RAPID FILTRATION BY A MULTI-COMPARTMENT UPFLOW AND DOWNFLOW PROCESS

By

Toni Fernàndez
B.Sc. (Hons) Geology
M.Sc. Water Treatment Technology

A thesis submitted in partial fulfilment of the
requirements of The Nottingham Trent University
for the degree of Master of Philosophy

January 2000

DECLARATION

This thesis is composed by myself. The work and the results reported herein are my own except otherwise stated.

No paper has yet been published on elements of the work described in this thesis.

ACKNOWLEDGMENTS

I am deeply indebted to Dr. Ihsan Sholji — First Supervisor, Department of Civil and Structural Engineering, the Nottingham Trent University— for his suggestions on the research and equipment for this work, continuous supervision, invaluable advice and constant encouragement; I would also like to express my sincere apologies to my First Supervisor for so many late meetings and neglected duties. Naturally, I feel grateful to Professor R. K. Hawkins —Head of the Department of Civil and Structural Engineering and Dean of the Faculty of Construction and the Environment— for the opportunity to carry out this research in his department.

Sincere gratitude and thanks are due to Professor K. J. Ives —Second Supervisor, Emeritus Professor of Public Health Engineering, University College, London— who advised on this work and Mr. M. J. Shapley —Third Supervisor, Department of Civil and Structural Engineering, The Nottingham Trent University— for his technical suggestions, contributions and help in general.

Sincere thanks are also due to all the technical staff —Department of Civil and Structural Engineering, The Nottingham Trent University— for their assistance and good sense of humour, particularly Mr. R. C. Gregory (who retired without seeing my thesis completed) and Mr. M. Flanagan for their invaluable help during manufacturing and installing of the experimental set up, and Mr. C. Chamber for his contribution during the course of experimental work. I should like to express my appreciation to all my research colleagues in the department —notably Paul Sharratt— for the support and motivation they gave me during this work. Special thanks to all my friends in Nottingham —in particular Mila Panizo, Anna Espert, Olga López, Anna Traianou, Peter Ovens and Mr. and Mrs. Buxton— for their precious love and encouragement, their immense patience and unfailing moral support of this project and myself, which helped to bring things quicker to completion without pushing everything else off to infinity. As always, I am indebted to them for nearly everything. Many thanks are expressed to Natalie Ulrich, and also my parents and sister Júlia for their love, encouragement, patience and financial support showed during the course of this work.

ABSTRACT

The object of this research project was to investigate the validity and practical applicability of rapid filtration by a multi-compartment up-flow and down-flow process in removing the suspended particles under various initial conditions.

This investigation concentrated on the behaviour of each material used and the characteristics of removal, and efficiency, when various uniform sizes of granular material (548—1086μm) are used and various degrees of concentration of synthetic turbidities are involved.

The net velocity acting upon a suspended particle is a product of hydrodynamic and settling velocities, depending on the direction of flow. These observations were used to monitor the separation of suspended matter in depth in downward and upward flow.

Various types of impurities (kaolin, PVC and Lycopodium powder), inlet concentration (100—400 mg/l) and rates of flow (0.2—1.0 l/min) were investigated to determine their impact on the removal efficiency. At nearly all filtration rates an effluent quality of less than 1 mg/l was obtained, thus effectively achieving a removal efficiency of over 99 percent in most of the experiments, which it could probably include the elimination of *Cryptosporidium* oocysts and *Giardia lamblia* cysts.

The observed results are expected to be a contribution towards a thorough understanding and further analysis of the up-flow and down-flow rapid filtration in the field of water treatment.

RESEARCH OBJECTIVES

The object of this research project was to investigate the validity and practical applicability of rapid filtration by a multi-compartment up-flow and down-flow process in removing the suspended particles under various initial conditions.

Due to the nature of this investigation, only basic filtration parameters were determined in order to assess filter's performance, these being headloss, turbidity readings, removal efficiency, concentration ratio, and water temperature.

Particular attention was given to the behaviour of the system during up-flow filtration, as very little information was found on the topic. The closest commercial example is the OFSYS filter system developed by Culligan Italiana Company, Bologna, Italy, consisting of two successive pressure filters. The first is up-flow coarse media, the second is down-flow fine media (Ives, 2000, oral communication).

The AWWA recommends an operating level of no more than 0.5 NTU of turbidity in filter plant effluent and a goal of no more than 0.2 NTU.

Therefore, based on filter's efficiency and nature of water to be treated, the potential application of such filter would be as follows:

- Direct filtration for water supply purposes of low-turbidity waters
- Polishing filter in either water or wastewater treatment.
- Conventional rapid filtration

DEDICATION

To
My Parents

LIST OF CONTENTS

CHAPTER 1 INTRODUCTION

CHAPTER 3 APPARATUS, MATERIALS, EXPERIMENTAL PROCEDURES AND OBSERVATIONS

CHAPTER 4 ANALYSIS OF THE EXPERIMENTAL WORK

CHAPTER 5 DISCUSSION

APPENDICES

LIST OF TABLES

LIST OF FIGURES

NOTATION

SYMBOL	DESCRIPTION	UNITS
A	Plan area of filter	m^2
b	Sembi and Ives' bulk factor	-
C	Hazen's coefficient of compactness	-
C	Concentration of suspended solids	vol/vol
C_D	Newton's drag coefficient	-
C_0	Inlet concentration of suspended solids	vol/vol
d	Grain size	m
d_{eq}	Diameter equivalent	m
d_g, D_g	Grain size— geometric mean diameter	m
D_h	Particles hydraulic diameter	m
d_{10}	Effective size grain diameter	m
f	Friction factor	-
f_σ	Ives' self-porosity of deposited solids	-
g	Gravitational acceleration	m/s^2
G	Velocity gradient	s^{-1}
Ga	Galileo number	kg/ms
H	Headloss	m
h_L	Headloss through bed of particles of uniform grain size	m
H_0	Headloss of clean filter	m
k	Fair and Hatch's filtration constant	-
K	Headloss factor	-
K_2	Sholji's dynamic shape coefficient	-
K_3	Sholji's experimental shape coefficient	-

L	Depth of the filter bed	m
L_e	Depth of the expanded filter bed	m
P	Power input	N·m/s
Δp	Pressure across a fluidized bed	N
Q	Volumetric flow rate	m^3/s
r	Sembi and Ives' ratio of headloss at any time to initial headloss	-
R	Reynolds number	-
RE	Filter's removal efficiency	-
S	Fair and Hatch's shape factor	-
T	Hazen's water temperature	°F
t	Filter run time	s
V	Volume of mixing basin	m^3
v_a	Superficial or approach velocity	m/s
v_b	Face velocity of the backwash water	m/s
v_h	Hazen's superficial velocity	m/s
v_i	Interstitial velocity	m/s
v_{mf}	Minimum fluidizing velocity	m/s
v_s	Particle settling velocity	m/s
x_{ij}	Weight fraction between adjacent sieve sizes	-
μ	Viscosity of water	$N \cdot m/s^2$
ϵ	Bed porosity	dimless
ϵ_0	Initial bed porosity	dimless
ϕ	Particle shape factor	dimless
ρ_w	Density of water	kg/m^3
ρ_s	Density of solids	kg/m^3
ρ_b	Bulk density	kg/m^3

γ	Kinematic viscosity of water	m^2/s
λ	Filter coefficient	m^{-1}
λ_0	Initial filter coefficient	m^{-1}
σ	Specific deposit	vol/vol
σ_u	Ives' ultimate saturation value of specific deposit	dimless
σ_m	Sembi and Ives' absolute specific deposit	mass/vol
β, x, y, z	Ives' empirically derived factors	dimless
ψ	Grains sphericity	dimless

CHAPTER 1

INTRODUCTION

1.1 CHARACTERISTICS OF WATER AND WASTEWATER

Water, perhaps because it is common in nature, seems a simple material. It is, however, quite complex in a chemical sense. The water molecule is not symmetric, and its unbalanced nature gives a polar character which allows individual molecules to join by hydrogen bonding, forming arrays which are in a constant state of flux. This hydrogen bonding and polar character of water explain its action as a nearly universal solvent. Some compounds such as sugars and alcohols are dissolved by hydrogen bonding. Others such as salts, which are ionized, are dissolved through the neutralization of ions of opposite charge by clusters of oriented water molecules (McGhee, 1991). But pure water does not exist in nature, it normally contains varying amounts of other materials in concentrations ranging from a few milligrams per litre in rain to about 35,000 mg/l in seawater. Wastewater usually contains most of the constituents of the water supply to the area with additional impurities from the waste producing process (Tebbutt, 1983).

The characteristics of water and wastewater are classified as physical, chemical and biological. Domestic wastewaters are not usually as complex as industrial wastewaters, where specific toxic and hazardous compounds may exist, i.e. phenols and toxic organics (Kiely, 1997). *Domestic* or *sanitary wastewater* refers to liquid discharge from residences, business buildings and institutions. *Industrial wastewater* is discharge from manufacturing plants. *Municipal* or *urban wastewater* is the general term applied to the liquid collected in sanitary sewers and treated in a municipal plant. Additionally, interceptor sewers direct dry weather flow from combined sewers to treatment, and unwanted infiltration and inflow enters the collector pipes. Storm runoff water can be collected in a separate storm sewer system — and conveyed to the nearest watercourse for discharge without treatment— or in the same piping together with sanitary wastes (Hammer, 1986).

Water supply sources may be surface waters *or* groundwaters. The surface water and groundwater resources of an area typically are closely related and are interconnected by the hydrologic cycle. Atmospheric water vapour condenses and falls to the earth as some form of precipitation. Once on the earth´s surface, water flows into streams, lakes, and eventually the oceans, or percolates through the soil and into aquifers that eventually discharge into surface waters. Through evaporation from surface waters or by evapotranspiration from plants, water molecules return to the atmosphere to repeat the cycle. Although the movement through some parts of the cycle may be relatively rapid, complete recycling of groundwater must often be measured in geologic time (Peavy *et al.*, 1985). Surface *water* is the term used to describe water on the land surface. It may be running, such as streams and rivers, or quiescent, such as in lakes or reservoirs. Surface water is produced by runoff of precipitation and by groundwater seepage (Reinert *et al.*, 1990*)*.

All water beneath the land surface is referred to as *underground water*, and its quality is influenced by the quality of its source. Groundwaters are often superior in quality to surface waters, but it must be noted that they may be contaminated by toxic or hazardous materials leaking from landfills, waste treatment sites, or other sources (some natural) which may not be known either to the public or regulatory agencies. To a certain extent, anything with which water comes in contact will be dissolved in or mixed with the flow. Pollutants, once introduced into groundwater, may be carried away for very long distances and will be very difficult to remove, although natural processes such as adsorption, biodegradation, radioactive decay, ion exchange, and dispersion may reduce concentrations to some extent (McGhee, 1991).

1.1.1 PHYSICAL CHARACTERISTICS

Physical characteristics of any kind water are those properties that respond to the senses of sight, touch, taste, or smell. Suspended solids (SS), turbidity, colour, taste and odour, and temperature fall into this category (Peavy *et al.*, 1985).

Solids suspended in water may consist of inorganic or organic particles or of immiscible liquids. Inorganic solids (such as clay, silt and other soil constituents) and organic material (such as plant fibres and biological solids) are both common in surface water. Because of the filtering capacity of the soil, suspended material is seldom a constituent of groundwater. Domestic wastewater usually contains large quantities of suspended solids that are mostly organic in nature. Industrial use of water may result in a wide variety of suspended impurities of either organic or inorganic nature. Immiscible liquids such as oils and greases are often constituents of wastewater (Peavy *et al.*, 1985). Suspended solids are, in any case, discrete particles which can be measured by filtering a sample through a fine filter paper (Tebbutt, 1983), and the results of this suspended solids test are expressed as dry mass per unit volume (mg/l).

Turbidity is a measure of the extent to which light is either absorbed or scattered by suspended material in water, but it is not a direct quantitative measurement of suspended solids. Turbidity meter readings are normally expressed as *formazin turbidity units* (FTU). The term *nephelometry turbidity units* (NTU) used in this investigation indicates that the tests were run according to the scattering of light principle. Most turbidity in surface waters results from the erosion of colloidal material such as clay, silt, rock fragments, and metal oxides from the soil. Vegetable fibres and microorganisms may also contribute to turbidity. Domestic and industrial wastewaters may contain a wide variety of turbidity-producing material, such as soaps, detergents, and emulsifying agents (Peavy *et al.*, 1985).

Pure water is colourless, but water in nature is often coloured by foreign substances. Water whose colour is partly due to suspended matter is said to have *apparent colour*. Colour contributed by dissolved solids that remain after removal of suspended matter is known as *true colour*. There are several methods of colour measurement, which normally involve comparison with standardized coloured materials. Results are expressed in *true colour units* (TCU) where one unit is equivalent to the colour produced by 1 mg/l of platinum in the form of chlorplatinate ions. Colour in natural waters is usually originated after contact with organic debris such as leaves, weeds or wood (yellowish-brown hues caused by tannins, humic acid and humates) or inorganic compounds such as iron oxides (reddish tones) or manganese oxides (brown or blackish colour). Industrial wastes may add substantial colouration to water in receiving streams (Peavy *et al.*, 1985).

Many substances with which water comes into contact in nature or during human use may impart perceptible *taste and odour.* This is due to dissolved impurities, often organic in nature, e.g. phenols and chlorophenols, but also minerals, metals, and salts from the soil, and constituents from wastewater. These are subjective properties, which are difficult to measure. Quantitative tests that employ the human senses of taste and smell can be used for this purpose.

Temperature is basically important for its effect on other properties, e.g. speeding up of chemical reactions, reduction in solubility of gases, amplification of tastes and odours, etc. (Tebbutt, 1983). Temperature also affects other physical properties of water: the viscosity of water increases with decreasing temperature. The maximum density of water occurs at 4°C, and density decreases on either side of that temperature, a unique phenomenon among liquids. The use of water for dissipation of waste heat in industry and the subsequent discharge of the heated water may result in dramatic, though perhaps localized, temperature changes in receiving streams (Peavy *et al.*, 1985).

1.1.2 CHEMICAL CHARACTERISTICS

Chemical parameters are related to the solvent capabilities of water and tend to be more specific in nature and are thus more immediately useful in assessing the properties of a sample. Total dissolved solids (TDS), pH, alkalinity, hardness, oxidation-reduction potential (ORP), dissolved oxygen (DO), oxygen demand (biochemical or BOD, and chemical or COD), nitrogen, phosphorus, chloride, metals and some specific organic compounds are chemical parameters of concern when dealing with water quality.

Total dissolved solids (TDS) is defined as all the matter that remains as residue upon evaporation at 103 to 105°C. Total solids, or residue upon evaporation, can be further classified as *non-filterable* (or *suspended*) or *filterable* by passing a known volume of water through a filter. The filterable-solids fraction consists of *colloidal* (size range from 0.001 to 1 μm) and *dissolved solids* (Tchobanoglous and Burton, 1991). Dissolved material results from the solvent action of water on solids, liquids, and gases. Referring to their origin, dissolved solids may be organic —such as decay products of vegetation, organic chemicals and gases— or inorganic, like minerals, metals and gases. Many dissolved substances may produce aesthetically displeasing colour, tastes, and odours. Some chemicals may be toxic or even carcinogenic. The TDS content is expressed as milligrams per litre on a dry-mass basis. An estimated analysis for TDS is often made by determining the electrical conductivity of the water, and for dilute solutions is approximately proportional to the TDS content (Peavy *et al.*, 1985).

pH is the measurement of the concentration of hydrogen ions present in any solution, and tells us about the intensity of acidity or alkalinity of a sample on the pH scale, from 0 to 14 with 7 as neutrality, below 7 being acid and above 7 being alkaline. Biological activity is usually restricted to a pH range of 6-8. Highly acidic or highly alkaline waters are undesirable because of corrosion hazards and possible difficulties in treatment (Tebbutt, 1983).

Alkalinity is defined as the quantity of ions in water that will react to neutralize hydrogen ions, in other words, it is the ability of water to neutralize acids (H.S. Peavy *et al.*, 1985). Most of the natural alkalinity in waters is due to the presence of carbonate (CO_3^{2-}), bicarbonate (HCO_3^-), or hydroxide (OH^-), produced by the action of groundwater on limestone or chalk. Alkalinity is useful in waters and wastewaters in that it provides buffering to resist changes in pH (Tebbutt, 1983). Alkalinity is measured by titrating the water with an acid and determining the hydrogen equivalent. The amount of alkalinity is then expressed in terms of milligrams per litre of $CaCO_3$.

Hardness is the property of a water, due to bicarbonate to form $CaCO_3$, which prevents lather formation with soap and produces scale in hot-water systems. Hardness may be then defined as the concentration of multivalent metallic cations in solution. At supersaturated conditions, the hardness cations will react with anions in the water to form a solid precipitate. Hardness is classified as *carbonate hardness* —metals associated with HCO_3^-, CO_3^{2-} + CO_2 — and *non-carbonate hardness* —metals associated with SO_4^{2-}, Cl^-, NO_3^-. The hardness that is equivalent to the alkalinity is termed carbonate hardness, with any remaining hardness being called non-carbonate hardness (Peavy *et al.*, 1985). Analysis for hardness is commonly made on natural waters and on waters intended for potable supplies and for certain industrial uses. Hardness can be measured by using spectrophotometric techniques or chemical titration to determine the quantity of calcium and magnesium ions in a given sample, and it is expressed as milligrams per litre of $CaCO_3$.

Oxidation-reduction potential (ORP). In any system undergoing oxidation there is a continual change in the ratio between the materials in the reduced form and those in the oxidized form. Many reactions of interest in wastewater treatment such as organic oxidation and methane fermentation, nitrification, and denitrification are of this type and are mediated by bacteria (Sawyer *et al.*, 1994). Operational experience has established ORP values likely to be critical for various oxidation reactions. Aerobic reactions show ORP values of >+200 mV, anaerobic reactions occur below +50 mV (Tebbutt, 1983).

Dissolved Oxygen (DO). In liquid wastes, DO is the factor that determines whether the biological changes are brought about by aerobic or by anaerobic organisms. The former use free oxygen for oxidation of organic and inorganic matter and produce innocuous end products, whereas the latter bring about such oxidations through the reduction of certain inorganic salts such as sulphates, and the end products are often very obnoxious (Sawyer *et al.*, 1994). Unfortunately oxygen is only slightly soluble in water (14.6 mg/l at 0°C and 9.1 mg/l at 20°C at sea level). Clean surface waters are normally saturated with DO, but such DO can be rapidly removed by the *oxygen demand* of organic wastes. Organic compounds are generally unstable and may be oxidized biologically or chemically to stable end products such as CO_2, NO_3, H_2O. *Biochemical Oxygen Demand* (BOD), is a measure of the oxygen required by micro-organisms whilst breaking down organic matter. *Chemical oxygen demand* (COD), is chemical oxidation using boiling potassium dichromate and concentrated sulphuric acid (Tebbutt, 1983).

Nitrogen is a constituent of proteins, chlorophyll, and many other biological compounds (Peavy *et al.*, 1985). It can exist in seven oxidation states. However, in aquatic systems only four of these forms are of interest to water quality:

- *Organic nitrogen* (as proteins, amino acids and urea)
- *Ammonia nitrogen* (as ammonium salts and free ammonia, it is oxidized by autotrophic nitrifying bacteria or *Nitrosomonas* group)
- *Nitrite nitrogen* (an intermediate oxidation stage not normally present in large amounts, it is oxidized by the *Nitrobacter* group of nitrifying bacteria)
- *Nitrate nitrogen* (final oxidation product of nitrogen).

Analyses for nitrogen in its various forms have been performed on potable and polluted waters ever since water was confirmed to be a vehicle for the transmission of disease. The determinations served as one basis of judging the sanitary quality for a great many years (Sawyer *et al.*, 1994). A water containing high organic nitrogen and ammonia nitrogen would be considered unsafe because of recent pollution. Waters in which most of the nitrogen was in the form of nitrates were considered safe as nitrification had already occurred and therefore offered little threat to the public health (Tebbutt, 1983).

Phosphorus appears exclusively as phosphate (PO_4^{3-}) in aquatic environments. Sources of phosphate are fertilizers, animal wastes, municipal wastewater —such as detergents, body wastes and food residues— and industrial wastes. Phosphates are not toxic and do no represent a direct health threat to humans or other organisms, but they do represent a serious indirect threat to water quality. Phosphate is a limiting nutrient in surface waters, and when its concentration increases so it does the growth of the aquatic plants. They are measured colorimetrically. Results of the analysis are reported as milligrams per litre of phosphate as phosphorus (Peavy *et al.*, 1985).

Chloride occurs in all natural waters in a widely varying concentration but it can also be an indicator of sewage pollution because of the chloride content of urine. Chloride in a reasonable concentration is not harmful to humans. At concentrations above 250 mg/l it gives a salty taste to water, although up to 1,500 mg/l it is considered to be safe. Chlorides are usually determined by means of volumetric procedures employing internal indicators. For most purposes the Mohr method employing silver nitrate as the titrant and potassium chromate as the indicator is satisfactory (Sawyer *et al.*, 1994).

All *metals* are soluble to some extent in water. Sources of metals in natural waters include dissolution from natural deposits and discharges of domestic, industrial or agricultural wastewaters. There are two basic types of metals: those that are harmful to humans and other organisms in relatively small amounts or *toxic metals* —such as arsenic, barium, cadmium, chromium, lead, mercury, iron, aluminium, copper, zinc, and silver— and those that do not present health hazards or *non-toxic metals*, such as sodium and manganese. However, copper and zinc are synergetic and when both are present, even in small quantities, may be toxic to many biological species. Measurement of metals in water is usually made by atomic absorption spectrophotometry (Peavy *et al.*, 1985).

Many *organic materials* are soluble in water. Most organics in natural water systems consist of the decay products of organic solids, while synthetic organics are usually the result of wastewater discharges or agricultural practices. Dissolved organics in water can be divided into two categories: *biodegradable* —such as starches, fats, proteins, alcohols, acids, aldehydes, and esters— and *non-biodegradable*, such as tannic and lignic acids, cellulose, phenols, some polysaccharides, and benzene. Biodegradable material consists of organics that can be utilized for food by naturally occurring microorganisms within a reasonable length of time. It is usually quantified by the biological oxygen demand (BOD) test. Most natural water and municipal wastewaters will have a population of microorganisms that will consume the organics. In sterile waters, microorganisms must be added and the BOD of the material containing the organisms must be determined and subtracted from the total BOD of the mixture. The presence of toxic materials in the water will invalidate the BOD results. Non-biodegradable organics are those that are resistant to biological degradation. They can be measured by the chemical oxygen demand (COD) test or the total organic carbon (TOC) analysis. Both COD and TOC measure the biodegradable fraction of the organics, so the BOD result must be subtracted from the COD or TOC to quantify the non-biodegradable organics (Peavy *et al.*, 1985).

1.1.3 BIOLOGICAL CHARACTERISTICS

Almost all organic wastes contain large numbers of microorganisms, sewage containing over 10^6 /ml, but the actual numbers present are not often determined. After conventional sewage treatment the effluent still contains large numbers of microorganisms as do many natural surface waters (Tebbutt, 1983). Based on their known tolerance for a given pollutant, certain organisms can be used as indicators of the presence of pollutants. Standards for microbiological quality are essentially based on the need to ensure that microorganisms indicative of pollution by human wastes are absent. Pathogens are those organisms not native to aquatic systems and that usually require an animal host for growth and reproduction. They are microorganisms capable of infecting, or of transmitting diseases to, humans. Many species of pathogens are able to survive in water and maintain their infectious capabilities for significant periods of time. These waterborne pathogens include species of *bacteria* (responsible for terrible diseases such as cholera and typhoid), *viruses* (they can cause poliomyelitis and infectious hepatitis), *protozoa* (giardiasis and other gastrointestinal disorders), and *helminths* (parasitic worms). Analysis of water for all the known pathogens would be a very time-consuming and expensive proposition. The purity of water is normally checked using indicator organisms. These organisms belong to the fecal coliform group and the most important one is *Escherichia coli*. It is a non-pathogenic organism and is believed to have a longer survival time outside the animal body than do most pathogens. It is the usual practice in the European Union and the United States to use the total coliform group (those of both fecal and non-fecal origin) as indicators of the sanitary quality of drinking water, while the indicator of choice for wastewater effluents is the fecal coliform group. The membrane-filter technique gives a direct count of coliform bacteria. The results are reported in number of organisms per 100 ml of water —after having been incubated at the right temperature for 24 h (Peavy *et al.*, 1985).

1.1.4 TYPICAL CHARACTERISTICS

It is not really possible to give details of what might be termed normal characteristics for a particular sample. As a guide, however, Table 1.1 gives typical analyses for major pollutant of a raw domestic wastewater (Kiely, 1997). Table 1.2 gives analyses for various types of water source and Table 1.3 is representative for domestic sewage analyses before and after treatment (Tebbutt, 1983).

Table 1.1 Typical major pollutant characteristics of a raw domestic wastewater

Parameter Type	Parameter	Concentration by phase (225 l/h/d)			
		Total Load (kg/h/d)	**Soluble (mg/l)**	**Particulate (mg/l)**	**Total (mg/l)**
Physical	Suspended Solids				
	Volatile	~80%			240
	Inert	~20%			60
	Total	~0.07 kg/h/d			300
	Dissolved Solids				
	Volatile	~40%			175
	Inert	~60%			265
	Total	~0.10 kg/h/d			440
	Temperature				10-20°C
	Colour				Fresh— grey Old— black
Chemical	BOD_5	~0.06 kg/h/d	(30%) 65	(70%) 135	250
	COD	~0.11 kg/h/d	130	370	500
	TOC				160
	Total Nitrogen	0.01 kg/h/d	25	15	40
	Organic N				15
	Free Ammonia				25
	Nitrites				0
	Nitrates				0
	Total Phosphorus	0.002 kg/h/d	5	4	9
	Organic				4
	Inorganic				6
	Alkalinity				100
	Fats, Oils and Grease				100
Microbiological	Total Coliforms				100-1000 million MPN/l
	Faecal Coliforms				10-100 million MPN/l
	Total Viruses				1000-10000 infectious units/l

After G. Kiely, 1997

Table 1.2 Characteristics of various water sources

Characteristic figure	Source		
	Upland catchment	Lowland river	Chalk aquifer
pH (units)	6.0	7.5	7.2
Total solids (mg/l)	50	400	300
Conductivity (μS/cm)	45	700	600
Chloride (mg/l)	10	50	25
Alkalinity, total (mg/l)	20	175	110
Hardness, total (mg/l)	10	200	200
Colour (°H)	70	40	<5
Turbidity (NTU)	5	50	<5
Amm. -N (mg/l)	0.05	0.5	0.05
NO_3 -N (mg/l)	0.1	2.0	0.5
DO (percent saturation)	100	75	2
BOD	2	4	2
22°C colonies/ml	100	30 000	10
37°C colonies/ml	10	5000	5
Coliform MPN/100 ml	20	20 000	5

After T.H.Y. Tebbutt, 1983

Table 1.3 Typical sewage analyses

Characteristic mg/l	Source		
	Crude	Settled	Final effluent
BOD	300	175	20
COD	700	400	90
TOC	200	90	30
SS	400	200	30
Amm.-N	40	40	5
NO_3 -N	<1	<1	20

After T. H. Y. Tebbutt, 1983

1.2 GENERAL INTRODUCTION TO WATER TREATMENT PROCESSES

The objective of municipal water treatment is to provide a water supply —one that is chemically and microbiologically safe for human consumption. Quality requirements for industrial uses are often more rigid than for domestic supplies. Thus, additional treatment may be required by the industry (Hammer, 1986).

Contaminants may be present as *floating* or *large suspended solids* —such as leaves and branches—, *small suspended and colloidal solids* —clay and silt particles, microorganisms—, *dissolved solids* —alkalinity, hardness, organic acids— , *dissolved gases* —carbon dioxide, hydrogen sulphide— or *immiscible liquids*, such as oils and greases (Tebbutt, 1983).

Common water sources for municipal supplies are deep wells, shallow wells, rivers, natural lakes, and reservoirs. Underground water is normally cool, uncontaminated and presents a uniform quality that is easily processed for municipal use. Only chlorination is normally required in order to provide both disinfection and residual protection against potential contamination in the water distribution system. The usual process in surface-water treatment is chemical clarification by coagulation, flocculation, sedimentation and filtration, followed by a final disinfection (Hammer, 1986).

There are three main classes of treatment processes:

- *Physical processes*— screening, sedimentation, filtration, gas transfer.
- *Chemical processes*— coagulation, flocculation, precipitation, ion exchange.
- *Biological processes*— aerobic and anaerobic biological filtration and activated sludge, etc.

Figure 1.1 shows typical operational ranges of various treatment processes. Flow diagram and typical design criteria for conventional water treatment plant are given in Figure 1.2 (Tebbutt, 1983).

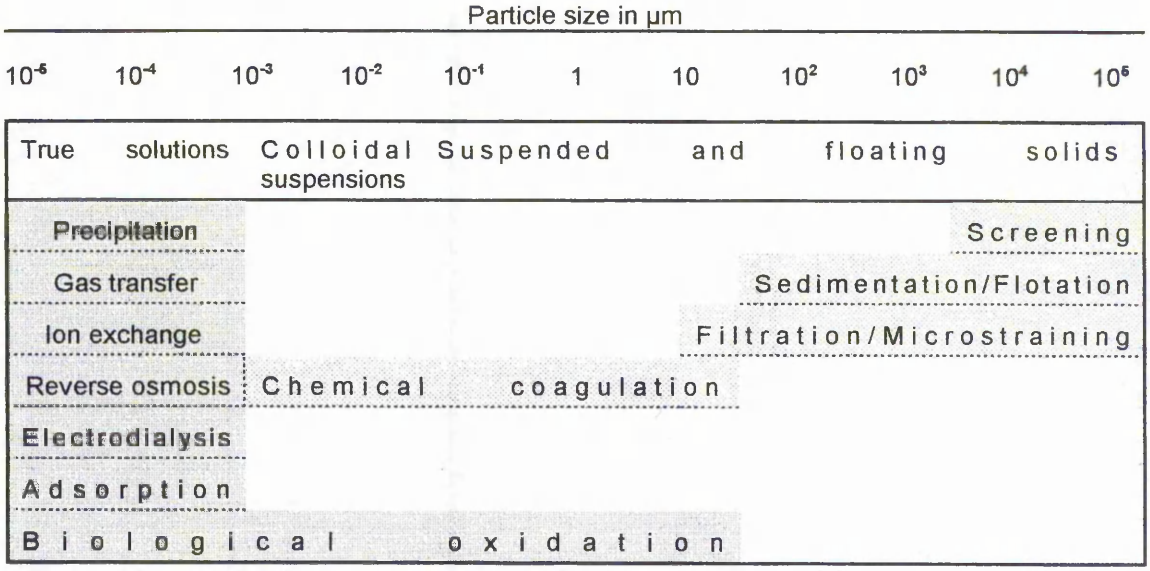


Fig 1.1 Applications of the main treatment processes, after T. H. Y. Tebbutt, 1983.

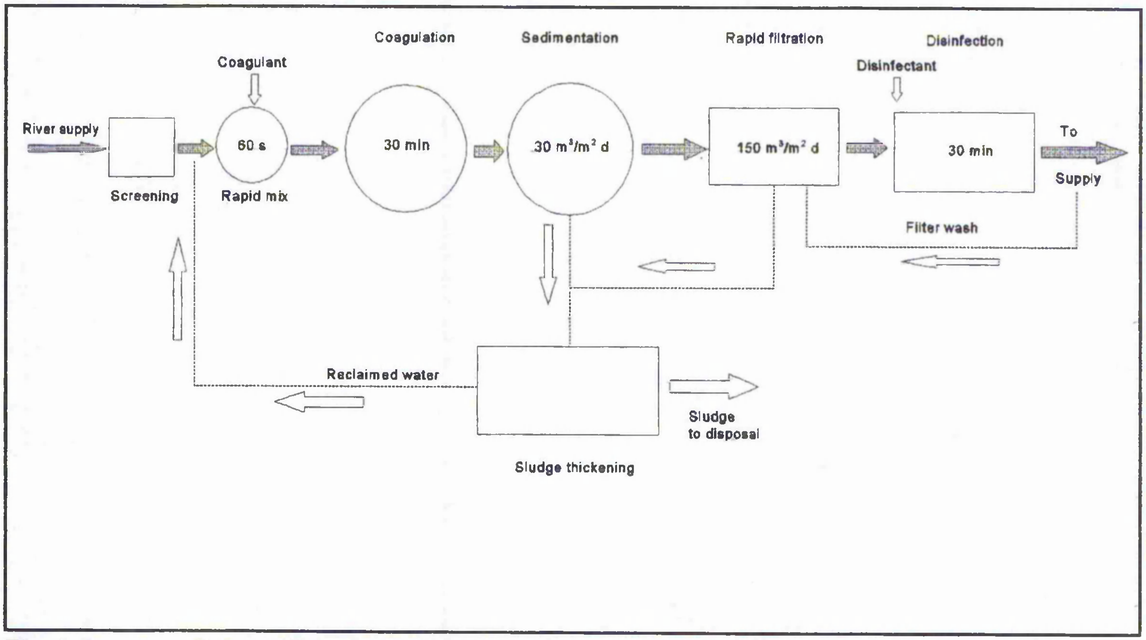


Fig 1.2 Schematic pattern of typical surface water treatment system, after T. H. Y. Tebbutt, 1983.

1.2.1 PRELIMINARY TREATMENT PROCESSES

River water frequently contains suspended and floating debris varying in size from logs to small rags. These solids can clog and damage pumps or impede the hydraulic flow in open channels and pipes.

Screening is the first step in treating water containing large solids (Viessman, Jr. and Hammer, 1993). Thus, some form of protective boom or coarse screen with openings of about 75 mm is used to prevent large objects reaching the intake. The main screens are usually provided in the form of a mesh with openings of 5-20 mm and arranged as a continuous belt, a disc or a rotating drum through which the flow must pass (Tebbutt, 1983). If storage is not provided, fine screens are fitted after the coarse screens. If there is storage then fine screens are placed at the outlet of the storage tanks. Fine screens are typically mesh with openings about 6 mm diameter or square (Kiely, 1997).

Microstraining is employed for removal of algae and similar-size particles from waters of otherwise good quality. It has also applications in wastewater treatment as a final tertiary stage to produce a high-quality sewage effluent. Because of the small mesh apertures (20-60 µm), clogging occurs rapidly so that the drum is rotated at a peripheral speed of about 0.5 m/s and the mesh continually washed clean by high-pressure sprays. Straining rates in normal usage are 750-2500 m^3/m^2d (Tebbutt, 1983).

Aeration is the supply of oxygen from the atmosphere to water to effect beneficial changes in the quality of the water. It is a common treatment process for groundwater and less common for surface waters. It usually is a simple mechanical process of spraying water into the air and allowing it to fall over a series of cascades. Aeration is used to release excess H_2S gas —which may cause undesirable tastes and odours—,and CO_2 — corrosive effect on concrete materials— and also to increase the O_2 content of water in order to improve its physical characteristics and to remove the presence of iron and manganese (Kiely, 1997).

1.2.2 COAGULATION — FLOCCULATION

Raw water, after screening, has impurities in suspension and solution. Particulate matter in suspension has a particle size range of 10^{-7} to 10^{-1} mm. Inorganic clay colloids range in size from 10^{-6} to 10^{-3} mm and form the dominant component of particulates in suspension. The minor component is that of organic colloids or microorganisms. Because of the very small size, the suspended matter has negligible settling velocity, whether it is organic or inorganic (Kiely, 1997). Their removal can be achieved by promoting agglomeration of such particles by flocculation with or without the use of a coagulant followed by sedimentation or flotation.

Inorganic material is usually easier to remove and the efficiency of removal can be measured without much difficulty using conventional laboratory analytical techniques and equipment. For example, inorganic matter in suspension can be measured before and after chemical treatment using parameters of turbidity and suspended solids, and metals such as iron, aluminium, zinc and lead can all be measured quickly using spectophotometric techniques. The more usual type of inorganic material encountered in water treatment settles easily, especially when it is of particulate size; when plain settling is used one or two hours retention will generally remove at least 40 to 50 % of particulate matter, and treatment with a coagulant such as aluminium sulphate or an iron salt will remove at least 99 % of the remainder (Twort *et al.*, 1994).

The coagulation process normally utilizes what is known as a chemical coagulant (aluminium or iron salt) which reacts with alkalinity in the water to produce an insoluble metal hydroxide floc which incorporates the colloidal particles. This fine precipitate is then flocculated to produce settleable solids. Before flocculation can take place it is essential to disperse the coagulant, usually required in doses of 30-100 mg/l, throughout the body of water. This is carried out in a rapid mixing chamber (normally between 30-120 s) with a high-speed turbine or by adding the coagulant at a point of hydraulic turbulence, e.g. at a hydraulic jump in a measuring flume (Tebbutt, 1983). Most common is mechanical mixing using a vertical-shaft impeller in a tank with stator baffles. The turbulent flow pattern in mixing is a function of the tank size and shape, number, shape and size of impellers, kind of location of stator baffles, water temperature and power input (Viessman, Jr. and Hammer, 1993).

The complexity of the coagulation process is evident as all the following factors affect it in diverse ways:

- coagulant dosage
- pH
- colloid concentration —often measured by turbidity
- TOC and COD or colour
- anions and cations in solution
- mixing effects
- electrophoretic mobility or zeta potential
- temperature

The optimum dosages of coagulants for a specific water are easily and effectively determined using the simple *jar test*, which simulates the processes of coagulation, flocculation and sedimentation in a laboratory batch test. The jar test may be used for the following:

- coagulant selection
- optimum dosage selection, which corresponds to the lowest residual turbidity measured after the jar test has been performed
- coagulant aid selection
- determination of optimum pH
- determination of point of addition of pH adjustment chemicals and coagulant aids

- optimisation of mixing energy and time for rapid mixing and slow mixing
- determination of dilution of coagulant and other similar measurement

In addition to residual turbidity in jar tests other parameters such as zeta potential, streaming current and particle size analysis may be used to supplement the data from jar tests as strategies for coagulation control (Amirtharajah and O'Melia, 1990).

Flocculation is a slow mixing process in which the destabilized particles and chemical precipitates resulting from coagulation are brought into contact in order to promote their agglomeration into larger heavier floc that settle out by gravity (Mc Ghee, 1991). Flocculation is a principal mechanism in removing turbidity from water. Floc growth depends primarily on two factors:

- intermolecular chemical forces
- physical action induced by agitation

Thus, the flocculation process relies on turbulence to promote collisions. The common mechanical mixing devices in water treatment are paddle (reel) flocculators, flat-blade turbines and vertical-turbine mixers.

Design parameters for rapid-mix units are mixing time *t* and velocity gradient *G*. The velocity gradient is a measure of the relative velocity of two particles of fluid and the distance between. A more useful concept of velocity gradient, however, is given in terms of power dissipation per unit volume as shown in equation [1-1] (Peavy, *et al.*, 1985):

$$G=\left(\frac{P}{V\mu}\right)^{½} \qquad [1.1]$$

where $G=$ velocity gradient, s^{-1}
$P=$ power input, W ($N \cdot m/s$, J/s)
$V=$ volume of mixing basin, m^3
$\mu=$ viscosity of water, $N{\cdot}s/m^2$

Optimum values exist for flocculation *t* and *G*. Below a minimum time, no flocculation occurs, and increasing time beyond maximum floc formation does not significantly improve flocculation. Based on experience, flocculator tanks with a minimum of three compartments in series reduce the time required for flocculation (Viessman, Jr. and Hammer, 1993). For the common coagulants of aluminium and iron salts the value of the velocity gradient is usually in the range of 20 to 75 s^{-1} with retention times in flocculation chambers varying from about 10 to 60 minutes (Twort *et al.*, 1994).

Large *G* values with short times tend to produce small, dense flocs, while low *G* values and long times produce larger, lighter flocs. Since large, dense flocs are more easily removed in the settling basin, it may be advantageous to vary the *G* values over the length of the flocculation basin (Peavy, *et al.*, 1985).

1.2.3 SEDIMENTATION — FLOTATION

Sedimentation and flotation are solid-liquid separation processes used in water treatment to lower the solids concentration, or load, on granular filters. As a result, filters can be operated more easily and cost effectively to produce acceptable-quality filtered water. Many sedimentation and flotation processes exist, and each has advantages and disadvantages. The most appropriate process for a particular application will depend on the water to be treated as well as local circumstances and requirements (Gregory and Zabel, 1990).

Flotation can be described as a gravity separation process —mainly used in water treatment— in which gas bubbles attach to solid particles to cause the apparent density of the bubble-solid agglomerates to float to the surface. The float is removed from the surface, and clarified water is taken from the bottom of the flotation tank. Different methods of producing gas bubbles give rise to two main types of flotation processes (Gregory and Zabel, 1990):

- *Electrolytic flotation* —generation of bubbles of hydrogen and oxygen in a dilute aqueous solution by passing a direct current between two electrodes.
- *Dissolved-air flotation* —bubbles are produced by the reduction in pressure of a water stream saturated with air.

Sedimentation (clarification) is defined as the removal of solid particles from suspension by gravity. In water treatment, the common application of sedimentation is after chemical treatment to remove flocculated impurities and precipitates (Viessman and Hammer, 1993). Usually these particles are denser than the surrounding liquid so that sedimentation takes place but with very small particles and with low-density particles flotation may offer a more satisfactory clarification process. Sedimentation units have a dual role —the removal of settleable solids and the concentration of the removed solids into a smaller volume of sludge (Tebbutt, 1983).

Some types of settlement tanks are in use for the clarification of flocculated waters. For large volumes of water containing a relatively heavy load of suspended solids a relatively dense floc is formed which settles easily, and in warm climates where the viscosity of water is lower, so permitting more rapid settlement of floc, the large horizontal —or *cross-flow*— sedimentation tank can be an economical solution for clarification (Twort *et al*, 1994). They tend to have a length-width ratio of about 2 and a depth of the order of 1.5 to 6 m. A sludge draw-off well is located at the upstream base, and the sludge is drawn to this by a travelling scraper board. Circular settling tanks´ dimensions typically are 10 to 50 m in diameter and 2.5 to 6 m in depth. Water enters to the central well either at the top or up through a central pipe. As the influent water settles, it spreads out and a sludge scraper moves the sludge towards a central sludge withdrawal hopper at bed level. The clarified water exits over a weir along a perimeter of the tank at surface level. The key parameters and typical values in the design of settling tanks are:

- Surface overflow rate —20-35 m^3/m^2 day
- Detention times —2-8 h
- Weir overflow rate —150-300 m^3/m day

The above values vary depending on whether the water treated is raw water for potable treatment or coagulated raw water (Kiely, 1997). Design of clarifiers is based on empirical data from the performance of full-scale sedimentation tanks. Mathematical relationships to predict settling of suspensions in actual treatment processes have not been successful in design, and the application of data from laboratory settling column tests has met only limited success (Viessman and Hammer, 1993).

Sedimentation may be classified into various types depending upon the characteristics and concentrations of suspended materials:

- *Discrete particles* —particles whose size, shape and specific gravity do not change with time.

- *Flocculating particles*—particles whose surface properties are such that they aggregate, or coalesce, with other particles upon contact, thus changing size, shape and perhaps specific gravity with each contact.

- *Dilute suspensions*—suspensions in which the concentration of particles is not sufficient to cause significant displacement of water as they settle or in which particles will not be close enough for velocity field interference to occur.

- *Concentrated suspensions*—suspensions in which the concentration of particles is too great to meet the above conditions.

These differences result in significantly different settling patterns and require separate analysis (Peavy *et al.,*1985). The basic theory of sedimentation assumes the presence of discrete particles. When such a particle is placed in a liquid of lower density it will accelerate until a limiting terminal velocity is reached, then the gravitational force will be equal to that of the frictional drag force. When dealing with flocculating suspensions it is not possible to apply the above principle because the agglomeration of floc particles results in increased settling velocity with depth due to the formation of larger and heavier particles. Many of the suspensions in the treatment of water are flocculent in nature. In the case of concentrated suspensions (> 2000 mg/l SS) hindered settlement occurs. In these circumstances there is a significant upward displacement of water due to the settling particles and this has the effect of reducing the apparent settling velocity of the particles (Tebbutt, 1983).

1.2.4 FILTRATION

Filtration is the process of passing water through a porous medium with the expectation that the filtrate has a better quality than the influent. The medium is usually sand but other granular materials can also be utilized, such as garnet or anthracite (Kiely, 1997). The granular-media gravity filter is the most common type used in water treatment to remove nonsettleable floc remaining after chemical coagulation and sedimentation (Hammer, 1986).

There are two main types of filtration in depth systems used in water treatment (*see Section 1.4.1 Types of Filters*):

- *Slow sand filtration* — or *cake filtration*, with filtration rates ranging from 0.1 to 0.3 m/h, mainly applied to waters which are very low in turbidity and thus require no pretreatment. This type of filter has been shown to improve the microbiological water quality, with removal rates of up to 99.9 per cent for *Giardia* and *Cryptosporidium* cysts.

- *Rapid filtration* — or *depth filtration*, with rates from about 5 to 20 m/h, it can be used to treat high-turbidity waters, generally after the proper application of coagulants to destabilize colloidal suspensions.

However, the fundamentals of water filtration will be discussed in greater detail in section 1,4 and in later chapters.

1.2.5 DISINFECTION

The other treatment procedures like coagulation and filtration should remove more than 90 per cent of bacteria and viruses. However, to meet the standards given by the European Union Water Standard Directives, and to provide protection against regrowth, additional disinfection has often been practised (Kiely, 1997). Thus, *disinfection* can be defined as the killing of potentially harmful micro-organisms and not as the complete destruction of all living matter (*sterilisation*).

Viruses and protozoal cysts are more resistant to disinfectants than bacteria and need additional exposure time and higher concentrations. Turbidity producing colloids, and iron and manganese deposits can shield organisms and use up the disinfectant. Chlorine (and its compounds) is by far the most commonly used disinfectant. It is a very strong oxidizing agent and, when added to water in the molecular form, will react with both organic and inorganic materials which are present. The efficiency of chlorine is adversely affected by high pH and low temperature. Dosages of 0.25 to 0.5 mg/l are usually adequate to maintain a residual level of chlorine in the reservoir. Higher dosages or subsequent addition within the distribution system may be required to maintain residual protection until the water reaches the user. Other methods of disinfection include the use of ozone —known to effectively reduce the *Cryptosporidium* cysts to non infective numbers— and UV radiation, both especially effective in killing viruses (Ghee, 1991). Since no residual remains, it is necessary to use small amounts of chlorine after ozonization or radiation to provide continuous protection against regrowth in the distribution system.

1.3 GENERAL INTRODUCTION TO WASTEWATER TREATMENT PROCESSES

The purpose of municipal wastewater treatment is to prevent pollution of the receiving watercourse. Characteristics of a municipal wastewater depend to a considerable extent on the type of sewer collection system and industrial wastes entering the sewers. The degree of treatment required is determined by the beneficial uses of the receiving stream or lake, such as water supply or recreation.

An industry has three possibilities for disposal of process wastewaters:

- separate treatment in an industrial waste treatment plant prior to discharge to a watercourse
- complete treatment of industrial wastewaters in a municipal treatment plant
- pretreatment at the industrial site prior to discharge in the municipal sewerage system

Wastewaters from households, industries and combined sewers are collected and transported to the treatment plant with the effluent commonly disposed of by dilution in rivers, lakes or estuaries. Other means of disposal include irrigation, evaporation from lagoons and submarine outfalls extending into the sea (Viessman, Jr. and Hammer, 1993).

Conventional wastewater treatment is a combination of physical and biological processes designed to remove organic matter from solution. The overall process can be viewed as thickening; pollutants removed from solution are concentrated in a small volume convenient for ultimate disposal (Figure 1.3).

The daily contribution of raw domestic wastewater is about 200-250 l/person with a total solids content of less than 0.1 per cent, 240 mg/l SS and 200 mg/l BOD. Liquid waste sludge withdrawn from primary and secondary processing amounts to approximately 2 l/person with a solids content of 5 percent by weight. This can be further concentrated to a handleable material by mechanical dewatering; the extracted water is returned for reprocessing. Cake from dewatering amounts to about 0.25 l/person with a 30 percent solids concentration. Ultimate disposal of dewatered solids may be landfill, incineration or land application if biologically stabilized.

Referring to Table 1.3, 50 percent of total volatile solids, 60 percent of total nitrogen and 70 percent of total phosphorus remain in the effluent after secondary biological treatment. Advanced wastewater treatment processes are needed to remove these refractory contaminants (Hammer, 1986).

Wastewater treatment processes can be summarized as follows:

- Preliminary Processes —include screening to remove large solids, grit removal to protect mechanical equipment against abrasive wear, flow measuring and pumping to lift the wastewater above ground.

- *Primary Treatment* —settleable organic matter (amounting to 30 to 50 percent of SS) and floatable materials removal.

- *Secondary Treatment* —aeration in open basins with return biological solids, or fixed-media filters, followed by final settling.

- *Tertiary or Advanced Treatment* —removal of refractory pollutants such as phosphorus or nitrogen by activated carbon filtration or biological nitrification and denitrification.

Table 1.3 Approximate composition of average domestic wastewater*

Parameter	Raw	After Settling	Biologically Treated
Total Solids	800	680	530
Total Volatile Solids	440	340	220
Suspended Solids	240	120	30
Volatile Suspended Solids	180	100	20
Biochemical Oxygen Demand	200	130	30
Inorganic Nitrogen as N	15	15	24
Total Nitrogen as N	35	30	26
Soluble Phosphorus as P	7	7	7
Total Phosphorus as P	10	9	8

(*) Expressed in mg/l and based on 450 l/person · day. After M. J. Hammer, 1986.

1.3.1 PRELIMINARY TREATMENT PROCESSES

By definition, preliminary treatment is the process or processes that prepare a wastewater to a condition that it can be further treated in conventional primary and secondary treatment processes. In municipal wastewater it means the removal of floating debris and grit and the removal of oily scums. These pollutants would inhibit the biological process and possibly damage mechanical equipment. Ideal influent parameters for municipal activated sludge, the principal biological treatment processes are in the range 100 to 400 mg/l BOD and SS (Kiely, 1997).

The following preliminary processes are used in municipal wastewater treatment:

- coarse screening (*bar racks*)
- medium screening
- flow measuring

- equalization (flow, organics, nutrients and pH balancing)
- pumping
- grit removal
- pre-aeration

Although not common in pretreatment, flotation, flocculation and chemical treatment are sometimes dictated by the industrial pollutants in the municipal wastewater. *Flotation* is used to remove fine suspensions, grease and fats, and is performed either in a separate unit or in a pre-aeration tank also used for grit removal. If adequate pretreatment is provided by petroleum industries and meat-processing plants, flotation units are not required at a municipal facility. *Flocculation,* with or without chemical additions, may be practised on high-strength municipal wastewaters to provide increased primary removal and prevent excessive loads on the secondary treatment processes. *Chlorination* of raw wastewater is sometimes used for odour control and to improve settling characteristics of the wastes (Viessman, Jr. and Hammer, 1993).

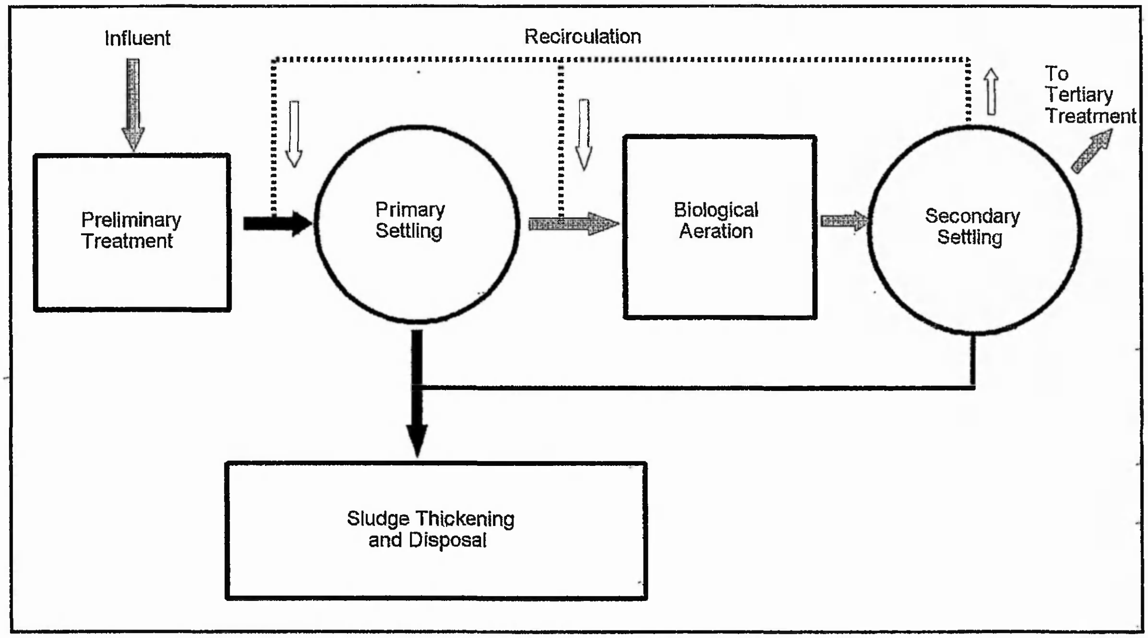


Fig 1.3 Schematic diagram of conventional wastewater treatment, after W. Viessman, Jr. and Hammer, 1993.

1.3.2 PRIMARY TREATMENT

Primary treatment is often called clarification, sedimentation or settling. This is the unit processes where the wastewater is allowed to settle for a period of about two hours in a settling tank and so produce a somewhat clarified liquid effluent in one stream and a liquid-solid sludge —primary sludge— in a second stream. The objective is to produce a liquid effluent of suitably improved quality for the next treatment stage and to achieve a solids separation resulting in a primary sludge that can be conveniently treated and disposed of (Kiely, 1997).

Clarification is performed in rectangular, square or circular basins where the wastewater is held quiescent to permit particulate solids to settle out of suspension (Hammer, 1986). These primary clarifiers are usually designed to remove particles with settling rates of 0.3 to 0.7 mm/s. Plants operated within that design range normally provide suspended solids (SS) removals from 30 to 60 percent, depending in part on the original concentration and the age of the sewage. Since the solids are largely organic, their removal also results in a reduction in biological oxygen demand (BOD). Retention times in primary clarifiers are generally short, from 1 to 2 h at peak flow. Combining this criterion with a surface overflow rate of 0.3 to 0.7 mm/s yields a depth of 1 to 5 m. Recent research indicates that great depth in primary clarifiers is not helpful and that shallower basins function best, subject to limitations imposed by the need to install mechanical equipment for sludge removal (Mc Ghee, 1991).

The addition of coagulant chemicals (iron salts, lime, alum) before primary sedimentation promotes flocculation of fine suspended matter into more readily settleable flocs. This increases the efficiency very substantially of SS and BOD removal rates, over a surface overflow rate range of 20 to 80 m^3/m^2day (Heinke and Tay, 1980). A disadvantage of chemical coagulants is an increase in primary sludge which is a chemical-type sludge quite different to the biological sludge, from primary sedimentation. The mechanism of chemically enhanced primary sedimentation is to use an aeration tank prior to the settling tank. The chemicals are flash-mixed with the flow and added to the aeration tank (Kiely, 1997). If their dosage is sufficiently large, significant quantities of phosphorus may also be precipitated (Mc Ghee, 1991).

1.3.3 SECONDARY TREATMENT

The effluent from primary treatment still contains 40 to 50 percent of the original SS and virtually all of the original dissolved organics and inorganics. This organic removal, referred to as *secondary treatment*, may consist of chemical-physical processes or biological processes. Combinations of chemical-physical operations such as coagulation, microscreening, filtration, chemical oxidation, carbon adsorption and other processes can be used to remove the solids and reduce the BOD to meet the minimum European Union standards for discharge. Biological processes are used in practically all municipal wastewater-treatment systems where secondary treatment is employed (Peavy *at al.,* 1985).

Biological treatment consists of application of a controlled natural process in which microorganisms remove soluble and colloidal organic material from the waste and are, in turn, removed themselves (Mc Ghee, 1991). This is achieved by using activated-sludge processes, trickling filters or rotating biological contactors.

In the activated-sludge method wastewater is fed continuously into an aerated tank, where microorganisms consume the organics. The resulting microbial floc —or *activated sludge*— is settled from the aerated mixed liquor under quiescent conditions in a final clarifier (solid-liquid separator) and returned to the aeration tank or biological reactor (Figure 1.4). The plant effluent is clear supernatant from secondary settling. Advantages of liquid aeration are high-BOD removals, ability to treat high-strength wastewater and adaptability for future use in plant conversion to advanced treatment.

Trickling filters and rotating biological contactors have media to support microbial films. These slime growths extract organics from the wastewater as it trickles over the surfaces. Oxygen is supplied from air moving through voids in the media. Excessive biological growth washes out and is collected in the secondary clarifier (Viessman, Jr. and Hammer, 1993). In northern climates fibre-glass covers have been employed to prevent ice formation since efficiency in high-rate filters is reduced by approximately 30 percent per 10°C, and freezing may cause partial plugging of the filter medium and overloading of the remaining open area.

Clarifiers following ordinary trickling filters are designed to remove relatively large particles of sloughed bacterial slime or humus. Surface overflow rates are 25 to 35 m/day at average flow and should not exceed 50 m/day at peak flow. Weir loading rates and retention times are similar to those used in primary clarifiers (Mc Ghee, 1991).

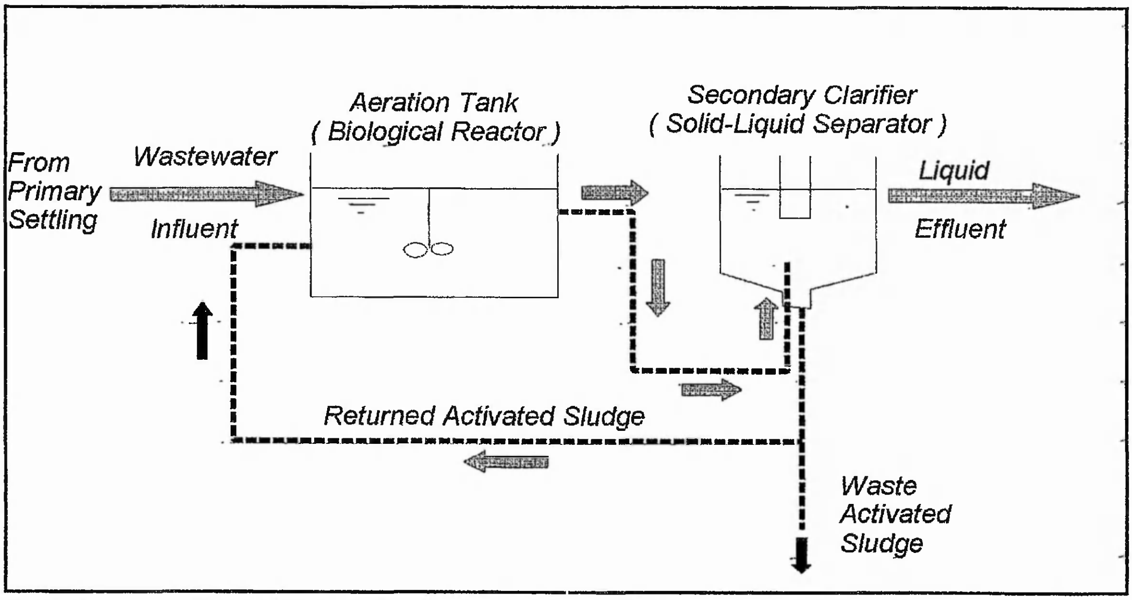


Fig 1.4 Typical layout of activated sludge system, after G. Kiely, 1997.

1.3.4 TERTIARY TREATMENT

Tertiary treatment refers to methods and processes that remove more contaminants from wastewater than are usually taken out by conventional techniques. Several of these pollutants not removed by secondary biological methods can adversely affect aquatic life in streams, accelerate eutrophication of lakes, hinder use of surface waters for municipal supplies, and restrict direct reuse of wastewater for irrigation, groundwater recharge or other beneficial applications (Viessman and Hammer, 1993).

Secondary treatment achieves effluent standards with BOD values around 20 mg/l SS concentrations around 30 mg/l, 300 mg/l of total solids, 200 mg/l of volatile solids, 20 mg/l of nitrogen and 7 mg/l of phosphorus. Treated water may also contain traces of organic chemicals, heavy metals, excreted pathogens and other contaminants (Viessman Jr. and Hammer, 1993). *Effluent polishing* is the term used to improve effluent quality for SS/BOD from 30:20 to, say, 10:10 (Kiely, 1997). This additional treatment often involves the removal of nitrogen and phosphorus compounds, plant nutrients associated with eutrophication. Further treatment may be required to remove additional SS, dissolved inorganic salts and refractory organics (Peavy *et al.,* 1985).

For effluent polishing and toxics removal the more readily available processes are:

- Granular media filtration —rapid filtration or slow filtration
- Micro-straining
- Activated carbon adsorption
- Upward-flow clarifier
- Land irrigation on grass plots
- Lagoons
- Chemical treatment —for pH correction, precipitation of phosphorus and enhancement of sedimentation
- Ammonia removal by air stripping, or biological nitrification-denitrification
- Chlorination

The most common tertiary treatment processes are filtration, phosphorus precipitation and biological nitrification-denitrification. But the method which finds most application in large works is that of rapid gravity filtration (Tebbutt, 1983). *Filtration* of wastewater is most commonly used for the removal of residual biological floc in settled effluents from secondary treatment before discharge to the receiving waters —i.e. when the effluent SS must be <10 mg/l (Kiely, 1997). Filtration is also used to remove residual precipitates from the metal salt or lime precipitation of phosphates and is used as a pretreatment operation before treated wastewater is discharged to activated-carbon columns. In reuse applications, filtration of treated wastewater is required for application to food crops, park and playground irrigation, and body-contact recreational impoundments (Tchobanoglous *et al.,* 1991).

The majority of phosphorus in wastewater is soluble, therefore phosphorus is only sparingly removed by plain sedimentation. Secondary biological treatment removes phosphorus by biological uptake. As a result, in Table 1.3, the total phosphorus of 10 mg/l in the raw water is reduced to 8 mg/l in the biologically treated effluent. *Chemical precipitation* using aluminum or iron coagulant or lime is common and effective in phosphate removal. The treatment scheme includes mixing, sedimentation and filtration. Recarbonation is required prior to filtration for stabilization of the wastewater. This tertiary lime treatment also precipitates heavy metals, removes some dissolved organic compounds and provides disinfection (Hammer, 1986).

Nitrogen is usually removed from wastewater by biological nitrification and denitrification and, when preceded by secondary treatment, over 90 percent removal of total nitrogen is achieved. *Nitrification* involves the oxidation of ammonia to nitrate, with nitrite as an intermediate. The reactions are carried out by the bacteria *Nitrosomonas* and *Nitrobacter*. *Denitrification* involves the conversion of nitrate to nitrogen gas and a number of facultative heterotrophs use nitrate instead of oxygen as the final electron acceptor during the breakdown of organic matter under anoxic conditions. Because nitrified effluent contains little carbon, a carbon source is normally added. This is frequently methanol because it is almost completely oxidised, thus producing less sludge for disposal (Mason, 1991).

1.4 FLOW THROUGH POROUS MEDIA IN DEPTH

The objective of this section is to review the basic principles determining the removal of particles by rapid gravity filtration. Usually rapid filtration is preceded by chemical treatment of the water; rapid filtration without chemical treatment is effective for relatively few waters, its main use being primary filtration before slow sand filtration (Twort *et al.,* 1994).

Filtration of suspensions through porous media (usually sand) is an important stage of the treatment of potable waters to achieve final clarification. Although about 90 percent of the turbidity and colour are removed in coagulation and sedimentation, a certain amount of floc is carried over from settling tanks and requires removal (Tebbutt, 1983).

Under certain conditions, filtration may serve as the primary turbidity-removal process (e.g. in direct filtration of raw water). Although filtration removes many pathogenic organisms from water, filtration should not be relied upon for complete health protection (Peavy *et al.,* 1985).

Sand filtration is also employed to provide tertiary treatment of 30:20 standard sewage effluents after secondary biological treatment (*polishing filter*). Other uses of flow through porous media include ion-exchange beds and absorption columns where the aim is not to remove suspended matter but to provide contact between two systems (Tebbutt, 1983).

1.4.1 TYPES OF FILTERS

A number of different types of filters are used in water processing, and they may be described in various classification schemes (Cleasby, 1990). The rate of flow through the filter can be constant or variable depending on the flow control method (Tchobanoglous and Burton, 1991).

One *physical classification* scheme is based on the type of granular medium used:

- *Granular-bed filters* —commonly use a substantial depth of sand or anthracite coal or combinations thereof
- *Precoat filters* —use a thin layer of vary fine medium such as diatomaceous earth that is disposed of after each filter cycle.

Filters may also be described by the *hydraulic arrangement* provided to pass water through the medium:

- *Gravity filters* —flow through the medium is achieved by gravity; these are filters open to the atmosphere
- *Pressure filters* —the filter medium is contained in a pressure vessel, water is delivered to the vessel under pressure and leaves the vessel at slightly reduced pressure.

Filters may also be classified by the *rate of filtration*, that is, the flow rate per unit area they can be operated at (their main characteristics are summarized in Table 1.4 and Table 1.5):

- *Rapid sand filters* —single medium (sand or anthracite), dual medium (sand and anthracite) and multimedia (usually garnet, sand and anthracite), length of run of 1 to 5 days

- *Slow sand filters* —operate at a much lower rate that favours surface removal on the top of the sand bed and biological action, length of run of 20 - 90 days

Or according to the *direction of flow*:

- *Down-flow filters* —this is by far the most commonly used type of filtration. The liquid to be filtered is passed downward through the filter bed.

- *Up-flow filters* —the liquid to be filtered flows upward through the filter bed. The principle used is to have progressively finer sand in the direction of flow, which allows the filter to carry a greater load of impurity before backwashing, because the larger particles tend to be held in the lower, coarser part of the filter, leaving the upper parts to deal with the smaller particles (Twort *et al.*, 1994).

- *Down-flow & up-flow filters* —this type of filtration is basically a sequence of up-ward flow filtration followed by down-ward flow filtration. It has often been referred to as *biflow filtration*. Up-flow and biflow filters may be particularly useful in filtering waters containing high suspended solids concentrations (Mc Ghee, 1991).

Finally, filtration can be classified as:

- *Depth filtration* —solids are removed within the granular material, such as in rapid filtration

- *Cake filtration* —solids are removed on the entering face of the granular material (e.g. slow sand filtration).

Table 1.4 Design Criteria for Slow Sand Filters

Parameter	UK Recommended Level
Design Life	10-15 years
Period of Operation	24 h/day
Filtration Rate	0.1-0.2 m/h
Filter Bed Area	0.1-0.5 ha per individual basin*
Height of Filter Bed	
Initial	0.8-0.9 m
Minimum	0.5-0.6 m
Sand Specification Effective Size	0.15-0.3 mm
Uniformity Coefficient	<3
Height of Under drains (including gravel layer)	0.3-0.5 m
Height of Supernatant water	1 m

After Visscher *et al.*, 1987

Table 1.5 Rapid Gravity Filter Media Characteristics

		Parameter			
Media Type	Granular Material	Depth (m)	Effective Size (mm)	Uniformity Coefficient	Filtration Rate (m/h)
Single	Sand	0.7	0.6	<2	10
	or anthracite	0.7	0.7	<2	10
Dual	Anthracite	0.6	1.0	<2	12
	and sand	0.15	0.5	<2	
Multi	Anthracite,	0.5	1.0	<2	15
	sand *and*	0.2	0.5	<2	
	garnet	0.1	0.2	<2	

After G. Kiely, 1997

1.4.2 THEORY OF FILTRATION IN RAPID FILTERS

A rapid filter contains a bed of a coarse medium, such as sand, ranging in depth from 300 mm to several metres. The kinetics of removal of particles smaller in size than the pore openings of the bed is extremely complex and very dependent on proper chemical conditioning (Mc Ghee, 1991).

The suspended matter which is removed from the flow, accumulates within the bed pores, thus the amount of the accumulated deposits must equal the amount of impurities removed from the flowing turbid water, according to the continuity equation. This deposited matter within the pore's structure will influence:

- interstitial velocity
- local porosity
- headloss
- quality of the filtrate.

The principal modes of action of rapid filters are mainly physical and *physico-chemical.* Biological processes are not considered and hardly take place. In rapid filtration the particles to be removed from suspension are smaller than the filter pores (Figure 1.5). It follows then that if particles followed the fluid streamlines, many of them would never have the chance of contact with a grain surface and be removed from the flow.

The removal itself has been described as consisting of a transport and an attachment-detachment process (Mc Ghee, 1991). *Mechanisms of transport* are those that move the particles across the streamlines, allowing the suspended particles to be in contact with the grain surface. When they are nearer to the surface the attractional Van der Waal's forces become dominant, and adhere on the grain surface.

Removed particles from the flowing suspension through the granular medium are influenced by the following transport mechanisms, depending on the fluid flow conditions, the geometry of the filter pores as well as the nature of the particles (size, shape, density, etc.).

Transport mechanisms are straining, interception, inertia, sedimentation, diffusion, hydrodynamic action and orthokinetic flocculation:

Straining may be further classified as *absolute straining* if particles large enough to be strained arrive at the filtration surface they will form a mat and clog it rapidly. This is undesirable, as the depth of the filter will not be used efficiently. Such surface clogging can also take place if the concentration of particles is too high; and *partial straining* this applies to particles penetrating the depth and flocculate within the pore, forming larger suspended particles and strained or accumulated within the pores at a deeper layer.

- *Interception*, this is a valid mechanism even for very small particles. If particles remain in streamlines which approach the grain surface at a distance smaller than a particle radius, the particle will contact with the grain surface. This is characterized by the ratio of the particle diameter to the grain diameter.

- *Inertia*, streamlines approaching a filter grain have to deflect as the flow passes round it. If deflected particles have sufficient inertia they will maintain a trajectory which causes them to collide with the grain's surface, and be removed.

- *Sedimentation* , If the particles are large enough, and have a density significantly greater than that of water, they will have a relative velocity moving them in the direction of gravity, and bringing them in contact with the grain surface due to the gravitational forces.

- *Diffusion*, for particles less than 1 μm the movement becomes increasingly significant with decreasing sizes. This mechanism is observed to impart a random movement to very small particles (less than 1 μm) in water, due to the thermal energy of the water molecules (known as *Brownian motion*).

- *Hydrodynamic action*, the flow in the filter pores is depends on the state of flow, and under steady flow conditions, particles will exhibit an apparently random, drifting motion across the streamlines which may cause them to collide with the grain surfaces.

- *Orthokinetic flocculation*, this is scarcely a mechanism for transporting particles to grain surfaces but could aggregate particles in the filter pores, thus enhancing their probability of removal due to interstitial flocculation enhancing their sizes and promoting further partial straining.

Attachment of the particles after their contact with the medium is chemical in nature and is influenced by the following parameters (Mc Ghee, 1991):

- pH
- ionic composition of the water
- age of the floc
- nature and dosage of the coagulant
- composition and surface condition of the medium

This *attachment* may be produced by the following mechanisms:

- *Electrical double layer interaction*, between surfaces in water the interaction of electrical double layers can lead to an attraction or repulsion, depending on whether the surfaces have electrokinetic (zeta) potentials of unlike or like sign respectively. As sand and other filter media surfaces, and the great majority of particulate impurities in water have negative zeta-potentials, the double layer interaction will usually inhibit attachment. This is greatly dependent on the concentration of dissolved salts. For sewage effluents, or lowland river waters, the range of this interaction will only be of the order of 10 nm. At such small distances, attractive forces of the van der Waal type become significant and may overcome the repulsion.

- *Van der Waal's forces*, this mechanism only applies for very close approach of particles and filter grains (usually less than 50 nm), where the separation distance is much less than the sphere diameter. These universal attractive forces between atoms and molecules are additive, and lead to an interaction of attraction between the grain surfaces and particles in water.

- *Mutual adsorption,* some polymers, such as polyelectrolytes or hydrolysis products of aluminium, can form links and bridges, having one end attached to the grain surface and the other to the particle.

Detachment mechanisms. The usual way of cleaning rapid filters is by reverse flow flushing with water, sometimes preceded or accompanied by scouring with air bubbles. However, even without reverse flow flushing, there is evidence that an increase of flow-rate through a filter will detach particles causing a more turbid filtrate. The intensity of this effect depends not only on the magnitude of the increase of flow rate, but also on the backwashing rate, settling velocity of medium, water temperature and initial bed porosity.

Removal of particulate matter in a rapid sand filter results from the transport and attachment processes described above. Removal tends to occur initially in the upper strata of the bed with particles penetrating deeper as the run progresses. As the interstices of the bed are filled, the superficial velocity of the water increases, resulting in resuspension of some particles, increased pressure loss in the bed and creation of relatively large flow channels in the upper surface. If the headloss becomes too large, a partial vacuum may be created, resulting in the formation of bubbles from gases drawn from solution. This phenomenon, called *air binding*, further restricts the area of flow, increases the velocity and headloss and may cause particles to be carried through the filter (Mc Ghee, 1991).

The operation of rapid filters is affected by a number *of* variables, some of them which are determined at the design stage and others which are significant during operation. These variables are (Ives and Sholji, 1965):

- depth of media
- grain size media
- grain material
- rate of filtration
- inflow concentration of suspended particles
- type of suspension
- water temperature
- initial porosity

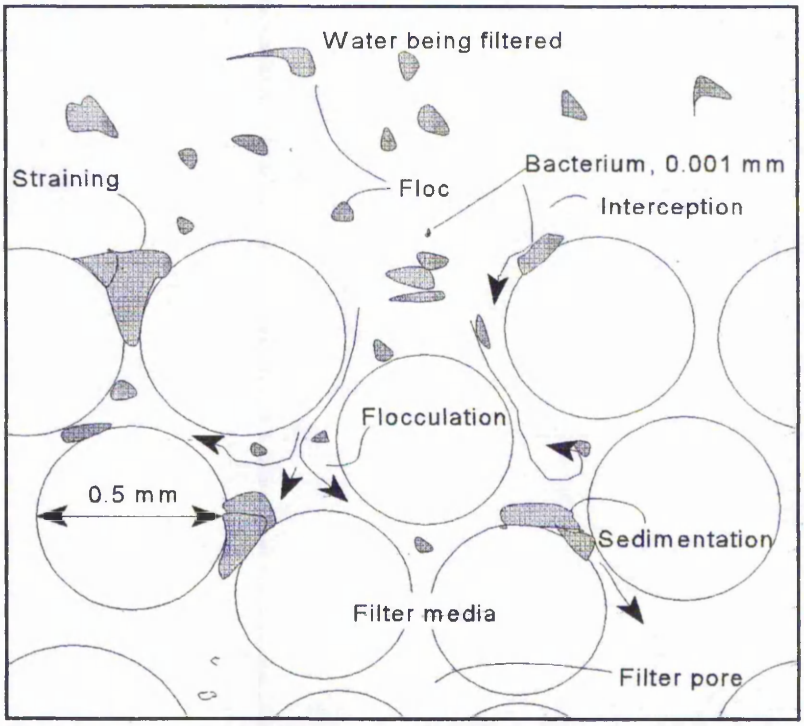


Fig 1.5 Schematic diagram illustrating straining, flocculation and sedimentation actions in a granular-media filter, after Viessman and Hammer, 1993.

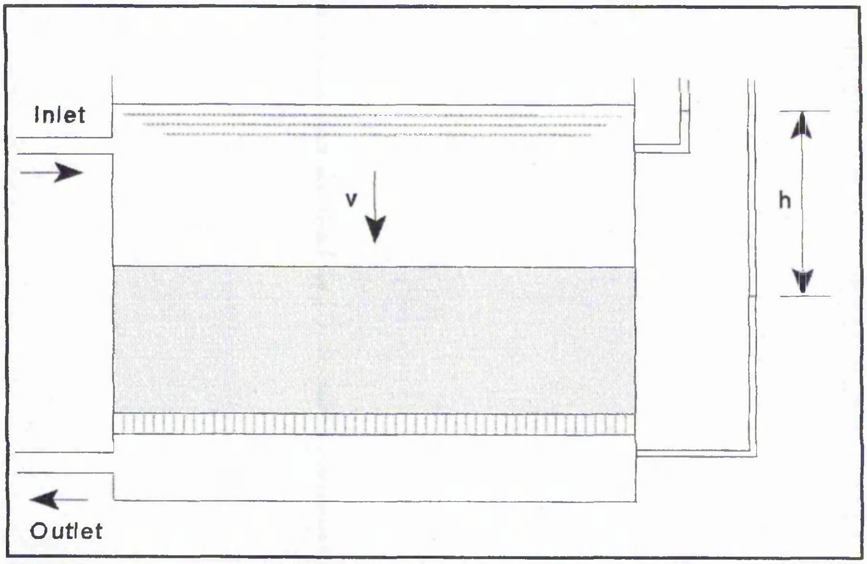


Fig 1.6 Schematic filter, after T. H. Y. Tebbutt, 1983.

1.4.3 HYDRAULICS OF FILTRATION

The basic formulae for the hydraulics of filtration assume a bed of unisize medium and refer to the schematic filter shown in Figure 1.6.

The head loss that occurs when clean water flows through a bed of clean filter medium can be calculated from well known equations. Carmen modified the Darcy — Wiesbach equations for head loss in a pipe to reflect conditions in a bed of porous media of uniform size. The resulting equation, known as the Carmen — Kozeny equation, is:

$$h_L = \frac{fL(1-\epsilon)v_s^2}{\epsilon^3 \phi g d_g} \quad [1.2]$$

where h_L = headloss through bed of particles of uniform grain size, m
L = depth of the filter, m
ϵ = bed porosity, i.e., the volume of voids divided by the total bed volume, dimensionless
v_s = superficial velocity of the water just above the bed (flow rate/bed area), m/s
g = gravitational acceleration (9.8 m/s^2)
d_g = grain size —geometric mean diameter, m
ϕ= particle shape factor (1.0 for spheres, 0.82 for rounded sand, 0.75 for average sand, 0.73 for crushed coal and regular sand)

The remaining term f is a *friction factor* related to the *drag coefficient* C_D around the particles. In the usual range of filter velocities (*laminar flow*) this can be calculated by

$$f=150\frac{(1-\epsilon)}{R}+1.75 \qquad [1.3]$$

where R (Reynolds number)

$$R=\frac{\phi\rho_w v_s d_g}{\mu} \qquad [1.4]$$

ρ_w = water density, kg/m^3
μ = water dynamic viscosity, N·s / m^2

Equation [1.2] can be rewritten for a bed of non-uniform medium as

$$h_s=\frac{L(1-\epsilon)v_s^2}{\epsilon^3\phi g}\sum f\frac{x}{d_g} \qquad [1.5]$$

This equation assumes that the bed is stratified by size and that the *porosity* ϵ is uniform throughout. The *mass fraction* x is determined from a sieve analysis of the medium. The *geometric mean diameter of grains* d_g is the square root of the product of the adjacent sieve sizes d_1 and d_2 (in metres).

Several other formulae have been developed, such as those of Rose, Fair and Hatch or Hazen. Rose used dimensional analysis to develop the equation

$$h_L=1.067C_D\frac{Lv_s^2}{gd_g\phi\epsilon^4} \qquad [1.6]$$

where C_D is Newton's *drag coefficient*

$$C_D=\frac{24}{R}+\frac{3}{\sqrt{R}}+0.34 \qquad [1.7]$$

In the case of a rapid filter it is necessary to take account of the variation of C_D with particle size due to its stratified bed packing nature. Filter beds in water and wastewater are usually graded beds stratified as a result of backwashing with the coarsest grains on the bottom and the finest on top. The grain-size distribution in a bed is defined by a sieve analysis.

For practical purposes, the thicknesses of substantially uniform layers in a stratified bed can be assumed to be proportional to the weights of the portions separated by the sieves. Hence, the total head loss is the sum of the head losses calculated by equation [1.6] for successive layers based on the weight gradation from a sieve analysis of the filter material.

Rose proposed the following equation for calculating head loss through each layer of a stratified filter bed:

$$h_s = 1.067 \frac{L v_s^2}{g \phi \epsilon^4} \sum C_D \frac{x}{d_g} \qquad [1.8]$$

Similarly, Fair and Hatch developed analogous equations for calculating head loss in a homogeneous granular-media bed [1.9] or in a stratified filter bed [1.10]:

$$h_h = k \gamma S^2 \frac{(1-\epsilon)^2 L v_s}{\epsilon^3 g d^2} \qquad [1.9]$$

$$h_s = k \gamma \frac{(1-\epsilon)^2 L v_s}{\epsilon^3 g} \left(\frac{6}{\phi}\right)^2 \sum \frac{x}{d_g^2} \qquad [1.10]$$

where $\gamma =$ kinematic viscosity of water, m^2/s

$S =$ shape factor, varies between 6.0 for spherical particles to 8.5 for crushed materials, dimensionless

$k =$ filtration constant, 5 based on sieve openings, 6 based on size of separation, dimensionless

Finally, Hazen proposed the following formula for calculating the clear water head loss through a granular medium (Tchobanoglous *et al.*, 1985):

$$h_h = \frac{60 \gamma L v_s}{C T + 10 g d_{10}^2} \qquad [1.11]$$

where v_s= superficial velocity, m/s

γ= kinematic viscosity of water, m^2/s

C= coefficient of compactness, varies from 600 for very closely packed sands that are not quite clean to 1200 for very uniform clean sand, dimensionless

T = water temperature, °F

d_{10}= effective size grain diameter, m

1.4.4 FILTER MEDIA

The ideal filter media should be coarse enough for large pore openings to retain large quantities of floc yet sufficiently fine to prevent passage of suspended solids; adequate depth to allow relatively long filter runs; and graded to permit effective cleaning during backwash with a minimum quantity of water.

A number of properties of filter media are important in affecting filtration performance and in defining the media (Cleasby, 1990). These properties include:

- *Grain Size and Size Distribution*. Grain size has an important effect on the filtration efficiency and on backwashing requirements for the medium (Cleasby, 1990). Granular media are specified by effective size and uniformity coefficient. *Effective size* (d_{10}) is the sieve size in mm which permits 10 percent of the medium by weight to pass. *Uniformity coefficient* (d_{60}/d_{10}) is the ratio between the sieve size which permits 60 percent by weight to pass and the effective size (McGhee, 1991).

- *Grain Shape and Roundness*. The shape and roundness of the filter grains are important because they affect the backwash flow requirements for the medium, the fixed-bed porosity, the head loss for flow through the medium, the filtration efficiency and the ease of sieving (Cleasby, 1990). The concepts of sphericity and particle shape factor have been throughly explained in section 1.4.3.

- *Grain Density or Specific Gravity.* Grain density, the mass per unit grain volume, is important because it affects the backwash flow requirements for the medium. Grains of higher density but of the same diameter require higher wash rates to achieve fluidization. Therefore, greater hydraulic shear forces occur during backwashing and the washing is more effectively accomplished. Specific gravity is the ratio of the grain mass to the mass of an equal volume of water at a temperature of 23°C.

- Grain Hardness. The hardness of filter grains is important to the durability of the grains during long-term service as a filter medium. Hardness is usually described by the Mohs hardness number, which is a scale of hardness based on the ability of various minerals to be scratched by another harder object. A minimum Mohs hardness of 2.7 or 3 is often specified for anthracite coal filter medium. Silica sand, garnet and ilmenite are very hard and their hardness need not be of concern.

- *Fixed-Bed (or Compacted-Bed) Porosity.* Fixed-bed porosity is the ratio of void volume to total bed volume, expressed as a decimal fraction or percentage. It is important because it affects the backwash flow required, the fixed-bed head loss and the solids-holding capacity of the medium. Fixed-bed porosity is affected by the grain sphericity: angular grains have higher fixed-bed porosity. The loose-bed porosity may be as much as 5 percent greater than porosities measured after gentle compaction of the bed.

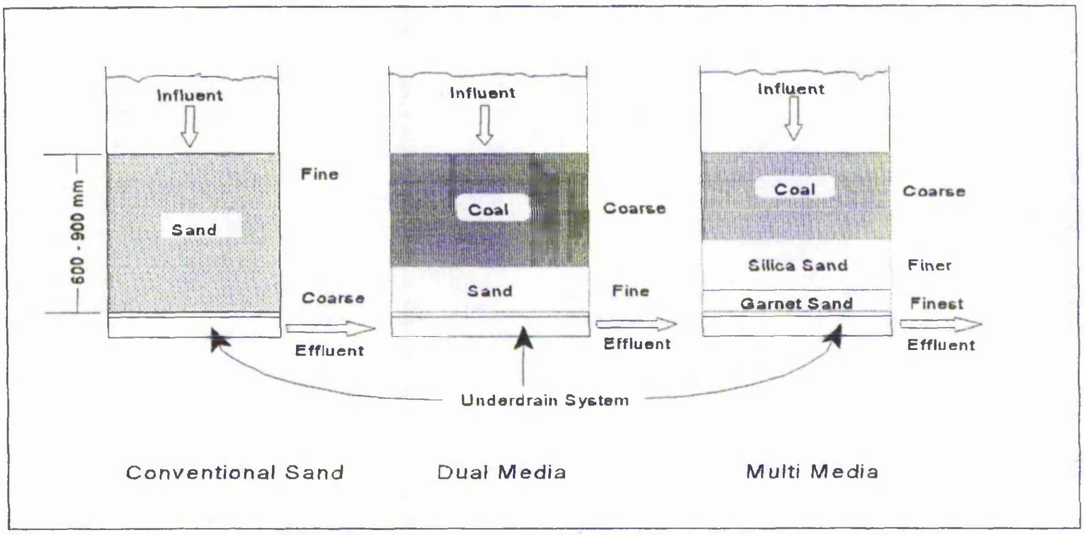


Fig 1.7 Schematic diagrams of typical filter configurations for rapid filtration, after J. L. Cleasby, 1990.

Table 1.6 illustrates typical measured values for some filter medium properties. Alluvial sands have the highest sphericity, and crushed materials such as anthracite, ilmenite and some garnet have lower sphericity. The loose-bed porosity is inversely related to the sphericity; that is, the lower the sphericity, the higher the loose-bed porosity.

Table 1.6 Typical Properties of common Filter Media for Granular-Bed Filters

	Silica Sand	Anthracite Coal	Granular Activated Carbon	Garnet Sand
Grain Density (g/cm^3)	2.65	1.45 — 1.73	1.3 — 1.5	3.6 — 4.2
Loose-Bed Porosity	0.42 — 0.47	0.56 — 0.60	0.5	0.45 — 0.55
Sphericity	0.7 — 0.8	0.46 — 0.60	0.75	0.6

After J. L. Cleasby, 1990

As noted previously, filters may be classified according to the number of filtering media used as mono-medium (homogeneous), dual-medium (heterogeneous) or multimedium (composite) beds.

Depending on the procedure used to backwash a filter, the filtering materials may become stratified or unstratified (Tchobanoglous *et al.,* 1991). Typical filter media configurations are shown in Figure 1.7. The most commonly used of these are the conventional sand and dual-media filters, but a substantial number of triple-media filters have been installed in the United States (Cleasby, 1990).

The smaller the size of granular media, the smaller the pore openings through which the water must pass. Small pore openings increase filtration efficiency not only because of straining but also because of other removal mechanisms. However, as size of pore openings decreases, head loss through the medium increases, resulting in a diminished flow rate. Larger media increase pore size, reduce head loss and increase flow rate, but at a sacrifice of filtration efficiency (Peavy *et al.*, 1985). Coarse materials require higher backwash velocities for fluidization (if this is produced by water flow) but are less likely to form large agglomerates called *mudballs* during backwash (McGhee, 1991).

The earliest filter medium used was a fine sand of nearly uniform size. Sand is normally the cheapest filter medium. The sand used in rapid filtration should be free from dirt, be hard and resistant to abrasion, and preferably be quartz or quartzite. It should not lose more than 5 percent by weight after 24 h immersion in 40 percent hydrochloric acid (McGhee, 1991).

After backwashing, a sand bed becomes hydraulically graded with the largest grains on the bottom and the finest on the top. This forms a bed where most of the removal takes place at or near the surface, which is undesirable.

The earliest modification to the single-medium filter was to increase the sand size to allow longer filter runs and greater depth of penetration while reducing the undesirable effect of surface straining. Dual-media beds of coarser anthracite overlying the sand filter provide an upper layer of increased porosity to reduce surface plugging. Common ranges in effective size and uniformity coefficient for single-medium sand filters are 0.40 to 0.55 mm and 1.3 to 1.7, respectively. The depth of the sand layer is 610 to 760 mm (Cleasby, 1990).

A coal-sand dual-media filter permits use of a relatively coarse anthracite medium with a specific gravity of 1.4 to 1.6 over a finer sand layer of 2.65. The effective size and uniformity coefficient of the coal medium are 0.9 to 1.1 mm and less than 1.7, respectively. Both media layers are usually 300 mm in thickness. The upper layer of coarser anthracite has voids about 20 percent larger than the sand, and thus a coarse-to-fine grading of media is provided in the direction of flow. After backwashing, the bed stratifies with the heavier sand on the bottom and the lighter, coarser coal medium on top. Larger floc particles are adsorbed and trapped in the surface coal layer, while finer material is held in the sand filter; therefore, the bed filters in greater depth, preventing premature surface plugging. A disadvantage of dual-media filters is that the filtered material is held rather loosely in the anthracite layer. Any sudden increase in hydraulic loading dislodges the material and transports it to the surface of the sand layer, resulting in rapid binding at this level (Peavy *et al.*, 1985).

Mixed-media beds using coal anthracite, silica sand and garnet or ilmenite very closely approach the ideal filter. Garnet the finest medium, has a specific gravity of about 4.2, which is greater than that of the sand or coal. The relatively high cost and limited availability of garnet sand and ilmenite make their use as a mono-medium impractical (McGhee, 1991). The three media are sized so that intermixing of these materials occurs after backwashing with no discrete interface between the three. This eliminates stratification and more closely approximates the idea of a uniform decrease in pore space with increasing filter depth. A typical filter has a particle size gradation decreasing from about 2 mm at the top to 0.2 mm at the bottom (Hammer, 1986).

Nevertheless, mixed-media filters are not true depth filters, but rather provide two or three filter surfaces with progressively smaller openings. This permits effective use of a larger portion of the volume of the filter, since coarse particles can be removed at the upper surface while the finer penetrate deeper into the bed (McGhee, 1991).

Dual- and mixed-media filters make possible the direct filtration of water of low turbidity without settling operations. Coagulating chemicals are often added to the influent of the filter to produce small, strong flocs to enhance turbidity removal (Peavy *et al.*, 1985).

A graded gravel underlayer usually 350—450 mm deep is placed over the pipe lateral system to prevent the filter media from entering the lateral orifices and to aid in distribution of the backwash flow (Culp *et al.*, 1978). The conventional gravel system begins with coarse-sized gravel at the bottom with progressively finer-sized gravel layers above up to the filter medium (Cleasby, 1990).

1.4.5 FILTER CLOGGING

The investigations of Ives are representative of the best available in the field of water filtration theory. Ives begins with two equations suggested by Iwasaki in 1937, equations [1.12] and [1.14].

It is usually assumed that the rate of removal of particles is proportional to their concentration. The first-order relation between concentration of suspended solids C and depth L (from inlet surface) is valid for the initial condition of filtration through a uniform bed of media, where λ is the *filter coefficient* (the proportion of particle concentration removed per unit thickness):

$$-\frac{\partial C}{\partial L}=\lambda C \qquad [1.12]$$

The equation is a partial differential because the solids concentration varies with time as well as position in the bed (Tebbutt, 1983). In a filter where all the layers of grains are identical through the bed thickness, the filter coefficient is constant at all depths and independent of L.

The decrease in suspension concentration is logarithmic with respect to depth. This is also evident from the integration of equation [1.12]. At the surface of the filter, at $L=0$, $C=C_0$, (inlet concentration); and at the beginning of the filtration process, at time $t=0$, when the filter grains are clean, $\lambda=\lambda_0$. Thus, equation [1.12] may be integrated to give:

$$C=C_0e^{-\lambda_0L} \qquad [1.13]$$

$$\ln\left(\frac{C}{C_0}\right)=-\lambda_0L \qquad [1.13a]$$

So the initial distribution of the concentration in the flow is expressed by a negative exponential curve through the filter depth. Note the important fact that for a uniform filter, the amount of deposit in each layer is not uniform, and the upper layers contain most of the deposit causing clogging (Ives, 1979).

As suspended particles are removed from the flow they accumulate in the filter pores. The relation between the suspended-solids concentration in the flow and the accumulating deposits in the filter is expressed by the *continuity equation*:

$$V_a\frac{\partial C}{\partial L}+\frac{\partial \sigma}{\partial t}=0 \qquad [1.14]$$

where v_a = approach velocity of the water to the filter surface
σ = specific deposit, volume of suspended solids retained per bulk filter volume.

Equation [1.14] states that the volume of removed suspended particles from the flow is equal to the volume of retained particles accumulated in the filter bed within the same period of time.

During the filter run the efficiency of the filter changes due to the accumulation of deposits in the pores. There are two principal theories describing these changes:

- *Mints (1966)* attributes changes to scour of deposits by the water flowing through the pores
- *Ives (1969)* attributes changes to geometric and velocity changes in the pores.

The relationship derived by Ives is:

$$\frac{\lambda}{\lambda_0}=(1+\frac{\beta\sigma}{\epsilon_0})^y(1-\frac{\sigma}{\epsilon_0})^z(1+\frac{\sigma}{\sigma_u})^x \qquad [1.15]$$

where λ_0 = filter coefficient at t = 0 (clean filter)
ϵ_0 = initial filter porosity
σ_u = ultimate saturation value of specific deposit
β, x, y, z are empirically derived factors.

It has been shown that equation [1.15] is general in that all theories, except Mints′, can be described by appropriate choices of x, y, and z. Note that λ_0 has dimensions of cm^{-1}, all other factors (β, σ_u, ϵ_0, x, y, z) are dimensionless.

Equations [1.12], [1.14] and [1.15] describe clarification of a suspension by filtration at constant rate (v), in a uniform filter of constant grain size (d). If the filter is operated at a different rate, or with a different grain size, then the factors λ_0, β, σ_u , x, y, z will change.

Although some theoretical investigations have been made of the effects of changing flow rate and grain size (Ives and Sholji, 1965; Ison and Ives, 1969), it is necessary to obtain experimental information because the relationships depend on the nature of the suspension (Mohanka, 1969).

As the pores accumulate deposited material the filtration coefficient λ will change. This causes changes in the transport, attachment and detachment mechanisms. Although these changes are shown by different filter coefficients at different filter depths, varying also with filter run time, they are uniquely related to the amount of specific deposit.

Thus, λ is a function of the following parameters (Ives, 1960):

- *Interstitial Velocity* — dependent on porosity
- *Grain Surface Area* — dependent on filter grain size
- *Stoke′s Law Parameters* of the water and suspended particles

The form of this function has been investigated by several authors, with differing results. They all have a common characteristic: their plotted λ versus σ curves commence from $\lambda = \lambda_0$ when $\sigma = 0$, that is for the clean filter containing no deposit; they finish at $\lambda = 0$ when $\sigma = \sigma_u$, when there is so much deposit in a filter layer that no more suspension is removed. At this latter stage, there is still flow through the pores, but the internal surface area is so reduced, and the interstitial velocity so high, that no particles can deposit in the remaining pore space, after which the concentration rises, until it reaches an unacceptably high value. This is called *filter breakthrough*, and if it is reached, the filter run must be stopped (Ives, 1979).

During the course of filtration this accumulation of suspended particles within the pore space causes an increase in head loss. This has also been throughly investigated by several research workers over the past 30 years.

However, all the various mathematical theories of head loss due to clogging caused by deposits can be reduced to the same approximation (Ives and Sholji, 1965):

$$\frac{\partial H}{\partial L}=\left(\frac{\partial H}{\partial L}\right)_0+K\sigma \qquad [1.16]$$

where K is a dimensionless head loss factor, dependent upon grain size (d) and flow rate.

This equation states that the total loss of head at any time t in the filter run is the sum of the initial head loss and a constant multiplied by the local specific deposit at time t.

The total head loss through the filter can be obtained by integrating equation [1.16] with respect to thickness L. Thus, the total head loss can be expressed as a linear function which for uni-size media is (Ives and Gregory 1967):

$$H=H_0+\frac{KvC_0t}{(1-\epsilon)} \qquad [1.17]$$

where H_0 = head loss from equation [1.2] (Carmen — Kozeny)
t = filter run time

K = dimensionless constant, dependent upon grain size (d) and flow rate.

For graded media the expression is similar, but with K replaced by another constant depending on the size grading and the variation of K with particle size and it is a different value to the K value shown in equation [1.16].

1.4.6 FILTER BACKWASHING

To maintain a constant filtering velocity, an increment in driving force must be applied to match each increment in head loss resulting from decreased porosity. Conversely, if a constant driving force is applied, the filtering velocity will diminish as the porosity decreases. As head loss increases across the bed, the lower portion of the filter is under partial vacuum. This negative head permits the release of dissolved gases, which tend to fill the pores of the filter, causing air binding and reducing the rate of filtration (Viessman and Hammer, 1993). In filter operations, a run is terminated when sufficient solids have accumulated to

- use up the available driving force,
- cause the filtering velocity to drop below a predetermined level
- or exhaust the storage deposit capacity of the bed so that solids begin to break through.

At this point, the filter must be backwashed (Peavy *et al.*, 1985).

With a slow filter, penetration of solids is superficial and cleaning is achieved by removing the upper layer of the medium at intervals of a few months, washing and replacing. The rapid filter clogs much more rapidly due to its higher hydraulic loading and the solids penetrate deeper into the filter bed. Cleaning is achieved by backwashing at a rate of about ten times the normal filtration rate (Tebbutt, 1983).

The goal of the backwashing operation should be to keep the filter acceptably clean, so that no progressive evidence of the development of dirty-filter problems, such as mud balls and filter cracks, occurs (Cleasby, 1990).

The washing of tertiary filters is critical. Not only must the filter be cleaned of the accumulated suspended solids, but biological growths must be prevented from developing on the filter grains. These growths of bacteria and protozoa (particularly *Opercularia* and *Vorticella*) are associated with a brown or dark-brown coating on grains of sand or anthracite. If these growths become established normal backwashing procedures will not remove them, and their continued growth leads to short filter cycles, due to severe reduction of the permeability of the filter material (Ives, 1980). It appears that the cleaning system used for tertiary treatment filters should therefore be more aggressive than that for potable water filters (Jago, 1977).

Washing consists of fluidising the filter media either with water, air, or a combination of the two so that the individual grains will be suspended, be subjected to abrasion by their contact with each other, and thus be cleansed of the material which has been accumulated during the filter run. The methods commonly used for backwashing granular-medium filter beds include (Mc Ghee, 1991):

- full fluidisation with or without auxiliary air scour
- surface wash with auxiliary air scour, and partial fluidisation
- combinations of the two.

Surface wash (common practice in the USA) is provided by directing jets of water downward against the surface of the filter before and during the first few moments of backwash. The jets may be mounted on a fixed network of pipe above the bed or on rotating arms which sweep the entire filter. The fixed nozzles apply water at a rate of 0.2 m/min to the entire bed and the surface wash is normally continued while the bed is expanded. The surface wash begins 1 minute before the backwash and continues during the expansion of the bed. Multimedia filters which provide more than one filtering surface must include surface wash for the internal surfaces as well (Mc Ghee, 1991). However, since it is only applied at the surface of the filter and has little penetration, it does not prevent progressive build-up of growth on the medium as a whole, especially in tertiary treatment filters (Jago, 1977).

Air scour is widely used in Europe and can be employed with plenum or pipe-type underdrains using nozzles. Air is introduced into the underdrain at rates ranging from 0.3 to 1.5 m/min (commonly 0.9 to 1.2 m/min), based on the filter plan area. After a few minutes of air agitation, water is added to the underdrain at rates ranging from 0.3 to 0.5 m/min. The air is left on in some designs until the rising flow reaches the effluent structure or troughs and, in others, throughout the washing process.

The air wash fluidizes the bed without expanding it, and the violent rolling action provided by the air scours deposits both on the upper surface and internal surfaces such as those in multimedia filters (Mc Ghee, 1991). Air scour appears to be the most acceptable auxiliary technique for cleaning tertiary treatment filters, but little work has been published on the optimum rate to ensure adequate cleaning (Jago, 1977).

Fluidization is defined as upward flow through a granular bed at a sufficient velocity to suspend the grains in the water. During the process of fluidization, the upward flow overcomes the gravitational force on the grains, and the energy loss is due to fluid motion.

The pressure loss through a fixed bed is a linear function of flow rate at low superficial velocities when flow is laminar (the superficial velocity is the quantity of flow divided by the cross-sectional area of the filter). As the flow rate increases further, the resistance of the grains to wash-water flow increases until the resistance equals the gravitational force and the grains are suspended in the water.

Any further increase in upward velocity results in additional expansion of the bed while maintaining a constant pressure drop equal to the buoyant weight of the media (Viessman and Hammer, 1993). It is assumed that collisions between grains are negligible and extremely limited, otherwise the grains would not be kept in suspension (Amirtharajah, 1978).

A sand bed is completely fluidized when the friction drag or pressure drop across the bed is just enough to support the weight of the filter media. Mathematically, this relationship is given by:

$$\Delta p = hg\rho_w = L_e(\rho_s - \rho_w)g(1 - \epsilon_e) \qquad [1.\ 18]$$

where Δp = pressure across the fluidized bed, N
h = head loss, as a water column weight, m
L_e = length of expanded bed, m
ρ_w = mass density of wash water, kg/m^3
ρ_s = mass density of solid grains, kg/m^3
ϵ_e = porosity of expanded bed, dimensionless
g = acceleration of gravity, m/s^2

The expression in [1.18] is only applied for sand filter beds (as shown by Ives on private correspondence).

Since the quantity of filter medium remains the same whether the bed is stationary or fluidized, the volume of grains initially can be equated to the volume of grains after expansion:

$$\frac{L_e}{L_0} = \frac{(1 - \epsilon_0)}{(1 - \epsilon_e)} \qquad [1.19]$$

where L_0 = length of stationary bed, m
ϵ_0 = porosity of stationary bed, dimensionless.

The quantity ϵ_e is a function of the terminal settling velocity of the particles and the backwash velocity. An increase in the backwash velocity will result in a greater expansion of the bed.

The expression commonly used to relate the bed expansion to backwash velocity and particle settling is (Fair *et al.*, 1968):

$$\epsilon_e=\left(\frac{v_b}{v_s}\right)^{0.22} \qquad [1.20]$$

where v_b = face velocity of the backwash water (backwash flow divided by the total filter area), m/s

v_s = particle settling velocity, m/s.

The depth of the fluidized bed and the backwash velocity for a given size medium (with known v_s) can now be related as follows:

$$\frac{L_e}{L_0}=\frac{(1-\epsilon_0)}{1-\left(\frac{v_b}{v_s}\right)^{0.22}} \qquad [1.21]$$

Equation [1.21] can also be modified for a stratified bed of nonuniform particles where:

$$x_{ij}=\frac{L_{ij}}{L_0} \qquad [1.22]$$

is the weight fraction between adjacent sieve sizes. Assuming uniform porosity in the packed bed, L_{ij} will be the depth of the layer of media represented by x_{ij}. The expansion of this layer is represented by:

$$L_{e,ij}=x_{ij}\frac{L_0(1-e_0)}{1-(\frac{V_b}{V_{s,ij}})^{0.22}} \qquad [1.23]$$

The total expansion is the sum of the individual layers

$$L_e=L_0(1-e_0)\sum\frac{x_{ij}}{1-(\frac{V_b}{V_{s,ij}})^{0.22}} \qquad [1.24]$$

Work in the UK by Sholji (1987) produced a new backwashing model based on an ideal condition utilizing unisize grains, particularly sand, and anthracite, which can be used for any type of hydraulic conditions.

Sholji developed a backwashing model that allows the evaluation and prediction of the appropriate values of backwashing velocity, the exact minimum fluidization velocity, or the point of incipient fluidization required for a predetermined type of unisize grains, dynamic shape coefficient or sphericity, without eliminating any of the major variables influencing the operation of backwashing, such as flow conditions and physical properties of the bed (length, porosity, etc.).

