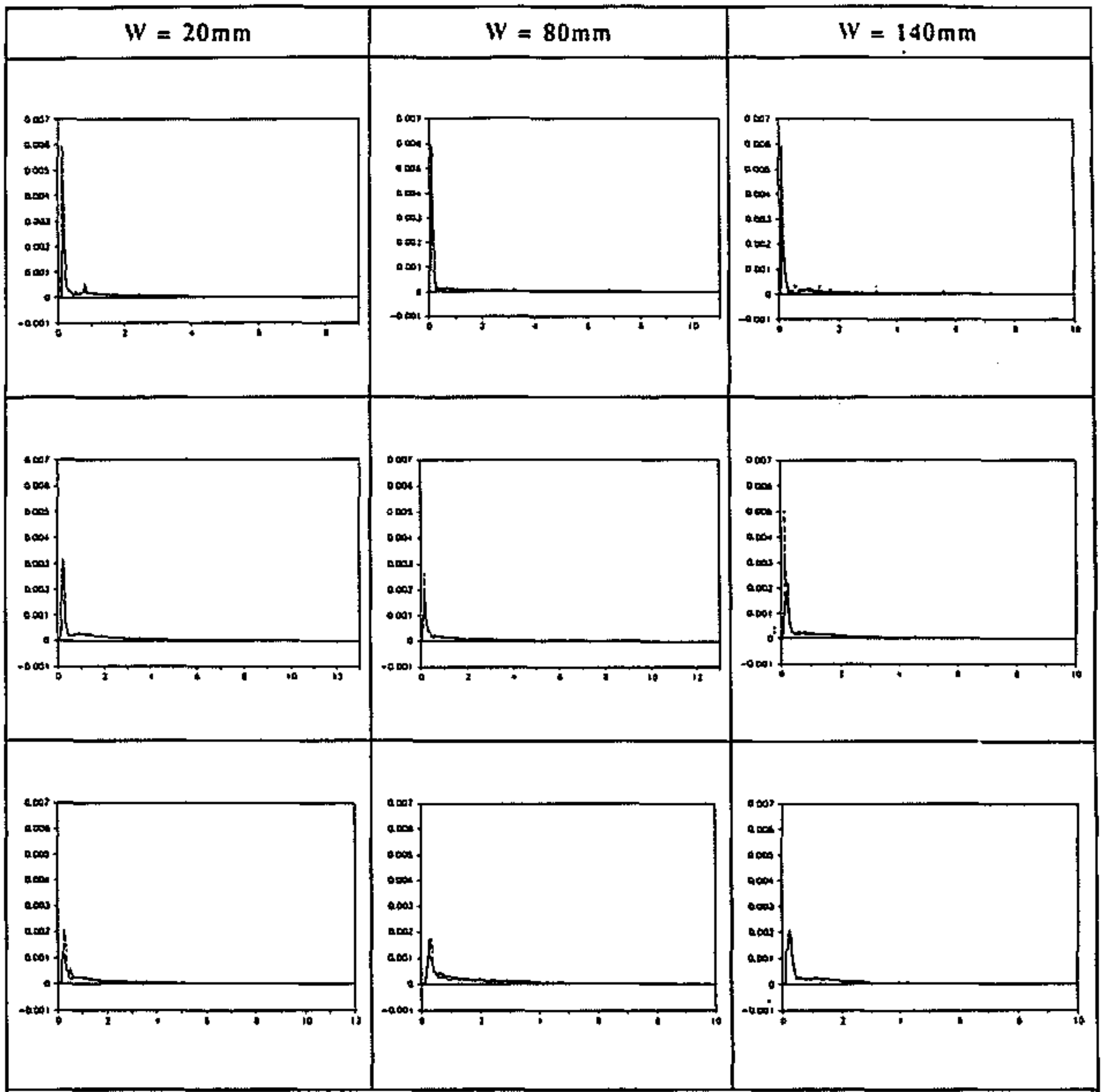


FIGURE 4.11 : C/C_0 vs T/T_0 at length L = 600mm (in Figure 4.10) (X axis = T/T_0 ; Y axis = C/C_0)