The evolved mathematical and experimental work led Sholji to his mathematical backwashing model:

$$\frac{L_e}{L_0} = \frac{K_2 C_0}{\left(1 - \frac{V_b}{V_s}\right)^{3/2}} \tag{1.25}$$

and experimental model:

$$\frac{L_e}{L_0} = \frac{K_2 C_0}{\left(1 - \frac{V_b}{V_s}\right)^{3/2}} \pm K_3 \tag{1.26}$$

where C_0 = initial concentration of grains forming the filter bed = $(1 - \epsilon_0)$

K_2 = dynamic shape coefficient, dimensionless

K_3 = variable experimental coefficient, which depends on the type, shape, size, and density of the uniform grains used, but independent of ϵ_0 and ϵ_e, dimensionless

The terms L_e / L_0 and $C_0 / [1- (v_b / v_s)]^{3/2}$ are both dimensionless groups, and their representation graphically should produce a linear relationship. The negative or positive sign, shown before K_3, is attributed to the round (+) or flaky (-) shape of the uniform grains used in the fluidized bed. The values for the dynamic shape coefficient, K_2, for the sand used in Sholji's investigation are from 2.000 to 2.230.

Knowing the mean settling velocity of individual grain size, v_s, at a particular temperature, the minimum fluidizing velocity is easily evaluated under the same temperature conditions, provided K_2 and K_3 are determined experimentally for each type of unisize medium. This velocity can be found from the term $C_0 / [1- (v_b / v_s)]^{3/2}$ when the expansion ratio L_e / L_0 is equal to 1.0.

Total expanded bed should range from 120 to 155 percent of the unexpanded depth (Baumann, 1978). Amirtharajah (1978) has shown that the optimum expansion for hydraulic backwashing occurs at expanded porosities of from 0.65 to 0.70.

Various studies in the literature had shown that several properties of fluidized beds maximized around a bed porosity of about 0.70. Those properties were turbulence, heat transfer, mixing, and mass transfer (Cleasby *et al.*, 1978). It was this evidence that led to the hypothesis that backwashing effectiveness would also be optimized at a porosity of about 0.70 (for maximum hydrodynamic shear in a fluidized bed). Excessive expansion is not desirable since the particles will be forced further apart, scouring action will be reduced and the backwash water consumption will be increased (Tebbutt, 1983).

In the case of a single-medium filter such as a sand filter, during backwash with fluidisation the grains tend to stratify by size with the finer grains on top and the coarser grains on the bottom. The tendency to stratify at a given backwash rate (above fluidization velocity) is driven by bulk density differences between the fluidized grains of different sizes. Smaller grains expand more and have a lower bulk density (grains plus fluidising water) and thus rise to the top of the bed.

The stratification is partially upset by nonuniform upflow velocities of the backwash water that creates localized regions of above-average upflow velocity (*sand boils* or *jet action*). Larger grains are transported upward rapidly into the upper bed, while in adjacent regions the sand is moving downward, carrying filter grains down into the bed (Cleasby, 1990).

Intermixing will tend to occur between adjacent layers of dual- and triple-media filters. For example, the upper, finer sand grains of a dual-media bed move up into the lower, coarser grains of anthracite bed that lies above.

The tendency to intermix increases with the backwash flow rate because the bulk densities tend to converge at higher flow rates. The *bulk density* is the mixed density of the grains and fluidizing water, calculated as follows (Cleasby *et al.*, 1975):

$$\rho_b = (1 - \epsilon)\rho_s + \epsilon\rho_w \qquad [1.27]$$

where ρ_b = bulk density, kg/m^3
ρ_s = density of medium, kg/m^3
ρ_w = density of fluid, kg/m^3

An example of this for ordinary silica sand and garnet sand shows that intermixing actually begins before the bulk densities become equal because of uneven flow distribution and mixing and circulation patterns that exist in the fluidized layers. Cleasby and Woods (1975) demonstrated, for this particular experiment, that mixing occurred when the bulk densities converged to within 1800 to 2000 kg/m^3.

The amount of water required for backwash depends on the design of the filter and the quality of the water being filtered. Filter runs on rapid filters may range from a few hours to several days.

The washing cycle requires from 15 to 30 minutes to lower the water level in the bed, begin surface wash, expand and wash the medium until it is clean, allow it to reconsolidate, and return the filter to service. The actual washing lasts 5 to 10 minutes. For typical systems, backwash use amounts to 1 to 5 percent of the flow produced (Mc Ghee, 1991). It represents a rather large volume of water with low solids content. Therefore, its recovery represents a savings in water resources and in the chemicals that were expended to treat it initially (Cleasby, 1990).

CHAPTER 2

LITERATURE REVIEW

2.1 GENERAL

Deep-bed filtration is a long-known but still very important process to eliminate most kind of turbidities from water. Filtration as such is one of the oldest and most widely used water-treatment operations. Since 1829, when sand filters were first used for municipal water treatment, much filtration research has been completed and the process has been considerably improved (Fox and Cleasby, 1966). Therefore, extensive literature on this technique is available containing observations and experience gained from large-scale- and pilot plants.

The literature dealing with filtration is so voluminous that discussing the subject in detail escapes the purpose of this chapter. However, extensive literature has been surveyed and reviewed in order to identify the current level of information regarding rapid filtration by using granular material, such as sand, in the field of water treatment. The main factors that influence the filtration process have already been described in the previous sections. Therefore, here the author shall only deal with the general theory which has been developed during the past 50 years regarding deep-bed filtration —a comprehensive review of all current advances would be too lengthy for the purpose of this work.

2.2 THEORY AND PRACTICE OF RAPID SAND FILTRATION

Soil grains help filter the ground water —picture the extremely clear water that flows out from *underground streams* as spring water—, and through the years environmental engineers have learned to apply this natural process in water treatment and supply systems, and have developed what we know as the *rapid sand filter* (Vesilind, 1975). The actual process of separating impurities from carrying liquid by rapid sand filtration involves two processes: filtration and backwashing. The main focus of this review is filtration. Reference has previously been made to the backwashing process (*see Chapter One, Section 1.4.6*), and will not be repeated here.

As explained earlier the purpose of filtration is to separate nonsettleable solids from water and wastewater by passing it through a porous medium. The most common system is filtration through a layered bed of granular media, usually a coarse anthracite coal underlain by a finer sand (Gimbel, 1983).

In order to understand and to be able to describe the observed behaviour of rapid filters, several theories have been proposed which should provide the basis for an optimum design and operation of such plants.

The science of rapid filtration is 100 years old. It was an inventive leap by Fuller in 1897 that replaced the slow sand filter (with its necessary biological action), by the rapid sand filter with its associated physicochemical action for dealing with turbid river water. Subsequent development refined its design, modified the cleaning process, and hung sophisticated equipment around it, to the form familiar to us today. It was an inventive leap that introduced in the USA in the 1930´s, anthracite coal, first as an alternative to sand, and later as an overlayer to the sand. This has been subsequently developed to multiple layers, and the use of several different alternative media such as garnet sand, expanded slate, pumice and granular activated carbon (GAC). It was an inventive leap by the Dutch, closely matched in the former USSR by scientific research, which lead to practical designs of up-flow filtration, to be followed closely by radial flow (Ives, 1978).

Modern theory of filtration, in the form of basic mathematical statements of the physical process of removal of suspension material, first appeared in the technical literature in 1937. Iwasaki claimed that:

- filtration is a dynamic process, its action being dependent on depth of the filter and time of the filter run
- the removal of suspended particles through the depth of the filter is proportional to the concentration of particles
- the constant of this proportionality increases linearly with the amount of clogging, which is time dependent
- the material removed from suspension clogs the filter pores

Iwasaki based his model upon a detailed and laborious microscopic examination of the penetration and distribution of microorganisms and fine particulate matter in a mixed bed of non-uniform size sand. Although he worked with a slow sand filter, his model as it turns out, would be applicable to rapid sand filters as well.

Iwasaki's basic model consisted of the following:

$$\frac{\partial C}{\partial L}=-\lambda C \qquad [2.1]$$

where C = quantity of microscopic material reaching a 1 cm^2 area of the filter bed at a certain depth L, in the sand bed per day
L = depth within the sand bed, cm
λ = impediment modulus (*filter coefficient*), cm^{-1}

and

$$v_a\frac{\partial C}{\partial L}+\frac{\partial \sigma}{\partial t}=0 \qquad [2.2]$$

where σ = volume of microscopic material, contained by deposition within 1 cm^3 bulk volume of sand at a certain depth L in a certain time, dimensionless
t = time of filtration, days
v_a = approach velocity of the water to the filter surface, m/s

Eliassen (1935) carried out an experimental study which served Stein, who was studying under Thomas Camp at the Massachusetts Institute of Technology, to modify in 1940 Iwasaki´s third statement to read: *the constant of this proportionality first increases linearly, then decreases non-linearly with the amount of clogging.* Excellent as it was Stein failed to produce mathematical statements to describe continuously the whole course of filtration through a run.

In 1951, Daniel Mints published a paper on the mathematical theory of filtration. Mints´ ideas where similar to Iwasaki´s, but he saw the filtration process as being composed of a constant deposition in the filter pores together with a shearing away of existing deposits. In 1955, Shekhtman again with arguments very similar to Iwasaki´s, published a book containing a theoretical exposition of how the filtrate quality and clogging depend on both filter depth and time of operation (Ives, 1964).

Both Mints and Shekhtman considered the detachment mechanism to be important, and at a constant rate of deposition (i.e. constant value of λ) is counteracted by a variable rate of detachment. In 1960, Mackrle presented his thesis dealing with physicochemical forces between filter grains and suspension particles (*Van der Waals forces*).

Meanwhile, research on filtration had started at University College London by the hand of Ives (1959), which produced a mathematical model opposed to Mints´ and Mackrle´s theories. Ives modified Iwasaki´s equations expressing the dependent variables in terms of volumetric concentration and rearranged Iwasaki´s second equation [2.2] to include filtration rate as an independent variable. Ives tested and verified the theory by conducting filtration experiments in which he filtered algae.

All of these theories were presented in the form of partial differential equations, and it was not easy to compare them without a long mathematical exposition. One aspect, however, that of filtrate variability with time, as predicted by the most important theories, was a point of coincidence. It must be borne in mind that there were several other aspects (e.g., filtrate variability with depth and head loss development) in which these theories also differed (Ives, 1964).

Ives´ basic model consisted of the following equations:

$$\frac{\partial C}{\partial L} = -\lambda C \qquad [2.3]$$

where
$C=$ concentration of suspended particles at depth L, expressed in terms of volume per unit volume of water
$L=$ depth of filter layer, m
$\lambda=$ distance rate factor (later referred to as *filter coefficient*), dimensionless

and

$$-\frac{\partial C}{\partial L} = \frac{1-f_\sigma}{v_a}\frac{\partial \sigma}{\partial t} \qquad [2.4]$$

where
$f_\sigma=$ self-porosity of deposited solids, dimensionless
$v_a=$ approach velocity, m/s
$\sigma=$ specific deposit (volume of deposit per unit volume of filter)
$t=$ time of filtration

and (Ives and Sholji, 1965):

$$\frac{\partial h}{\partial L}=\left(\frac{dh}{dL}\right)_0+k\sigma \qquad [2.5]$$

in which

$h=$ head loss, cm

$(dh/dL)_0=$ initial (clean bed) head loss per unit depth

$k=$ head loss constant, dimensionless

$L=$ depth of filter layer, cm

In addition to the preceding equations, Ives'filter model also included various empirical expressions for defining or predicting λ (see Chapter One, Section 1.4.5). Equation [2.3] states that the amount of suspension material removed in any layer of the filter is proportional to the local concentration of suspension in the flow.

As it was observed before, this has been the basic assumption of all filtration theoreticians since Iwasaki first formulated it in 1937 but it was not rigorously experimentally confirmed until Sholji (1963) and Ison and Ives (1969) demonstrated it at University College London. It may be regarded as having a statistical basis (Litwiniszyn, 1963; Hsiung and Cleasby, 1968) with λ representing the probability of removal of a particle, by the filter.

Usually values of λ represent a mean for the whole suspension, even though it may be heterodisperse. Only Mackrle and Mackrle (1962) have attempted to create a mathematical model for λ (in a series form) for heterodisperse suspensions (Ives, 1970).

Equation [2.2] states that the volume of material removed from suspension in a unit layer during an interval of time is equal to the volume of deposit in the same layer during the same time. It is, therefore, an *equation of continuity* for the suspended solids (Ives, 1970).

Solutions of equations [2.3] and [2.4] are obtained by using a certain assumed functional relationship between λ and σ. Many such functional relationships between λ and σ are listed by Tien and Gimbel (1982). Out of these, the relationship proposed by Ives (1969) is more general in nature (Ives, 1975) and has been successfully used by Sembi and Ives (1983) — *See Chapter One, Section 1.4.5, Equation [1.15].*

However, the use of equation [1.15] requires σ as an input. While it is possible to compute σ in units of mass per unit volume of media bed, its conversion to σ as a *bulk specific deposit* (i.e. volume of deposit per unit volume of filter) is not straightforward due to inherent uncertainty with the determination of the value of the porosity of the deposited solids. Sembi and Ives (1983) tackled this problem by assuming

$$\sigma=b\sigma_m \qquad [2.6]$$

where σ_m= absolute specific deposit in mass (concentrations) per unit volume of media

b = bulk factor, which converts mass concentration of suspension deposited into volume occupied by deposits.

Sembi and Ives (1983) expressed $b\sigma_m$ as:

$$b\sigma_m=A_1\sigma_m+A_2\sigma_m^2+\ldots+A_n\sigma_m^n \qquad [2.7]$$

where A_1, A_2, ..., A_n are constants and n is the order of polynomial. Use of equation [2.7] was made by these authors to model their experiments.

Sembi (1981) and Sembi and Ives (1983) used equations [1.15], [2.7] and the Kozeny head loss model as given below:

$$r=[1+\frac{b\sigma}{(1-e_0)}]^{c_1}[\frac{e_0}{(e_0-b\sigma)}]^{c_2} \qquad [2.8]$$

where r= ratio of head loss at any time to initial head loss

c_1, c_2= exponent coefficients assumed to be 4/3 and 3 from theoretical considerations.

In order to use the equations developed by Ives, it is necessary for the filter designer to determine experimentally a set of five coefficients for each filter bed, flow rate, and pretreatment condition he may wish to consider.

Ives and Sholji (1965) investigated the effects of certain physical filtration variables (namely, grain size media, grain material, depth of media, flow rate, inflow concentration of suspended particles, initial porosity, type of suspension, and water temperature) on the filter coefficient constants, and their study represented a valuable comparison of the filtration theories at that time (O'Melia *et al.*, 1967).

Through the simulation of filter performance, five parameters of equation [1.15] along with six coefficients in equation [2.7] were evaluated by Sembi (1981). To describe the filter performance, Sembi and Ives (1983) suggested to evaluate as many as eight terms in equation [2.7] in addition to the five parameters of equation [1.15]. Thus, the modelling efforts were aimed at obtaining as many as 11-13 parameters. If the constants in equation [2.7] are assumed, the modelling efforts will be reduced to the evaluation of seven parameters only, i.e. five parameters in equation [1.15] and two parameters in equation [2.8].

This last approach has been most widely used to model the filter runs. However, instead of using a non-linear variation of the bulk factor with the absolute specific deposit, a constant value of the bulk factor has been considered in several investigations. But the use of a constant value of bulk factor has not been favoured by Camp (1964) and Sembi and Ives (1983), as it does not explain the depth-dependent variation of the bulk factor, as manifested by most of the filtration results (Ojha and Graham, 1993).

Some other mathematical models were suggested at that time, but study of them was unrewarding as their theoretical bases were obscure, and they were unsupported by any experimental evidence (Hall, 1957; Boreli and Jovasovic, 1961).

Work produced by Ling (1952) at the University of Minnesota underlined Eliassen´s conclusions regarding the depth and time dependence of filter operation, and showed most conclusively that the filtrate drawn at any depth improves in quality before it begins to deteriorate. This point was the significant difference between the European theories; for the latter held that the filtrate deteriorates continuously from the very beginning of the filter run.

In 1955, Stanley was the first to utilise radioactive tracers in filtration. His experiments showed that floc, once deposited in the filter pores, does not become detached and migrate in the filter; a direct experimental reputation of Mints´breakaway hypothesis (Ives, 1964).

The dominant mechanisms of deep-bed filtration relate to the particles that are significantly smaller than the pores, hence the intrinsic nature of filtration in depth, that is, not straining. However, it is recognised in practice that a distribution of particle sizes exist in suspensions, and some may exceed the pore size. This is one of the causes of surface layer straining which forms a mat on the inlet face of the filter causing an undesirably high, exponential increase in pressure drop (Ives, 1986).

Regardless of the formation of such surface clogging, particles may also flocculate within the grain pore and form to larger suspended particles and strain or accumulate within the pores at a deeper layer. This phenomenon has come to be known as *partial straining*.

In 1962, Cleasby and Baumann of Iowa State University published some results of their experiments. In searching for an optimum filter rate, Cleasby and Baumann identified the importance of this floc mat that may form on the filter surface; a topic that had received scant attention from the theorists. Opinions differed on the form of structure adopted by the surface mat: one view was that it is a continuous compressible cake (Cleasby and Baumann, 1962), the other that it is a mat partly on the surface with holes in it (Ives, 1963).

As explained earlier, most mathematical filter models were highly empirical simplifications of the operation and only two models of the rapid sand filter offered a rational, though somewhat empirical, application of many of the basic principles and mechanisms of filtration. The most significant point was that the experimental studies conducted both in America and Europe showed a great similarity. Their results, therefore, could be unified by the mathematical statements resulting from the theories (Ives, 1964).

One of these models was the one proposed by Ives (1960) and represented an extension of a mathematical model of the slow sand filter developed by Iwasaki in 1937. The other model was proposed by Camp (1964) and consisted of a modified form of the Kozeny equation combined with a variation of the Iwasaki model similar to that employed by Ives.

It remains to be seen whether these models, which have been tested against experience with potable-water filters, are capable of describing the behaviour of filters treating sewage-works effluents, where the biological activity and the concentrations of contaminants are so much greater (Jago, 1977).

2.2.1 REMOVAL MECHANISMS

The early theories for deep-bed filtration in the initial phase were mainly based on the classical particle transport mechanisms such as interception, sedimentation, diffusion and inertia. These mechanisms —which have already been described in Chapter One, Section 1.4.2— can be defined by characteristic dimensionless numbers, which can be combined into a phenomenological equation such as, for example, the general equation of Ives (1970).

The problem with all data and information gained in previous research works is that no one of them fully replicates the real life situation with regard to the following (Ives, 1985):

- the media, which is not uniform and usually varies from layer to layer in the filter
- the raw water suspension, which is heterogeneous with ill-defined particle characteristics
- and pretreatment, filtration and backwashing are a continuous process which are not operated on an intermittent or batch mode basis.

The removal of suspended particles within a filter is considered to involve at least two separate and distinct steps:

- first, the *transport* of suspended particles to the immediate vicinity of the solid-liquid interface presented by the filter (i.e. to a grain of the media or to another particle previously retained in the bed), which is a hydrophysical process

- second, the *attachment* of particles to this surface (O´Melia and Stumm 1967; Ives and Gregory, 1967), which is a physicochemical process.

- finally, there is a step also taking place in the bed —a *detachment step* first suggested by Mints in 1964, who stated that retained particles are being detached as long as new particles are being supplied.

The transport and attachment steps and their mathematical descriptions are commonly available (e.g., Yao *et al.*, 1971; O´Melia and Crapps, 1964; Ives, 1975; O´Melia, 1985; Cleasby, 1972; Rajagopalan and Tien, 1979; Spielman and FitzPatrick, 1973) and have already been described in the previous chapter; thus, they will not be considered here in any detail.

In general, three phenomena —Brownian motion, interception, and gravity sedimentation— have been identified as the mechanisms responsible for the transport of suspended particles from the bulk liquid to the medium surface. The importance of each mechanism depends on the suspended-particle size and flow conditions (Sprouse and Rittmann., 1990).

Chemical conditions and surface charge of particles tend to be most influential in the attachment step. However, there is experimental and practical evidence that increasing the flow in a deep-bed filter, when deposited particles are present in the pores, leads to detachment of some of these particles causing a locally increased suspension concentration. Mints (1966) already suggested that such detachment takes places even at constant flow rate, because deposits in the pores cause local increases in interstitial velocity.

2.2.2 PARTICLE COUNTING

Particles in treated water provide a surface area for the adherence of toxic chemicals, shielding of microorganisms from disinfection (e.g., *Cryptosporidium* and *Giardia*) and their presence also reduces the aesthetic appearance of the treated water (Ives *et al.*, 1993). Therefore, the removal of particulates is one of the main objectives in the treatment of drinking water.

A number of studies have reviewed possible correlations between turbidity and particle counts with varying conclusions. A study carried out for the AWWA Research Foundation (Hargeshiemer *et al.*, 1991) found a very good linear relationship between particle counts (Number of Particles (*NP*) > 0.7 µm) and turbidity.

Hargeshiemer *et al.* (1991) found that turbidity followed the same trends as particle counts (*NP* 1 to 5 µm) and Cleasby *et al.* (1989) reported good correlation for particle counts (*NP* > 1.0 µm) and turbidity. Beard and Tanaka (1977) found correlation in the trends of the two measures but not a point to point correlation.

The studies with good correlations measured the particles in the low and sub-micron particle size ranges which may be an important factor in correlation with turbidity. Turbidity is based on a measure of light scatter which is caused by all particles, the majority of particles in filtered water are likely to be in the sub-micron size (Hargesheimer *et al.*, 1991), and a measurement which includes more of this size is likely to show a better correlation.

Particle counts show large changes for only a small change in turbidity thus highlighting when filter performance is changing. The question of suitability of the different particle counters for natural and flocculated suspensions cannot be discussed here but it is an important factor when applying particle size analysis.

The particle counter counts and sizes individual particles, offering a quantitative measure of the particulates in water and their removal by filtration, compared to a relative turbidity measure.

2.2.3 FILTER RIPENING, PARTICLE SIZE AND BREAKTHROUGH

As explained earlier, depth filtration removes particles by attaching them to the media or to previously retained particles. Continued capture of particles leads to an increasing surface area for the attachment of further particles and improvement in removal efficiency of the filter. After backwashing this phenomena is particularly important to improve the efficiency of the clean bed filter and is referred to as *ripening*. *Breakthrough* refers to the time when the filter removal efficiency decreases, letting an increasing number of particles through, and is closely related to the particle size owing to the domination of hydrodynamic forces.

Ripening, the effects of the particle size distribution (PSD) and breakthrough are three of the many variables which need to be addressed in any effective modelling of filter performance. Models for the ripening of packed-bed filters are more empirical (O'Melia, 1985); included among important studies are those by Ives and Gregory (1967), Tien and Gimbel (1982), and O'Melia and Ali (1973).

Lack of understanding of the complex nature and interaction of the variables affecting filter performance has to date hindered the development of a practical model for prediction of filter performance. Even models developed for conditions of clean bed and single sized media have tended to underpredict removals of small particles and over predict removals of large particles (Moran *et al.*, 1993).

Characterization of a suspension by its PSD is a very valuable information for the interpretation of solid separation performance. The term *particle size*, however, is not clearly defined. It depends on the shape of the particulates and, above all, on the method of particle counting, i.e. how the particle size is defined in terms of diameter, surface or volume measurements (Boller, 1993).

Early work by Kozeny (1927) and Fair and Hatch (1933) on models for the performance of rapid sand filters resulted in descriptions of the effects of media size, filtration rate, bed depth, bed porosity, and fluid temperature on the head loss through clean filters. Later, the removal accomplished by clean filters was described in terms of the physical characteristics of the suspended particles, the suspending fluid, and the filter bed by Friedlander (1958), Yao *et al.* (1971), Spielman and FitzPatrick (1973), and others with some success. However, these models for head loss and removal efficiency become incorrect as soon as a filter run begins.

Quantitative concepts describing changes in head loss and filtrate quality as filtration proceeds are needed. As explained earlier, Ives *et al.* (1967, 1969) developed good mathematical characterizations of these changes, but these efforts use several empirical coefficients that are not yet related to properties of the suspension to be treated or the filter providing treatment.

O´Melia and Ali (1973) assumed that particles removed from the flowing fluid and retained within a filter bed can act as collectors or filter media for particles applied subsequently to the bed. The result was a model for filter ripening in which removal of suspended particles by packed-bed filters has an autocatalytic character (Lawler *et al.*, 1980).

Ripening and breakthrough in filters has generally been evident from turbidity and suspended solids monitoring but with the introduction of particle counting the influence of particle size on ripening and breakthrough has been studied by a number of researchers (Moran *et al.*, 1993). From considerations of the mechanisms of filtration, it can be concluded that efficient filters are ripened filters (O´Melia, 1985).

Experiments conducted by Yao *et al.* (1971), for single spheres found that 1 μm sized particles had the lowest removal efficiency. They were able to predict this critical size with their model; they found it was between the effective size zones of two separate transport mechanisms.

There is evidence from Sholji (1963) and Yao (1971) that the dominant transport mechanism for smaller, sub-micron, particles is diffusion with gravity —due to their greater Brownian motion—and interception/ sedimentation for larger particles, which leaves the 1 μm sized particles in between with no dominant transport mechanism.

Breakthrough of particles can either be caused by particles passing straight through the filter, or as a number of studies have concluded through attachment and break-off (Moran *et al.*, 1993).

In reality there is probably a combination of reduced particle attachment efficiency and particle break-off. Both are related to specific deposit (volume attached particles per unit volume of bed), greater deposit causes higher interstitial velocity and less attachment, greater deposit results in greater projection of attached particles into the flow stream and risk of shearing (Moran, *et al.*, 1993). In addition the shearing forces will increase with the interstitial velocity supporting the attachment and break-off theory.

In all cases of filtration the flow is laminar (Reynolds number is approximately 6.0), that is pressure drop is proportional to flow rate (*Darcy's Law*). Cleasby and Baumann (1962) at Iowa State University showed that even with filters considerably clogged with deposits, causing pressure drops many times greater than the clean medium value, Darcy's Law was still obeyed. The flow regime has some similarity with Poiseuille flow in a capillary; indeed the Kozeny — Carman model of flow, which describes the flow of clean fluids through porous media, is based on this similarity.
This means that there is a velocity gradient in each pore, with zero velocity at the boundary with the grain surface, and a maximum velocity near the pore centre.

Also in the mid 70s Ives' experiments at University College London visualising flow with dye streams, round 5 mm grains showed no disturbance of the streamline flow at rates considerably higher than those encountered in practice. Ives also described the hydrodynamic transport of suspended particles to the medium as a function of the Reynolds number of the filter based on medium diameter, overflow rate, and fluid density and viscosity.

The head loss required to maintain a given rate of flow through a clean bed of granular media is determined by the geometry of the media, i.e., media effective size, media shape, and media size distribution; fluid viscosity; bed depth; bed porosity; and the ratio of filter diameter to media effective size (Lang *et al.*, 1993).

The *effective size* was a definition produced by Hazen in 1892 because he observed that for most natural sands the ten percentile by weight (i.e. 10% by weight was smaller than this size) appeared to control the hydraulic effectiveness of the sand, irrespective of the spread of the size. However, this observation was confined to the hydraulic resistance of the clean filter and was not intended to describe the removal characteristics of the media. Furthermore the observation applied to filters in which the grains were distributed homogeneously. Obviously, this is not the case of the typical rapid sand filter designed to retain suspension particles, which after backwashing its filtering material becomes size-stratified and not homogeneous (Ives, 1963).

The *uniformity coefficient*, was also first defined by Hazen as the ratio of the size of the sixty percentile to the size of the ten percentile by mass, is really a non-uniformity coefficient, for as it increases from unity the sand is becoming less uniform (Ives, 1963).

2.2.4 FILTER OPTIMISATION

It has been felt since the late 1960's that knowledge of the detailed operation of filters could be applied to produce more efficient designs. One aspect of this has been the concept of *optimisation* of the design so that both the hydraulic capacity and the clarification capacity of a filter could be used fully (Ives, 1970).

An optimum is achieved when the filter design and operation cause the filter to reach its head loss limit (i.e. when the pressure line just reaches atmospheric pressure in the filter media) at the same time as the filtrate quality deteriorates to an unacceptable value. Formal procedures for achieving this was published by Mints (1966) and extended by Ives (1968). A review of optimisation of rapid filters was also produced as a thesis by Gur (1969). These methods achieve an optimum design by appropriate choice of filter thickness, filtration rate and filter grain size (Ives, 1970).

The above studies indicated that filter operation can be optimised by use of deeper beds with coarser media and higher velocities, that is to minimize additional head loss, use greater depth of bed for removals and maintain overall removal efficiency.

There will be trade-off studies in any optimisation, and Sembi and Ives (1983) theoretically reviewed filter performance optimisation. They aimed to reach the additional head loss limit and the effluent quality limit (i.e. breakthrough) at the same time, but also to maximize the production of filtrate between backwash cycles. The results indicated that the dual media filters outperformed single media filters and triple media filters only performed slightly better than dual media filters.

The effects on filter performance of intermixing of media at media interfaces have also been studied (Baumann *et al.*, 1974; Cleasby *et al.*, 1975; Tchobanoglous, 1970).

Cleasby *et al.* (1975) suggested there is nothing to be gained from intermixing, but pointed out that some intermixing was inevitable. Anthracite used on its own has been shown to be no more efficient for removal of solids than sand of the same size (Pettet *et al.*, 1951; Tchobanoglous, 1970; Tebbutt, 1971). However, In Sholji's experimental results (1963) it can be observed that the use of anthracite was more efficient than that of sand. Tebbutt found anthracite more difficult to keep clean. It has been reported that both mixed-medium and up-flow filters allow longer runs and capable of higher flow rates than conventional rapid gravity sand filters (Jago, 1977).

In addition to altering the physical variables it is possible to achieve optimum conditions by modifying the nature of the suspension with polyelectrolytes (Mints, 1969), or varying chemical dose (Kreissl *et al.*, 1968) or with polyphosphates. This is the most flexible method but it is difficult to formalize in any predictive manner. Constant monitoring of the performance of a small pilot filter gives a guide to the polyelectrolyte dose required at the main filter inlet (Conley, 1965).

Some results were obtained with wastewater filtration after precipitation of phosphates with ferric salts and polyelectrolyte (Morgeli, 1979), as for example 80 percent removal of suspension to a limit head loss of 2 m, in a 1 m deep bed of 2-3 mm quartz sand operating at 10 m/h. This produced a maximum quantity of filtrate during the filter cycle (23 h).

The objective of an optimum design is to produce filtrate at least cost per m^3, and this depends on capital costs, energy and maintenance costs as well as operational costs of chemicals, washwater and compressed air. The most important factor is the flow rate as this will determine the plan size of the filter installation (Ives, 1980).

Based on a simplified mathematical model of the filtration process, Herzig *et al.*, (1970) studied the theory of optimisation, including an initial comment on the economic optimum. Taking standard conditions of uniform 0.45 mm sand, with a head loss limit of 5.2 m, they calculated a maximum production of filtrate with a filter 1.35 m depth at a filtration rate of 10.8 m/h, giving a filter run of 2 h. This is not necessarily the optimum solution as they preset their grain size and headloss limit (Ives, 1978).

Using a standard form of filter, Huang and Baumann (1974) showed that as the sand size increased, unit cost of filtrate diminished, with increasing filtration rate, and sand depth, but with lower head losses. Unfortunately, they did not continue their calculations to reach a minimum in unit cost, but stopped at 1.3 mm sand size, at 14 m/h, 0.9 m depth, and 2.5 m head loss. Their minimum lies at a coarser, deeper filter, operating at a higher rate (Ives, 1978).

An integrative approach to water treatment plant design and operation, from raw water quality through to filter bed performance, will facilitate process evaluation and has the potential for providing a basis for optimal design (O´Melia, 1985).

CHAPTER 3

APPARATUS, MATERIALS, EXPERIMENTAL PROCEDURES AND OBSERVATIONS

3.1 INTRODUCTION

The experimental apparatus, materials and chemicals used and the procedures and observations will be described in this chapter, to investigate the validity and practical applicability of rapid filtration by a multi-compartment up-flow and down-flow process in removing the suspended particles under various initial conditions.

Thus, this investigation concentrates on the behaviour of each material used and the characteristics of removal, and efficiency, when various uniform sizes of granular material are used and various degrees of concentration of synthetic turbidities are involved. These observations are used to monitor the separation of suspended matter in depth in downward and upward flow.

Various types of impurities, inlet concentration and rates of flow have been carried out to determine the removal efficiency of the system. The observed results are expected to be a contribution towards a thorough understanding and further analysis of the up-flow and down-flow rapid filtration in the field of water treatment.

Three types of synthetic turbid materials namely kaolin, lycopodium powder and polyvinyl chloride microspheres (PVC) were used in this research work. Experiments were conducted in a perspex filter box receiving synthetic turbid water of a known concentration on a constant flow rate basis. Sampling ports were allocated to collect the sample, which were installed at varying heights of the filter, to determine the residual turbidity at different intervals of time. Likewise these sampling ports acted as manometric pressure probes that provided the measurement of the head loss of the system. Particular attention was given in flushing the pipes before collecting the sample to avoid particle deposition in the sample collecting pipes.

The model filter apparatus, peripheral equipment and instruments, synthetic turbidity materials, sampling ports and technique, preparation of suspension, schedule of the experiments, experimental procedures and observations are all described in the following subsequent sections.

3.2 EXPERIMENTAL APPARATUS

3.2.1 DESIGN CONSIDERATIONS

Normally the flow is downwards through the media. The water percolates down through the saturated granular medium at a rate of flow between 5 and 15 $m^3/m^2 \cdot h$ and generates an initial head loss due to the permeability of the clean sand.

The filter model unit has been designed to allow both downward and upward flow filtration. The deposition of particles from the homogeneous suspension into the sand pores causes a progressive head loss; consequently an increasing pressure difference is necessary to maintain a given flow rate.

Flow is laminar and this means that the flow follows Darcy's Law (*flow rate is proportional to pressure difference*), and the streamlines are ordered round the grain surfaces and in the pores. Flow is reversible: the same pattern of streamlines occurs whether flow is upwards, or downwards. There is a maximum fluid shear stress at the boundaries (grain surface) and a shear gradient exists across all pores. The net velocity or the resultant velocity acting upon a suspended particle is a product of hydrodynamic and settling velocities, depending on the direction of flow. The rate of flow was kept constant for all the experiments, this actually being 0.2, 0.4, 0.6, 0.8 and 1.0 l/min (1.9, 3.8, 5.6, 7.5 and 9.4 $m^3/m^2 \cdot h$) per each type of synthetic material and concentration of turbid material.

Pre-filtration turbidities were selected between 100 and 400 mg/l, for each turbidity material and rate of flow. At nearly all filtration rates mentioned earlier, an effluent quality of less than 1 mg/l can be obtained, thus achieving a removal efficiency of more than 99 percent in most of the cases, which it could probably include the elimination of cryptosporidium cysts and viruses. The object of drinking water filtration is to reduce the initial turbidity values of the filtrate to 0.1 mg/l or less.

3.2.2 FILTER MODEL

The model treatment unit as shown in Figure 3.2, consists of a filter box made of 12.0 mm thick perspex sheet with an operating water height of about 2000 mm. This box is 700 mm deep, 932 mm wide and 104 mm long, divided into five compartments of 160 mm wide each one, with a net column section of 80 x 80 mm.

The model was made with perspex material to enable the author to observe the behaviour of the media during filtration and backwashing. The total depth of sand after compaction was 3023 mm (as specified in Table 4.1). This was supported on 125 mm graded gravel per subcompartment (1250 mm total depth).

This model treatment unit was designed by Dr. Sholji, who is the first supervisor of this research work, and manufactured by the Heavy Structures Civil Engineering Workshop, Department of Civil and Structural Engineering, The Nottingham Trent University, UK.

The whole assembly of the experimental apparatus used in this investigation consists of various component parts which will be described separately in the following subsections with reference to the schematic diagram shown in Figure 3.2. Photographs of the arrangement are also shown in Figure 3.4 and Figure 3.5.

a) *Multi-Compartment Filter Bed*— as described above, it consists of a filter box made of 12.0 mm thick perspex plastic sheet divided into five compartments, each having both downward and upward filtration, thus making a total of ten compartments. The object of this is to introduce a larger size of unisize grain in the first compartment, followed by smaller unisize grains in the subsequent compartments that follow. The coarser unisize media carry the burden of removing larger suspended colloidal matter from the flow, throughout the depth of the filter layer, leaving the fine subsequent unisize compartments of the filter bed to follow, suit and share the burden of the removal of suspended impurities in depth, for down-flow and up-flow in a constant rate of flow.

The uniform sand grain sizes used in the experiments were 548 µm (compartments 1 and 2), 653 µm (compartments 3 and 4), 777 µm (compartments 5 and 6), 922 µm (compartments 7 and 8), and 1086 µm (compartments 9 and 10), respectively. The filter box has been specially designed to facilitate the water to flow from compartment number 1 to compartment number 10 by means of a slight gradient as shown in Figure 3.8. All joints of this box were sealed with resin to prevent any water leakage. A 18 mm internal diameter inlet pipe with controlling valve was connected to the upper part of the filter compartment number 1 at a height of 670 mm from the base. The outlet pipe of 18 mm diameter was connected to the upper part of sub-compartment number 10 at a height of 660 mm from the base.

A backwash piping system was connected to the lower part of the filter box, providing a valve-controlled backwash water inlet to each one of the filter´s ten compartments. Six 18 mm diameter outlet pipes with controlling valves were connected to the upper part of the filter box to drain the waste water produced during the backwashing process.

An overflow weir arrangement was built on to the upper filter box wall at a height of 500 mm with a sump to receive the overflow backwash water reaching beyond the predetermined water level, and this sump was connected to the main outlet pipe.

Backwash water was introduced into the filter box through an inlet pipe made of perspex with 3 mm diameter orifices located at the bottom of each compartment. These backwash water supply points were designed to provide a uniform flow distribution throughout the filter bed. To secure both a perfect flow distribution and support the filter medium, these outlets were covered with two layers of gravel. These layers were graded from coarse on the bottom (1070 g per unit, comprising 2 compartments) to fine at the top (560 g per unit).

b) *Sampling Ports*— sixteen sample collecting pipes were inserted into the filter box to collect the samples for residual turbidity measurement. Details of the sampling ports have been shown in Figure 3.3 and described in section 3.9. Six of them were allocated at the upper part of the filter box, and the rest at the bottom of each compartment.

Each sampling port consisted of two parts: the inner part was a 3.0 mm diameter flexible PVC pipe inserted in a way to ensure a proper length from the inner wall ; the outer projected part of the sampling port was connected to a short 5.0 mm diameter rubber tube equipped with a pinch-cock to enable the collection of water samples.

c) *Mixing Tanks*— two 610 x 610 x 450 mm water tanks made of plastic material were installed under the filter box bench in order to supply the suspension required to run the experiments. The suspension requirements were met by using synthetic turbidity materials dispersed in tap water, being stirred continuously by means of two stainless steel rotors. The details of the stirring rotors are shown in Figure 3.1.

The synthetic raw water input to the filter was a suspension of PVC microspheres (0.5— 1.5 μm diameter), kaolin (2— 53 μm diameter; 2 μm 80% and up to 53 μm 20%) or lycopodium powder (35 μm diameter) filtered at various initial concentrations and rates of flow. The suspension was pumped out to the constant head tank as shown in Figure 3.2.

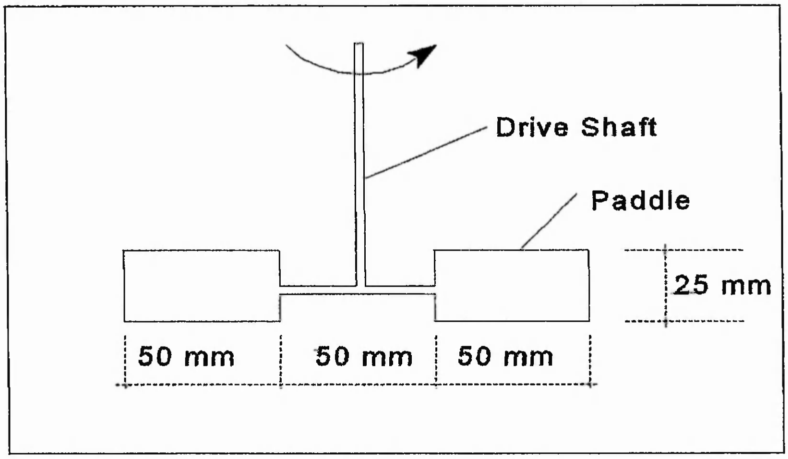


Fig 3.1 Details of the stirring rotors

d) *Constant Head Tank*— this was a 500 mm deep circular receptacle made of 6.0 mm thick perspex plastic sheet with an internal diameter of 200 mm allowing an operating water height of 2000 mm. The suspension was pumped out to the constant head tank above the surface of the filter bed from the two mixing tanks below. An overflow weir arrangement was built on to the tank wall with a receiving sump and this sump connected to an overflow pipe to receive the overflow water reaching beyond the predetermined water level. This overflow pipe returned the overflow water to the mixing tanks as shown in Figure 3.2. The constant head tank supplies the turbid water to the filter by gravity and the flowing water emerges at the top of compartment number 10. The details of the constant head tank are shown in Figure 3.5.

e) *Manometer*— twenty-six manometric pressure probes were installed at varying heights of the filter to enable the measurement of the head loss of the system. Details of the head loss ports have been shown in Figure 3.9. Six of them were allocated at the upper part of the filter box, ten at the bottom of each compartment and the remaining ten halfway between the inlet and the outlet of each compartment. Each head loss gauge consisted of a 6.0 mm diameter flexible PVC pipe equipped with a pinch-cock to enable its shutting should the head loss rise beyond a maximum value and this way prevent the development of air suction into the system.

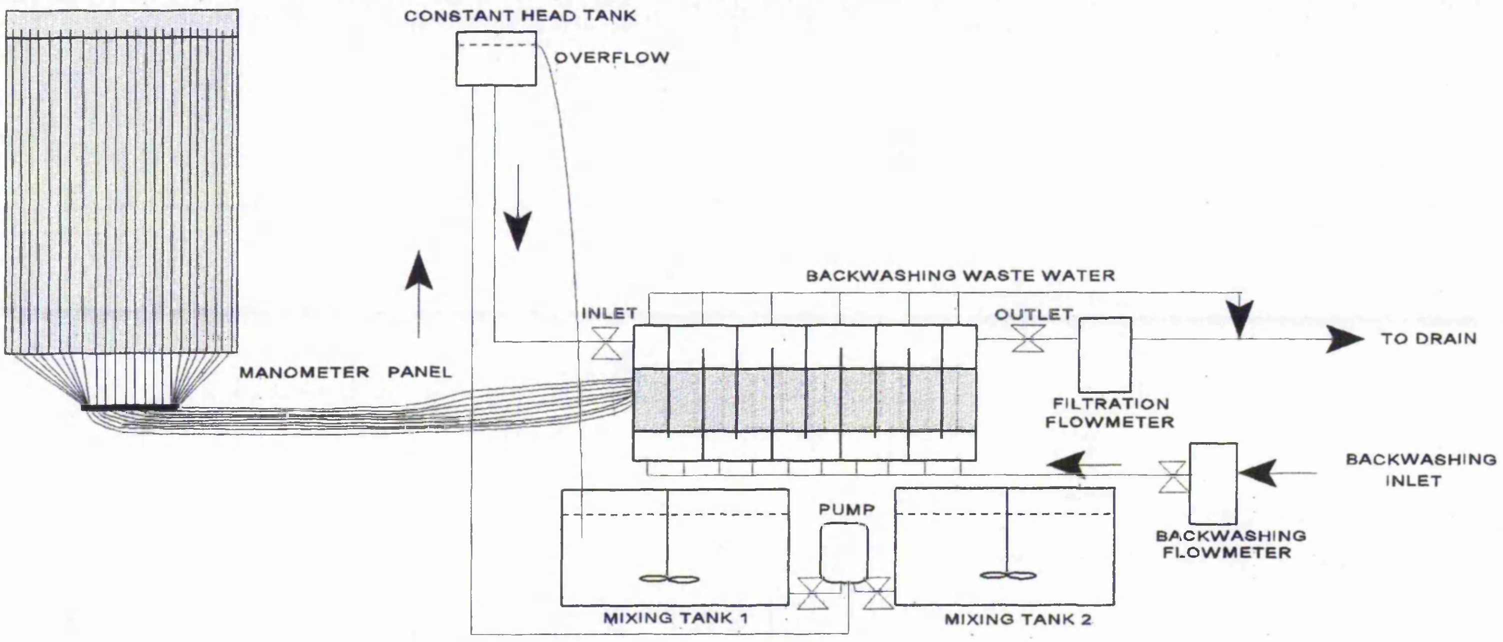


Fig 3.2 Schematic Diagram of the Filtration System.

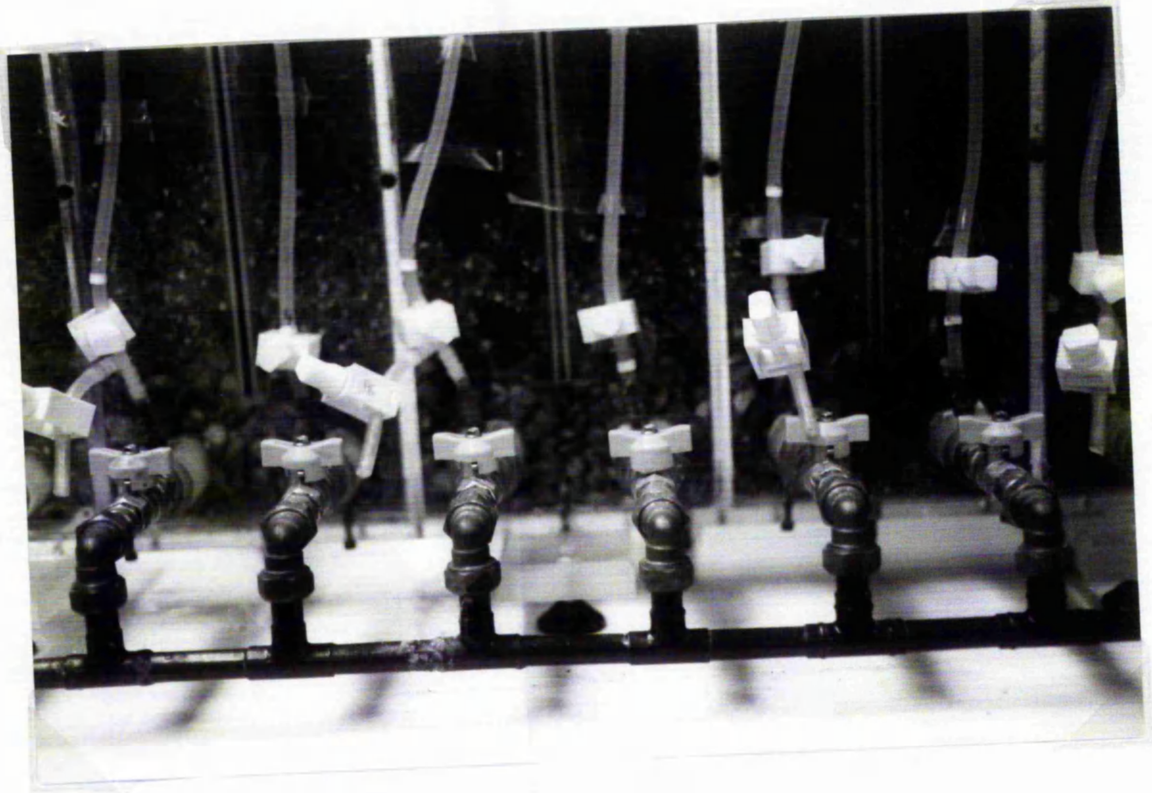


Fig 3.3 Photograph showing the details of the sampling ports.

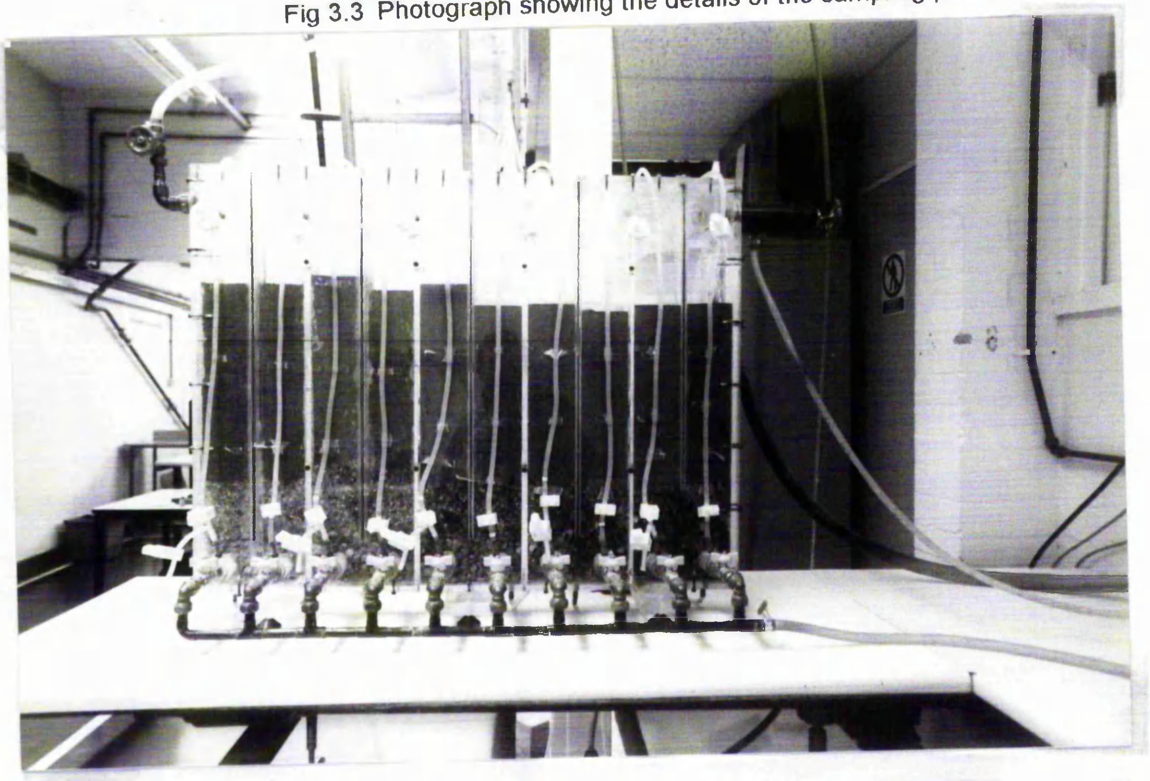


Fig 3.4 Photograph showing a general view of the filter box.

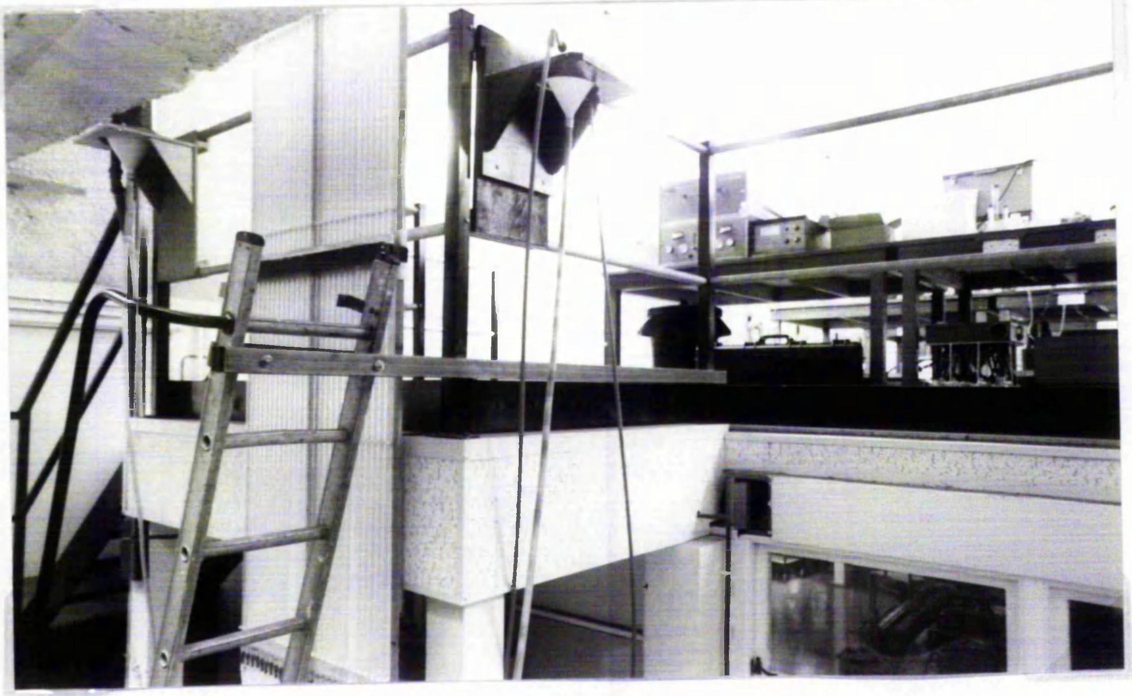


Fig 3.5 Detail photograph of the manometer and constant head tank.

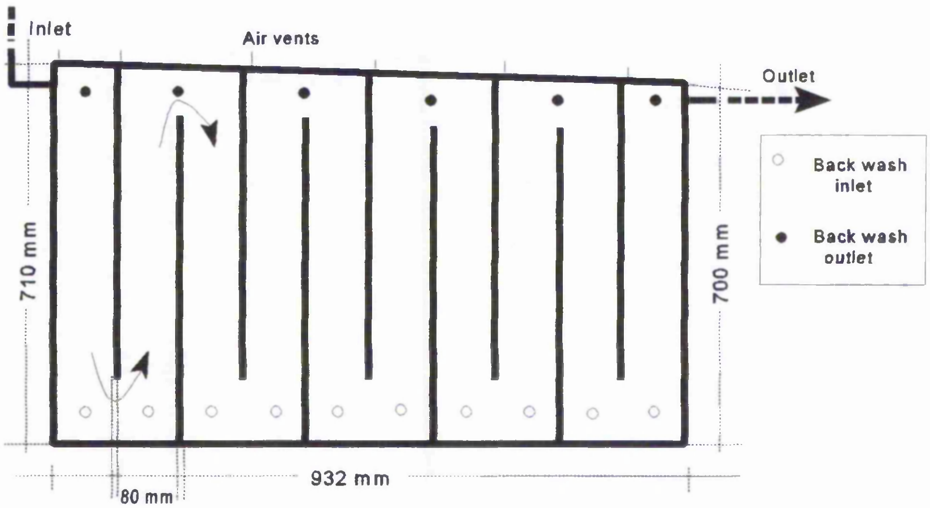
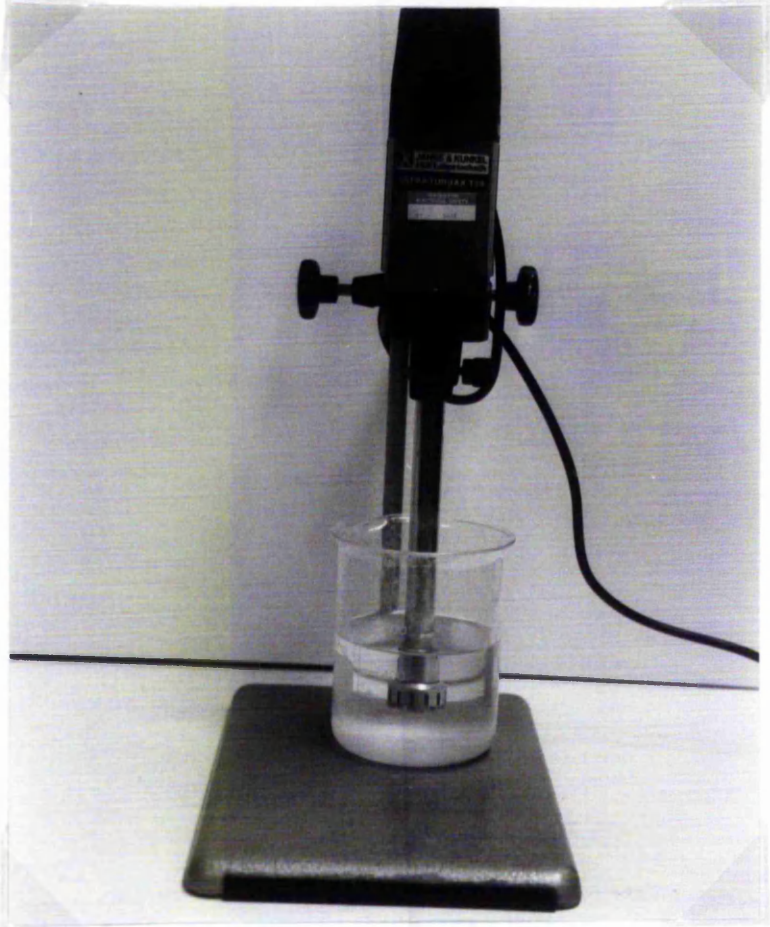


Fig 3.6 Plan of the model filter apparatus.



3.7 Photograph of the homogeniser

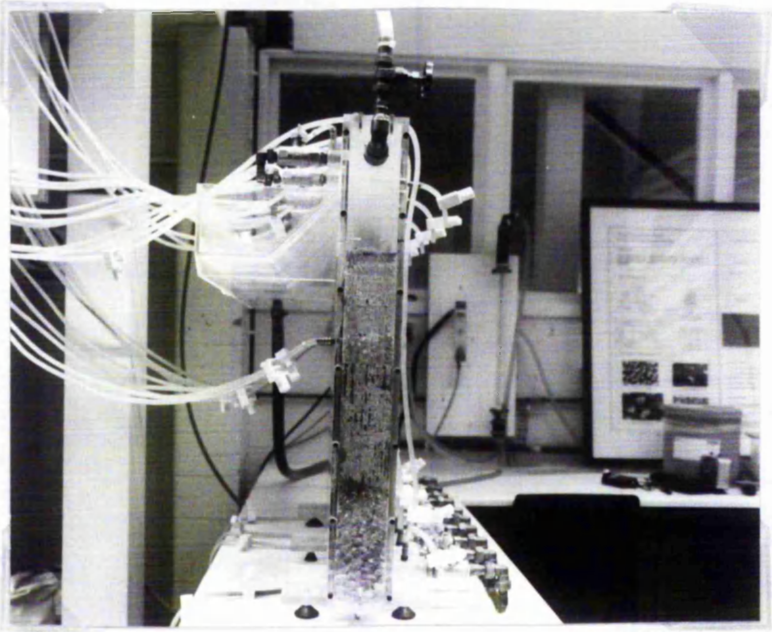


Fig 3.8 Vertical section of the model filter apparatus.

3.2.3 PERIPHERAL EQUIPMENTS

The peripheral equipments associated with the experimental work of this investigation will be described in this section.

a) *Flowmeters*— two flowmeters were necessary during the course of this work. A 500 mm long RA 90432-5 rotameter supplied by GEC. Elliot Process Inst., Purley Way, Croydon, UK was used during the backwashing process. A 0.2—3.0 l/min Platon flowmeter supplied by Platon Flow Control Ltd, Platon Park, Viables, Basingstoke, UK was used to control the inlet flow rates. The RA 90432-5 flowmeter had to be calibrated in order to rest assured of its reliability; this was accomplished by gravimetric weighing. No calibration was required for the Platon rotameter as this instrument had been previously calibrated prior to its acquisition. The calibration curve and statistical data for RA 90432-5 are contained in the appendix section.

b) *Homogeniser*— type Ultra-Turrax T25 S7, 200—250 volts, 50/60 Hz, 600 W (input) speed range 8,000—24,000 rpm, manufactured by Janke & Kunkel GmbH & Co. KG, Ikalabortechnik, Staufen, Germany. This homogeniser was used to disperse the synthetic turbidity discretely without adding any dispersing chemical agents, to ensure thorough dispersion of the turbid water. The suspension was sucked in from the lower part of the mixing tube by the rotor and discharged out at the sides of the tube through the static slots. The rotor speed was selected and controlled electronically. A speed of 8,000 rpm was selected and used to prepare the suspension during the experimental course of this investigation, following the practical advice of Dr. M. Kazi who had previously carried out similar tests with the very same homogeniser and synthetic turbidities. Figure 3.7 shows a photograph of this homogeniser.

c) *Turbidimeter*— this equipment was a product of Hach Chemical Company, Ames, Iowa, USA, microprocessor-based Model 2100 AN, furnished with six sample cells which were maintained in a clean condition. The sample size for all turbidity measurements were approximately 30 ml. Figure 3.11 shows the photograph of this instrument. Turbidity ranges that could be measured were 0—2, 0— 20, 0— 200, 200— 4,000 and 4,000— 10,000 NTU (Nephelometric Turbidity Units). This instrument was operated on the principle that light, passing through a substance, is scattered by particulate matter suspended in the substance. The optical system (shown in Figure 3.9) consisted of a tungsten-filament lamp, lenses and apertures to focus the light, a 90° detector to monitor scattered light, a forward-scatter light detector, a transmitted-light detector and a back-scatter light detector (Ratio method). The benefits of using the Ratio Measurement Method include improved linearity, calibration stability, wide measurement range and the ability to measure turbidity in the presence of colour.

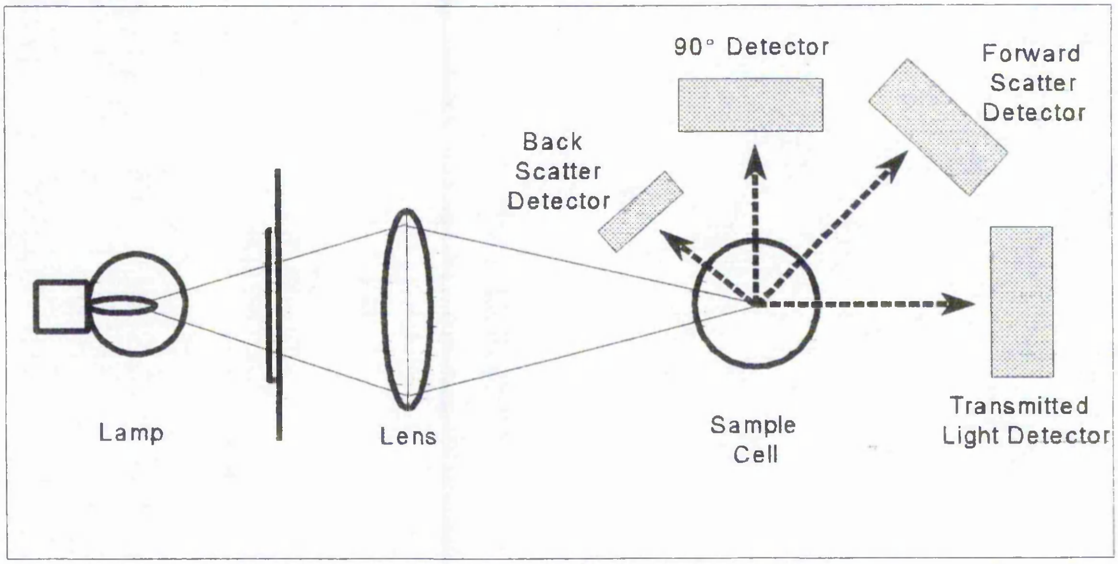


Figure 3.9 Optical system of Model 2100 AN Hach Turbidimeter, after the optical diagram shown in the apparatus instruction manual.

There are two standard specifications for turbidity measurement which are generally in use worldwide. These are the international standard ISO 7027 (1984 ed) and the USEPA (method 181.1). The specification for the ISO standard is more stringent and requires the use of a monochromatic light source. This specification allows for greater reproducibility of the measured values and greater agreement between other measuring instruments. Unfortunately, the 2100 AN Model used during the course of this investigation only meets the design criteria of the USEPA standard.

As the instrument was very sensitive to dirt and finger prints left on the glass tube during handling of the sample, or even minor imperfections and scratches, all the sample cells were cleaned both inside and outside and the outside was wiped with tissue paper before each reading. Instrument calibration was achieved with the aid of five pre-calibrated Gelex Secondary Standards, supplied with the turbidimeter. This substance was a metal-oxide particle suspension formulated to correspond to Formazin primary turbidity standards in their light scattering characteristics.

Turbidity is a very complex analytical measurement which can be affected by many factors, as already described. An important consideration is the comparison of readings between different instruments calibrated with the same standard. For reasons described in this thesis, it is unreasonable to measure the same sample with more than one turbidimeter. On the other hand, measurement of turbidity generally provides an immediate estimate of the relative quantity of suspended solids. For direct suspended solids measurement in mg/l for each type of turbidity material, turbidity readings must be supported by auxiliary measurements and appropriate calibration curves had to be established (*see Appendix 3*).

d) In addition, a pump, thermometers, burettes and pipettes, beakers, dishes, glass bottles, electronic balances, sieves, etc. were used during the experimental course of this investigation.

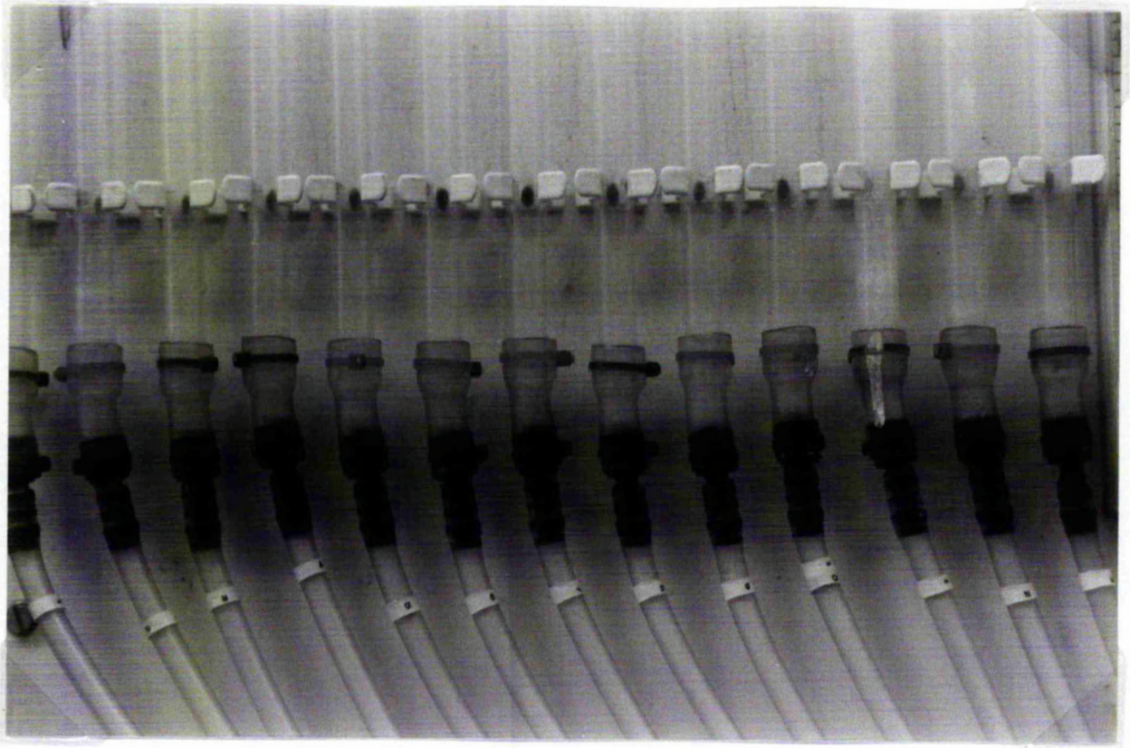


Fig 3.10 Detail photograph of head loss gauges and tubes connection at bottom level.

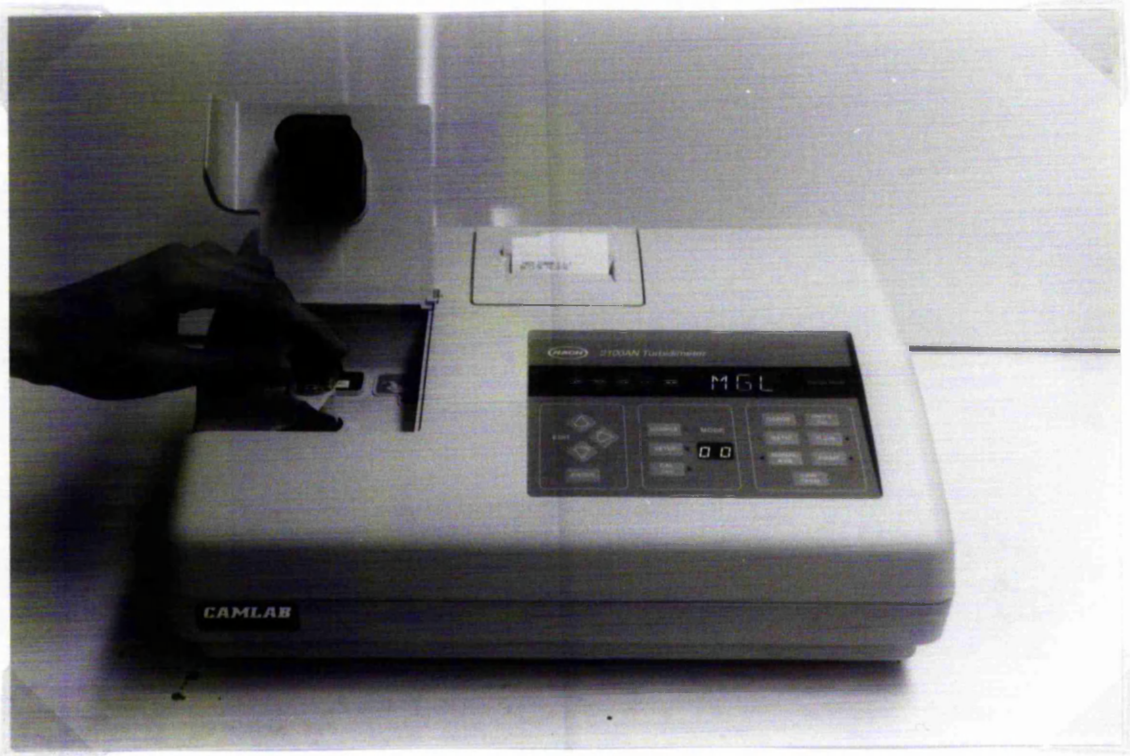


Fig 3.11 Photograph of the turbidimeter

3.3 SYNTHETIC TURBIDITY MATERIALS

Three types of synthetic turbidity materials were used in this investigation of rapid filtration. These turbidity materials were: Kaolin, Lycopodium powder and Polyvinyl Chloride powder (PVC). The physicochemical properties of these materials will be described in the following subsequent sections and are summarised in Table 3.1.

Table 3.1 Physicochemical Properties of the Turbidity Materials

Turbidity Material	Size in µm	Density in kg/m^3	Description
Kaolin	2— 53 (2 µm 80%; up to 53 µm 20%)	2,600	Heavy clay
Lycopodium powder	35	1,180	Dead fungus cells
PVC powder	0.5— 1.5	1,400	Long-chain polymer

3.3.1 POLYVINYL CHLORIDE POWDER

This material is commercially known as Corvic PVC Polymer. It was supplied by EVC UK Ltd., Kings Court, Manor Farm Road, Runcorn, UK. This material consisted of long-chain molecules, particles are irregular in shape and the size range is 0.5 to 1.5 µm with a density of 1,400 kg/m^3.

3.3.2 KAOLIN

This material was supplied by Fisons Scientific Equipments, incorporating Griffin & George, Bishop Meadow Road, Loughborough, UK. Kaolin is a heavy clay powder of which *kaolinite* (a finely crystalline form of hydrated aluminium silicate, formed mainly by the weathering of feldspar) is the main constituent, 2 to 53 µm in size (2 µm 80% and up to 53 µm 20%). The particles are irregular in shape with a density of 2,600 kg/m^3.

3.3.3 LYCOPODIUM POWDER

This material was supplied by Philip Harris Education, catalogue no. S 5103015, Lichfield, UK. Lycopodium powder is a fine yellowish flammable powder consisting of the spores of a lycopod (esp. *Lycopodium clavatum*). The size of the particles is uniform, at 35 µm and its absolute density is measured as 1,180 kg/m^3. Particles are floated on water without being wetted. It is generally used to demonstrate *Brownian motion* (i.e. erratic movement of extremely small particles in a colloid solution caused by the impact of molecules in the surrounding medium).

3.4 PREPARATION OF THE UNISIZE SAND MATERIAL

A number of properties of filter media are important in affecting filtration performance and in defining the media. These properties include size, shape, density, hardness and porosity (see *Chapter 1, Section 1.4.4*).

Uniformity in sand size was obtained by mechanical sieving; the standard procedure for conducting sieve analysis of a filter medium is based on the BS 410:1986 Specification for Test Sieves from the British Standards Institution. This standard lists tolerances and gives relevant definitions and an outline of inspection procedures, aperture sizes for wire cloth and perforated plate (including 7 non ISO sizes for round holes) in sieves for testing the size distribution of granular products in the particle size range from 125 mm to 32 µm. The mean diameter was taken as the geometric mean of the two sieve sizes (material passing through the higher sieve size and retained on the lower).

The filtering material used in this investigation was *Leighton Buzzard Sand*, supplied by EEC Quarries (George Garside Sands), UK. As mentioned above, the filtering media were sieved through the relevant adjacent sizes of sieves placed in a motor driven shaker. The procedure for the preparation of the unisize sand is briefly described next.

Table 3.2 Characteristics of Leighton Buzzard Sand

BSS Sieve No	D_G (µm)	T (°C)	µ (kg/m·s)	ρ_W (kg/m^3)	C_D	V_s (m/s)	R	D_h (µm)	ψ
600/500	548	16.25	0.001088	998.926	1.643	0.076997	31.304	443	0.65
710/600	653	16.75	0.001095	998.839	1.385	0.089669	41.424	506	0.60
850/710	777	17.00	0.00109	998.795	1.198	0.102369	53.546	571	0.54
1000/850	922	16.25	0.001088	998.926	1.094	0.111659	63.558	620	0.45
1180/1000	1086	16.00	0.001	998.970	0.9580	0.124	82.986	669	0.38

Characteristics of filtering media and settling velocity test for each unisize grain type used in this research work, where D_G is the geometric diameter, T temperature of water, μ water dynamic viscosity, ρ_w density of water, C_D Newton's drag coefficient, V_s settling velocity, R Reynolds number, D_h hydraulic diameter, ψ sphericity. The density of sand (ρ_s) was 2650 kg/m^3.

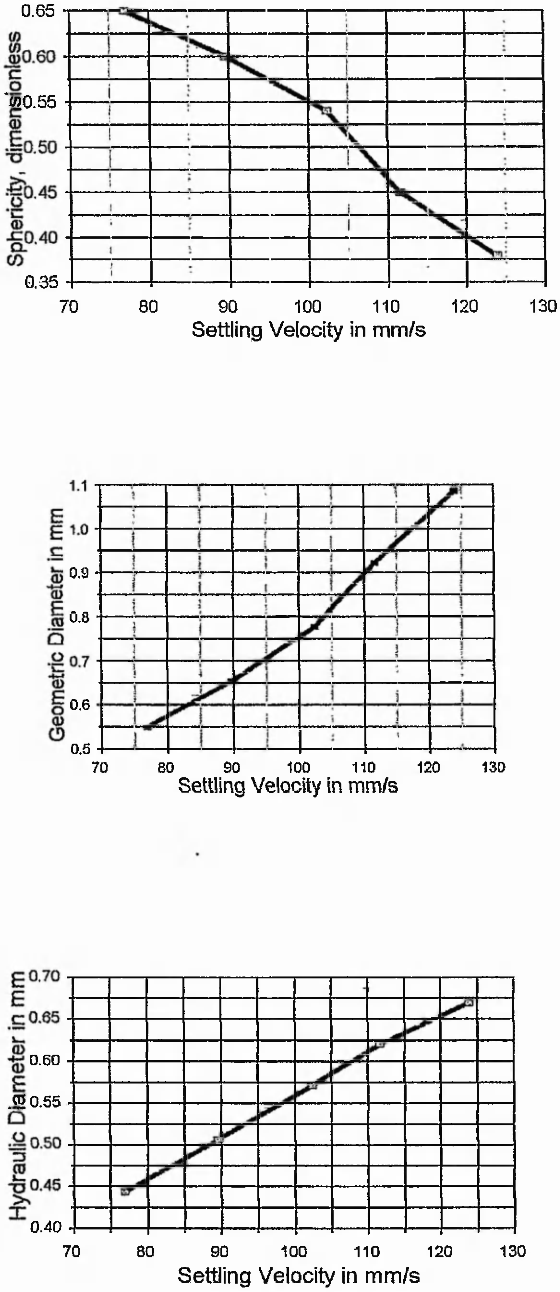


Fig 3.12 Effect of sand physical characteristics on settling velocity, after data from Table 3.2

After the filtering material had been obtained by mechanical sieving, weighed and carefully introduced into the various filter compartments, the settling velocities of the particles in water were determined by taking the average of 100 sand particles for each type of unisize grains by allowing the unisize sand grain to settle freely in a test column without the interference from adjacent particles (i.e. *unhindered settling*) or wall effect of the settling tube.

Since the Reynolds number R was found to be greater than one and less than 10^4 for each set of grains (as shown on Table 3.2) it was assumed that the grain particles were moving down under transitional flow conditions, and the *general settling velocity formula* was then applied:

$$V_s=[\frac{4}{3}\frac{g}{C_D}\frac{\rho_s-\rho_w}{\rho_w}D_h]^{½} \qquad [3.1]$$

where C_D = Newton's *drag coefficient* related to R and defined as follows:

$$C_D=\frac{24}{R}+\frac{3}{R^{½}}+0.34 \qquad [3.2]$$

The *hydraulic diameter* D_h for each group of unisize grains was calculated by applying the trial and error method (D_h calculated had to match D_h assumed and up to three decimal figures). D_h calculated represents the actual hydraulic diameter of the equivalent sphere of identical velocity to the grain with geometric mean diameter D_G. The value of the *sphericity ratio* Ψ for each group of unisize grains had to be calculated as follows:

$$\Psi = \left(\frac{D_h}{D_G}\right)^2 \leq 1 \qquad [3.3]$$

Figure 3.12 shows the effect of sand physical characteristics (sphericity, geometric diameter and hydraulic diameter) on settling velocity. Porosity was calculated from the dry weight of material used, the volume occupied in the filter, and specific gravity. The initial porosity was determined after vibration, so as to obtain a fully consolidated condition. Theoretically, for the same type of packing, the porosity should be independent of the diameter. But in an actual random packing the porosity increases with the decreasing diameter of particles. This has been found to be the case for the sand used in the experiments, as shown in Table 4.1.

3.5 PREPARATION OF SUSPENSION

The synthetic turbidity requirements were met by using kaolin, lycopodium powder and PVC powder, as described in section 3.3. A high-speed dispersing and emulsifying apparatus, the IKA Ultra Turrax T 25 homogeniser was used to prepare the homogeneously dispersed suspension of discrete particles. The medium was sucked in from below by the rotor and came out at the sides through the static slots. The speed range of the homogeniser was 8,000 to 24,000 rpm. A rotor speed of 8,000 rpm was used for 120 seconds to disperse the particles homogeneously. The procedure for the preparation of the suspension is described below:

- A known quantity of the turbidity material was weighed in a 250 ml beaker.
- 800 ml tap water was put into a 1000 ml beaker secured in the homogeniser with a strap clamp.

- The dispersing tool of the homogeniser was immersed into the water in such a manner that its clearance from the bottom of the beaker was 15 mm to allow the suction and suspension entry to the rotor.

- The synthetic turbidity was added to the beaker water and the drive unit of the homogeniser was then started at a speed of 8,000 rpm and allowed 120 seconds to disperse the turbidity materials homogeneously.

- The prepared suspension was then added immediately to the water in one of the mixing tanks and kept in a constant agitating state in order to maintain the homogeneous dispersion. The same procedure was then repeated again for the second mixing tank. The experimental set up was then ready for the start of the filter run.

3.6 RATES OF FLOW

The rate of flow through a filter may be expressed as follows:

$$\textit{Rate of flow} = \frac{\textit{volumetric flow rate}}{\textit{plan area of filter}} \qquad [3.4]$$

At the start of the filter run, the driving force must overcome only the resistance offered by the clean filter bed and the underdrain system. As suspended solids start to accumulate within the filter, the driving force must overcome the resistance offered by the clogged filter bed and the underdrain system (G. Tchobanoglous *et al.*, 1991).

The range of volumetric flow rates used in this investigation was 0.2 to 1.0 l/min. The flow through the filter was maintained at a constant rate for all the experiments, this actually being 0.2, 0.4, 0.6, 0.8 and 1.0 l/min (1.9, 3.8, 5.6, 7.5 and 9.4 $m^3/m^2 \cdot h$) for each material and initial concentration. The rate of filtration was carefully controlled by means of an outlet flowmeter and an outlet needle valve that allowed the author to manually adjust the flow through the filter unit.

At the beginning of the run, a large portion of the available driving force was dissipated at the valve, which was almost closed. The valve was opened as the head loss began to build up within the filter during the run. The rate of flow was then manually adjusted by using the rotameter described in section 3.2.3.

3.7 TEMPERATURE

Temperature plays an important role in filtration. Early experiments indicated that the results were sensitive to quite small temperature changes (Sholji, 1963), consequently the water temperature was recorded as it changed from day to day with the environmental changes throughout the year.

Temperature affects efficiency by influencing the rate of chemical reactions, the viscosity of water and hence the particle settling velocity. Temperature can also be a surrogate for change in other parameters that occur on a similar seasonal basis. Changes in alkalinity, colour, turbidity and orthophosphate concentration affect coagulation reactions and the properties of and rate of settling of resulting floc particles (Gregory and Zabel, 1990).

If the viscosity of water at 20 °C is taken as a reference, decreasing the temperature to 3.5 °C increases the viscosity by 58% and increasing the temperature to 33.0 °C decreases the viscosity by 26%. Because of the fairly steep negative gradient of the viscosity-temperature graph there are substantial viscosity changes between summer and winter temperatures (Figure 3.13). Controlling the temperature in this laboratory was not entirely possible. However, certain steps were taken in order to minimize these seasonal temperature changes by filling the water tanks the night before allowing this way the suspension to reach room temperature by the following day; the whole laboratory was air-conditioned to a constant 20 °C in winter, when the outside temperature could range from 10 °C plus to well below freezing.

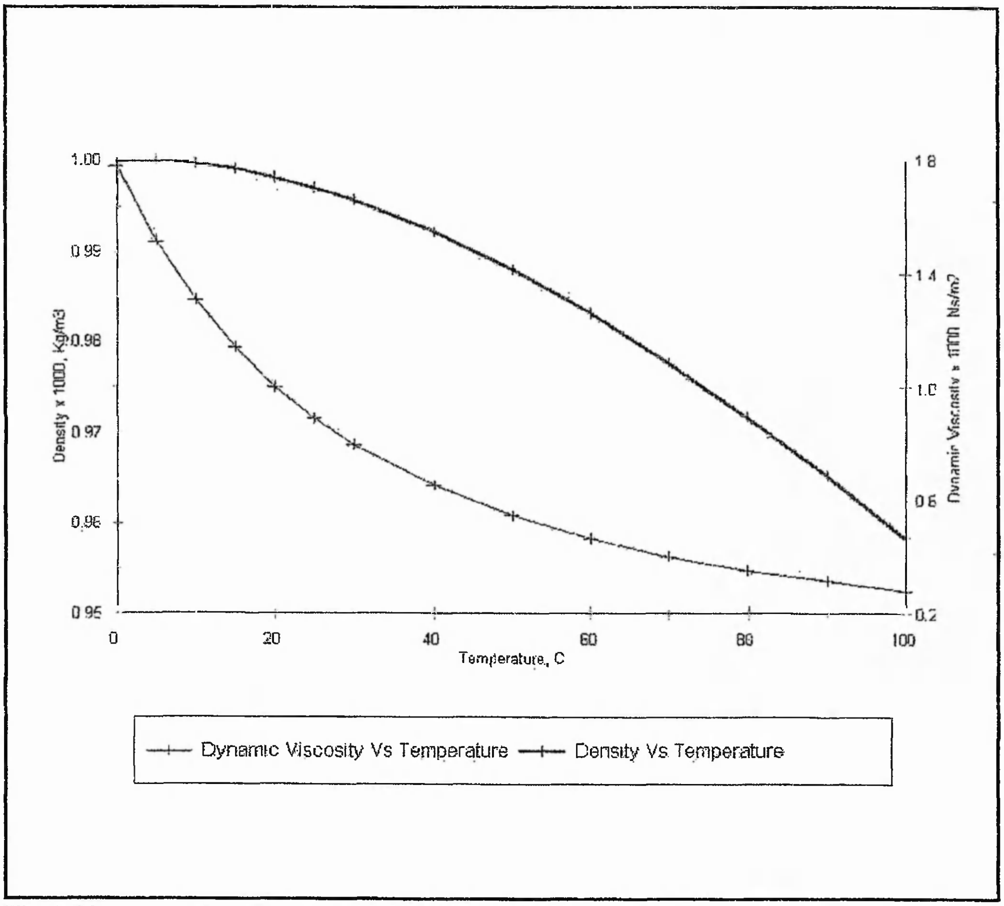


Fig 3.13 Density and dynamic viscosity of liquid water as a function of temperature, after M. Crawford, 1976.

3.8 TAP WATER CHEMISTRY

The chemistry of tap water plays a significant role in filtration. The supplied water analysis reports for the period 18/10/1995 to 16/10/1996 are not available from Severn Trent Water Plc, for the area where the laboratory of this work is situated.

However, a comparison of the analysis of tap water used for a previous investigation was made with the supplied water for the period 01/01/1993 to 31/08/1994. The results of this comparison are shown in Table 3.3, and they are the average value of around 300 samples tested during the course of the experiment (Kazi, 1995, Ph.D. thesis on pneumatic flocculation at the Nottingham Trent University, UK). In 1994 Severn Trent Water Plc achieved a record 99.8 per cent compliance with EU drinking water standards and this could be taken to mean that the company's standard of water treatment remains strong.

Table 3.3 Analysis of Tap Water

Parameters	Severn Trent Water Plc				Nottingham Trent University		
	Period	Min.	Max.	Average	Min.	Max.	Average
Turbidity (NTU)	1993	0.1	1.0	0.2	0.4	0.4	0.4
	1994	0.3	0.3	0.3			
pH	1993	7.3	8.3	7.8	7.01	8.10	7.66
	1994	7.4	8.2	7.8			
Temperature (°C)	1993	6.5	22.9	12.0	8.70	22.60	14.00
	1994	5.5	20.1	12.5			
Alkalinity mg/l as $CaCO_3$	1993	110	146	124	110	150	125
	1994	85	125	108			

Modified after Kazi, 1995.

3.9 SAMPLING TECHNIQUE

The sample collecting ports were designed bearing the consideration in mind to collect truly representative samples as far as possible. Sixteen sample collecting pipes were inserted into the filter box to collect the samples for residual turbidity measurement. Details of the sampling ports have been shown in Figure 3.3. Six of them were allocated at the upper part of the filter box (the inlet of these sampling ports was projected 40.0 mm from the inside wall of the filter box), and the rest at the bottom of each compartment.

Each sampling port consisted of two parts: the inner part was a 3.0 mm diameter flexible PVC pipe inserted in a way to ensure a proper length from the inner wall ; the outer projected part of the sampling port was connected to a short 5.0 mm diameter rubber tube equipped with a pinch-cock to enable the collection of water samples.

The sample collecting pipes were flushed every time before collecting the sample to avoid any deposition of turbidity material inside the pipes. Sampling of the suspension was carefully conducted, so that no disturbance was caused to the filter or the deposits accumulating in the pores. However, disturbance to the inlet pipe could happen during the collection of samples and any deposited turbidity material on the upper part of the projected sampling tube could alter the next downward sampling port.

For all filtration rates, samples were collected at equal intervals of 15 minutes for the first hour; afterwards, this interval would be increased to 30 minutes over a total filtration time of more than 300 minutes, except for 1.0 l/min and 0.8 l/min (due to a higher filtration rate, the suspension contained in the mixing tanks would not allow the experiment to run longer than 120-240 minutes).

Collection of samples was started from the lower port in compartment number 1 ending at the upper one (or outlet) in compartment number 10 (Figure 3.14). The displacement time, or time needed for suspension to travel from compartments 1 and 2, and so on, was taken into account. Thus, the displacement time was calculated for every filtration rate in order to establish a proper timing for the collection of samples.

3.10 SCHEDULE OF EXPERIMENTS

Seventy-five experimental runs were conducted to investigate the efficiency and applicability of rapid filtration by a multi-compartment up-flow and down-flow process at different filtration rates, type and concentration of synthetic turbidity materials. The details of the experimental runs are described in Table 3.4.

Table 3.4 Details of the experimental runs

PVC			Kaolin			Lycopodium		
Run No.	Initial Concentration (mg/l)	Rate of Flow (l/min)	Run No.	Initial Concentration (mg/l)	Rate of Flow (l/min)	Run No.	Initial Concentration (mg/l)	Rate of Flow (l/min)
1	400	1.0	26	400	1.0	51	400	1.0
2		0.8	27		0.8	52		0.8
3		0.6	28		0.6	53		0.6
4		0.4	29		0.4	54		0.4
5		0.2	30		0.2	55		0.2
6	300	1.0	31	300	1.0	56	300	1.0
7		0.8	32		0.8	57		0.8
8		0.6	33		0.6	58		0.6
9		0.4	34		0.4	59		0.4
10		0.2	35		0.2	60		0.2
11	200	1.0	36	200	1.0	61	200	1.0
12		0.8	37		0.8	62		0.8
13		0.6	38		0.6	63		0.6
14		0.4	39		0.4	64		0.4
15		0.2	40		0.2	65		0.2
16	150	1.0	41	150	1.0	66	150	1.0
17		0.8	42		0.8	67		0.8
18		0.6	43		0.6	68		0.6
19		0.4	44		0.4	69		0.4
20		0.2	45		0.2	70		0.2
21	100	1.0	46	100	1.0	71	100	1.0
22		0.8	47		0.8	72		0.8
23		0.6	48		0.6	73		0.6
24		0.4	49		0.4	74		0.4
25		0.2	50		0.2	75		0.2

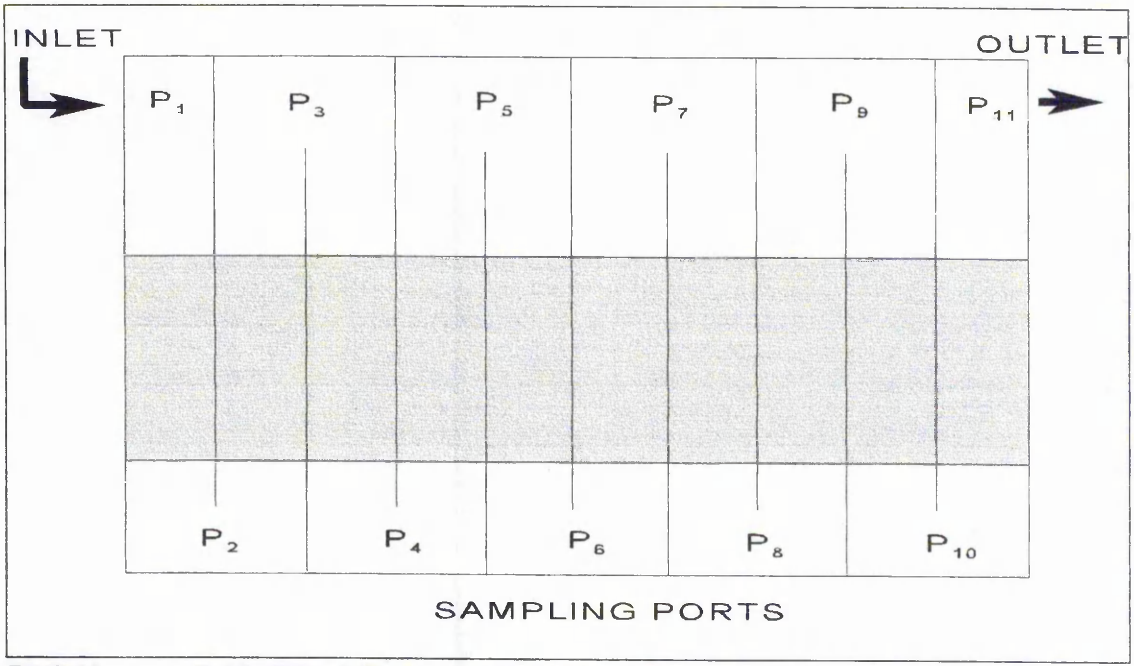


Fig 3.14 Schematic of filter's sampling ports.

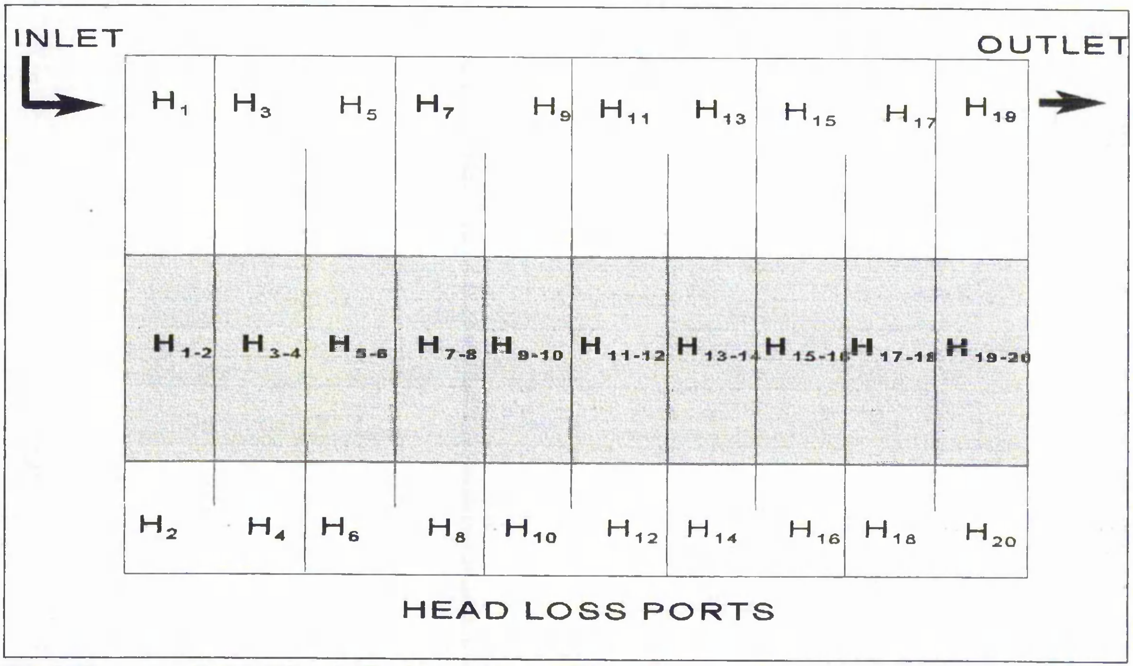


Fig 3.15 Schematic of filter's head loss ports.

3.11 EXPERIMENTAL PROCEDURE

The experimental runs were conducted following the schedule described in section 3.10 and shown in Table 3.4, to investigate the removal efficiency of rapid filtration by a multi-compartment up-flow and down-flow process.

Three types of turbidity materials namely PVC powder, kaolin and lycopodium powder were used to prepare the suspensions which were introduced into the system under different initial conditions as presented in Table 3.4.

The tap water from the Environmental Engineering laboratory —supplied by Severn Trent Water Plc— was used to create the homogeneously dispersed inflowing suspension consisting of turbidity particles of known characteristics, as described in sections 3.3 and 3.5.

Samples were collected from eleven number of collecting ports installed, as shown in Figure 3.14, to determine the residual turbidity at different intervals of time: every 15 minutes during the first hour of the filter run, and then at an interval of 30 minutes. Likewise these sampling ports acted as manometric pressure probes that provided the measurement of the head loss of the system.

Every 30 minutes during a filter run readings of water temperature and head loss across the filter were taken, and water samples from the filter were analysed for residual turbidity, as described in section 3.2.3, over a filtration period of 300 minutes on average.

The step by step procedure is described as follows (the filter system is shown in Figure 3.1):

- Following the procedure described in section 3.5, 0.8 litre of suspension was made in a 1 litre beaker to make prepare the turbid water at a concentration of 100, 150, 200, 300 or 400 mg/l in the mixing tanks and kept ready for the test. A homogeniser with a speed of 8,000—24,000 rpm was used to disperse the synthetic turbidity material discretely, without adding any dispersing agent.

- Turbid water was kept constantly agitated by a motor-driven stirrer inside the mixing tanks, from where it was pumped to the top of the filter or constant head tank. The filter box was filled with turbid water to an operating water height of 2000 mm above the inlet level. Initial concentration, water temperature and porosity were measured during the filling of the filter box. These parameters were also measured at the end of the filter run. The purging of air from the filter unit preceded the addition of the homogenised prepared synthetic suspensions.

- The outlet needle valve was opened slowly; the constant head tank supplied the suspension to the filter by gravity at the required rate of flow. After 5 minutes a reading of the head loss of the system was taken; this value was considered to be the initial head loss due to the hydraulic resistance of the clean filter.

- The rate of flow was checked and manually adjusted with a needle valve with reference to a rotameter to maintain the flow rate required for the test. Filtration rates, constant for any one experiment, were used as follows: 0.2, 0.4, 0.6, 0.8 and 1.0 l/min.

- At the time of collecting samples great care was taken to ensure a slow rate of draw-off so that the particles depositing inside the filter bed were least disturbed. Readings of the head loss were taken every 30 minutes. The filter runs averaged about 300 minutes (depending on the rate of flow used). A filter run was terminated due to an excessive head loss or after having the turbid water supply completely depleted. At this point, the filter had to be backwashed.

- The sequence of cleaning the bed was started by closing the influent valve and opening the wash-water outlet valve, thus draining out the water overlying the bed and decompressing the filter box. After the close of the outlet valve, the wash-water inlet valve can be opened, allowing wash water to flush impurities out of the fluidized bed for a period of 20 minutes. The clean media are allowed to settle down in quiescent water before closing and opening appropriate valves to restore filtration. From the obtained data, it was possible to plot the variation of the parameters such as head loss or concentration ratio with time for each filtering velocity, and evaluate the filter's removal efficiency following the procedure described in section 3.12.

3.12 EXPERIMENTAL OBSERVATIONS

Samples were collected for analysis from different depths of the filter bed at equal intervals of 30 minutes over a total filtration period of 300 minutes on average except for the first hour when samples were collected every 15 minutes in order to achieve a better characterization of the filter performance during the early stages of the process; this will also be discussed in chapter 4. The *displacement time*, or time required for suspension to displace the clean water initially present in the pores, was logically taken into account, as described in section 3.9 (Sholji, 1963).

Residual turbidity was measured in NTU which was based on the optical effect of the suspended matter present in the sample. These readings in NTU were then automatically converted into the actual amount of synthetic turbidity materials in mg/l using the calibration curves shown in Appendix 1 which were constructed during the early stages of this investigation. From these readings the concentration changes with times through the various compartments were plotted. Also, the head loss curves varying with time and depth of filter.

Turbidity removal efficiency of collected samples at different depths and times was then calculated. The removal efficiency (RE) of each of the filtering compartments is given by:

$$RE = 100\left(1 - \frac{C_1}{C_0}\right) \qquad [3.5]$$

where RE= removal efficiency, as a percentage

C_1 = effluent concentration, mg/l

C_0 = influent concentration, mg/l

Analysis of samples taken from the sampling ports of the filter indicated that a progressively smaller percentage of turbidity particles were removed as the time of filtration progressed, and this removal was confined to compartments 1 to 4 of the filter. This pattern was common to all runs. This will be discussed in more detail in chapter 4.

There is insufficient space to present all the experimental data sheets and the concentration and head loss curves for both time and depth. However, a few experimental data sheets covering representative filtration runs of all the three types of synthetic turbidity materials, are shown in Appendix 2. The corresponding concentration and head loss curves are also presented as shown in Appendix 3.

CHAPTER 4

ANALYSIS OF THE EXPERIMENTAL WORK

4.1 GENERAL

The main factors affecting rapid filtration are (Ives and Sholji, 1965):

- depth of media
- grain size media
- grain material
- rate of filtration
- inflow concentration of suspended particles
- type of suspension
- water temperature
- initial porosity

The effects of the above parameters on efficiency of turbidity removal will be analysed and discussed in this chapter.

The conventional down-flow rapid filter has remained unchanged in its basic form for more than 80 years. Characteristically, the filter is made of fine heterogeneous granular material such as sand.

The reverse flow of water during backwashing tends to grade the sand grains and hydraulic stratification takes place, so that the finest are at the top and the largest at the bottom of the bed. Raw water therefore meets the finest sand first and goes out through the coarsest. This causes the separation of suspended solids from the liquid when filtration takes place mainly within the top surface layer of the bed, with the bottom of the filter bed acting largely as a support, which stays clean without sharing the burden of the removal of suspended solids.

Over the past decades, water filtration researchers have used a variety of approaches in order to deal with this problem. The obvious one is up-flow filtration, where the clarification of water is achieved by filtering the influent through successively coarser to finer layers of sand, such as the *contact clarifier* or the *biflow filter*. In the up-flow filters, however, fluidisation of the bed has to be prevented, either by a grid in the sand surface, or by applying a rate of flow less than the incipient velocity or the minimum fluidisation velocity.

Another way of tackling the problem is to use two or more media of different densities so that light but coarse grains settle above a layer of fine but heavy ones (*composite beds* or *multilayer filters*). The difference in density between anthracite and sand is large (*see Table 1.6, page 49*) and therefore these two materials have been used extensively in multilayer filtration.

However, the composite type of filter may present a major problem, and that is media intermixing. During and after backwashing, the composite bed may get mixed and remain so even after the termination of backwashing, providing no absolute separation of the different types of granular material used. In this mixed layer the small denser grains pack into the spaces between the large lighter ones. The bed will have therefore a different form of packing, mean porosity, and pore size that may influence the clogging of the filter and produce greater head losses.

As described in Section 3.2, the model filter apparatus was designed to have a larger size of unisize grain in the first two compartments, followed by smaller unisize grains in the subsequent compartments that follow, in order to maintain a coarse to fine grain filter.

The coarser unisize media carry the burden of removing larger suspended colloidal matter from the flow, throughout the depth of the filter layer, leaving the subsequent unisize compartments of the filter bed to follow, suit and share the burden of the removal of suspended impurities in depth, for both down-flow and up-flow under a constant rate of flow. The final section of the bed (compartments 6 to 10) was used as a polishing filter to remove any fine particulate matter that could penetrate the porous medium; this will be discussed in the next sections. In addition to this, the total filter bed is about 3,000 mm deep, significantly deeper than most common filters used in practice.

Thus, the model filter apparatus is a novel filter, combining both down-flow and up-flow filtration, and with a particle size gradation decreasing from about 1.2 mm in the first two compartments to about 0.5 mm in the final two. This eliminates both stratification and media intermixing (in case different materials were employed) and more closely approximates the idea of a uniform decrease in pore size space with increasing filter depth.

Extensive experimental work was scheduled and carried out during the course of this investigation in order to provide sufficient information to evaluate and examine the validity of rapid filtration by a multi-compartment up-flow and down-flow process.

Seventy-five experimental runs were conducted in conjunction with three different types of synthetic turbidity materials, five different inlet concentrations either 100, 150, 200, 300 or 400 mg/l and different rates of flow ranging from 0.2 to 1.0 l/min. The details of the experimental runs are shown in Table 3.4 and in Appendix 1.

Sampling of the filtered suspension was accomplished by collecting the discharge from 11 sampling ports, installed at the inlet and outlet points of every compartment; this was done very carefully so that sampling would not cause any disturbances to the filter or the accumulated deposits in the filter pores. The turbidity at each sampling port was determined every 15 minutes during the first hour of a filter run and at longer time increments thereafter.

Concentration-ratio curves were produced and drawn from the experimental data. All the NTU turbidity readings were converted to turbidity in mg/l by using calibration curves as mentioned in Section 3.12 and shown in Appendix 2. The efficiency of turbidity removal was determined as illustrated in Section 3.12.

Water temperature and head loss were also taken periodically during each filter run. The changes in head loss with time for various compartments were plotted as mentioned in Section 4.4.

Rapid filtration finds its greatest application to clarification of dilute suspensions (less than 500 mg/l) of particles ranging in size from about 0.1 μm to about 50 μm. Higher concentrations would be better treated by cake filtration; smaller particles should be flocculated, larger particles should be strained or sedimented (Ives, 1970).

It is widely accepted that the physicochemical properties of the synthetic turbidity materials influence the filtration process in removing the turbidity from water. Three types of synthetic turbidity materials were used: polyvinyl chloride powder (PVC), kaolin and lycopodium powder.

The choice of these materials was a deliberate one, since they range in size from 0.5 μm (PVC) to 53 μm (kaolin), and present various shapes and densities which allowed the assessment of the filter's performance when turbidities of different particle size, particle size range, shape and density were used (Table 3.1 on page 101 summarises the physicochemical properties of the turbidity materials used in this investigation).

The suspensions were prepared as described in Section 3.5, and were kept in suspension at constant agitation in the mixing tanks. To avoid any physicochemical change in the nature of the turbidity, the suspensions were used immediately after preparation.

It follows that particle properties, particularly suspended particle size, and suspended particle size distribution, are expected to play important roles in deep granular bed filtration. This is the reason for the contrasting behaviour of PVC, kaolin and lycopodium observed during the experimental course of this investigation.

For the general purposes of this work, no routine quantitative measurements were made to determine particle size distributions of the synthetic turbidities. However, the materials used in this research are identical (same batch) to the ones used in a previous investigation carried out by Kazi (1995) in his experiments in this same University's Environmental Engineering Laboratory on pneumatic flocculation. Consequently, the results of the measurements obtained by Kazi were taken as valid and representative of the three types of turbidity materials used by the author (*see Table 3.1*).

As explained in Section 2.2.1, three phenomena (Brownian motion, interception, and gravity sedimentation) have been identified as the principal mechanisms responsible for the transport of suspended particles from the bulk liquid to the medium surface. The importance of each transport mechanism also depends on the suspended-particle size and flow conditions.

Cleaning of the filter bed was accomplished by using a water backwash only. A 1.2—7.8 l/min RA-90432-5 rotameter, as described in section 3.2.3, was utilised to control the flow applied to the filter media during the backwashing process. Since the analysis of the backwashing of the filter was not an objective of this study, the cleaning of the various filter compartments was achieved with no attempt to evaluate the efficiency of backwashing or describe the backwashing technique.

However, great care was taken to ensure a proper cleaning of the filtering media during backwashing. High rates and large quantities of water had to be used to flush the filter completely whenever significant deposits of suspended matter accumulated at the inlet surface of the first compartment. This difficulty did not occur in the remaining compartments. Besides, washwater becoming clear, emerging from the filter did not necessarily mean that all deposits had been removed; stopping of backwashing and restarting washing after a lapse of time showed further washwater cloudiness, although this was less likely if there had been a prior air scour, as reported by Ives (1980).

It is worth mentioning that the overall backwashing process had to be carried out in five successive turns of 10 minutes each, two compartments of identical granulometry, due to insufficient water pressure. Backwashing rates ranged from 7.8 l/min (73 m^3/m^2h) to 3.1 l/min (29.1 m^3/m^2h) for the coarsest and finest medium, respectively. This flow range was found to be in agreement with Kawamura's predictions. However, problems of underbed gravel disturbance were observed, caused by uneven backwash distribution.

Kawamura worked out in 1975 a formula for simply calculating the *appropriate* rate of backwashing with water for both sand and anthracite. For sand, the *appropriate* rate is the 60 percentile weight size of medium in mm expressed in m/min. Applying this concept to the unisize media employed in this research results a backwashing velocity range of 72—30 m^3/m^2h. The *appropriate* velocity of backwashing with tap water was calculated to be that which was most effective in cleaning the medium by the optimum combination of shear and scour forces of the fluidising water. Kawamura suggested that a lower rate would be acceptable if an auxiliary scouring system was used.

However, Kawamura's approach did not differentiate between the type of solids present in potable water (chemically coagulated flocs) and that present in secondary sewage effluent (biological flocs formed in the biological oxidation of sewage). The different behaviour of these two types of solids has a profound effect upon their ease of removal (Jago, 1977).

Although porosity as a variable has a marked effect on filter performance, as shown by Sholji (1963), in this investigation the initial porosity of each type of unisize sand has been assumed to be constant for all runs, as shown in Table 4.1. The initial porosity was determined after applying a hand tap pulsation on the filter shell, so as to obtain a predetermined initial porosity reading.

Also, from this table it can be seen that porosity differences between compartments containing the same sand grain size can be significant. It appears that slight differences in initial porosities (about 1 per cent) can have a pronounced effect on the filtrate quality, as described in Chapter 1, whether between compartments or between runs. This would explain the lower turbidity removal efficiency observed in up-flow filtration, as it will be discussed in more detail in Sections 4.4 and 4.5.

To minimize the differences in the initial porosity values, the filter bed was packed to a uniform porosity for each filter run by carefully tapping the filter shell of each compartment to a predetermined sand bed depth, as shown in Table 4.1 . This was used because it is the most easily reproducible. However, in an actual filter the porosity might be higher than that determined in these tests, due to the free fall of filter grains after backwashing, depending on the gentle or sudden shut-down of washwater valve.

In this investigation, the removal efficiency and the assessment of the filtration process were based on the dimensionless concentration-ratio versus time and depth. The head loss analyses were considered for each filter compartment (depth) versus time. The author did not elaborate on the evaluation of the filter coefficients λ_0, c and ϕ, as shown in Equations 1.15 and 2.8.

Table 4.1 Filter-Medium Characteristics

Grain Size D_G (μm)	1086		922		777		653		548	
Depth of Bed (mm)	310	295	281	290	305	300	309	304	309	320
Initial Compacted- Bed Porosity	0.38	0.35	0.32	0.34	0.37	0.36	0.38	0.37	0.44	0.46
Compartment	1	2	3	4	5	6	7	8	9	10

4.2 CONCENTRATION VERSUS TIME

The filter effluent quality throughout a filtration run is considered. During the filter run the efficiency of the filter changes due to the accumulation of deposits in the pores. Initially, the quality of filtered water is improving until it reaches acceptable standard level, and may also deteriorate near the end of the cycle if the run is prolonged for a sufficient time. Then, if water quality does not match the required standard, the filter must be stopped and backwashed.

As expected, increases in the influent suspension concentration were found to cause corresponding increases in the outlet measured turbidity levels at each rate of flow. Filtration efficiency was found to increase with a decrease in filter flow rate in accordance with the generally accepted theory.

The filtration tests demonstrated a high efficiency of removal, and the filtrate readings (filter's effluent) were very often too low to permit an accurate plotting of the concentration-ratio curves. The average concentration ratio (C/C_0) varied from 0.57 to 0.12 for the first compartment (initial concentration ratio), and from 0.04 to less than 0.01 at the outlet of compartment number 10 (ultimate concentration ratio).

The greater variability in the initial concentration values when compared with those obtained at the outlet of the filter apparatus gives an indication of the trends observed during the course of this investigation. The lower values correspond to the removal of PVC turbidity, and this may be attributed to the fineness and smaller size range shown by the PVC particles, as it will be discussed in greater detail in the next sections.

Initially, and in order to assess the filter's performance properly, the experimental data collected during this investigation were first plotted as C/C_o changes with time for various rates of flow and turbidity materials for each compartment, as shown in Figures 4.41—4.52. The initial conditions for the tests for which the typical graphs are provided are included in the figures. From these curves it can be seen that the variation of concentration ratio versus time is of a nonlinear nature, and tends to become asymptotic after 15 or 20 minutes.

On some curves, the concentration ratio decreases consistently with time, particularly when using PVC suspensions; the graphs drawn in Figures 4.43 and 4.44 consist of smooth asymptotic curves for all compartments, approaching to zero as the filter run progressed.

It was observed that removal occurred initially in the upper part of the bed with particles penetrating deeper as the run progressed. This was noticeable for compartments 1 and 2 containing the coarser media.

The improvement and deterioration of filtrate quality phenomena are demonstrated in compartment number 1 at the highest filtration rate of 1.0 and 0.8 l/min in the case of PVC powder, when the highest inlet concentration was being used, and this is obvious as shown in Figure 4.41 and Figure 4.42.

During these tests the medium in the first compartment takes the burden of removing the suspended solids from the inflow until the filter's available storage capacity of deposit is depleted and the medium starts to let the suspended particles pass through; this point is marked by a steep deterioration of the filtrate quality, with the concentration ratio rapidly approaching to one, i.e. inlet concentration equals outlet concentration, and in practice this is not reached.

Under the conditions specified in Figures 4.41 and 4.42, this decline in filter effluent quality happened after only 60 minutes from the start of the filtration run. If the experiment were prolonged for a sufficient time, the saturation front would be carried on to the next compartment. In Figure 4.42, for example, an incipient saturation of compartment number 2 is already noticeable towards the end of the filter run.

It is evident from Figure 4.41 to Figure 4.52 that the best removal efficiency was obtained in most cases from compartments 1 and 2 (and also 3 and 4, for kaolin and lycopodium powder) and that afterwards there was proportionally little change in the improvement of filtrate quality. This aspect of filtration is discussed in detail later, in Section 4.3.

However, the results represented in these graphs did not highlight clearly enough the differences in the concentration ratio among the different compartments, particularly in those that contained the finer media, and therefore a different approach was needed.

Figures 4.1—4.15 show plots of filtrate concentration changes with time for P_{11} (i.e. filter's effluent) at varying inlet concentrations for any one rate of filtration, for PVC, kaolin or lycopodium turbidities. These semi-log graphs illustrate perfectly the effect of the initial concentration of suspended particles on the efficiency of the filtration process for any given flow rate: The higher the initial concentration the lower the overall filtrate quality, irrespective of the type of synthetic turbidity used. Furthermore, the lower the rate of flow the higher the overall filtrate quality.

Additionally, it can also be seen that the difference in the overall filtrate concentration of suspended particles between various initial concentrations for a given suspension is smaller as the rate of flow decreases (Figures 4.1—4.15).

Thus, for example, the overall filtrate concentration for PVC powder at 1.0 l/min was consistently above 10 mg/l for an initial concentration of 400 mg/l, whereas for an inlet concentration of 100 mg/l, the ultimate filtrate concentration was as little as 0.2 mg/l; when the rate of flow was 0.2 l/min, the overall filtrate concentration range was about 4.5—0.2 mg/l, under identical initial filtration conditions. The same pattern was observed for both kaolin and lycopodium turbidities, even if the high—low gap was much narrower than that for PVC. As an example, and under the same initial conditions, the observed range in the overall filtrate concentrations was 2—0.5 mg/l and 0.9—0.1 mg/l for kaolin, and 1.5—0.7 mg/l and 1.5—0.3 mg/l for lycopodium powder.

With the PVC suspension previously described, the filter efficiency (taking the filter apparatus as a whole) was found to vary from 96.7 to practically 100 per cent at the slowest rate of filtration and lowest inlet suspended solids concentration (Figures 4.1—4.5). In addition, the detected level of turbidity decreased in almost every case for kaolin and lycopodium powder. Kaolin showed the best removal efficiency, close to that of lycopodium powder (Figures 4.6—4.15). The reason for the higher overall filter efficiency for these two types of suspensions could be the higher density and greater size range of the kaolin particles and the larger size and lighter nature of the lycopodium particles, which would enhance the particles' chances of removal, as it will be discussed in greater detail in Section 4.5.

The removal of kaolin turbidity at high rates of filtration (i.e. 1.0—0.8 l/min) gave some odd results, as it can be seen from Figures 4.6 and 4.7, where better filtrate quality was achieved at higher inlet concentration. A possible explanation for this phenomenon could be the effect of the temperature on the removal efficiency, as described in Section 3.7.

Thus, looking at the experimental data for Run 26 (kaolin, C_0=400, Q=1.0 l/min, T=12°C) and Run 46 (kaolin, C_0=100, Q=1.0 l/min, T=16°C), a significant difference in temperature between runs (4°C) was noticed. This was consequence of a failure in the heating of the laboratory registered during the execution of the mentioned experiments.

Finally, the same data were plotted differently also in semi-logarithmic scale to compare the behaviour of the different materials under identical initial conditions, as reflected in Figures 4.16—4.40. Semi-log plots of C/C_0 versus time for typical runs are found in Appendix 1.

The overall filter efficiency of removing the kaolin and lycopodium powder suspended solids was found to be over 99 per cent in all the cases, as shown in Table 4.2 to Table 4.6. As mentioned before, PVC results were consistently above 96 per cent. To a first impression, the removal efficiency of the filter for both kaolin and lycopodium is seemingly independent of the applied concentration of particles. Furthermore, as the rate of filtration and inlet concentration of suspended solids decreased the initial differences shown in efficiency removal among the three synthetic turbidities became less apparent (see Figures 4.16—4.40).

One possible contributory reason for the marked improvement in the filter efficiency for both kaolin and lycopodium powder is considered to be the influence of their physicochemical characteristics on the filtering media; PVC particles are finer and more uniform, 0.5—1.5 µm in size, with a density of 1,400 Kg/m^3, compared with kaolin (2—53 µm, 2 µm 80%, up to 53 µm 20%, density 2,600 Kg/m^3) and lycopodium (35 µm, density 1,180 Kg/m^3). The importance of this effect on removal efficiency will be discussed in more detail in Sections 4.5 and 5.4.

Table 4.2 Overall removal efficiency versus rate of flow and inlet concentration for PVC, kaolin and lycopodium powder.

Inlet Concentration (mg/l)	Rate of Flow (l/min)	Type of Suspension	Removal Efficiency (%)	Run Number
400	1.0	PVC	96.7	1
		Kaolin	99.9	26
		Lycopodium	99.6	51
	0.8	PVC	96.8	2
		Kaolin	99.8	27
		Lycopodium	99.7	52
	0.6	PVC	97.1	3
		Kaolin	99.8	28
		Lycopodium	99.6	53
	0.4	PVC	98.2	4
		Kaolin	99.9	29
		Lycopodium	99.8	54
	0.2	PVC	99.3	5
		Kaolin	99.9	30
		Lycopodium	99.8	55

Table 4.3 Overall removal efficiency versus rate of flow and inlet concentration for PVC, kaolin and lycopodium powder.

Inlet Concentration (mg/l)	Rate of Flow (l/min)	Type of Suspension	Removal Efficiency (%)	Run Number
300	1.0	PVC	98.2	6
		Kaolin	99.8	31
		Lycopodium	99.8	56
	0.8	PVC	98.5	7
		Kaolin	99.7	32
		Lycopodium	99.8	57
	0.6	PVC	98.0	8
		Kaolin	99.8	33
		Lycopodium	99.8	58
	0.4	PVC	99.3	9
		Kaolin	99.7	34
		Lycopodium	99.8	59
	0.2	PVC	99.1	10
		Kaolin	99.9	35
		Lycopodium	99.8	60

Table 4.4 Overall removal efficiency versus rate of flow and inlet concentration for PVC, kaolin and lycopodium powder.

Inlet Concentration (mg/l)	Rate of Flow (l/min)	Type of Suspension	Removal Efficiency (%)	Run Number
200	1.0	PVC	98.4	11
		Kaolin	99.7	36
		Lycopodium	99.6	61
	0.8	PVC	98.8	12
		Kaolin	99.5	37
		Lycopodium	99.7	62
	0.6	PVC	99.6	13
		Kaolin	99.8	38
		Lycopodium	99.7	63
	0.4	PVC	99.5	14
		Kaolin	99.8	39
		Lycopodium	99.4	64
	0.2	PVC	99.4	15
		Kaolin	99.8	40
		Lycopodium	99.1	65

Table 4.5 Overall removal efficiency versus rate of flow and inlet concentration for PVC, kaolin and lycopodium powder.

Inlet Concentration (mg/l)	Rate of Flow (l/min)	Type of Suspension	Removal Efficiency (%)	Run Number
150	1.0	PVC	99.4	16
		Kaolin	99.7	41
		Lycopodium	99.3	66
	0.8	PVC	99.4	17
		Kaolin	99.5	42
		Lycopodium	99.5	67
	0.6	PVC	99.7	18
		Kaolin	99.7	43
		Lycopodium	99.7	68
	0.4	PVC	99.6	19
		Kaolin	99.7	44
		Lycopodium	99.1	69
	0.2	PVC	99.8	20
		Kaolin	99.8	45
		Lycopodium	99.1	70

Table 4.6 Overall removal efficiency versus rate of flow and inlet concentration for PVC, kaolin and lycopodium powder.

Inlet Concentration (mg/l)	Rate of Flow (l/min)	Type of Suspension	Removal Efficiency (%)	Run Number
100	1.0	PVC	99.7	21
		Kaolin	98.8	46
		Lycopodium	98.5	71
	0.8	PVC	99.3	22
		Kaolin	99.3	47
		Lycopodium	99.2	72
	0.6	PVC	99.6	23
		Kaolin	99.7	48
		Lycopodium	99.5	73
	0.4	PVC	99.4	24
		Kaolin	99.7	49
		Lycopodium	99.5	74
	0.2	PVC	99.5	25
		Kaolin	99.4	50
		Lycopodium	99.4	75

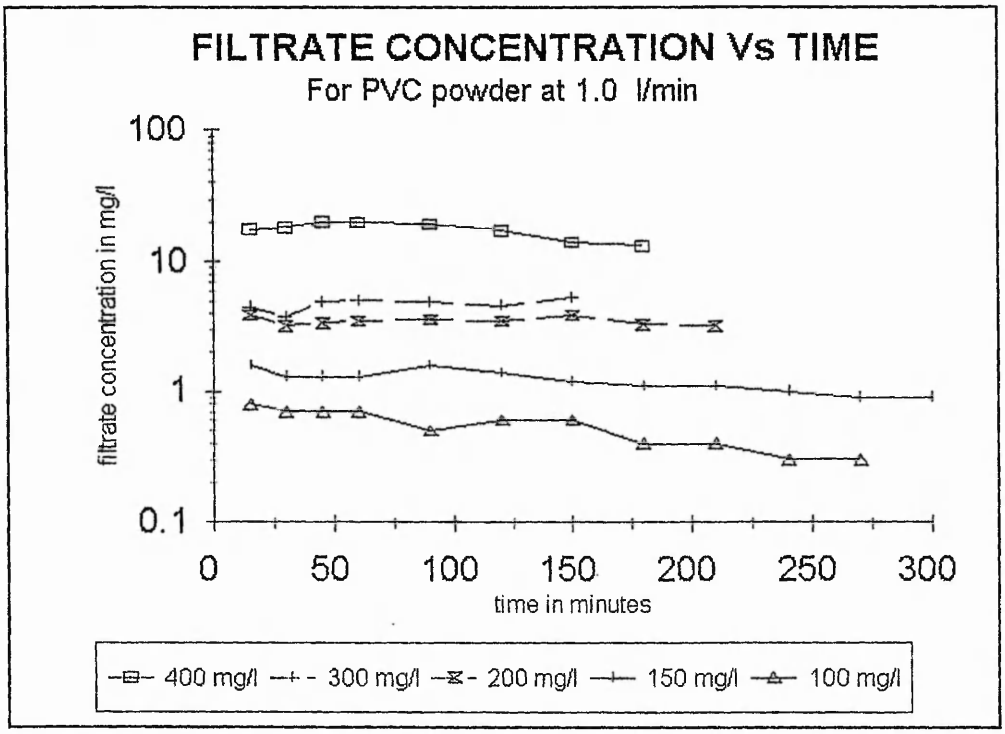


Fig 4.1 Suspension concentration in the filtrate with time and various inlet concentrations for PVC powder at 1.0 l/min.

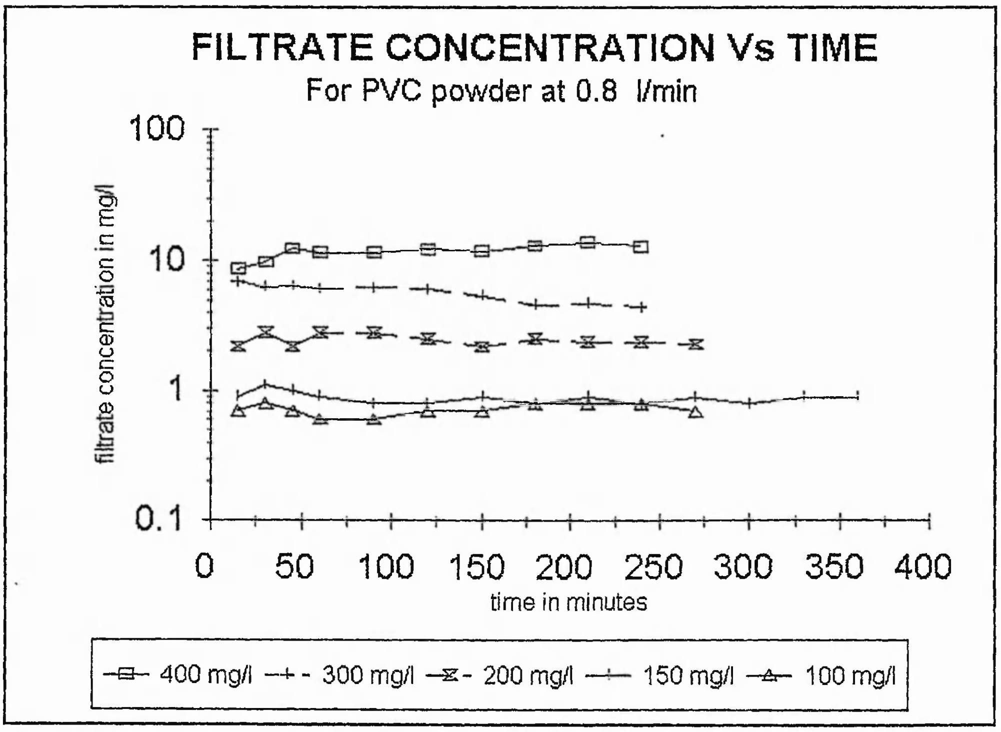


Fig 4.2 Suspension concentration in the filtrate with time and various inlet concentrations for PVC powder at 0.8 l/min.

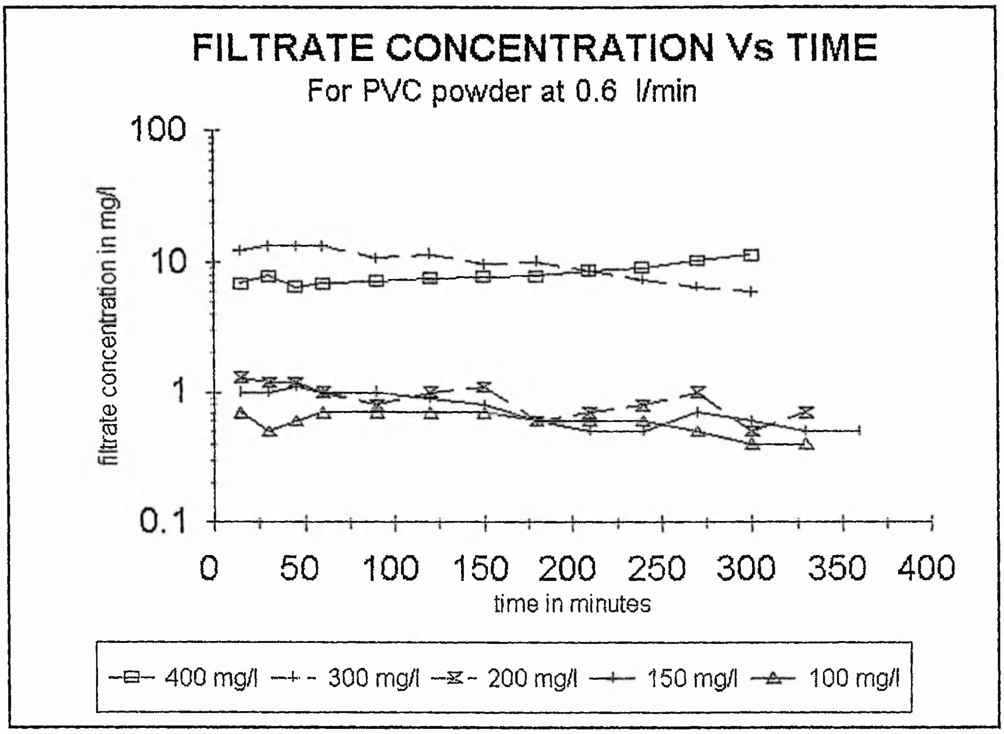


Fig 4.3 Suspension concentration in the filtrate with time and various inlet concentrations for PVC powder at 0.6 l/min.

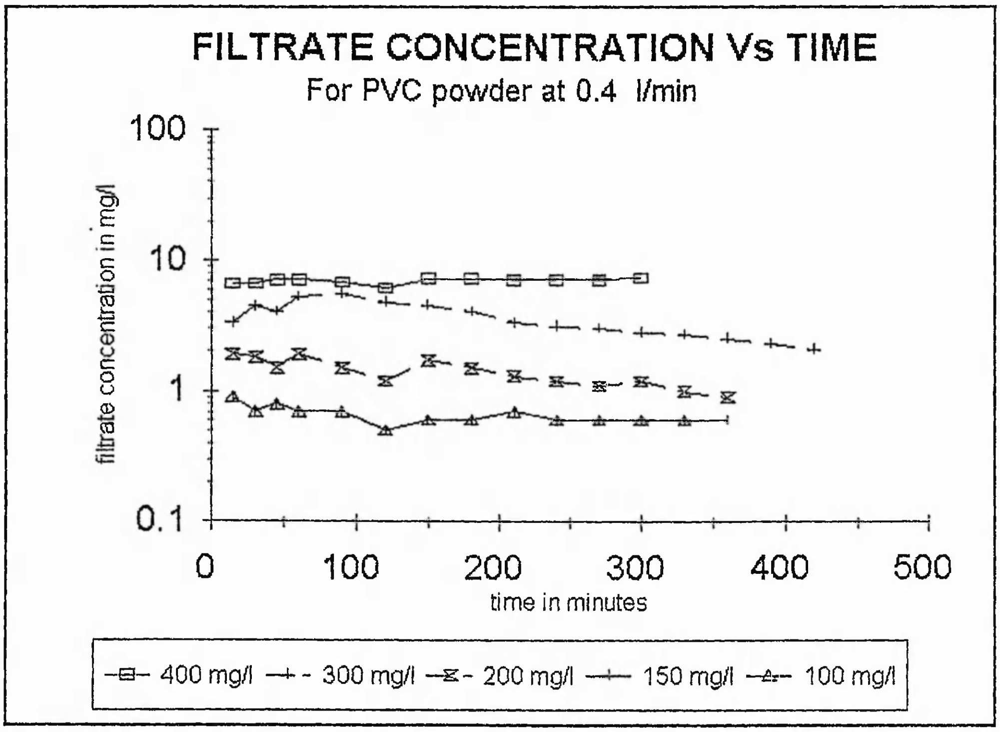


Fig 4.4 Suspension concentration in the filtrate with time and various inlet concentrations for PVC powder at 0.4 l/min.

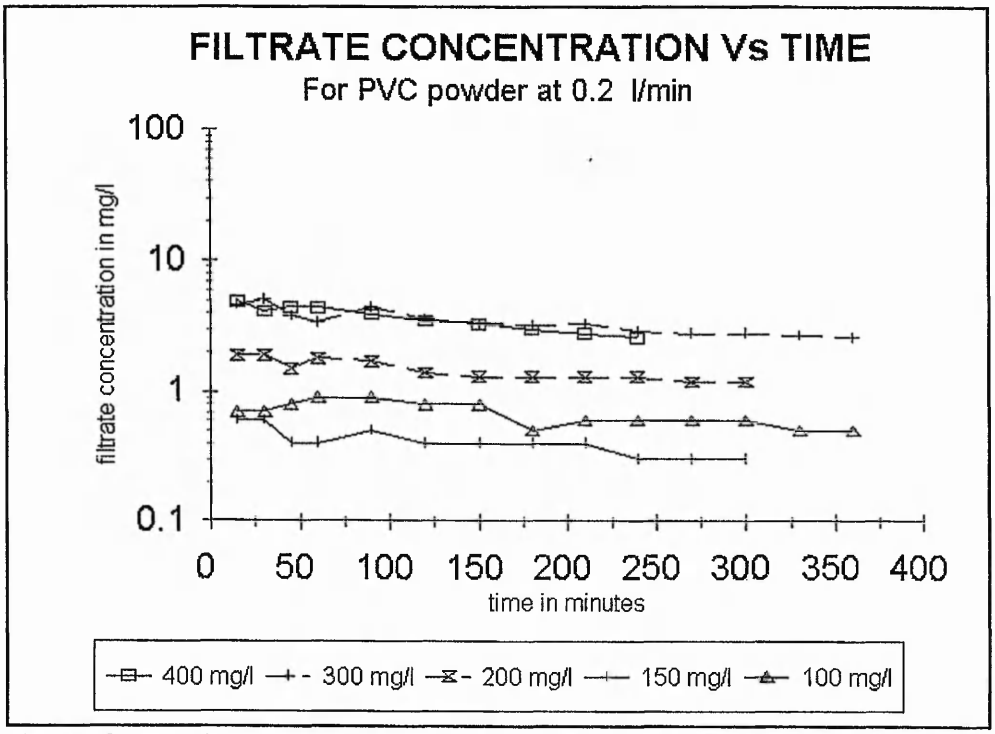


Fig 4.5 Suspension concentration in the filtrate with time and various inlet concentrations for PVC powder at 0.2 l/min.

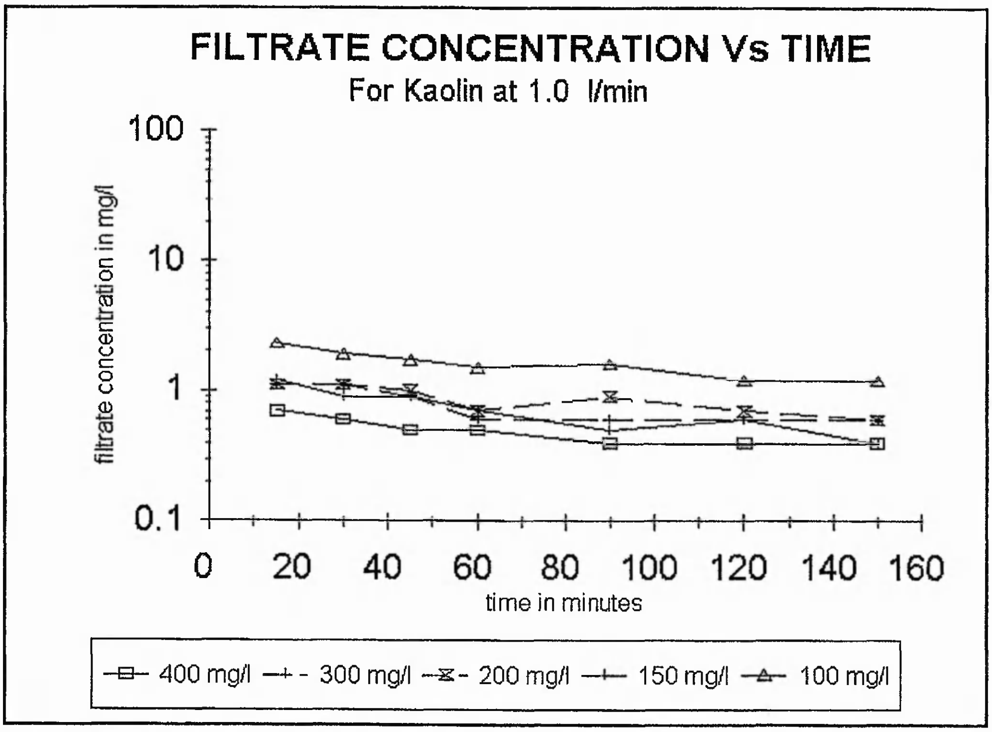


Fig 4.6 Suspension concentration in the filtrate with time and various inlet concentrations for Kaolin at 1.0 l/min.

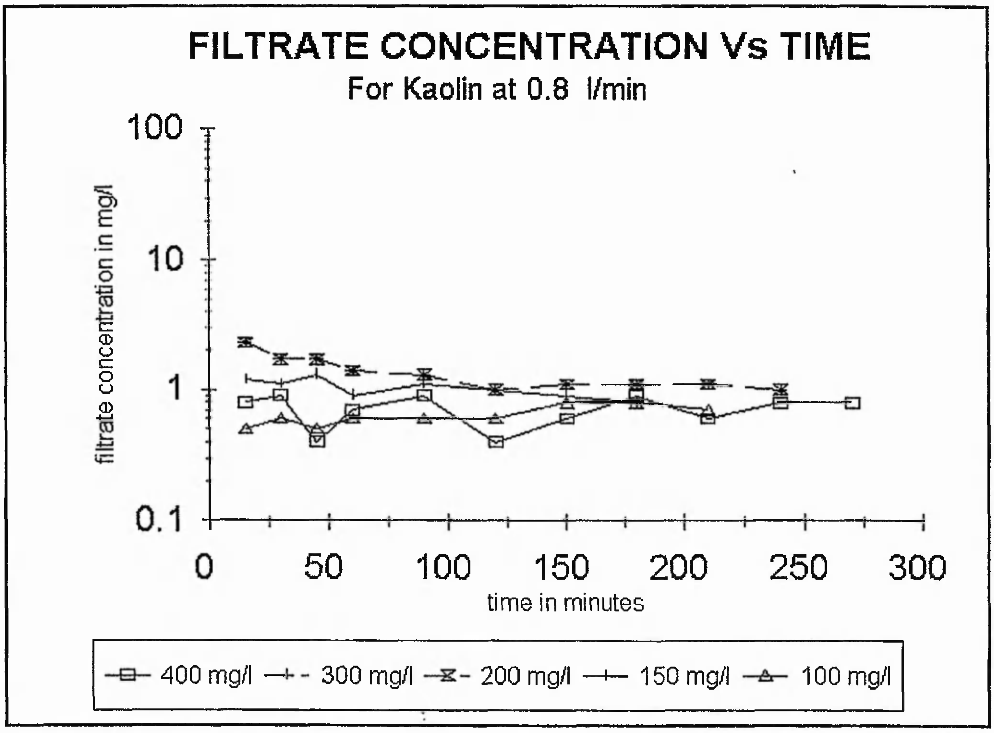


Fig 4.7 Suspension concentration in the filtrate with time and various inlet concentrations for Kaolin at 0.8 l/min.

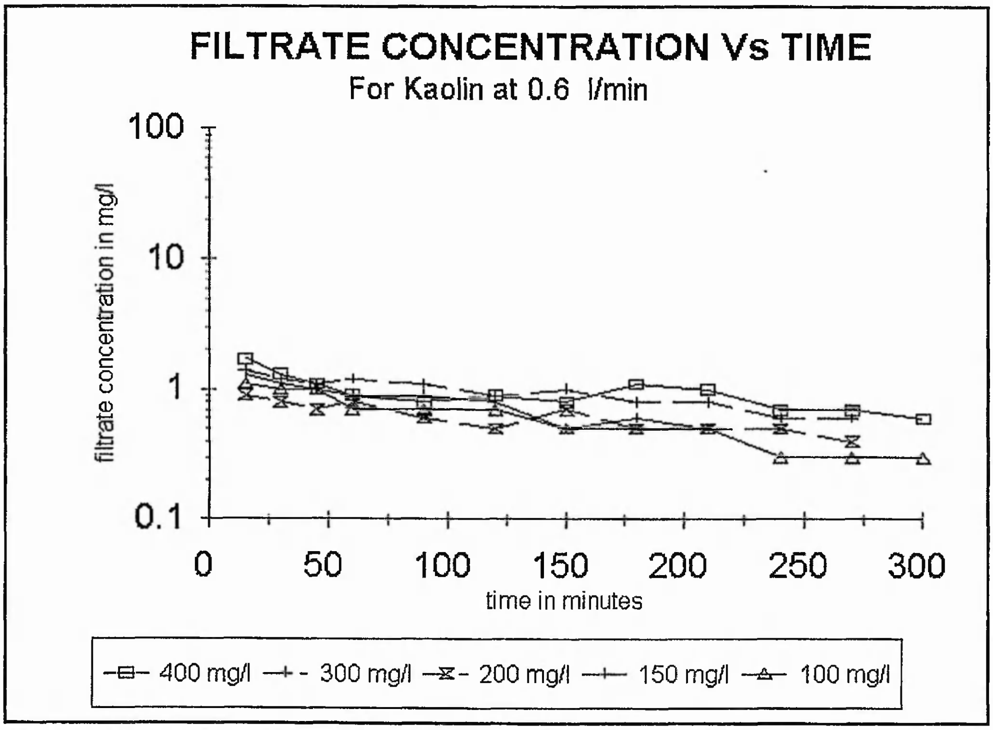


Fig 4.8 Suspension concentration in the filtrate with time and various inlet concentrations for Kaolin at 0.6 l/min.

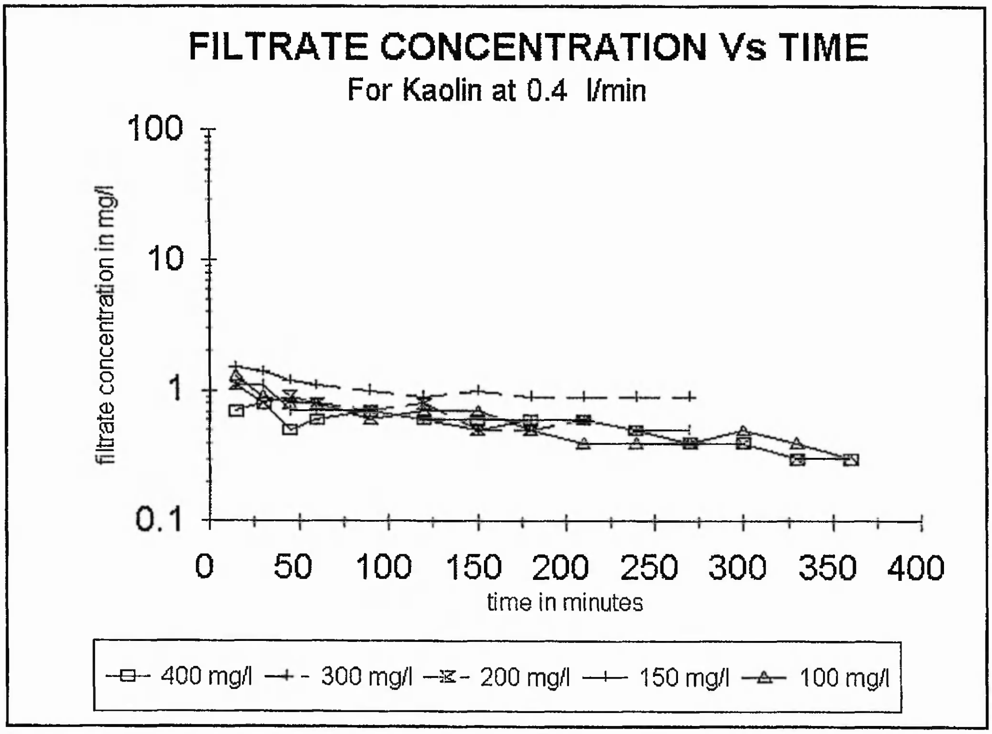


Fig 4.9 Suspension concentration in the filtrate with time and various inlet concentrations for Kaolin at 0.4 l/min.

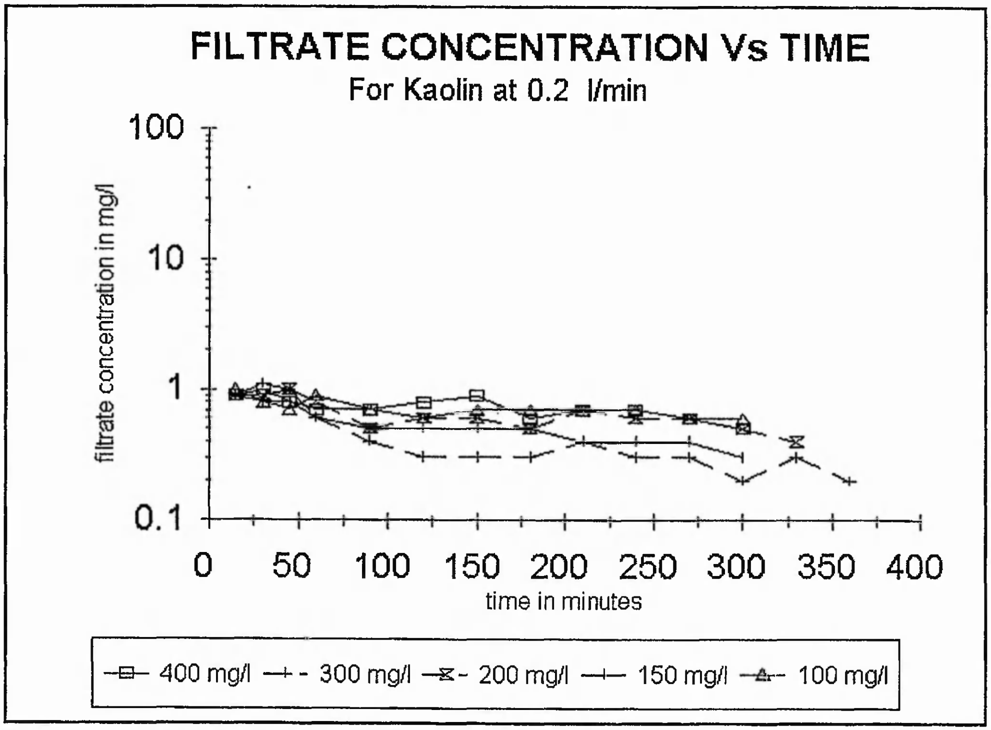


Fig 4.10 Suspension concentration in the filtrate with time and various inlet concentrations for Kaolin at 0.2 l/min.

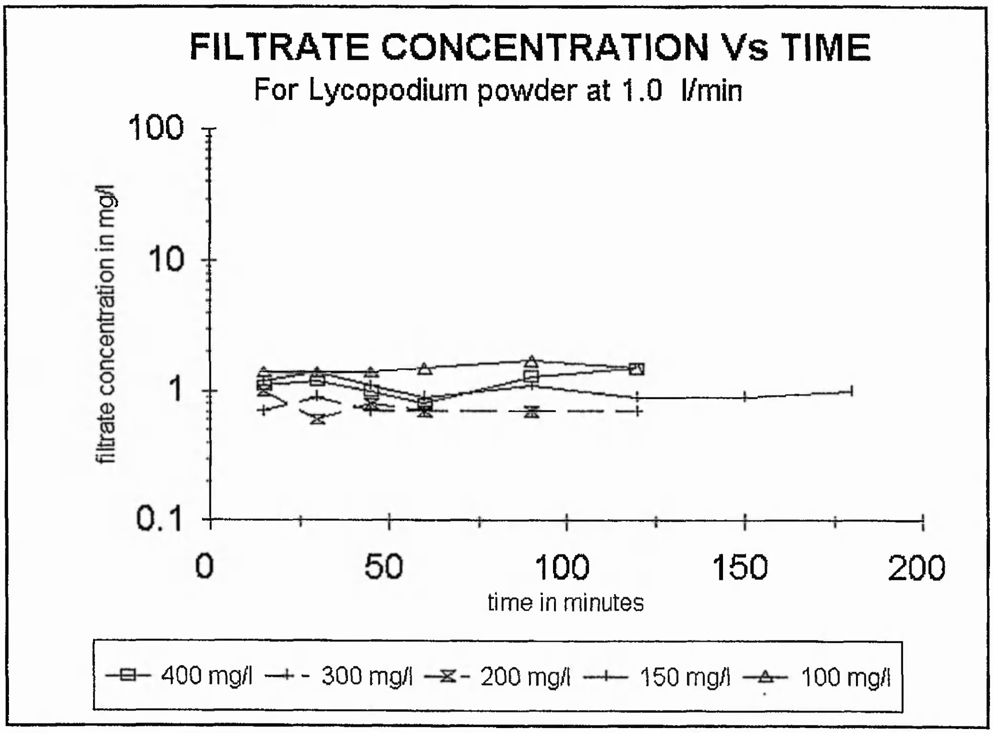


Fig 4.11 Suspension concentration in the filtrate with time and various inlet concentrations for Lycopodium powder at 1.0 l/min.

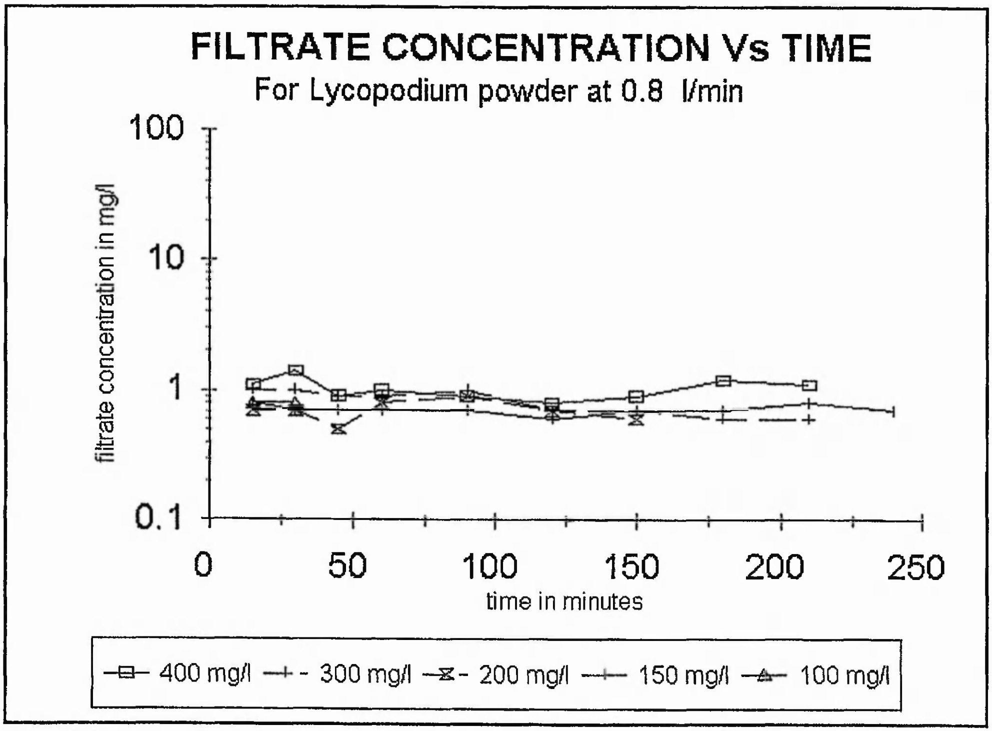


Fig 4.12 Suspension concentration in the filtrate with time and various inlet concentrations for Lycopodium powder at 0.8 l/min.

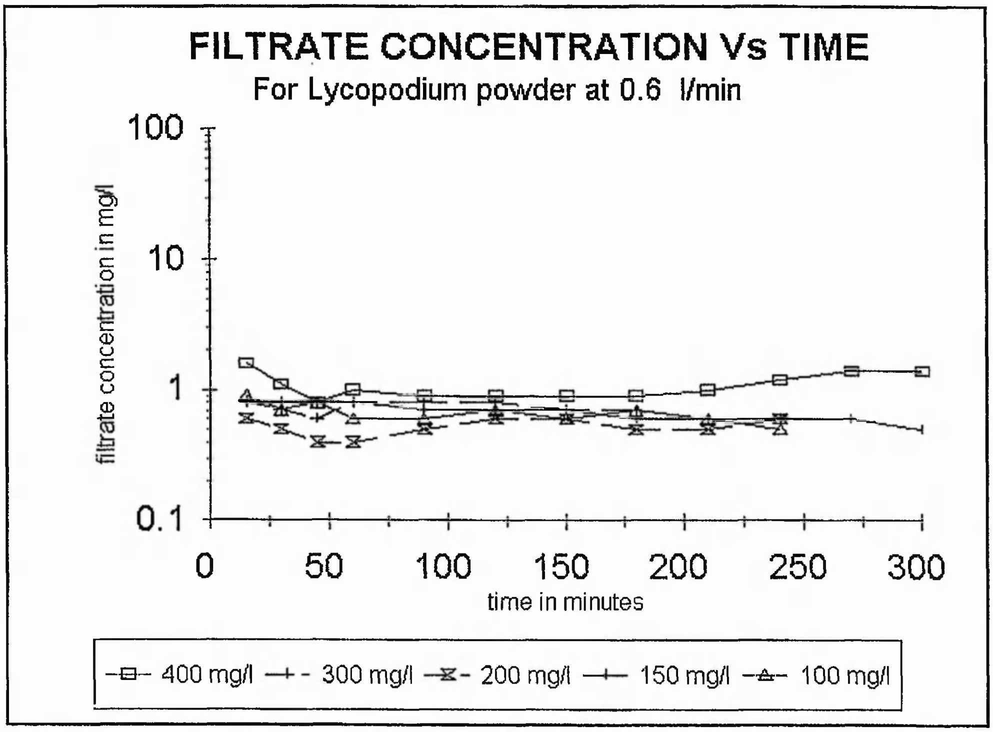


Fig 4.13 Suspension concentration in the filtrate with time and various inlet concentrations for Lycopodium powder at 0.6 l/min.

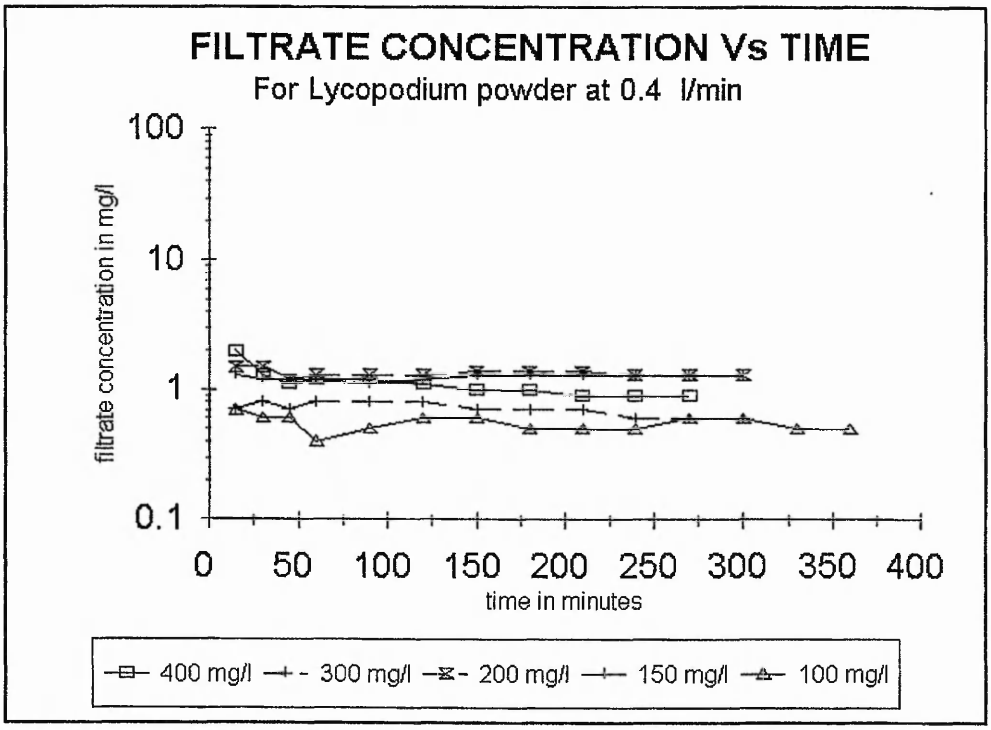


Fig 4.14 Suspension concentration in the filtrate with time and various inlet concentrations for Lycopodium powder at 0.4 l/min.

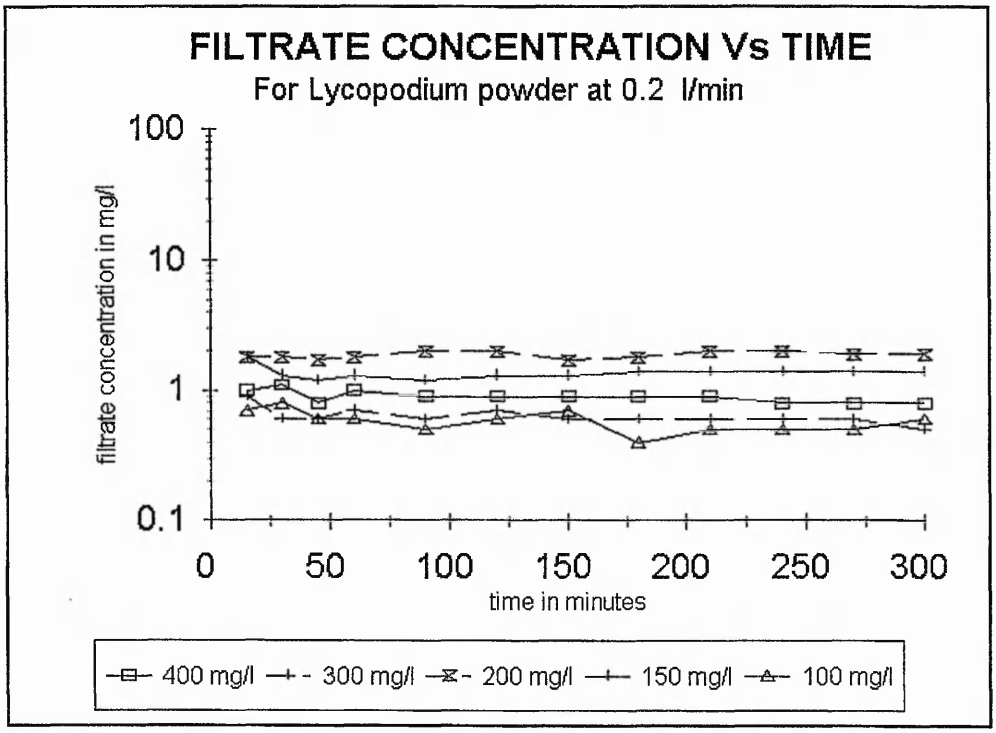


Fig 4.15 Suspension concentration in the filtrate with time and various inlet concentrations for Lycopodium powder at 0.2 l/min.

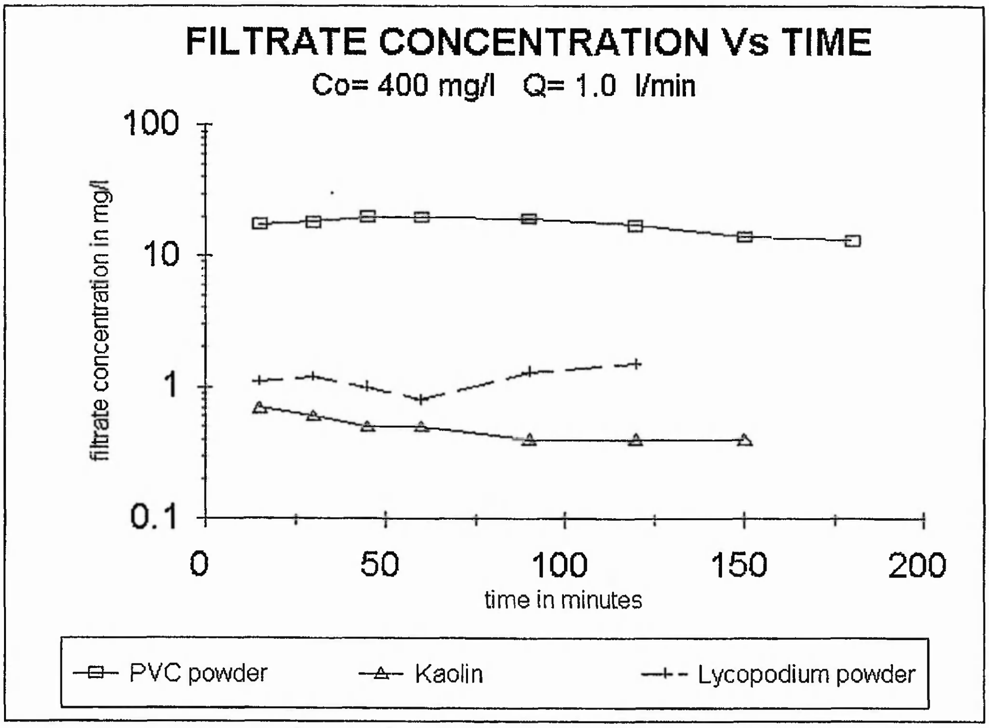


Fig 4.16 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 400 mg/l and Q= 1.0 l/min.

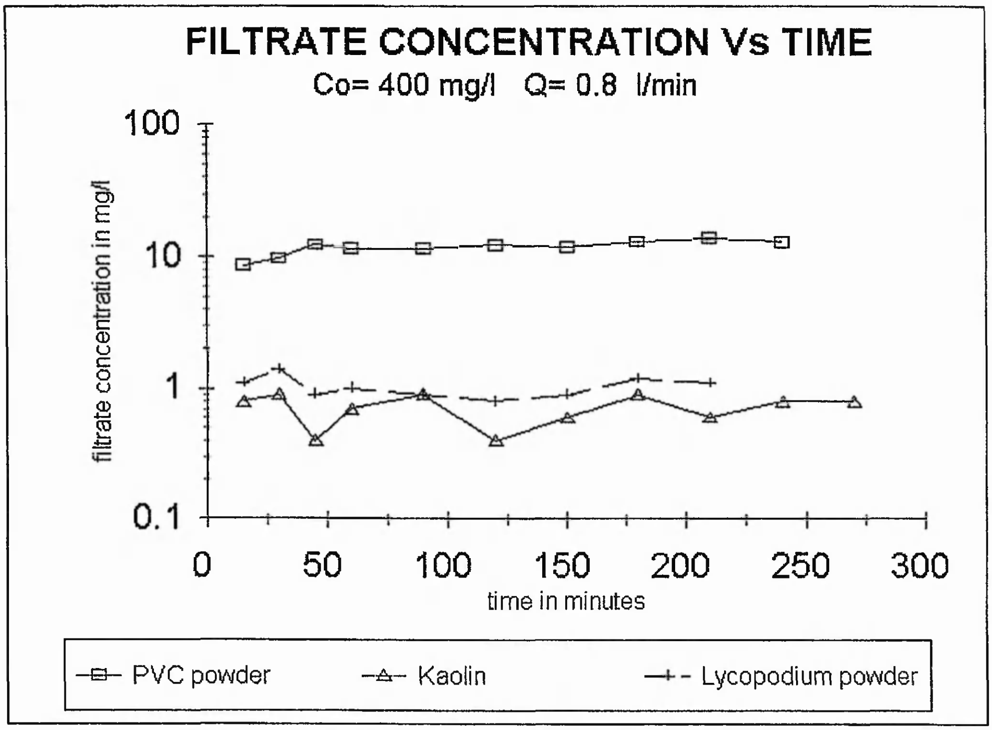


Fig 4.17 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 400 mg/l and Q= 0.8 l/min.

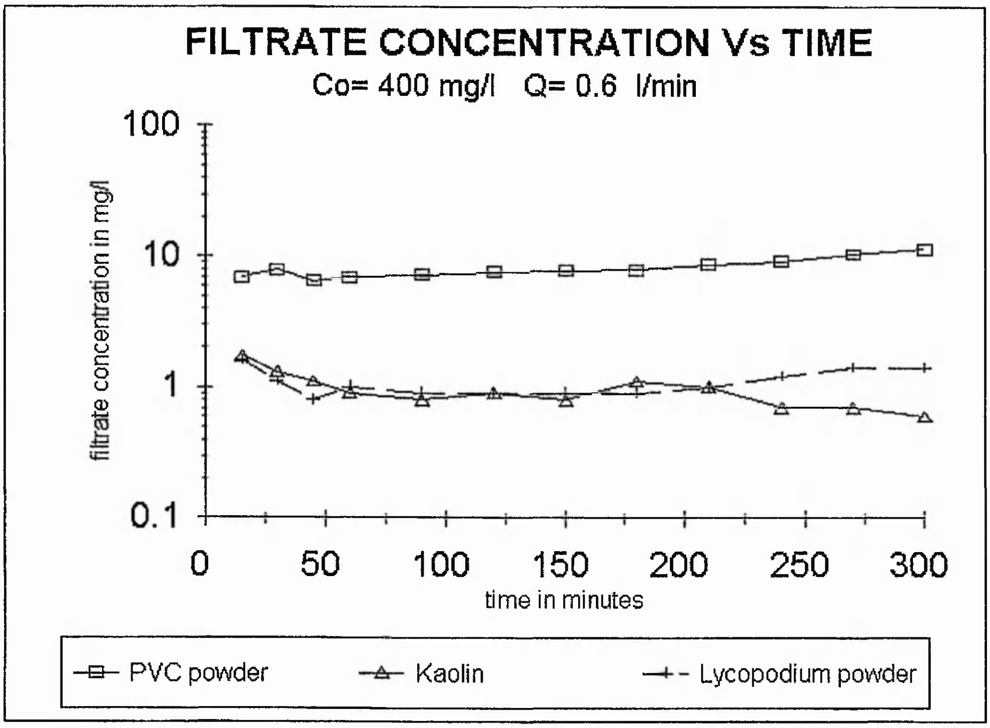


Fig 4.18 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 400 mg/l and Q= 0.6 l/min.

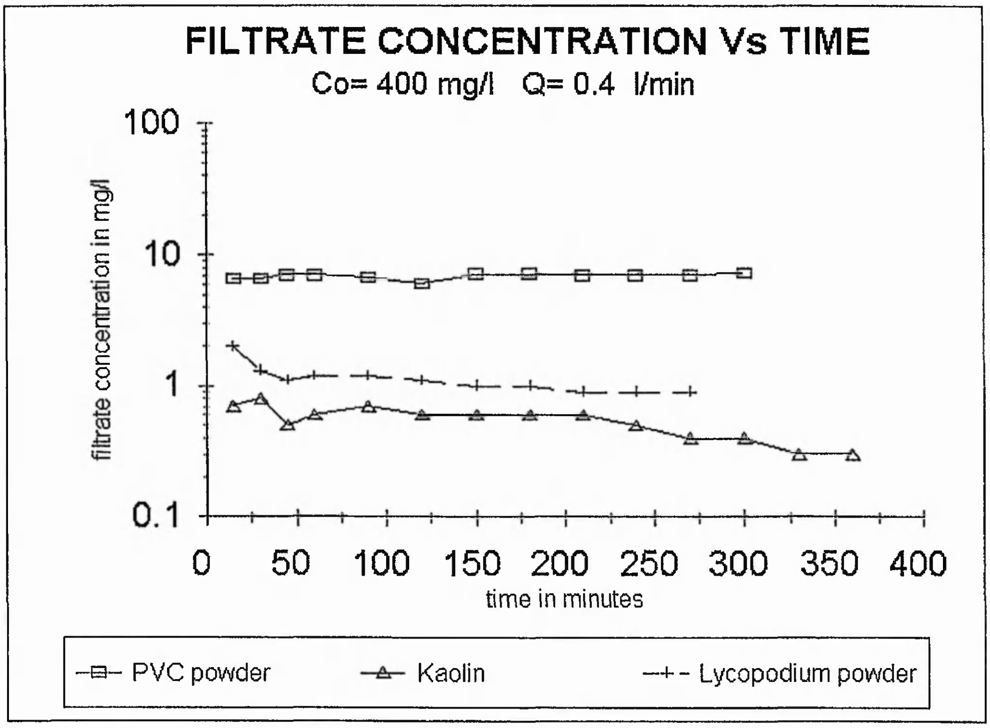


Fig 4.19 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 400 mg/l and Q= 0.4 l/min.

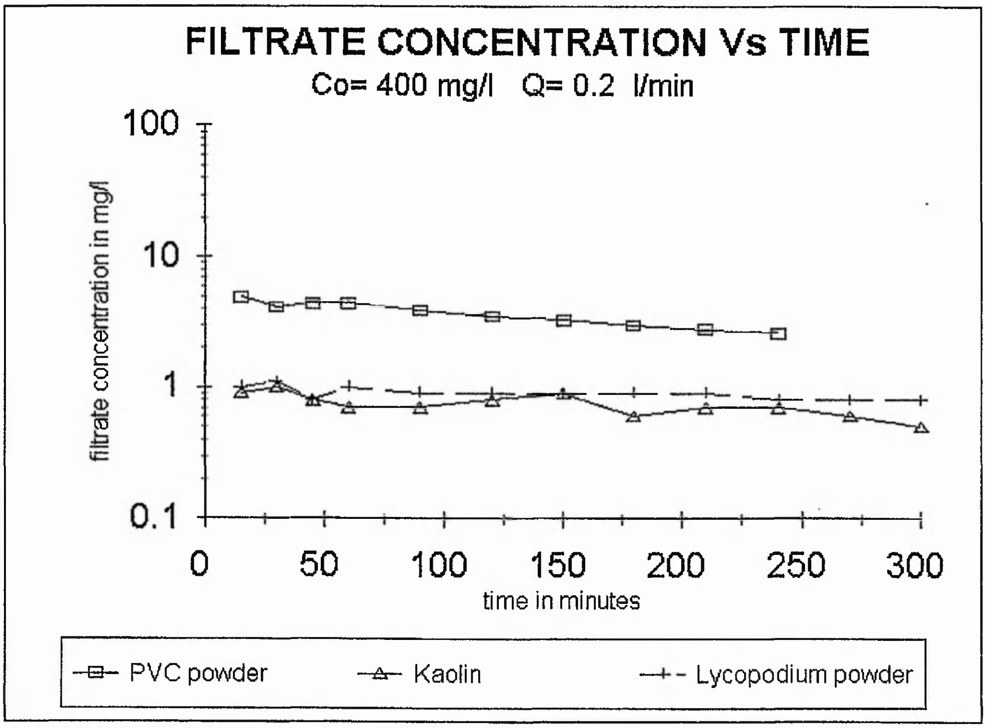


Fig 4.20 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 400 mg/l and Q= 0.2 l/min.

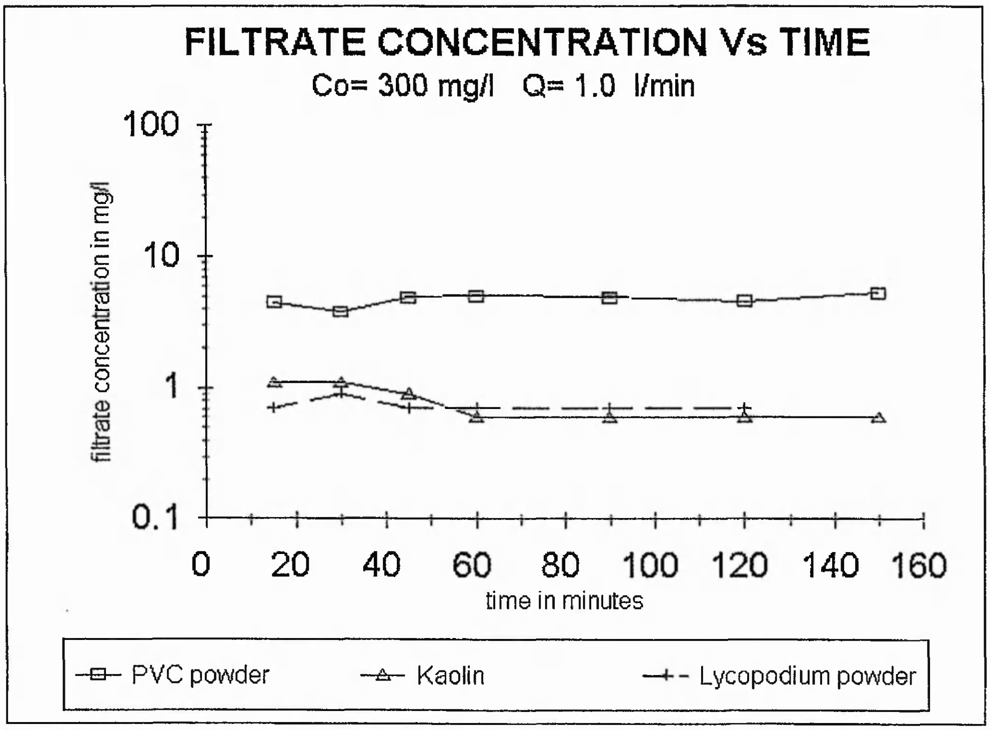


Fig 4.21 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 300 mg/l and Q= 1.0 l/min.

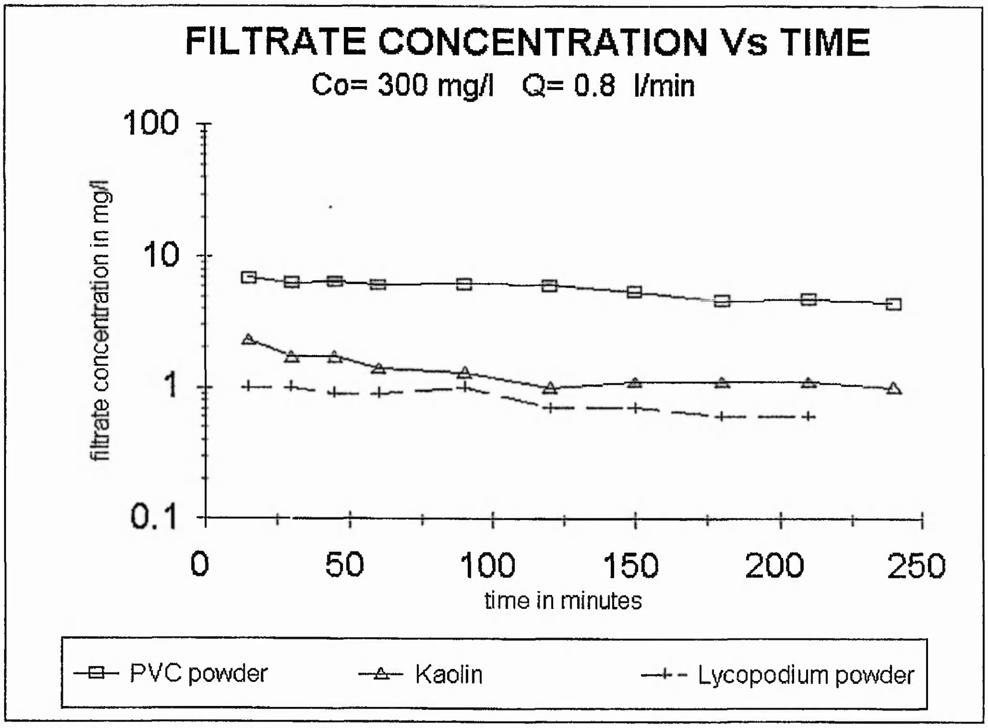


Fig 4.22 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 300 mg/l and Q= 0.8 l/min.

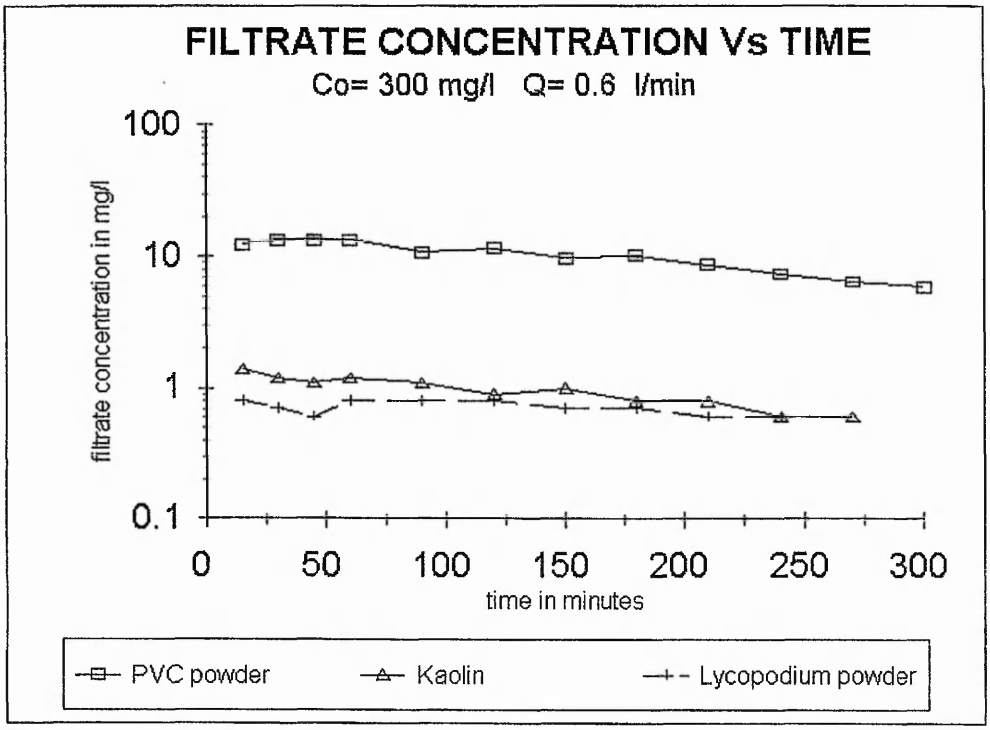


Fig 4.23 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 300 mg/l and Q= 0.6 l/min.

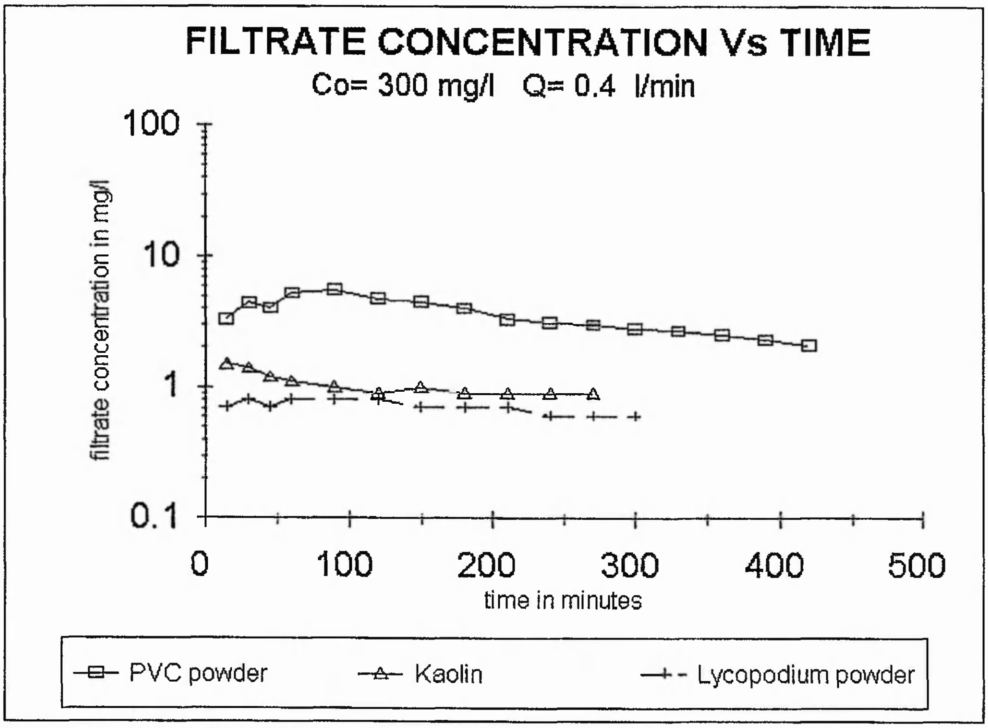


Fig 4.24 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 300 mg/l and Q= 0.4 l/min.

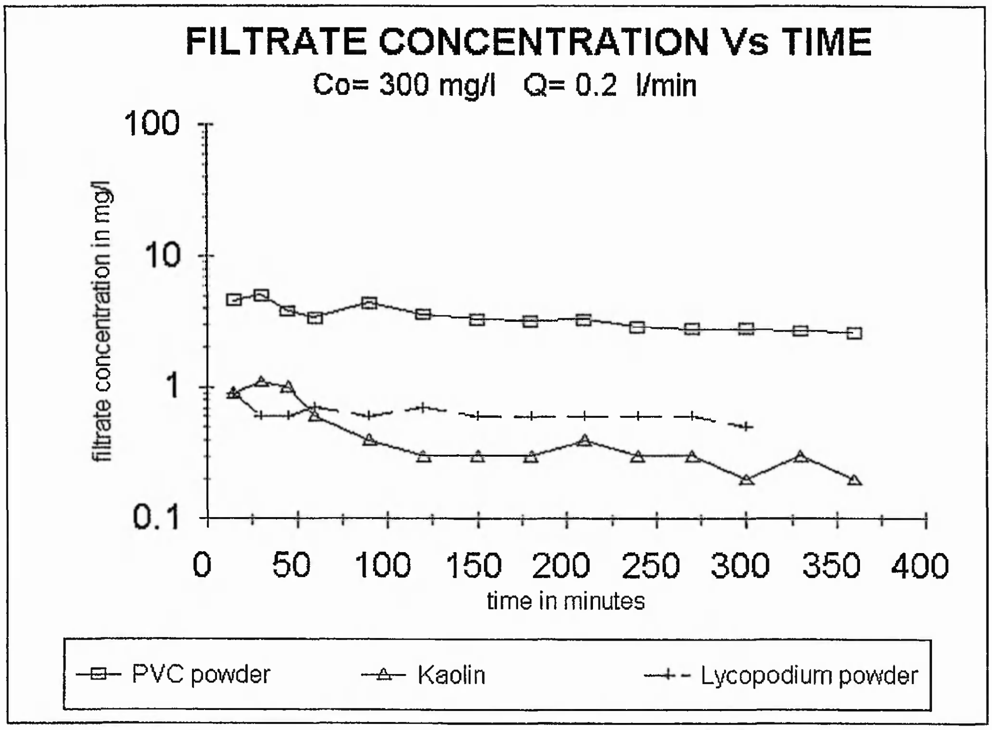


Fig 4.25 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 300 mg/l and Q= 0.2 l/min.

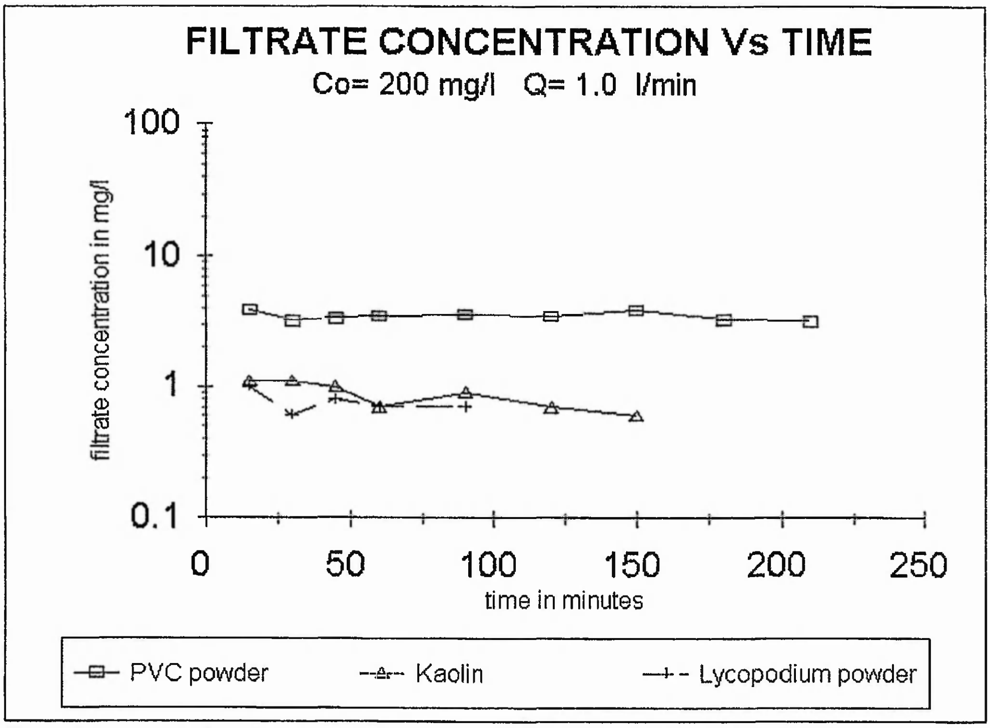


Fig 4.26 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 200 mg/l and Q= 1.0 l/min.

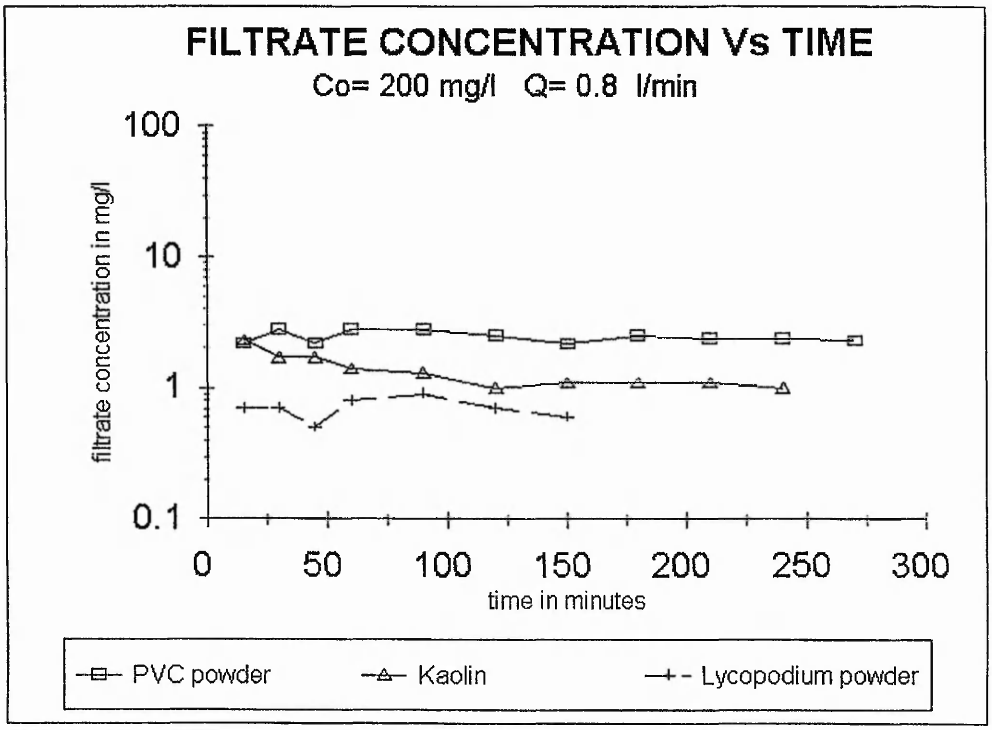


Fig 4.27 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 200 mg/l and Q= 0.8 l/min.

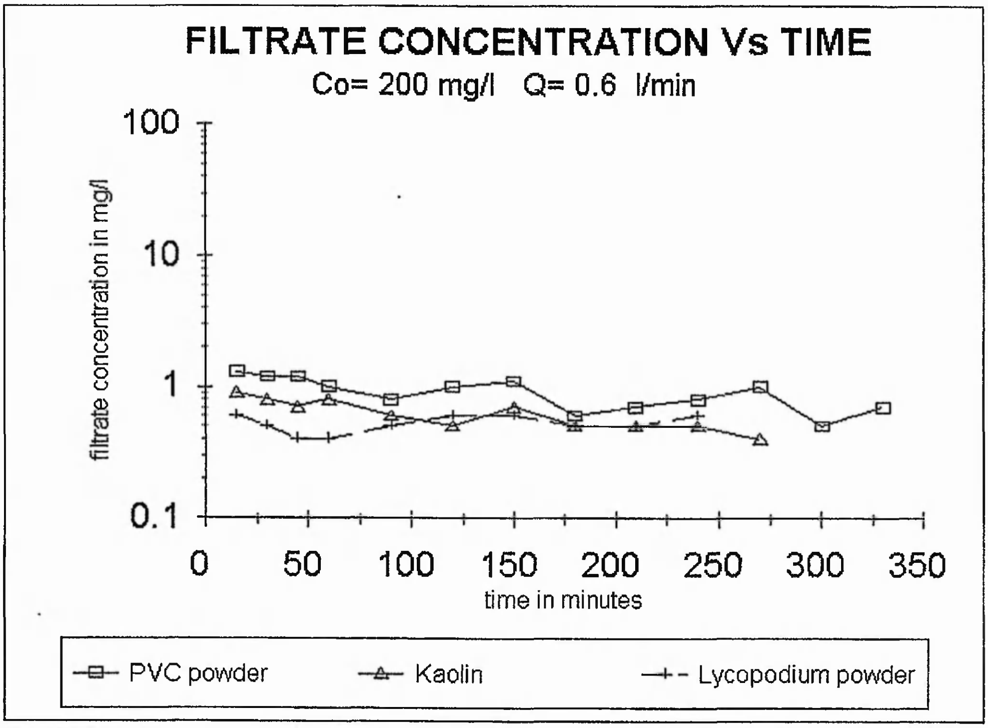


Fig 4.28 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 200 mg/l and Q= 0.6 l/min.

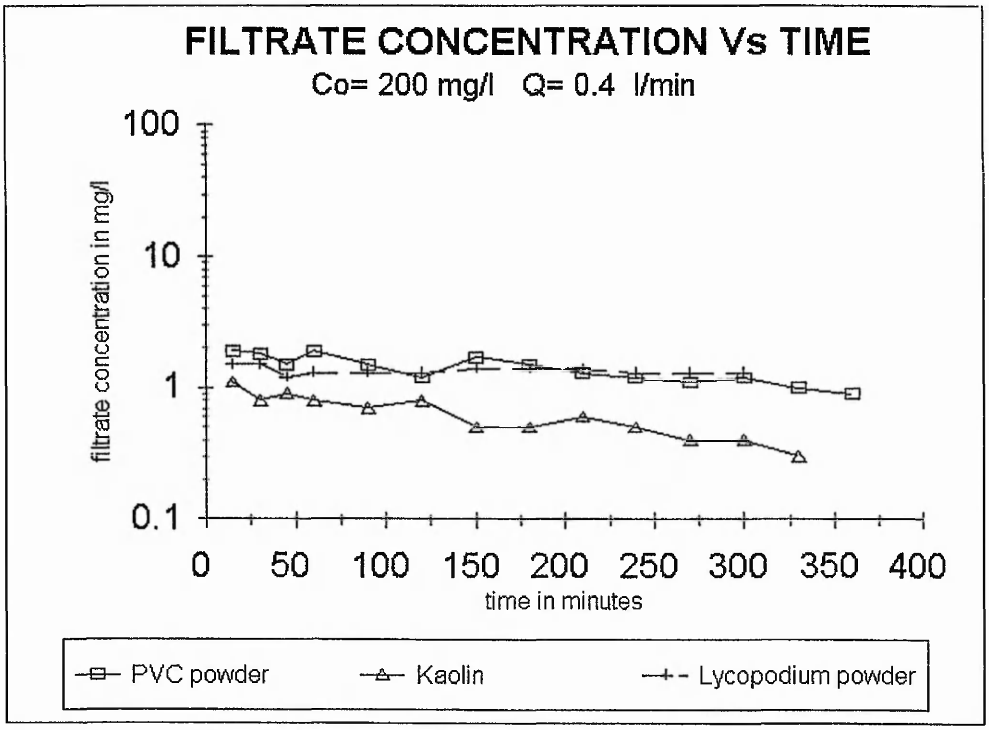


Fig 4.29 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 200 mg/l and Q= 0.4 l/min.

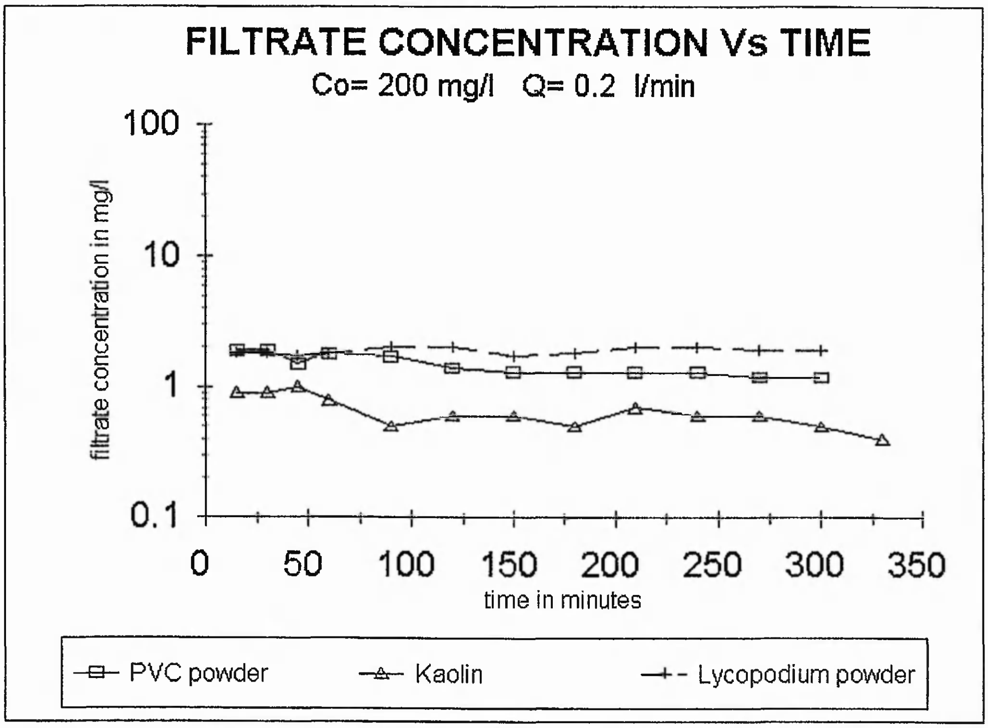


Fig 4.30 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 200 mg/l and Q= 0.2 l/min.

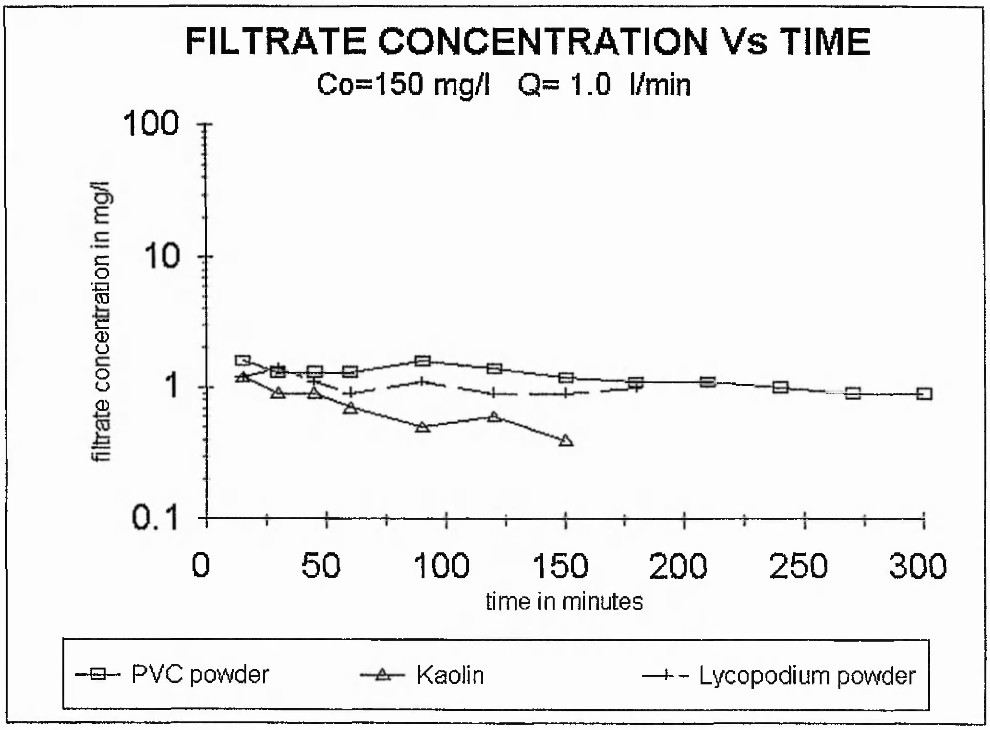


Fig 4.31 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 150 mg/l and Q= 1.0 l/min.

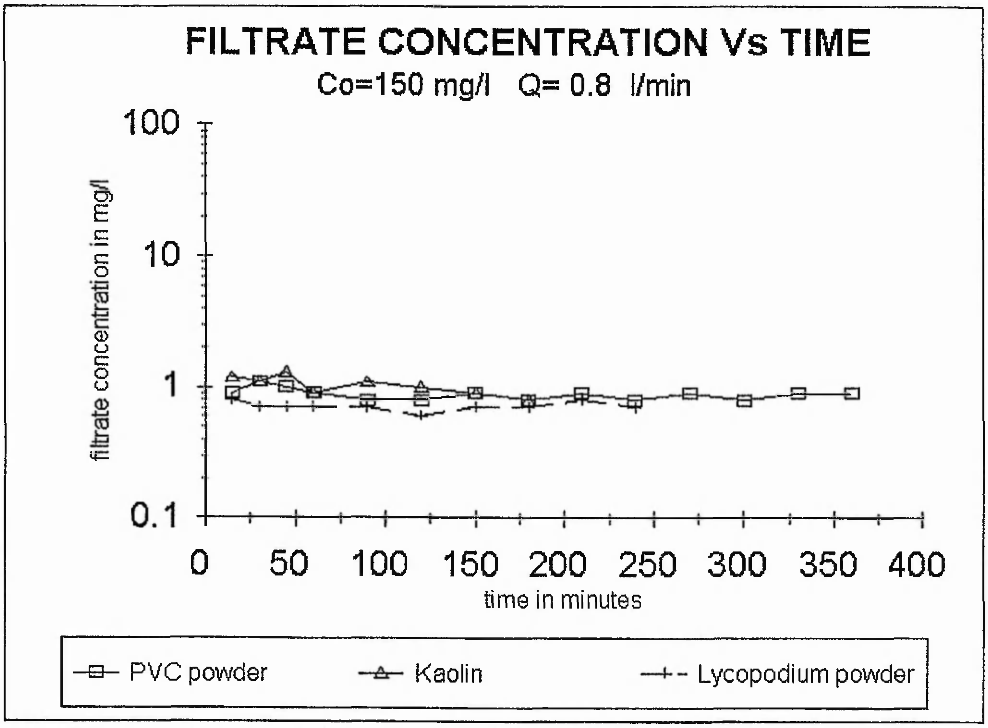


Fig 4.32 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 150 mg/l and Q= 0.8 l/min.

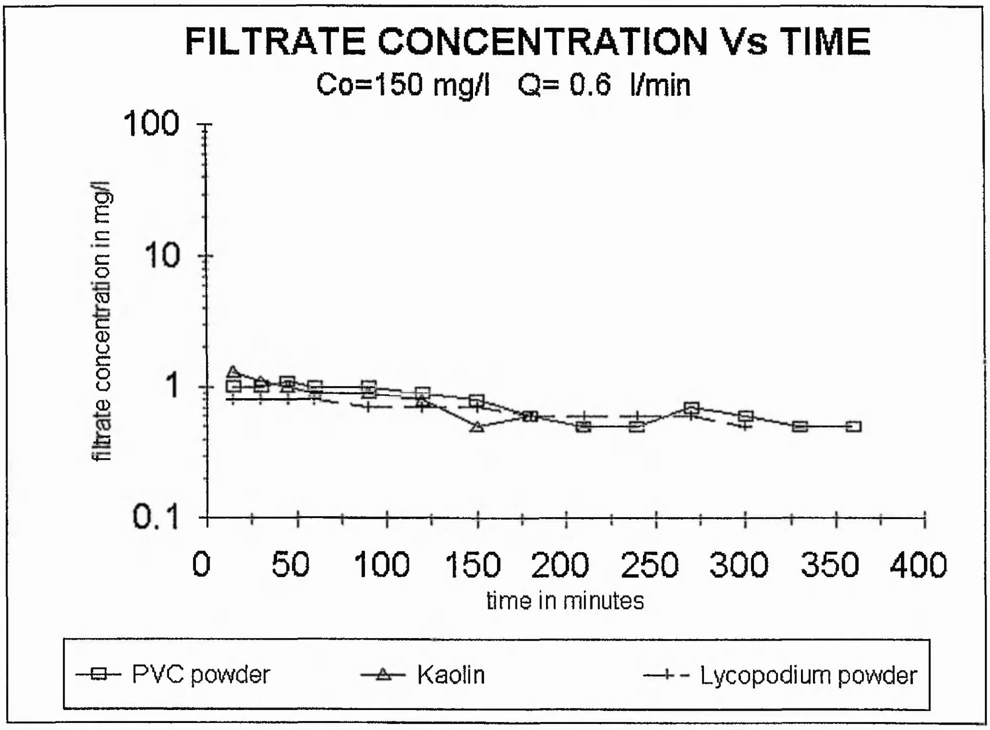


Fig 4.33 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 150 mg/l and Q= 0.6 l/min.

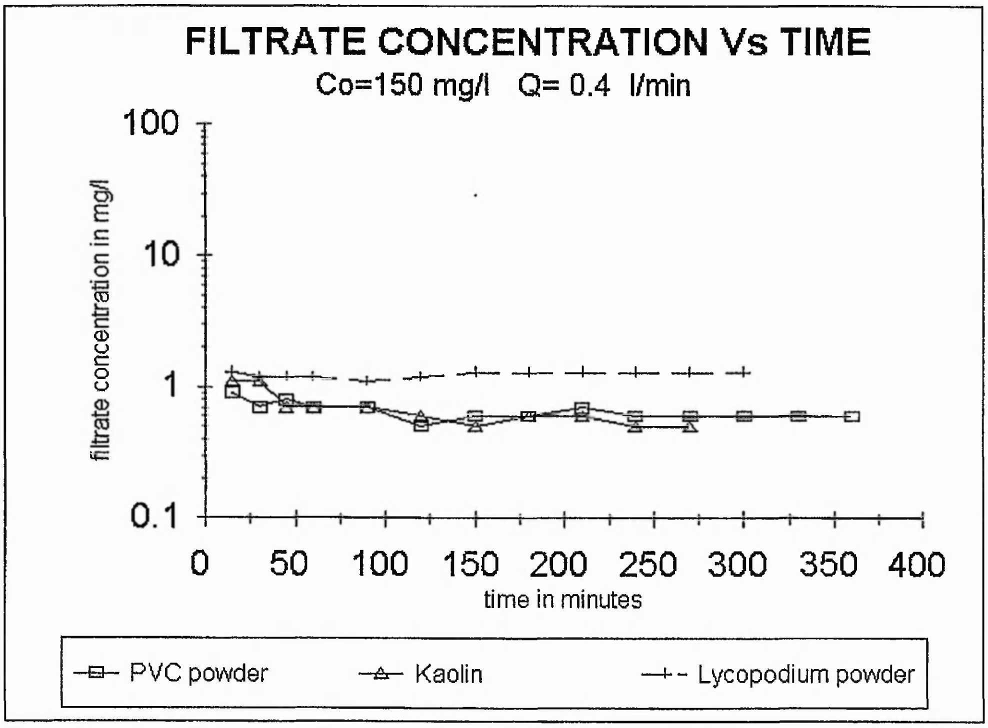


Fig 4.34 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 150 mg/l and Q= 0.4 l/min.

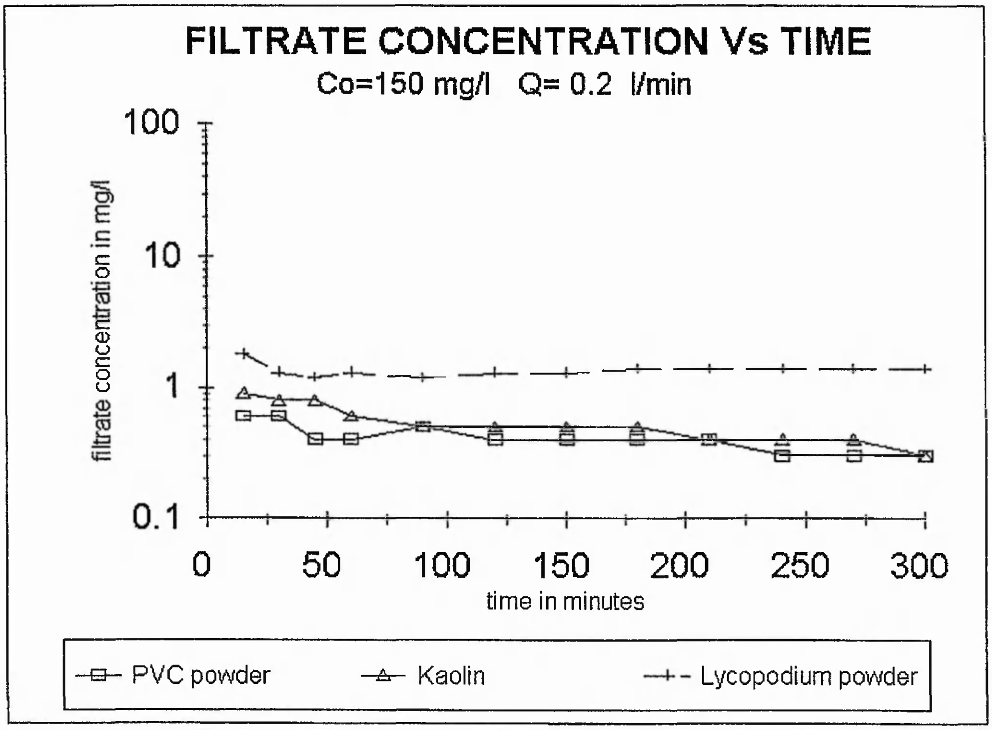


Fig 4.35 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 150 mg/l and Q= 0.2 l/min.

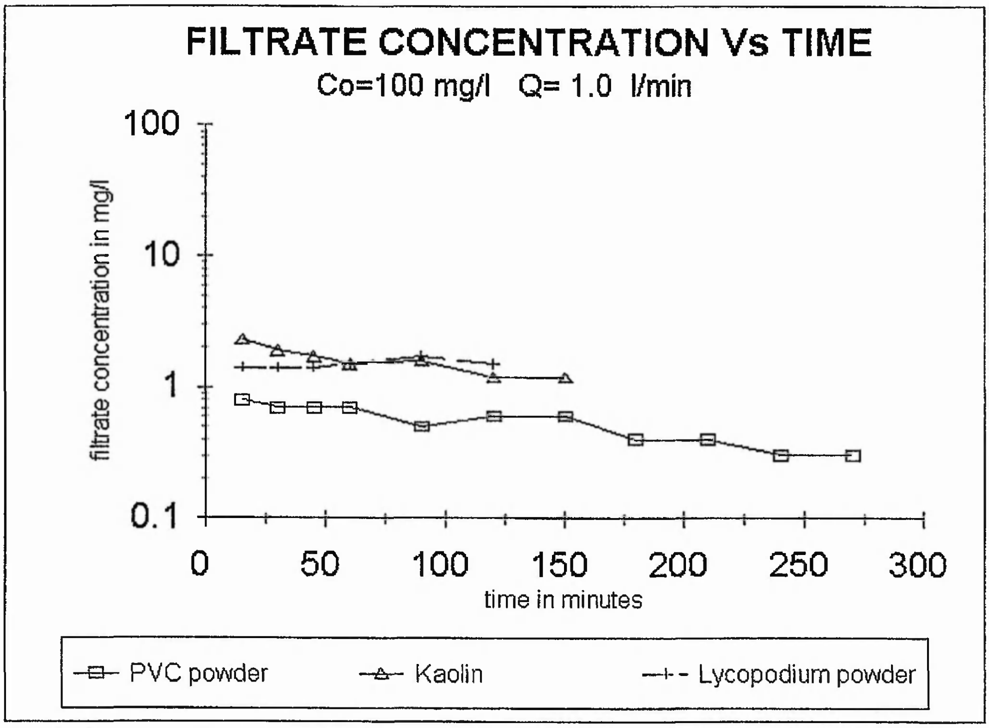


Fig 4.36 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 100 mg/l and Q= 1.0 l/min.

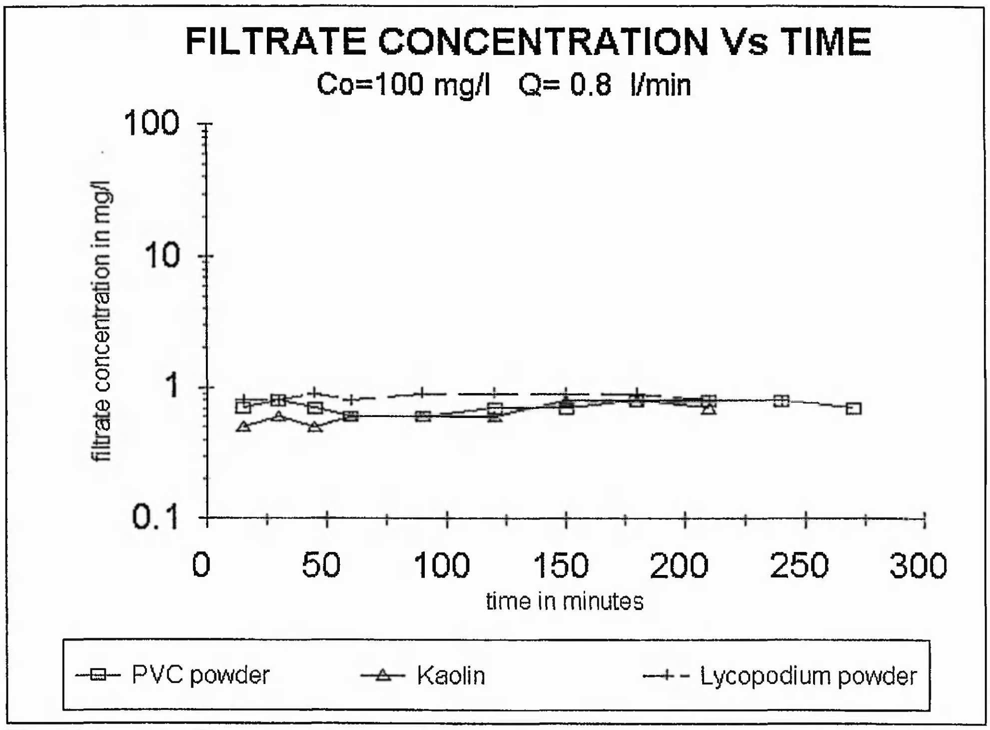


Fig 4.37 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 100 mg/l and Q= 0.8 l/min.