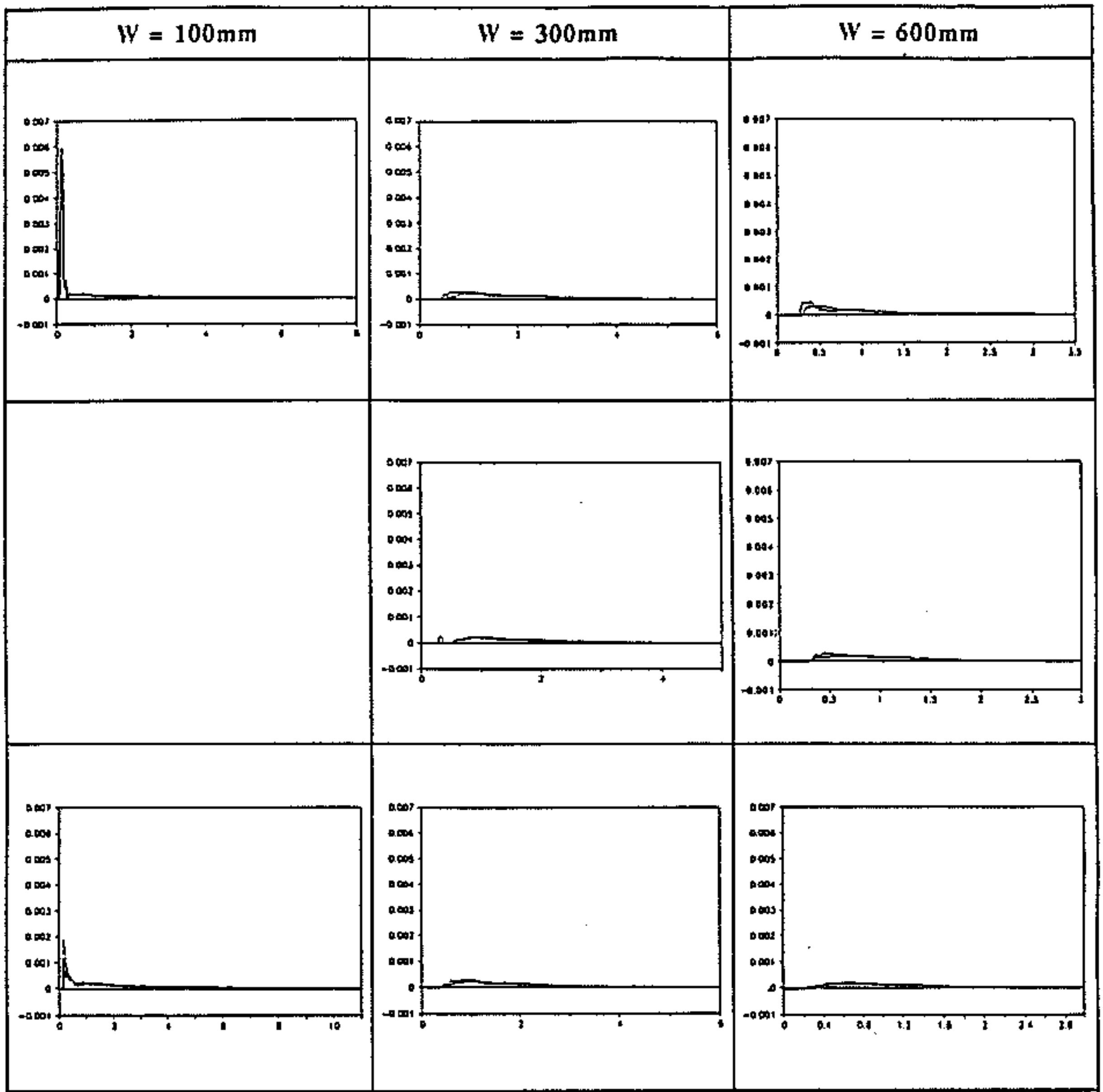


FIGURE 4.12 : C/C_0 vs T/T_0 at length L = 700mm (in Figure 4.10) (X axis = T/T_0 ; Y axis = C/C_0)