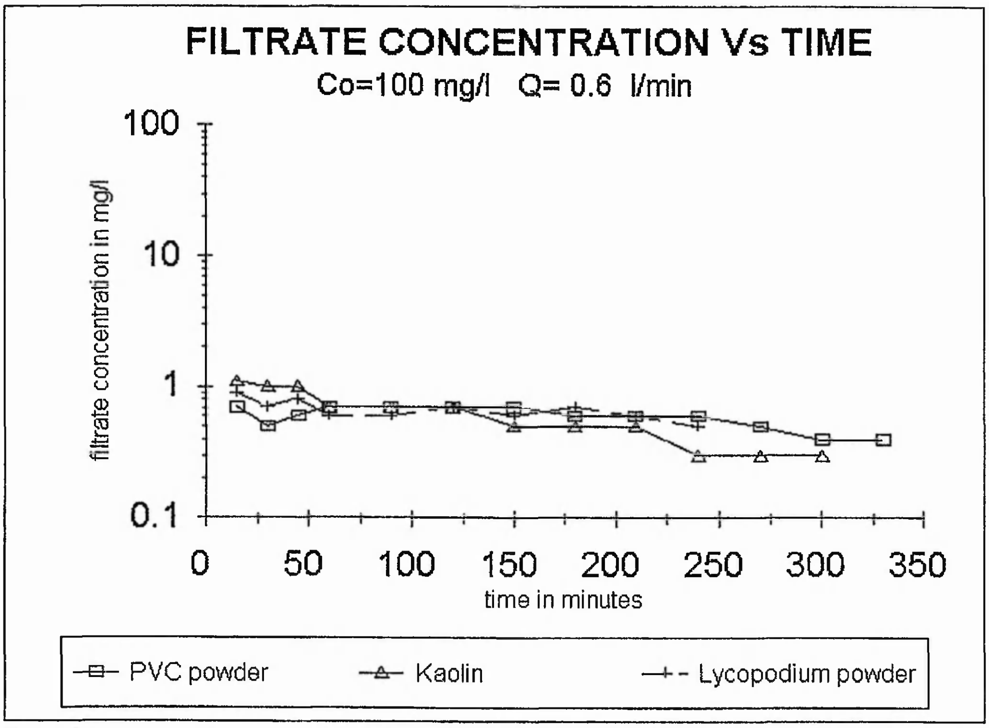


Fig 4.38 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 100 mg/l and Q= 0.6 l/min.

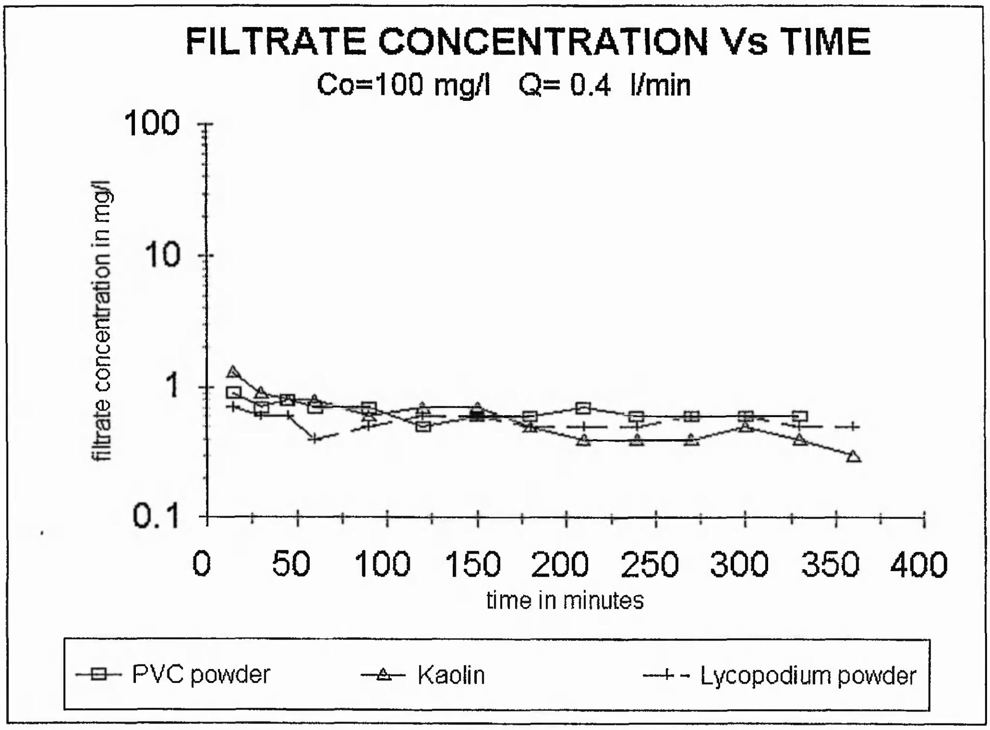


Fig 4.39 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 100 mg/l and Q= 0.4 l/min.

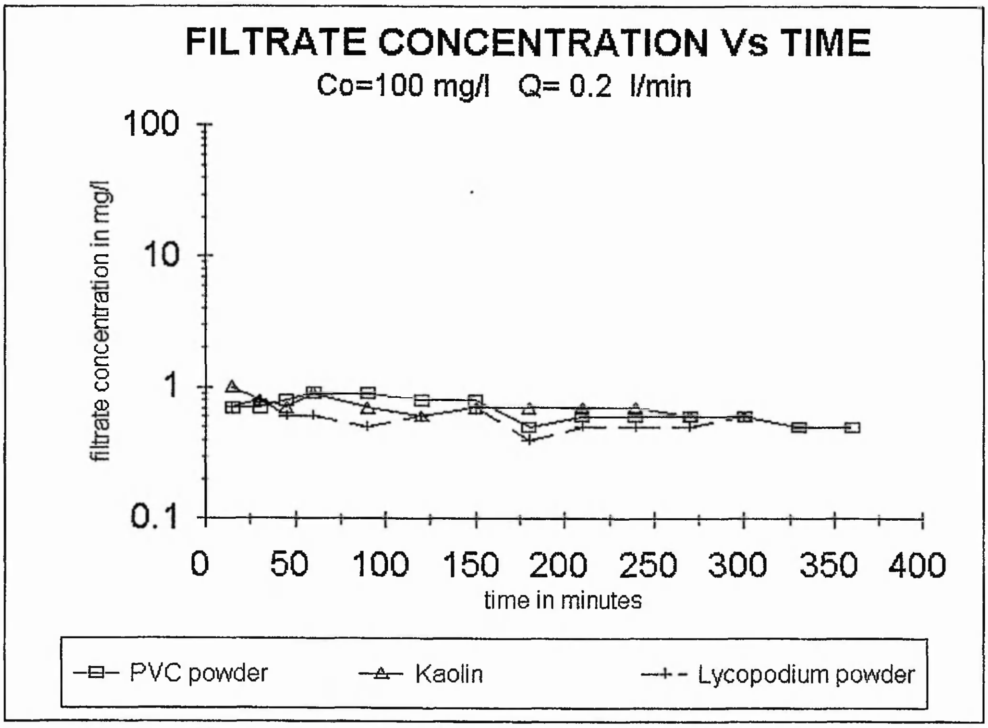


Fig 4.40 Time variation of suspension concentration in the filtrate for various turbidity materials with C_0= 100 mg/l and Q= 0.2 l/min.

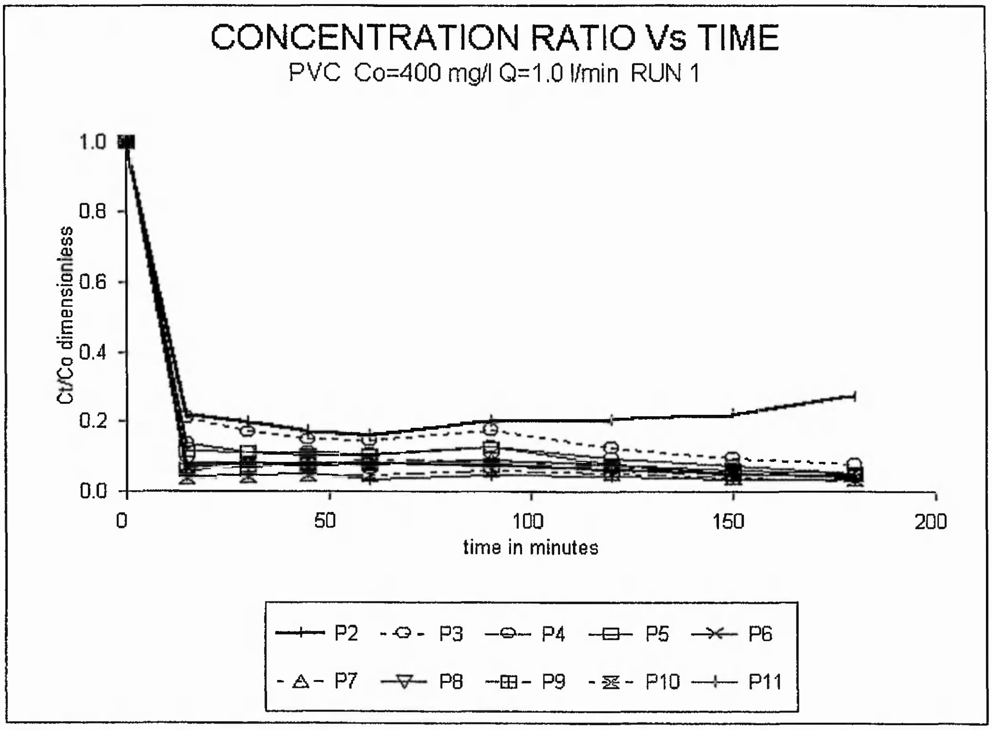


Fig 4.41 Curves of concentration ratio varying with time of a filter run for PVC with C_0= 400 mg/l and Q= 1.0 l/min.

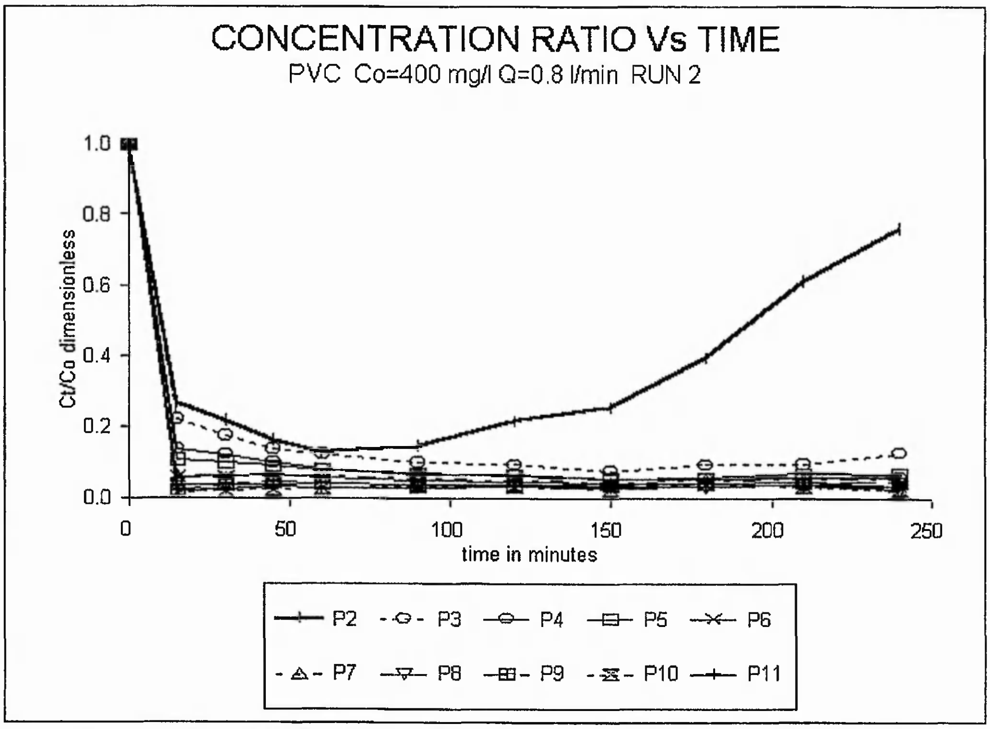


Fig 4.42 Curves of concentration ratio varying with time of a filter run for PVC with C_0= 400 mg/l and Q= 0.8 l/min.

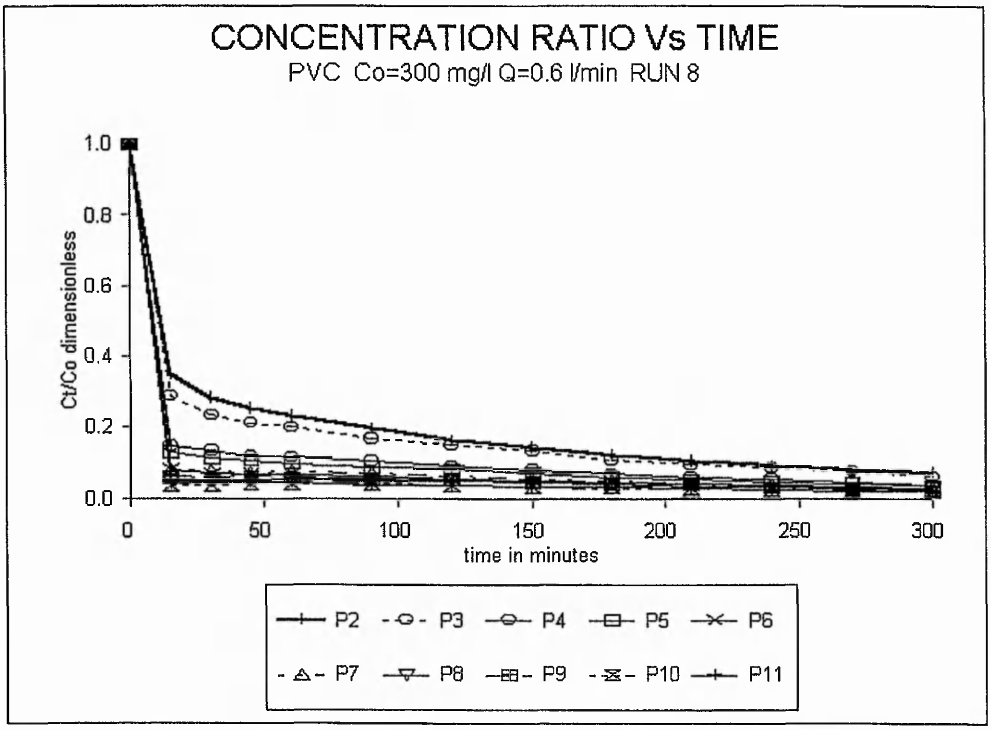


Fig 4.43 Curves of concentration ratio varying with time of a filter run for PVC with C_0= 300 mg/l and Q= 0.6 l/min.

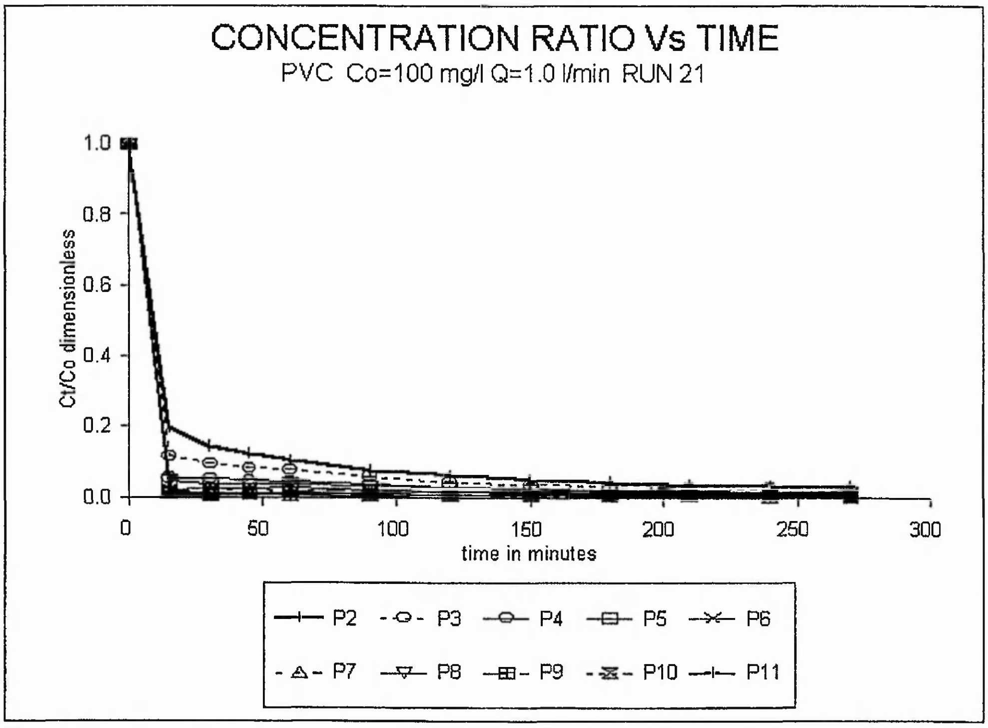


Fig 4.44 Curves of concentration ratio varying with time of a filter run for PVC with C_0= 100 mg/l and Q= 1.0 l/min.

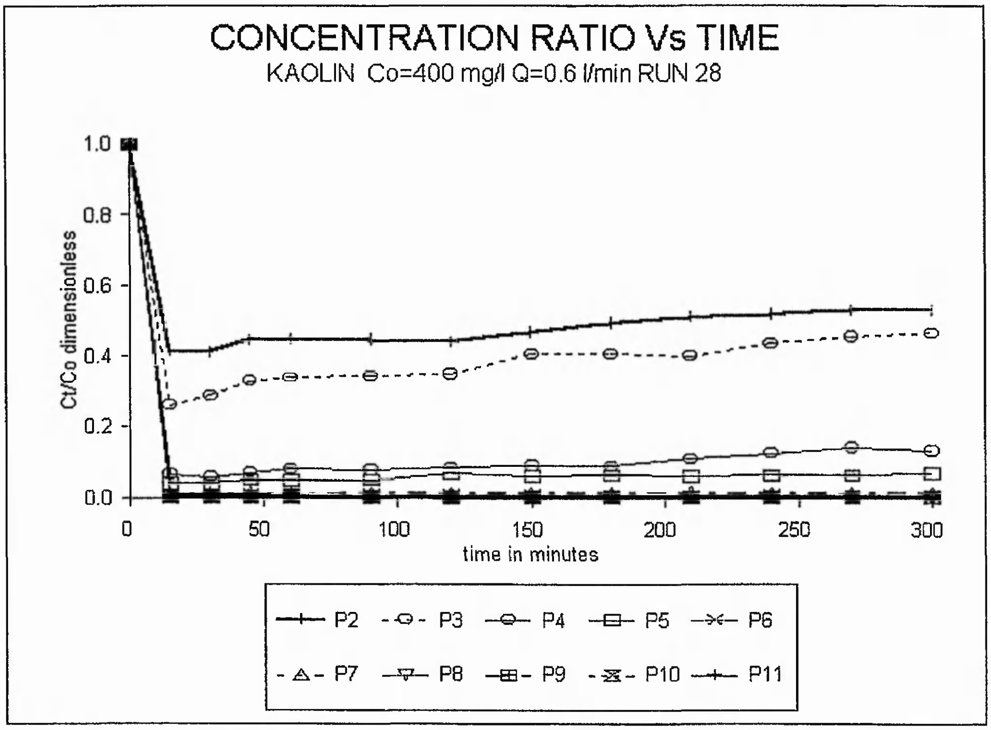


Fig 4.45 Curves of concentration ratio varying with time of a filter run for KAOLIN with C_0= 400 mg/l and Q= 0.6 l/min.

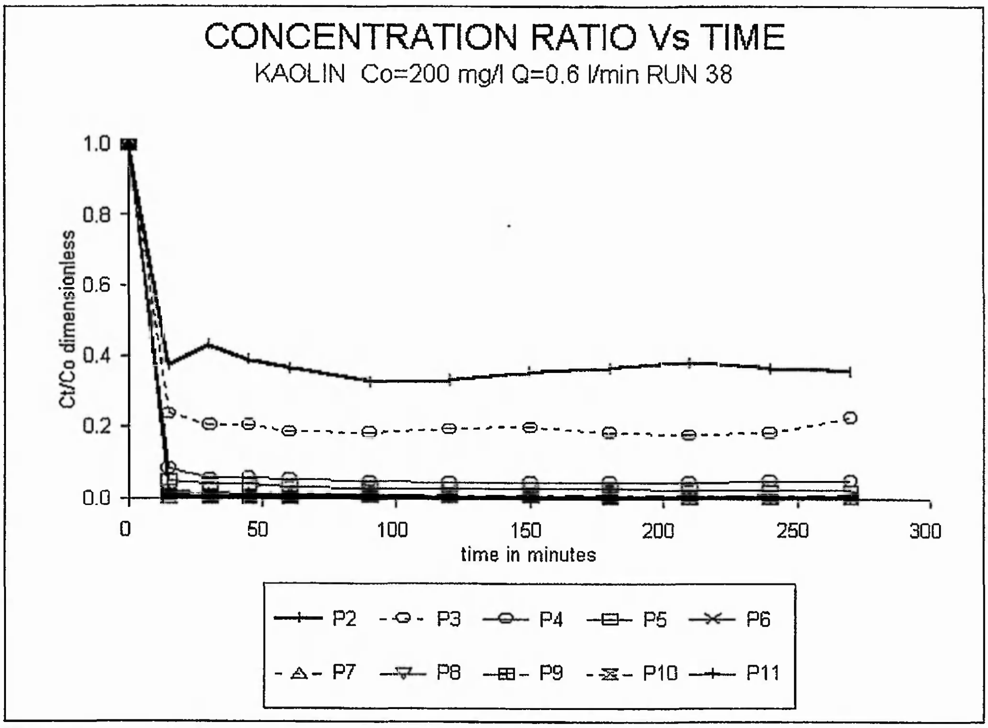


Fig 4.46 Curves of concentration ratio varying with time of a filter run for KAOLIN with C_0= 200 mg/l and Q= 0.6 l/min.

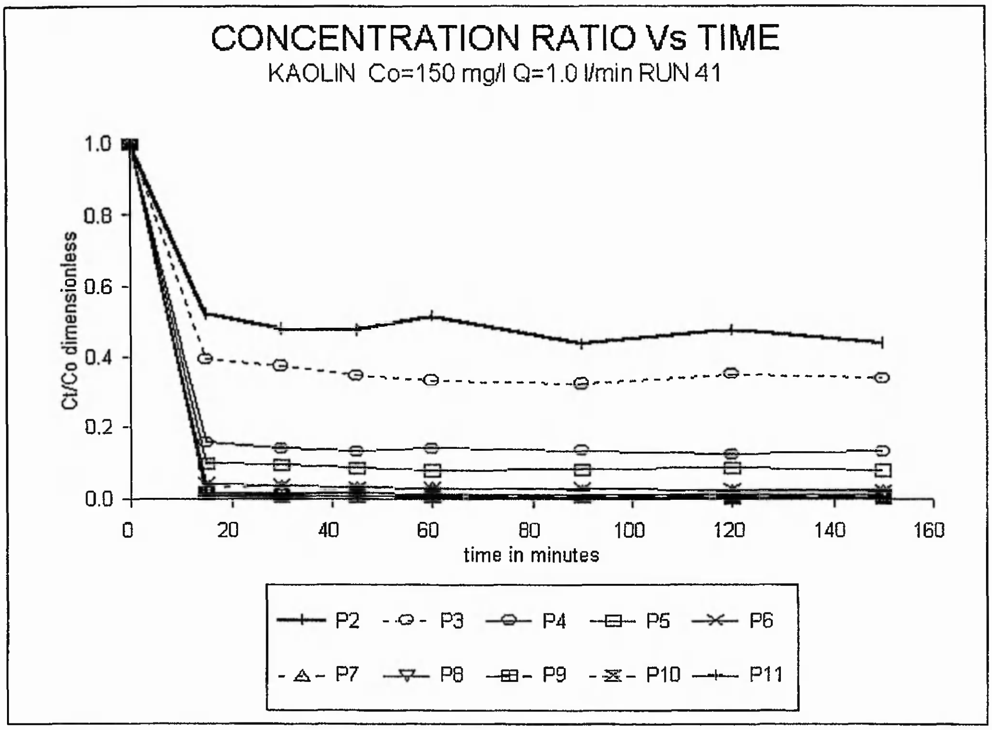


Fig 4.47 Curves of concentration ratio varying with time of a filter run for KAOLIN with C_0= 150 mg/l and Q= 1.0 l/min.

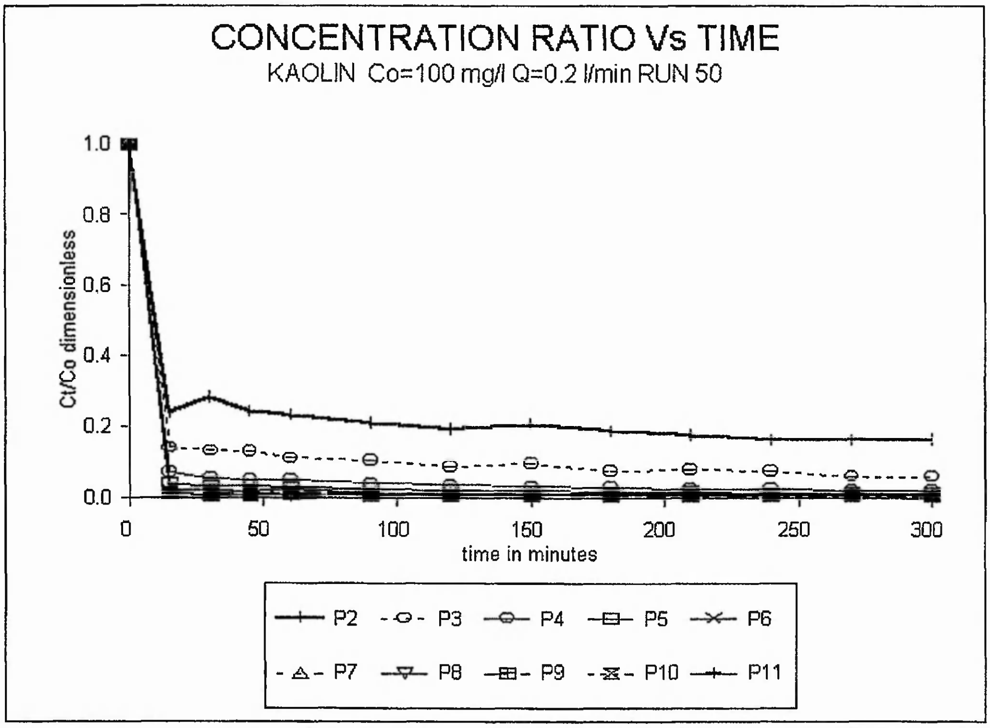


Fig 4.48 Curves of concentration ratio varying with time of a filter run for KAOLIN with C_0= 100 mg/l and Q= 0.2 l/min.

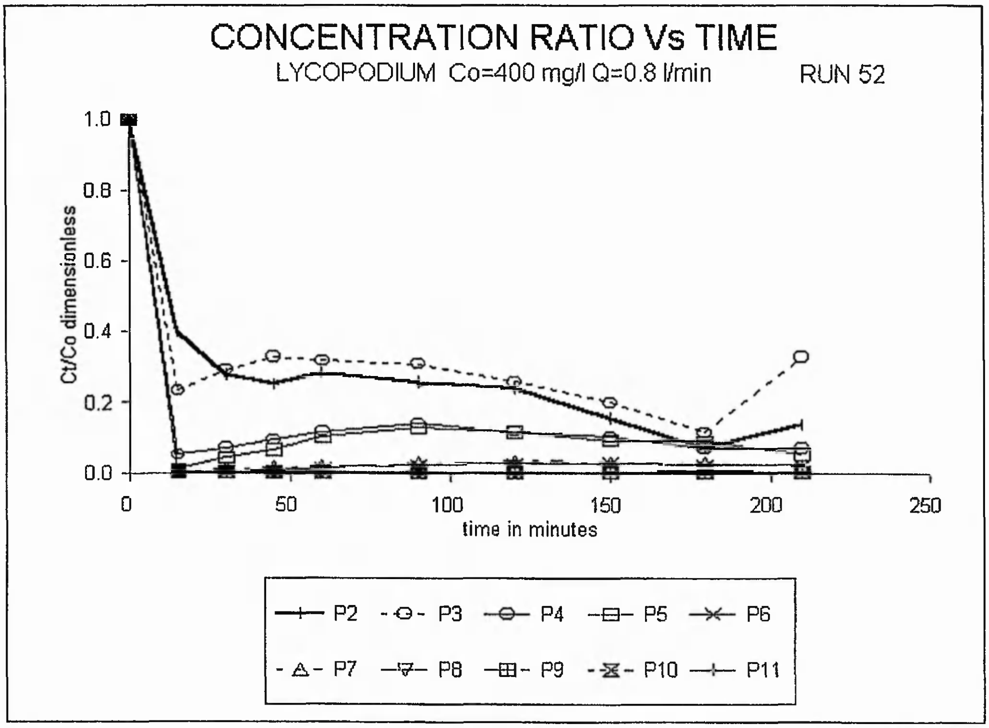


Fig 4.49 Curves of concentration ratio varying with time of a filter run for LYCOPODIUM powder with C_0= 400 mg/l and Q= 0.8 l/min.

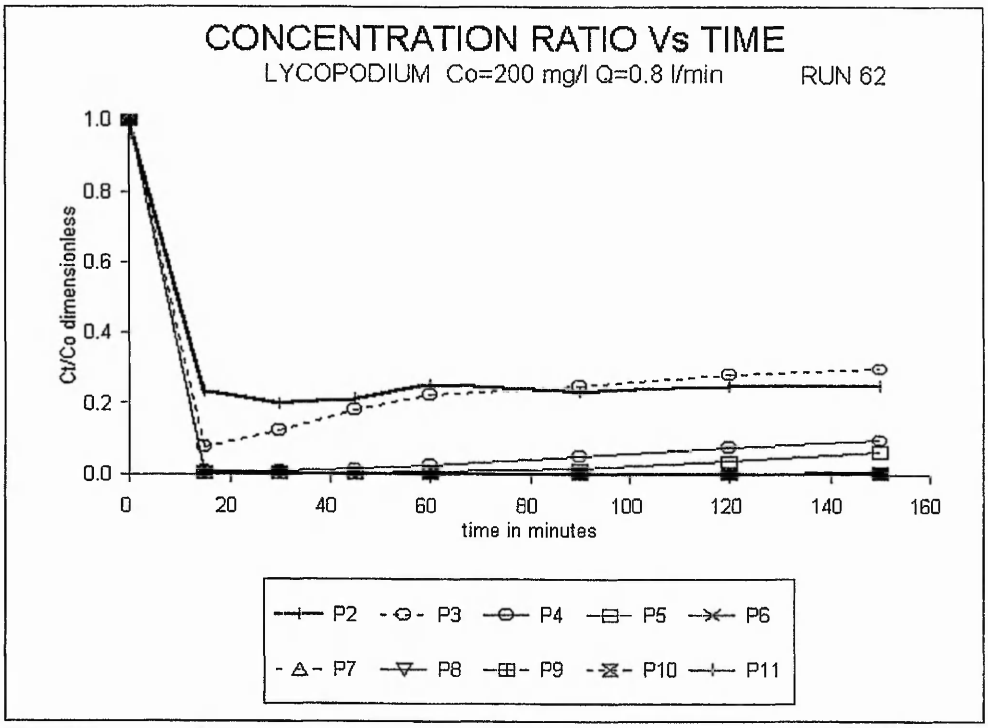


Fig 4.50 Curves of concentration ratio varying with time of a filter run for LYCOPODIUM powder with C_0= 200 mg/l and Q= 0.8 l/min.

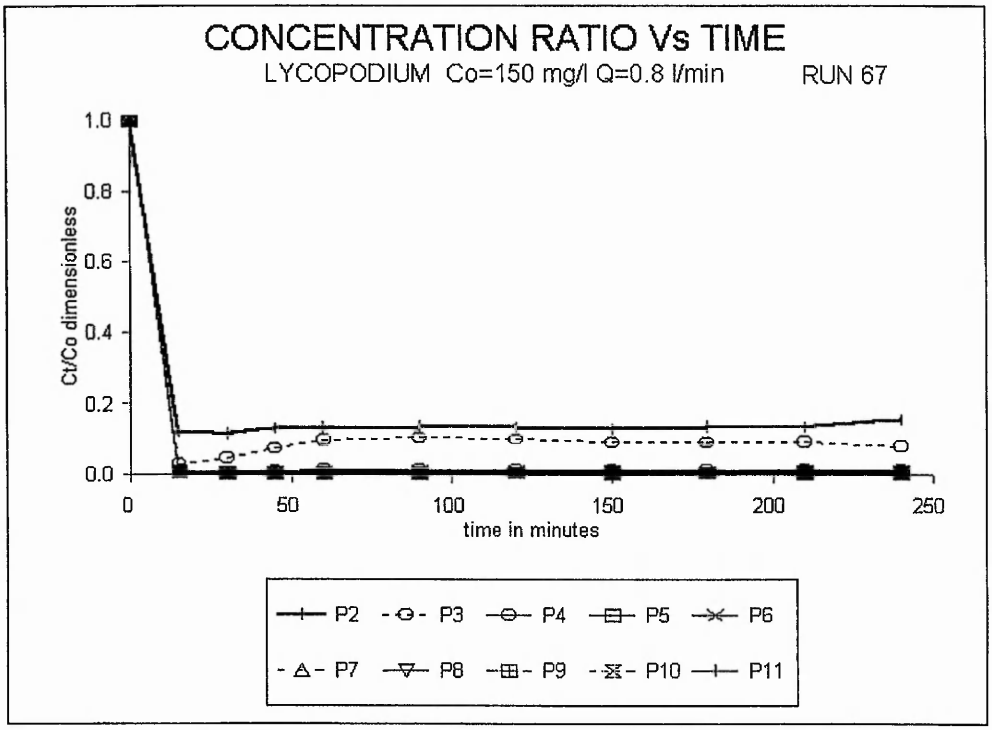


Fig 4.51 Curves of concentration ratio varying with time of a filter run for LYCOPODIUM powder with C_0= 150 mg/l and Q= 0.8 l/min.

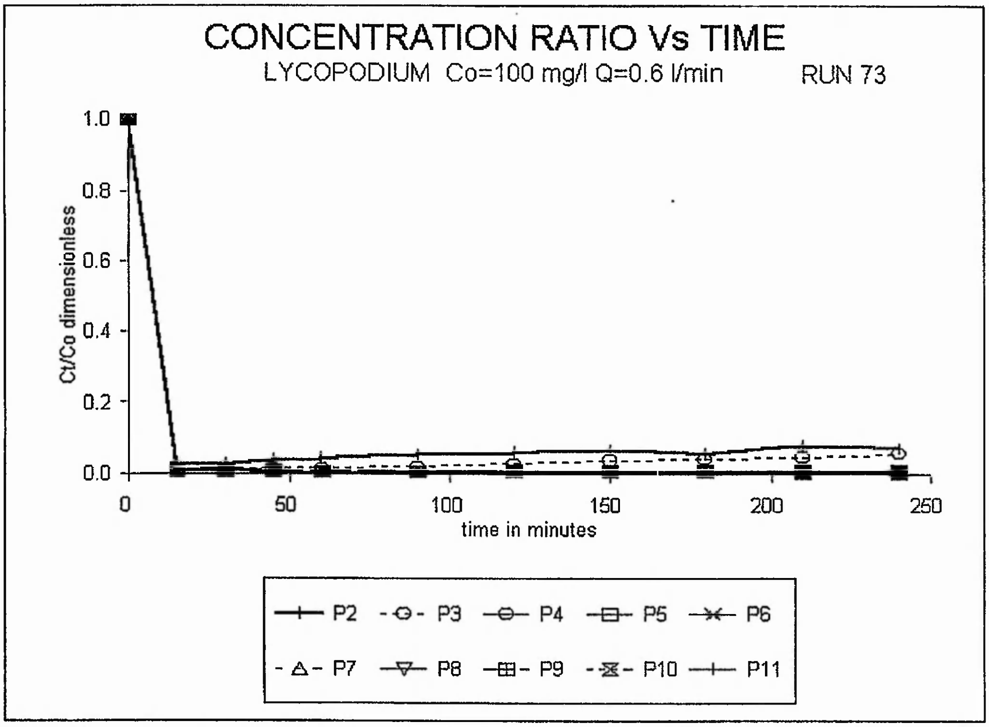


Fig 4.52 Curves of concentration ratio varying with time of a filter run for LYCOPODIUM powder with C_0= 100 mg/l and Q= 0.6 l/min.

4.3 CONCENTRATION RATIO VERSUS DEPTH

As it has been stated before, the dominant mechanisms of deep bed filtration relate to the particles that are significantly smaller than the pores, hence the intrinsic nature of filtration in depth, that is, no straining; action will take place at the inlet face of the bed. If it were a predominant mechanism, the openings or pores in the first layer of grains of the filter bed would soon become blocked, and the subsequent layers never be used (Sholji, 1963; Ives, 1970).

However, it is recognised in practice that a distribution of particle sizes exists in suspensions, and some may exceed the pore size. When the retained suspended solids accumulate entirely on the surface of the filter bed, filtration may then be described *surface filtration* or *absolute straining* as opposed to *depth filtration.* In practice, for heterogeneous filter beds and after backwashing, hydraulic stratification takes place and finer grains are normally stacked on top. This is one of the causes of *surface layer deposition.*

It was observed during the course of the experimental work that with some agglomerated particles of the inflow, absolute straining on ingress occurred when testing with high inlet concentrations of PVC and lycopodium suspensions which led to the build-up of a mat on the inlet surface of the filter media (i.e. compartment number 1). This thin mat was usually only a millimetre or two thick, but gave rise to head losses. The conclusion is that the rate of increase of head loss with time due to surface deposition is exponential (Ives, 1965), as described in more detail in Section 4.4 and Chapter 5.

Ives and Pienvichtr (1965) observed, with the aid of a hand lens, holes through the surface deposit, after stopping filter runs and draining down the superficial water. They also observed that the holes were uniformly distributed over the surface and appeared to have a diameter comparable with that of the filter pores.

As the thickness of the surface deposit mat developed in the first compartment is increased, the loss of head is increased, and the interstitial velocity and hydrodynamic forces in that compartment of the filter are also increased, resulting in a reduction in the number of particles attaching themselves to the mat. It seems plausible, therefore, that by increasing the size of the bed media in the first compartment the formation of a thin surface mat may be avoided and thus improved performance may be achieved.

The clogging front just forms within the upper layer before its advancement into the bed. The penetration of suspended matter then moves to a subsequent compartment of the filter, and the particle removal front moves down in the bed to further compartments.

As described in Sections 3.11 and 4.2, numerous analyses of the concentration readings through the depth of the filter (inlet and outlet concentration readings for each compartment) at various times were made for each filter run. These values were then plotted as changes in the concentration ratio (C/C_0) against depth with time, and conform with the generally accepted curves for rapid gravity filtration, that is, an initial increase in the removal efficiency followed by a slow reduction in the filtrate quality as time progresses, except for PVC turbidity, which showed a slow but steady improvement of filtrate quality over time.

Figures 4.53— 4.58 show typical graphs representing the removal of solids with depth and time during filtration tests for PVC, kaolin and lycopodium turbidities. The conditions for the tests for which the typical graphs are provided are included in the figures. The abscissas show filter depth in millimetres, and the ordinates represent the concentration ratio C/C_0. The limits between compartments are marked by the dotted lines, from P_1 to P_{11} (inlet and outlet sampling ports, respectively, as indicated in Figure 3.15 on page 114), and the direction of flow is represented by the arrows at the top of each graph.

It is evident from the curves shown on the previously-mentioned figures that up-flow filtration gives the worst efficiency results of removing turbidity. Turbidity penetrates the entire depth of the media under almost all conditions of flow, and there is a gradual reduction in turbidity all along the depth, higher in the first five compartments, represented by the shape of the changes in the concentration ratio curve with time and depth.

Although conventional size-graded filters can be operated much more efficiently in up-flow than in the traditional down-flow direction (Diaper and Ives, 1965), the opposite is true in unisize media filtration, that is, when the filtering media is made of unisize grains (no hydraulic stratification occurs), the up-flow filter is clearly less efficient than its down-flow counterpart, under the same initial conditions.

For a given rate of flow at a given temperature, the lower removal efficiency in up-flow filtration, compared with down-flow filtration, may be attributed to any of the following factors or to a combination of them:

- differences in the initial operating porosity values under initial hydrodynamic forces (clean bed)
- hydrodynamic forces due to operating interstitial porosity
- physicochemical properties of the synthetic turbidities
- the weight force of suspended matter
- unstable structure of deposit formed on the lower surface area of the grains

Higher *medium porosity* results in lower filtration efficiency. This would help explain the lower removal efficiency obtained in compartments number 4 and 10, where a recorded difference in initial porosity values of 2 per cent was recorded. The porosity factor does not provide, however, a satisfactory answer for compartments number 2, 6 and 8, where initial porosity was actually lower than in the previous compartment (*see Table 4.1*).

Another factor to be considered is the *hydrodynamic effect* upon both the filter media and the suspended particles. In up-flow filtration it is possible for the head loss above any level to become equal in magnitude to the weight of sand above that level (however this is untrue for any material other than sand and medium porosity above or under 40 per cent). When this happens, the sand bed starts to lift, increasing the porosity and allowing deposited and suspended material to become unstable and be carried away with the filtrate. This phenomenon was observed in compartments number 8 and 10, containing the finer media, and even right from the beginning of the filter run when operating at high rates of flow. On the other hand, the resultant velocity vector in up-flow filtration may also play an important role in the turbidity removal efficiency, as a result of its smaller up-flow net velocity vector, when compared with the down-flow, as it will be described in greater detail in Section 4.5.

The nature of the synthetic turbidities, particularly particle density, is also a key factor in affecting the removal efficiency of the system in up-flow filtration. Thus, for example, a lower density will result in lower removal efficiency, due to a comparatively greater net velocity vector, which will reduce the particles' chances to be removed by gravitational forces (*see Section 4.5*).

Finally, the weight force of suspended matter may help detach the accumulated deposits from the lower surface of the grains facing the flow in upward filtration due to the unstable nature of the deposits' structure on the bottom face of the sand grains. The bouncing of the turbidity particles on such a hanging surface makes it possible for the particles to be easily swept away by the hydrodynamic forces acting in the flow.

In terms of depth, both PVC powder and kaolin suspended particles were removed at a gradually decreasing rate because of their settling velocities values, whereas the rate of removal of lycopodium powder suspension showed the greatest differences: although the overall removal efficiency was very high, the removal efficiency was significantly lower in compartments 6—10.

In other words, the removal efficiency decreases from compartment number 1 to compartment number 10, and also between up-flow and down-flow, even if the quality of the filtrate shows a steady improvement in depth, from inlet to outlet, as it is obvious from the shape of the concentration ratio curves in Figures 4.53—4.58. The effect of the physicochemical properties of the synthetic turbidities on the removal efficiency will be discussed in more detail in Section 4.5.

This *stepped-curve pattern* or serrated outlook of the concentration ratio curves versus depth appears to be characteristic of this novel filtration apparatus as a direct consequence of the alternating use of down-flow and up-flow in the filtration process. However, this trend is not usually noticeable after P_4 or P_6 (about 850—1,500 mm deep), especially as time elapses, and the concentration ratio approaches near zero. That was why the semi-log scale was adopted to highlight the phenomena of the steep (down-flow) and gentle (up-flow) improvement of the filtrate quality with time and depth (see graphs in Appendix 1). Still, the overall performance of the filter model shows a high filtration efficiency at removing suspension particles: in excess of 99 per cent efficiency in 85 per cent of all filter runs.

Figures 4.59a—4.59e show the changes in the turbidity removal efficiency (*see Equation 3.5*) throughout the filter bed for a typical run for PVC, kaolin and lycopodium after 15 minutes of filtration (non-incremental), based on the calculation of the removal efficiency *RE* for each filter compartment, regarded as an independent filter bed:

$$RE = 100\ (1 - \frac{C_1}{C_0}) \qquad [3.15]$$

Where C_0 is the incoming concentration and C_1 the outgoing concentration for each filter compartment. C_1 will be the C_0 of the following compartment in turn.

As expected, Figures 4.59f—4.59h show that the higher the initial concentration of suspended particles the lower the removal efficiency. Again, from all of these graphs it is obvious that up-flow filtration gave lesser removal results for the three types of turbidities and different initial conditions. Both PVC and kaolin presented a similar behaviour, whereas lycopodium showed a somewhat less defined pattern, more erratic behaviour, which may be attributed to the light density and larger size and settling velocity of the lycopodium particles, giving better results as a direct consequence of the straining action.

In some cases, and Figure 4.59a provides a good example, a negative removal efficiency was observed in up-flow filtration (i.e. the effluent concentration was higher than the influent concentration), particularly in those compartments containing the finest media. Such phenomenon could be a result of the medium's instability caused by the uplift hydrodynamic forces, when the up-flow velocity matches the incipient fluidising velocity and dislodging of deposited particles may take place.

Experience had shown that intermittent sampling could lead to disturbances of the deposits in the filter pores. For this reason, great care was taken over the sampling technique used in order to prevent the local scour of deposits that could affect the validity of the measurements. Thus, the sampling ports were placed outside the filter bed (at the inlet and outlet of each compartment) in order to avoid disturbance of the filter bed during the collection of samples. However, and as described above, some times the results were unreliable and inconsistent, with higher turbidity levels than previous ones, especially those regarding the samples collected after up-flow filtration, and even for down-flow samples, as shown in Figures 4.59a— 4.59h.

Additionally, despite the measures taken to prevent the bed expanding, the finer media contained in compartment number 10 (and number 8 to a lesser degree) suffered from a phenomenon colloquially referred to as *sand boiling effect*, a sign of incipient bed fluidisation, as a result of the uplift force caused by the pressure difference across the sand bed during up-flow filtration at higher rates. It was assumed therefore, that a detachment of particles had occurred to some degree, thus negatively affecting the filtrate quality, as it is obvious from Figure 4.53 to Figure 4.58.

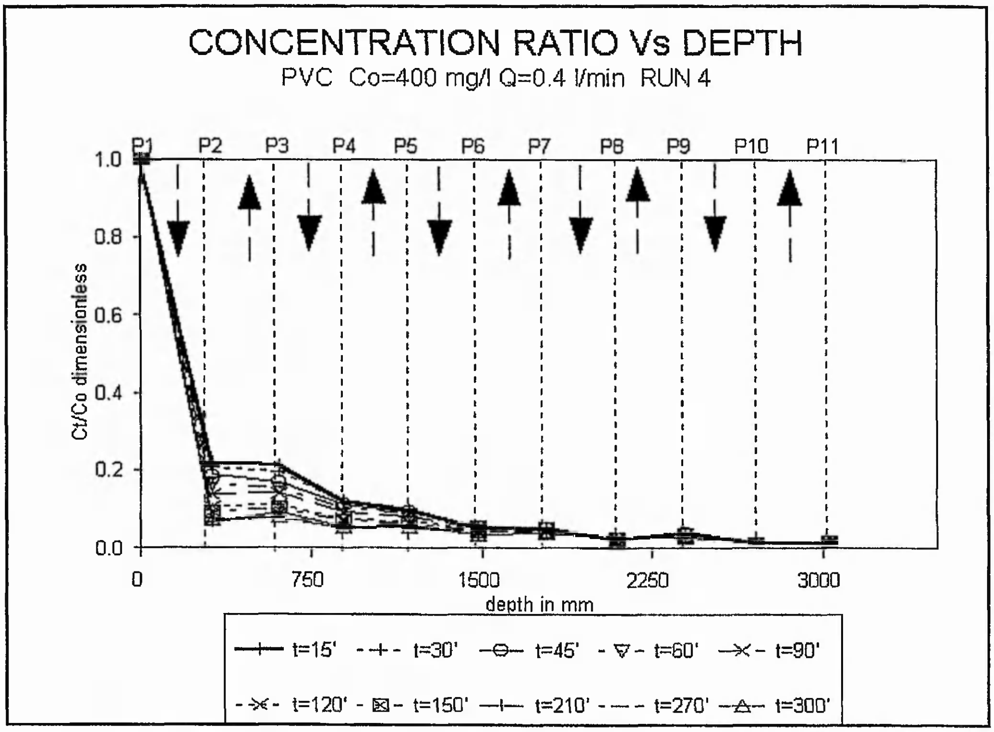


Fig 4.53 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for PVC C_0=400 mg/l at 1.0 l/min.

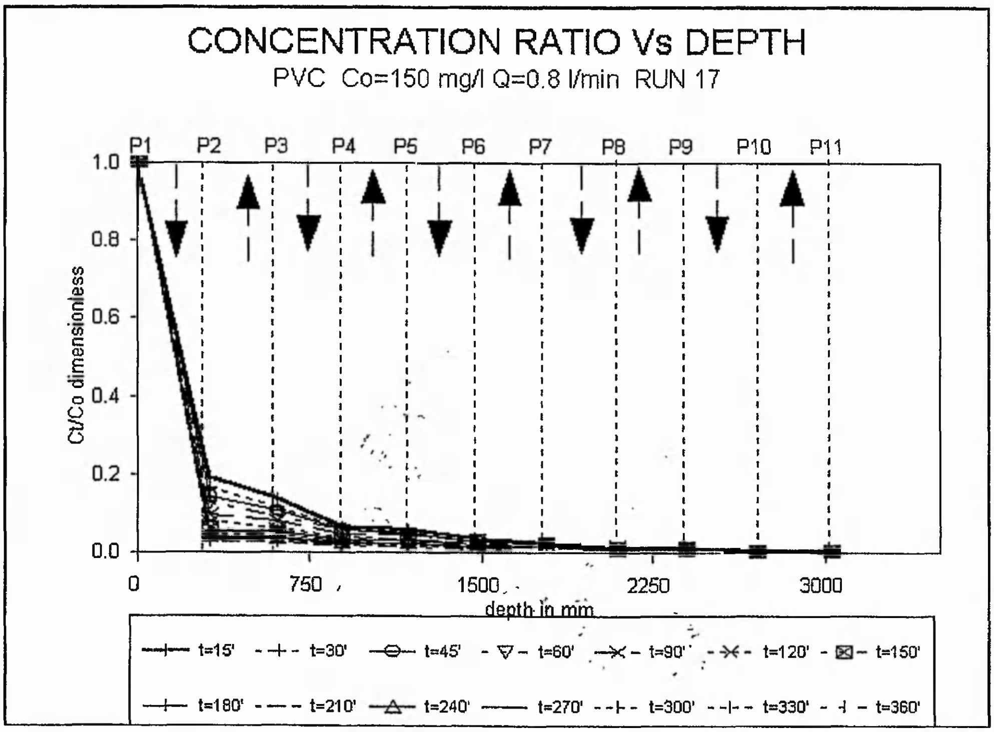


Fig 4.54 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for PVC C_0=150 mg/l at 0.8 l/min.

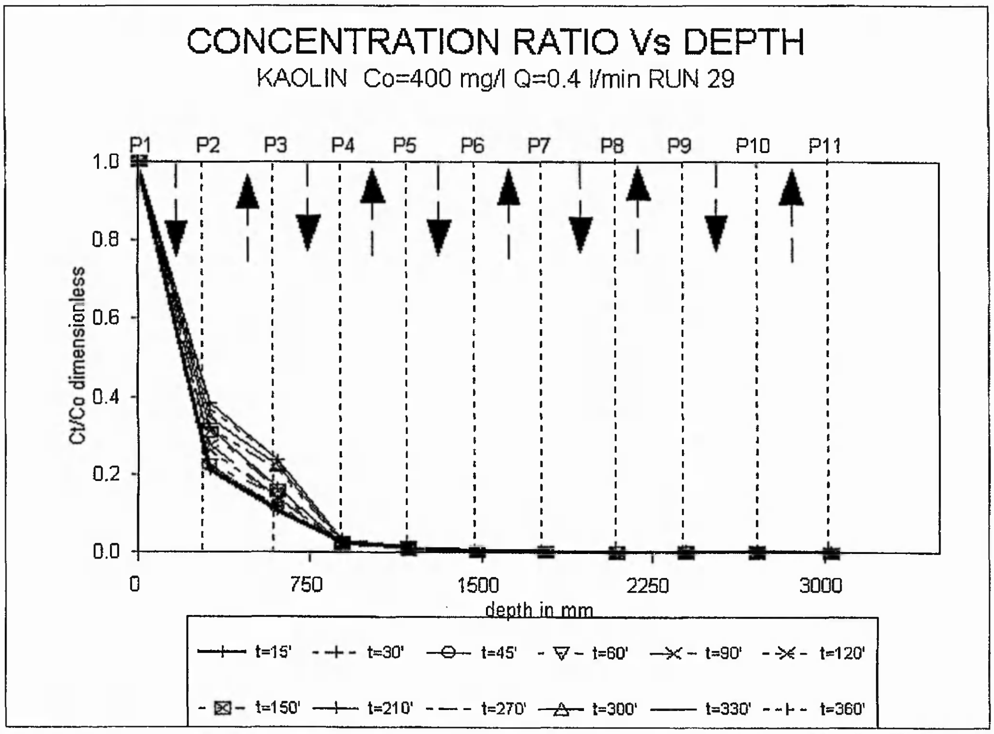


Fig 4.55 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for Kaolin C_0=400 mg/l at 0.4 l/min.

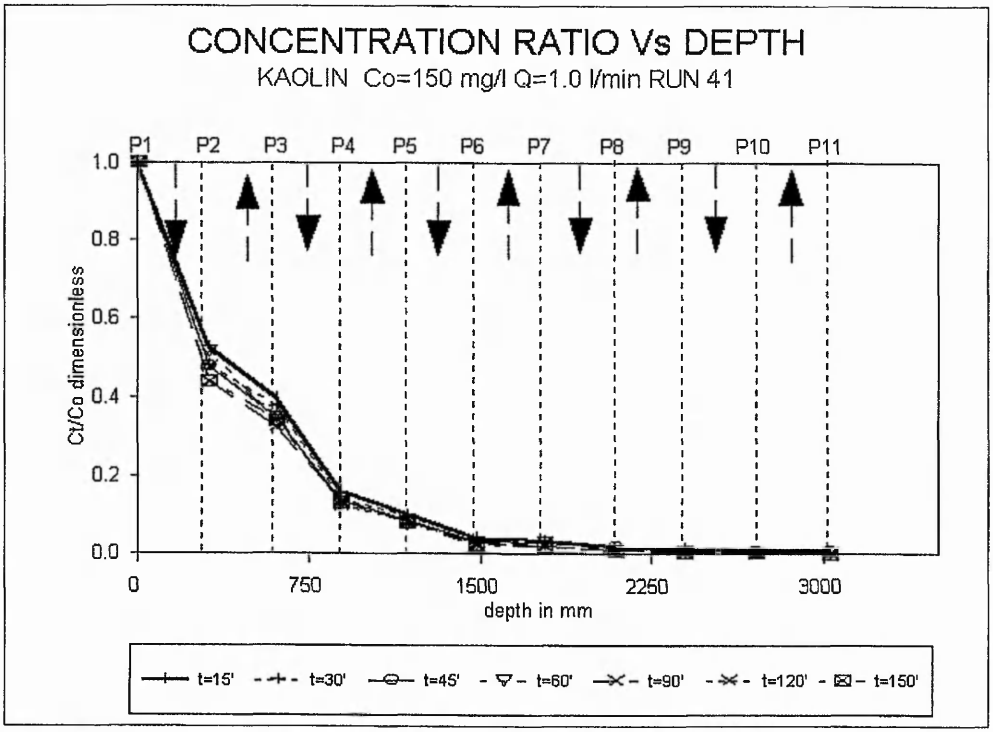


Fig 4.56 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for Kaolin C_0=150 mg/l at 1.0 l/min.

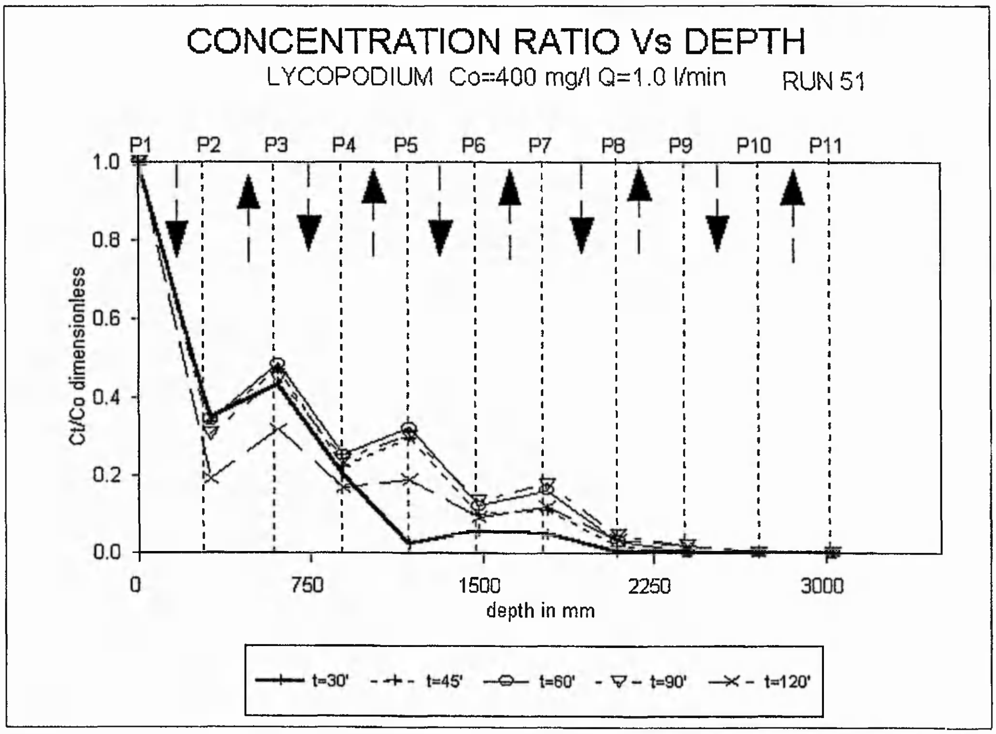


Fig 4.57 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for Lycopodium powder C_0=400 mg/l at 1.0 l/min.

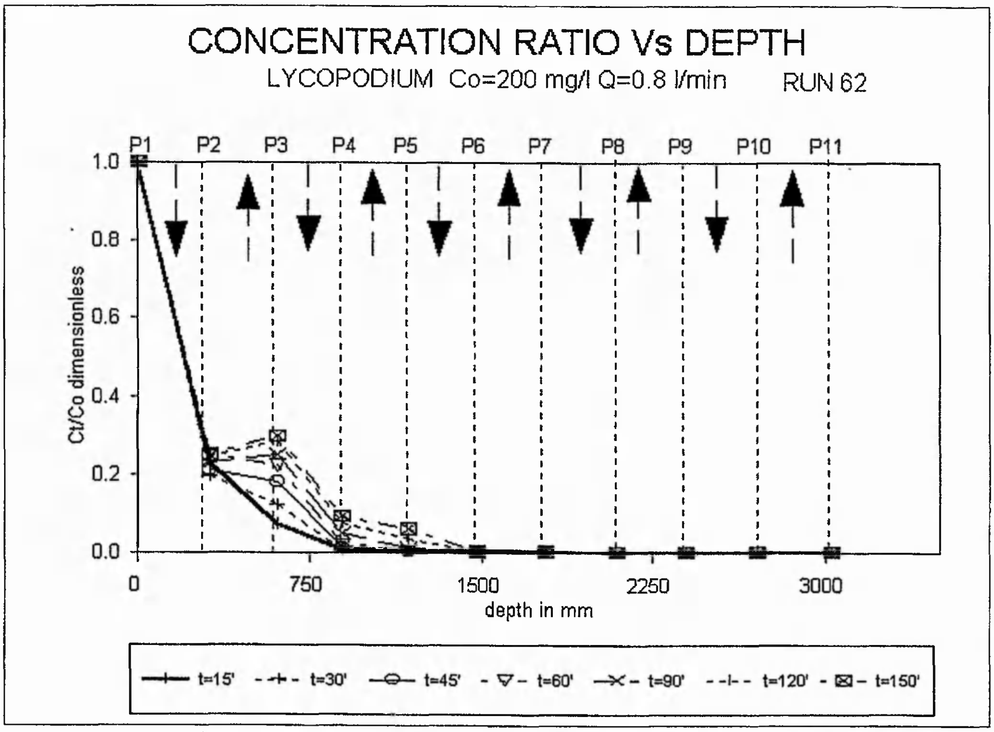


Fig 4.58 Curves of suspension concentration ratio varying with depth in the filter media and time of filter run for Lycopodium powder C_0=200 mg/l at 0.8 l/min.

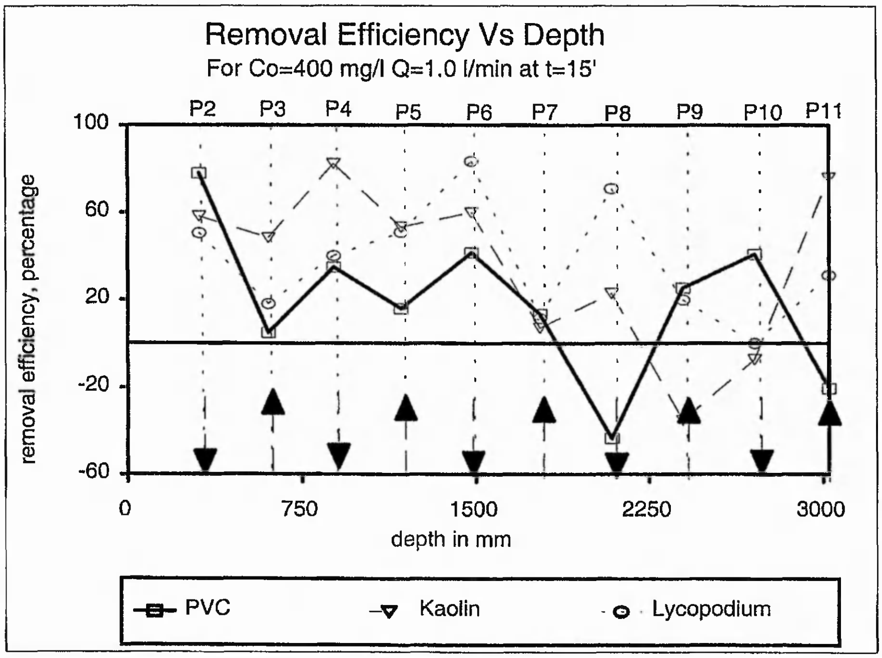


Fig 4.59a Removal efficiency versus depth of representative filtration runs for PVC, kaolin and lycopodium powder suspensions after 15 minutes of filtration time (non-incremental).

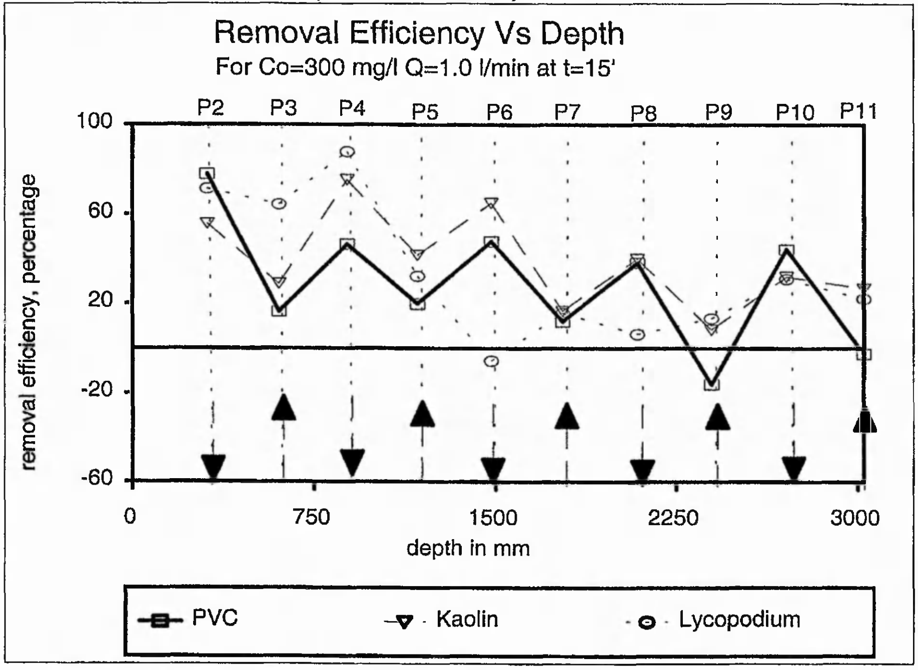


Fig 4.59b Removal efficiency versus depth of representative filtration runs for PVC, kaolin and lycopodium powder suspensions after 15 minutes of filtration time (non-incremental).

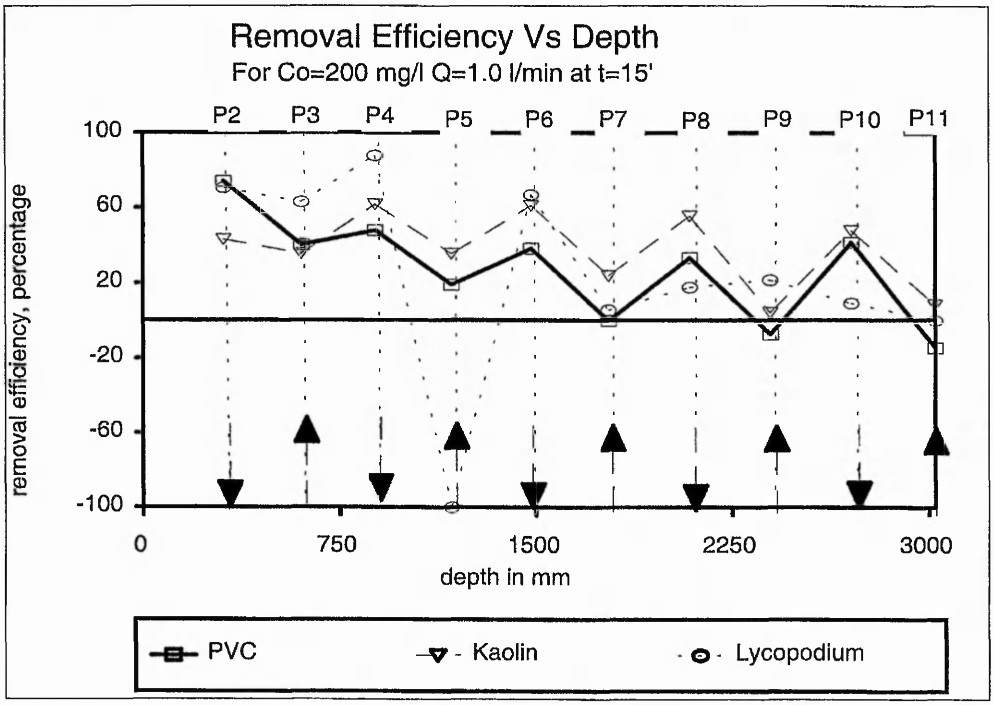


Fig 4.59c Removal efficiency versus depth of representative filtration runs for PVC, kaolin and lycopodium powder suspensions after 15 minutes of filtration time (non-incremental).

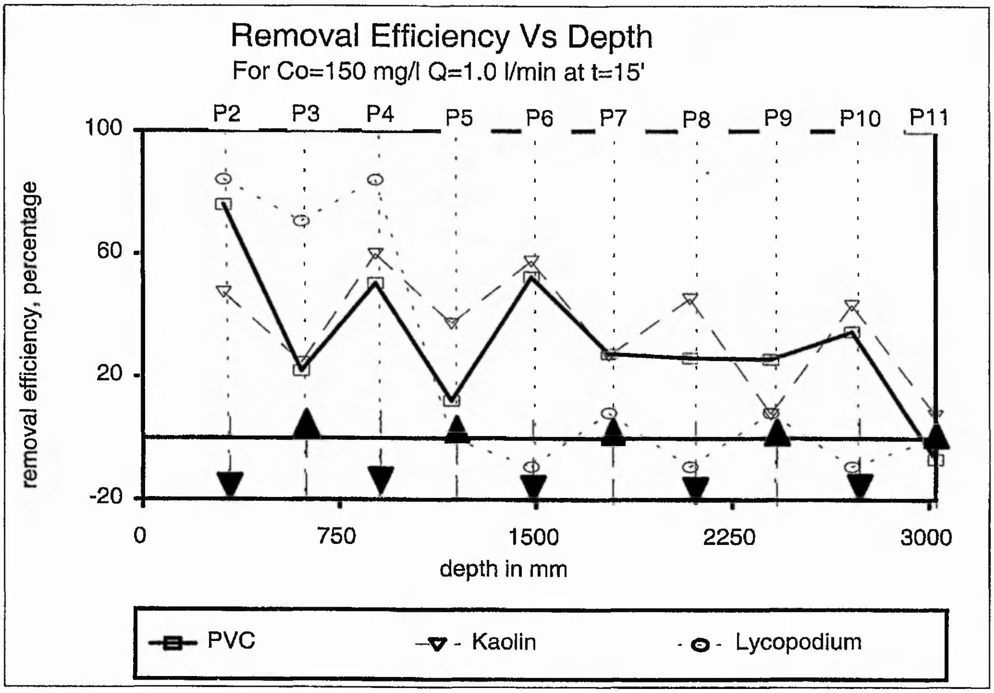


Fig 4.59d Removal efficiency versus depth of representative filtration runs for PVC, kaolin and lycopodium powder suspensions after 15 minutes of filtration time (non-incremental).

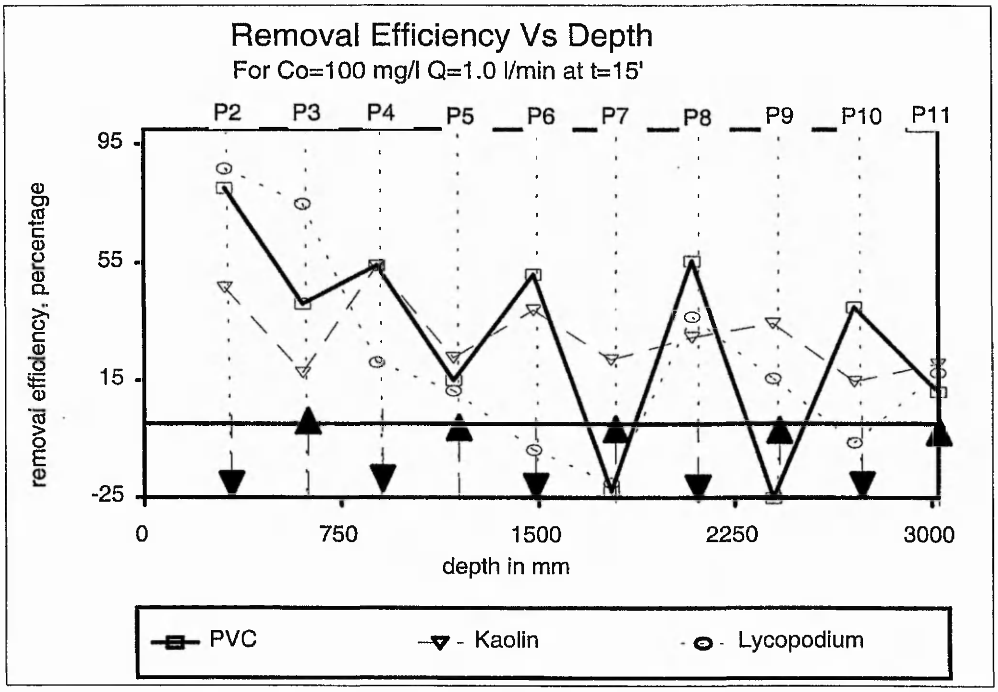


Fig 4.59e Removal efficiency versus depth of representative filtration runs for PVC, kaolin and lycopodium powder suspensions after 15 minutes of filtration time (non-incremental).

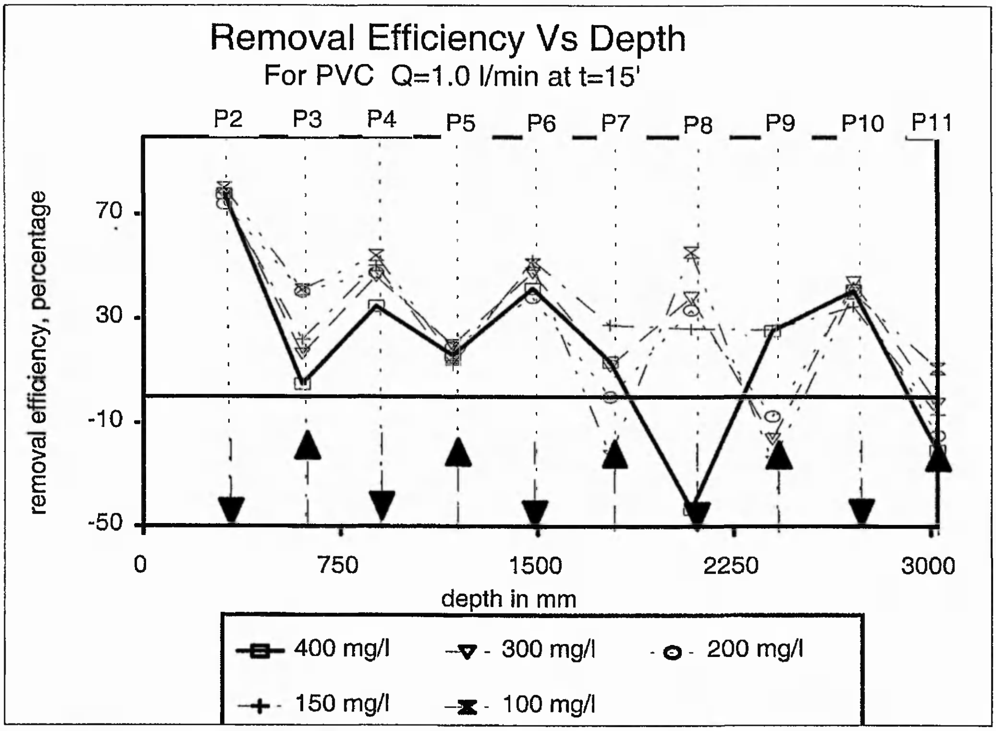


Fig 4.59f Removal efficiency versus depth of representative filtration runs for PVC at varying inlet concentrations after 15 minutes of filtration time (non-incremental).

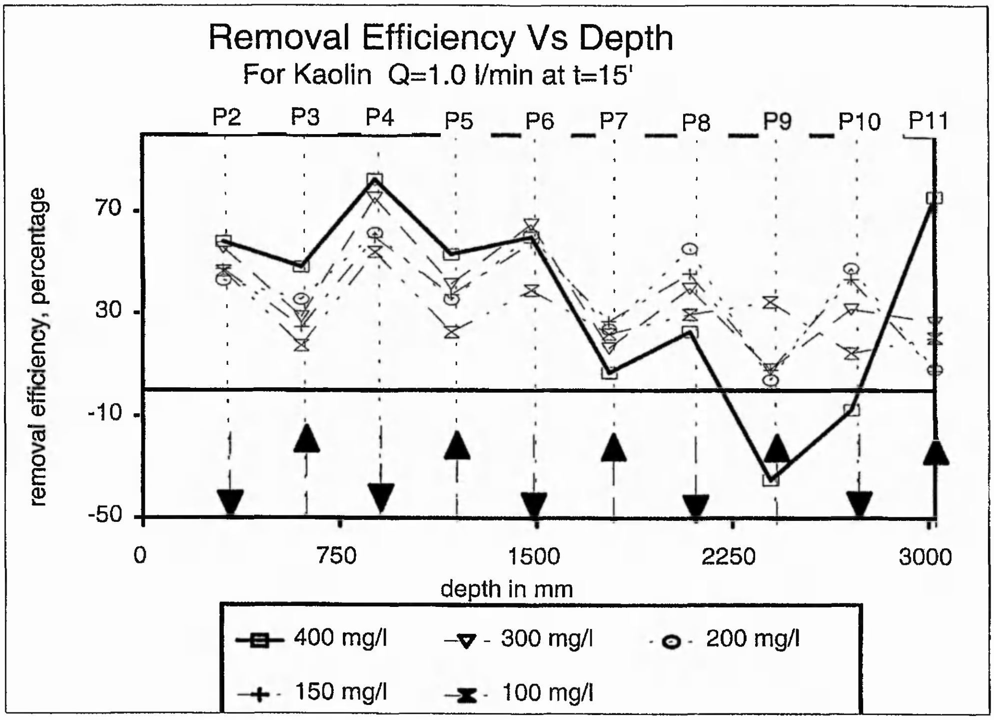


Fig 4.59g Removal efficiency versus depth of representative filtration runs for KAOLIN at varying inlet concentrations after 15 minutes of filtration time (non-incremental).

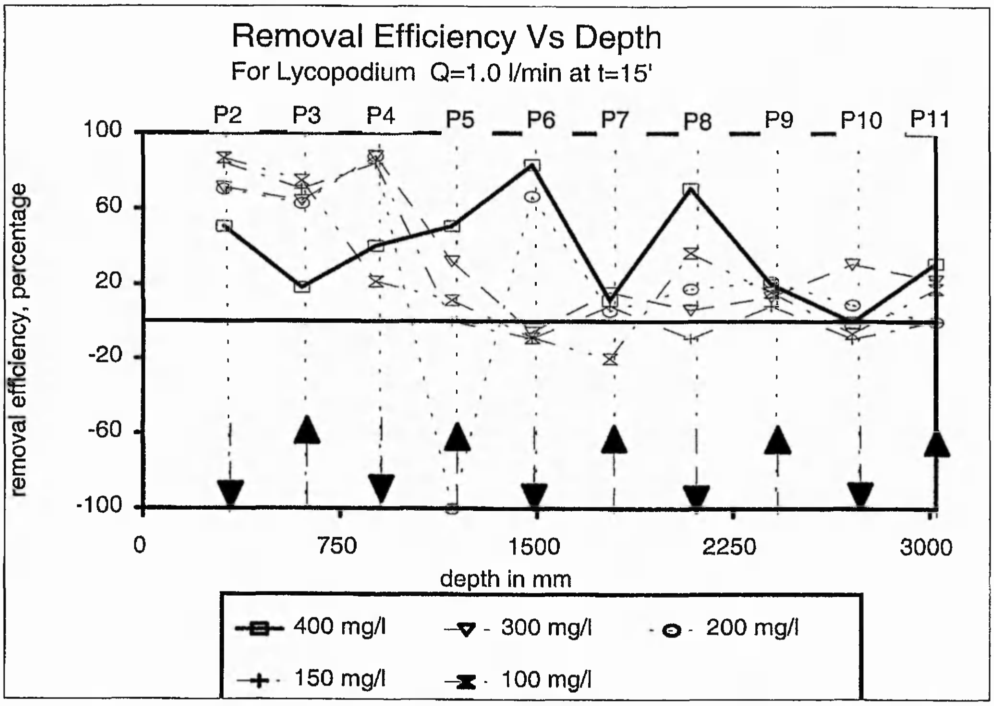


Fig 4.59h Removal efficiency versus depth of representative filtration runs for LYCOPODIUM at varying inlet concentrations after 15 minutes of filtration time (non-incremental).

4.4 HEAD LOSS VERSUS TIME

Unlike the quality of the filtrate, the head loss developed during filtration is very dependent upon media size, filtration rate, fluid temperature, initial porosity, and the concentration of suspended particles to be filtered.

Slight changes in the initial porosity can have a pronounced effect on the filtration efficiency (*see Section 4.1 and Table 4.1*). In addition to this, the deposition of turbidity material from suspension into the sand pores causes a progressive loss of permeability. Porosity changes due to further deposition and increase in the specific deposit value; consequently an increasing pressure difference was necessary to maintain a given flow rate.

The depth of penetration of solids into the filter bed can be assessed indirectly by the pressure drop that develops in each compartment. Thus, as described in Section 3.2, the filter was equipped with pressure taps and manometers to measure the head loss distribution within the filter's compartments.

Augmented by simple observation, the head loss data may permit graphical determination of the frontal advancement of the solids, and also of the filter holding capacity for each compartment.

Therefore, the head loss readings for each compartment were plotted in order to evaluate the head loss development as the filter run progressed. Typical graphs representing the head loss variations across the filter's depth during filtration tests for PVC, kaolin and lycopodium turbidities are shown in Figures 4.60— 4.71. The depth of the filter is represented by each one of the pressure taps (noted as H_1, H_{1-2}, H_2 ... as indicated in Figure 3.15, page 114). The abscissas show filtration time in minutes, and the ordinate represents the head loss in millimetres water gauge.

These figures clearly indicate that the head loss developed across the filter bed with time increased gradually from top to bottom, i.e. from compartment number 1 to compartment number 10, for all turbidities and filtration conditions.

A close look at typical results of one of the filter runs, can serve as an illustration. Run number 21 used a homogeneous suspension of 100 mg/l of PVC powder filtered at a rate of 1.0 l/min. Figures 4.63 and 4.44 show the changes of head loss and turbidity observed during this experiment. It is evident that turbidity removal and loss of head are directly related. Figure 4.44 shows clearly the typical characteristics of rapid filtration, that is, a gradual and steady improvement of the filtrate quality over time and throughout the filter's depth (before filter breakthrough takes place towards the end of a normal-length filter run).

The formation of surface mats and quick clogging of pore spaces within the bed were two major problems encountered when operating at high initial concentrations (400—300 mg/l) and flow rates (1.0—0.8 l/min), especially with PVC and lycopodium turbidities

The effect of surface deposition upon the head loss can be graphically recognised as shown on Figures 4.60, 4.61 and 4.68. It is therefore possible to determine how much of the filter material is being utilised. The pressure line which reaches the atmospheric pressure value defines the limit of head loss (i.e. 2000 mm) and therefore it has not been plotted beyond that value, as the pressure in the filter should not fall below atmospheric.

Under these circumstances, the rate of head loss development increased rapidly, as it can be noticed on Figure 4.61, when the input to the filter was a 400 mg/l PVC suspension supplied at a rate of flow of 0.8 l/min. Within a very short time compartment number 1 reached saturation and the concentration ratio for that compartment quickly approached to one, while the remainder of the filter bed remained relatively clean, even though some signs of saturation were already noticeable in compartment number 2 by the end of the experiment, as represented in Figure 4.42.

Figures 4.43 and 4.62 show data for PVC obtained from a run at an inlet concentration of 300 mg/l and filtration rate 0.6 l/min and provide an example of the trends observed for turbidity removal and head loss development throughout the study, that is, a progressive decrease of the final suspended solids concentration (Figure 4.43) and a steady increase of the loss of head (Figure 4.62) with time. Additionally, it can be seen that most of the removal occurred in the top third of the filter, i.e. compartments 1 to 4 containing the coarser media: The concentration ratio varies from 1.0 to 0.1 within the first four compartments of the filter bed, at a depth of 1,176 mm over a total of 3,023 mm.

Typical head loss curves for kaolin are represented in Figures 4.64—4.67. On the other hand, Figures 4.45—4.48 show the concentration ratio curves for the same filter runs. As with PVC, most of the particles were captured within the first compartments. However, the concentration ratio curves for compartments 1, 2 and 3 indicate a lower filtrate quality than that obtained with PVC. Head loss data for kaolin suggest that no straining or surface deposition occurred, more probably due to the greater particle size range showed by the kaolin turbidity (2µm 80%; up to 53 µm 20%), which caused a greater particle distribution across the filter depth.

Contrary to the results obtained with PVC, no significant improvement in the filtrate quality of the kaolin turbidity was achieved as the filter run progressed, as it can be deduced from the low gradient showed by the head loss curves represented in Figures 4.64—4.67, as well as the greater distance between the concentration ratio curves in Figures 4.45—4.48.

In order to compare the total head loss development for the three types of synthetic turbidities under identical initial conditions, total head loss readings were plotted as shown in Figures 4.72—4.80. These figures show data obtained from typical runs for PVC, kaolin and lycopodium at a filtration rate of 1.0 l/min and varying inlet concentration, and provide an example of the trends that were observed for total head loss development throughout the study.

Lycopodium turbidity showed a similar trend to that of kaolin in terms of removal efficiency even if at a cost of greater head losses. Usually, lycopodium turbidity showed the highest *initial* head loss, when compared with kaolin and especially PVC suspensions (Figures 4.72—4.80), and this may be attributed to the larger size of the lycopodium particles (35 µm), initially clogging the filter pore spaces faster than the smaller PVC and kaolin particles. This first head loss value obtained after five minutes of the commencement of the filtration test was taken as the *initial* head loss of the system.

Besides, surface deposition was observed when clarifying lycopodium suspension at high rates of flow and inlet concentrations. This caused a greater rate of increase of the head loss development, with an upward curve as the filter run progressed, characteristic of surface deposition, as observed during run 52, for an initial particle concentration of 400 mg/l and flow rate 0.8 l/min (Figure 4.68).

Again, this behaviour may be due to the physicochemical properties of the lycopodium turbidity (lower density, higher particle size and uniform particle size distribution) and the porosity of the lycopodium deposits.

The increase in the applied concentration also causes a corresponding shortening of filter runs. This is evident from the head loss curves on Figures 4.72 and 4.73 when compared to those on Figures 4.74 and 4.75. At the highest inlet concentration and filtration rate, the test had to be terminated after only 2 hours, whereas for the lower inlet concentrations significantly longer runs were achieved.

Besides, from the above-mentioned figures it can be seen that usually both kaolin and lycopodium turbidities present a similar trend in their head loss increase patterns in comparison to that of PVC.

Although the head loss for PVC turbidity is initially lower than that for kaolin and lycopodium, as time elapses, PVC head loss curves rise steeply and ultimate reach, and often exceed, the values obtained with the other two synthetic turbidities. The reason for this can be attributed to their distinctly dissimilar physicochemical properties (particle size range, shape and density); PVC particles are finer and can fill up the pore spaces and produce greater head loss due to frictional losses and more surface area per unit volume of deposit for resistant to flow, as it will be discussed in more detail in Chapter 5.

If the head loss across the filter increases proportionally with time, the filtration is taking place within the filter pores in the depth of the filter. If it increases as an upward curve with time, then some of the head loss is due to a surface mat on the inlet face of the filter material, thus shortening the filter cycle as shown in Figure 4.60. This surface mat was observed to be a layer with holes in it, which persisted in spite of an appreciable thickness (1—3 mm) of the surface layer (Ives, 1965).

Figures 4.81—4.83 show the effect of the rate of flow on total head loss development for PVC, kaolin and turbidity. As it would be expected, these figures show that a higher filtration rate gives a shorter filter run, as a result of higher head losses.

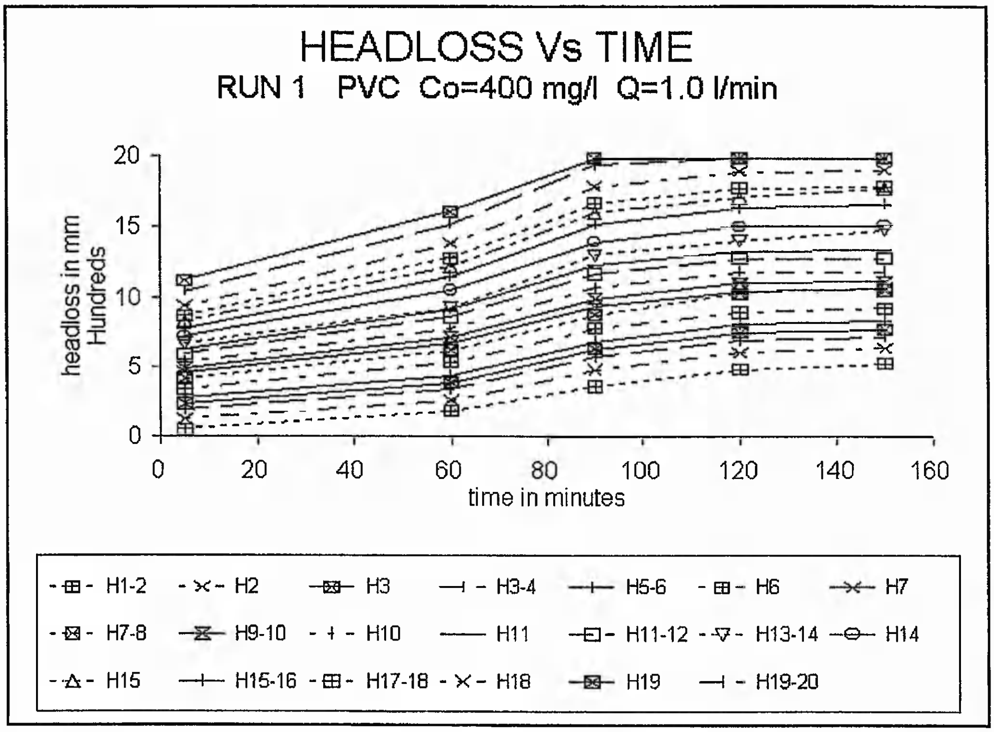


Fig 4.60 Head loss varying with depth and time for PVC with C_0=400 mg/l and Q= 1.0 l/min.

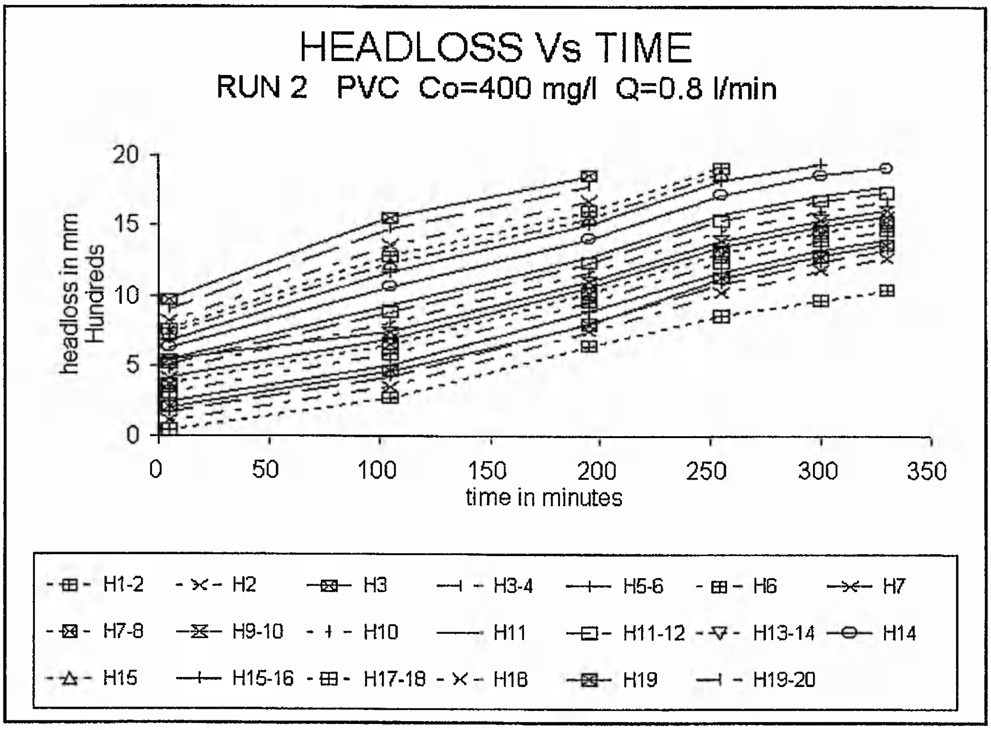


Fig 4.61 Head loss varying with depth and time for PVC with C_0=400 mg/l and Q= 0.8 l/min.

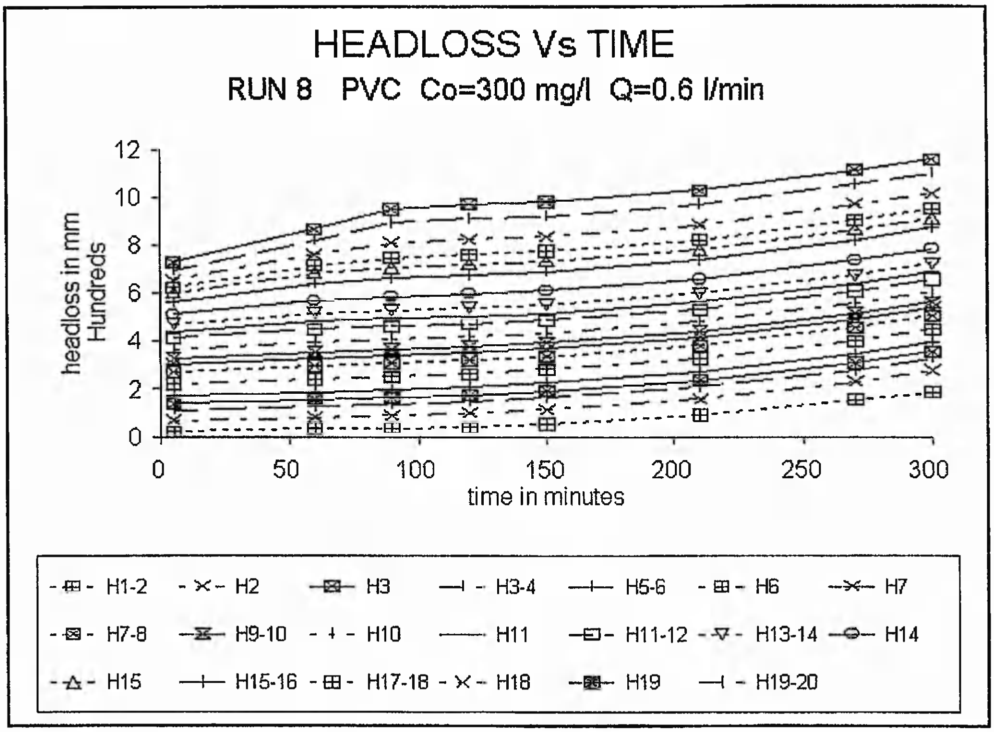


Fig 4.62 Head loss varying with depth and time for PVC with C_0=300 mg/l and Q= 0.6 l/min.

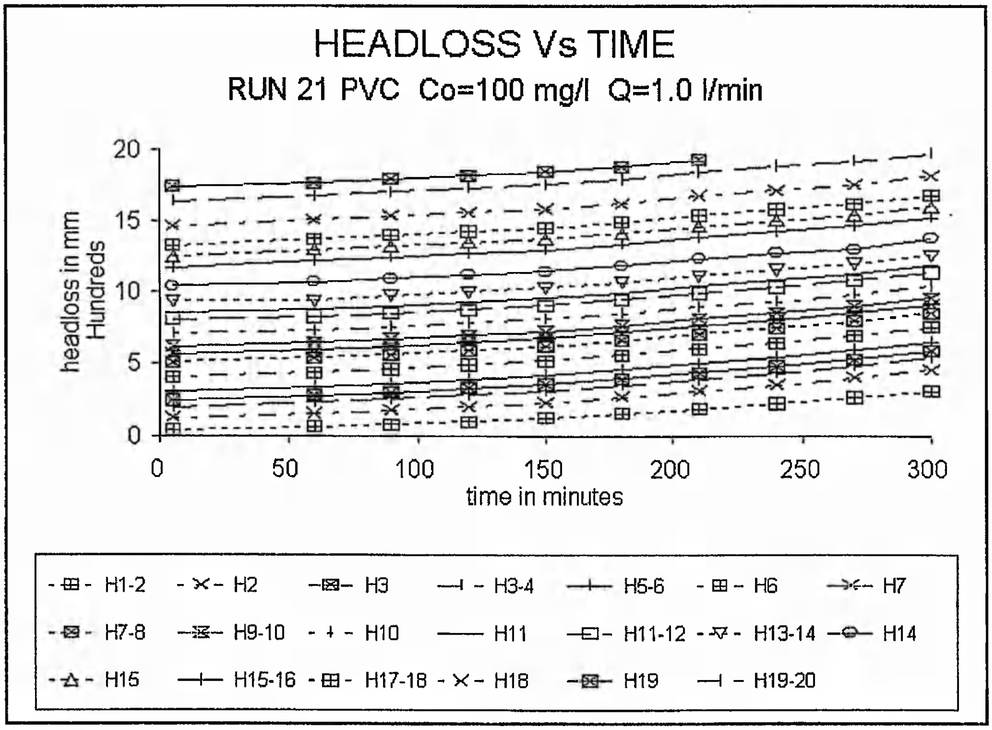


Fig 4.63 Head loss varying with depth and time for PVC with C_0=100 mg/l and Q= 1.0 l/min.

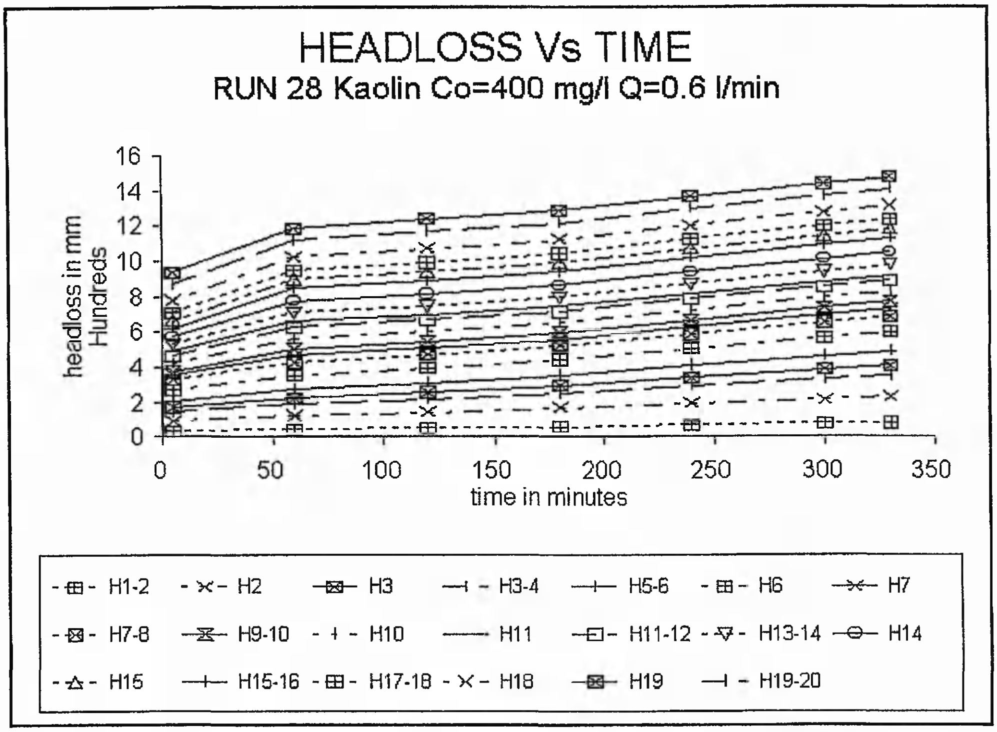


Fig 4.64 Head loss varying with depth and time for kaolin with C_0=400 mg/l and Q= 0.6 l/min.

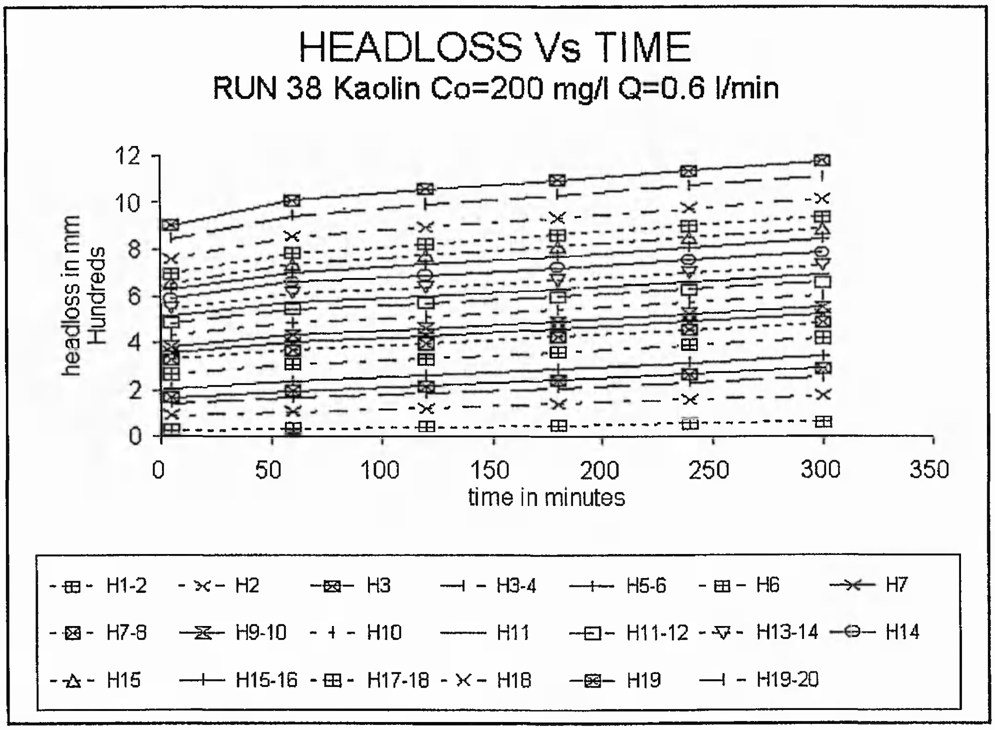


Fig 4.65 Head loss varying with depth and time for kaolin with C_0=200 mg/l and Q= 0.6 l/min.

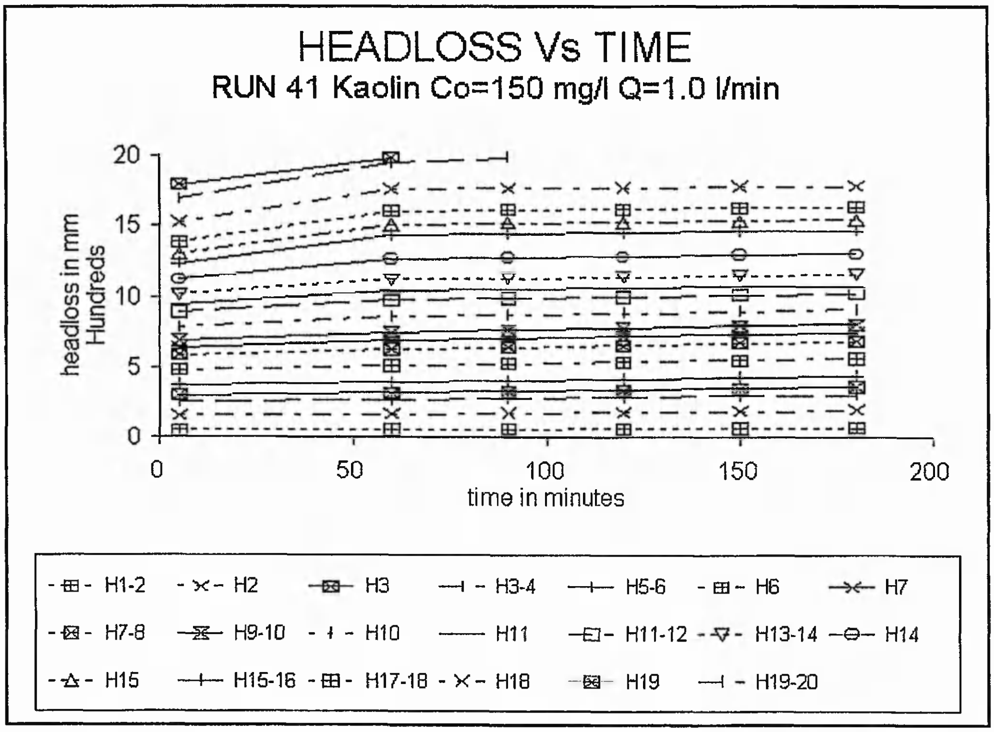


Fig 4.66 Head loss varying with depth and time for kaolin with C_0=150 mg/l and Q= 1.0 l/min.

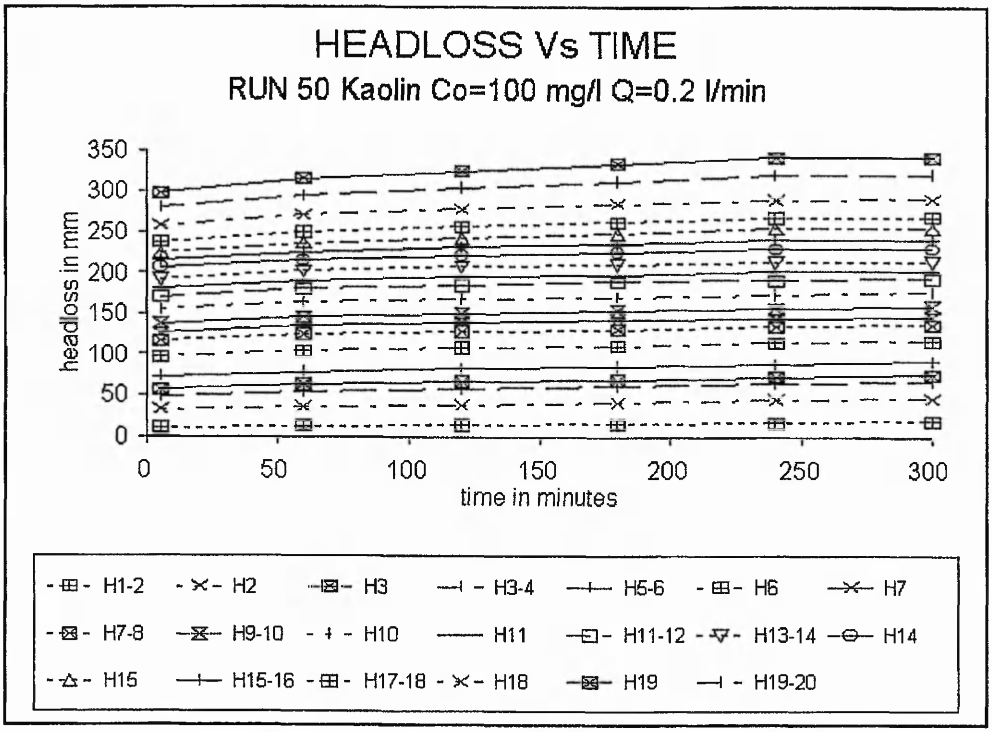


Fig 4.67 Head loss varying with depth and time for kaolin with C_0=100 mg/l and Q= 0.2 l/min.

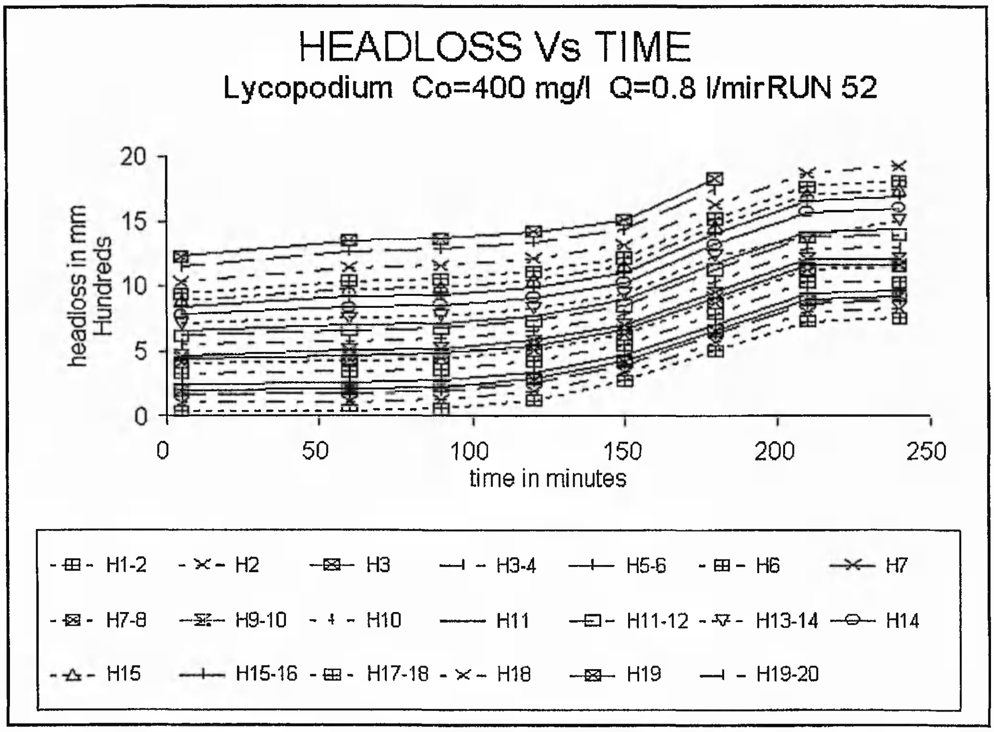


Fig 4.68 Head loss varying with depth and time for lycopodium powder with C_0=400 mg/l and Q= 0.8 l/min.

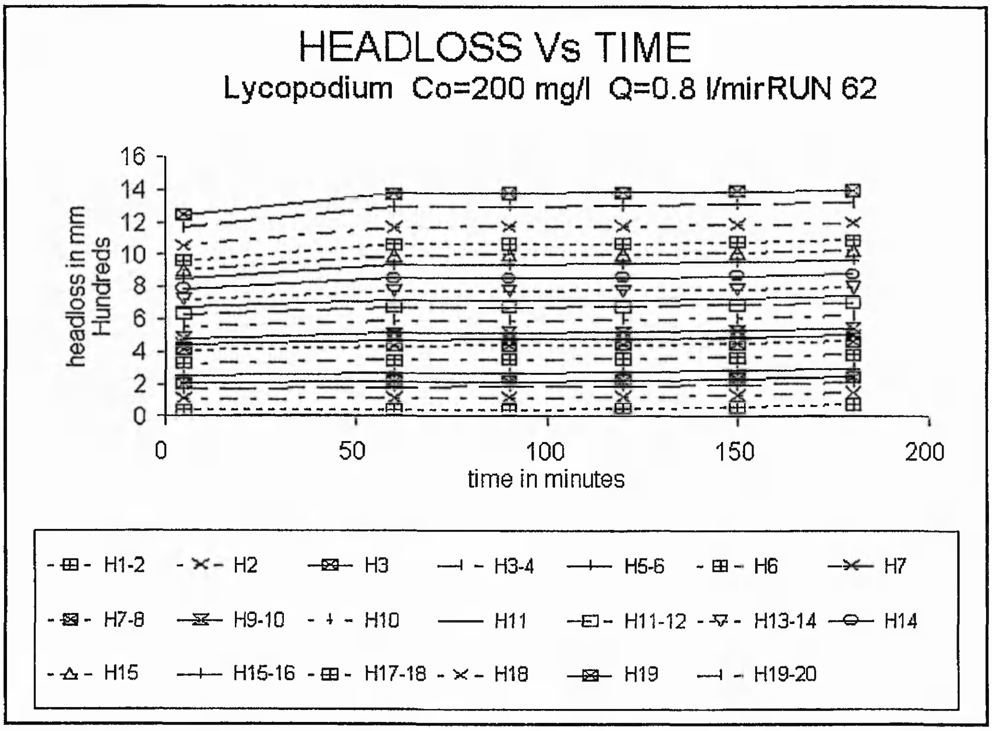


Fig 4.69 Head loss varying with depth and time for lycopodium powder with C_0=200 mg/l and Q= 0.8 l/min.

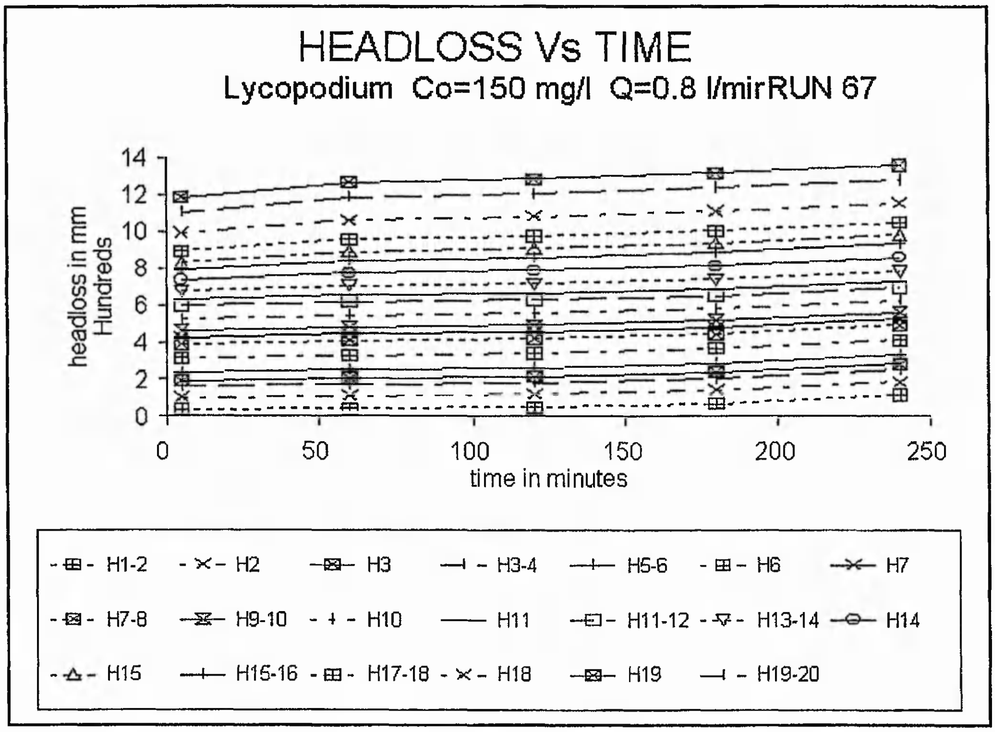


Fig 4.70 Head loss varying with depth and time for lycopodium powder with C_0=150 mg/l and Q= 0.8 l/min.

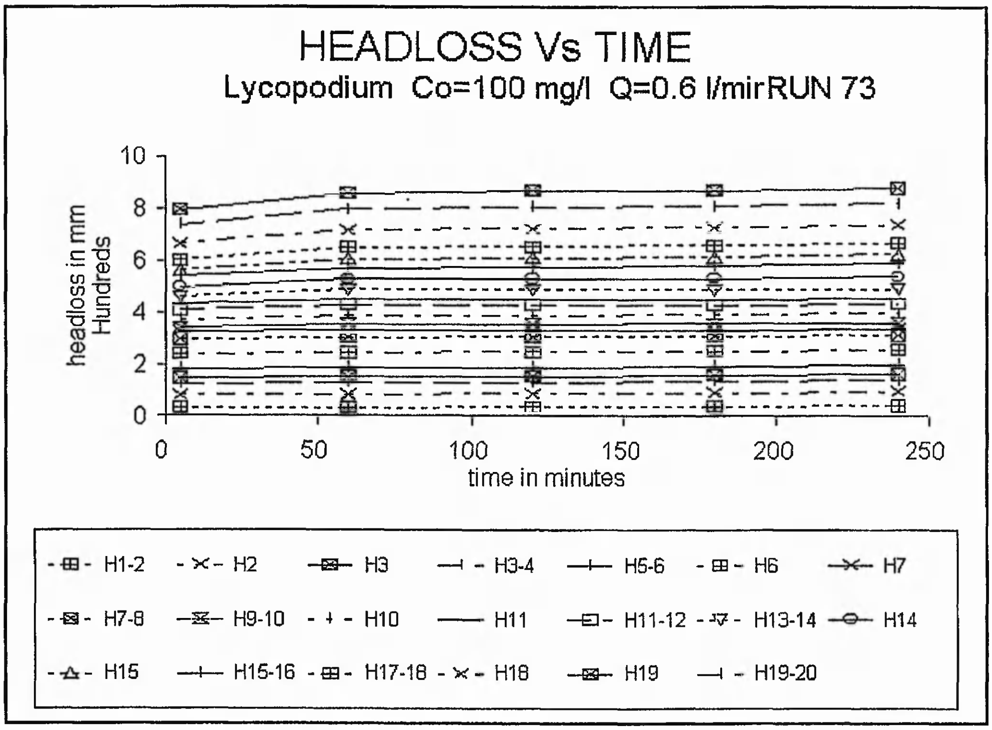


Fig 4.71 Head loss varying with depth and time for lycopodium powder with C_0=100 mg/l and Q= 0.6 l/min.

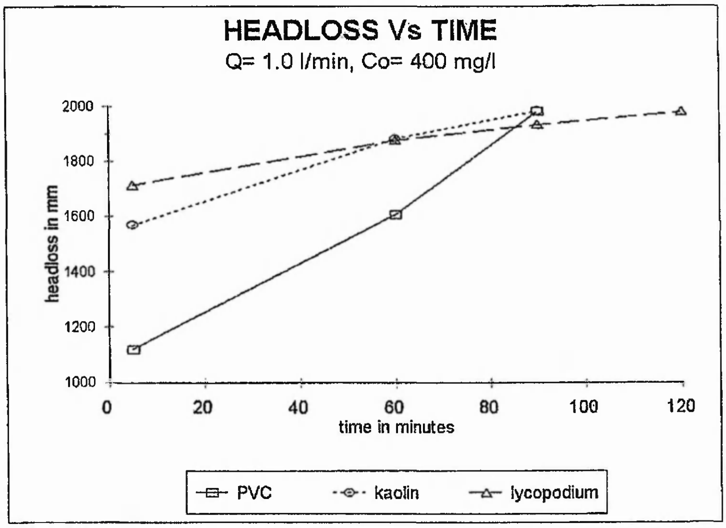


Fig 4.72 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 1.0 l/min and initial concentration 400 mg/l.

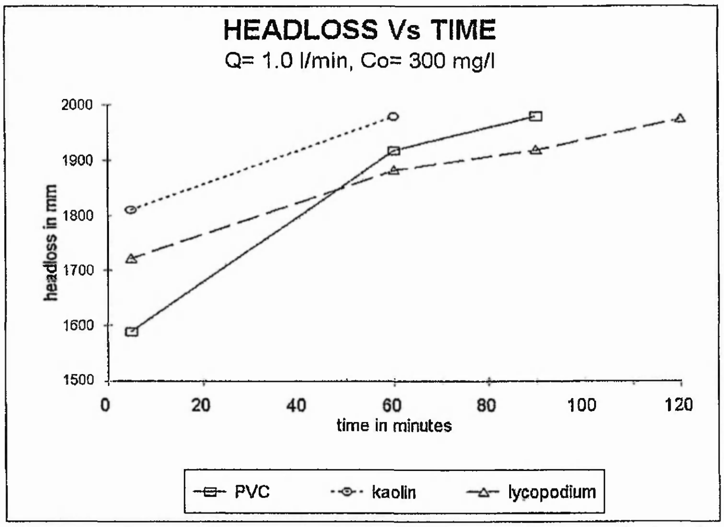


Fig 4.73 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 1.0 l/min and initial concentration 300 mg/l.

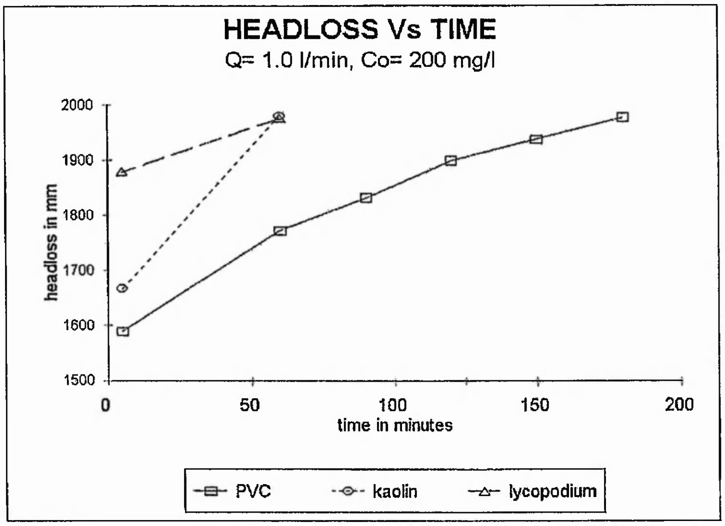


Fig 4.74 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 1.0 l/min and initial concentration 200 mg/l.

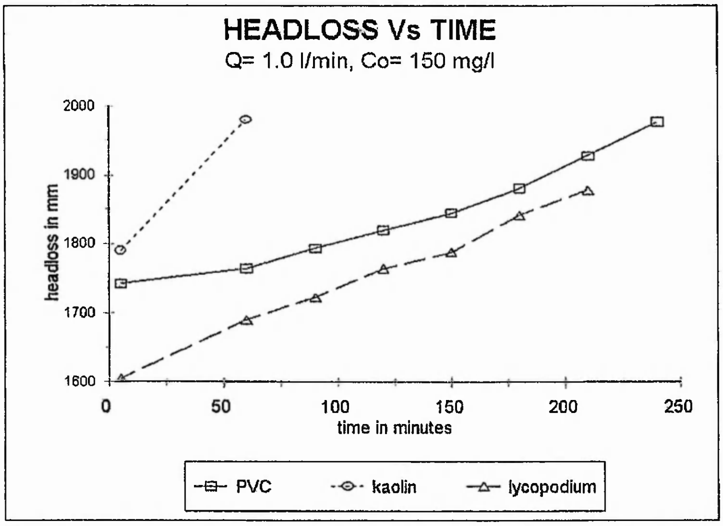


Fig 4.75 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 1.0 l/min and initial concentration 150 mg/l.

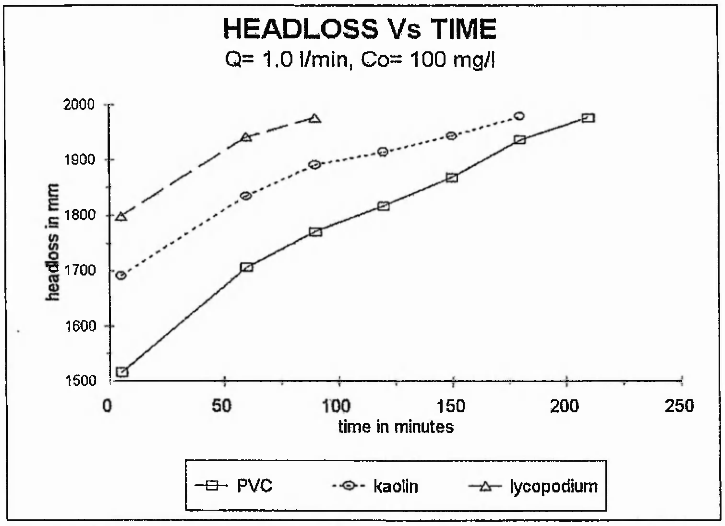


Fig 4.76 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 1.0 l/min and initial concentration 100 mg/l.

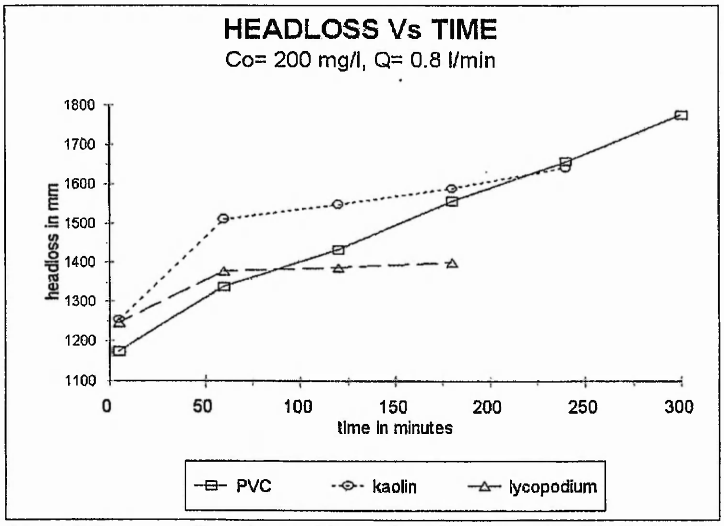


Fig 4.77 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 0.8 l/min and initial concentration 200 mg/l.

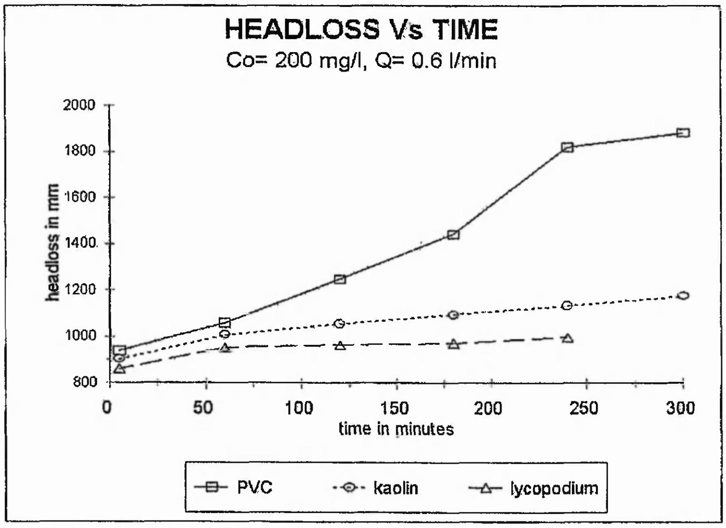


Fig 4.78 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 0.6 l/min and initial concentration 200 mg/l.

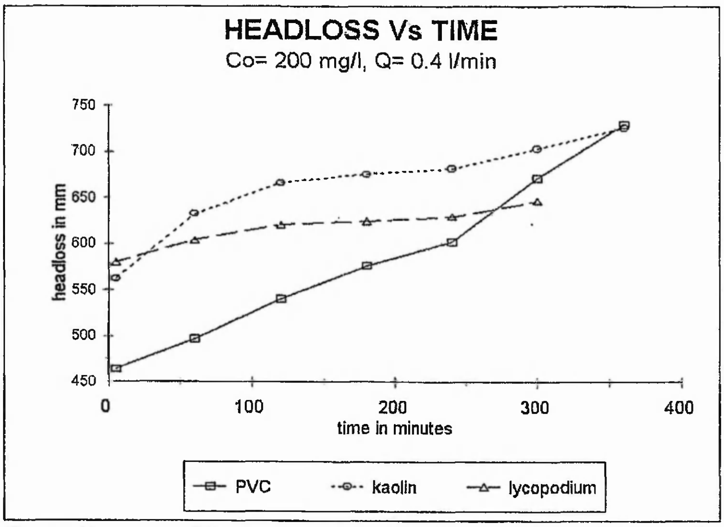


Fig 4.79 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 0.4 l/min and initial concentration 200 mg/l.

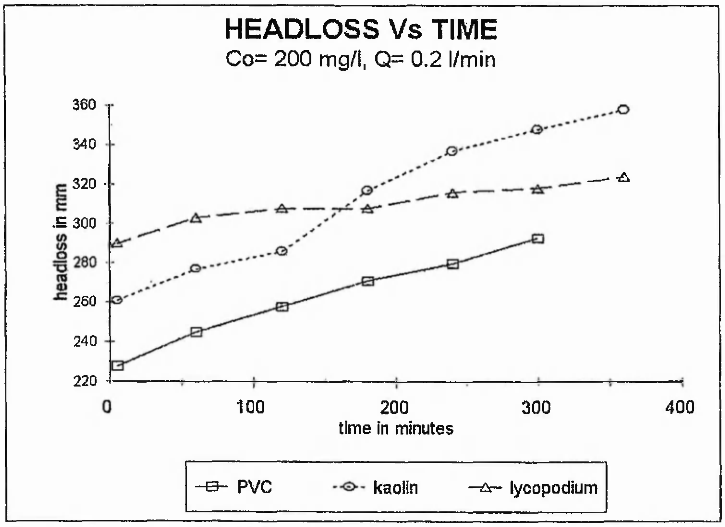


Fig 4.80 Total head loss development for PVC, kaolin and lycopodium, at a filtration rate of 0.2 l/min and initial concentration 200 mg/l.

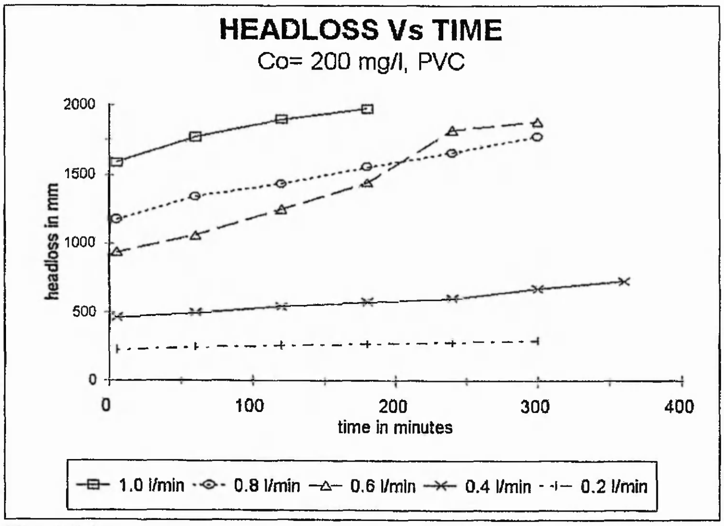


Fig 4.81 Effect of increasing filtration rate on total head loss for PVC; initial concentration 200 mg/l.

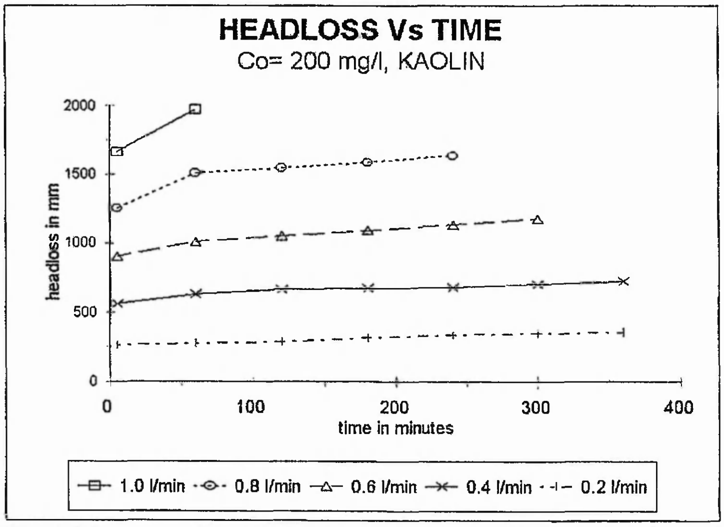


Fig 4.82 Effect of increasing filtration rate on total head loss for kaolin; initial concentration 200 mg/l.

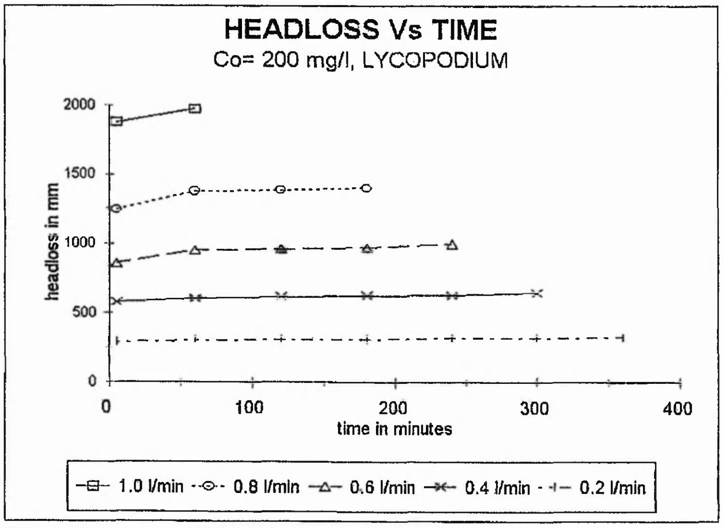


Fig 4.83 Effect of increasing filtration rate on total head loss for lycopodium powder; initial concentration 200 mg/l.

4.5 THE NATURE OF SYNTHETIC TURBIDITY AND THE REMOVAL EFFICIENCY

As described in previous sections, and in order to show variations in fine suspension particles removal patterns for PVC, kaolin and lycopodium powder under various initial conditions, ratios of effluent/influent concentrations were calculated and plotted against elapsed filtration time and depth.

Figures 4.1—4.15 show changes of the filtrate concentration with time for the three types of synthetic turbidities tested in this investigation when the model filter apparatus was run under various inlet concentrations and filtration rates. The experimental data was plotted on a semi-log scale to highlight the differences observed between the different initial concentrations used in the filter runs.

These curves show clearly that for PVC suspension (Figures 4.1—4.5), a decrease in the inlet concentration leads to a proportionally greater increase in the filtrate quality under the same varying filtration rates, when compared with kaolin or lycopodium turbidities. So, for example, with an inlet concentration of 400 mg/l the overall filtrate quality is about 10 mg/l at 0.6 l/min, whereas for 100 mg/l the overall filtrate quality is of the order of 0.2 mg/l, that is, an overall 96 per cent reduction in the final effluent concentration of suspended matter from the highest to the lowest initial concentration. This pattern was common for all runs.

Differences in the concentration curves versus time for both kaolin and lycopodium turbidities (Figures 4.6—4.15) are not so pronounced; these curves indicate a general tendency (as more time is elapsed) to shade off and merge with one another in spite of the fact that the differences in the initial concentration values were considerably high (from 400 mg/l to 100 mg/l). For example, with the kaolin and lycopodium suspensions the overall reduction in the final effluent concentration of suspended particles when the initial concentration varies from 400 to 100 mg/l at a filtration rate of 0.6 l/min is only 57 and 40 per cent respectively.

As expected, these curves show that the variation of the overall concentration ratio versus time is of a nonlinear nature, since the filtrate quality keeps improving with time before reaching the filter's breakthrough point towards the end of the filter run. The graphs depicted in this work are based on experimental data collected over a filtration time of only 2 to 6 hours, therefore filter breakthrough did not occur during the course of this investigation, and only the improvement period could be represented.

In order to compare the overall removal efficiency of the system for PVC, kaolin and lycopodium powder, the concentration ratio changes were plotted on a semi-log scale against time for every type of suspension under identical initial conditions. The results are shown in Figures 4.16—4.40. From these figures it can be seen that the maximum overall removal efficiencies were achieved with the kaolin and lycopodium turbidities in comparison to PVC powder, as described in section 4.2, especially for the higher initial concentrations, although there was a tendency for these differences to become less apparent with a decrease in the inlet concentration and rate of flow. This merging phenomenon is especially pronounced for the lowest inlet concentration (100 mg/l) at any flow rate. Thus, for example, the filtrate quality for PVC, kaolin and lycopodium powder varies from about 10.5, 1.1 and 0.5 mg/l, respectively, at Q= 1.0 l/min and C_0= 400 mg/l (Figure 4.16) to values of the order of 0.8—0.4 mg/l for the three turbidity materials at Q= 0.2 l/min and C_0= 100 mg/l (Figure 4.40).

After a general examination of the curves shown on the above-mentioned figures it is evident that the tests performed at higher rates of flow gave the worst overall efficiency results of removing PVC turbidity, whereas the kaolin and lycopodium suspensions offered the best results at any one flow rate. In addition to this, increasing the inlet concentration from 100 to 400 mg/l showed no significant improvement over the filter's removal efficiency for kaolin and lycopodium turbidities (varying between a high of 2.0 mg/l and a low of 0.2 mg/l in the final filtrate concentration) at any given filtration rate, as shown in Figures 4.16—4.40.

On the other hand, PVC turbidity showed the highest initial removal efficiency (i.e. effluent from compartment number 1, where the concentration ratio varies from about 0.2 to less than 0.1, slowly approaching to zero), whereas lycopodium, and particularly kaolin, produced the lowest initial efficiencies of removing suspended solids, with concentration ratio values between 0.5 and 0.2 for kaolin and 0.3 and less than 0.1 for lycopodium, with no significant improvement over time, as it can be seen from Figures 4.41 to 4.52. This could be attributed to the more uniform particle size range of PVC (0.5—1.5 μm) as well as to the fineness of its particles, a fact that gives the particles the largest specific surface area to react with the filtering media.

Analogously, these figures also show that, due to the greater particle size range and particle shape variation of kaolin, turbidity removal spread more evenly across the various filter compartments, particularly compartments 1, 2 and 3. This trend was also noticed for lycopodium, even though not as markedly as with kaolin.

Figures 4.53—4.58 show typical concentration ratio changes with depth and time. Again, kaolin and lycopodium turbidities present similar trends: while the filtrate quality for PVC kept improving as time elapsed, both kaolin and lycopodium turbidities showed no significant variations over time, and even a general tendency towards a drop in the filtrate quality was detected as the filter run progressed (especially noticeable for lycopodium turbidity), as Figures 4.45—4.52 seem to suggest. Due to kaolin's relatively high density (2,600 kg/m^3), the general trend was for particles to deposit at the inlet surface of the filter's first compartment, thus confirming that the predominant removal mechanism for kaolin suspension is gravity.

However, no significant head losses due to surface deposition were observed, as it can be deduced from the shape of representative head loss curves shown in Figures 4.64—4.67. This phenomenon can be explained if the particle size range and shape are considered.

Kaolin particles are irregular in shape and present a wide variation in size, 2 to 53 μm, although with a clear prevalence of the smallest sizes over the largest (2 μm

80%, and up to 53 μm, 20%). Thus, the larger fraction was retained mainly at the coarser media, and the finer particles were trapped by the following compartments. The particle shape irregularity also added to the greater distribution extent achieved within the various filter compartments. In other words, the kaolin suspension was made up of particles of various sizes and shapes, a fact that together with the media gradation from coarser to finer, undoubtedly helped to increase the particles' chances to adhere to the filter media and make a more efficient use of the filter's storage capacity.

On the other hand, the particles of lycopodium powder are comparatively coarser with a uniform size range (35 μm) and shape, and present the lowest density (1,180 kg/m^3). Due to the relative large size of the lycopodium particles, surface deposition at the inlet face of the first compartment was observed. This resulted in shorter filter runs with great head losses caused by the progressive build-up of a surface mat, especially when operating at high rates of flow and inlet concentrations (Figure 4.68). The light nature of the lycopodium particles also help to explain the high variability (scattered results) in filtrate quality and removal efficiency recorded in the lower compartments, especially in up-flow filtration (Figures 4.49 and 4.50, 4.59a to 4.59e, Figure 4.59h, and Figures contained in Appendix 1).

Any particle in a streamline is frequently subjected to two different velocity vectors:

- the fluid velocity vector
- the gravitational velocity vector (settling velocity)

In the down-flow direction, the resultant velocity vector is the sum of the fluid velocity vector and the gravitational velocity vector. While in up-flow filtration, the net velocity vector is the difference between the flow velocity vector and the particle's gravitational velocity vector.

The lower removal efficiency for lycopodium turbidity in up-flow filtration was believed to be a consequence of its lower density, which resulted in a significantly lower settling velocity. Thus, the lycopodium particles were transported across the filter bed by a comparatively higher net up-flow velocity (interstitial velocity), which resulted in a lower probability of turbidity removal by sedimentation or gravitational effect.

It is interesting to note that there was a nearly equal percentage removal of particles per compartment in the lower size range, that is, PVC powder. As the suspended particle size range increased, the percentage removal per compartment also increased, as it is evident from the graphs shown in Figures 4.41, 4.47 and 4.49.

Figure 4.84 shows the typical turbidity removal and head loss increase patterns developed during this investigation, at the outlet of compartments number 1 (down-flow) and number 2 (up-flow), for PVC, kaolin and lycopodium powder. The initial concentration was 300 mg/l and the rate of flow 0.6 l/min.

Removal of turbidity for PVC increased as the time of filtration progressed, that is, the filtrate quality improved as time elapsed; for kaolin the turbidity removal increased to an apparent maximum and remained generally constant for the remainder of the run, whereas for lycopodium powder removals were greatest at the beginning of the filtration period and then decreased slowly with time.

As described in section 4.4, head losses were initially higher for both kaolin and lycopodium, when compared with that for PVC. However, the head loss increase for PVC was considerably higher than that for kaolin or lycopodium, as a result of the fineness of the PVC particles that can fill the pore spaces and produce greater head loss due to frictional losses and more surface area per unit volume of deposit for resistance to flow. Both PVC and lycopodium showed a greater tendency towards surface mat formation, mainly due to their lower particle size range.

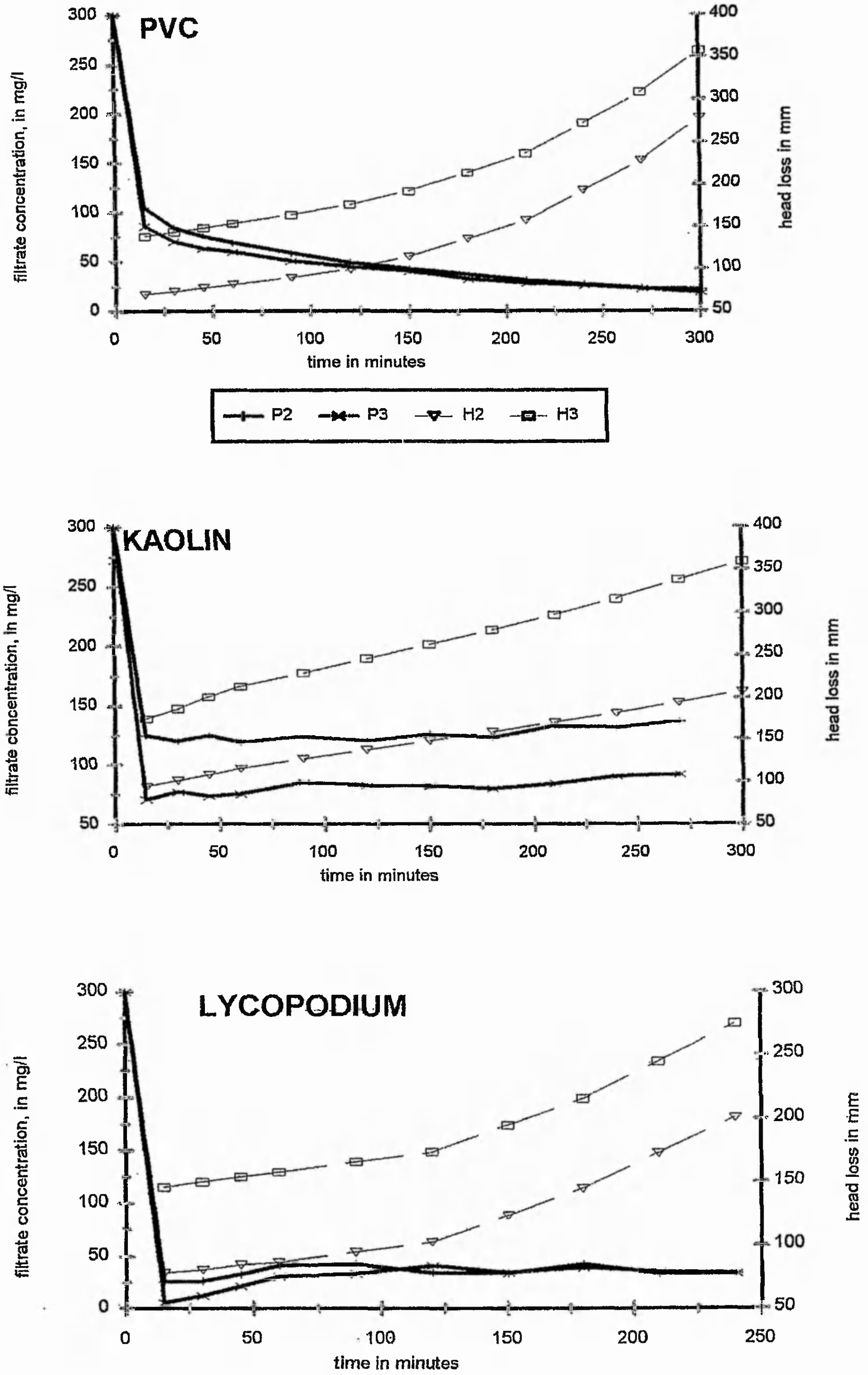


Fig 4.84 Filtrate concentration and head loss changes with time for PVC, kaolin and lycopodium, for C_0= 300 mg/l and Q= 0.6 l/min.

4.6 EFFECT OF THE RATE OF FLOW

After examining all the optimum values of flow rate to produce the optimum removal of turbidity by rapid filtration, as shown in Table 4.2 to Table 4.6, filtration efficiency was found to increase with a decrease in filter flow rate in accordance with theory; an increase in filtration rate caused a deterioration in filter water quality and at the same time caused a proportionally greater reduction in filter run length (especially valid for the lower particle size range, i.e. PVC). Lower flow rates cause less shear and allow more and smaller particles to settle in the filter.

The experimental results clearly indicate that for obtaining a filtrate of low turbidity, the most effective measure appears to be the reduction of the velocity of flow.

As an example, it was found that the data obtained at a rate of flow of 0.4 and 0.2 l/min prevailed over the others achieving a high filtration efficiency in excess of 99 per cent efficiency in 28 runs out of 30 for the three types of synthetic turbidities. Still, the overall performance of the filter model showed a high filtration efficiency in removing suspension particles for all filter runs: in excess of 99 per cent efficiency in 64 tests out of 75.

The higher rate of flow gave shorter filter runs, because of the increased penetration of solids, leading to deterioration of the filtrate and because higher head losses occurred, as a result of the increased velocity (Figures 4.81 to 4.83). Thus, for example, Figure 4.83 shows that the total filtration time could be as much as six times longer when operating at the lowest rate of flow (0.2 l/min) and still not surpassing the head loss value of 250 mm (over a total head loss limit of 2,000 mm). This was a common trend in the filtration of PVC, kaolin and lycopodium turbidities, for any given inlet concentration.

Where the suspended solids formed at the inlet surface, as in some tests with PVC and lycopodium, the higher rate produced a more rapid increase in straining efficiency but again resulted in an even shorter filter run because of higher head losses (Figures 4.1 and 4.41).

As reported in Section 4.5, the filter apparatus showed relatively little differences in the removal efficiencies of both kaolin and lycopodium turbidities when the rate of filtration was considered.

When the input to the filter was PVC turbidity, significantly higher filtrate qualities were achieved with a decrease in the rate of flow, when compared to those for kaolin and lycopodium (Figures 4.16—4.40). Again, the reason for this may be attributed to the different physicochemical properties of the synthetic turbidities used in this investigation, as explained in more detail in the previous sections.

The *approach velocity* of filtration, v_a, can be defined as follows:

$$v_a = \frac{Q}{A} \qquad [4.1]$$

where $Q =$ volumetric flow rate, m^3/s
$A =$ plan area of filter, m^2

The approach velocity was constant at either 2, 4, 6, 8 or 10 m/h for any one filter run. These values are in the range of normal waterworks filtration rates.

The above figures were not, however, representative of the mean flow rate in the pores, which was higher due to the restricted open area between the grains. This is referred to as the *interstitial velocity* v_i:

$$v_i = \frac{v_a}{\epsilon} \qquad [4.2]$$

where v_a = approach velocity
ϵ = porosity

As the porosity is changing, in depth and in time, due to the accumulating deposits, the interstitial velocity v_i was not used due to the difficulty to obtain valid results. The approach velocity, is more simply defined and is linearly related to v_i, as shown in Table 4.7. In addition, v_i is only a mean value as it varies from pore to pore due to the packed grain geometry; also the velocity varies across a pore due to the laminar flow velocity distribution (Ives, 1987).

Table 4.7 Interstitial and approach velocities per compartment

Compartment Number	Rates of Flow in l/min									
	0.2		0.4		0.6		0.8		1.0	
	v_a*	v_i*	v_a	v_i	v_a	v_i	v_a	v_i	v_a	v_i
1	2.03	5.34	4.07	10.7	6.10	16.1	8.12	21.4	10.16	26.7
2	2.02	5.77	4.05	11.6	6.07	17.3	8.09	23.1	10.11	28.9
3	2.04	6.38	4.08	12.8	6.11	19.1	8.15	25.5	10.19	31.8
4	1.99	5.85	3.98	11.7	5.98	17.6	7.97	23.4	9.96	29.3
5	1.99	5.38	3.98	10.8	5.97	16.1	7.97	21.5	9.96	26.9
6	2.00	5.56	4.00	11.1	6.00	16.7	8.00	22.2	10.00	27.8
7	1.99	5.24	3.97	10.4	5.96	15.7	7.95	20.9	9.93	26.1
8	2.00	5.41	4.01	10.8	6.01	16.2	8.01	21.6	10.01	27.1
9	2.03	4.61	4.06	9.23	6.09	13.8	8.13	18.5	10.16	23.1
10	2.08	4.52	4.15	9.02	6.23	13.5	8.31	18.1	10.38	22.6

(*) Initial interstitial velocity, V_i, calculated from the porosity values presented in Table 4.1.
Velocity values are expressed in m/h.

4.7 EFFECT OF THE TEMPERATURE

As explained in section 3.7, the viscosity of water is strongly dependent on temperature; the viscosity increases approximately by 60 per cent when water temperature drops from 20°C to 0°C (Figure 3.13). Viscosity changes are important in filtration, whereas the relatively small density changes over the same temperature range have little effect.

Laboratory and pilot-scale studies under controlled conditions have evaluated the effects of viscosity on filter performance (Cleasby *et al.*, 1963; Ives and Sholji, 1965; Hsiung and Cleasby, 1968). Ives and Sholji (1965) determined proved theoretically and experimentally demonstrated that the filter coefficient λ was inversely proportional to the filtration rate, grain size and the square of the viscosity (Equation 2.3). At higher temperatures viscosity is less, particles settle faster and filtration is enhanced.

However, the laboratory where the filtration tests were conducted was air-conditioned to a constant 20°C in winter. In order to minimize the temperature effect, the mixing tanks were filled the night before. The water contained in these tanks reached ambient temperature and the seasonal fluctuations in water temperature were kept to a minimum.

Despite all the measure taken, there were a few days when the heating of the laboratory did not function properly and a difference in water temperature was observed between runs. The experiments performed under those conditions illustrate the effect of temperature on the removal efficiency. Figures 4.6—4.7 show that a difference of 4°C in the water temperature was determinant in the odd results obtained when filtering kaolin suspension, when better filtrate quality was achieved at higher inlet concentration. Sholji (1963) proved for the first time the effect of the water temperature on the filtrate concentration ratio versus time for constant initial conditions (within a range of 3-33°C).

CHAPTER 5

DISCUSSION

5.1 GENERAL

The objective of this experimental work was to determine the viability of rapid filtration by a multi-compartment up-flow and down-flow process. Although rapid sand filtration has been studied extensively for more than 60 years, the present research is the first in which up-flow and down-flow have been used in conjunction with five different sand grain sizes on a finer to coarser basis.

In addition, five initial concentrations were examined: 400, 300, 200, 150 and 100 mg/l, at filtration rates of 1.0, 0.8, 0.6, 0.4 and 0.2 l/min, using the model filter apparatus as described in Chapter 3.

As previously indicated and in order to establish performance characteristics, laboratory studies were carried out using a model of effluent of a known inlet concentration of particles consisting of polyvinyl chloride, kaolin or lycopodium suspensions. These turbidity materials were selected for their specific physicochemical characteristics, namely particle size, size range, shape and density, which provide a varied range of properties, within the scope of this investigation, as described in sections 3.3 and 4.5. The use of such a model water allowed for direct comparisons between experimental runs without variations in raw water quality that may be encountered when using a natural water.

Initially, thirty-five filtration tests were needed to develop and evaluate the apparatus and the experimental technique, and eventually modify the system to ensure proper functioning.

Seventy-five filter runs were conducted to establish the efficacy of the process and provide comparative data to that obtained by conventional rapid filtration.

Analysis of the results obtained during the experimental course of this investigation tend to confirm what intuition suggests and modern theories of filtration predict, namely that in a deep bed of unisize filter material, if all other parameters remain unaltered, filtration improves if the flow rate is decreased, if the size of the suspended particles increases, and if the grain size of the media is decreased.

Some filtration tests demonstrated a very high efficiency of particle removal, therefore final filtrate turbidities were too low to permit accurate concentration-ratio determinations. Consequently, concentration changes with time and depth were systematically expressed in mg/l or plotted logarithmically with respect to time and depth in order to emphasize the differences. Some of these graphs can be found in Appendix 1.

However, the testing procedures described and their results do not allow precise deductions regarding an optimal filter concept. Based on mathematical models, and in a future research, selected pilot tests may yield the data for designing a system.

Four factors must be taken into account in practice (Ives, 1969):

- granulometry
- sand bed depth
- filtration rate
- filter running time

Additionally, the nature and concentration of suspended particles in the raw water which constitute the input to the filter, are also variables one must take into account.

5.2 FILTER MEDIA CHARACTERISTICS

Rapid filtration by a multi-compartment up-flow and down-flow process has been shown to be highly effective at reducing turbidity levels at filtration rates less than 1.0 l/min (≈10 m^3/m^2h). The overall removal efficiency of the system has been systematically above the 96 per cent mark, as reported in Chapter 4.

The advantage of such a compartmented filter box is that the coarse granulometries at the inflow arrest the coarse particles while the fine grains at the outflow give the water the requisite quality. In the coarse medium the probability of a particle being held in any plane is reduced, and the suspended particles therefore penetrate deeper into the bed. Provided the grain size is appropriate, the first two compartments remove a fair proportion of the suspended matter without a high pressure loss, and the finest medium contained in the last compartment deals only with a very small concentration of suspended matter. If this were the case, then the filtration rate could be increased, without loss of quality.

The disadvantage of using different granulometries can be the adverse influence of the transitional zones from one compartment to the other on the removal capacity of the system. If these *breaks* in the filtration process affect filtration significantly, then the overall removal efficiency is influenced also.

Regarding the use of multilayer filters, practical investigations have shown that the length of the transitional zone should not exceed about 20% of the layer thickness, otherwise the benefit of separate layer filtration is lost (Ives, 1979). However, this space is required in order to obtain a proper backwash. Drawing a parallel between multilayer and multi-compartment filters, this could add to the reasons for the comparatively lower filtration efficiency obtained in up-flow filtration. Future research on the effect of this transitional zone will help to answer this question.

The only real difficulty encountered during filtration was caused by the difference in size of the media used in the various compartments, i.e. their relative expansion originated by the incipient fluidisation observed during up-flow filtration in those compartments containing the finest sand, as reported in section 4.3, and fully discussed in section 5.3.

As explained in section 1.4.6, the point of incipient fluidisation, or minimum fluidising velocity v_{mf}, is the superficial fluid velocity required for the onset of fluidisation. The bed is completely fluidised when the friction drag, or pressure drop across the bed, is just enough to support the weight of the filter media. Mathematically, this relationship is given by equation [1.18], on page 60.

Sholji (1987) showed that v_{mf} depends upon the initial concentration of fluidised grains, C_0, or initial porosity, ϵ_0, and on the temperature of the fluidised water, which all influence the settling velocity, v_s.

Knowing the mean settling velocity of individual grain size, v_s, at a particular temperature, the minimum fluidising velocity is easily evaluated under the same temperature conditions (Sholji, 1987), provided coefficients K_2 (dynamic shape coefficient) and K_3 (variable experimental coefficient) are determined experimentally for each type of unisize medium (see equation [1.26] on page 63). The values for K_2 for the sand used in Sholji's investigation ranged from 2.000 to 2.230, and K_3 varied from -0.335 to -0.455. These negative values refer to the flaky shape of the uniform grains used in the fluidised bed.

Thus, the minimum fluidising velocity can be found by drawing a line asymptotic to the x-axis, or $C_0 / [1- (v_b / v_s)]^{3/2}$ in equation [1.26], when the expansion ratio L_e / L_0 in the y-axis is equal to 1.0, where v_b is the face velocity of the backwash water (backwash flow divided by the total filter area, in m/s), v_s is the particle settling velocity (m/s) and C_0 is the initial volumetric concentration of grains forming the filter bed or $(1-\epsilon_0)$, where ϵ_0 is the initial porosity of the bed.

Another way of calculating v_{mf} is by using the following equation (Wen and Yu, 1966):

$$v_{mf} = \frac{\mu}{\rho_w d_{eq}}(33.7^2 + 0.0408Ga)^{0.5} - 33.7\frac{\mu}{\rho_w d_{eq}} \qquad [5.1]$$

where *Ga* is the *Galileo number*.

$$Ga = d_{eq}^{\ 3}\,\frac{\rho_w(\rho_s - \rho_w)g}{\mu^2} \qquad [5.2]$$

where d_{eq}= diameter equivalent, m
ρ_w= mass density of water, kg/m^3
ρ_s= mass density of sand, kg/m^3
g= acceleration due to gravity, m/s^2
μ= water dynamic viscosity, kg/m·s

The theoretical minimum fluidising velocity values reveal that incipient fluidisation did occur in compartments 8 and 10 containing the finer sand in the up-flow mode, since these figures are close enough to those of the actual filtration rates, as observed experimentally and shown in Table 5.1

Table 5.1 Minimum Fluidising Values

Comp No.	BSS Sieve Openings (µm)	Filtration Rate (l/min)	ρ_w (Kg/m^3) at 16°C	ρ_s (Kg/m^3)	µ (Kg/m·s) at 16°C	Minimum Fluidising Velocity (m/s)		Minimum Fluidising Flow Rate (l/min)	
						Wen & Yu	Sholji	Wen & Yu	Sholji
8	710/600	1.0	998.97	2650	0.001	0.00402	0.0046	1.5	1.8
10	600/500					0.00288	0.0042	1.1	1.6

Minimum fluidising values for filter compartments 8 and 10 containing the finest granular material (after Wen and Yu (1966) and Sholji (1987) equations,), with initial porosity values of 0.37 and 0.46 respectively.

The statements by Mints (1966) and Ives (1969) describe the optimisation of a uniform filter by comparing the head loss and filtrate quality as functions of the filter depth. With a given granulometry and rate of flow the optimal filter depth is obtained when the cleaning capacity of the filter is exhausted upon reaching the hydraulic capacity of the bed (Ives, 1979). The length of any filter run was restricted by both the amount of turbid water available in the mixing tanks and the rapid increase in the total head loss. Many experiments during this investigation had to be terminated once the head loss limit had been reached (about 2000 mm). The coarser medium in compartment number 1 gave the effect of a surface filter by taking the burden of removing larger suspended colloidal matter from the flow, which resulted in a steep increase in the head loss of the system, while the final compartments showed comparatively little effect on the overall filtration efficiency.

Thus, the filter running times achieved in these experiments ranged from a mere 120 minutes, in the case of an excessive head loss (for the highest inlet concentrations and rates of flow), to up to 420 minutes, as a consequence of having depleted the volume of turbid water available in the mixing tanks. Too short a running time could be also the result of an unsuitable media combination.

Research by Ives *et al.* in the 50's and 60's demonstrated mathematically and in practice that the smaller the grain, the larger the total surface area of the filtering medium; also that the lower the sphericity (i.e. more angular grains), the larger the surface area. It follows then that finer sand gives a finer filtration because the size of the channels is decreased, and because far more area of medium is available.

However, recent thinking has suggested that a larger medium size provides for deeper penetration of suspended solids into the filter, using more of the medium and giving longer filter runs. In addition to this, the thickness of the filter media is inversely related to the quality of filtered water: a thicker bed produces better quality water than a thin one (Hudson, 1963, among others), due to an increase of the available area of filter medium.

As described in section 1.4.3, the theoretical head loss of a clean filter obeys a formula in which the h/L varies as v_s^2/d_g, where h is the head loss, L the depth of bed, v_s the flow velocity and d_g the geometric mean diameter of the grains. We did not concentrate on this issue during this investigation: It is a non-linear relationship which is more complex than it seems at first sight.

A larger media size results in lower pore velocities, at equivalent filtration rates, and hence lower shear forces enhancing particle to media attachment. The lower pore velocities result in reduced head loss effects and it has been reported that deeper bed penetration occurs, approaching what may be considered true depth filtration (Adin and Rebhun, 1974; Moran *et al.*, 1993). Coarse media will therefore permit deeper beds and higher rates to be used. In addition to this, studies by Baumann and Huang (1975) demonstrated that high rate, deep bed filters containing coarse grain sizes are more economic, for uniform filters.

Thus, increasing bed media size and depth are considered means by which performance at higher filtration rates and initial turbidities may be improved. The use of polyelectrolyte within such a treatment system is considered to offer higher removal efficiencies (Adin and Rebhun, 1974; O'Melia, 1974).

As shown in Figures 4.53—4.58, a significant proportion of the filter is never used efficiently. It is therefore further deduced that in order to increase the overall efficiency of the filter, the final compartments of the filter must be persuaded to remove a greater proportion of the suspended solids reaching them than the first compartments. The previous analysis indicates that this could be achieved by increasing the cross-section area of the filter bed in the direction of flow so that the velocity progressively decreases with increasing depth, or by increasing the size of particles as they travel through the bed by encouraging them to flocculate (Treanor, 1971).

Gaps in the filter media (i.e. pockets due to arching of the grains) were observed in an unconsolidated bed following backwashing. These gaps were significantly reduced but not completely eliminated after bed consolidation by vibration. Similar observations were made by Ives (1986), and with the aid of fibre optic endoscopes he concluded that the influence of these gaps is to produce a greater local flow, which certainly exert an influence on the filtration efficiency.

There is experimental evidence (Gimbel, 1983) that surface roughness and the hydrodynamic processes inside the filter bed have a marked effect on particle adhesion. The latter is demonstrated, for example, by the extremely low deposit capacity of filter beds consisting of relatively smooth materials such as glass spheres. On the grounds of these results, the move to a more angular filter medium could be considered in order to achieve a better removal efficiency. This is also confirmed by Ives (1988). However, grains should not be too flaky (not less than 0.6 by sphericity testing) in order to improve filter's performance.

5.3 UP-FLOW AND DOWN-FLOW FILTRATION

Comparing the directions of flow, up-flow filtration showed lower removal efficiency than the corresponding down-flow compartments, as it is obvious from the typical shape of the concentration ratio curve with depth and time (Figures 4.53—4.58 and 4.59). This stepped-curve pattern or serrated outlook of the concentration ratio curves with depth appears to be characteristic of this novel filtration apparatus as a direct consequence of the alternating use of down-flow and up-flow in the filtration process, as mentioned in section 4.3.

From the author's experimental results and previous investigations, it is obvious that the effect of gravity is important in filtration because the removal efficiency is decreased in the up-flow compartments, that is, the chances of a particle to be removed by sedimentation are lower.

Straining action was evident in downward filtration, whereas this was not present in up-flow. Surface deposition was more likely to occur in the higher rate of flow and inlet concentrations, with larger suspended solids. Again, this comes to show that in filtration the direction of flow relative to a free surface, and the particle size and density of suspended solids are important factors to take into account. These findings are consistent with the observations of Diaper and Ives (1965).

However, it is possible that the lower removal efficiency in up-flow may have been at least partly induced by a local sampling-related scour of deposits, as it has been reported in section 4.3. Also, some signs of initial fluidisation of the layers were observed in compartments 8 and 10, when operating at high rates of flow (*see Section 5.2*).

Previous investigations showed that an important step in liberating the suspended solids from the depth of the bed is the initial fluidisation of the layers, which effects a physical break-up of the deposit (Hulbert and Herring, 1929; Diaper and Ives 1965; and Cleasby and Woods, 1975). This could have contributed to the general lower removal efficiency in up-flow filtration, as it seems plausible to assume that a slight detachment of particles had occurred while taking filtrate samples, which may explain the negative removal efficiency obtained during the course of certain experiments in the up-flow compartments, particularly after compartment number 5 (Figures 4.59a—4.59h).

This incipient fluidisation of the filter bed in compartment number 10 (and to some extent in compartment number 8) is undesirable as it produces a final filtrate of lower quality than otherwise expected. Consequently, in practical design and operation it should be eliminated or minimised.

There are three different, and viable, solutions to this problem. This could be achieved by either:

A.- eliminating the last compartment, this way the final filtrate would be collected after down-flow filtration (Figure 5.1)

B.- increasing the cross-section area of the filter bed contained in the last compartment, in the direction of up-flow, so that the velocity decreases with increasing area; this way the interstitial velocity will be below the minimum fluidising velocity of the filter medium in that compartment

C.- removing the partition in the last compartment, thus resulting in a bigger downward compartment filter surface area and reduced approach velocity altogether (Figure 5.2).

However, these signs of incipient fluidisation, or creeping of the water flow near the wall, may not be so important in practice, as the natural roughness of the material used for the construction of the filter walls may largely restrain this most unwanted effect (e.g. concrete). Additionally, fluidisation impact on filtrate quality can be avoided by changing the grain size. By increasing the grain size alone, lower pore velocities can be attained and thus avoid filter bed fluidisation completely. Even so, and due to deeper bed penetration taking place in coarser media, this solution would not be suitable for an influent containing a high proportion of fine particles, such as the PVC suspension used in these experiments.

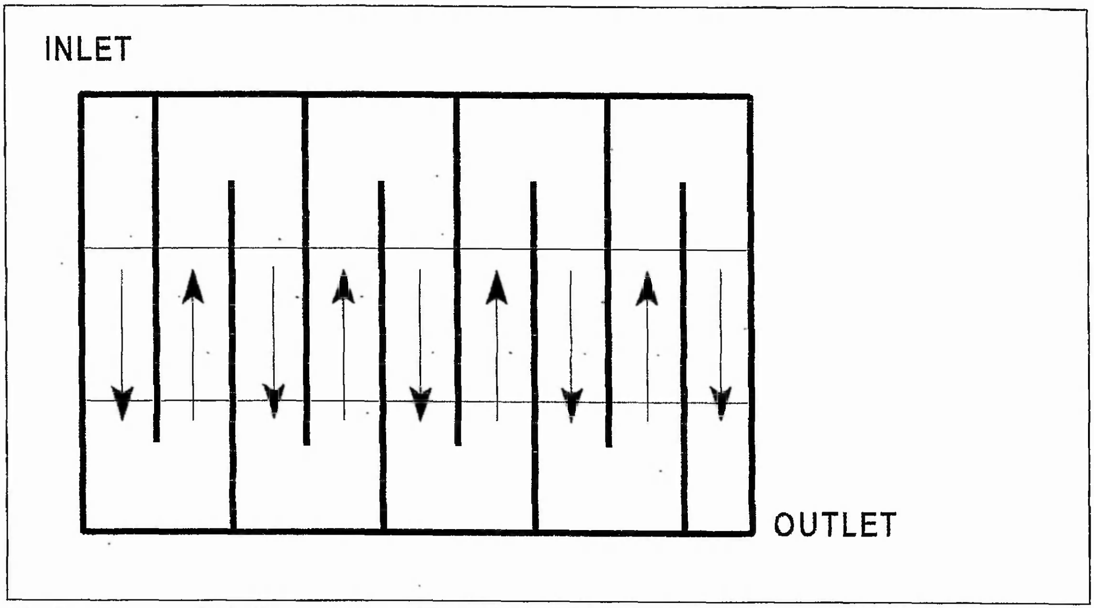


Fig 5.1 Solution A to avoid incipient filter bed fluidisation: elimination of the last compartment, resulting in a nine compartment filter and filtrate collecting point at the bottom of the apparatus.

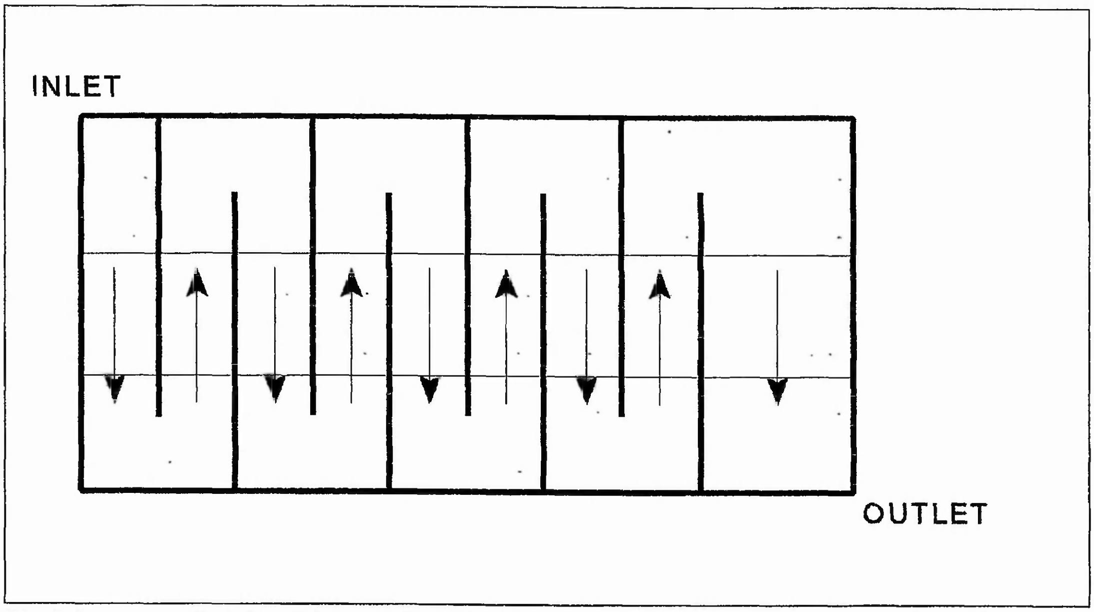


Fig 5.2 Solution C to avoid incipient filter bed fluidisation: suppression of partition in the last compartment, resulting in a nine compartment filter and filtrate collecting point at the bottom of the apparatus.

5.4 EFFECT OF THE PHYSICOCHEMICAL PROPERTIES OF THE SYNTHETIC TURBIDITIES

In wastewater suspensions, particle sizes range from well below 1µm to more than 1000 µm, whereas after sedimentation maximum diameters of usually 150 µm are reached. In settled domestic wastewater, the organic content can roughly be fractionated into 45-55 per cent dissolved components (< 0.001 µm), 20-35 per cent colloidal particulates (0.001—1 µm) and 25-35 per cent supra-colloidal particulates (1—100 µm) (Dégrémont, 1979).

The input to the filter was, however, a homogeneously dispersed suspension of different turbidity materials of known physicochemical properties. The idea behind was to test the filter's response at clarifying turbid water containing suspended solids of various shapes within a particle size and density range of 0.5—53 µm and 1,400—2,600 kg/m^3, respectively.

The process of filtration takes place in three different steps, *transport* to a grain surface, *attachment* to the surface and detachment from the surface, as described in section 1.4.2. Many filtration mechanisms making up these steps have been studied for over 30 years (Ives and Gregory, 1966; Ison and Ives, 1969; Tchobanoglous, 1970;Yao *et al*,1971; Habibian and O'Melia, 1975;Kawamura,1975).

As explained in section 1.4.2, modern theory of filtration assumes that initially, when the filter medium is clean, every layer of the filter is equally efficient at removing particles from suspension and that in every layer the suspension entering it and leaving it is uniformly dispersed. Another basic principle is the mass balance for the suspension particles. It is simply a statement that particles removed from suspension are deposited in the filter pores.

It follows that the retained particles will progressively clog the pore spaces between the grains and either the flow rate will diminish, or increasing pressure difference will be needed to maintain the flow rate. In this experimental work, constant flow rate was maintained and the head loss started from an initial low value (due to the flow resistance of the clean granular material) and increased to an upper limit of around 2000 mm (set by the fluid mechanics of the filter system).

As filtration proceeds and as the pores become restricted by deposited solids, the removal efficiency decreases due to the progressive increase in interstitial velocity. Coad and Ives (1981) demonstrated experimentally (conductance technique) that the interstitial velocities are less uniform as the filter pores become progressively clogged.

As expected, particle size, density and shape have been shown to be of great significance to filter performance. Ives and Gregory (1967) have suggested that different mechanisms of transport are responsible for removing particles of different sizes. According to previous investigations, an increase in particle size will improve filtration. This is true for both kaolin and lycopodium suspensions, as it is evident from the experimental data represented in Figures 4.16 — 4.31. Also, smaller grain sizes and smaller filtration velocities cause less shear and allow more and smaller particles to settle in the filter thus producing a better filtrate quality. So, for example, the overall removal efficiency for the PVC suspension shows similar values to those of kaolin or lycopodium's when the rate of flow decreases significantly to the point to be practically identical, if not higher, when the filtration rate is set below the 0.6 l/min mark (Tables 4.2 — 4.6).

There is a particle size which gives a minimum filtration efficiency, where they are too large for diffusion, but too small for interception and sedimentation. Yao *et al.* (1971), Habibian and O'Melia (1975) and Adin *et al.* (1989) found a critical particle size of 1 μm at which filtration was least efficient, and this may explain why with PVC turbidity (0.5—1.5 μm) the system showed the lowest overall removal efficiency (Tables 4.2 — 4.6), despite the good initial quality (i.e. outlet from compartment number 1) of the filtrate caused by the surface deposition at the inlet face of compartment number 1 (mainly due to its lower particle size range). Figure 4.23

provides a good example of the kind of results obtained when operating with PVC suspension at high concentrations and rates of flow. The suspension concentration in the filtrate for PVC powder is patently higher than that obtained for the kaolin or lycopodium suspensions.

Straining is a function of suspended particle / grain size ratio which, according to Herzig *et al.* (1970), has to be at least 0.05 to be important (i.e. particle size of 25 µm when 500 µm grains are being used). This ratio can only be found within the coarser range of the kaolin particles (up to 53 µm, 20 %) for the finest filtering material (548 and 653 µm, in compartments 7—10) and for the lycopodium particles (35 µm in size) throughout the filter's depth, meaning that turbidity removal by straining is likely to take place for both kaolin and lycopodium powder particles. Another fact to take into account is that pore sizes are reduced and their shape changed continuously during filtration. Therefore, the higher the particle concentrations, the more likely is the removal of smaller particles by straining (Adin *et al.*, 1979). Bearing in mind that PVC particles are about 2 orders of magnitude smaller than the mean filter pore size, straining can be dismissed as a significant removal mechanism (Ives, 1965).

Besides, there is evidence from Sholji (1963) and Yao (1971) that the dominant transport mechanism for smaller, sub-micron particles is diffusion with gravity (due to their greater Brownian motion) and interception/sedimentation for larger particles, which leaves the 1 µm sized particles with no dominant transport mechanism. Ives and Sholji (1965) studied the effect of Brownian motion indirectly and Yao (1968) by direct experimentation. Thus, for particles less than 1 µm the movement becomes increasingly significant with decreasing sizes (Spielman and Goren, 1970; Spielman and FitzPatrick, 1973, Rajagopalan and Tien, 1979; Adamczyk *et al.*, 1983).

According to O'Melia(1985), removal of suspended particles by diffusion can be important in contact filtration for turbidity removal where small particles are not aggregated prior to filtration and in contact or direct filtration of waters containing humic substances. Particles in the size range of one to a few microns are most likely to penetrate a filter bed. Sub-micron particles can produce excessive head losses and usually require aggregation prior to filtration. PVC particle size range (between 0.5 and 1.5 µm) is, again, on the border line. The experimental results show that PVC turbidity produced the lowest initial head losses in most runs, compared with kaolin and lycopodium turbidities (Figures 4.74— 4.80). However, and as it has previously been described in Section 4.4, as time elapsed, PVC head loss curves rose steeply and ultimately reached, and even exceeded, the values obtained for kaolin and lycopodium. Both PVC and kaolin turbidities penetrated throughout the entire depth of the medium, mainly due to the fineness of their particles (2 µm in 80% of the kaolin particles).

Research by Adin and Elimelech (1989) showed that the removal efficiency of particles larger than 10 µm in direct granular filtration was relatively high and increased with filter grain size and decreased with filtration velocity. In fact, the experimental results show that the overall removal efficiency for both kaolin and lycopodium turbidities is consistently higher than 99.1 % and increases with decreasing flow rate and inlet concentration, even if not significantly (Tables 4.2— 4.6). In the experiments by Adin *et al.* (1989) larger grain sizes clearly showed better removal efficiency of very fine particles (1—10 µm). Figures 4.59a— 4.59h are examples of typical removal efficiency values obtained for each compartment; it is obvious from these graphs that the best filtration results (in excess of 60 per cent removal efficiency) correspond to compartment number 1, containing the coarser medium.

Adin and Elimelech (1989) also found that for particles larger than 10 µm and up to 60 µm there was no difference in the rate of particle removal. This was also observed during the experimental runs for both kaolin (up to 53 µm, 20%) and lycopodium (35 µm) turbidities (Figures 4.45— 4.52). The tendency of the removal curves toward a minimal efficiency for PVC turbidity (Figures 4.43— 4.44) supports the experimental and theoretical works of other investigators (Yao *et al.* 1971; Habibian and O'Melia, 1975 and Adin Elimelech, 1989).

According to Ives and Gregory (1967), the combined action of electrical double layer and van der Waal's forces play a fundamental role in removing PVC suspended particles from suspension. However, no studies based on direct particle size measurements were made on the behaviour of the suspended particles as it was beyond the scope of this investigation.

As described in section 3.3, kaolin is a heavy clay powder (absolute density 2,600 kg/m^3) and 2 to 53 μm in size (2 μm 80% and up to 53 μm 20%). Kaolin showed good dispersion and stability in water, although the particles lacked spherical isotropy. No particle size and distribution analyses were available for kaolin. However, Graham (1988) reported in his experiment that the particle size distribution for kaolin powder, based on a log-normal weight distribution, had a log mean of 4.5 μm, determined by a two-tube Coulter Counter analysis. Such a measurement takes place in a flow of liquid through a small orifice so it may be considered that the analysis is meaningful in a hydrodynamic sense (Ison and Ives, 1969).

Ison and Ives (1969) carried out a series of tests using a kaolin suspension in both upward and downward flow through a randomly packed bed of 50 mm glass ballotini and concluded that deposition occurred on the upper surface of the grains in both cases, thereby indicating that gravitation was almost certainly the dominant removal mechanism. Their experiments also showed that the absence of any deposition on the downstream face of the sphere implies that physicochemical adhesion cannot be regarded as a dominant force, and the absence of deposition on the upstream face in upward flow indicates that neither contact action nor diffusion can be regarded as important.

Although a number of removal mechanisms were operative, gravity was assumed to be the principal mechanism responsible for the removal of kaolin turbidity. According to O'Melia (1985), filtration efficiency increases with increasing suspended particle size and with decreasing filtration rate if gravity forces dominate particle transport. Ison's results demonstrated with some success the significance of interception, sedimentation and hydrodynamic mechanisms for kaolin in water.

The results obtained by Ison and Ives clearly illustrate the importance of the gravitational effect for suspensions with a density appreciably greater than water. Also, due to its greater particle size range, removed kaolin particles were found to be more uniformly distributed throughout the depth of the filter, as it is deduced from the concentration ratio curves in Figures 4.45—4.48 and 4.55—4.56.The particles of lycopodium turbidity were comparatively coarser, with a uniform size of 35 µm, as described in section 3.3. The removal of lycopodium particles was found to show somehow a similar behaviour to that of kaolin (Section 4.5). However, at high inlet concentrations a surface mat at the inlet face of compartment number 1 developed quicker and thicker than in the case of kaolin. This may be attributed to the larger particle size and the lower density (1,180 kg/m^3) of the lycopodium turbidity.

The development of head loss in the various filter compartments followed in some cases a curved relationship, which is characteristic of straining (Figure 4.68). The fact that the concentration should not change with time if straining was the operative removal mechanism was demonstrated first by Stein in 1940. Additional supporting evidence was presented by Ives (1963) and Eliassen and Tchobanoglous (1970).

Ives (1960) developed analytical theory of surface deposition assuming two modes of deposition:

- coating of the pores leading to a constriction of the flow
- build-up of deposit between the pores, lengthening the flow path

These two modes of deposition were observed with fibre optic endoscopes and recorded on video tape (Ives, 1986). He concluded that deposition on the surface is primarily caused by the sedimentation transport mechanism, forming caps on the upper surface of the filter grains, as previously suggested and noticed by Sholji (1963).

The concentration of suspended material in the filtrate is directly related to the concentration in the applied water or inlet concentration. The application of a high inlet concentration of suspended material to a filter causes increased particle penetration and an increased concentration in the filter effluent.

No significant differences were observed among the three types of synthetic turbidities in terms of initial concentration of suspended particles. Increases in the influent suspension were found to cause corresponding increases in the value of head loss at each rate of flow for all suspensions, as shown in Figures 4.72—4.76.

From the experimental results and previous investigations, it is evident that as the inlet concentration decreases the removal efficiency tends to become the same (Figures 4.37—4.40), particularly at lower flow rates (above 99.1 per cent in all cases at a filtration rate of 0.2 l/min), irrespective of the type of synthetic turbidities, as shown in Tables 4.2—4.6.

The laboratory results also indicate a number of ways in which the design of rapid filtration by a multi-compartment up-flow and down-flow process may be improved. One of these is the quality of the water applied to the filter bed.

Pretreatment is, of course, important in improving this characteristic: by the application of polyelectrolyte to the incoming water a higher removal efficiency may be achieved. These coagulant aids help to increase the shear strength of the deposits and may prevent detachment, as it was first suggested by Mints in 1966.

According to Ives and Rajapakse (1988), flocculation agents like aluminium sulphate assist the adhesion of suspended particles to the granular media and subsequent growth of these newly formed particles, thus improved water results, while filter runs of satisfactory length are obtained at higher filtration rates.

Thus, a suitable application for this new type of filter could be direct filtration. Direct filtration is a water treatment scheme that excludes sedimentation. All natural and added solids removal and storage occurs in the filter media. The objective of chemical pretreatment is to develop a floc that can be filtered rather than settled.

Direct filtration can be used when the raw water turbidity, colour and algae levels are low enough to allow reasonable run lengths. It has been suggested as optimum values a turbidity range of 0—10 NTU, a colour range of 0—15 APHA colour units, and 0—1000 units/ml of algae. The usual filtration design rate is 12.2 m/h. (Hilmoe and Cleasby, 1986).

Conditions inside a filter medium are good for coagulation. Coagulation consists of bringing small particles together, creating surfaces which will adhere to one another and avoiding shear forces which tend to rip the particles apart again.

Modern coagulant aids have increased the adhesion between coagulated particles and, in addition to this, in deep bed filters the average size of suspended particles can be made to grow through the medium. A coarse filter medium can therefore serve a dual function: to coagulate the particles and to hold the bigger flocs which are of course more easily retained by the medium.

Finally, the experimental results indicate that rapid filtration by a multi-compartment up-flow and down-flow process could achieve an improved removal of *Cryptosporidium* oocysts. However, a number of considerations need to be made (Logsdon, 1981; LeChevalier and Norton., 1992-1995; Ives *et al.*, 1993).

A critical factor in the ability of a filter to remove *Cryptosporidium* oocysts is the size of the oocysts. Oocysts at approximately 4 μm diameter lie near the minimum efficiency for sand filters of approximately 1 μm (Ives *et al.*, 1993). Chemical additions to aid coagulation and flocculation are necessary to achieve effective removals of oocysts. Settling with chemical additions has been found to remove up to 99.8% of *Cryptosporidium* oocysts. The effluent turbidity needs to be low (<0.3 NTU) and backwashing should be initiated by a small increase in turbidity (0.1—0.2

NTU). In conclusion, the laboratory results give a promising indication that, sizewise, the filter apparatus may provide effective removals of *Cryptosporidium* oocysts.

5.5 FILTRATION RATE

The effect of the rate of flow on removal efficiency has been briefly described in section 4.6. Typical data resulting from the experimental runs for PVC, kaolin, lycopodium powder, at different filtration rates and influent concentrations are represented in Figures 4.16—4.40, and 4.81— 4.83.

There is evidence that increase of flow rate through a filter will detach particles causing a more turbid filtrate (Tuepker, 1968). The effect is diminished if polyelectrolytes are applied to the suspension to be filtered.

As it has been previously stated, the rate of flow has a considerable influence on the fineness of filtration; low flow rates cause less shear and allow more and smaller particles to settle in the filter. Filtration efficiency was found to increase with a decrease in filter approach velocity, in accordance with theory. Thus, for example, for PVC the experimental results show a minimum filtration efficiency of 96.7 per cent turbidity removal at 1.0 l/min and 400 mg/l inlet concentration; this value rose to 99.3 per cent removal efficiency when the rate of filtration was kept at 0.2 l/min (Tables 4.2—4.6).

Nevertheless, perhaps one of the most obvious consequences of operating at higher rates of filtration is the head loss development and run length of the filter system. Rates of filtration were constant at 0.2, 0.4, 0.6, 0.8 or 1.0 l/min for any one filter run, all of which are in the range of normal waterworks filtration rates. Typical filter runs lasted 150 to 360 minutes, with total head loss rising from about 700 mm up to 2000 mm depending on the degree of clogging and rate of flow employed during the experiment. Extreme cases where those runs performed at 1.0 l/min, with filtration significant body of evidence to suggest that a move to a larger bed media size would improve performance.

5.6 FURTHER WORK

Although the process is far from being optimised and has only been tested with ideal suspensions and inflow concentrations of 400 mg/l and below, it shows great promise for practical application in most conventional water treatment plants either as a single stage treatment or as a polishing filter. However, further investigations are needed on the optimal design of the multi-compartment up-flow and down-flow filter.

These can be summarised as follows:

- Particle size - Redesign of model filter apparatus to provide larger pores which will result in both longer filter runs and higher removal efficiency. Experiment with different granulometries within the various compartments.
- Filtering material - Assess filter performance with different materials or even combination of materials in the various compartments.
- Inlet turbidity - Test the system with different levels and types of turbidity.
- Use of polyelectrolyte - This would provide a more direct and complete comparison of filtrate quality and operating characteristics with filtration tests conducted elsewhere.
- Up-flow filtration investigations
- Use of other parameters in order to assess filter's performance more accurately - pH, filter coefficient, particle counting, etc.

CHAPTER 6

CONCLUSIONS

1. Under the recorded laboratory conditions, rapid filtration by a multi-compartment up-flow and down-flow process appeared to be effective as a single stage treatment for relatively low turbidity waters (100—300 mg/l) in the flow range 0.2—0.8 l/min (2—8 m^3/m^2h).

2. The variation of head loss and of concentration ratio with depth and time conform with the generally accepted curves for rapid gravity filters. However, due to the apparatus' characteristics, the concentration ratio curves with depth and time show a typical pattern caused by the filter's various compartments and the influence of the direction of flow. Turbidity penetrated the entire depth of media under almost all experimental conditions, and there was a gradual reduction in turbidity all along the depth, even though the burden of removing the suspended solids took place mainly in compartments 1 and 2 (coarser media), and 3 and 4 to some extent.

3. Filtration efficiency increased with a decrease in filter flow rate, in accordance with the generally accepted theory.

4. Removal mechanisms of suspended matter are greatly affected by the physicochemical properties of the particles. Therefore, due to the differences in physicochemical properties of different types of synthetic turbidities, contrasting behaviour of the PVC, kaolin and lycopodium turbidities have been observed. These properties, especially particle size range, shape and density influence the process of filtration in removing the turbidity from water. The highest overall removal efficiency was achieved with the kaolin and

lycopodium turbidities in comparison to PVC suspension. However, the experimental results indicate that the model filter apparatus showed the highest initial removal efficiency (i.e. effluent from compartments 1 and 2) when the input to the system was PVC turbidity, due to the fineness of the suspended particles (0.5—1.5 μm).

5. The range of optimum overall removal efficiencies are 96.7 to 99.8%, 98.8 to 99.9% and 98.5 to 99.8% for PVC (0.5—1.5 μm), kaolin (2 μm 80%; up to 53 μm 20%) and lycopodium (35 μm) turbidity, respectively.

6. The conditions for the best turbidity removals (irrespective of filtration running times) are:

For PVC turbidity:	inlet concentration 150 mg/l, flow rate 0.2 l/min (1.9 m^3/m^2h)
For kaolin turbidity:	inlet concentration 400 mg/l, flow rate 1.0 l/min (9.4 m^3/m^2h)
For lycopodium turbidity:	inlet concentration 400 mg/l, flow rate 0.2 l/min (1.9 m^3/m^2h)

7. Unlike the quality of the filtrate, the head loss developed during filtration is very dependent upon media size, filtration rate, water temperature and the concentration of particles to be filtered.

8. The model filter apparatus should be redesigned to provide larger pores, particularly in compartments 1 and 2. Preliminary studies using the laboratory experimental system outlined previously have indicated that by increasing the size of the bed media alone (particularly in compartments 1 and 2), improved performance may be achieved at higher filtration rates and initial suspension concentrations. However, sufficient data are not yet available for firm conclusions to be drawn. Some other modifications may be also required. For example, it seems plausible that direct filtration of secondary effluents could provide both long filter runs and higher efficiency removal if compartments 1 and 2 contained media having a size of about 1500 μm and the system was operated at filtration rates in the order of 0.8 l/min (≈7.4 m^3/m^2h).

9. In future studies, the influent to the filter should be applied in conjunction with a polyelectrolyte. This would provide a more direct and complete comparison of filtrate quality and operating characteristics with filtration tests conducted with coagulant aids.

10. The exceptionally good and consistent performance from all filter units (in terms of solids removal) could mean that they have a greater capacity to hold solids than allowed for in design. If this is the case, higher loadings might be applied in general or used in future designs to allow smaller filter areas to be used.

11. The main disadvantage of up-flow filtration was the tendency for the bed to expand because of pressure differences overcoming the weight of the finer media during up-flow filtration. In those compartments containing the finest sand, means should be provided to prevent the bed expanding in up-flow filtration at high rates of flow.

12. The laboratory results with regard of size of suspended matter give a promising indication that the filter bed may provide effective removals of *Cryptosporidium* oocysts and *Giardia lamblia* cysts.

13. Although the process is far from being optimised and has only been tested with ideal suspensions and inflow concentrations of 400 mg/l and below, it shows great promise for practical application in most conventional water treatment plants either as a single stage treatment or as a polishing filter. However, further investigations are needed on the optimal design of the multi-compartment up-flow and down-flow filter.

BIBLIOGRAPHY

Adamczyk, Z., Dabros, T., Czarnecki, J. and Van de Ven, T.G.M. (1983). **Particle transfer to solid surfaces**. Advances in colloid and interface science. Vol. 19, pp. 183-252.

Adin, A., Baumann, E.R. and Cleasby, J.L. (1979). **The application of filtration theory to pilot-plant design**. Journal AWWA, *Water Technology / Quality*, January 1979, pp. 17-27.

Adin, A. and Elimelech, M. (1989). **Particle filtration for wastewater irrigation**. Journal of Irrigation and Drainage Engineering, Vol. 115, No. 3, pp. 474-486.

Adin, A. and Rebhun, M. (1987). **Deep-bed filtration: accumulation-detachment model parameters**. Chemical Engineering Science Journal, Vol. 42, No. 5, pp. 1213-1219.

Adin, A. and Rebhun, M. (1977). **A model to predict concentration and headloss profiles in filtration**. Journal AWWA, *Water Technology / Quality*, August 1977, pp. 444-453.

Adin, A. and Rebhun, M. (1974). **High-rate contact flocculation-filtration with cationic polyelectrolytes**. Journal AWWA, Vol. 66, No. 2, pp. 109.

American Water Works Association (1990). **Water quality and treatment: a handbook of community water supplies**. Fourth edition. Mc Graw-Hill Book Co., U. S. A.

Amirtharajah, A. (1978). **Optimum backwashing of sand filters**. Journal Environmental Engineering Division ASCE, Vol. 104, No 5, pp. 917.

Amirtharajah, A. and O'Melia, C. R. (1990). **Coagulation Processes: Destabilization, Mixing, and Flocculation**. Water quality and treatment. Fourth Edition. Mc Graw-Hill Book Co., U. S. A.

Arboleda, J. (1974). **Hydraulic control systems of constant and declining flow rate in filtration**. Journal AWWA, *Water Technology/Quality*. February, pp. 87-94.

Baghaei-Yazdi, N. and Ives, K.J. **Similitude studies of clarification of turbid water in granular fluidised beds**.

Baumann, E. R. (1978). **Granular media deep bed filtration**. Ann Arbor. USA.

Baumann, E.R. and Huang, J.Y.C. (1974). **Granular filters for tertiary waste water treatment**. Journal Water Pollution Control Fed., Vol. 46, pp. 1958-1973.

Baumann, E.R. and Ives, K.J. (1987). **The evidence of wormholes in deep bed filters**. *Proceedings of the Filter Technology Conference*, Utrecht, September 1987, Vol. 1, pp. 151-164.

Beard, J.B. and Tanaka, T.S. (1977). **A comparison of particle counting and nephelometry**. Journal of AWWA, Vol. 69, pp. 533-538.

Bellamy, W. D., Cleasby, J. L., Logsdon, G. S. and Allen, M. J. (1993). **Assessing treatment plant performance**. Journal AWWA, Management and operations, pp. 34-38. Dec.

Bhole, A.G. (1993). **Performance of static flocculators**. Journal Water Science Technology, Vol. 27, No. 10, pp. 181-194.

Bhole, A.G. and Ives, K.J. (1973). **Theory of flocculation for continous flow system**. ASCE Journal of the Environmental Engineering Division, Vol. 99 (EE1), pp. 17-34.

Bhole, A.G. and Ughade, V.A. (1981). **Study of a surface contact flocculator**. Presented at the 13th Annual Convention of IWWA, Pune, 30th January 1981, Journal of the IWWA, Vol. XIII, No. 2, pp. 179-183.

Boby, W. M. T. and Alpe, G. (1967). **Practical experiences using upward flow filtration**. Other details n/a.

Boller, M. (1993). **Removal of organic matter by physico-chemical mechanisms in wastewater treatment plants**. Water Science Technology Journal, Vol. 27, No. 11, pp. 167-183.

Boreli, M. and Jovasovic, D. (1961). **Clogging of porous media**. Int. Ass. Hydraulic Research. 9 th Convention, Dubrovnik, pp. 516-520.

Borrill, R. J. and Mc Kean, J. (1993). **Improving the operation of drinking water filters using particle size analysis**. *Paper to AWWA Water Technology Conference in Miami, U. S. A.* Severn Trent Water Ltd, U. K.

Camp, T. R. (1964). **Theory of water filtration**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, Vol. 90, SA4, pp. 1-30.

Carman, P.C. (1937). **Fluid flow through granular beds**. Transactions-Institution of Chemical Engineers, pp. 150-166.

Caro, P. (1992). **L'aigua**. Edicions de la Magrana, SA, Catalonia.

Catalán, J. (1981). **Química de l'aigua**. Second edition. Bellisco, Libreria Editorial, Spain.

Chadwick, B. and Leech, J. (1988). **Granular activated carbon for rapid gravity filtration**. Presented at the *Symposium on Rapid Filtration of Potable Waters*, 16th February 1988, Huntingdon, U.K.,Institution of Water and Environmental Management, Scientific Section, pp. 8-29.

Chadwick, A. and Morfett, J. (1993). **Hydraulics in civil and structural engineering**. Second edition. E & FN Spon, U. K.

Clark, R.G. (1988). **Experience in filtration for iron and manganese removal**. Presented at the *Symposium on Rapid Filtration of Potable Waters*, 16th February 1988, Huntingdon, U.K.,Institution of Water and Environmental Management, Scientific Section, pp. 52-59.

Cleasby, J.L. and Baumann, E.R. (1962). **Selection of sand filtration rates**. Journal AWWA, Vol. 54, pp. 579.

Cleasby, J. L., Williamson, M. M. and Baumann, E. R. (1963). **Effect of filtration rate changes on quality**. Journal AWWA, July, Vol. 55, Part 7, pp. 869-877.

Cleasby, J. L. and Woods, C. F. (1975). **Intermixing of dual media and multimedia granular filters**. Journal AWWA, Vol. 67, pp. 197-203.

Cleasby, J. L., Stangl, E. W. and Rice, G. A. (1975). **Developments in backwashing of granular filters**. Journal Environmental Engineering Division, ASCE, Vol. 101, EE5, pp. 713-727.

Cleasby, J.L., Arboleda, J., Burns, D.E., Prendiville, P.W. and Savage, E.S. (1977). **Backwashing of granular filters**. Journal AWWA, Research, February, pp. 115-126.

Cleasby, J. L. (1990). **Filtration**. Water quality and treatment. Fourth Edition. Mc Graw-Hill Book Co., U. S. A.

Cleasby, J. L. (1993). **Status of declining rate filtration design**. Water Science Technology, Vol. 27, No. 10, pp. 151-164.

Clough, G. and Ives, K. J. (1986). **Deep bed filtration mechanisms observed with fibre optic endoscopes and CCTV**. pp. 13.21-13-26.

Coad, M. A. and Ives, K. J. (1981). **Investigation of deep bed filtration using tracers**. Filtech Conference, pp. 131-136.

Coad, M. A. and Ives, K. J. (1981). **Deposition and flow patterns in the pores of deep bed filters**. *Proceedings of the Second World Congress on Chemical Engineering*, Montreal, pp. 111-114.

Coleman, J. J., Krovvidy, S., Suidan, M., Summers, R. S., Rossman, L. and Wee, W. G. (1994). **Intelligent sequence planning for wastewater treatment systems**. IEEE Expert, December, pp. 15-20.

Conley, W. R. (1965). **Integration of the clarification process.** Journal AWWA, Vol. 57, No. 10, pp. 1333.

Conley, W.R. and Pitman, R.W. (1960). **Test program for filter evaluation at Hanford**. Presented at the *Pacific North-west Section Meeting*, 24th April 1959, Vancouver, Canada. Journal AWWA, February, pp. 205-215.

Crawford, M. (1976). **Air pollution control theory**. McGraw-Hill, USA.

Croll, B.T. (1992). **Purification of water using carbons**. Presented at *Environmental aspects of coal utilisation and carbon science*, 31st March - 2nd April 1992, University of Newcastle upon Tyne, U.K. Anglian Water Services Ltd., U. K.

Croll, B.T. (1988). **The changing role of rapid gravity filtration in lowland surface water treatment**. Presented at the *Symposium on Rapid Filtration of Potable Waters*, 16th February 1988, Huntingdon, U.K., Institution of Water and Environmental Management, Scientific Section, pp. 1-3.

Culp, G.L., Culp, R.L. and Mack Wesner, G. (1978). **Filtration**. Handbook of advanced wastewater treatment, Second Edition, pp. 101-165. Van Nostrand Reinhold Company, U.S.A.

Davis, E. and Borchardt, J. A. (1966). **Sand filtration of particulate matter**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, October, Vol. SA 5, pp. 47-59.

DÉGRÉMONT. **Manual tècnic de l'aigua**. Bilbo, Basque Country, 1979.

Di Bernardo, L. (1987). **Designing declining-rate filters**. Filtration & Separation, September/December, pp. 338-341.

Di Bernardo, L. and Cleasby, J. L. (1980). **Declining-rate versus constant-rate filtration**. ASCE Journal of the Environmental Engineering Division. Vol. 106 (EE1), pp. 1023-1041.

Diaper, E. W. J. and Ives, K. J. (1965). **Filtration through size-graded media**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, Vol. 91 (SA3), pp. 89.

Ebie, K. and Amano, S. (1993). **Fundamental behaviour of humic acid and kaolin in direct filtration of simulated natural surface water**. Journal Water Science Technology, Vol. 27, No. 11, pp. 61-70.

Eliassen, R. (1935). **And experimental and theoretical investigation of the clogging of a rapid sand filter**. Sc.D. Thesis. MIT, USA.

Eliassen, R. and Tchobanoglous, G. (1970). **Filtration of treated sewage effluent**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*. Vol. SA2, pp. 243-265.

Fair, G.M. and Hatch, L.P. (1933). **Fundamental factors governing the streamline flow of water through sand**. Journal AWWA, Vol. 25, pp. 1551-1565.

Fair, G. M., Geyer, J. C. and Okun, D. A. (1968). **Water and wastewater engineering**. Wiley, New York. USA.

Fitzpatrick, J. A. and Swanson, C.L. (1986). **Performance tests on tertiary granular media filters**. Effluent and Water Treatment Journal, June, pp. 179-184.

Folkard, G.K., Al-Khalili, R. and Sutherland, J.P. (1996). **Contact flocculation filtration using a natural polyelectrolyte for the treatment of low turbidity surface water in developing countries**. Proceedings of the 7th Gothenburg Symposium on Chemical water and wastewater treatment, Edinburgh, Scotland.

Fox, D. M. and Cleasby, J. L. (1966). **Experimental evaluation of sand filtration theory**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, October, Vol. SA 5, pp. 61-81.

Friedlander, S.K. (1958). **Theory of aerosol filtration**. Industrial and Engineering Chemistry, Vol. 50, pp. 1161-1164.

Gates, W. C. B. (1993). **New wells: safeguard scenic Tahoe**. Civil Engineering Journal, May, pp. 60-62. U.S.A.

Ghosh, M.M., Jordan, T. A. and Porter, R. L. (1975). **Physicochemical approach to water and waste water filtration**. Journal of the Environmental Engineering Division, EE1, February, pp. 71-86.

Ghosh, G. (1958). **Mechanism of rapid sand filtration**. Water and Water Engineering Journal, April, pp. 147-153.

Ghosh, G. (1958). **Media characteristics in water filtration**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE.*, Vol. SA 1, February, pp. 1533-1-1533.25.

Gimbel, R. **Influence of filter grain surface structure on the transport and adhesion mechanisms in deep-bed filters**. pp.1.23-1.33.

Graham, N.J.D. (1988). **Filter pore flocculation as a mechanism in rapid filtration**. Water Research Journal, Vol. 22, No. 10, pp. 1229-1238.

Graham, N. J. D. and Hayes, C. R. (1994). **Water recycling in manufacturing industries and water pollution control**. Why recycle ?, pp. 123-131. Rainbow Ed., Balkema, Rotterdam. ISBN 90 54 10 367 1.

Gregory, R. and Zabel, T. F. (1990). **Sedimentation and flotation**. Water quality and treatment. Fourth Edition. Mc Graw-Hill Book Co., U. S. A.

Gur, A. (1969). **Theory and optimization of water filtration**. PhD Thesis, University of London.

Habibian, M.T. and O'Melia, C.R. (1975). **Particles, polymers and performance in filtration**. ASCE Journal Environmental Engineering Division, Vol. 101, No. EE4, pp. 567.

Hall, W.A. (1957). **An analysis of sand filtration**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, Vol. 83, SA3, pp. 1276.1-1276.9.

Hammer, M. J. (1986). **Water and wastewater technology**. Second edition. Prentice Hall, Inc., U. S. A.

Hargeshiemer, E.E., Lewis, C.M., Yentsch, C.M., Satchwill, T. and Mieklke, J.L. (1991). **Pilot scale evaluation of filtration processes using particle counting**. Proceedings of the AWWA, , pp. 323-344.Water Quality Technology Conference, San Diego, USA

Heinke, G. and Tay, J. A. (1980). **Effects of chemical addition on the performance or settling tanks**. Journal Water Pollution Control Federation. Vol. 52, pp. 12.

Herzig, J.P., Le Clerc, D.M. and LeGoff, P. (1970). **Flow of suspensions through porous media**. Ind. Eng. Chem. Vol. 62, No. 5, pp.8.

Hilmoe, D. J. and Cleasby, J. L. (1986). **Comparing constant-rate and declining-rate direct filtration of a surface water**. Journal AWWA, *Research & Technology*. December, Vol. 78, pp. 26-34.

Hsiung, K.Y. and Cleasby, J.L. (1968). **Prediction of filter performance**. Journal Sanitary Engineering Division, Proceedings ASCE, Vol. 94, SA6, pp. 1043-1069.

Hudson, H. E. (1963). **Functional design of rapid sand filters**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*. January, Vol. 89, No. SA1, pp. 17-28.

Hulbert, R. and Herring, F.W. (1929). **Studies on the washing of rapid filters**. Journal AWWA, Vol. 21, pp. 1445-1513.

Hutchinson, W. and Foley, P.D. (1974). **Operational and experimental results of direct filtration**. Journal AWWA, Water Technology / Quality. February, pp. 79-87.

Ison, C.R. and Ives, K.J. (1969). **Removal mechanisms in deep bed filtration**. Journal Chemical Engineering Science, Vol. 24, pp. 717-729. Pergamon Press. U.K.

Ives, K. J. (1959). **Rapid filtration**. Robert Blair Fellowship: Research report. Harvard University.

Ives, K.J. (1988). **Rapid filtration: past, present and future**. Presented at the *Symposium on Rapid Filtration of Potable Waters*, 16th February 1988, Huntingdon, U.K.,Institution of Water and Environmental Management, Scientific Section, pp. 60-68.

Ives, K.J. (1987). **Filtration of clay suspensions through sand**. Clay Minerals, *The Mineralogical Society*, Vol. 22, pp. 49-61.

Ives, K. J. and Horner, R. M. W. (1973). **Radial filtration**. *Proceedings of the Institution of Civil Engineers*, Vol. 55, Part 2, pp. 229-249.

Ives, K.J. (1979). **The basis for the application of multiple layer filters to water treatment**. Zeitschr. für Wasser und Abwasser Forschung, Vol. 12, No. 3/4/79, pp. 106-110.

Ives, K.J. (1969). **Deep bed water filters: new developments**. Filtration & Separation, January/February, pp. 42-49. *Proceedings of the Filtration Society.*

Ives, K. J. (1974). **Solid-liquid separation**. Effluent and Water Treatment Journal, November, pp. 636-645.

Ives, K.J. (1964). **Filtration - fact and fiction**. Presented at the *Effluent and Water Treatment Convention,* 1963. Manual 1964, pp. 43-52, Thunderbird Enterprises, London, U.K.

Ives, K. J. (1976). **Water filtration: the Mints-Ives controversy 1960-73**. Filtration & Separation, March/April, pp. 131-133.

Ives, K.J. (1966). **A theory of filtration**. Water, Vol. 50, No. 8, pp. 108-114. The Hague, Netherlands.

Ives, K. J. (1964). **Aquifer recharge with waste water**. Effluent and Water Treatment Journal, April, pp. 184-189.

Ives, K. J. (1972). **Capture mechanisms in filtration**. pp. 183-201. Other details n/a.

Ives, K. J. (1978). **Optimization model for tertiary treatment rapid filtration**. Mathematical models in water pollution control, chapter 16. pp. 339-351. John Wiley & Sons, U. K.

Ives, K. J. (1985). **Deep bed filters in mathematical models and design methods in solid-liquid separation.** Ed. A. Rushton, Nijhoff, Dordrecht.

Ives, K.J. (1968). **The functioning of deep-bed filters**. Trans. Institution of Chemical Engineers, Vol. 46, pp. T283-T286.

Ives, K. J. (1985). **Mathematical models in solid-liquid separation.** NATO ASI series.

Ives, K.J. (1960). **Rational design of filters**. Proceedings of the Institution of Civil Engineers, Vol. 16, pp. 189-193.

Ives, K. J. and Gregory, J. (1967). **Basics concepts of filtration**. Proceedings Society of Water Treatment and Exam., Vol. 16, No. 3, pp. 147-169.

Ives, K. J. (1965). **Research on deep filters**. *Trans. Institution of Chemical Engineers*, Vol. 43, Part 8, pp. T 238.

Ives, K.J. (1980). **Filtration in waste water treatment**. La Tribune du CEBEDEAU, No. 444, 33, pp. 455-461. Liège, Belgium.

Ives, K.J. (1980). **Deep bed filtration: theory and practice**. Filtration and Separation, Vol. 17, No. 2, pp. 157-168. *Proceedings of the Filtration Society.*

Ives, K.J. (1964). **Progress in filtration**. Presented at the *Annual Conference*, 4th June, Toronto, Canada. Journal AWWA, Vol. 56, pp. 1225-1232.

Ives, K. J. **The significance of theory**. Filtration. Other details n/a.

Ives, K.J. (1970). **Advances in deep-bed filtration**. Trans. Institution of Chemical Engineers, Vol. 48, pp. T94-T100.

Ives, K. J. (1961). **Filtration using radioactive algae**. Journal of the Sanitary Engineering Division, Proceedings of the ASCE, Vol. 87 (SA3), pp. 23-37.

Ives, K. J. (1979). **La filtration dans le traitement des eaux résiduaires**. La Tribune du CEBEDEAU, No. 426, pp. 187-192

Ives, K. J. (1979). **Derzeitiger Stand der tertiären Schnellfiltration in Großbritanien**. Gas Wasser Abwasser, Vol. 32, Part 1, pp. 31-34.

Ives, K. J. (1970). **Depth filtration of liquids**. Filtration & Separation, November/December, pp. 700-704. *Proceedings of the Filtration Society.*

Ives, K. J. (1986). **Developments in deep bed filtration**. Proceedings of the 3rd World Congress of Chemical Engineering, September, Tokyo, Japan. Vol III, pp. 69-72.