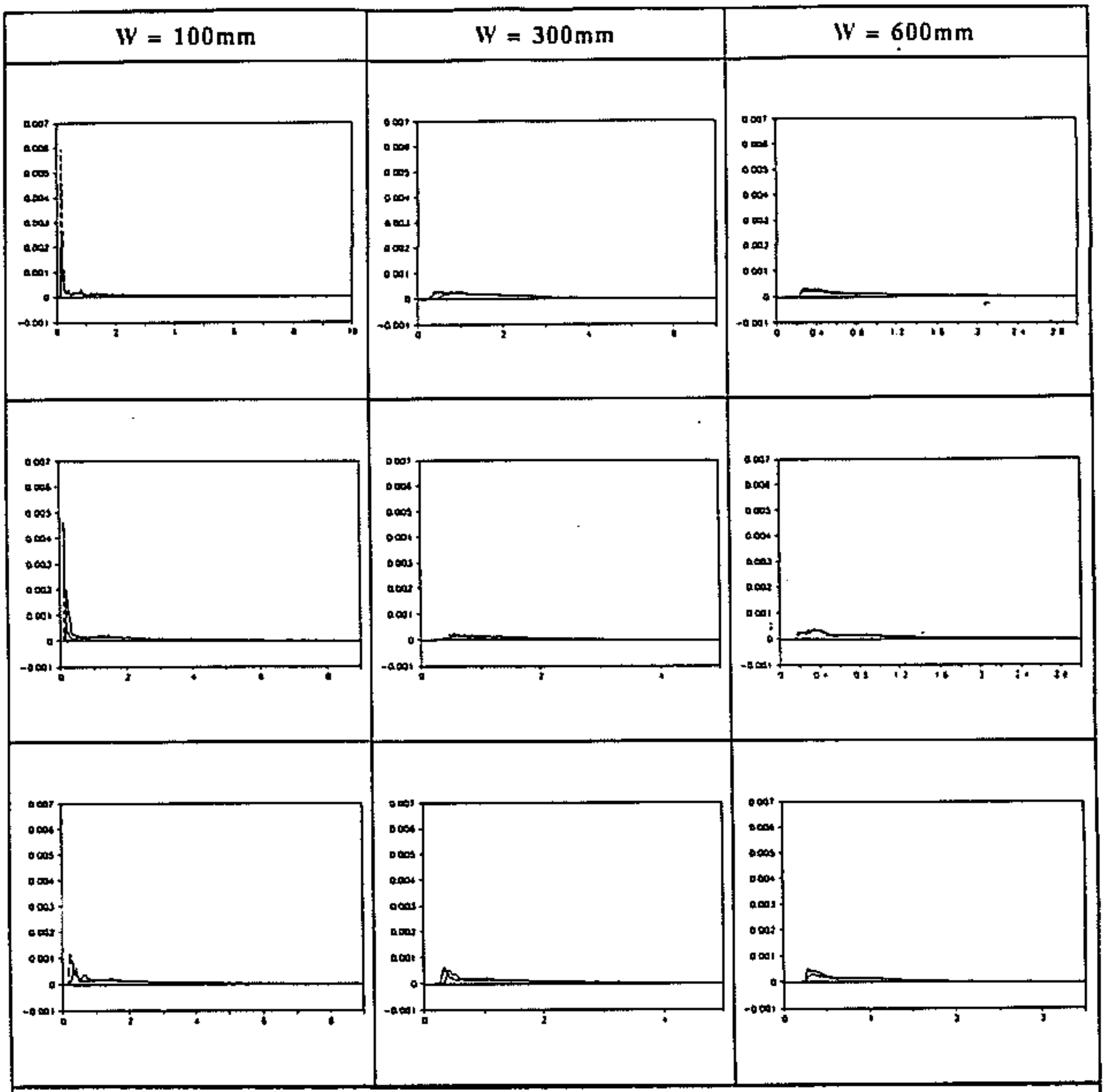


FIGURE 4.13 : C/C_0 vs T/T_0 at length L = 900mm (in Figure 4.10) (X axis = T/T_0 ; Y axis = C/C_0)