Ives, K. J. (1963). **Simplified rational analysis of filter behaviour**. Proceedings of the Institution of Civil Engineers, Vol. 25, pp. 345-364.

Ives, K. J. (1970). **Rapid filtration**. Water Research Journal, Vol. 4, pp. 201-223.

Ives, K. J. (1960). **Filtration through a porous septum: a theoretical consideration of Boucher's law**. Proceedings of the Institution of Civil Engineers, *Paper No. 6459*, Vol. 17, pp. 333-337.

Ives, K. J. (1978). **A new concept of filterability**. Submitted for IAWPR, Stockholm.

Ives, K. J. and Gregory, J. (1966). **Surfaces forces in filtration**. Proc. Soc. Water Treatment and Exam., Vol. 15, No. 2, pp. 93-116.

Ives, K. J. and Fourie, J. (1983). **Optical techniques in process control**. Paper presented at the *International Conference on Optical Techniques in Process Control*, The Hague, The Netherlands, 14th-16th June 1983.

Ives, K. J., Atkin, J. R. and Thompson, R. P. (1968). **Measurement of turbidity**. Part 1. Journal Effluent Water Treatment, Vol. 8, No. 7, pp. 342-348.

Ives, K. J. and Fitzpatrick, C.S. B. (1989). **Detachment of deposits from sand grains**. Colloids and Surfaces, No. 39, pp. 239-253. *Elsevier Science Publishers BV*, The Netherlands.

Ives, K. J. and Rajapakse, J. P. (1988). **Pretreatment with pebble matrix filtration**. Ed. N. J. D. Graham, Ellis Horwood Chichester, pp. 141-152.

Ives, K. J. and Rajapakse, J. P. (1990). **Pre-filtration of very highly turbid waters using pebble matrix filtration**. Journal Inst. Water Environ. Management, April, Vol. 4, pp. 140-147.

Ives, K.J. and Pienvichitr, V. (1965). **Kinetics of the filtration of dilute suspensions**. Journal Chemical Engineering Science, Vol. 20, pp. 965-973. Pergamon Press, U.K.

Ives, K.J. and Sholji, I. (1965). **Research on variables affecting filtration**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*. Vol. 91 (SA4), pp. 1-19.

Ives, K. J. and Mörgeli, B. (1979). **New media for effluent filtration**. Water Research Journal, Vol. 13, pp. 1001-1007.

Ives, K.J., Gregory, J. Scutt, J. and Pugh, H. (1993). **A microsphere in water:** ***Cryptosporidium*** **parvum**. Presented at 6th World Filtration Congress, Nagoya, Japan.

Iwasaki, T. (1937). **Some notes on sand filtration**. Journal AWWA, Vol. 29, pp. 1591.

Jago, P. H. (1977). **Review of filtration techniques for tertiary treatment of sewage effluents**. Technical Report TR64, December, *Water Research Centre*, U. K.

Kawamura, S. (1975). **Design and operation of high-rate filters**. Parts 1-3. Journal AWWA, Vol. 67, pp. 544-708.

Kazi, N. M. (1995). **Pneumatic flocculation in water treatment**. Ph.D. Thesis. Nottingham Trent University, UK.

Kiely, G. (1997). **Environmental engineering**. Mc Graw-Hill Publishing Co., U. K.

Klei, H.E. and Sundstrom, D.W. (1979). **Theory of granular filters**. Wastewater treatment, pp. 223-229. Prentice-Hall, Inc., U.S.A.

Korte, F. (1977). **What is water?** Water Quality, *Proceedings of an International Forum*, Ed. Coulston, F. & Mrak, E. Academic Press, U. S. A.

Kozeny, J. (1927). Sitzber. Akad. Wiss. Wien, Math-naturw., Kl. Abt. Iia, Vol. 136, pp. 271-306.

Kreissl, J.F., Robeck, G.G. and Sommerville, G.A. (1968). **Use of pilot filters to predict optimum chemical feeds**. Journal AWWA, Vol. 60, No.3 , pp. 299.

Lambert, S. D. and Graham, N. J. D. (1995). **A comparative evaluation of the effectiveness of potable water filtration processes**. Journal Water SRT-Aqua, Vol. 44, No. 1, pp. 38-51.

Lander, J. (1994). **Wastewater rapid-gravity filtration in Severn Trent Water**. Journal IWEM, June, Vol. 8, Part 3, pp. 256-268.

Lang, J. S., Giron, J. J., Hansen, A.T., Rhodes Trussell, R. and Hodges, W. E. (1993). **Investigating filter performance as a function of the ratio of filter size to media size**. Journal AWWA, Research and Technology. October, pp.122-130.

Lawler, D. F., O'Melia, C. R. and Tobiason, J. E. (1980). **Integral water treatment plant design: from particle size to plant performance**. American Chemical Society Journal, *Particulates in Water*, 16, pp. 353-388.

LeChevalier, M.W. and Norton, W.D. (1992). **Examining the relationships between particle counts and** *Giardia, Cryptosporidium* **and turbidity**. Journal AWWA, Vol. 84, No. 12, pp. 52.

LeChevalier, M.W. and Norton, W.D. (1995). *Giardia* **and** *Cryptosporidium* **in raw and finished water**. Journal AWWA, Vol. 87, No. 9, pp. 54.

Lekkas, T. D., Fox, G. T. J., and Mc Naughton, J. G. (1978). **Comparison of two and three layer granular bed filters for treating reservoir water**. pp. 239-248.

Ling, J.T. (1955). **A study of filtration through uniform sand filters**. Proceedings ASCE, 81. No. 751.

Litwiniszyn, J. (1963). **Colmatage considered as a certain stochastic process**. Bulletin Acad. Polon. Sci., Ser. Sci. Tech., Vol. 11, pp. 81-85.

Logsdon, G. S. (1981). **Filtration operation and maintenance for small systems**. *Black & Veatch, Cincinnati, U. S. A.* pp. 601-618.

Mackrle, V. (1960). **Study of the adhesion phenomenon. Clogging of the porous medium**. D. Sc. Thesis. University Grenoble, France.

Mackrle, S. and Mackrle, V. (1962). **Adhesion in filters**. American Society of Civil Engineers-Transactions, Vol. 127, Part III, pp. 266-281.

Mason, C. F. (1991). **Biology of freshwater pollution**. Second edition. Longman Group Ltd. U. K.

Matthews, P. (1995). **Innovations in wastewater treatment: a European perspective**. Presented at the *Water Environment Federation Specialty Conference:* New and emerging environmental technologies and products conference for wastewater treatment and stormwater collection, 4th June 1995, Toronto, Québec, Canada.

Mc Ghee, T. J. (1991). **Water supply and sewerage**. Sixth edition. Mc Graw-Hill Book Co., U. S. A.

Mints, D. M. and Krishtul, V. (1960). **Investigation of the process of filtration of a suspension in a granular bed**. Zhurnal Prikladnoi Khimii, Vol. 33, No. 2, pp. 304-316

Mints, D.M. (1969). **Preliminary treatment of water before filtration**. Special Report No. 6. International Water Supply Congress Proceedings, Vienna. IWSA, London, UK.

Mints, D. M. (1966). **Modern theory of filtration**. International Water Supply Association Seventh Congress, Barcelona, Catalonia, Vol. 1, Special Subject No. 10, P3-P29.

Mints, D. M. (1951). **Kinetics of filtration of low-concentration water suspensions in water purification filters**. Dokl. Akad. Nauk. Vol. 78, pp. 315-318.

Moffett, J.W. (1968). **The chemistry of high-rate water treatment**. Presented at the AWWA *Annual Conference*, 5th June 1968, Cleveland, U.S.A.

Mohanka, S.S. (1969). **Multilayer filtration**. Journal AWWA, Vol. 61, pp. 504.

Moran, M.C., Moran, D.C., Cushing, R.C. and Lawler, D.F. (1993). **Particle behaviour in deep-bed filtration**: Part II--Particle Detachment. Journal AWWA, Research and Technology. December, pp. 61-81.

Muslu, Y. (1987). **Shape factor and degree of packing in fluidization**. Journal of Environmental Engineering, Vol. 113, No. 2, pp. 311-329.

Oeben, R. W., Haines, H.P. and Ives, K. J. (1968). **Comparison of normal and reverse-graded filtration**. Journal AWWA, Vol. 60, No. 4, pp. 429-439.

Ojha, C. S. P. and Graham, N. J. D. (1993). **Theoretical estimates of bulk specific deposit in deep bed filters**. Water Research Journal, Vol. 27, No. 3, pp. 377-387.

Oron, G., Shelef, G. and Zur, B. (1980). **Storm water and reclaimed effluent in trickle irrigation**. ASCE Journal of the Irrigation and Drainage Division, *Proceedings of the ASCE*, December, Vol. 106, No. IR4, pp. 299-310.

Ott, C. R. and Bogan, R. H. (1970). **Theoretical evaluation of filter modeling experiments**. Journal of the Sanitary Engineering Division, *Proceedings of the ASCE*, April, Vol. SA 2, pp. 455-465.

O'Melia, C.R. (1974). **The role of polyelectrolytes in filtration processes**. EPA-670/ Vol. 2, No. 74, pp. 32.

O'Melia, C.R. (1985). **Particles, pretreatment and performance in water filtration**. ASCE Journal of Environmental Engineering, December, Vol. 111, No. 6, pp. 874-890.

O'Melia, C.R. and Stumm, W. (1967). **Theory of water filtration**. Journal AWWA, November, pp. 1393-1412.

O'Melia, C.R. and Ali, W. (1973). **The role of retained particles in deep bed filtration**. Progress in Water Technology, Vol. 10, pp. 167-182.

Patwardhan, S.V. (1981). **Water treatment**. Presented at the *Conference on R&D in engineering*, 61st Annual Convention, 8th-9th February 1981, Hyderabad.

Peavy, H. S., Rowe, D. R. and Tchobanoglous, G. (1985). **Environmental engineering**. International edition. Mc Graw-Hill Book Co., U. S. A.

Peirce, J.J., Vesilind, P.A. and Weiner, R.F. **Water treatment**. Environmental engineering, third edition, pp. 117-136. Butterworth-Heinemann, U.K.

Pettet, A.E.J., Collett, W.F. and Waddington, J.I. (1951). **Mechanical filtration of sewage effluents, III, Removalof humus: further experiments at Luton**. Journal Proceedings Inst. Sew.Purif., Pt. 2, pp. 195-202.

Powelson, D. K. and Gerba, C. P. (1994). **Virus removal from sewage effluents during saturated and unsaturated flow through soil columns**. Water Research Journal, Vol. 28, No. 10, pp. 2175-2181. Elsevier Science Ltd. U.K.

Powelson, D. K., Simpson, J. R. and Gerba, C. P. (1990). **Virus transport and survival in saturated and unsaturated flow through soil columns**. Journal of Environmental Quality, Vol. 19, pp. 396-401.

Rajagopalan, R. and Tien, C. (1979). **The theory of deep bed filtration**. Progress in filtration and separation, Vol. 1, R.J. Wakeman, Ed. Elsevier, pp. 179-269. USA

Resnick, W. and Wolf, D. (1963). **Residence time distribution in real systems**. I & EC Fundamentals, Vol. 2, No. 4.

Sawyer, C. N., Mc Carty, P. L. and Parkin, G. F. (1994). **Chemistry for environmental engineering**. Fourth edition. Mc Graw-Hill Book Co., U. S. A.

Schulz, C. R. and Okun, D. A. (1984). **Surface water treatment for communities in developing countries**. John Wiley & Sons, U. S. A.

Sembi, S. (1981). **Optimisation of size-graded filters**. PhD Thesis, University of London, UK.

Sembi, S. and Ives, K. J. (1983). **Optimisation of size-graded water filters**. Paper presented at the *Symposium on Water* Filtration, Antwerp, Belgium, April 1982. Filtration & Separation, September/October, pp. 398-402. *Proceedings of the Filtration Society*.

Sholji, I. (1963). The Filtration of Suspensions through Deep Granular Filters, Ph. D. University of London, UK.

Sholji, I. (1987). **Expansion of granular filters during backwashing**. Journal of Environmental Engineering, Vol. 113, No. 3, pp. 516-531. ASCE.

Sholji, I. and Johnson, F.A. (1987). **Comparison of backwash models for granular media**. Journal of Environmental Engineering, Vol. 113, No. 3, pp. 532-549. ASCE.

Spielman, L.A. and Goren, S.L. (1970). **Capture of small particles by London forces from low-speed liquid flows**. Environmental science and technology. Vol. 4, pp. 135-140.

Spielman, L.A. and FitzPatrick, J.A. (1973). **Theory of particle capture under London and gravity forces**. Journal of Colloid and Interface Science, Vol. 42, pp. 607-623.

Sprouse, G. and Rittmann, B. E. (1990). **Colloid filtration in fluidized beds**. Journal of Environmental Engineering, Vol. 116, No. 2, pp. 299-312. ASCE, U.S.A.

APHA, AWWA and WEF (1998). **Standard Methods for the examination of water and wastewater**, 20th Edition. 2130 Turbidity, Section 2-9. Edited by Clesceri, Greenberg and Eaton.

Stanley, D.R. (1955). **Sand filtration studied with radiotracers**. Proceedings ASCE, 81, No. 592.

Stein, P. C. (1940). **A study of the theory of rapid filtration of water through sand**. Sc.D.Thesis, MIT, USA.

Stetterfield, G. H. (1983). **Water treatment trials at Burham**. Effluent and Water Treatment Journal, January, pp. 18-23.

Stoodley, A.E. (1989). **Operational experiences with package filter sewage treatment plants**. Journal IWEM, 3, December, pp. 583-587.

Taylor, G. S., Hillis, P. and Walker, I. (1993). **Pilot-plant trials on River Dee Water at Huntington**. Journal IWEM, 7, August, pp. 333-343.

Tchobanoglous, G. (1970). **Filtration techniques in tertiary treatment**. Journal Water Pollution Control Fed., Vol. 42, pp. 604-623.

Tchobanoglous, G. and Burton, F. L. (1991). **Wastewater engineering: treatment, disposal and reuse**. Third edition. Metcalf & Eddy. Mc Graw-Hill Book Co., U. S. A.

Tebbutt, T.H.Y. (1971). **An investigation into tertiary treatment by rapid filtration**. Water Research Journal. Vol 5, pp. 81-92.

Tebbutt, T. H. Y. (1983). **Principles of water quality control**. Third edition. Pergamon Press. University of Birmingham, U. K.

Tien, C. and Gimbel, R.D. (1982). **On the development of a comprehensive model of deep bed filtration**. Proceedings of the International Symposium on Water Filtration, Weiler & Janssens Eds., Koninklijke Vlaamse Ingenieursvereniging, Antwerp, Belgium.

Treanor, A.I. (1971). **Filtration through deep beds**. Process Biochemistry, September, pp. 21-42.

Tuepker, J.L. and Buescher, C.A. (1968). **Operation and maintenance of rapid sand mixed-media filters in a lime softening plant**. Journal AWWA, Vol. 60, No. 12, pp. 1377.

Twort, A. C., Law, F. M., Crowley, F. W. and Ratnayaka, D. D. (1994). **Water supply**. Fourth edition. Edward Arnold, U. K.

Vesilind, P. (1975). **Environmental pollution and control**. Ann Arbor Science. USA.

Viessman, W. and Hammer, M. J. (1993). **Water supply and pollution control**. Fifth edition. Harper Collins College Publishers, U. S. A.

Visscher, J.T., Paramasiaam, R., Raman, A. and Hijnen, A. (1987). **Slow sand filtration for community water supply planning, design, construction and maintenance**. Technical Paper 24, IRC, U.K.

Watson, A. (1988). **The penetration of rapid gravity filters by plankton**. Presented at the *Symposium on Rapid Filtration of Potable Waters*, 16th February 1988, Huntingdon, U.K.,Institution of Water and Environmental Management, Scientific Section, pp. 34-49.

White, I. D., Mottershead, D. N. and Harrison, S. J. (1992). **Environmental systems: an introductory text**. Second edition. Chapman & Hall, U. K.

Wilson, E. M. (1990). **Engineering hydrology**. Fourth edition. Macmillan Education Ltd., U. K.

Yao, K. M., Habibian, M. T. and O'Melia, C. R. (1971). **Water and waste water filtration; concepts and applications**. Environmental Science & Technology, November, Vol. 5, No. 11, pp. 1105-1112.

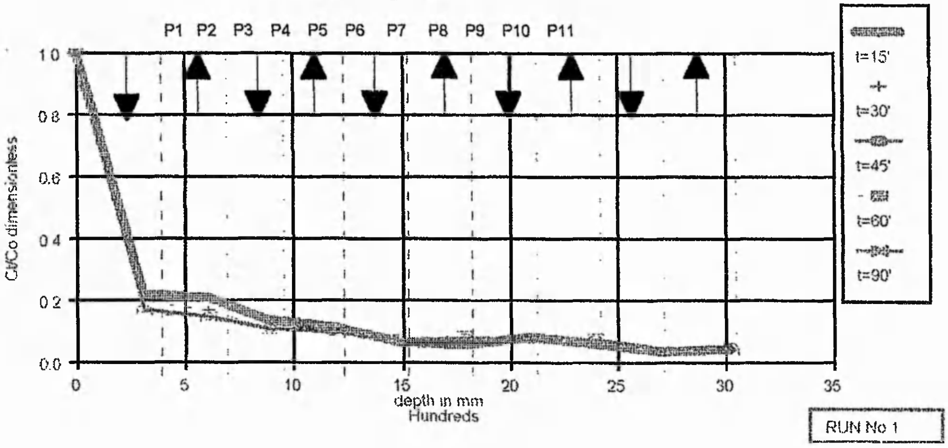
Appendix 1 Experimental Data Graphs

IMPORTANT

Please note that due to lack of printing resolution, C/C_0 appears to be 1.0 at t=0. This should not be so. Even if only minimally, $C/C_0 < 1.0$ at t=0, (from Iwasaki's equation).

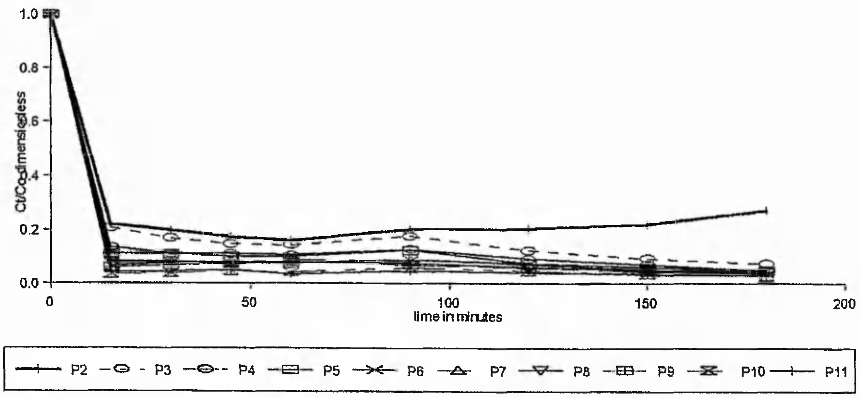
CONCENTRATION RATIO Vs DEPTH

PVC Co=400 mg/l Q=1.0 l/min



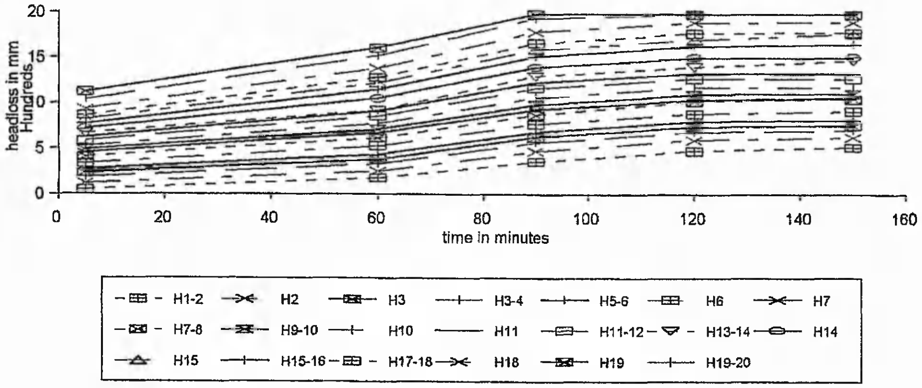
CONCENTRATION RATIO Vs TIME

PVC Co=400 mg/l Q=1.0 l/min RUN 1



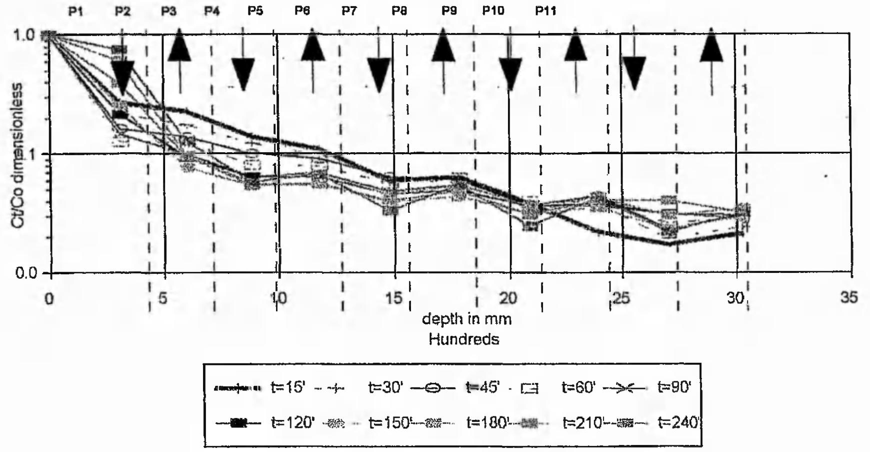
HEADLOSS Vs TIME

RUN 1 PVC Co=400 mg/l Q=1.0 l/min



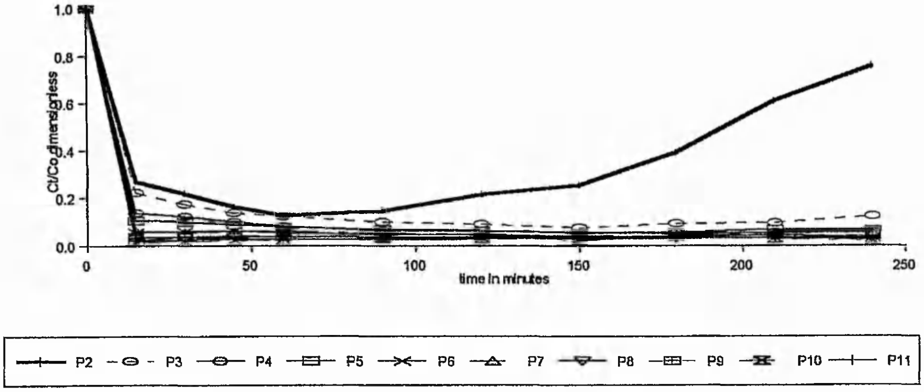
CONCENTRATION RATIO Vs DEPTH

RUN No 2 PVC Co=400 mg/l Q=0.8 l/min



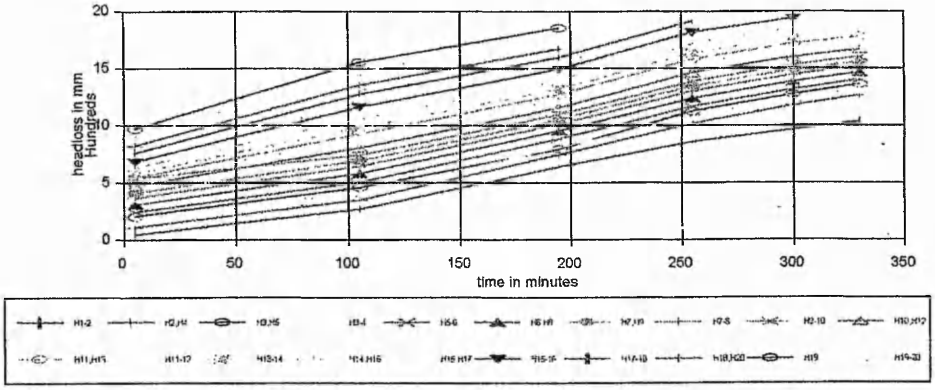
CONCENTRATION RATIO Vs TIME

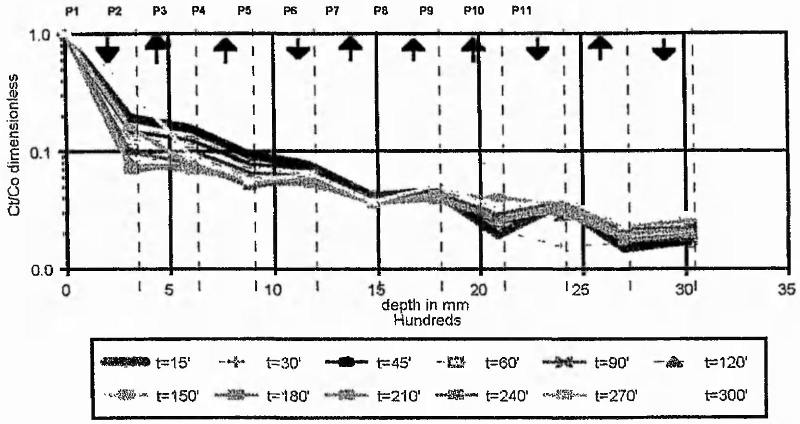
PVC Co=400 mg/l Q=0.8 l/min RUN 2



HEADLOSS Vs TIME

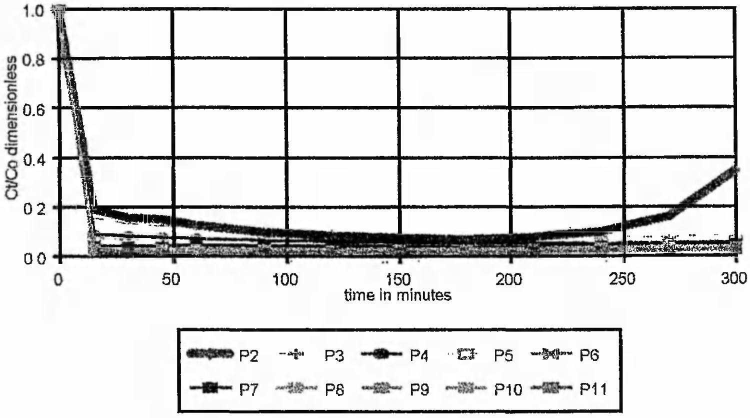
RUN No 2 PVC Co=400 mg/l Q=0.8 l/min





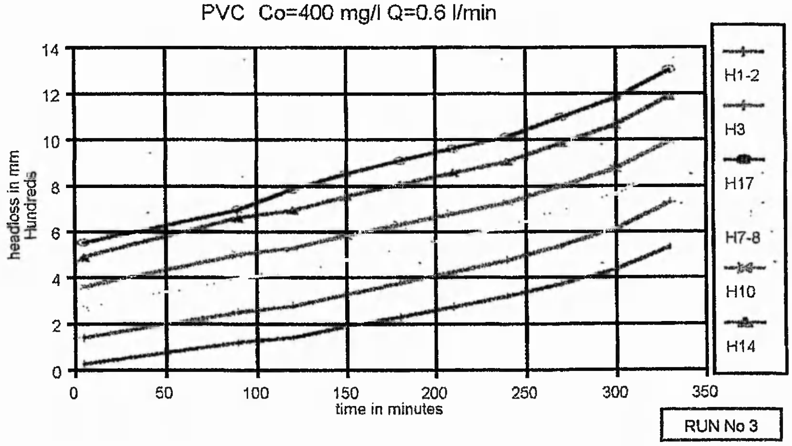
CONCENTRATION RATIO Vs DEPTH

RUN No 3 PVC Co=400 mg/l Q=0.6 l/min



CONCENTRATION RATIO Vs TIME

RUN No 3 PVC Co=400 mg/l Q=0.6 l/min

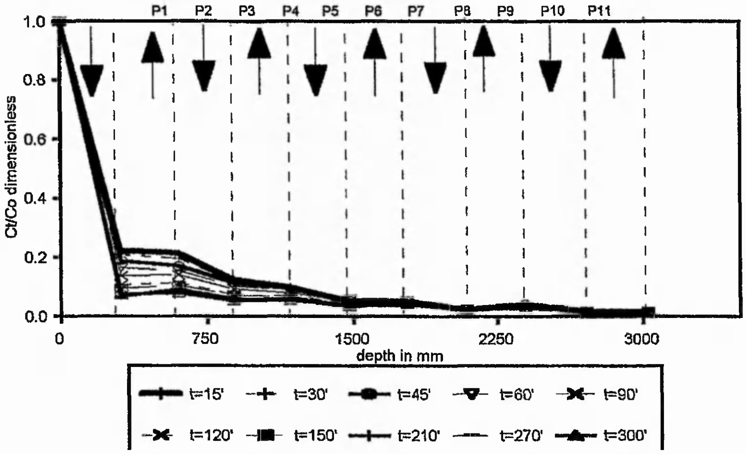


HEADLOSS Vs TIME

PVC Co=400 mg/l Q=0.6 l/min

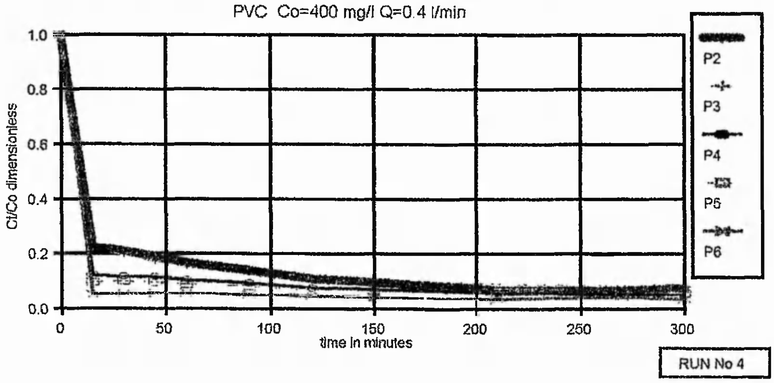
CONCENTRATION RATIO Vs DEPTH

PVC Co=400 mg/l Q=0.4 l/min RUN 4



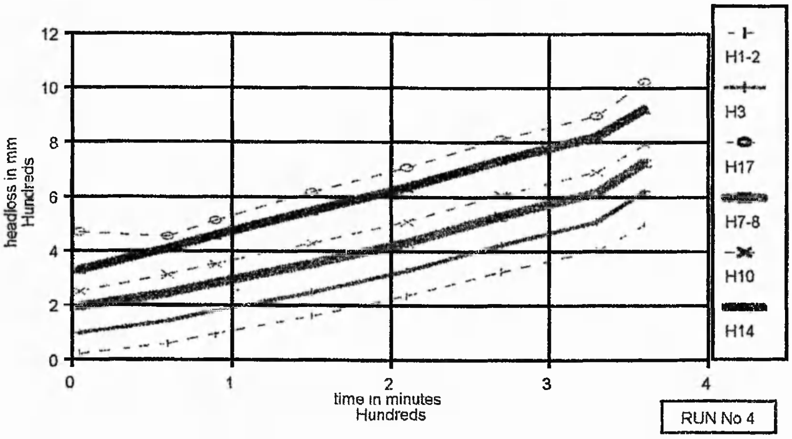
CONCENTRATION RATIO Vs TIME

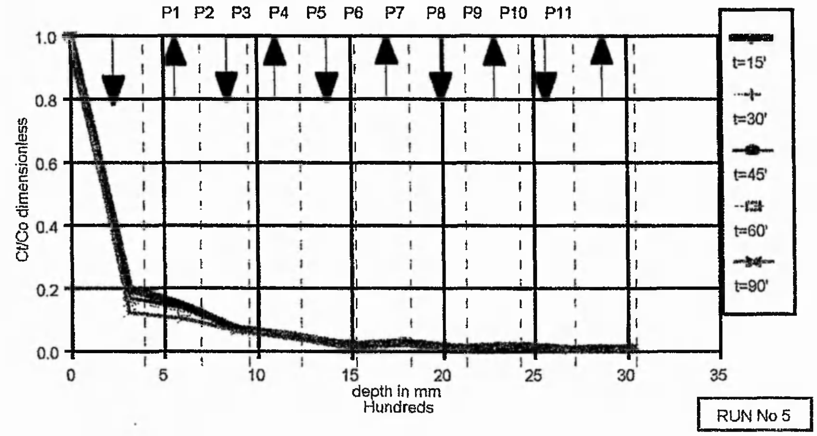
PVC Co=400 mg/l Q=0.4 l/min



HEADLOSS Vs TIME

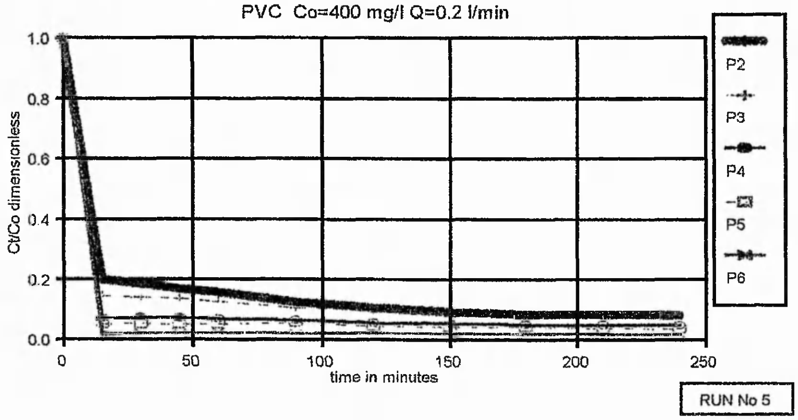
PVC Co=400 mg/l Q=0.4 l/min





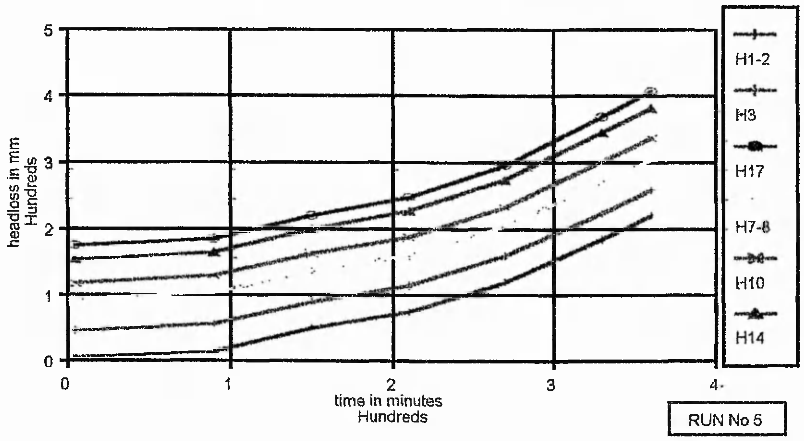
CONCENTRATION RATIO Vs DEPTH

PVC Co=400 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

PVC Co=400 mg/l Q=0.2 l/min

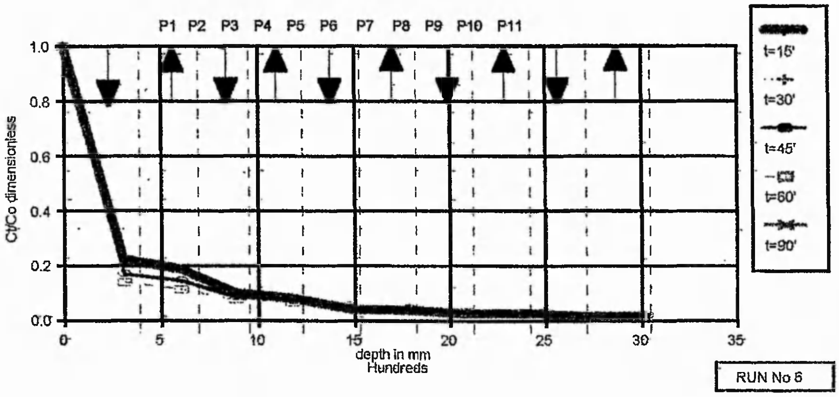


HEADLOSS Vs TIME

PVC Co=400 mg/l Q=0.2 l/min

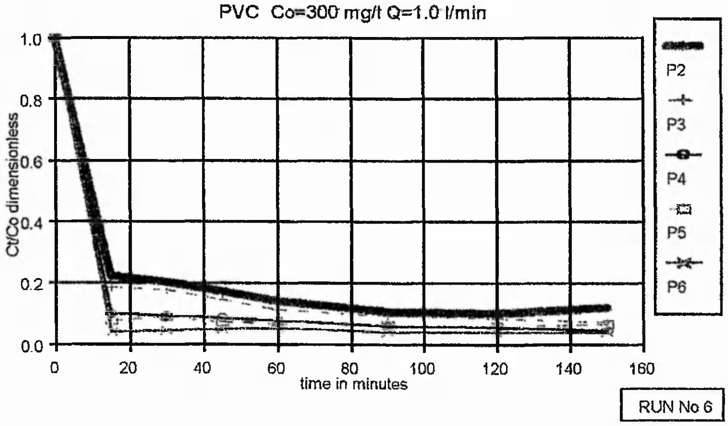
CONCENTRATION RATIO Vs DEPTH

PVC Co=300 mg/l Q=1.0 l/min



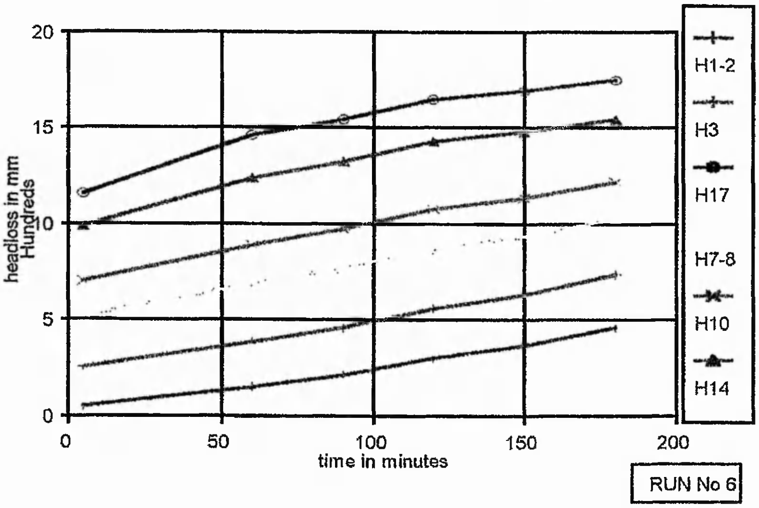
CONCENTRATION RATIO Vs TIME

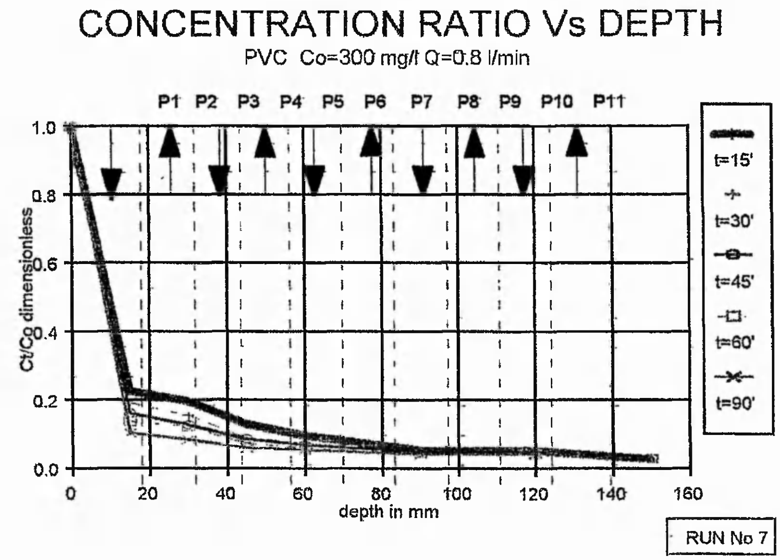
PVC Co=300 mg/l Q=1.0 l/min

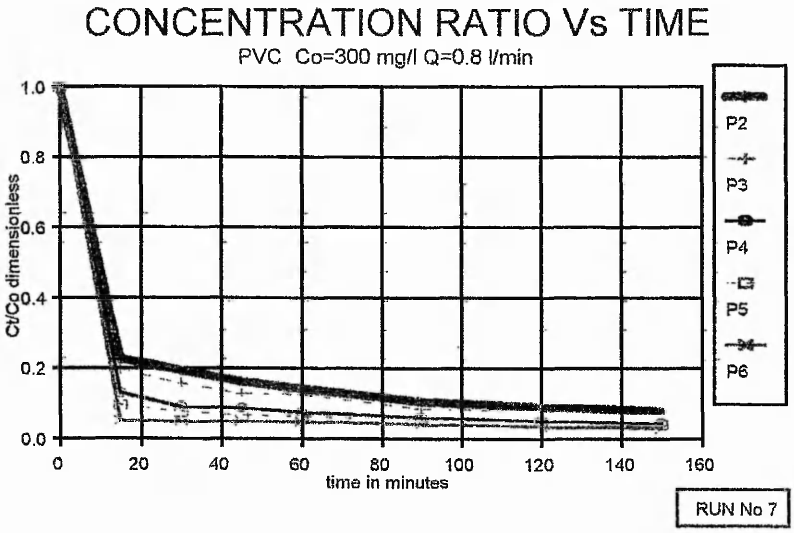


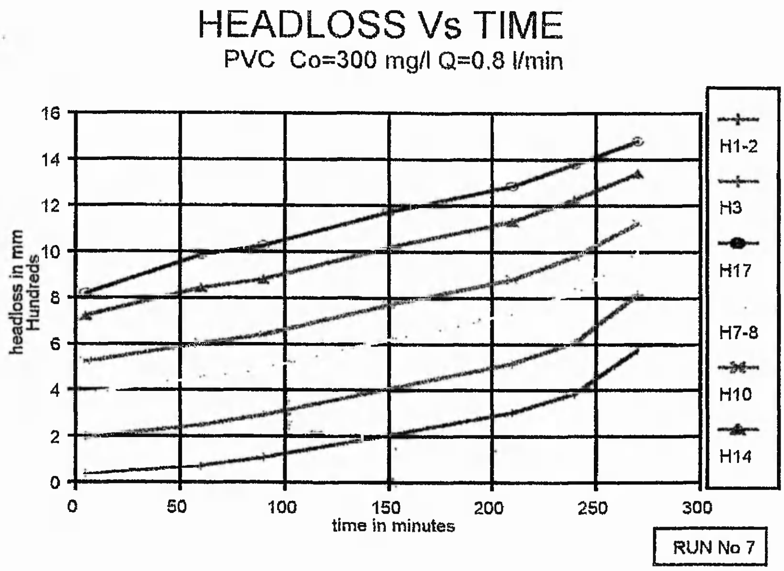
HEADLOSS Vs TIME

PVC Co=300 mg/l Q=1.0 l/min



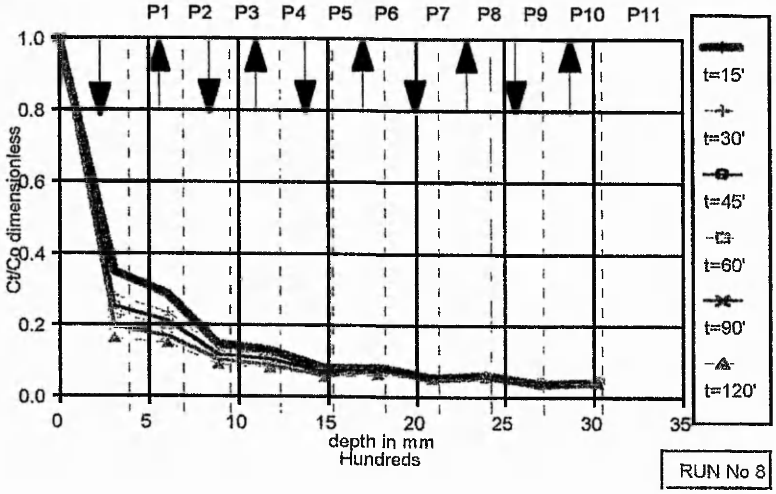

CONCENTRATION RATIO Vs DEPTH
PVC Co=300 mg/l Q=0.8 l/min
P1 P2 P3 P4 P5 P6 P7 P8 P9 P10 P11
Ct/Co dimensionless
1.0
0.8
0.6
0.4
0.2
0.0
0
20
40
60
80
100
120
140
160
depth in mm
t=15'
t=30'
t=45'
t=60'
t=90'
RUN No 7


CONCENTRATION RATIO Vs TIME
PVC Co=300 mg/l Q=0.8 l/min
Ct/Co dimensionless
1.0
0.8
0.6
0.4
0.2
0.0
0
20
40
60
80
100
120
140
160
time in minutes
P2
P3
P4
P5
P6
RUN No 7


HEADLOSS Vs TIME
PVC Co=300 mg/l Q=0.8 l/min
headloss in mm
Hundreds
16
14
12
10
8
6
4
2
0
0
50
100
150
200
250
300
time in minutes
H1-2
H3
H17
H7-8
H10
H14
RUN No 7

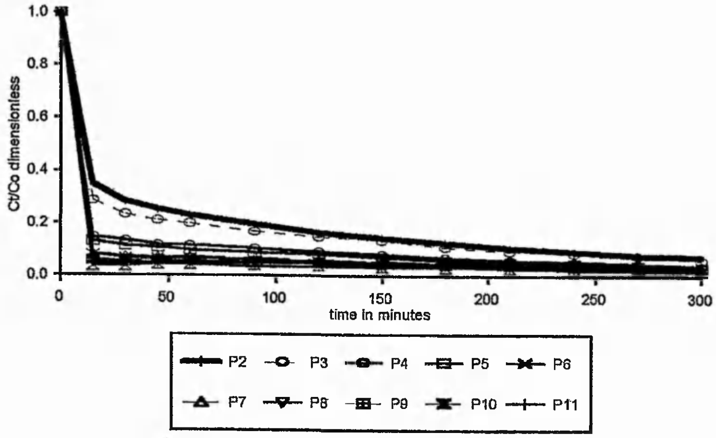
CONCENTRATION RATIO Vs DEPTH

PVC Co=300 mg/l Q=0.6 l/min



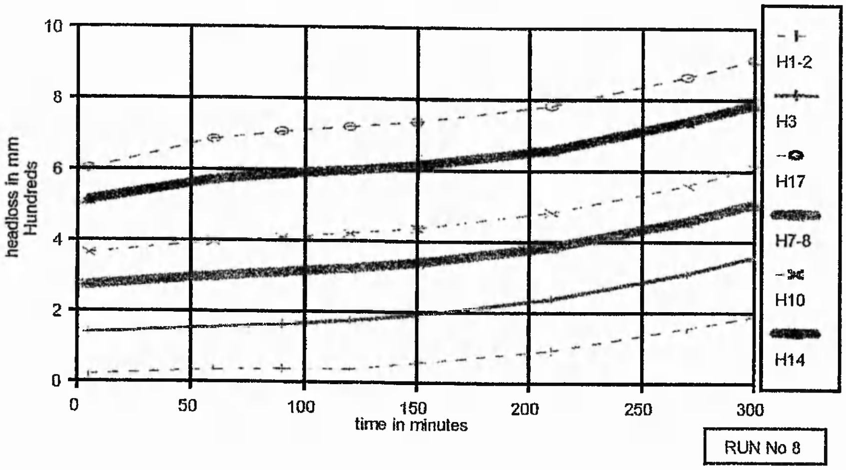
CONCENTRATION RATIO Vs TIME

PVC Co=300 mg/l Q=0.6 l/min RUN 8



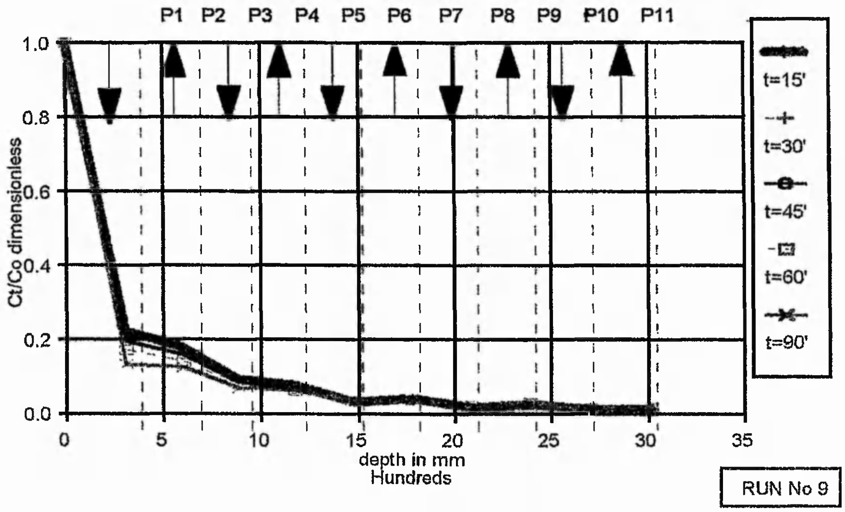
HEADLOSS Vs TIME

PVC Co=300 mg/l Q=0.6 l/min



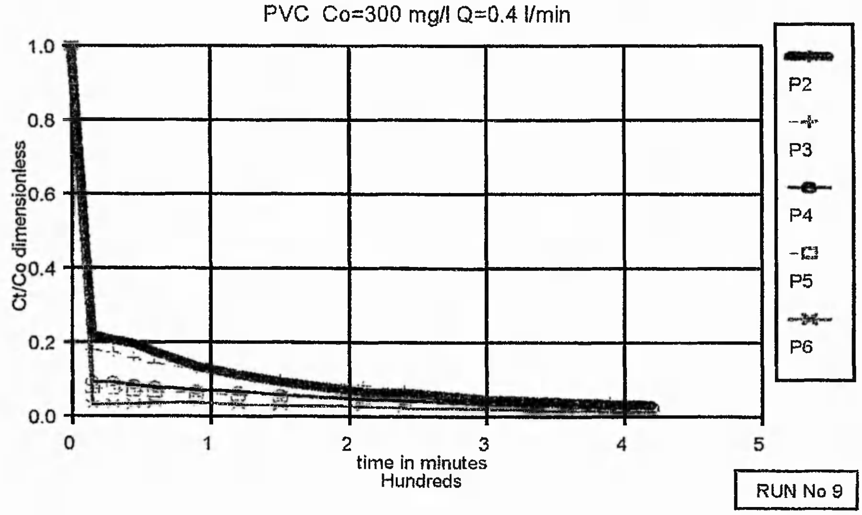
CONCENTRATION RATIO Vs DEPTH

PVC Co=300 mg/l Q=0,4 l/min



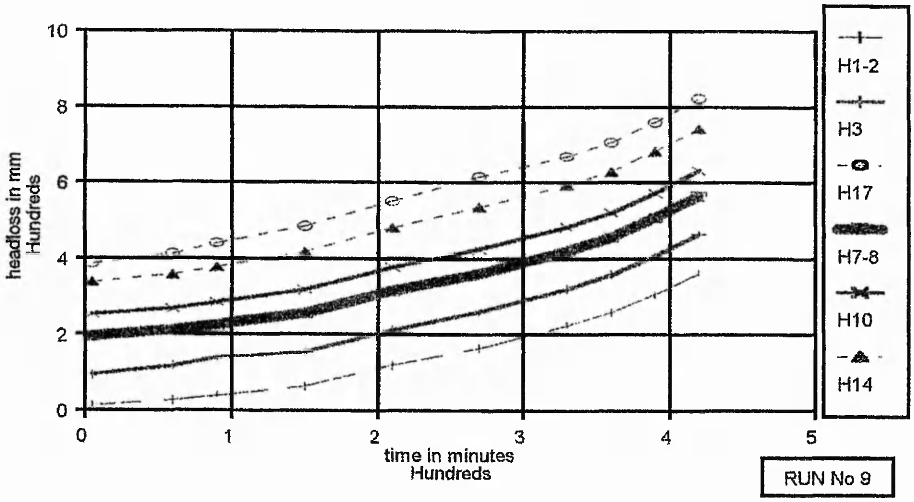
CONCENTRATION RATIO Vs TIME

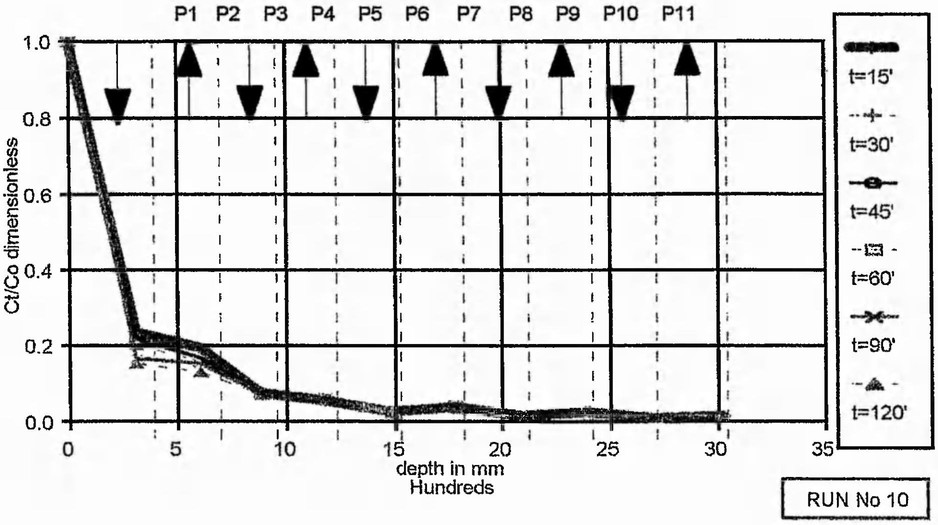
PVC Co=300 mg/l Q=0.4 l/min



HEADLOSS Vs TIME

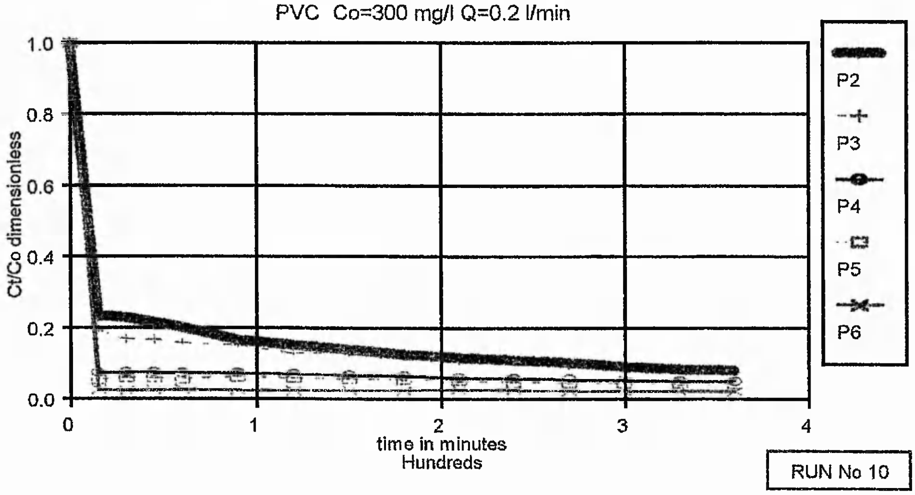
PVC Co=300 mg/l Q=0.4 l/min





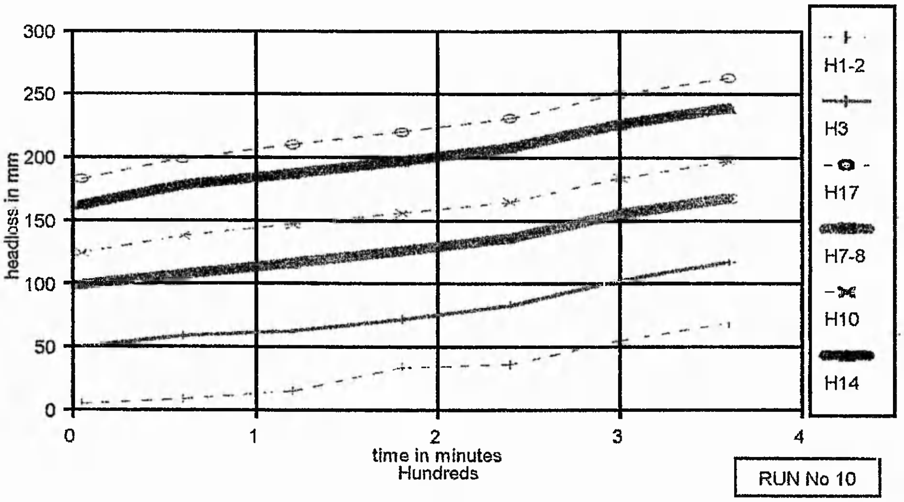
CONCENTRATION RATIO Vs DEPTH

PVC Co=300 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

PVC Co=300 mg/l Q=0.2 l/min

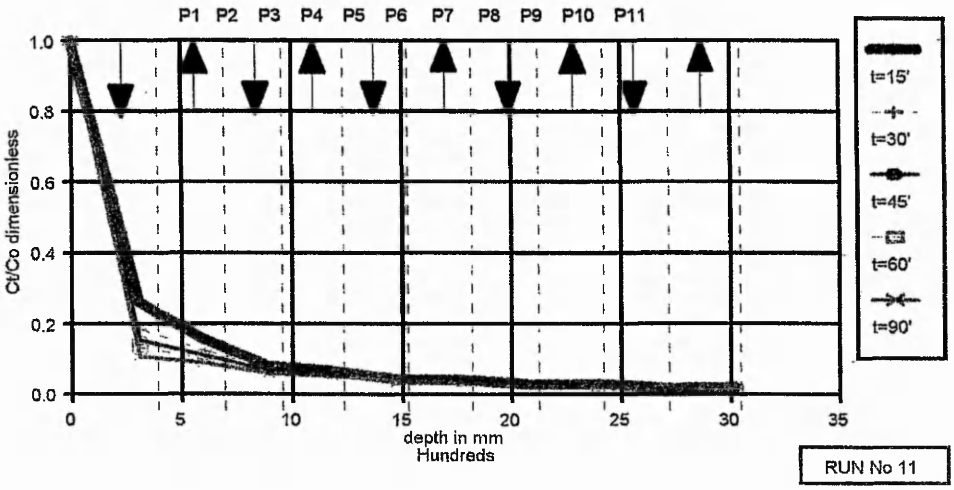


HEADLOSS Vs TIME

PVC Co=300 mg/l Q=0.2 l/min

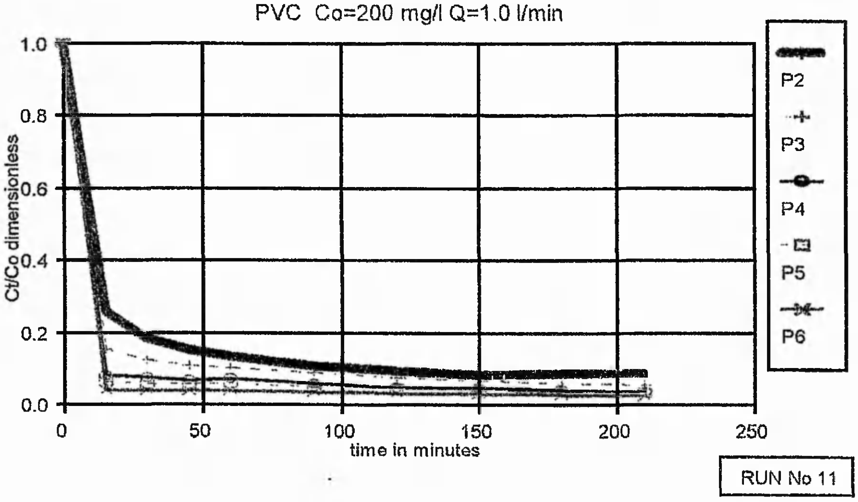
CONCENTRATION RATIO Vs DEPTH

PVC Co=200 mg/l Q=1.0 l/min



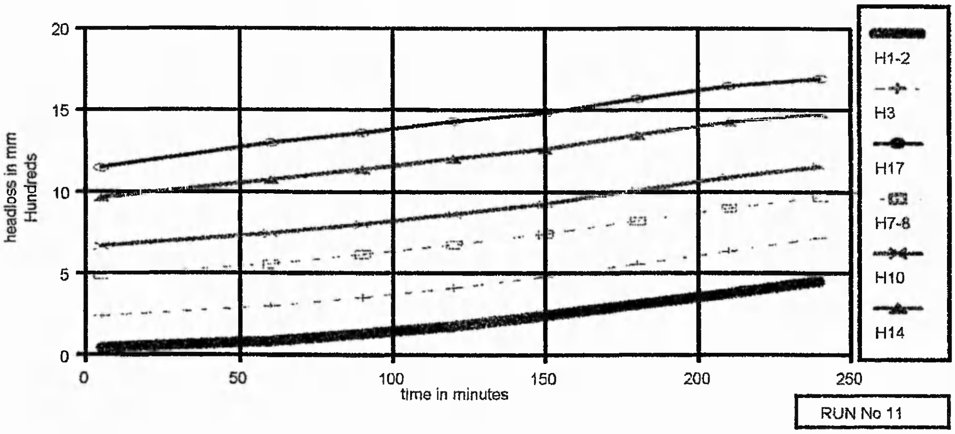
CONCENTRATION RATIO Vs TIME

PVC Co=200 mg/l Q=1.0 l/min



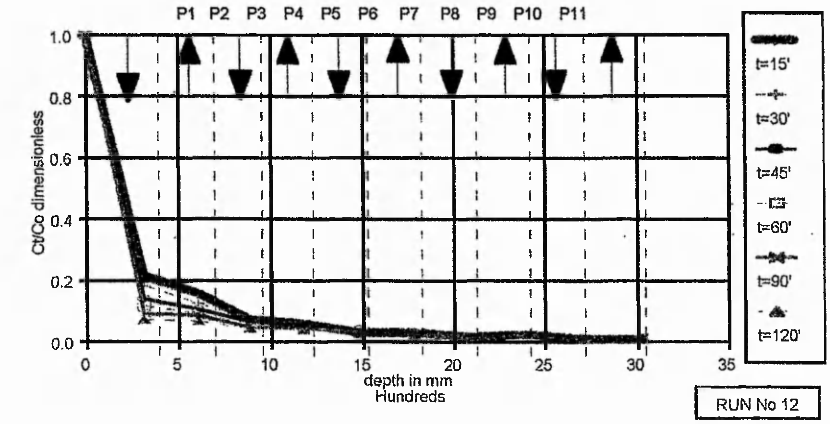
HEADLOSS Vs TIME

PVC Co=200 mg/l Q=1.0 l/min



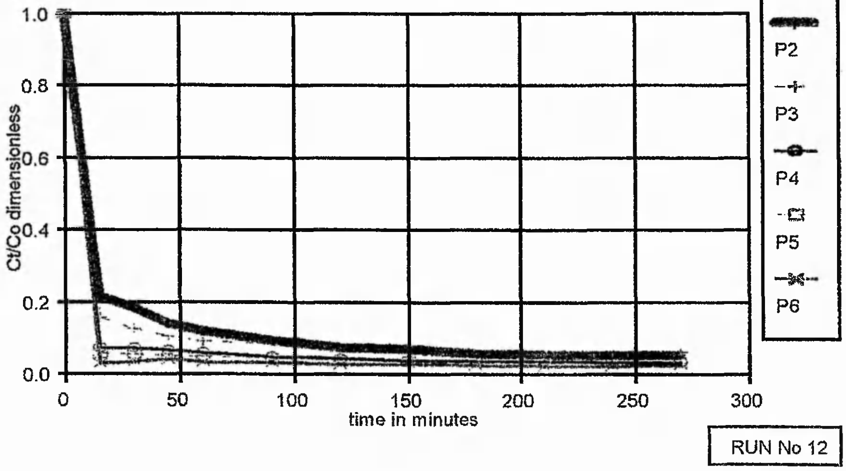
CONCENTRATION RATIO Vs DEPTH

PVC Co=200 mg/l Q=0.8 l/min



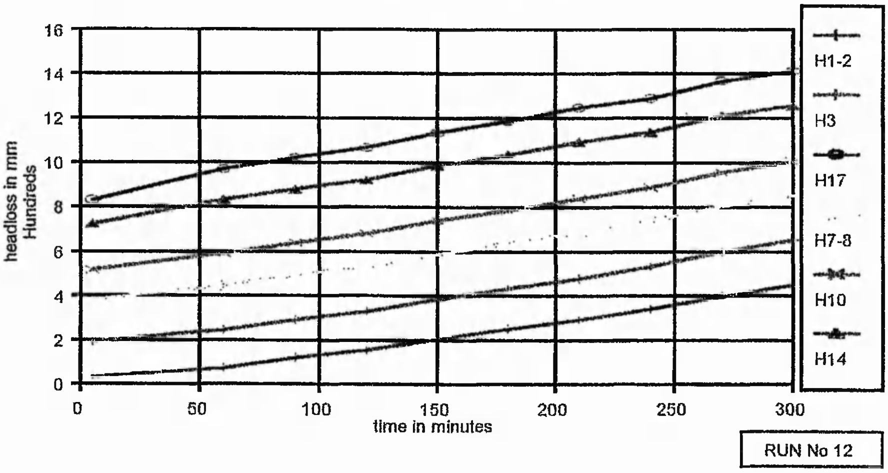
CONCENTRATION RATIO Vs TIME

PVC Co=200 mg/l Q=0.8 l/min



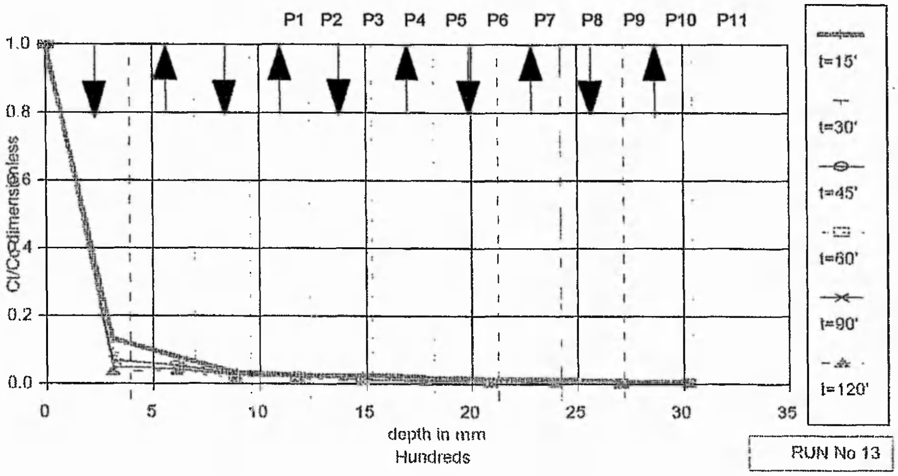
HEADLOSS Vs TIME

PVC Co=200 mg/l Q=0.8 l/min



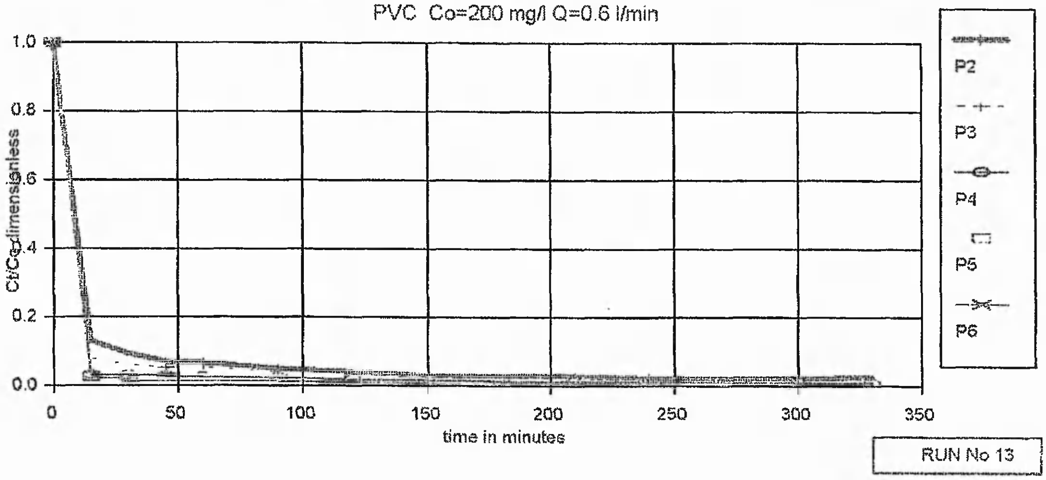
CONCENTRATION RATIO Vs DEPTH

PVC Co=200 mg/l Q=0.6 l/min



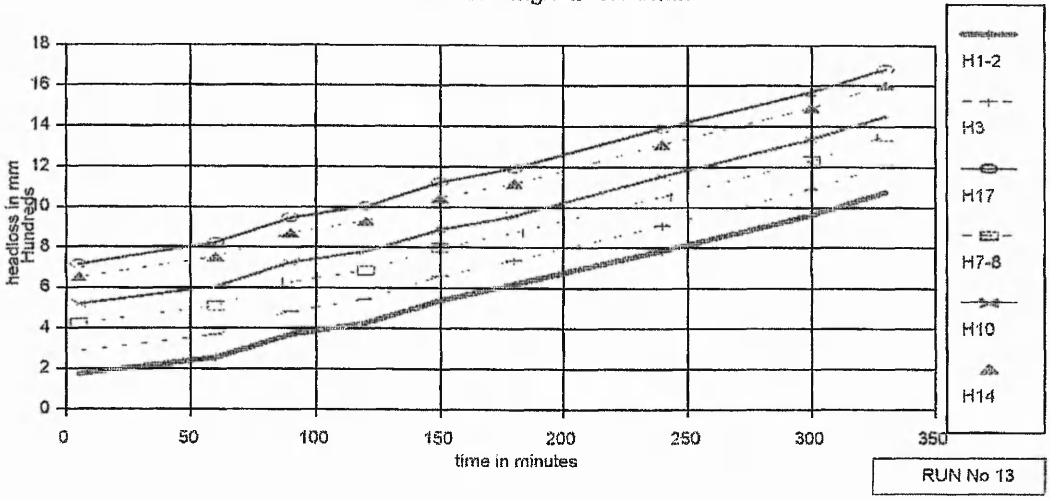
CONCENTRATION RATIO Vs TIME

PVC Co=200 mg/l Q=0.6 l/min



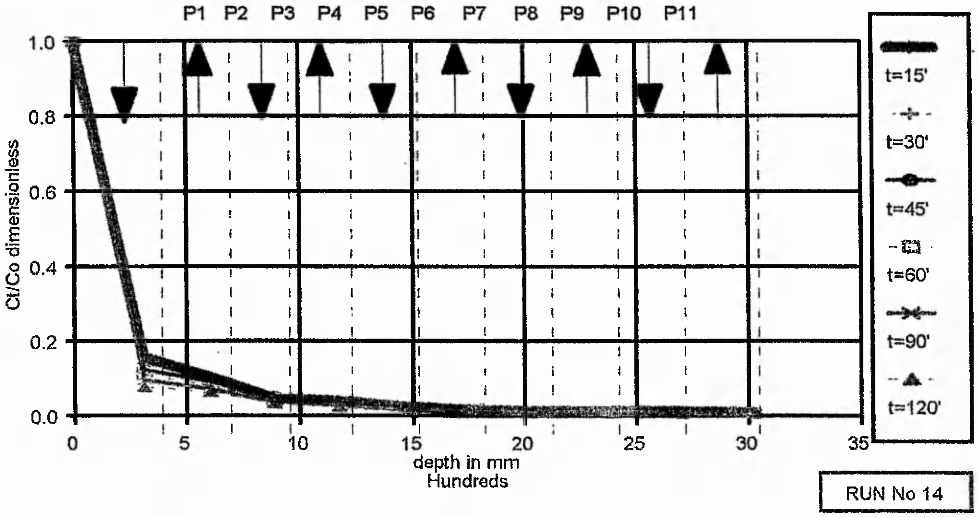
HEADLOSS Vs TIME

PVC Co=200 mg/l Q=0.6 l/min



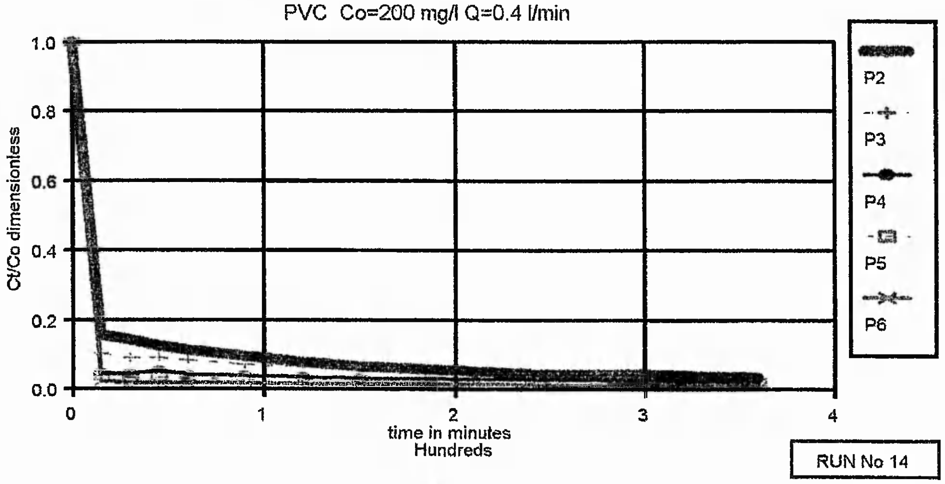
CONCENTRATION RATIO Vs DEPTH

PVC Co=200 mg/l Q=0.4 l/min



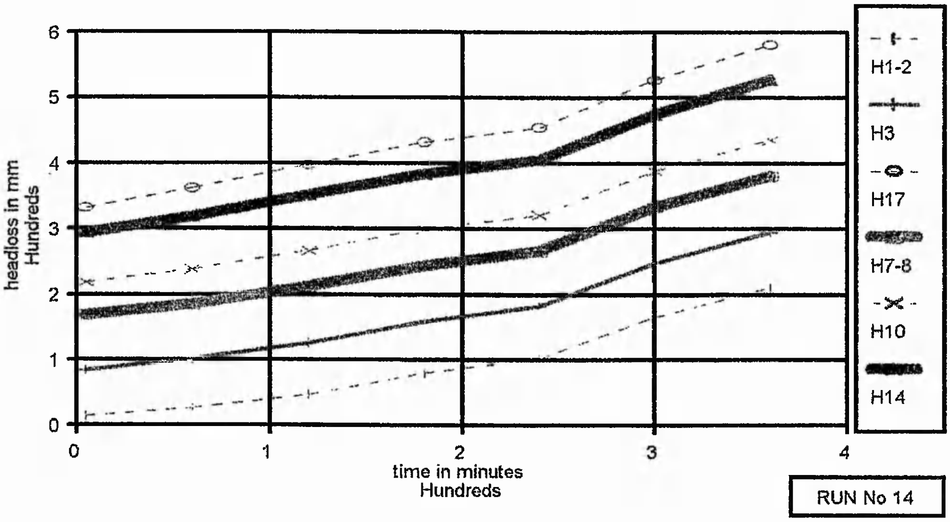
CONCENTRATION RATIO Vs TIME

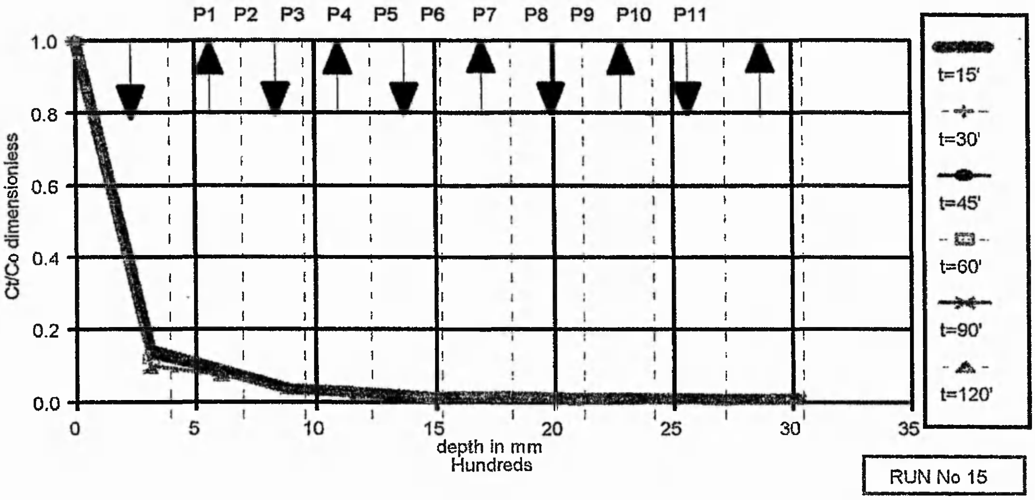
PVC Co=200 mg/l Q=0.4 l/min



HEADLOSS Vs TIME

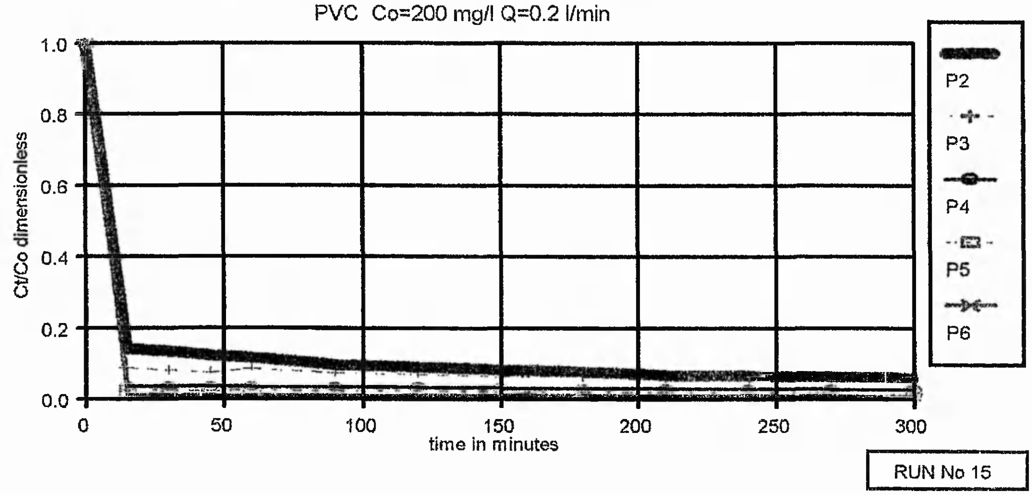
PVC Co=200 mg/l Q=0.4 l/min





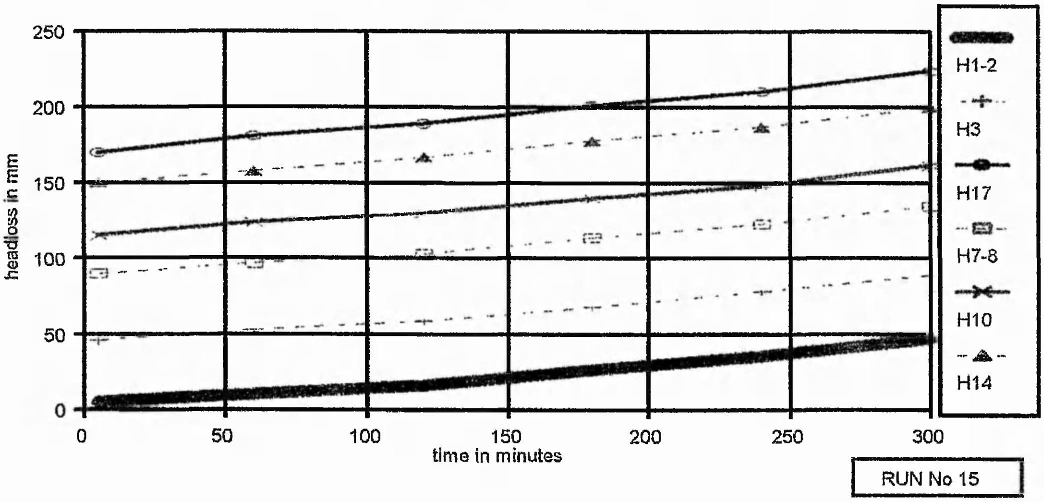
CONCENTRATION RATIO Vs DEPTH

PVC Co=200 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

PVC Co=200 mg/l Q=0.2 l/min

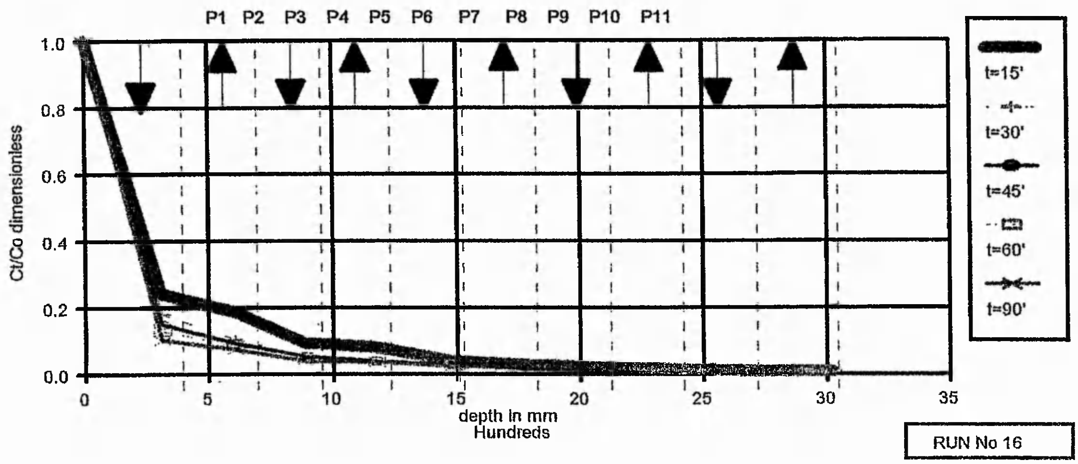


HEADLOSS Vs TIME

PVC Co=200 mg/l Q=0.2 l/min

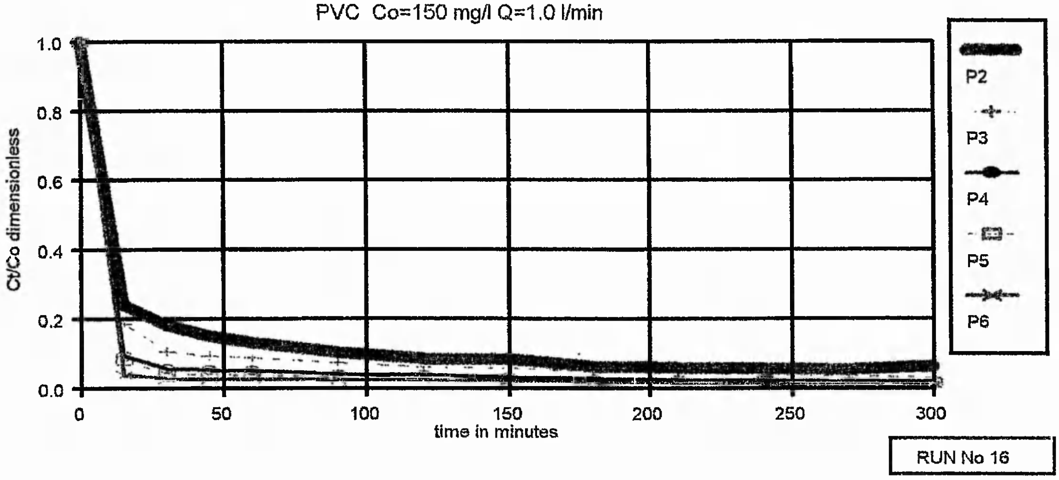
CONCENTRATION RATIO Vs DEPTH

PVC Co=150 mg/l Q=1.0 l/min



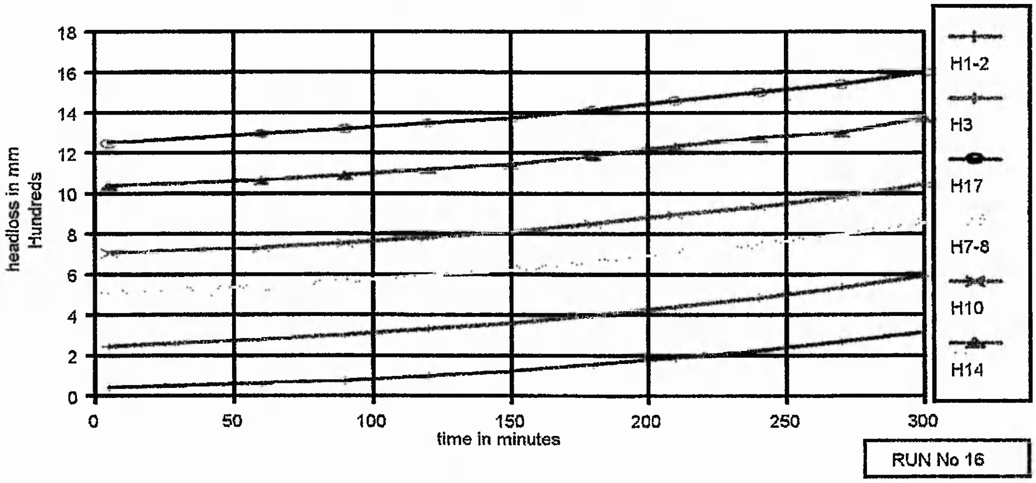
CONCENTRATION RATIO Vs TIME

PVC Co=150 mg/l Q=1.0 l/min



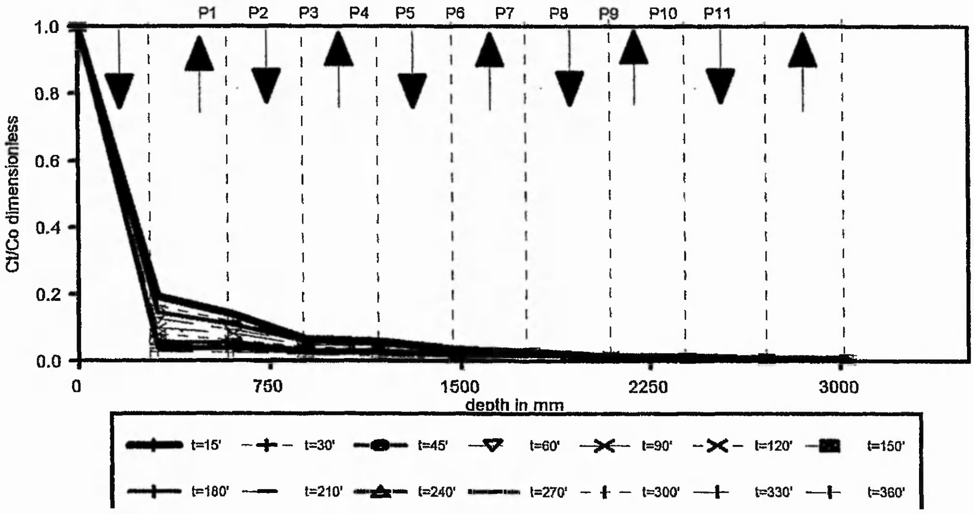
HEADLOSS Vs TIME

PVC Co=150 mg/l Q=1.0 l/min



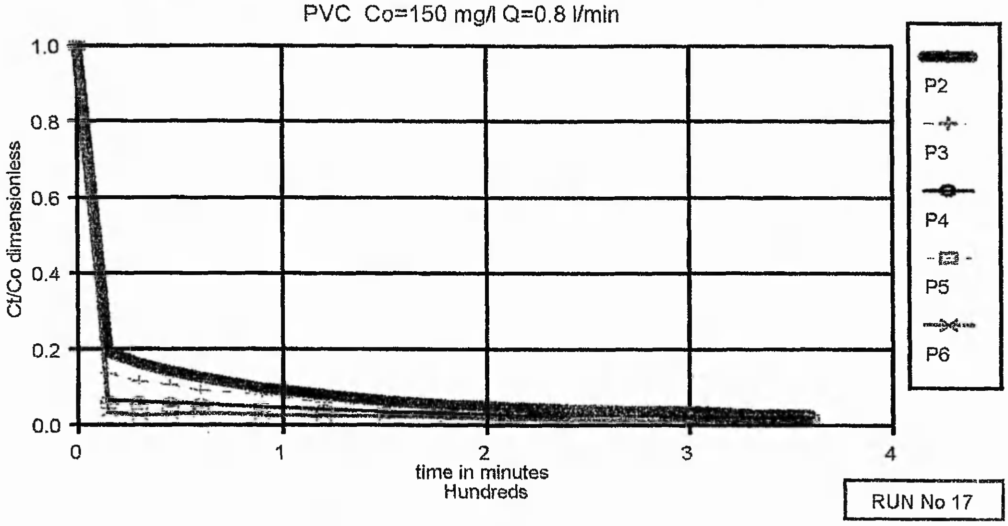
CONCENTRATION RATIO Vs DEPTH

PVC Co=150 mg/l Q=0.8 l/min RUN 17



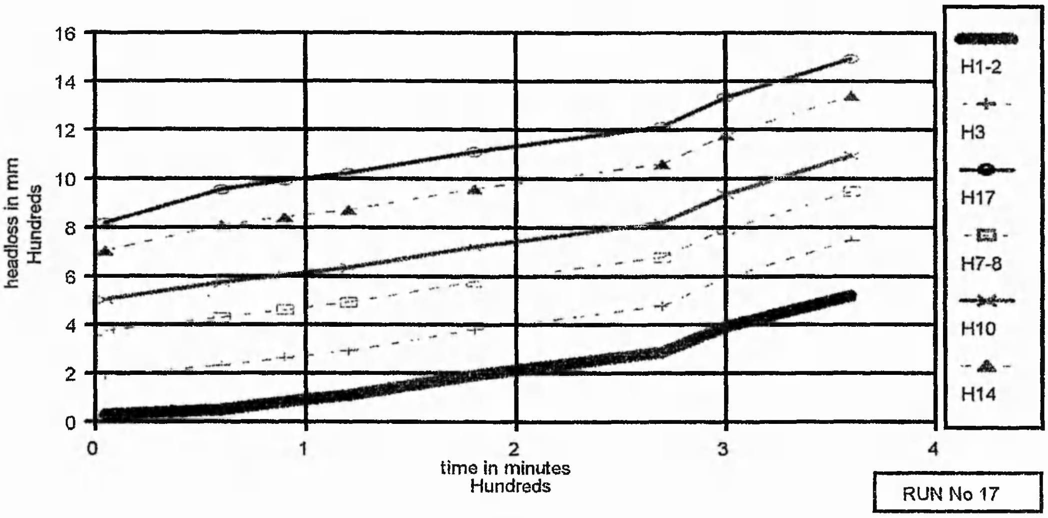
CONCENTRATION RATIO Vs TIME

PVC Co=150 mg/l Q=0.8 l/min



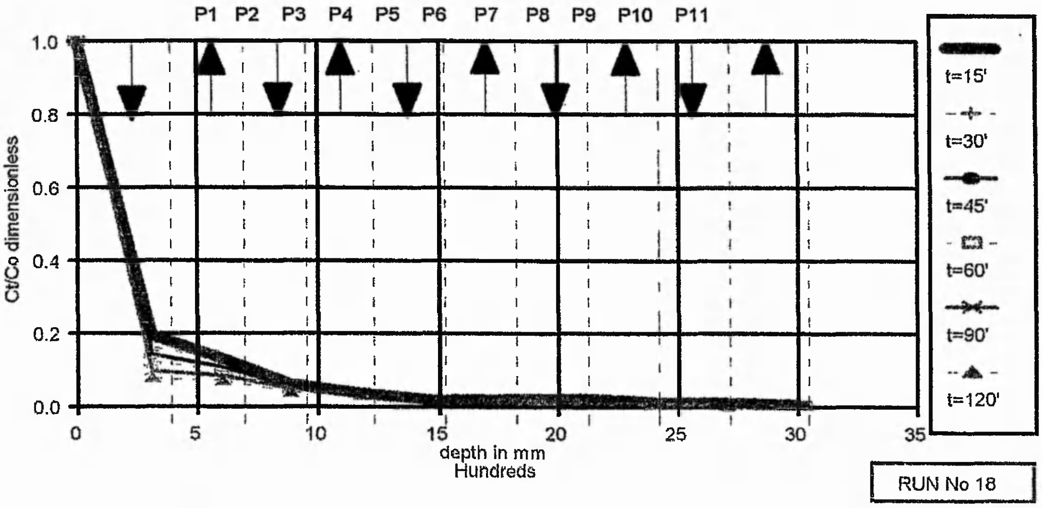
HEADLOSS Vs TIME

PVC Co=150 mg/l Q=0.8 l/min



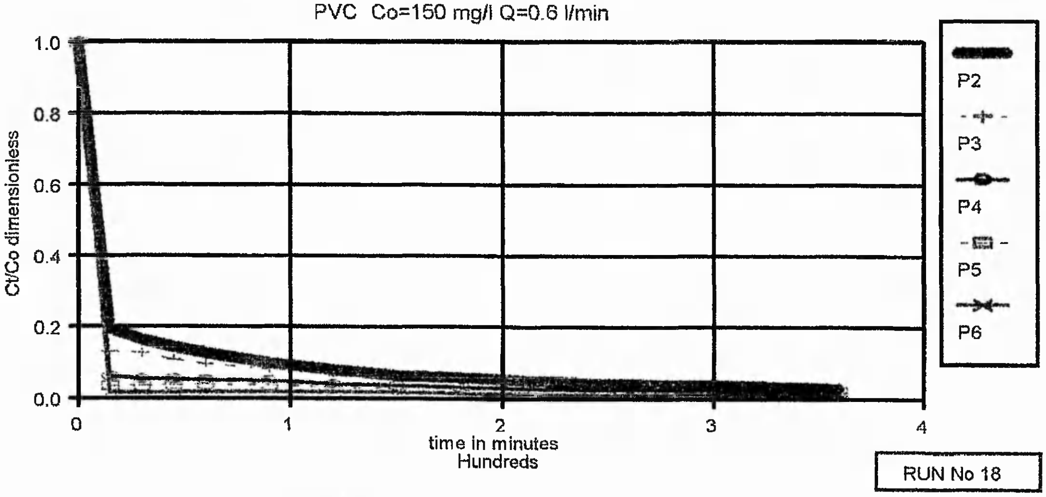
CONCENTRATION RATIO Vs DEPTH

PVC Co=150 mg/l Q=0.6 l/min



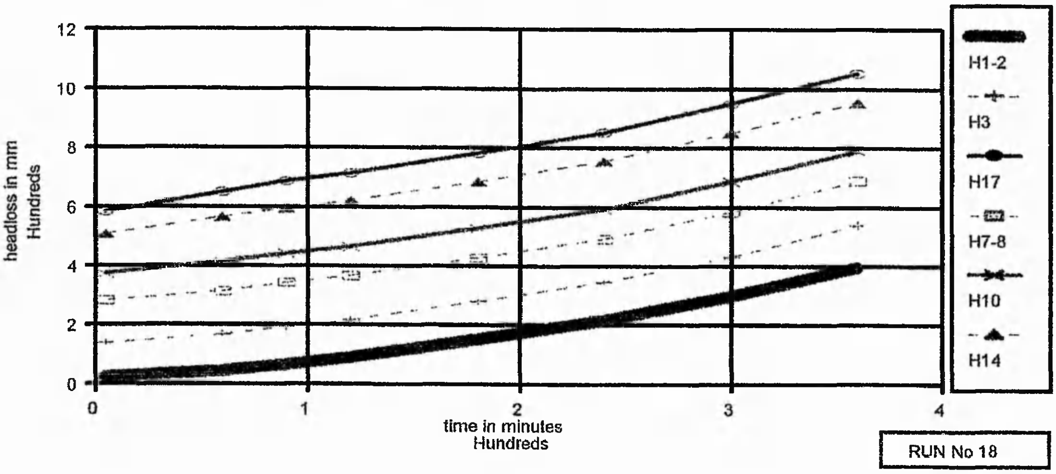
CONCENTRATION RATIO Vs TIME

PVC Co=150 mg/l Q=0.6 l/min



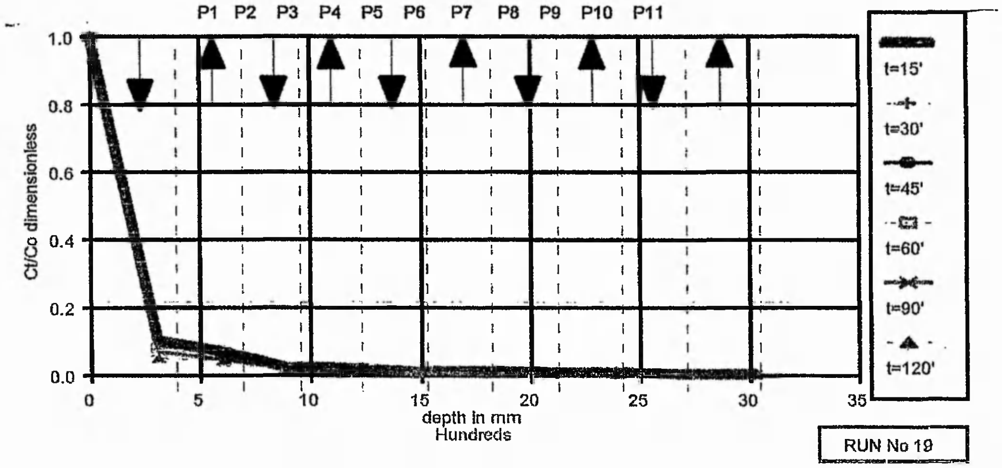
HEADLOSS Vs TIME

PVC Co=150 mg/l Q=0.6 l/min



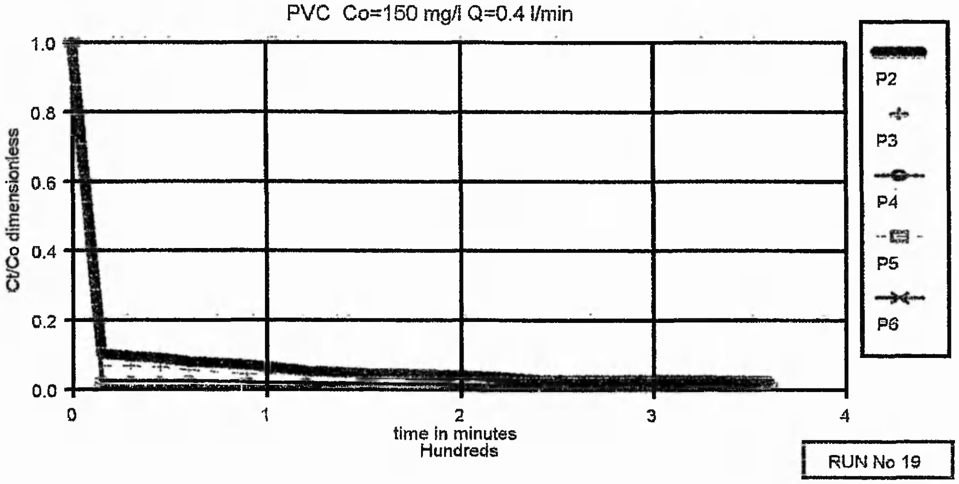
CONCENTRATION RATIO Vs DEPTH

PVC Co=150 mg/l Q=0.4 l/min



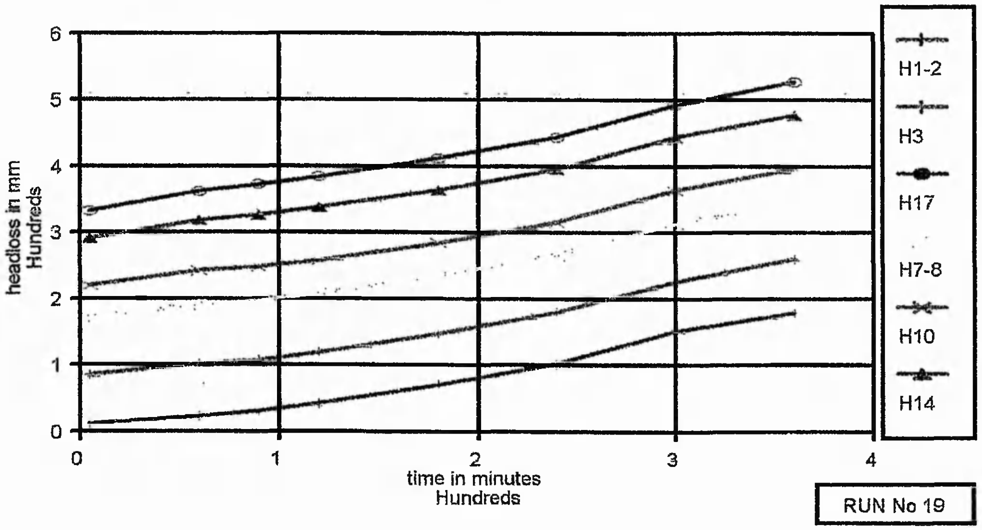
CONCENTRATION RATIO Vs TIME

PVC Co=150 mg/l Q=0.4 l/min



HEADLOSS Vs TIME

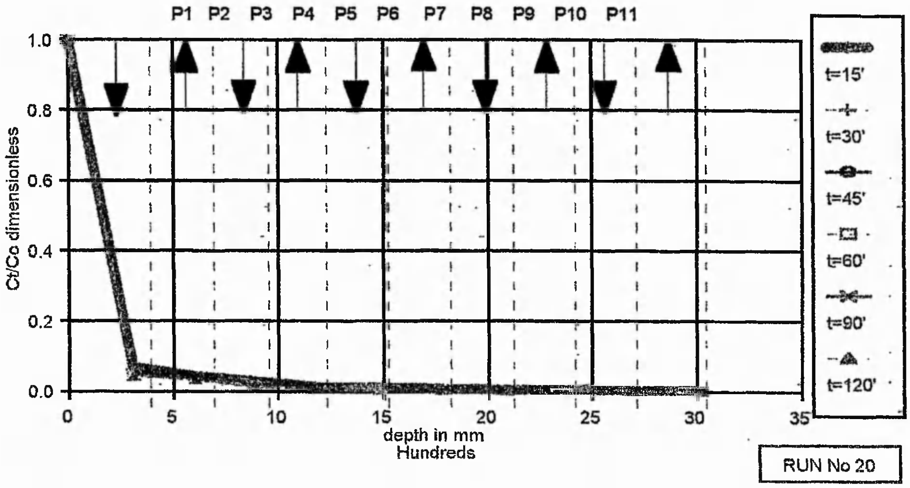
PVC Co=150 mg/l Q=0.4 l/min



Appendix 2 Experimental Data Sample

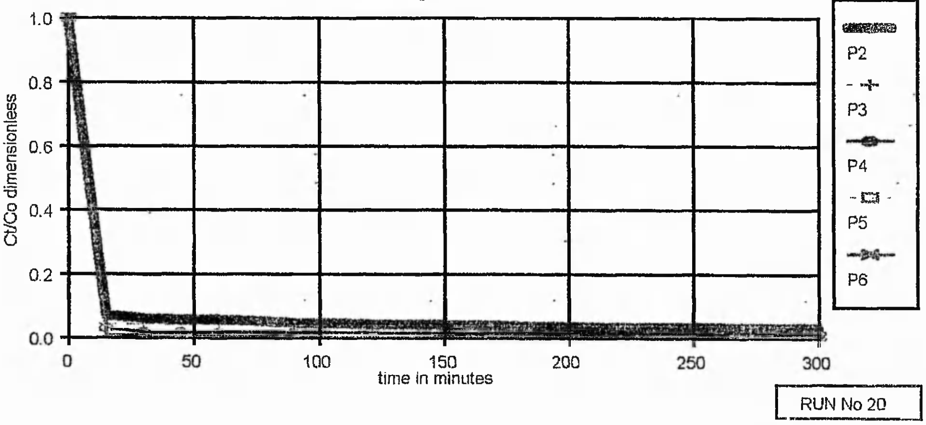
CONCENTRATION RATIO Vs DEPTH

PVC Co=150 mg/l Q=0.2 l/min



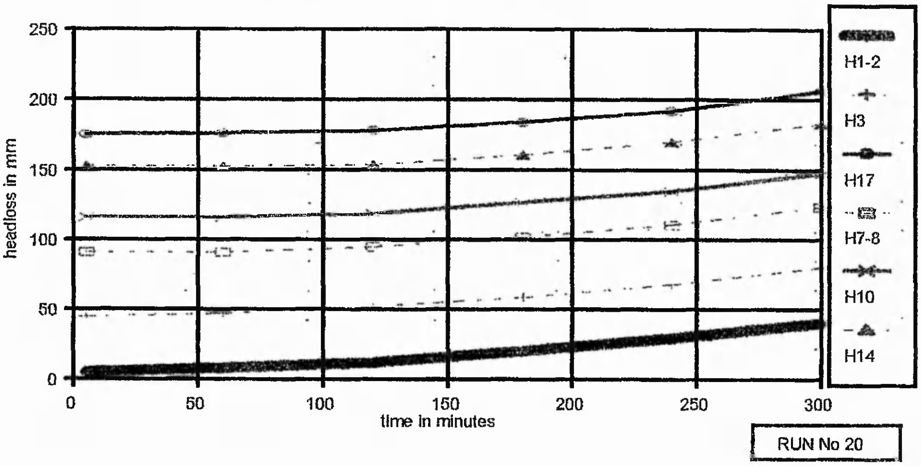
CONCENTRATION RATIO Vs TIME

PVC Co=150 mg/l Q=0.2 l/min



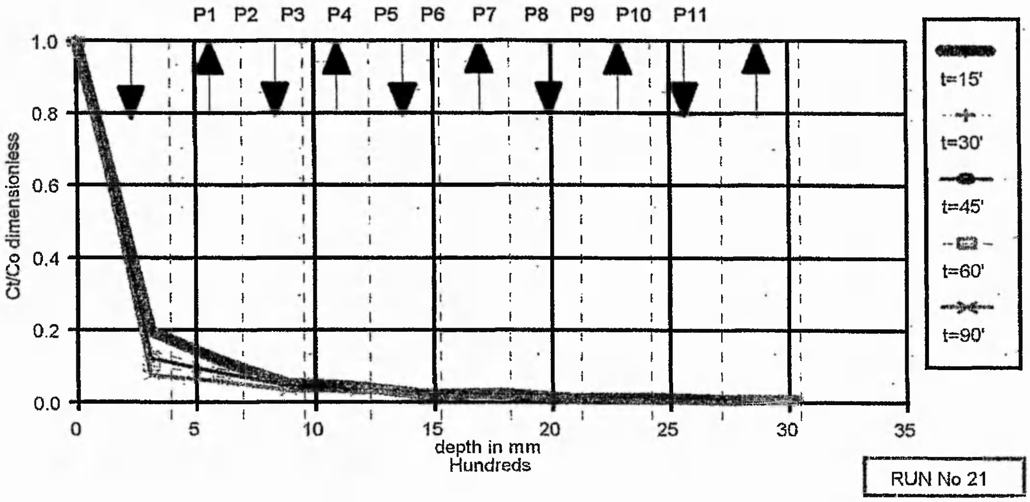
HEADLOSS Vs TIME

PVC Co=150 mg/l Q=0.2 l/min



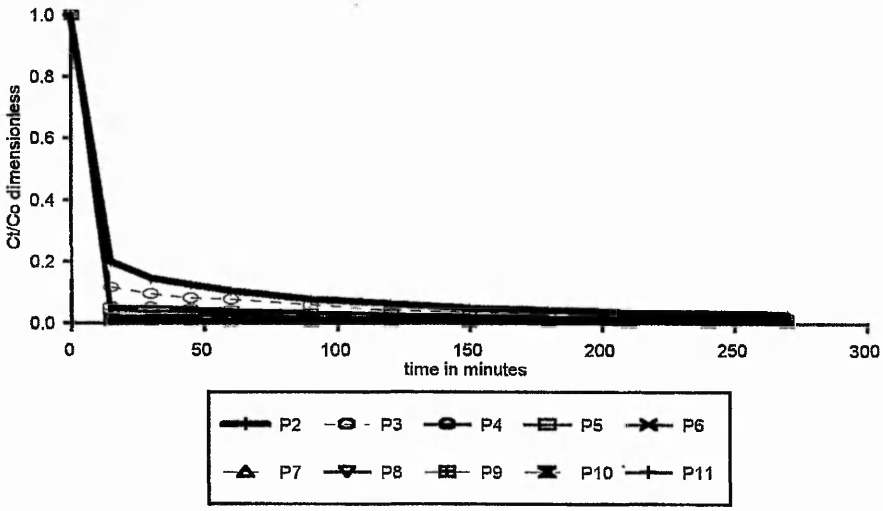
CONCENTRATION RATIO Vs DEPTH

PVC Co=100 mg/l Q=1.0 l/min



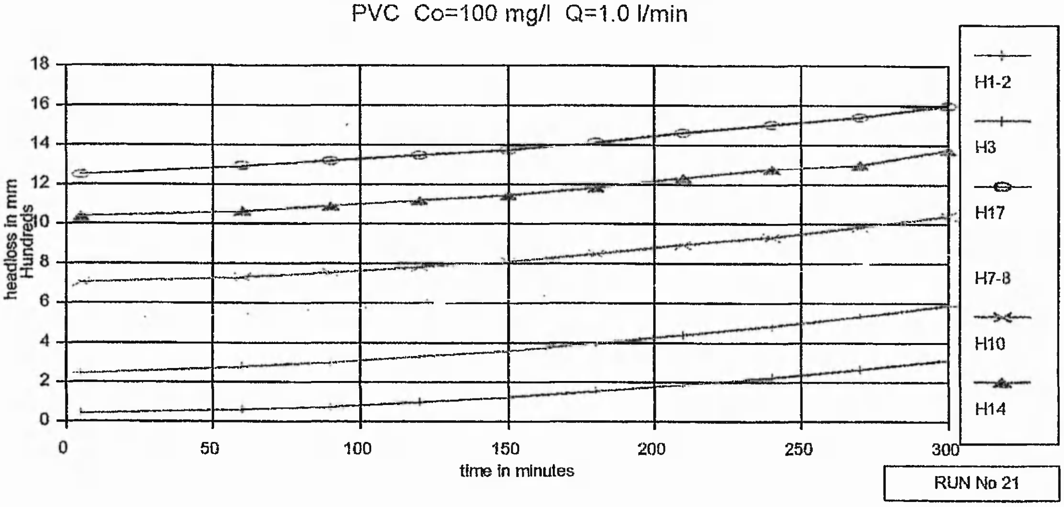
CONCENTRATION RATIO Vs TIME

PVC Co=100 mg/l Q=1.0 l/min RUN 21



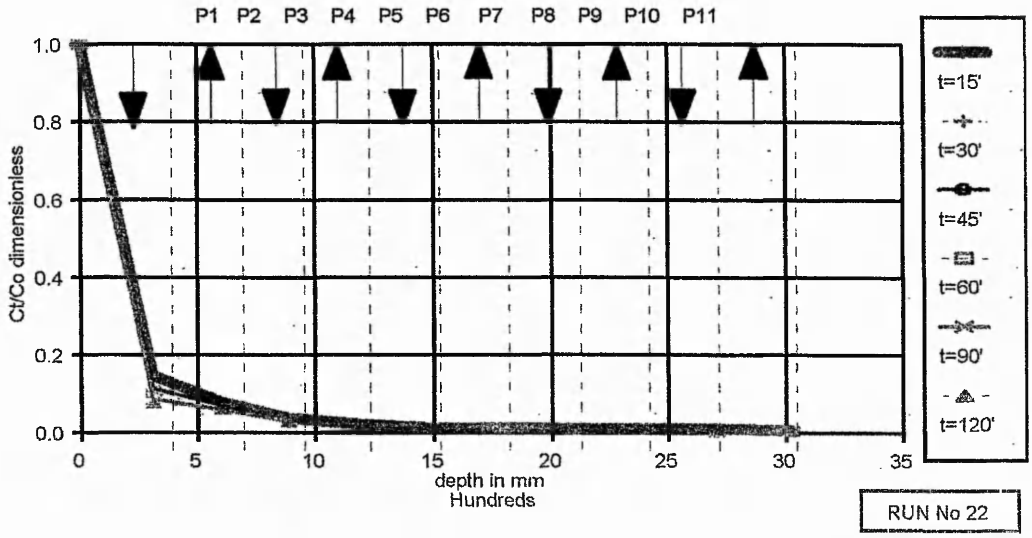
HEADLOSS Vs TIME

PVC Co=100 mg/l Q=1.0 l/min



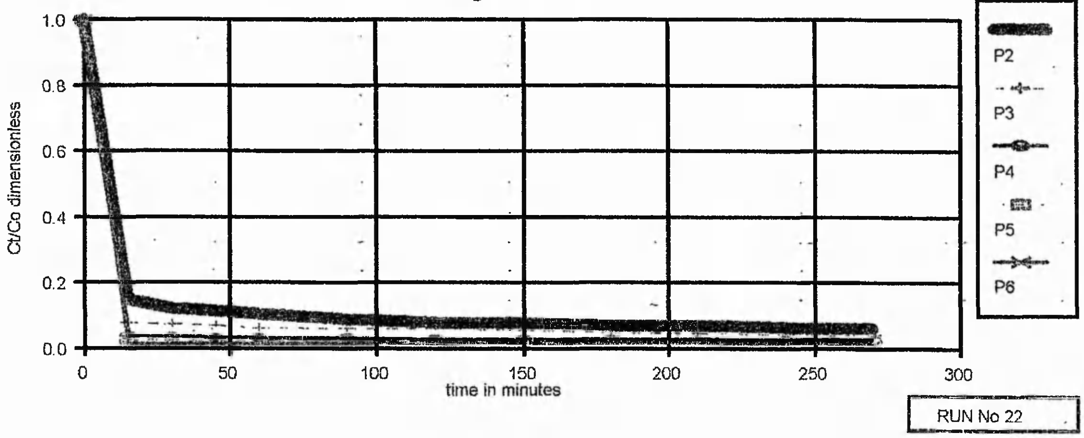
CONCENTRATION RATIO Vs DEPTH

PVC Co=100 mg/l Q=0.8 l/min



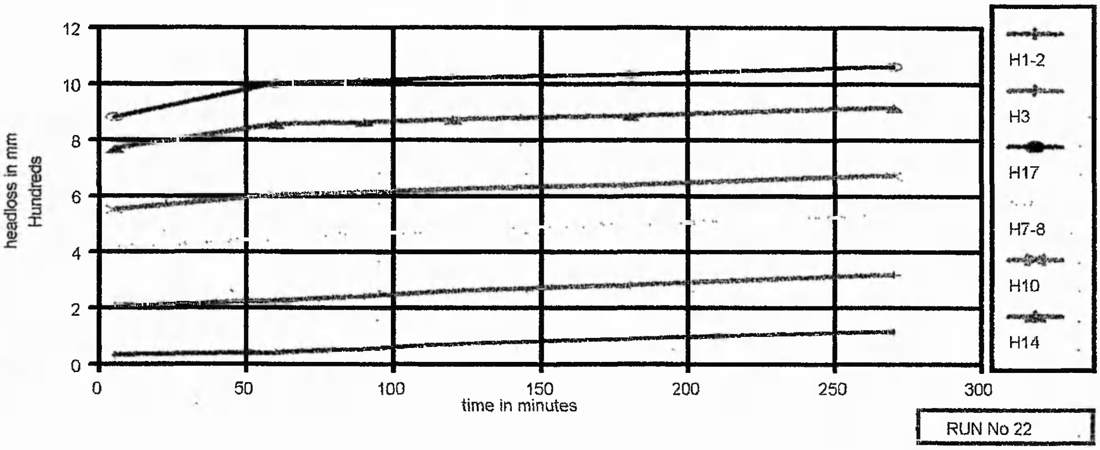
CONCENTRATION RATIO Vs TIME

PVC Co=100 mg/l Q=0.8 l/min



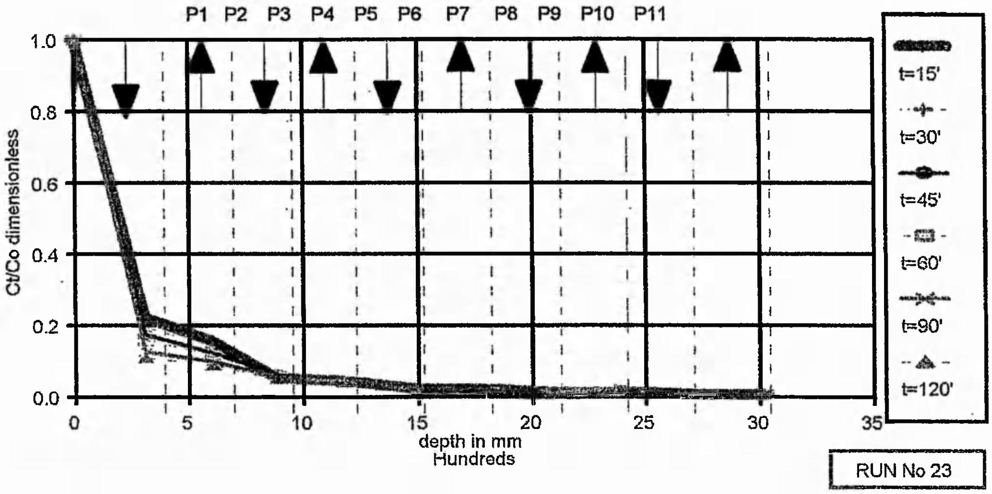
HEADLOSS Vs TIME

PVC Co=100 mg/l Q=0.8 l/min



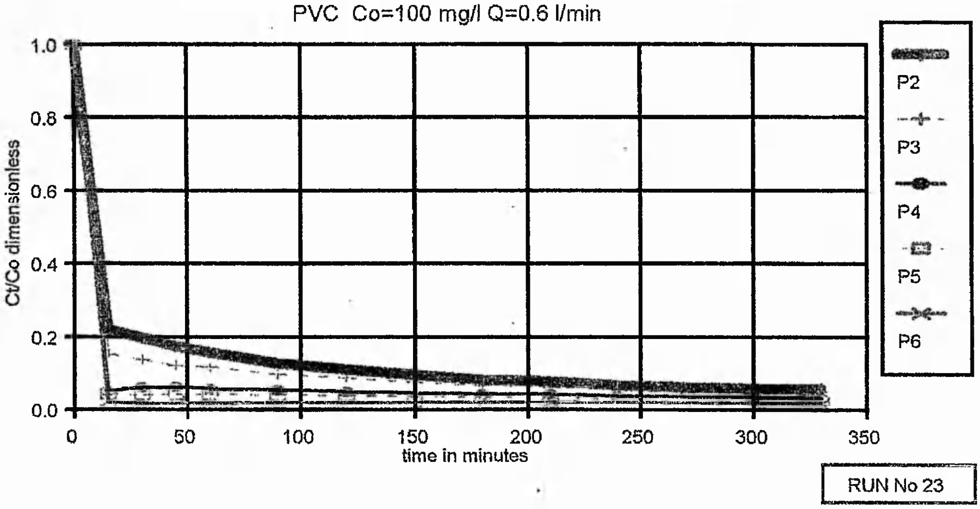
CONCENTRATION RATIO Vs DEPTH

PVC Co=100 mg/l Q=0.6 l/min



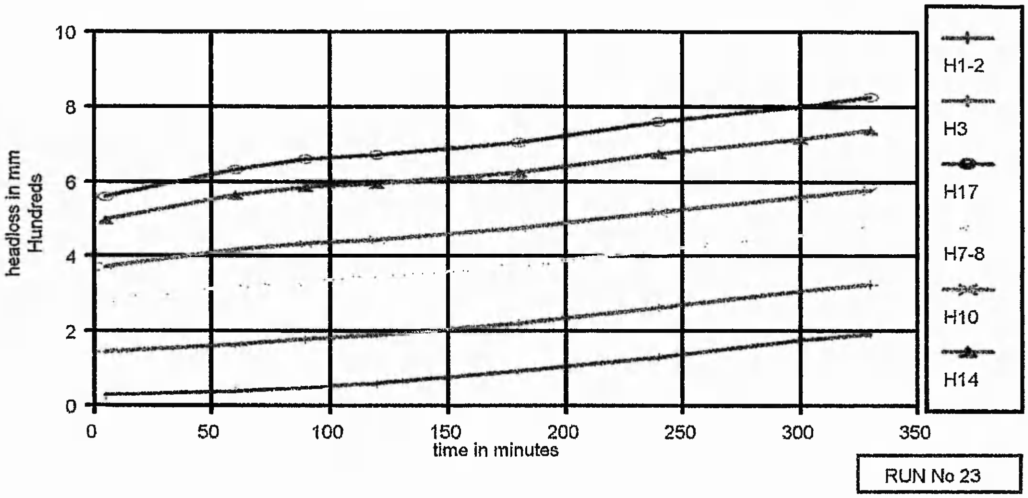
CONCENTRATION RATIO Vs TIME

PVC Co=100 mg/l Q=0.6 l/min



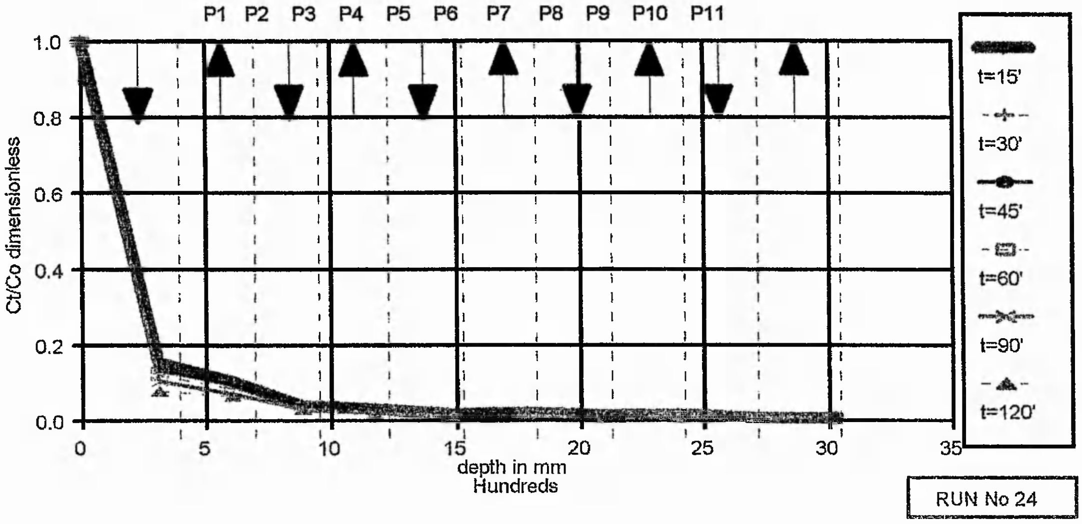
HEADLOSS Vs TIME

PVC Co=100 mg/l Q=0.6 l/min



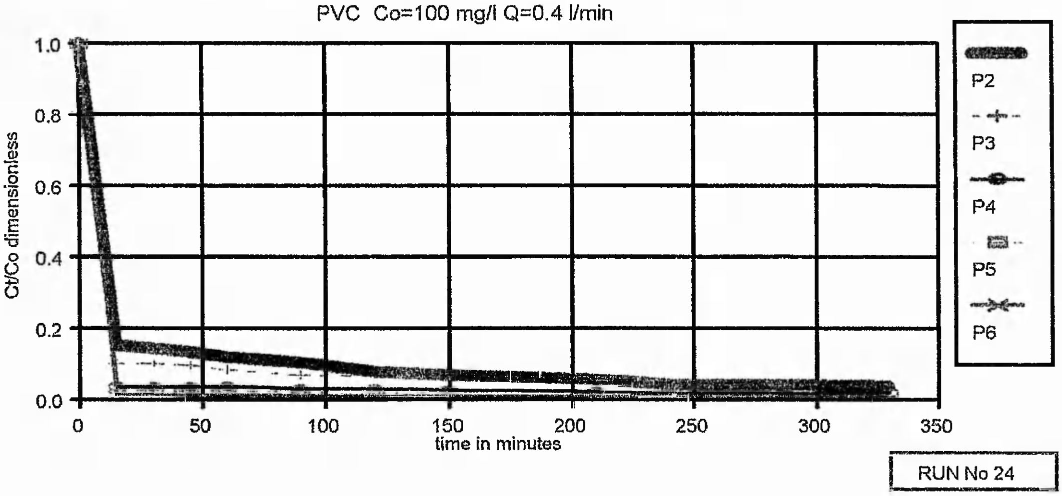
CONCENTRATION RATIO Vs DEPTH

PVC Co=100 mg/l Q=0.4 l/min



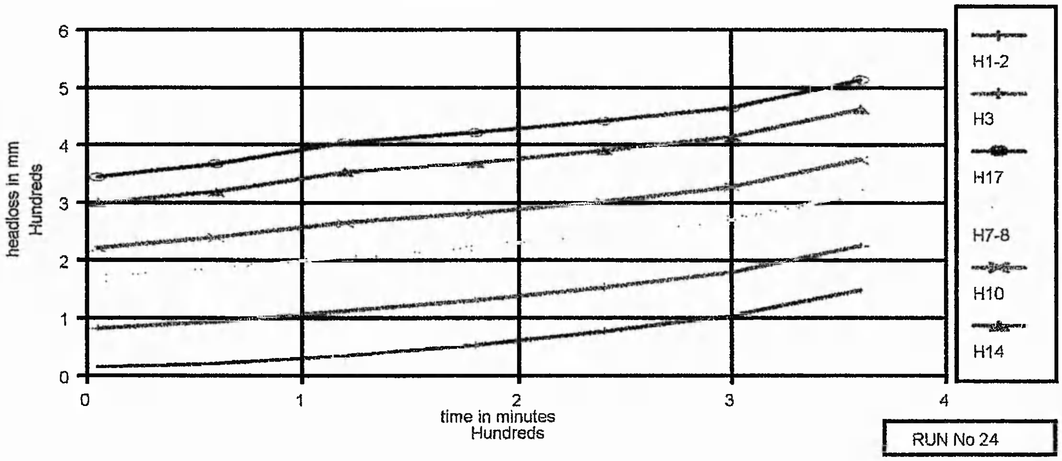
CONCENTRATION RATIO Vs TIME

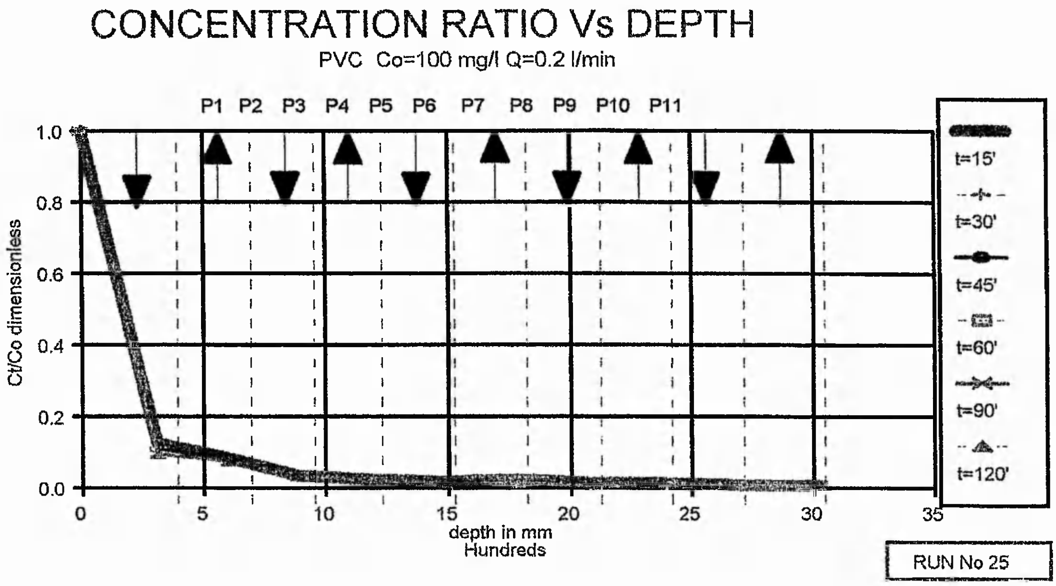
PVC Co=100 mg/l Q=0.4 l/min

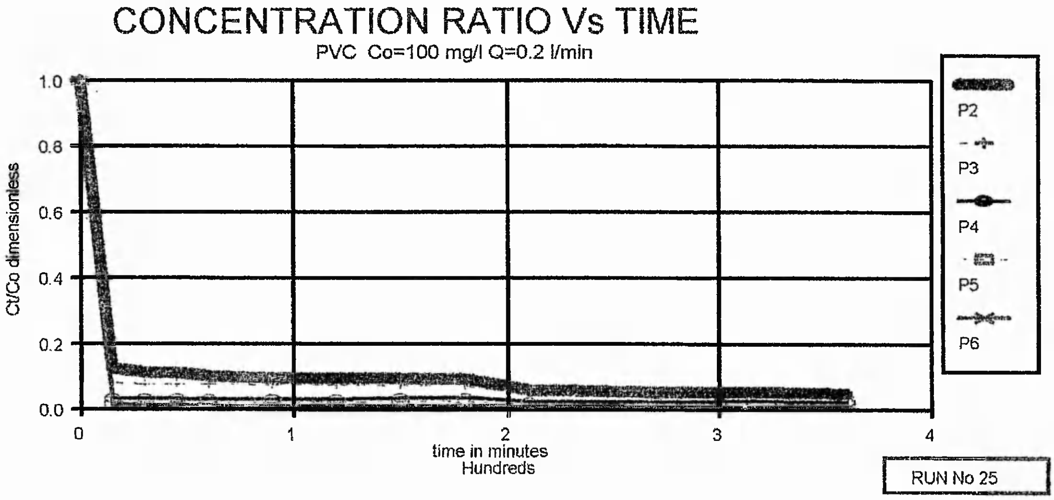


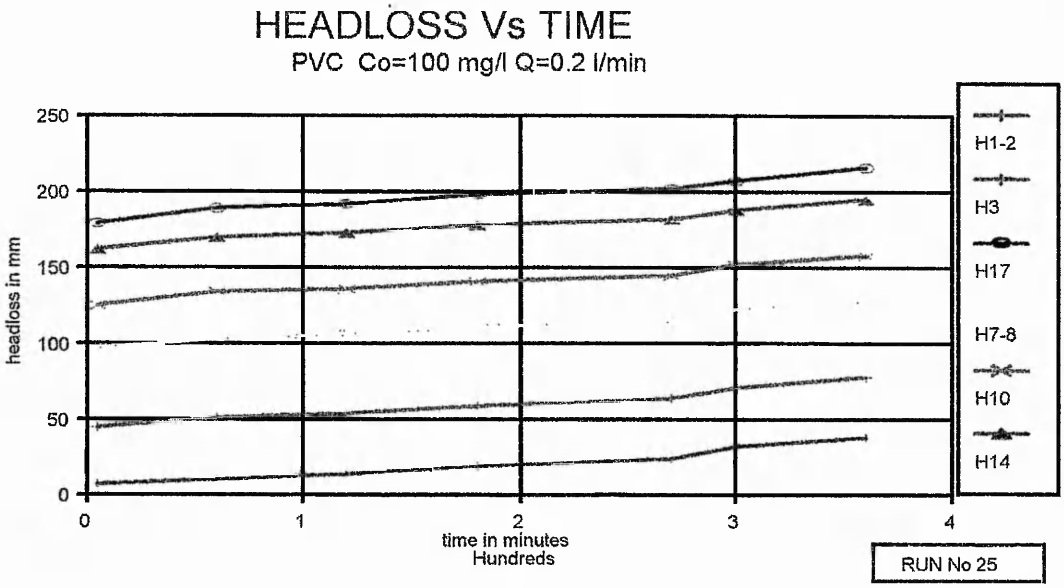
HEADLOSS Vs TIME

PVC Co=100 mg/l Q=0.4 l/min



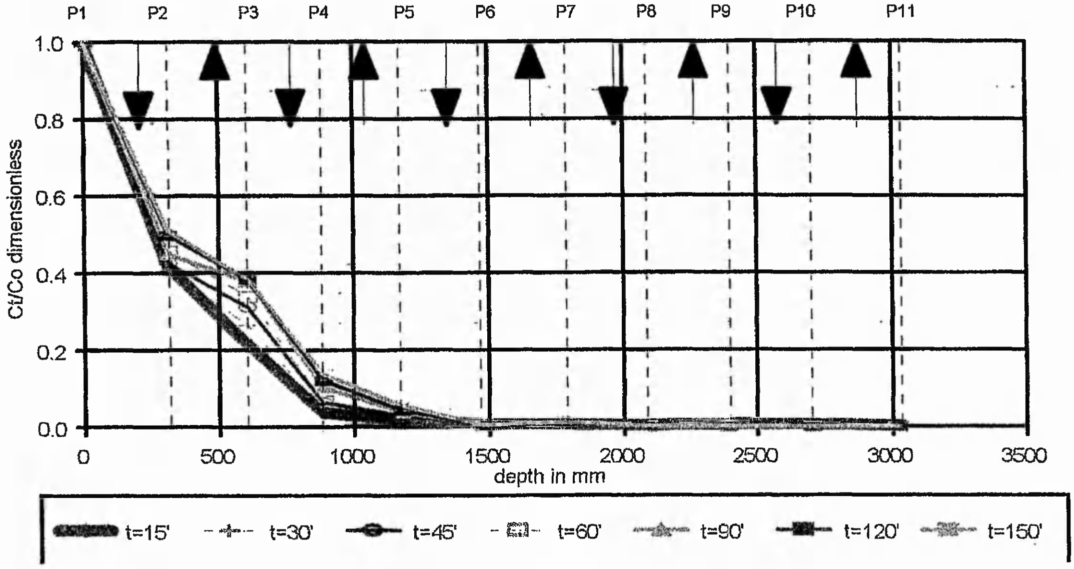

CONCENTRATION RATIO Vs DEPTH
PVC Co=100 mg/l Q=0.2 l/min
P1 P2 P3 P4 P5 P6 P7 P8 P9 P10 P11
Ct/Co dimensionless
1.0
0.8
0.6
0.4
0.2
0.0
0
5
10
15
20
25
30
35
depth in mm
Hundreds
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
RUN No 25


CONCENTRATION RATIO Vs TIME
PVC Co=100 mg/l Q=0.2 l/min
Ct/Co dimensionless
1.0
0.8
0.6
0.4
0.2
0.0
0
1
2
3
4
time in minutes
Hundreds
P2
P3
P4
P5
P6
RUN No 25


HEADLOSS Vs TIME
PVC Co=100 mg/l Q=0.2 l/min
headloss in mm
250
200
150
100
50
0
0
1
2
3
4
time in minutes
Hundreds
H1-2
H3
H17
H7-8
H10
H14
RUN No 25

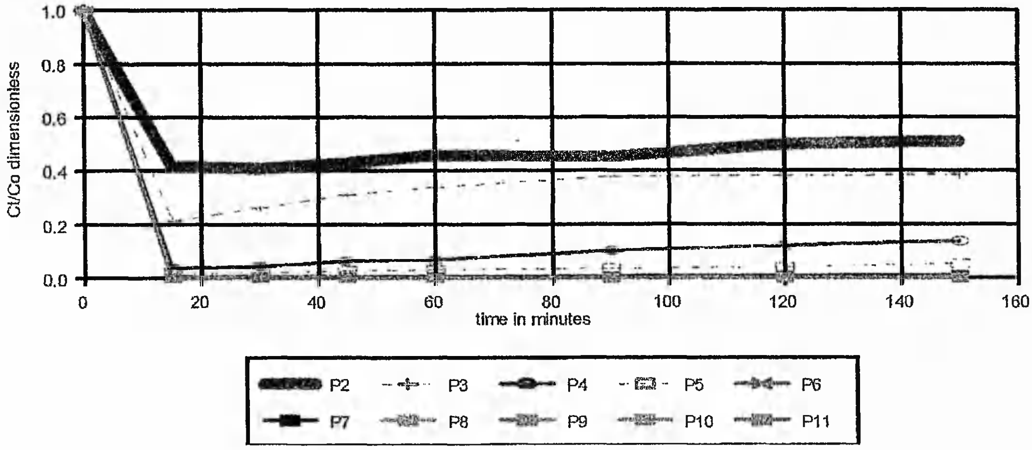
CONCENTRATION RATIO Vs DEPTH

RUN 26 KAOLIN Co=400 mg/l Q=1.0 l/min



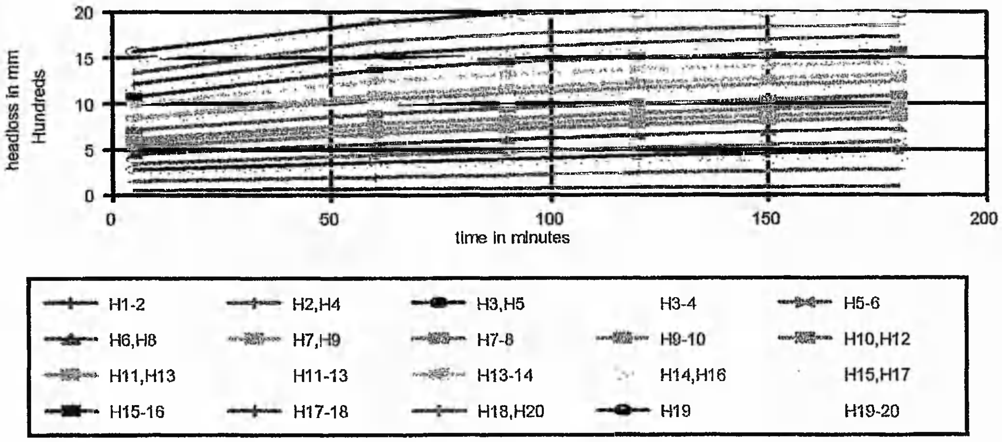
CONCENTRATION RATIO Vs TIME

RUN 26 KAOLIN Co=400 mg/l Q=1.0 l/min



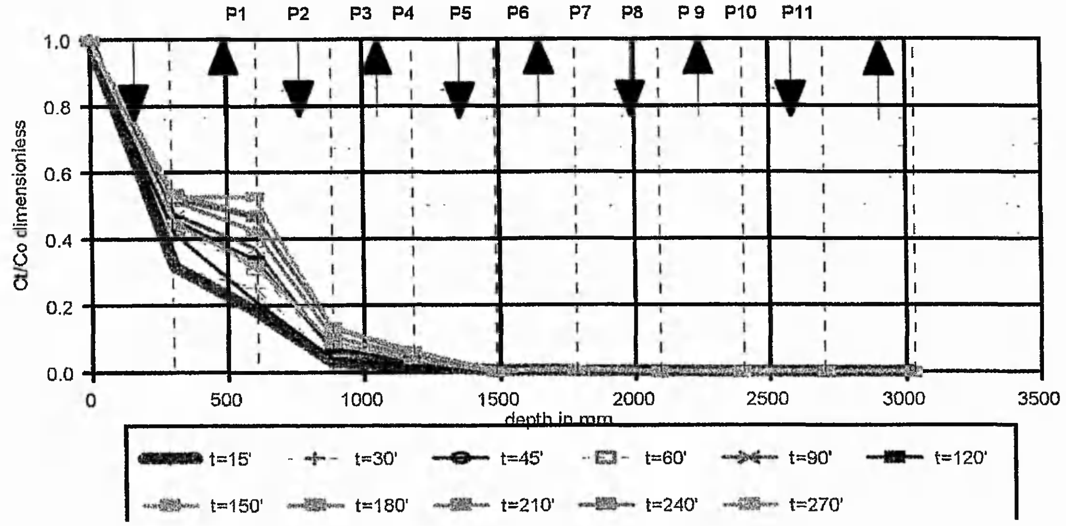
HEADLOSS Vs TIME

RUN 26 KAOLIN Co=400 mg/l Q=1.0 l/min



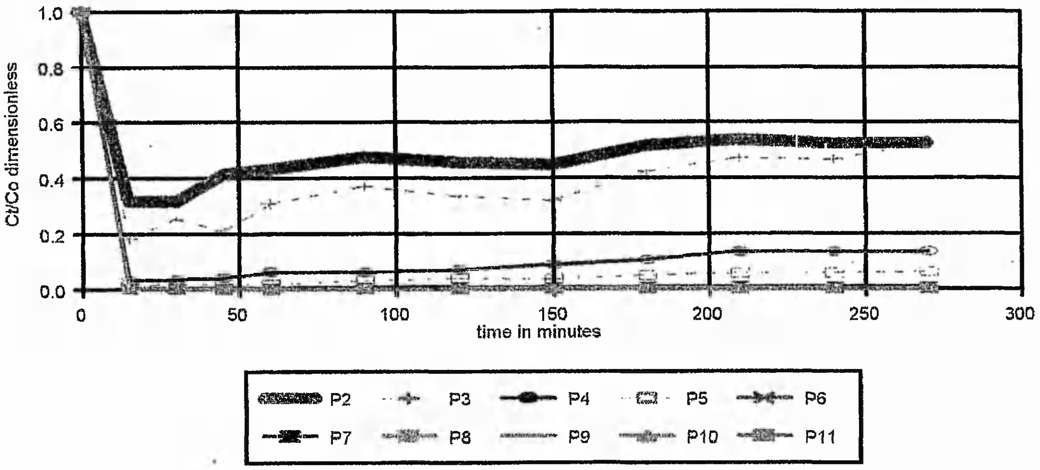
CONCENTRATION RATIO Vs DEPTH

RUN 27 KAOLIN Co=400 mg/l Q=0.8 l/min



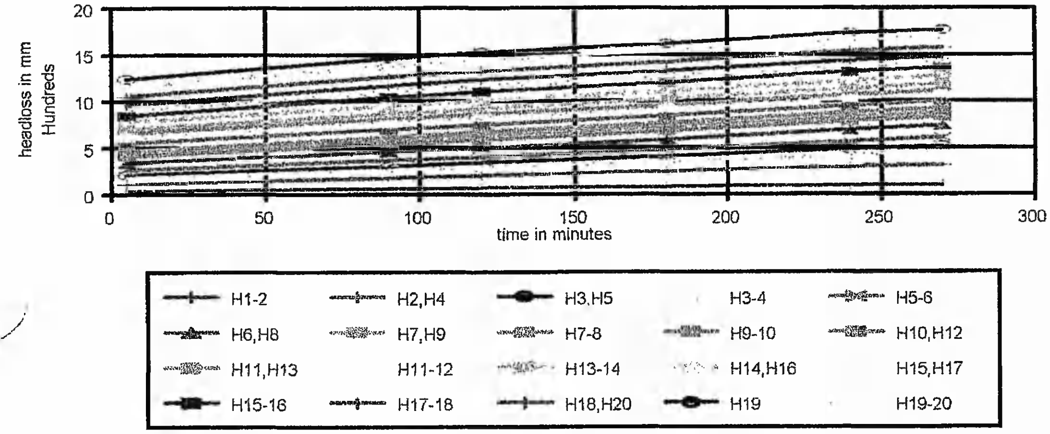
CONCENTRATION RATIO Vs TIME

RUN 27 KAOLIN Co=400 mg/l Q=0.8 l/min



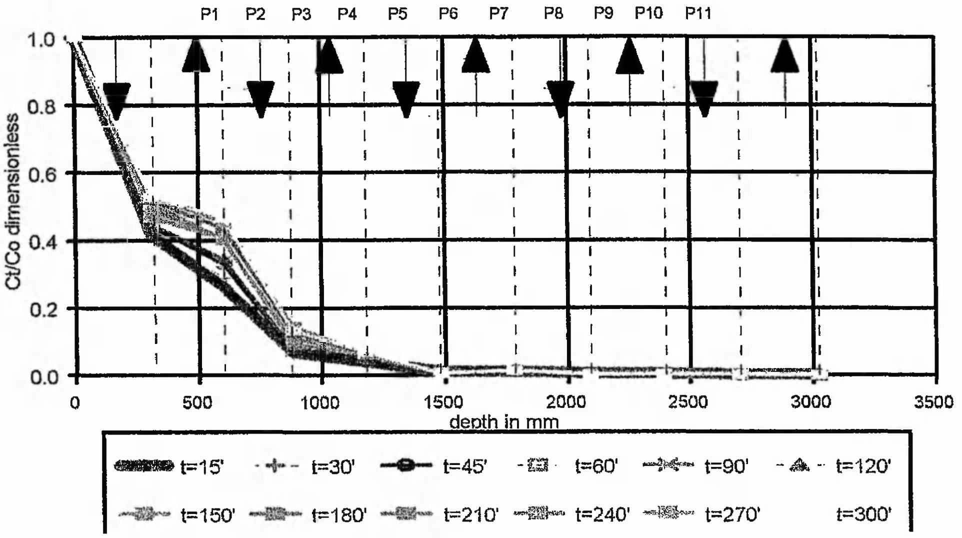
HEADLOSS Vs TIME

RUN 27 KAOLIN Co=400 mg/l Q=0.8 l/min



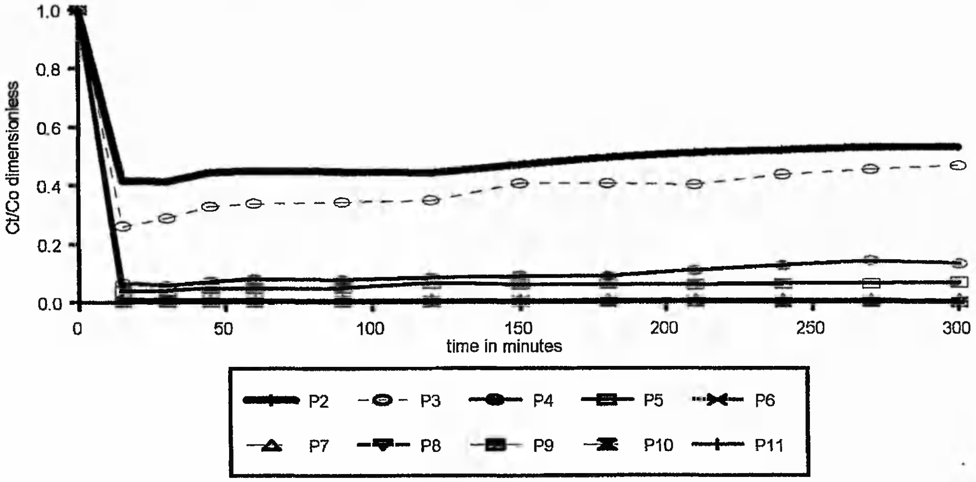
CONCENTRATION RATIO Vs DEPTH

RUN 28 KAOLIN Co=400 mg/l Q=0.6 l/min



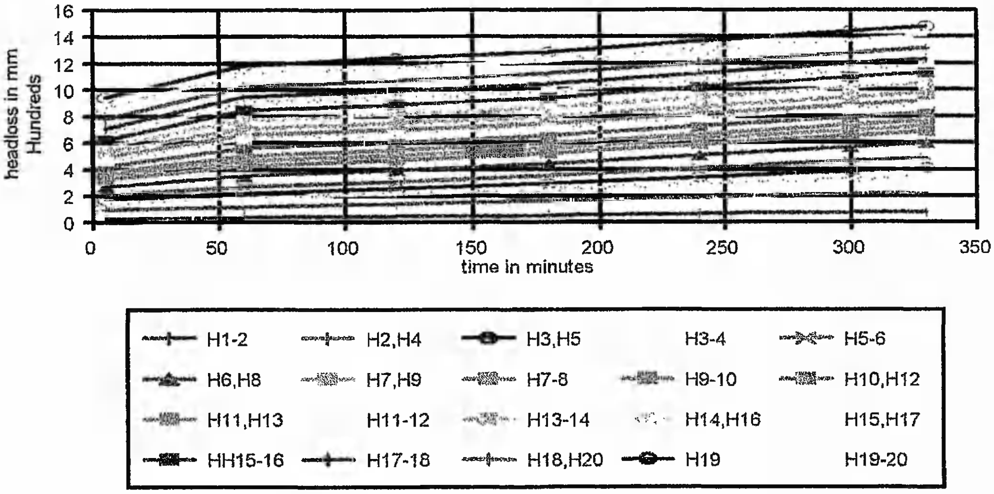
CONCENTRATION RATIO Vs TIME

KAOLIN Co=400 mg/l Q=0.6 l/min RUN 28



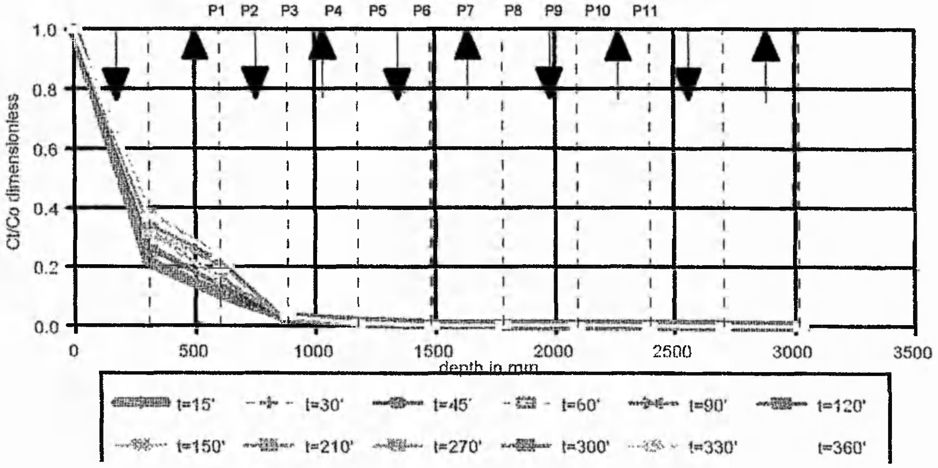
HEADLOSS Vs TIME

RUN 28 KAOLIN Co=400 mg/l Q=0.6 l/min



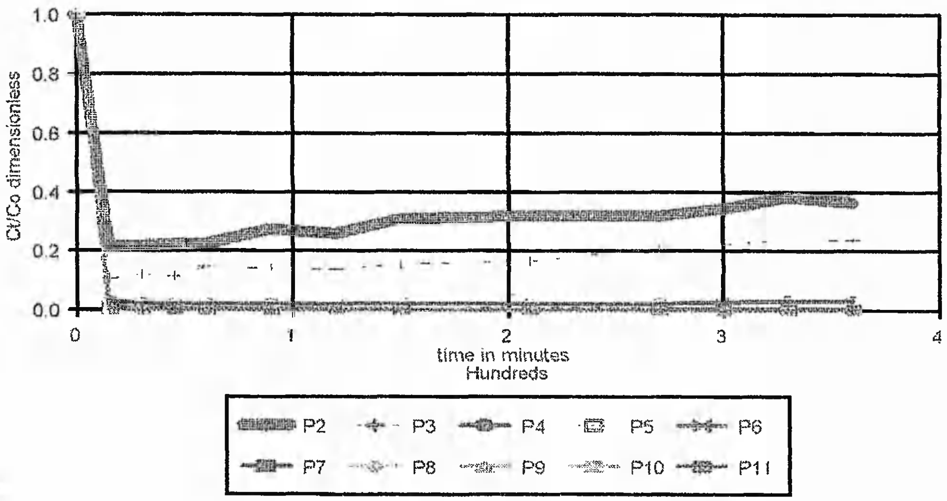
CONCENTRATION RATIO Vs DEPTH

RUN 29 KAOLIN Co=400 mg/l Q=0.4 l/min



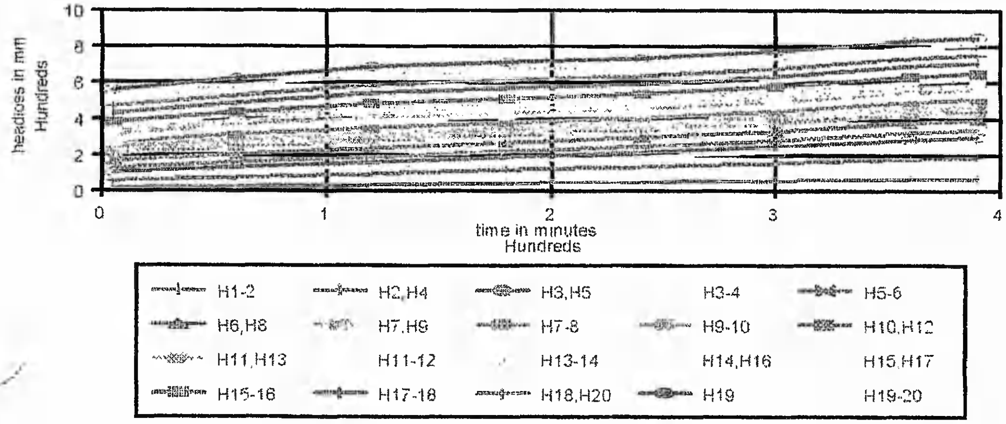
CONCENTRATION RATIO Vs TIME

RUN 29 KAOLIN Co=400 mg/l Q=0.4 l/min



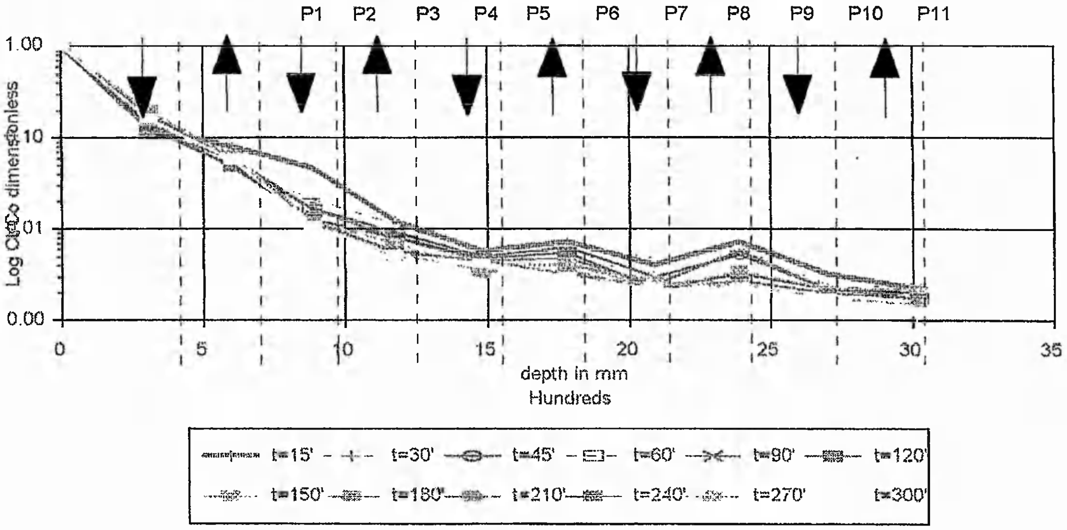
HEADLOSS Vs TIME

RUN 29 KAOLIN Co=400 mg/l Q=0.4 l/min



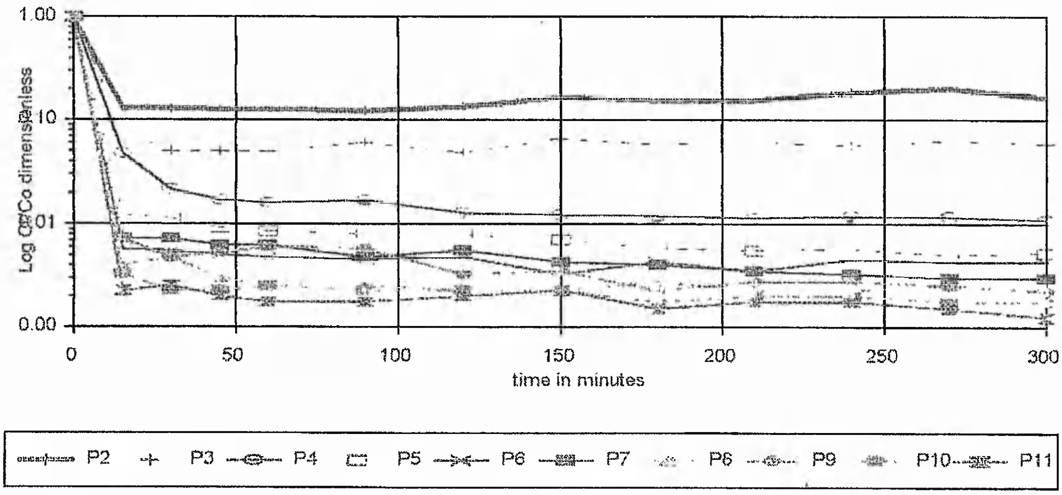
CONCENTRATION RATIO Vs DEPTH

RUN 30 KAOLIN Co=400 mg/l Q=0.2 l/min



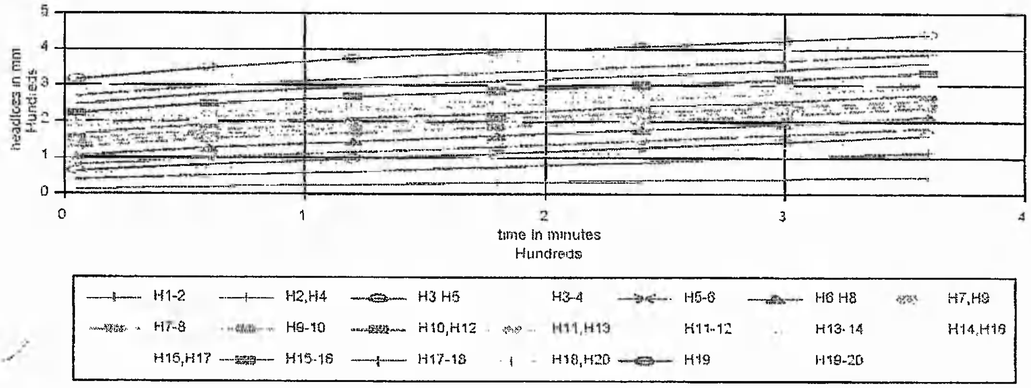
CONCENTRATION RATIO Vs TIME

RUN 30 KAOLIN Co=400 mg/l Q=0.2 l/min



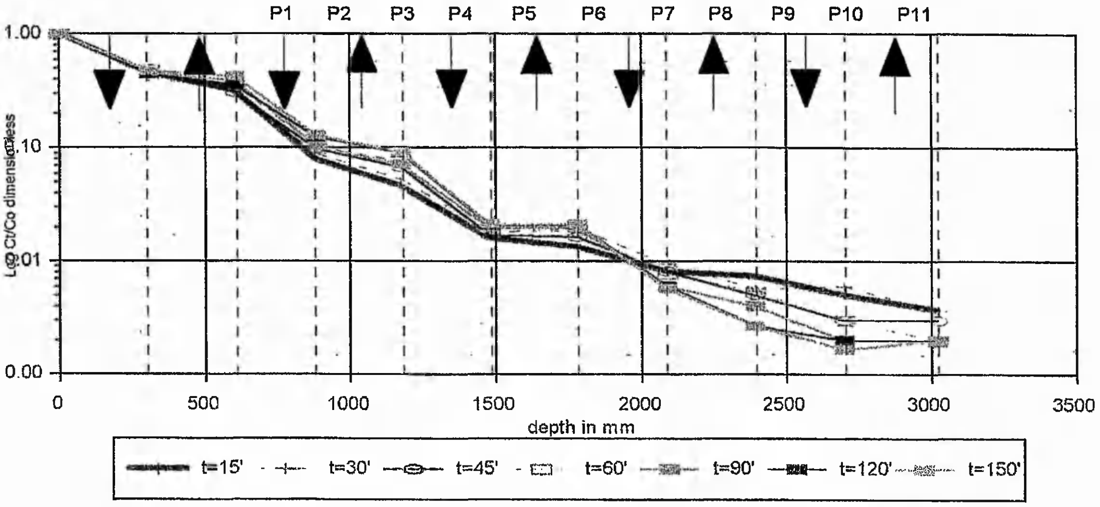
HEADLOSS Vs TIME

RUN 30 KAOLIN Co=400 mg/l Q=0.2 l/min



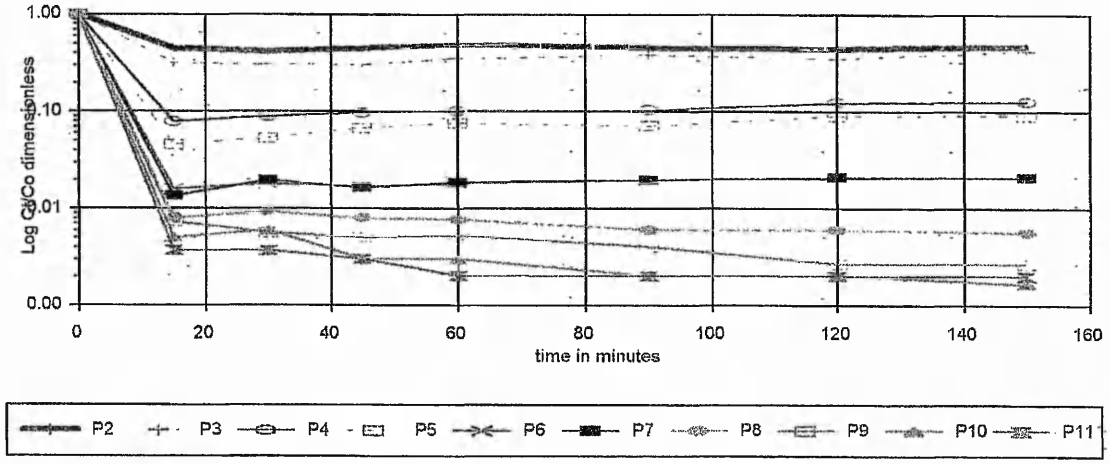
CONCENTRATION RATIO Vs DEPTH

RUN 31 KAOLIN Co=300 mg/l Q=1.0 l/min



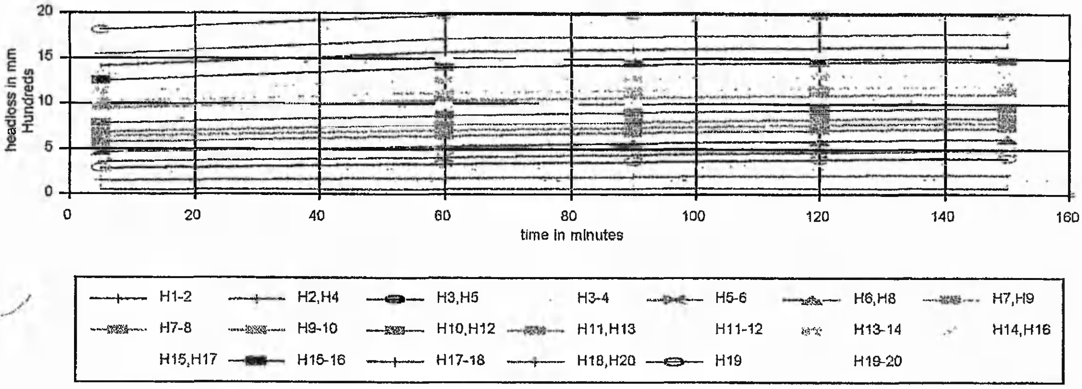
CONCENTRATION RATIO Vs TIME

RUN 31 KAOLIN Co=300 mg/l Q=1.0 l/min



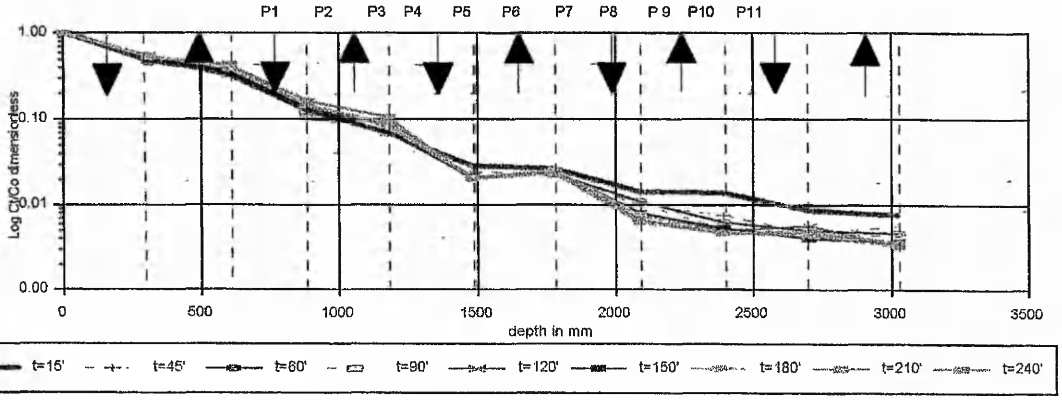
HEADLOSS Vs TIME

RUN 31 KAOLIN Co=300 mg/l Q=1.0 l/min



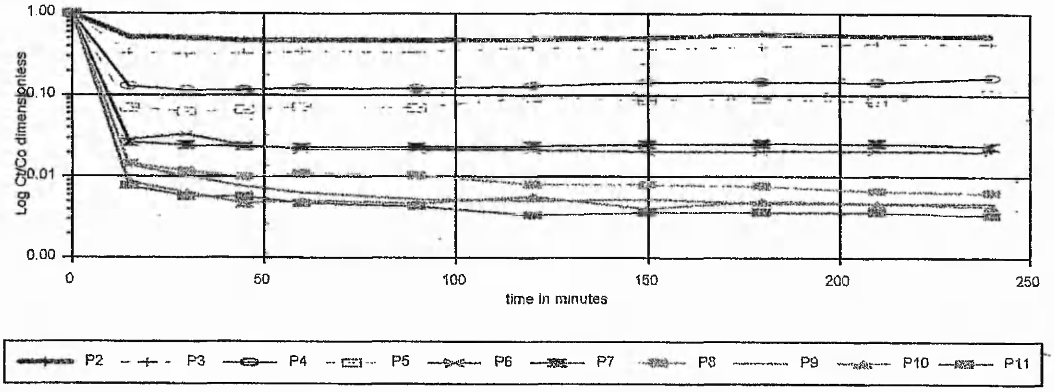
CONCENTRATION RATIO Vs DEPTH

RUN 32 KAOLIN Co=300 mg/l Q=0.8 l/min



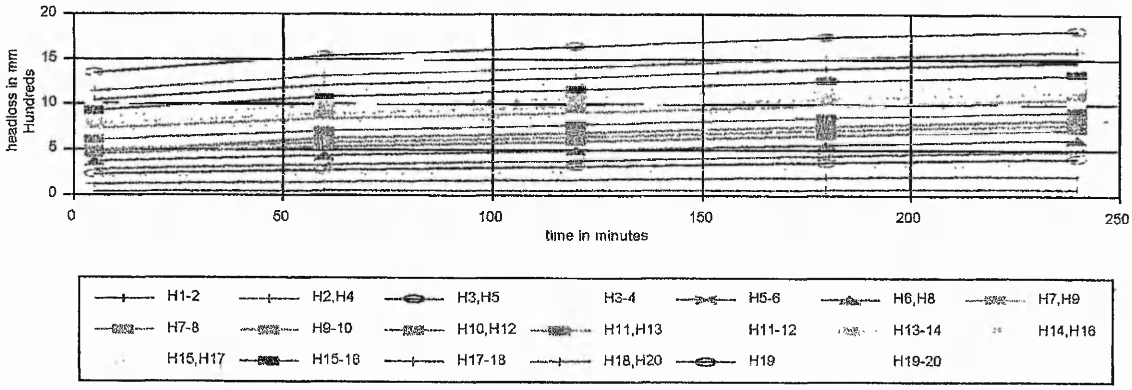
CONCENTRATION RATIO Vs TIME

RUN 32 KAOLIN Co=300 mg/l Q=0.8 l/min

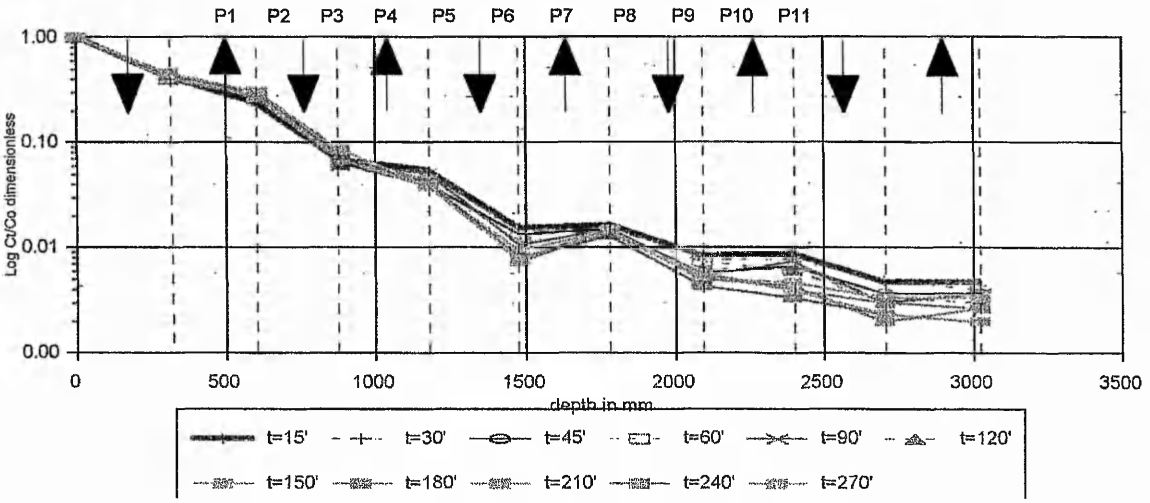


HEADLOSS Vs TIME

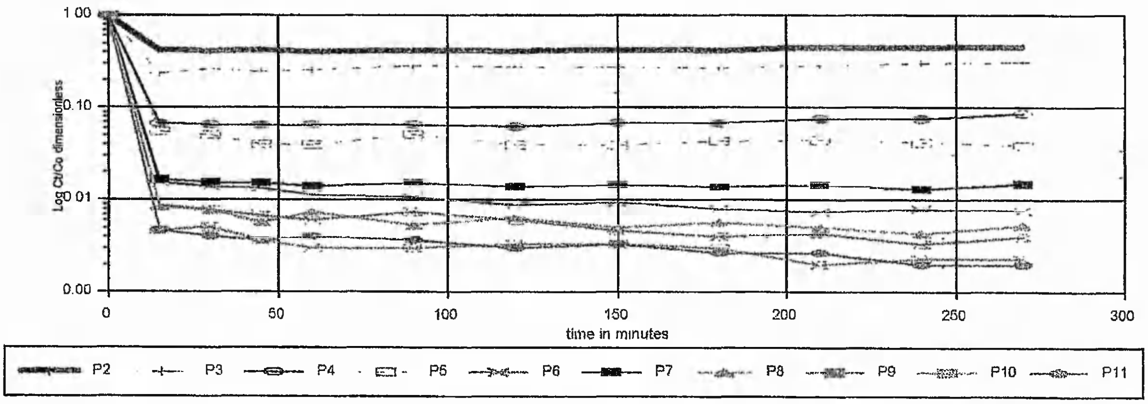
RUN 32 KAOLIN Co=300 mg/l Q=0.8 l/min



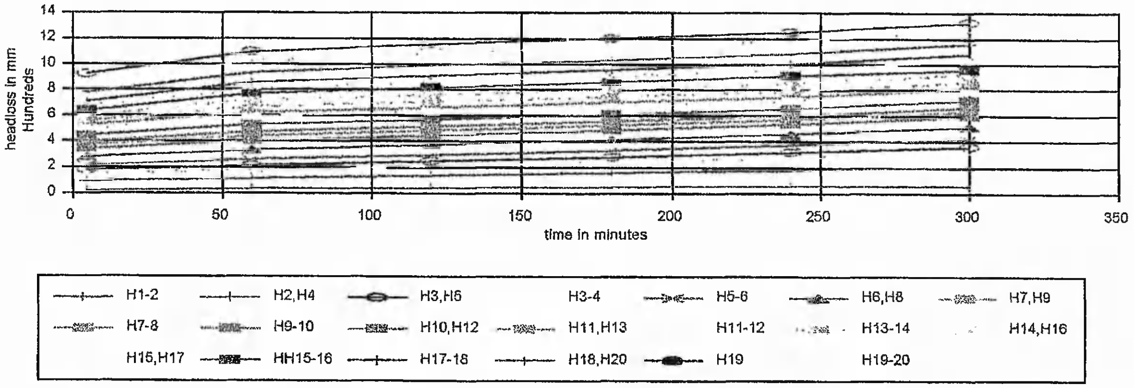
CONCENTRATION RATIO Vs DEPTH
RUN 33 KAOLIN Co=300 mg/l Q=0.6 l/min



CONCENTRATION RATIO Vs TIME
RUN 33 KAOLIN Co=300 mg/l Q=0.6 l/min

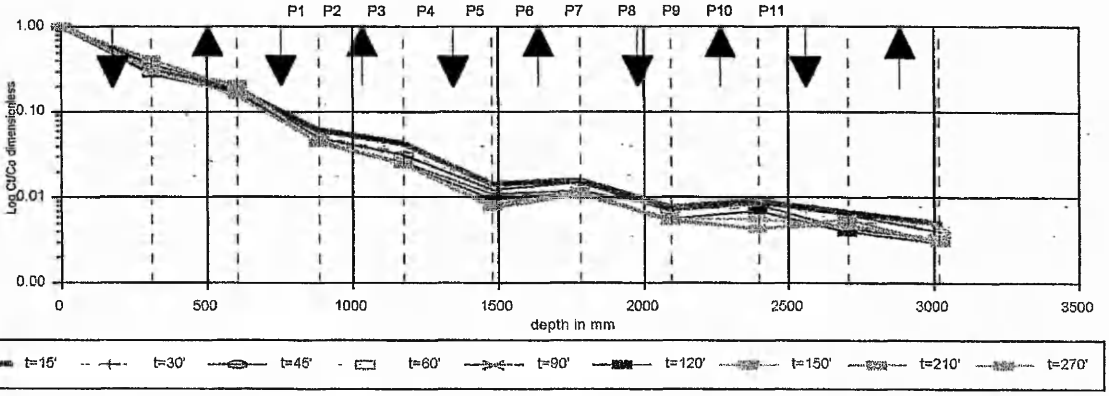


HEADLOSS Vs TIME
RUN 33 KAOLIN Co=300 mg/l Q=0.6 l/min



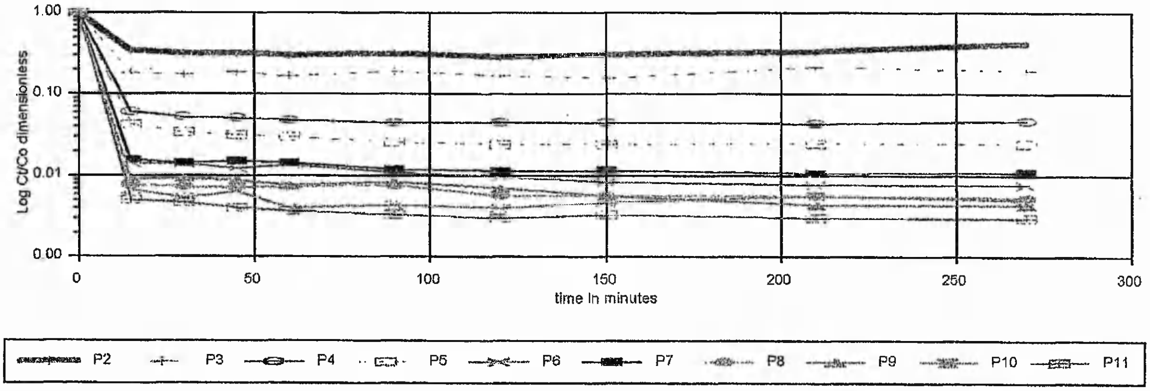
CONCENTRATION RATIO Vs DEPTH

RUN 34 KAOLIN Co=300 mg/l Q=0.4 l/min



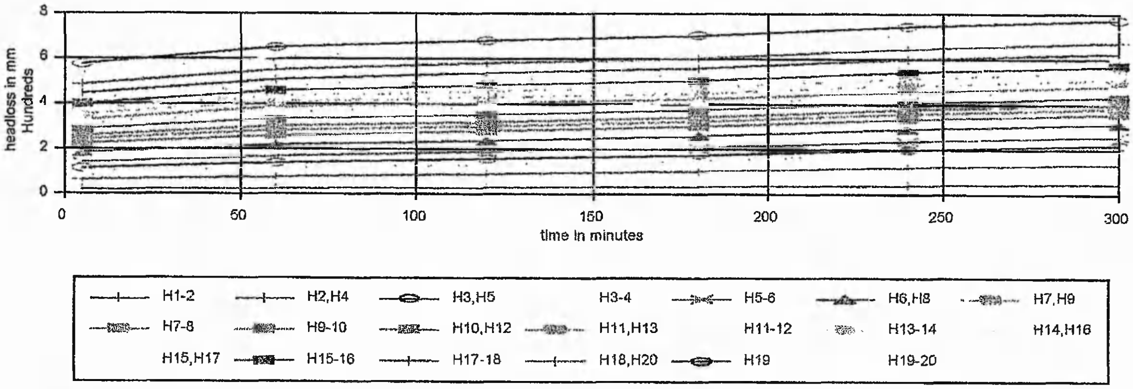
CONCENTRATION RATIO Vs TIME

RUN 34 KAOLIN Co=300 mg/l Q=0.4 l/min



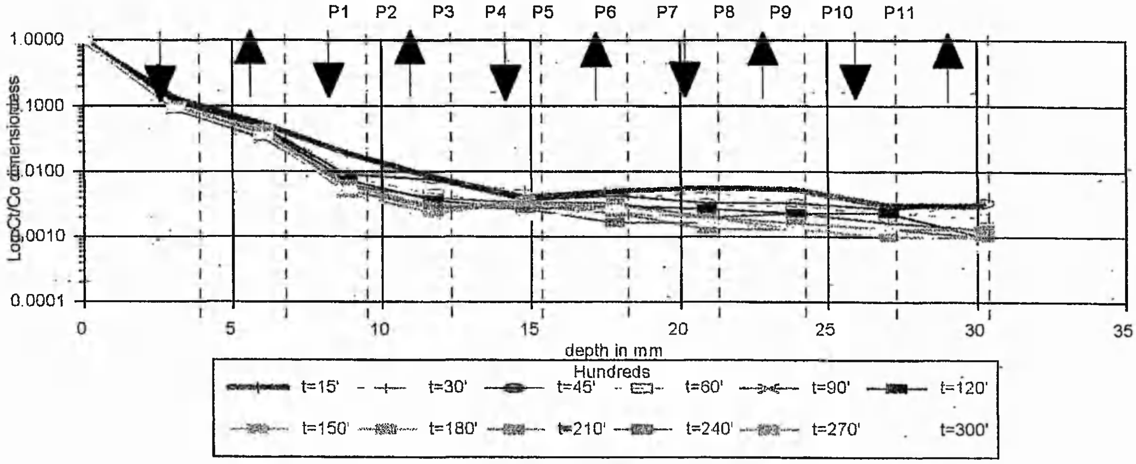
HEADLOSS Vs TIME

RUN 34 KAOLIN Co=300 mg/l Q=0.4 l/min



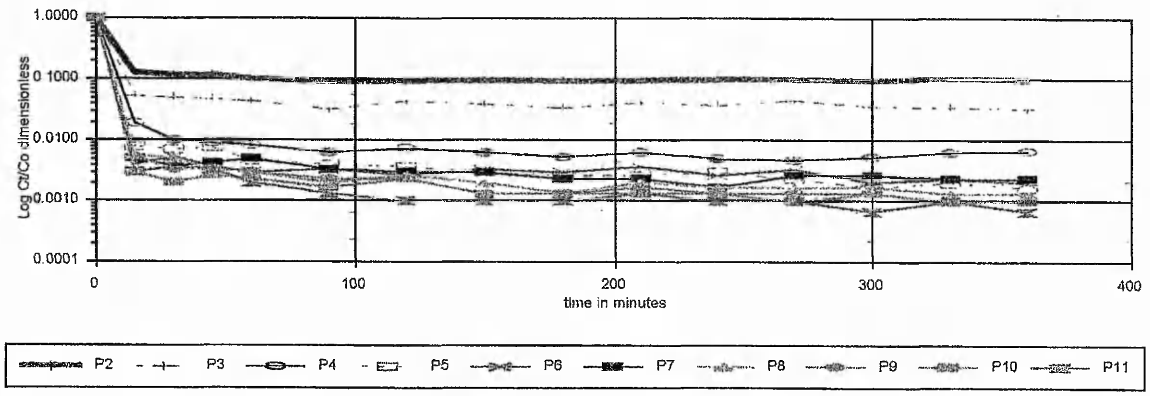
CONCENTRATION RATIO Vs DEPTH

RUN 35 KAOLIN Co=300 mg/l Q=0.2 l/min



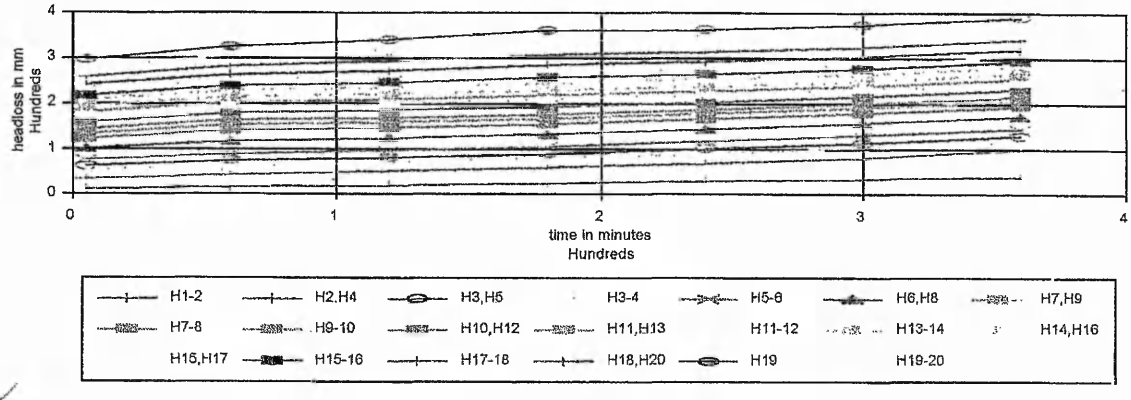
CONCENTRATION RATIO Vs TIME

RUN 35 KAOLIN Co=300 mg/l Q=0.2 l/min



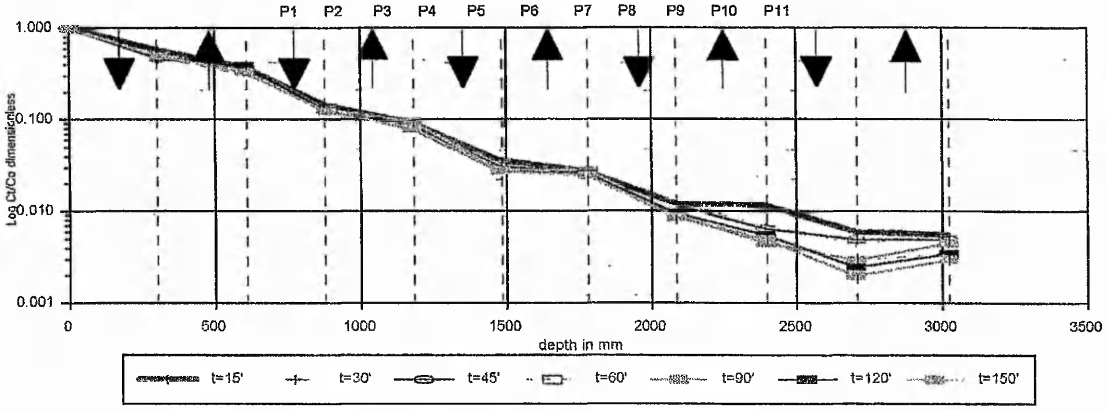
HEADLOSS Vs TIME

RUN 35 KAOLIN Co=300 mg/l Q=0.2 l/min



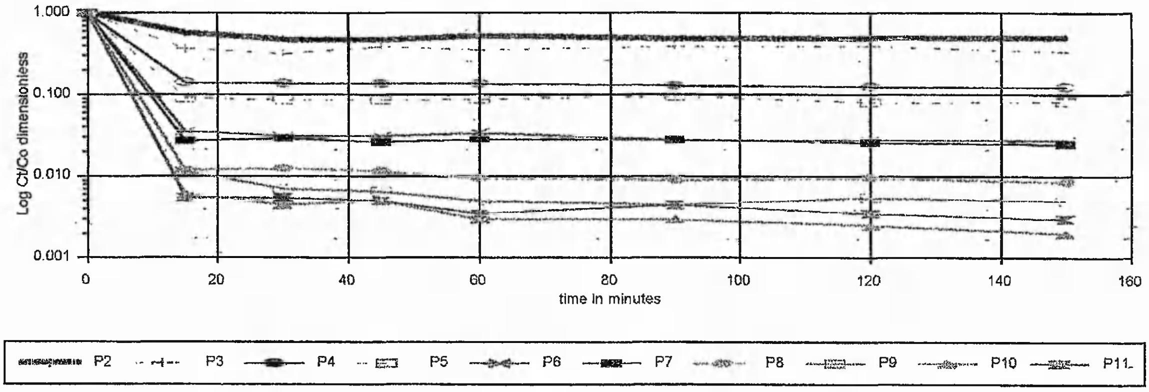
CONCENTRATION RATIO Vs DEPTH

RUN 36 KAOLIN Co=200 mg/l Q=1.0 l/min



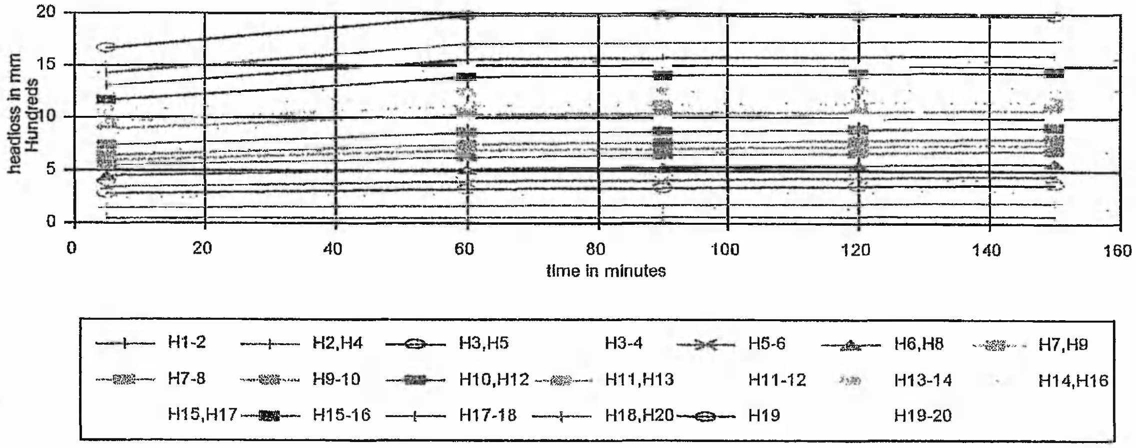
CONCENTRATION RATIO Vs TIME

RUN 36 KAOLIN Co=200 mg/l Q=1.0 l/min



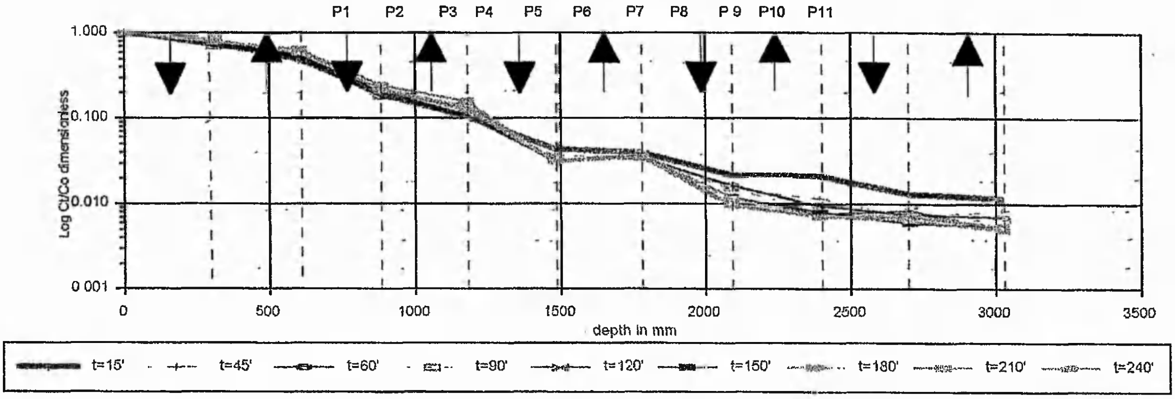
HEADLOSS Vs TIME

RUN 36 KAOLIN Co=200 mg/l Q=1.0 l/min



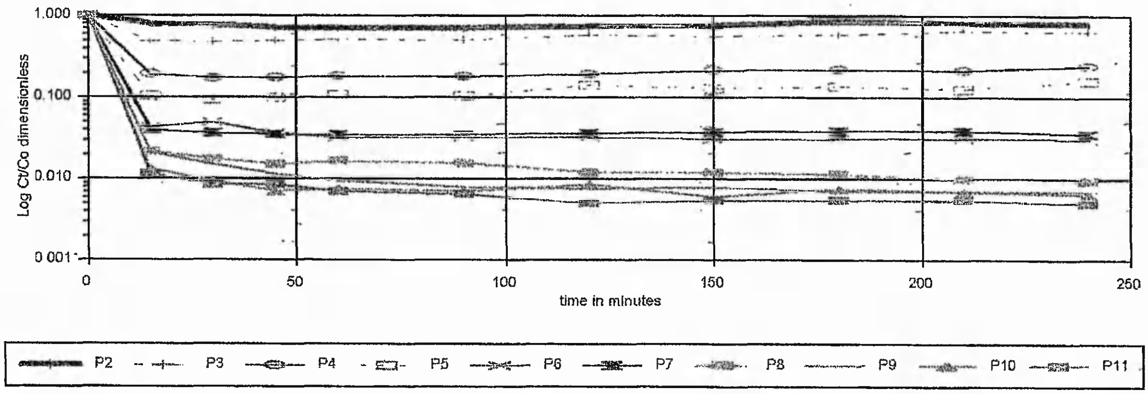
CONCENTRATION RATIO Vs DEPTH

RUN 37 KAOLIN Co=200 mg/l Q=0.8 l/min



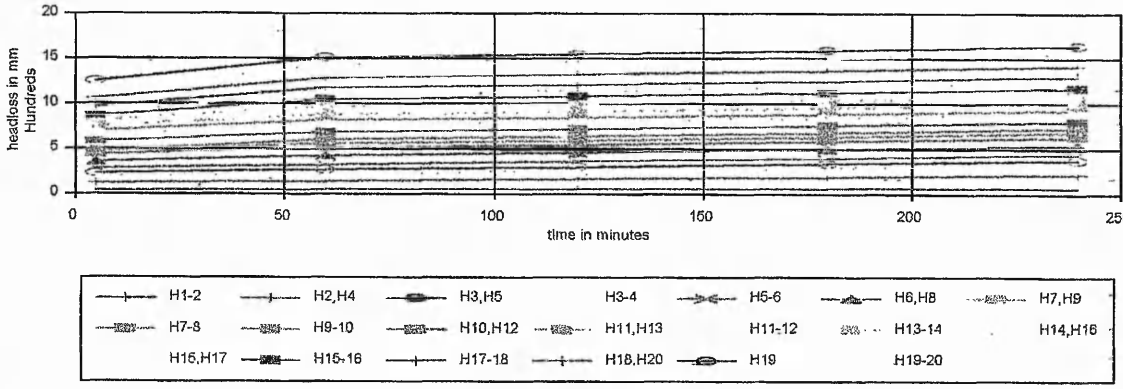
CONCENTRATION RATIO Vs TIME

RUN 37 KAOLIN Co=200 mg/l Q=0.8 l/min



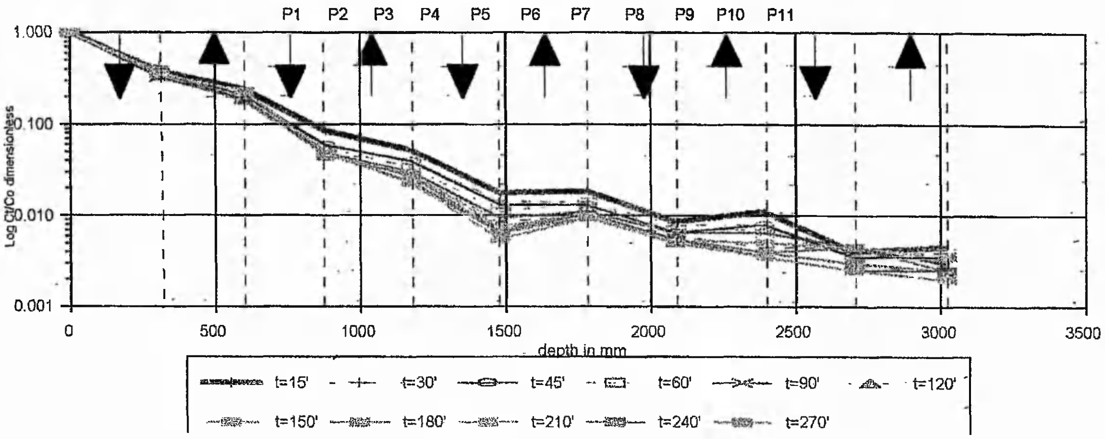
HEADLOSS Vs TIME

RUN 37 KAOLIN Co=200 mg/l Q=0.8 l/min



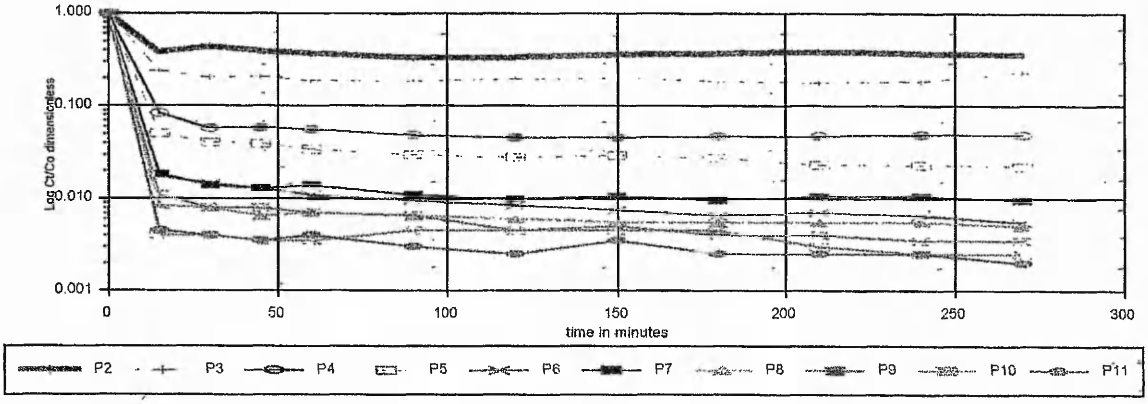
CONCENTRATION RATIO Vs DEPTH

RUN 38 KAOLIN Co=200 mg/l Q=0.6 l/min



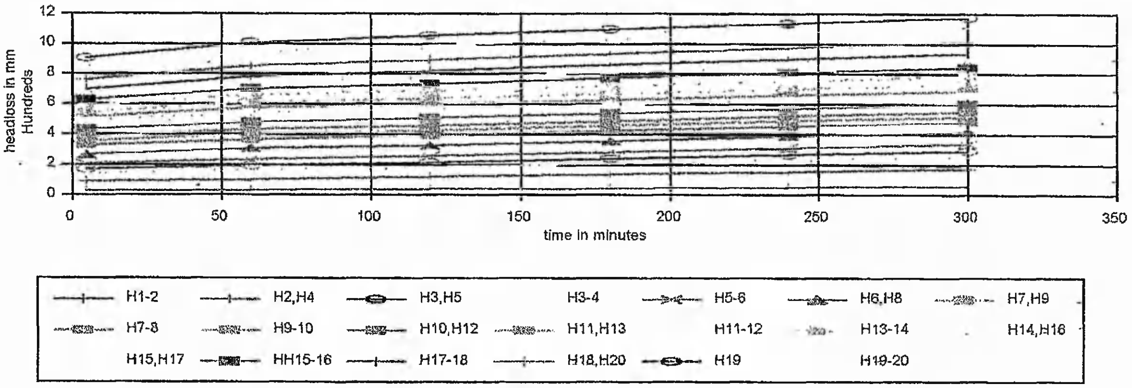
CONCENTRATION RATIO Vs TIME

RUN 38 KAOLIN Co=200 mg/l Q=0.6 l/min



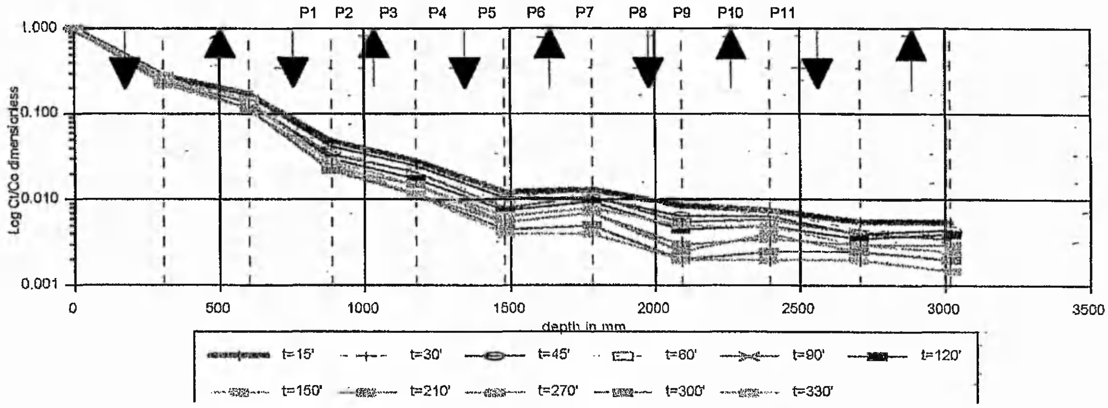
HEADLOSS Vs TIME

RUN 38 KAOLIN Co=200 mg/l Q=0.6 l/min



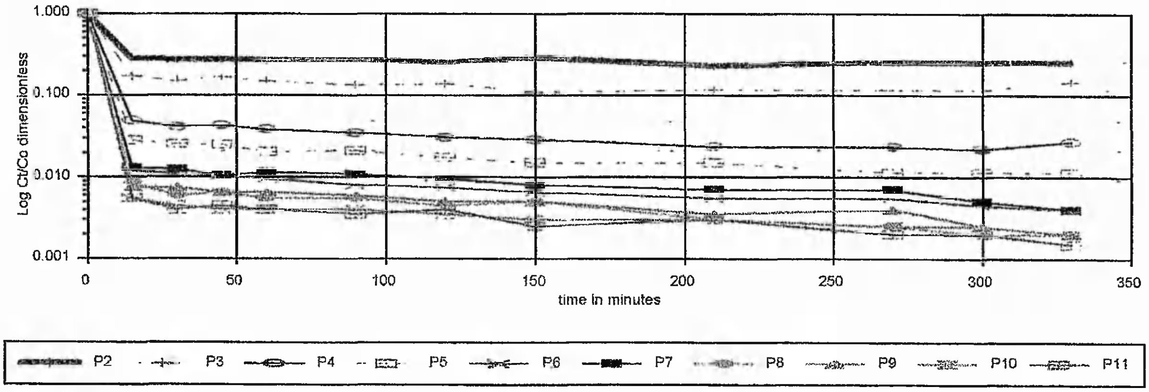
CONCENTRATION RATIO Vs DEPTH

RUN 39 KAOLIN Co=200 mg/l Q=0.4 l/min



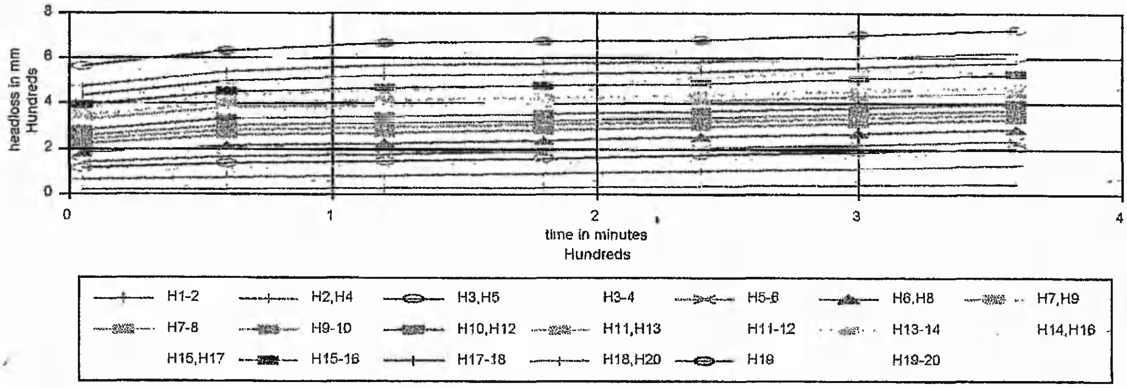
CONCENTRATION RATIO Vs TIME

RUN 39 KAOLIN Co=200 mg/l Q=0.4 l/min



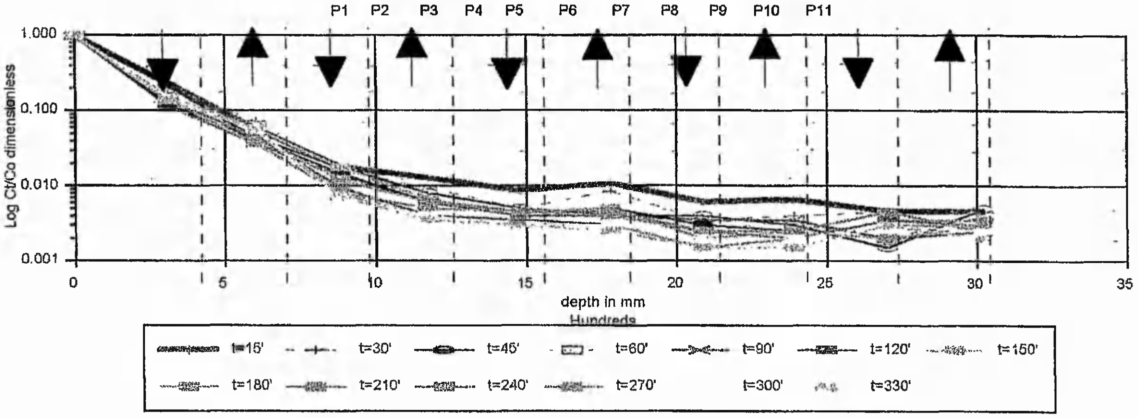
HEADLOSS Vs TIME

RUN 39 KAOLIN Co=200 mg/l Q=0.4 l/min



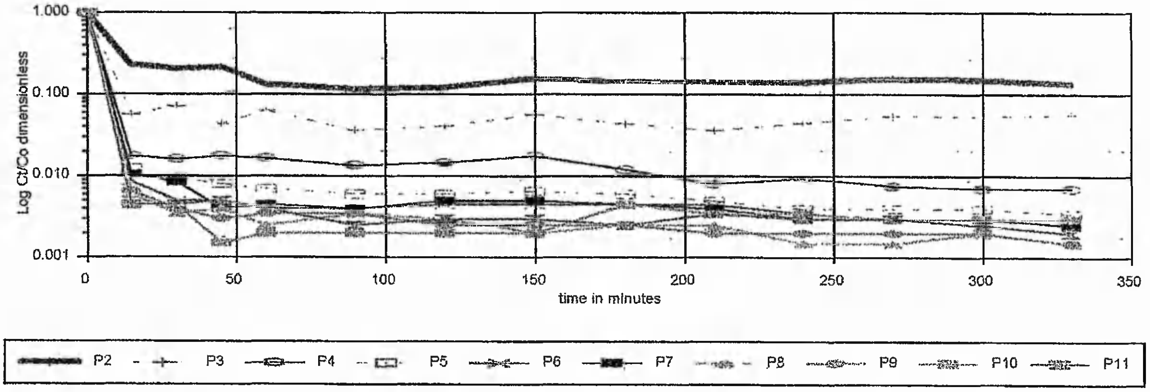
CONCENTRATION RATIO Vs DEPTH

RUN 40 KAOLIN Co=200 mg/l Q=0.2 l/min



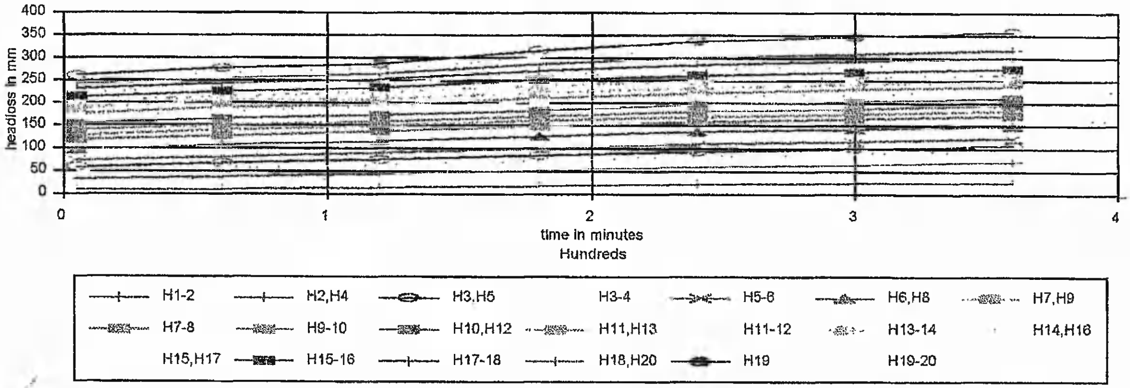
CONCENTRATION RATIO Vs TIME

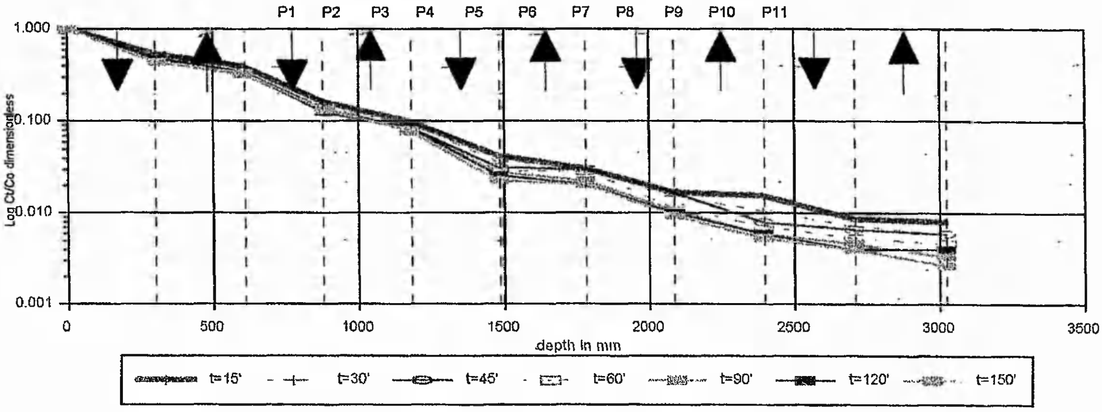
RUN 40 KAOLIN Co=200 mg/l Q=0.2 l/min



HEADLOSS Vs TIME

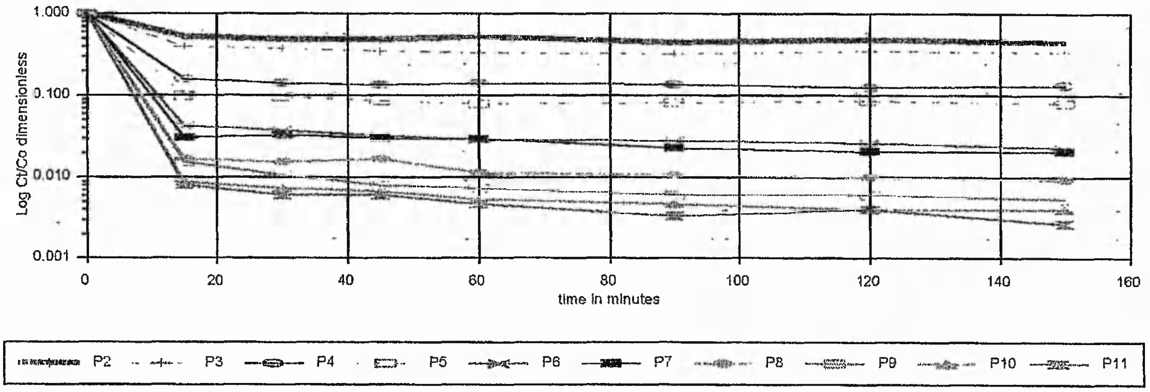
RUN 40 KAOLIN Co=200 mg/l Q=0.2 l/min





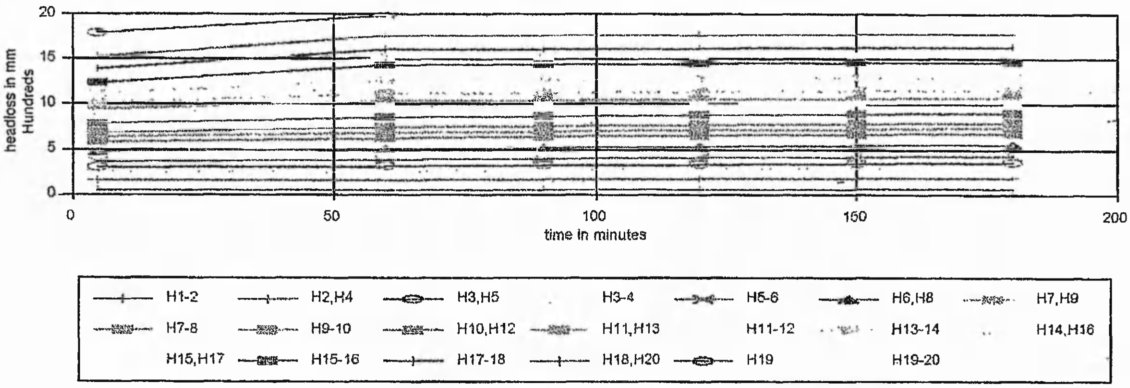
CONCENTRATION RATIO Vs DEPTH

RUN 41 KAOLIN Co=150 mg/l Q=1.0 l/min



CONCENTRATION RATIO Vs TIME

RUN 41 KAOLIN Co=150 mg/l Q=1.0 l/min

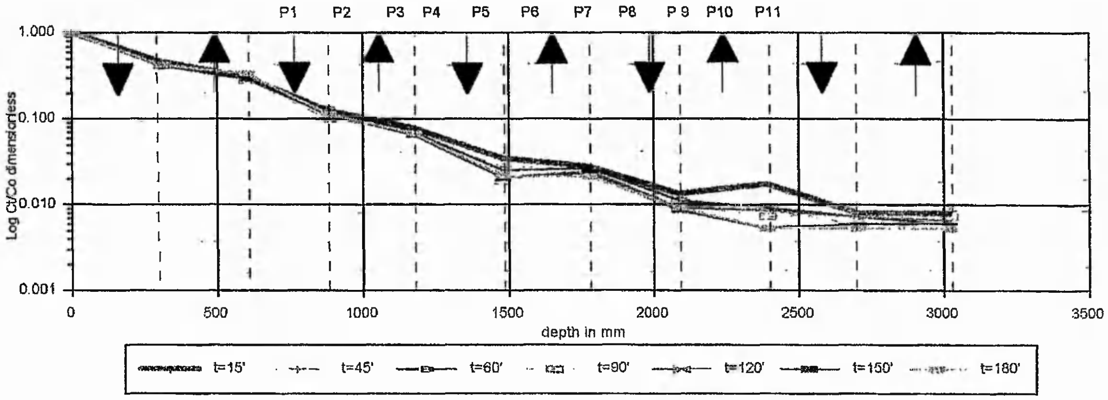


HEADLOSS Vs TIME

RUN 41 KAOLIN Co=150 mg/l Q=1.0 l/min

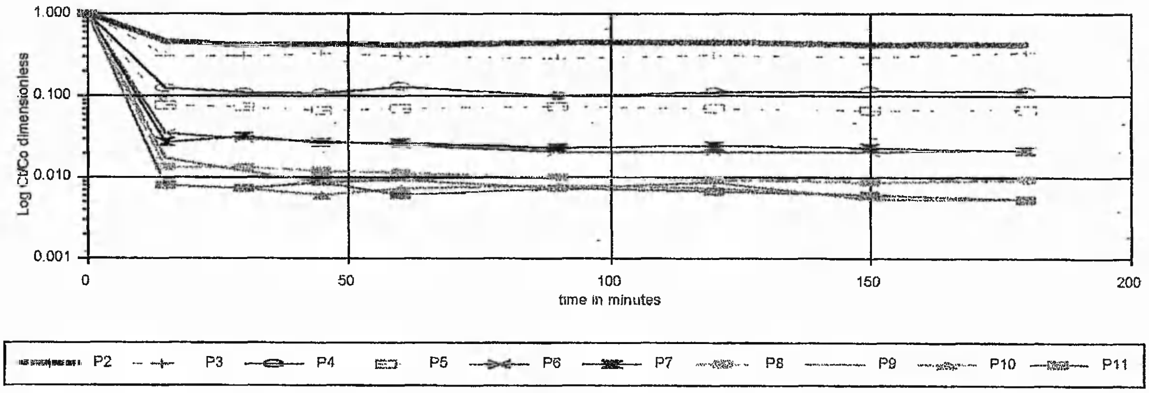
CONCENTRATION RATIO Vs DEPTH

RUN 42 KAOLIN Co=150 mg/l Q=0.8 l/min



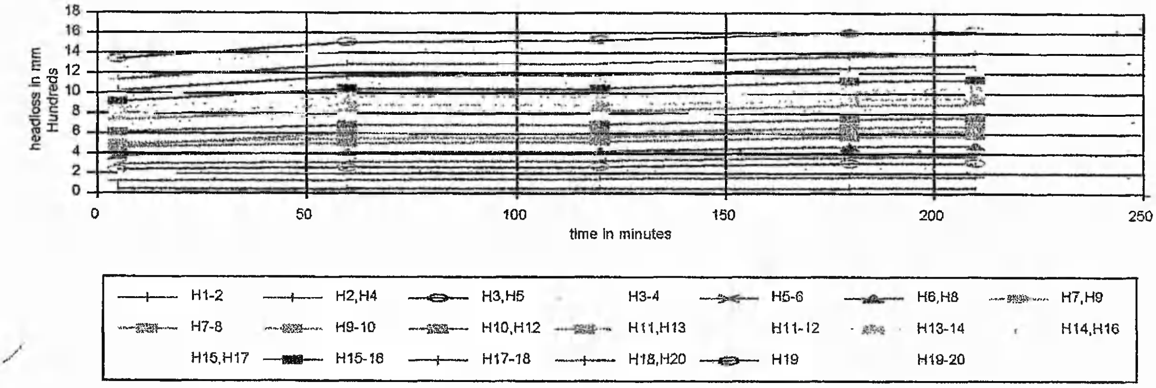
CONCENTRATION RATIO Vs TIME

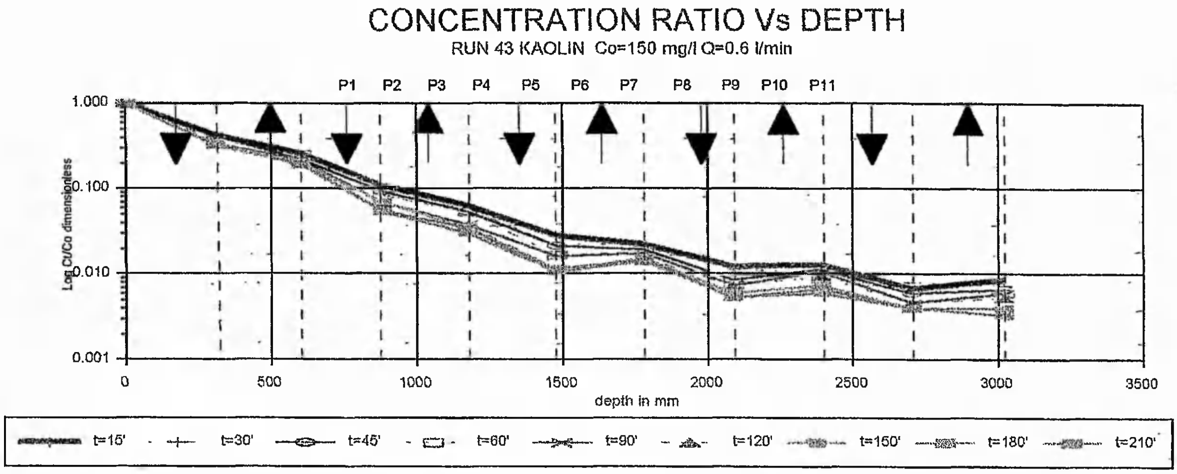
RUN 42 KAOLIN Co=150 mg/l Q=0.8 l/min

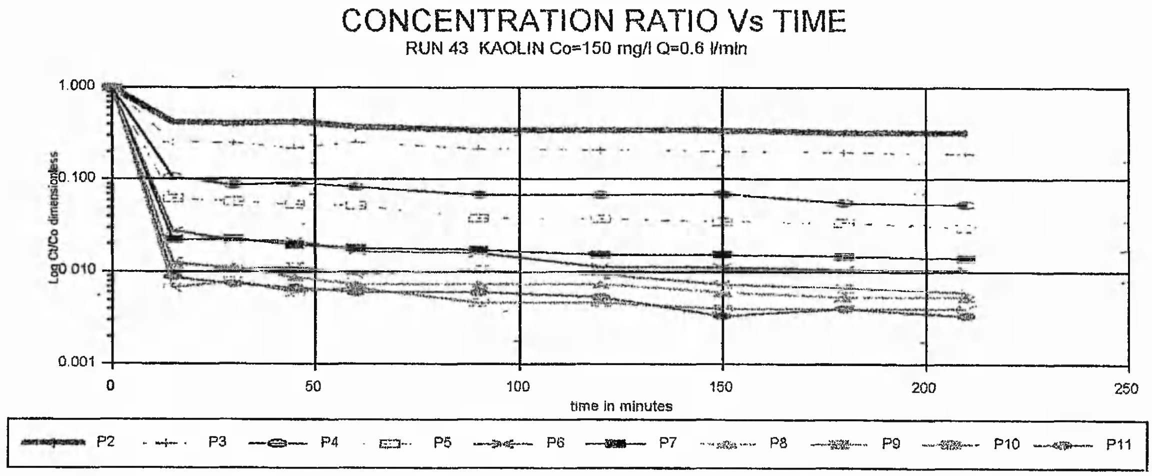


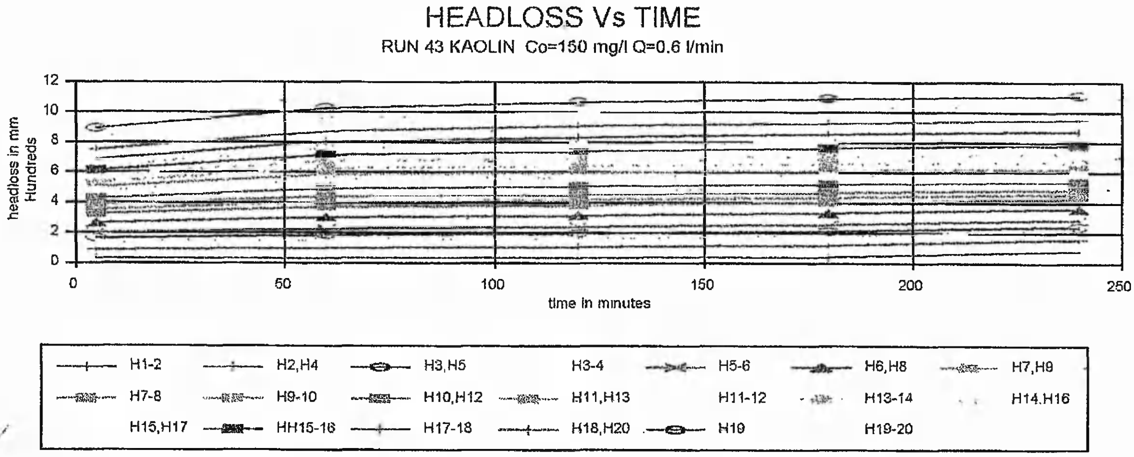
HEADLOSS Vs TIME

RUN 42 KAOLIN Co=150 mg/l Q=0.8 l/min



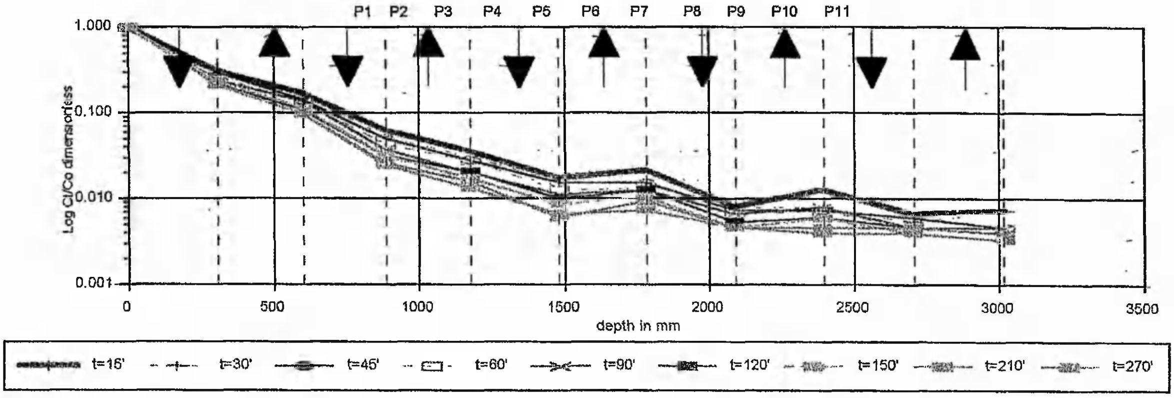

CONCENTRATION RATIO Vs DEPTH
RUN 43 KAOLIN Co=150 mg/l Q=0.6 l/min
P1 P2 P3 P4 P5 P6 P7 P8 P9 P10 P11
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
500
1000
1500
2000
2500
3000
3500
depth in mm
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
t=150'
t=180'
t=210'