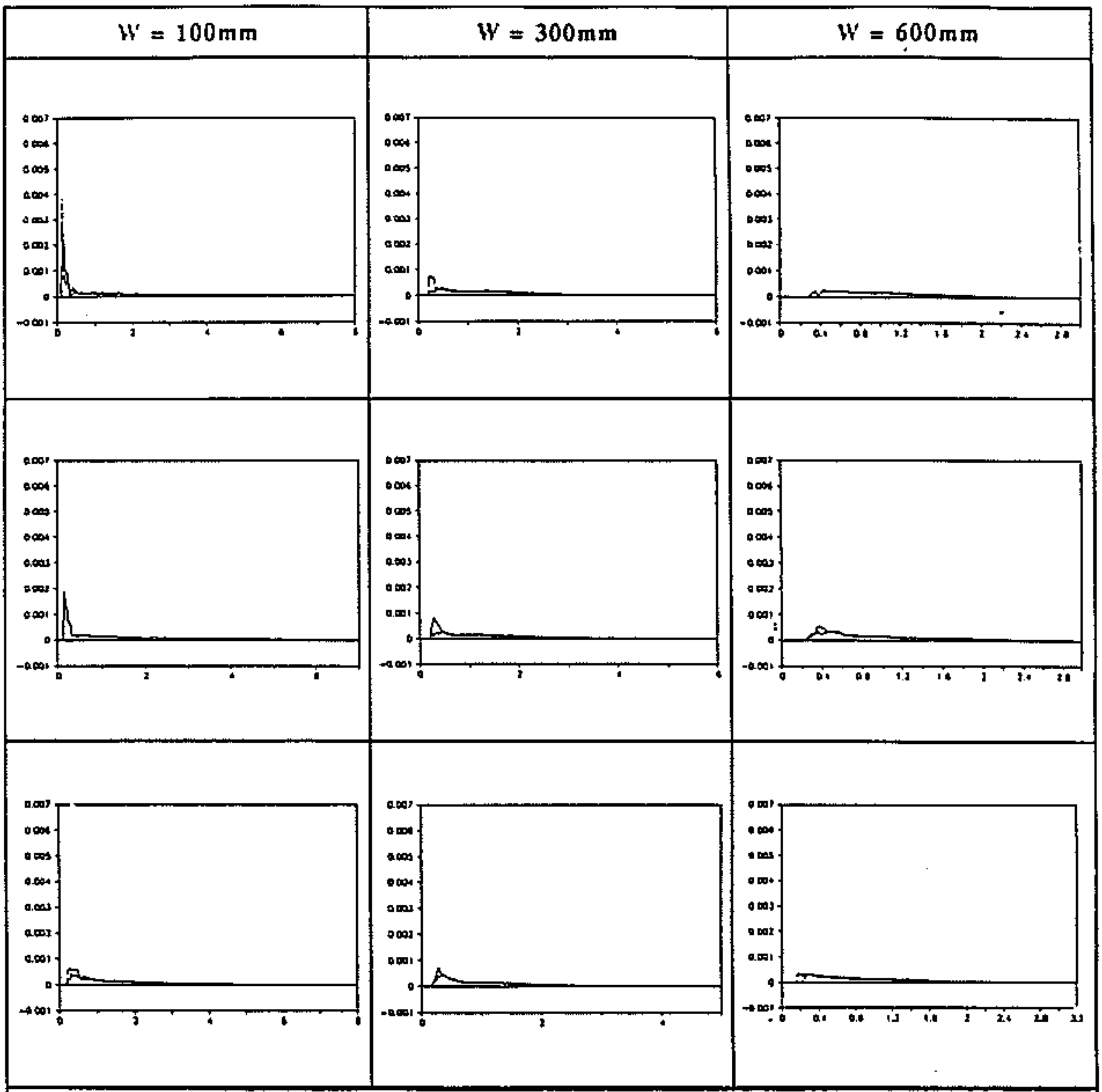


FIGURE 3.14 : C/C_0 vs T/T_0 curves at L = 1050mm (in Figure 4.10) (X axis = T/T_0 ; Y axis = C/C_0)

4.3.3 Residence Time Tests on the Plant Hydraulic Jump.

In order to verify the residence time distributions obtained on the model it is necessary to perform residence time distribution (RTD) tests on the plant. A formal request was made to Umgeni Water to allow me to perform RTD. Permission was granted and RTD tests are currently in progress.