CONCENTRATION RATIO Vs TIME
RUN 43 KAOLIN Co=150 mg/l Q=0.6 l/min
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
50
100
150
200
250
time in minutes
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11


HEADLOSS Vs TIME
RUN 43 KAOLIN Co=150 mg/l Q=0.6 l/min
12
10
8
6
4
2
0
headloss in mm Hundreds
0
50
100
150
200
250
time in minutes
H1-2
H2,H4
H3,H5
H3-4
H5-6
H6,H8
H7,H9
H7-8
H9-10
H10,H12
H11,H13
H11-12
H13-14
H14,H16
H15,H17
HH15-16
H17-18
H18,H20
H19
H19-20

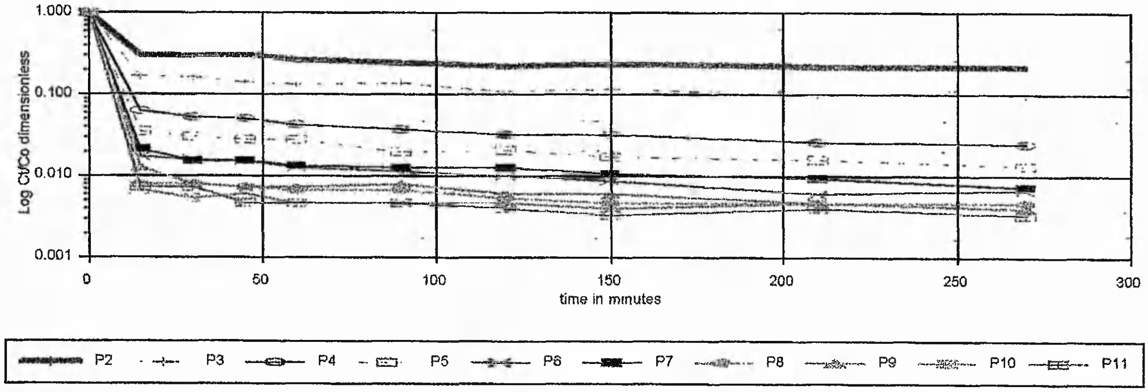
CONCENTRATION RATIO Vs DEPTH

RUN 44 KAOLIN Co=150 mg/l Q=0.4 l/min



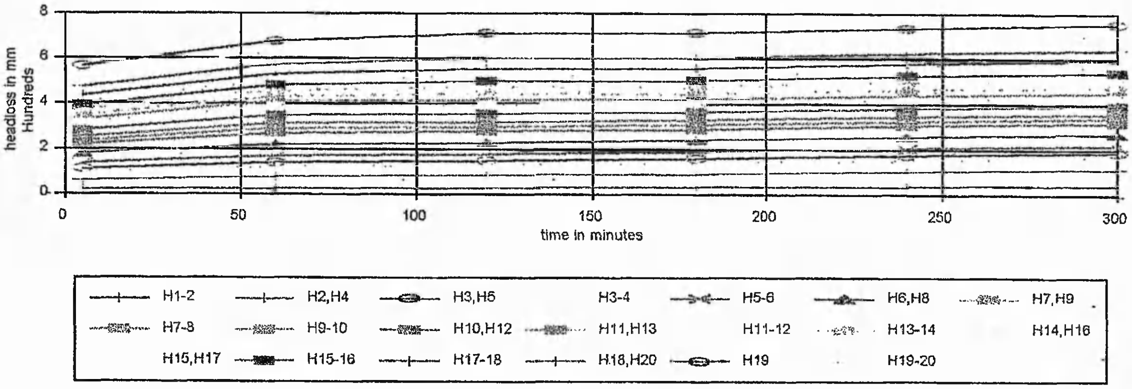
CONCENTRATION RATIO Vs TIME

RUN 44 KAOLIN Co=150 mg/l Q=0.4 l/min



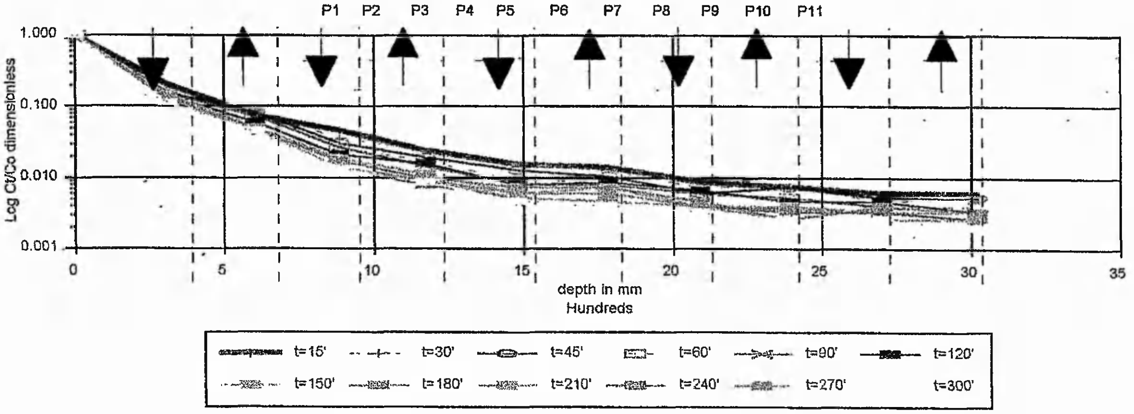
HEADLOSS Vs TIME

RUN 44 KAOLIN Co=150 mg/l Q=0.4 l/min



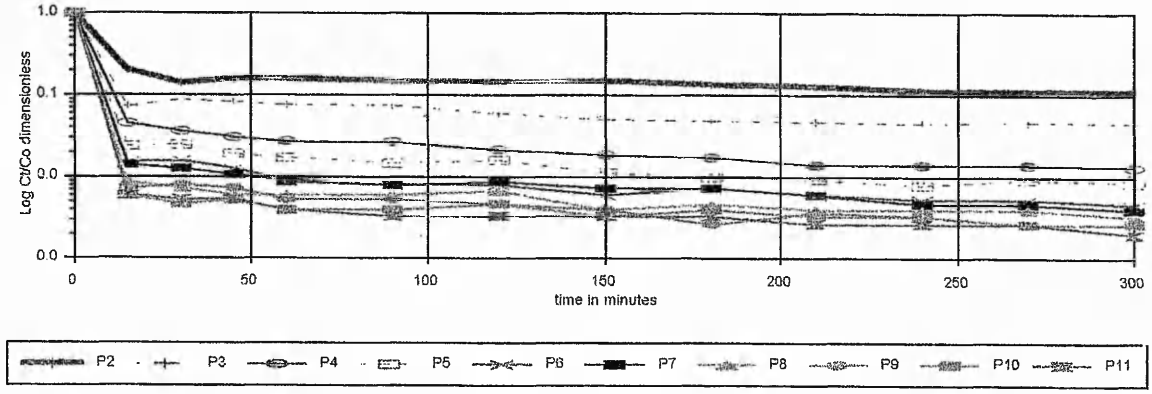
CONCENTRATION RATIO Vs DEPTH

RUN 45 KAOLIN Co=150 mg/l Q=0.2 l/min



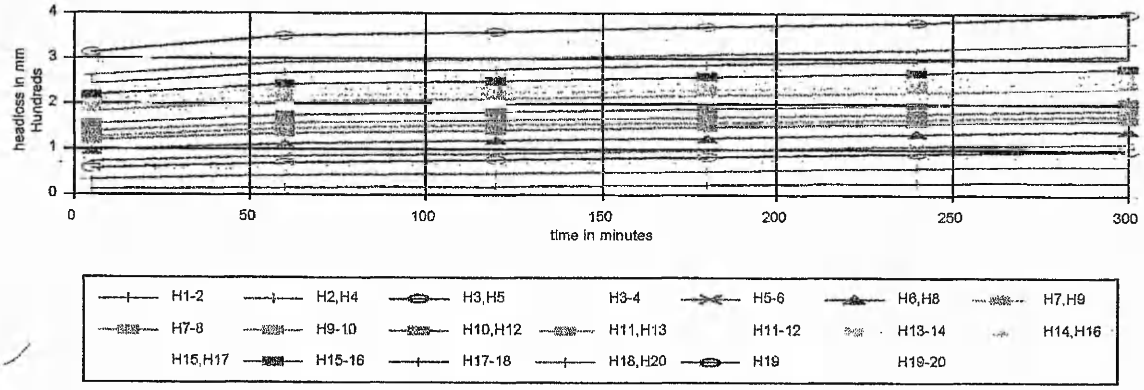
CONCENTRATION RATIO Vs TIME

RUN 45 KAOLIN Co=150 mg/l Q=0.2 l/min



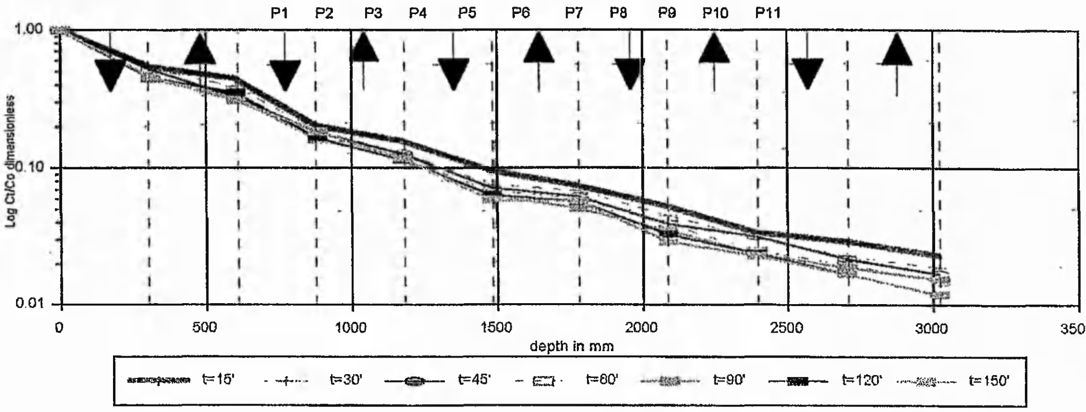
HEADLOSS Vs TIME

RUN 45 KAOLIN Co=150 mg/l Q=0.2 l/min



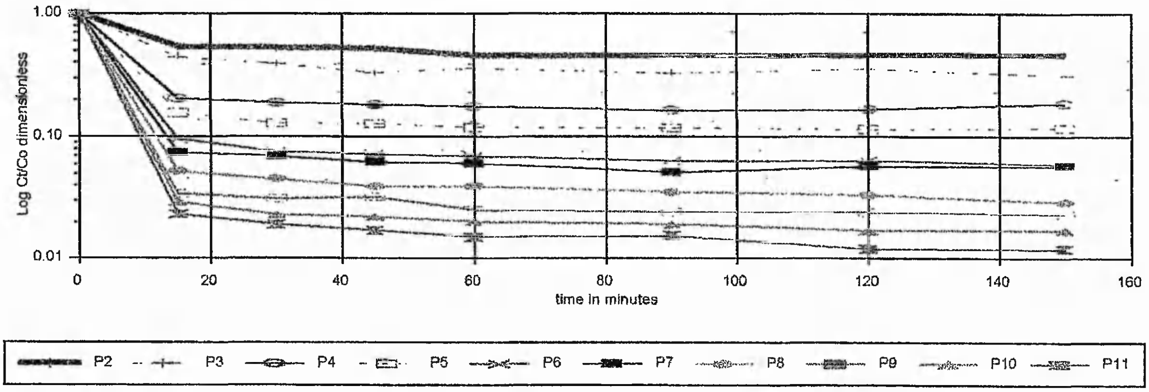
CONCENTRATION RATIO Vs DEPTH

RUN 46 KAOLIN Co=100 mg/l Q=1.0 l/min



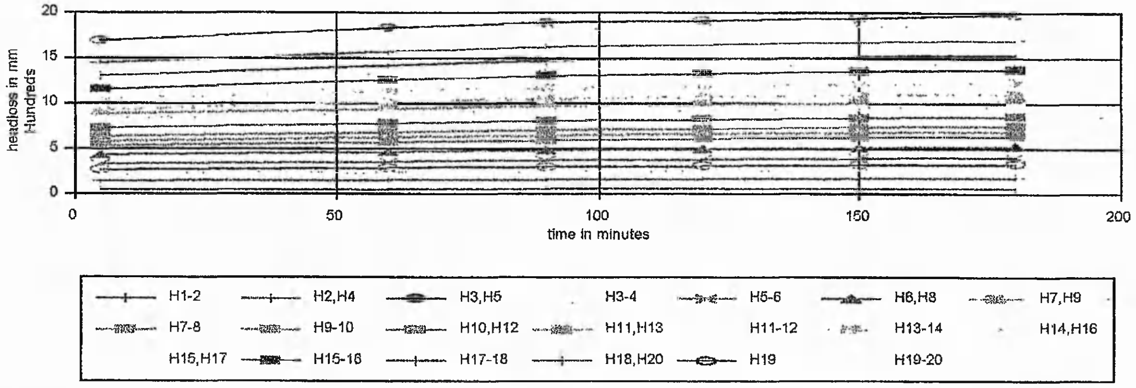
CONCENTRATION RATIO Vs TIME

RUN 46 KAOLIN Co=100 mg/l Q=1.0 l/min



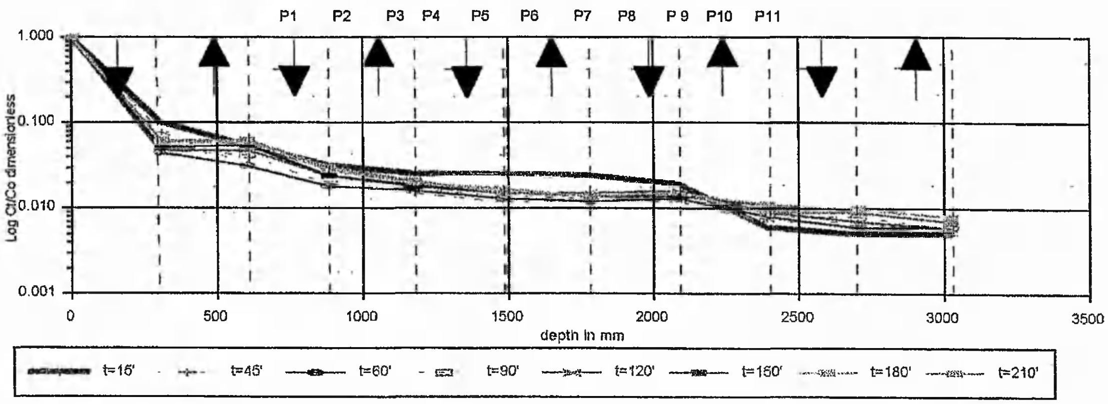
HEADLOSS Vs TIME

RUN 46 KAOLIN Co=100 mg/l Q=1.0 l/min



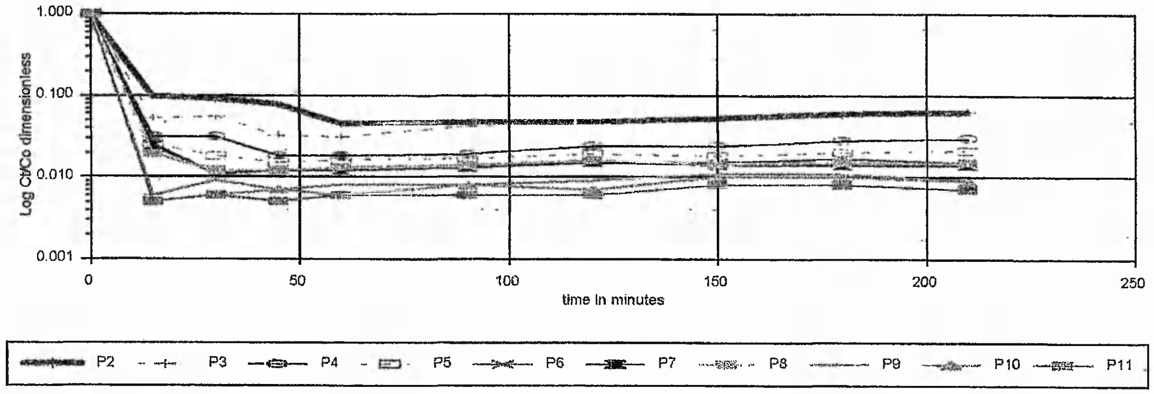
CONCENTRATION RATIO Vs DEPTH

RUN 47 KAOLIN Co=100 mg/l Q=0.8 l/min



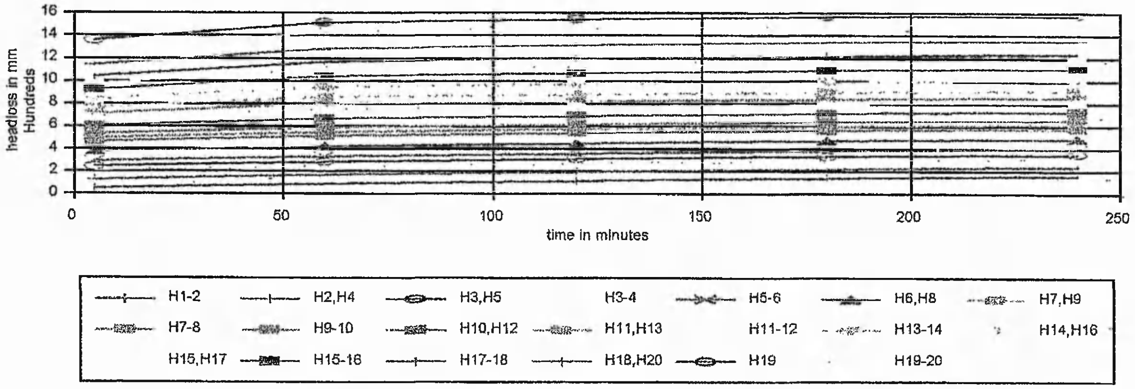
CONCENTRATION RATIO Vs TIME

RUN 47 KAOLIN Co=100 mg/l Q=0.8 l/min



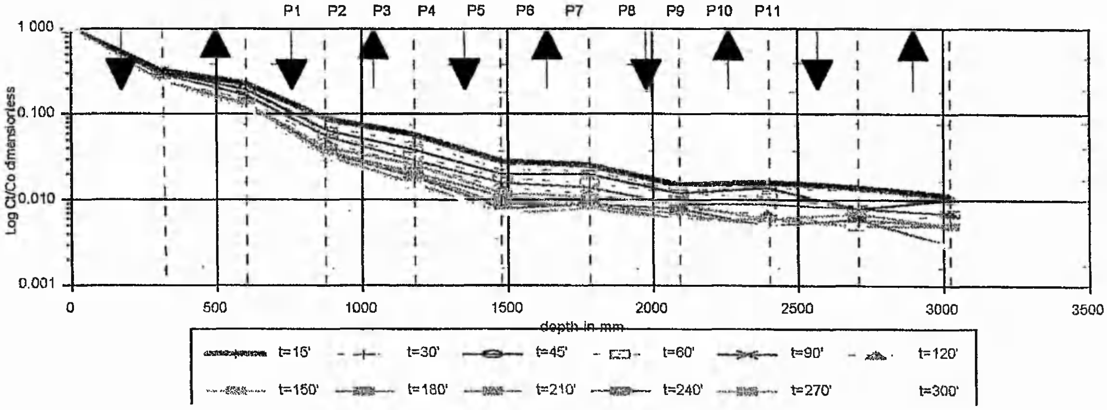
HEADLOSS Vs TIME

RUN 47 KAOLIN Co=100 mg/l Q=0.8 l/min



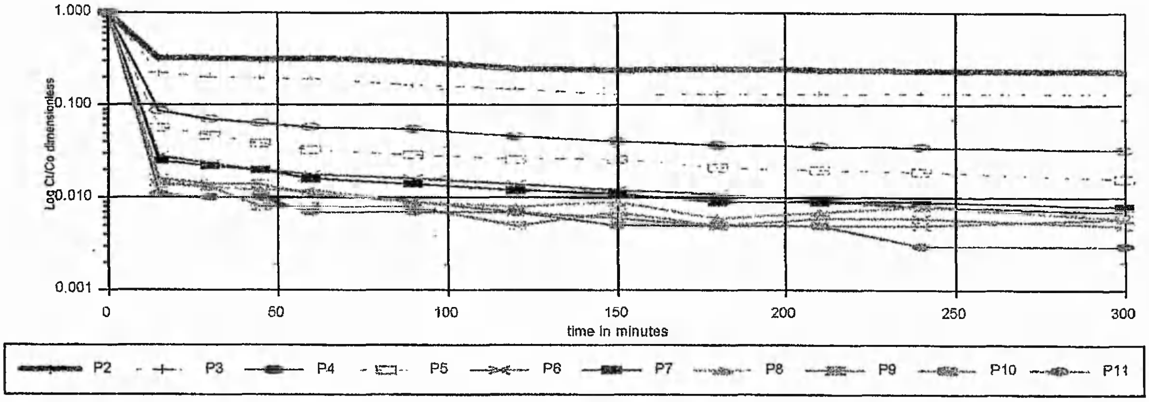
CONCENTRATION RATIO Vs DEPTH

RUN 48 KAOLIN Co=100 mg/l Q=0.6 l/min



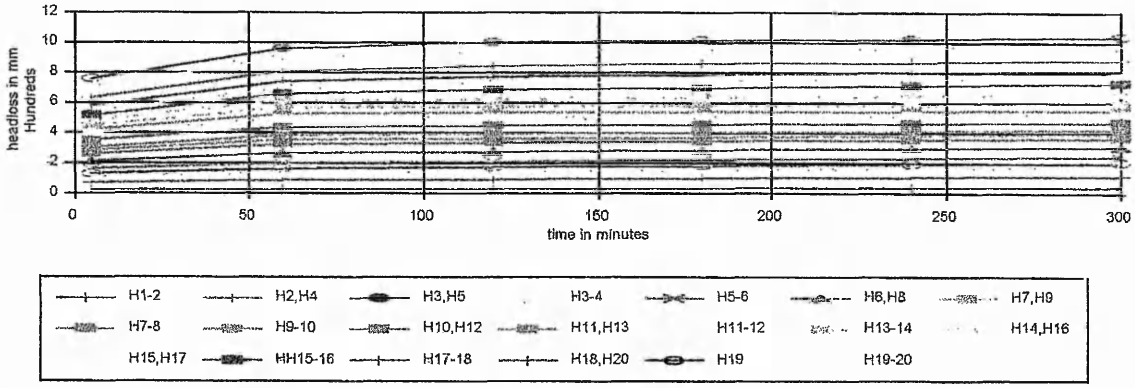
CONCENTRATION RATIO Vs TIME

RUN 48 KAOLIN Co=100 mg/l Q=0.6 l/min



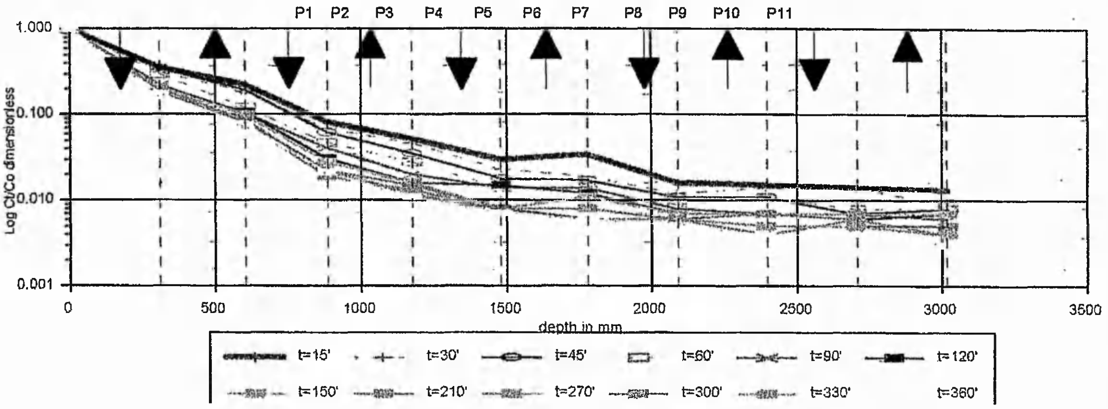
HEADLOSS Vs TIME

RUN 48 KAOLIN Co=100 mg/l Q=0.6 l/min



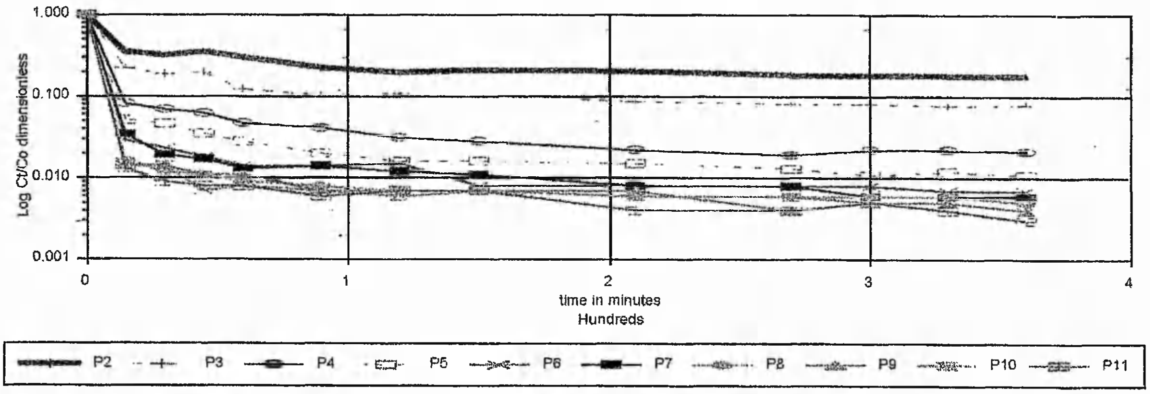
CONCENTRATION RATIO Vs DEPTH

RUN 49 KAOLIN Co=100 mg/l Q=0.4 l/min



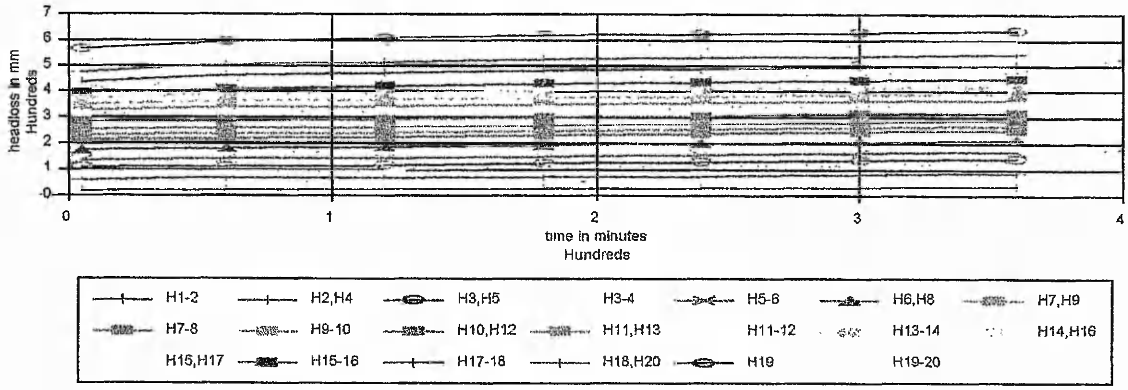
CONCENTRATION RATIO Vs TIME

RUN 49 KAOLIN Co=100 mg/l Q=0.4 l/min



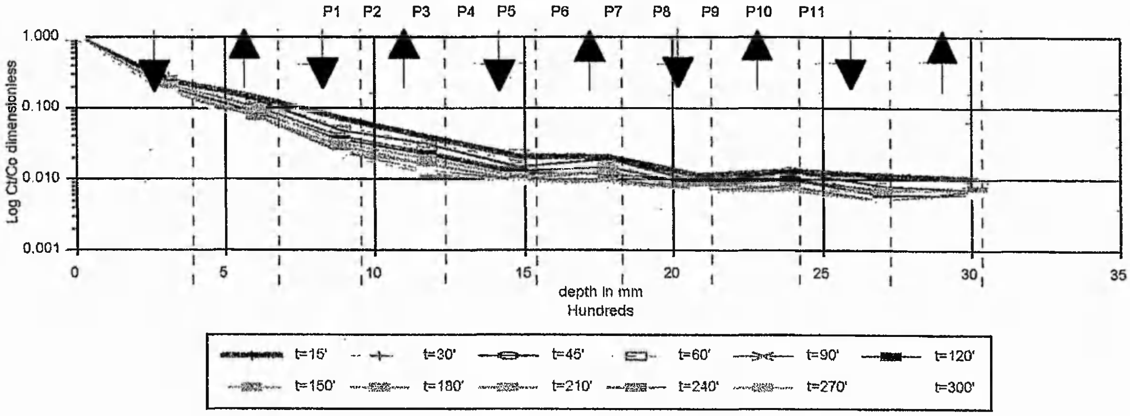
HEADLOSS Vs TIME

RUN 49 KAOLIN Co=100 mg/l Q=0.4 l/min



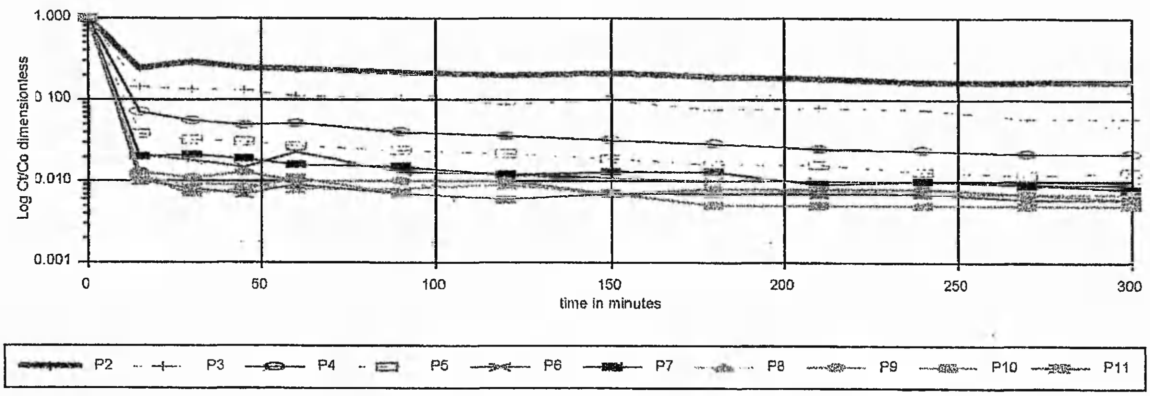
CONCENTRATION RATIO Vs DEPTH

RUN 50 KAOLIN Co=100 mg/l Q=0.2 l/min



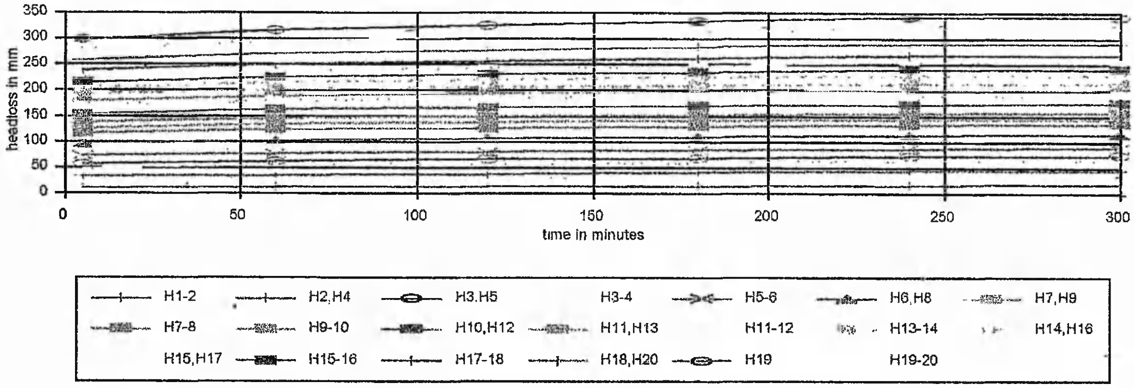
CONCENTRATION RATIO Vs TIME

RUN 50 KAOLIN Co=100 mg/l Q=0.2 l/min



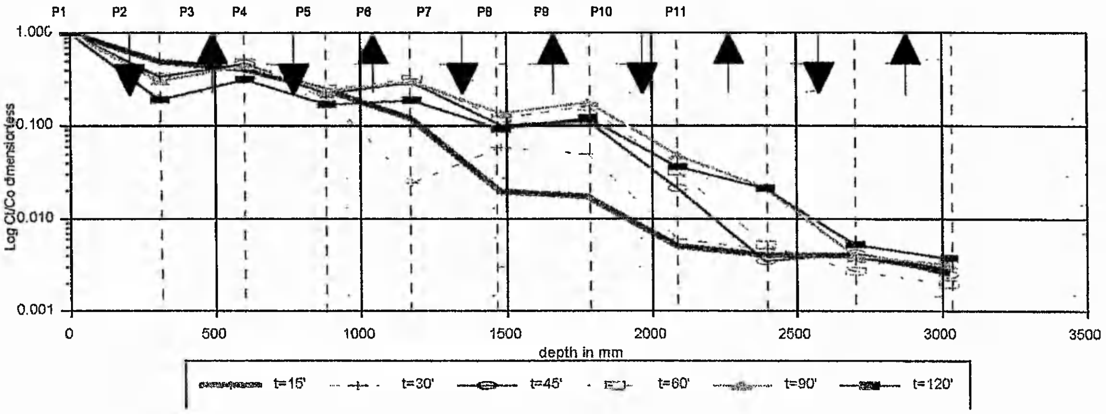
HEADLOSS Vs TIME

RUN 50 KAOLIN Co=100 mg/l Q=0.2 l/min



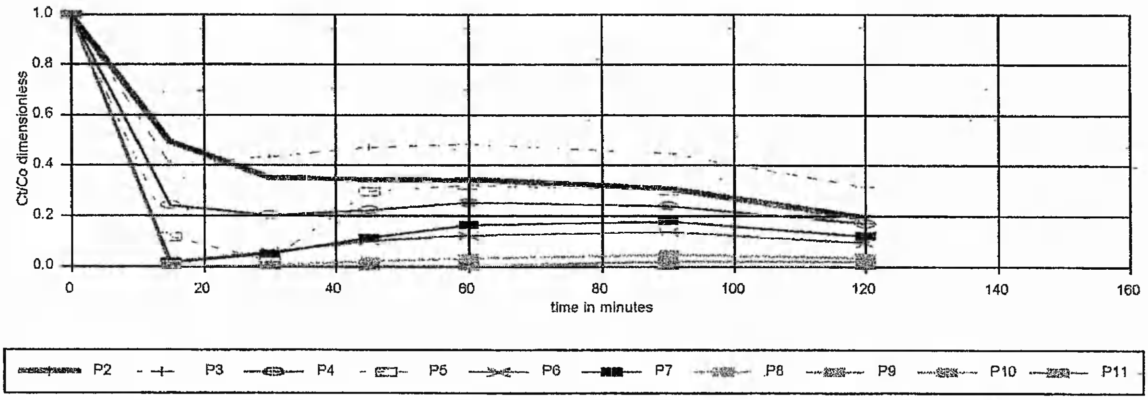
CONCENTRATION RATIO Vs DEPTH

RUN 51 LYCOPODIUM Co=400 mg/l Q=1.0 l/min



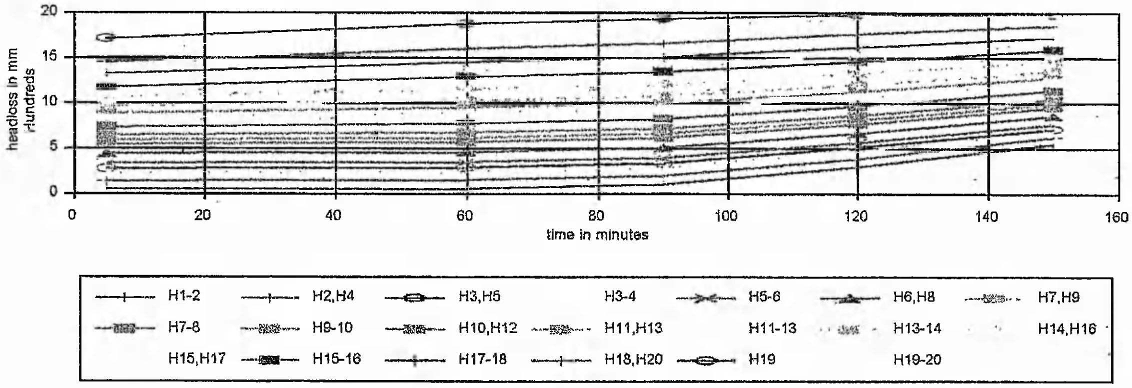
CONCENTRATION RATIO Vs TIME

LYCOPODIUM Co=400 mg/l Q=1.0 l/min RUN 51



HEADLOSS Vs TIME

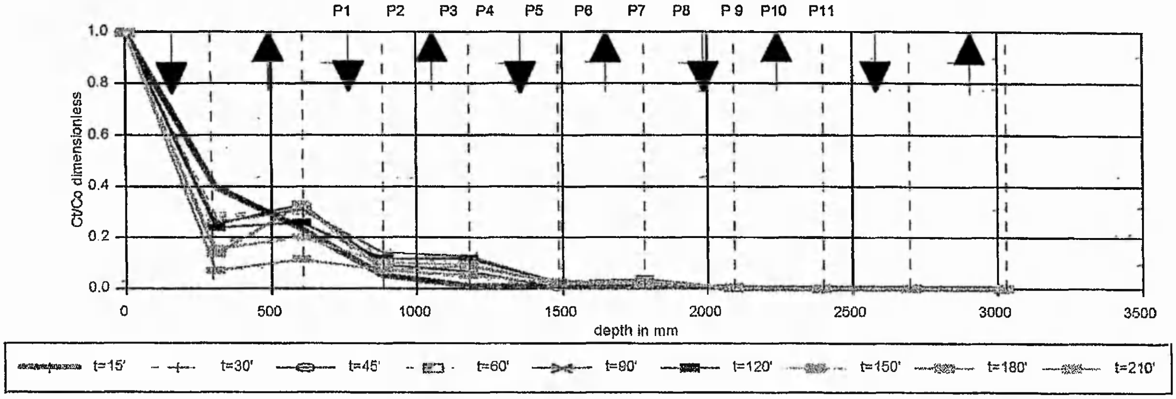
LYCOPODIUM Co=400 mg/l Q=1.0 l/min RUN 51



CONCENTRATION RATIO Vs DEPTH

RUN 52

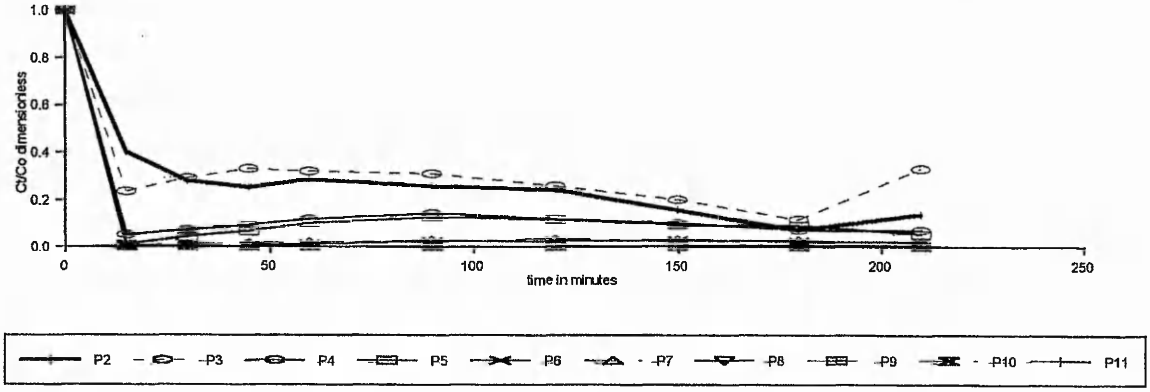
LYCOPODIUM Co=400 mg/l Q=0.8 l/min



CONCENTRATION RATIO Vs TIME

LYCOPODIUM Co=400 mg/l Q=0.8 l/min

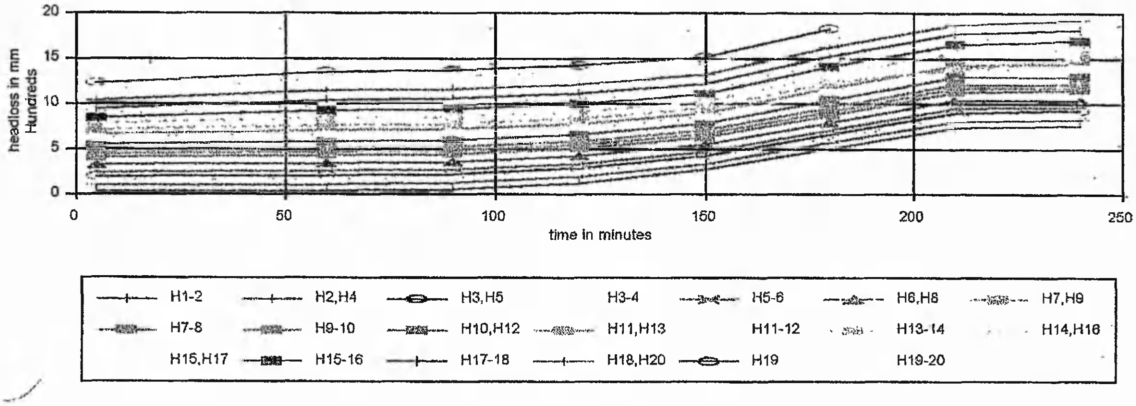
RUN 52



HEADLOSS Vs TIME

LYCOPODIUM Co=400 mg/l Q=0.8 l/min

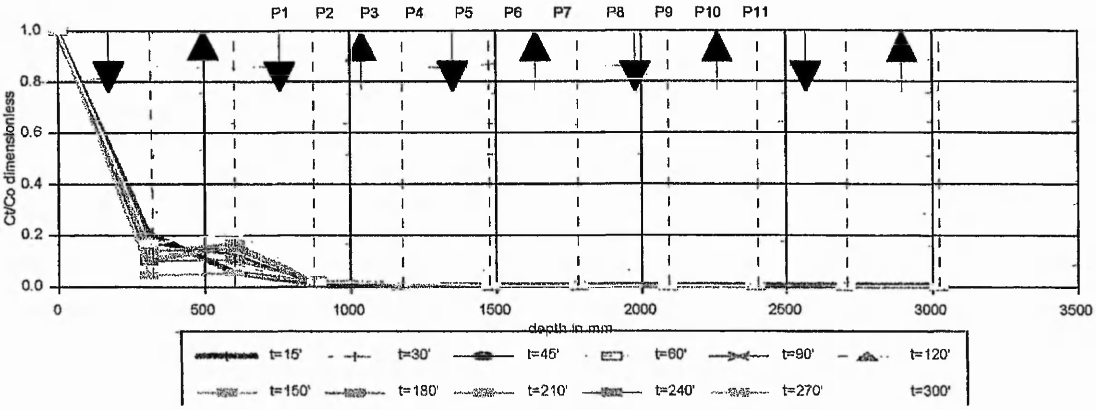
RUN 52



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

RUN 53 Co=400 mg/l Q=0.6 l/min

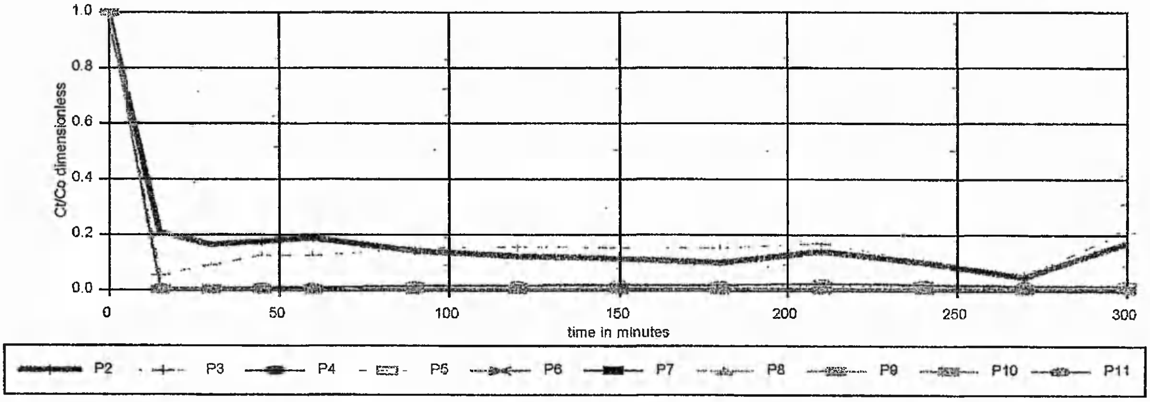


CONCENTRATION RATIO Vs TIME

LYCOPODIUM

RUN 53 Co=400 mg/l Q=0.6 l/min

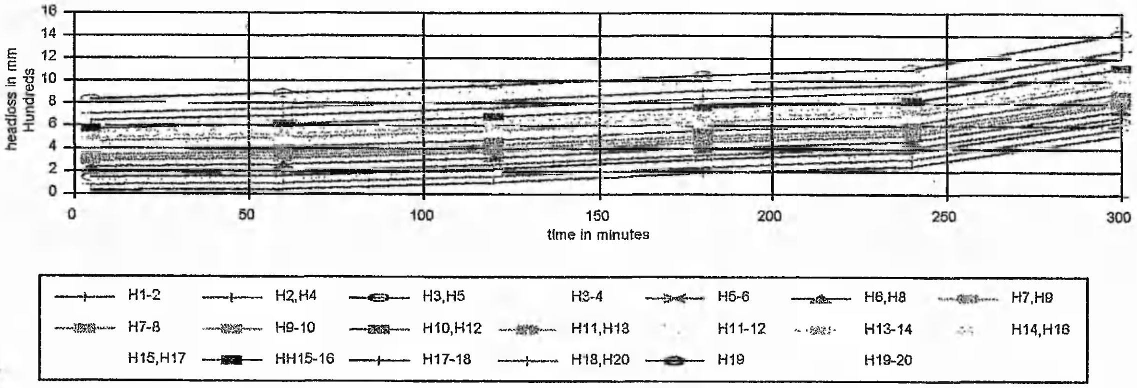
LYCOPODIUM



HEADLOSS Vs TIME

RUN 53 Co=400 mg/l Q=0.6 l/min

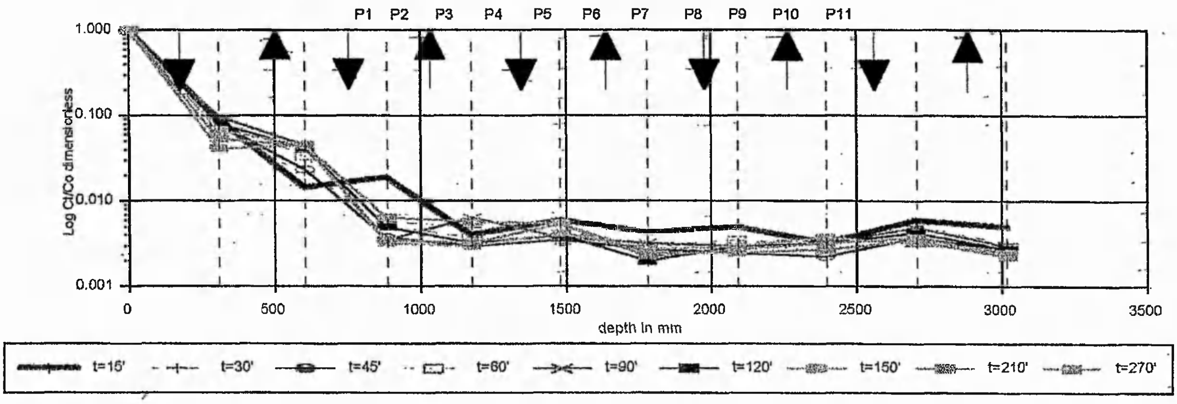
LYCOPODIUM



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

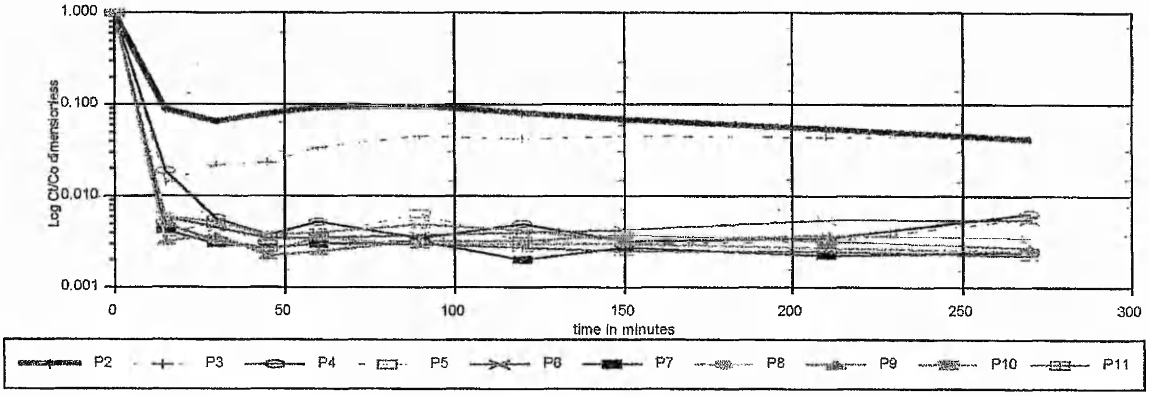
RUN 54 Co=400 mg/l Q=0.4 l/min



CONCENTRATION RATIO Vs TIME

RUN 54 Co=400 mg/l Q=0.4 l/min

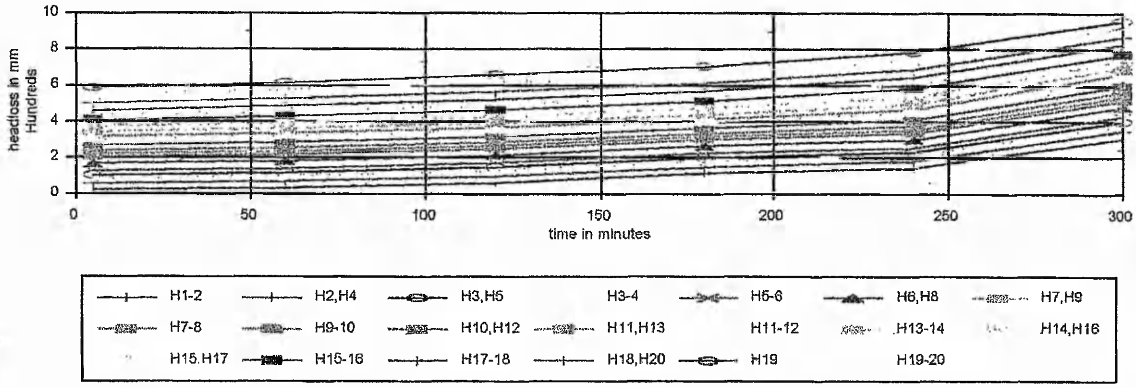
LYCOPODIUM

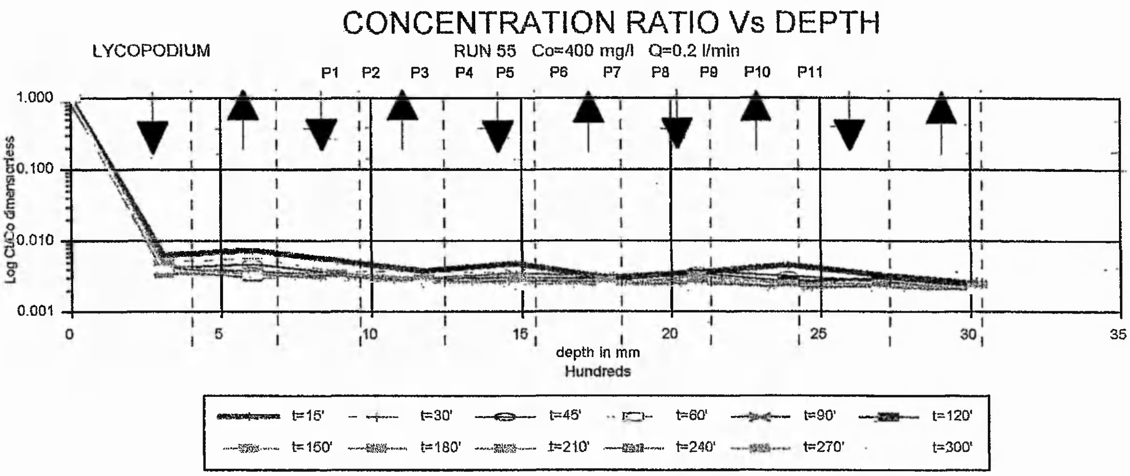


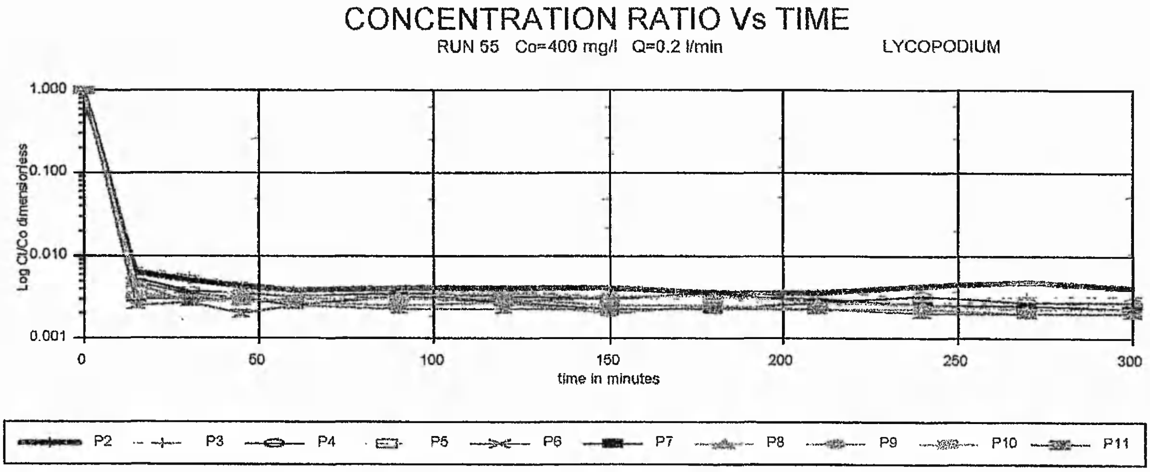
HEADLOSS Vs TIME

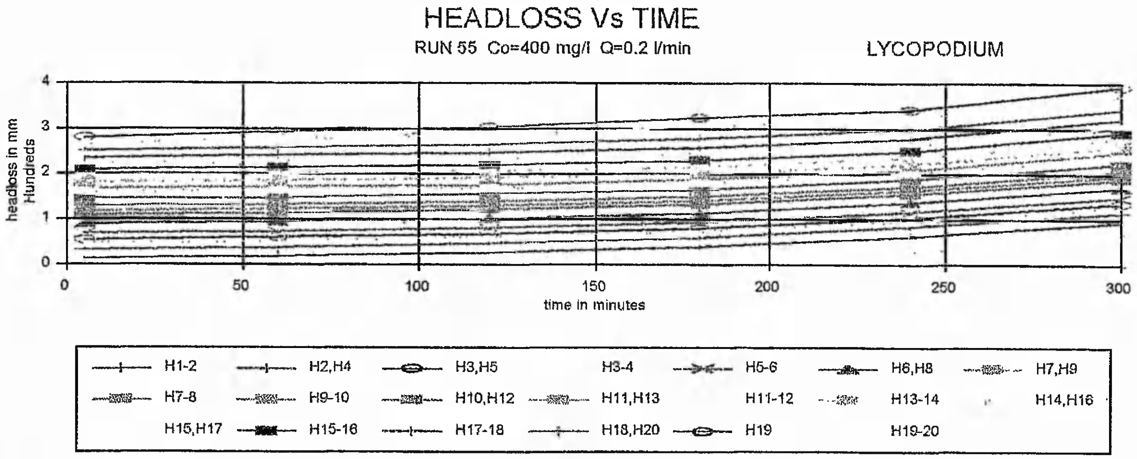
RUN 54 Co=400 mg/l Q=0.4 l/min

LYCOPODIUM



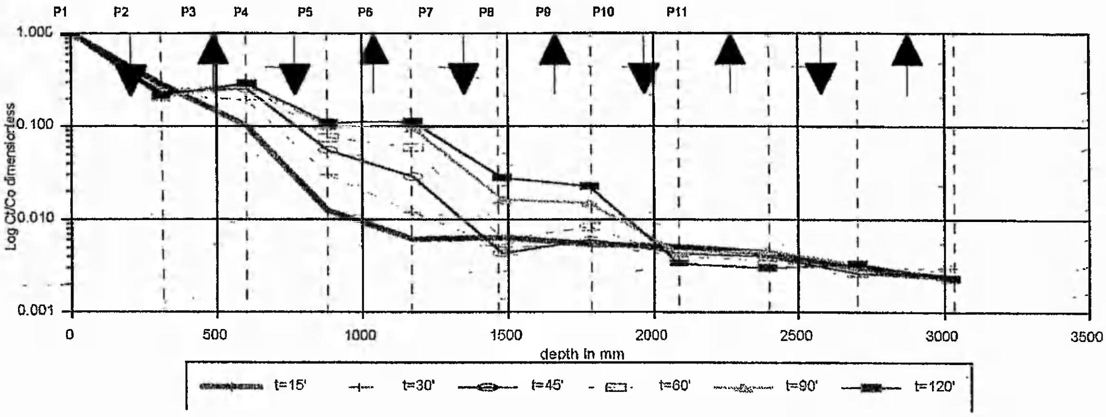

CONCENTRATION RATIO Vs DEPTH
LYCOPODIUM
RUN 55 Co=400 mg/l Q=0.2 l/min
P1
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11
Log Ct/Co dimensionless
1.000
0.100
0.010
0.001
0
5
10
15
20
25
30
35
depth in mm
Hundreds
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
t=150'
t=180'
t=210'
t=240'
t=270'
t=300'


CONCENTRATION RATIO Vs TIME
RUN 55 Co=400 mg/l Q=0.2 l/min
LYCOPODIUM
Log Ct/Co dimensionless
1.000
0.100
0.010
0.001
0
50
100
150
200
250
300
time in minutes
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11


HEADLOSS Vs TIME
RUN 55 Co=400 mg/l Q=0.2 l/min
LYCOPODIUM
headloss in mm
Hundreds
4
3
2
1
0
0
50
100
150
200
250
300
time in minutes
H1-2
H2,H4
H3,H5
H3-4
H5-6
H6,H8
H7,H9
H7-8
H9-10
H10,H12
H11,H13
H11-12
H13-14
H14,H16
H15,H17
H15-16
H17-18
H18,H20
H19
H19-20

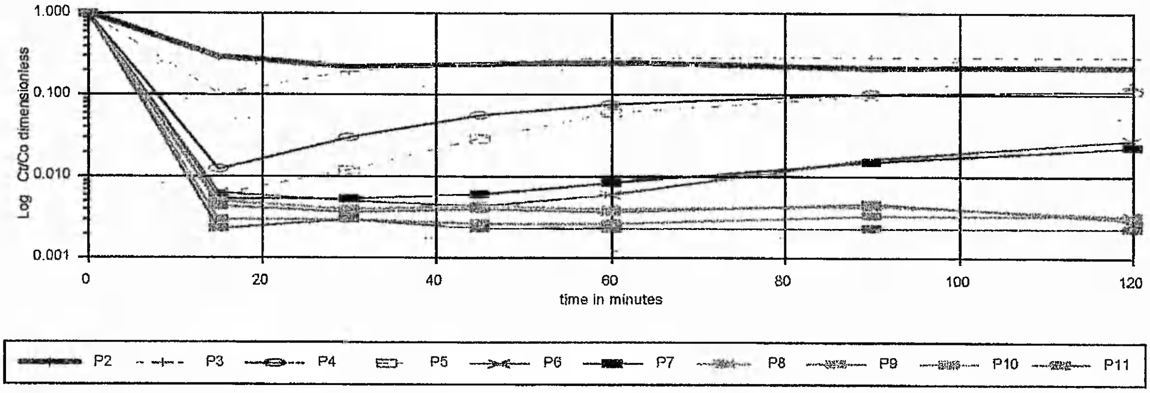
CONCENTRATION RATIO Vs DEPTH

RUN 56 LYCOPODIUM Co=300 mg/l Q=1.0 l/min



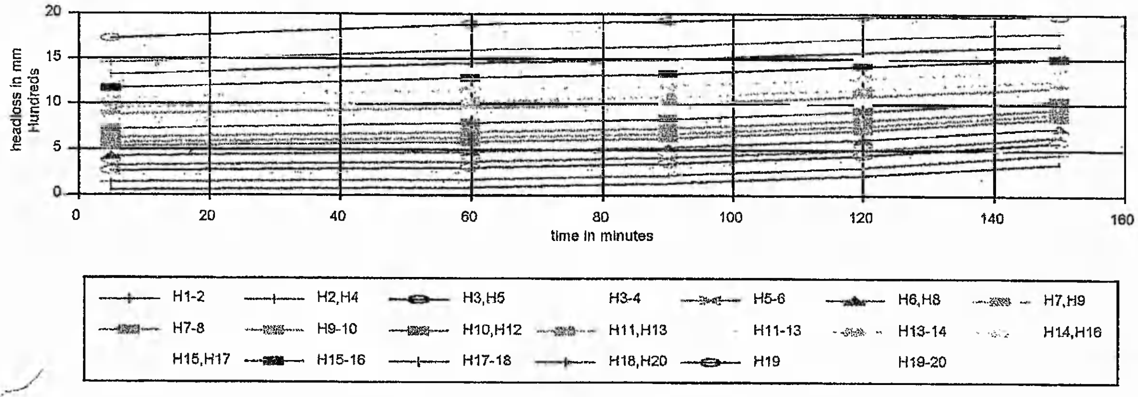
CONCENTRATION RATIO Vs TIME

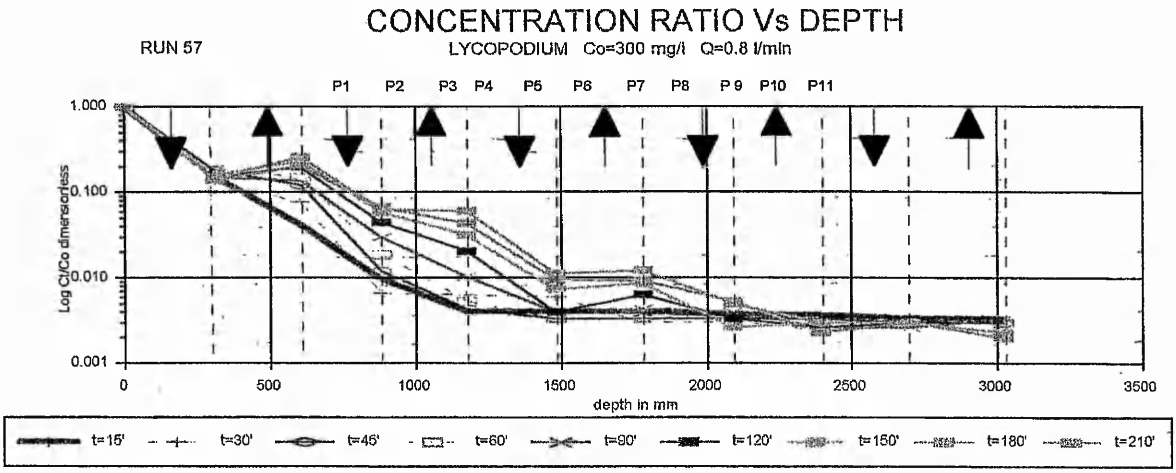
RUN 26 KAOLIN Co=400 mg/l Q=1.0 l/min

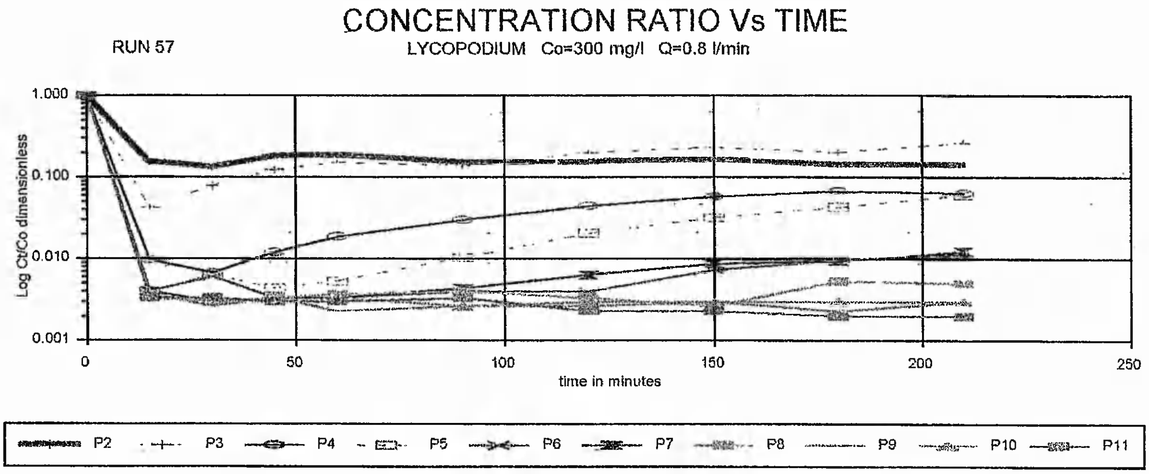


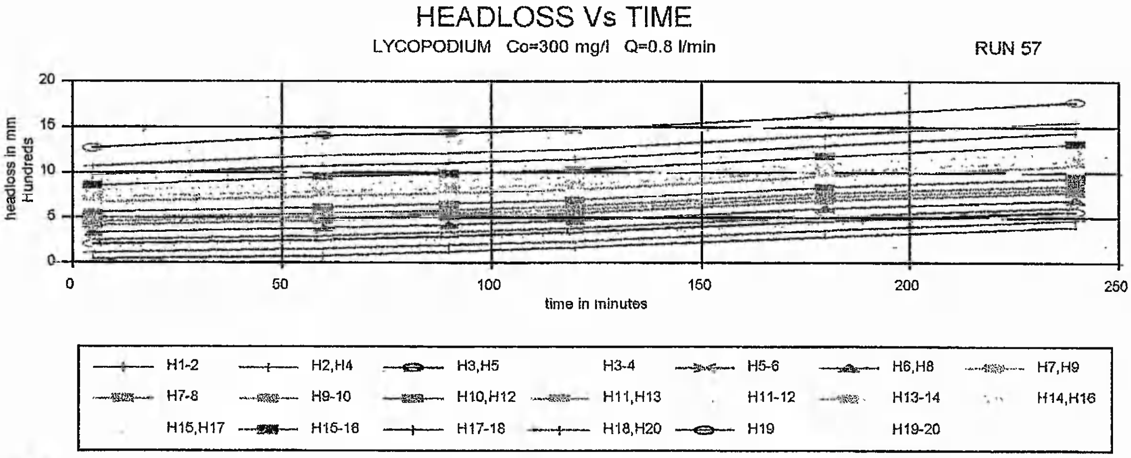
HEADLOSS Vs TIME

LYCOPODIUM Co=300 mg/l Q=1.0 l/min RUN 56




CONCENTRATION RATIO Vs DEPTH
RUN 57
LYCOPODIUM Co=300 mg/l Q=0.8 l/min
P1
P2
P3
P4
P5
P6
P7
P8
P 9
P10
P11
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
500
1000
1500
2000
2500
3000
3500
depth in mm
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
t=150'
t=180'
t=210'

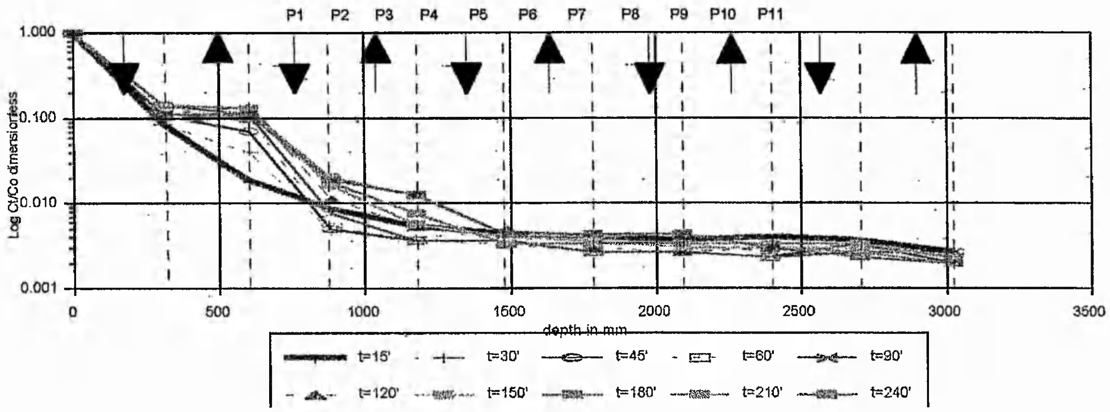

CONCENTRATION RATIO Vs TIME
RUN 57
LYCOPODIUM Co=300 mg/l Q=0.8 l/min
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
50
100
150
200
250
time in minutes
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11


HEADLOSS Vs TIME
LYCOPODIUM Co=300 mg/l Q=0.8 l/min
RUN 57
20
15
10
5
0
headloss in mm
Hundreds
0
50
100
150
200
250
time in minutes
H1-2
H2,H4
H3,H5
H3-4
H5-6
H6,H8
H7,H9
H7-8
H9-10
H10,H12
H11,H13
H11-12
H13-14
H14,H16
H15,H17
H15-16
H17-18
H18,H20
H19
H19-20

CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

RUN 58 Co=300 mg/l Q=0.6 l/min

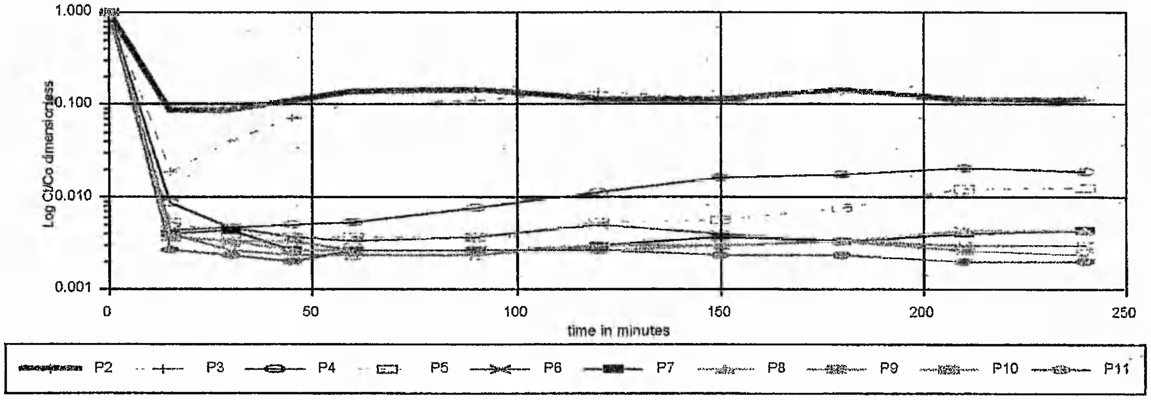


CONCENTRATION RATIO Vs TIME

LYCOPODIUM

RUN 58 Co=300 mg/l Q=0.6 l/min

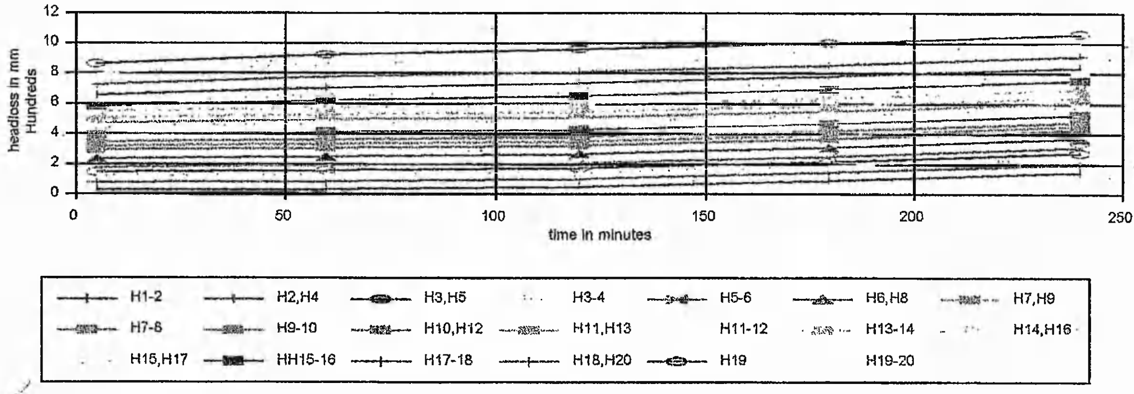
LYCOPODIUM



HEADLOSS Vs TIME

RUN 58 Co=300 mg/l Q=0.6 l/min

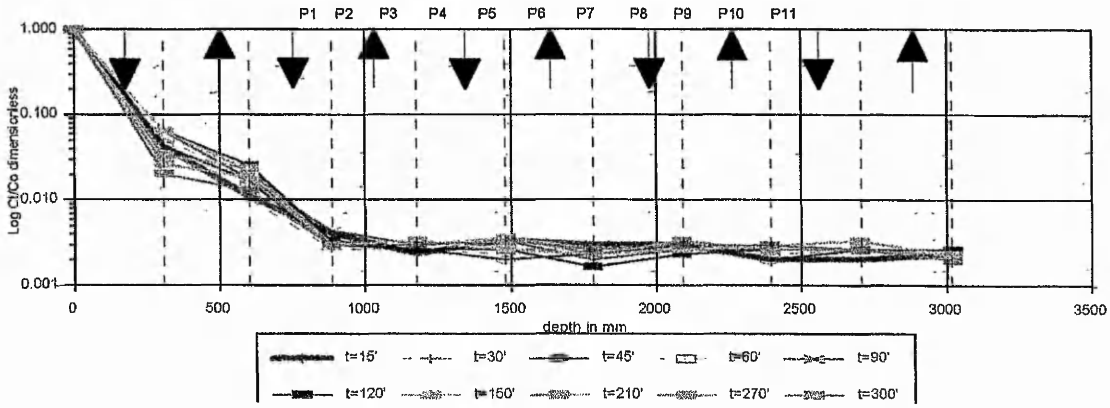
LYCOPODIUM



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

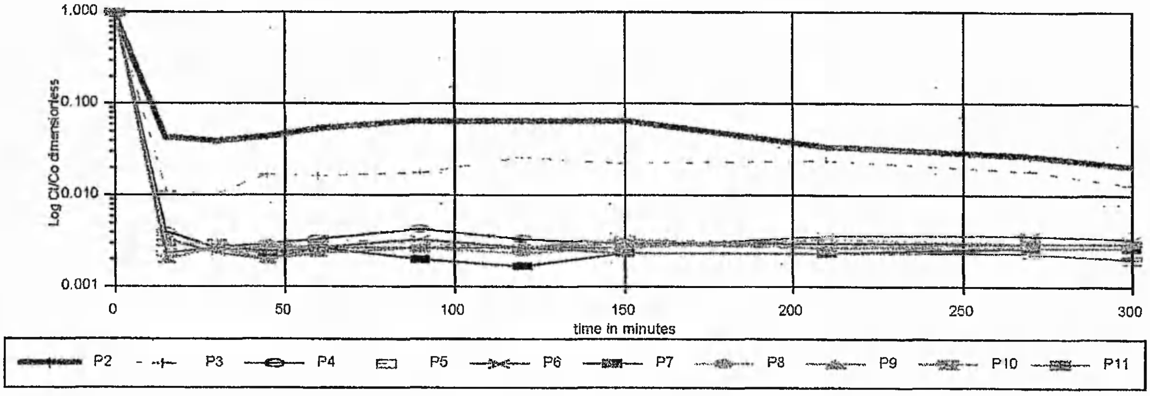
RUN 59 Co=300 mg/l Q=0.4 l/min



CONCENTRATION RATIO Vs TIME

RUN 59 Co=300 mg/l Q=0.4 l/min

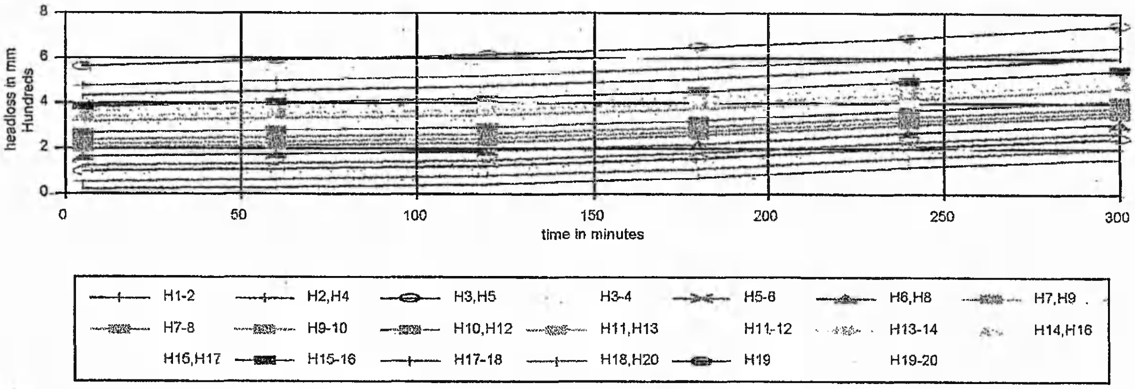
LYCOPODIUM



HEADLOSS Vs TIME

RUN 59 Co=300 mg/l Q=0.4 l/min

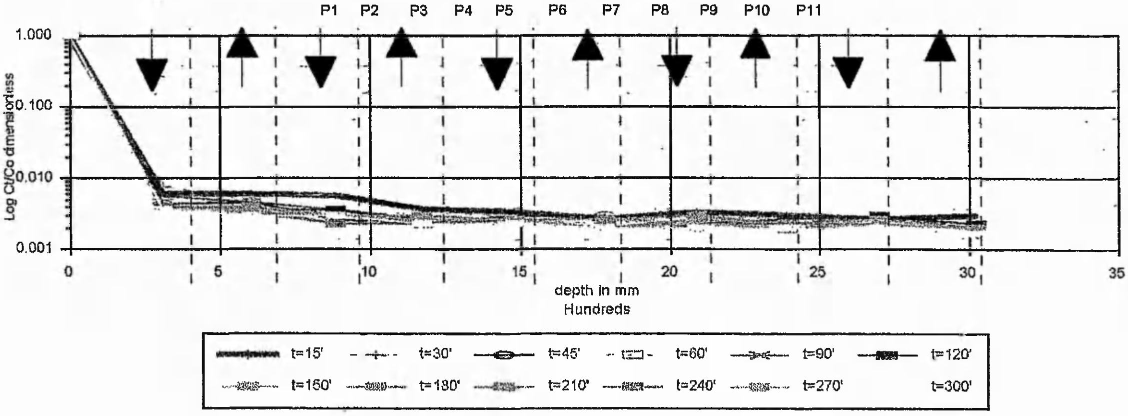
LYCOPODIUM



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

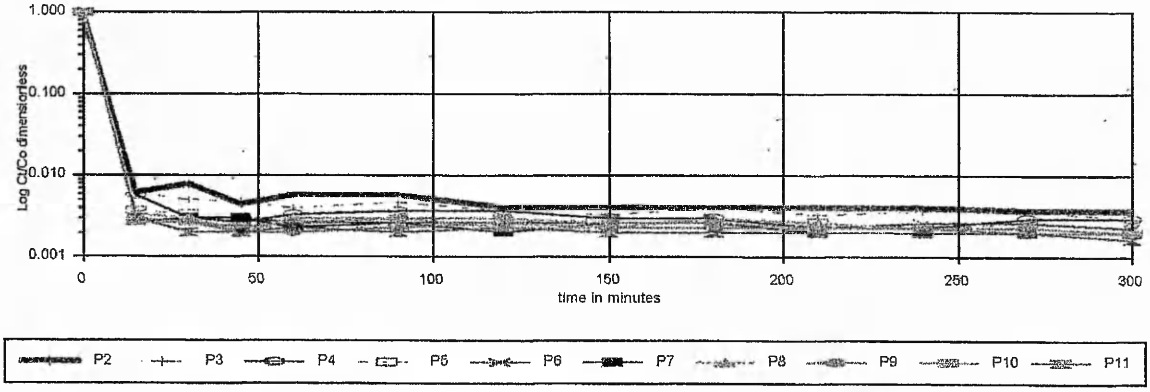
RUN 60 Co=300 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

RUN 60 Co=300 mg/l Q=0.2 l/min

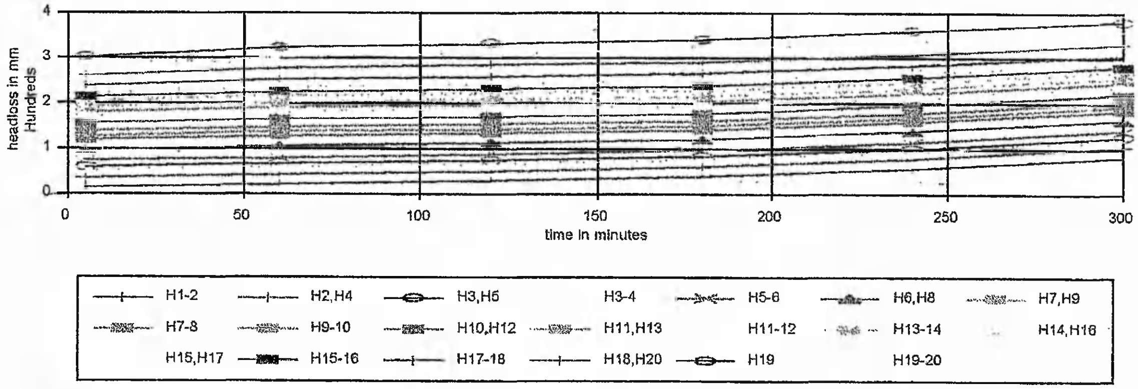
LYCOPODIUM



HEADLOSS Vs TIME

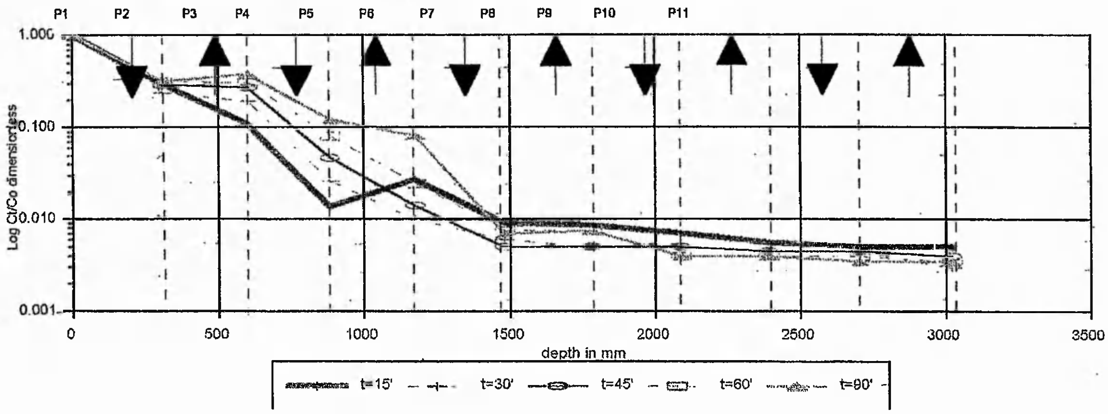
RUN 60 Co=300 mg/l Q=0.2 l/min

LYCOPODIUM



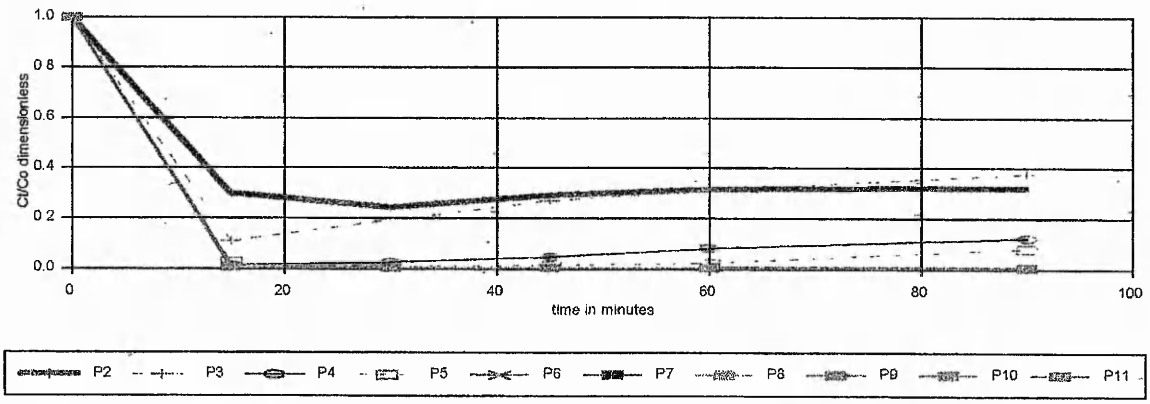
CONCENTRATION RATIO Vs DEPTH

RUN 61 LYCOPODIUM Co=200 mg/l Q=1.0 l/min



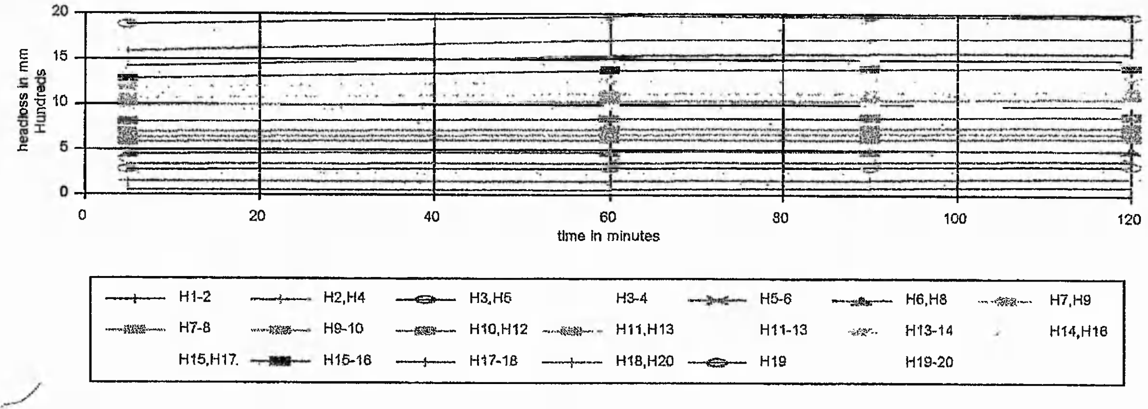
CONCENTRATION RATIO Vs TIME

LYCOPODIUM Co=200 mg/l Q=1 0 l/min RUN 61



HEADLOSS Vs TIME

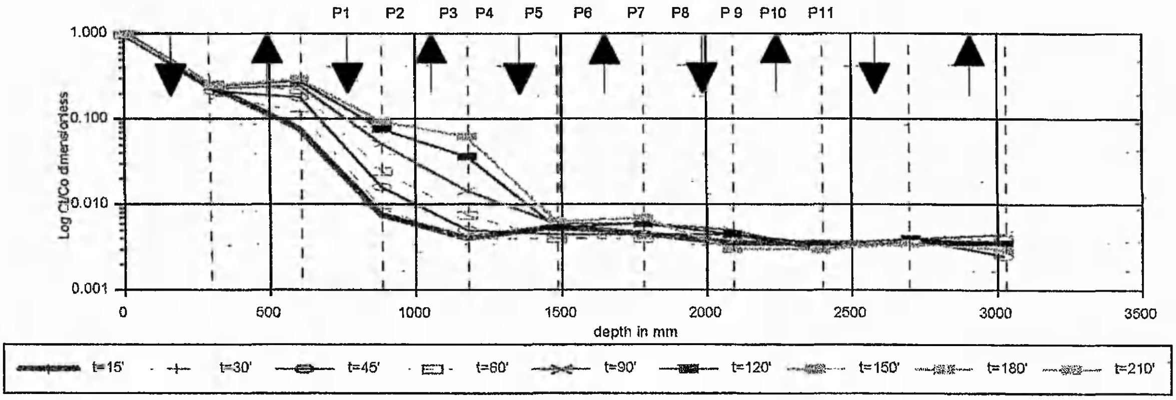
LYCOPODIUM Co=200 mg/l Q=1.0 l/min RUN 61



CONCENTRATION RATIO Vs DEPTH

RUN 62

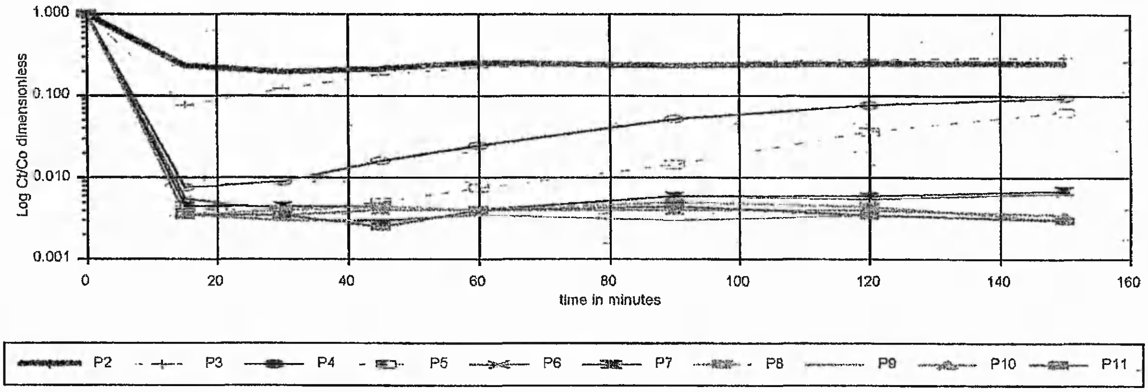
LYCOPODIUM Co=200 mg/l Q=0.8 l/min



CONCENTRATION RATIO Vs TIME

RUN 62

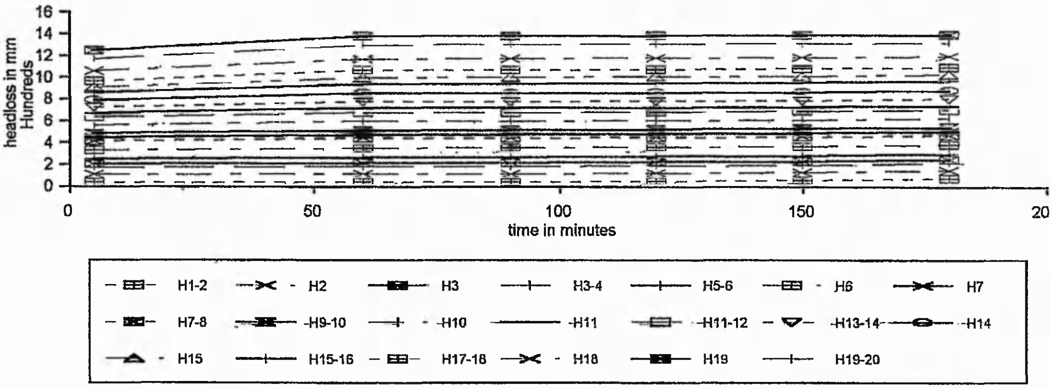
LYCOPODIUM Co=200 mg/l Q=0.8 l/min



HEADLOSS Vs TIME

Lycopodium Co=200 mg/l Q=0.8 l/min

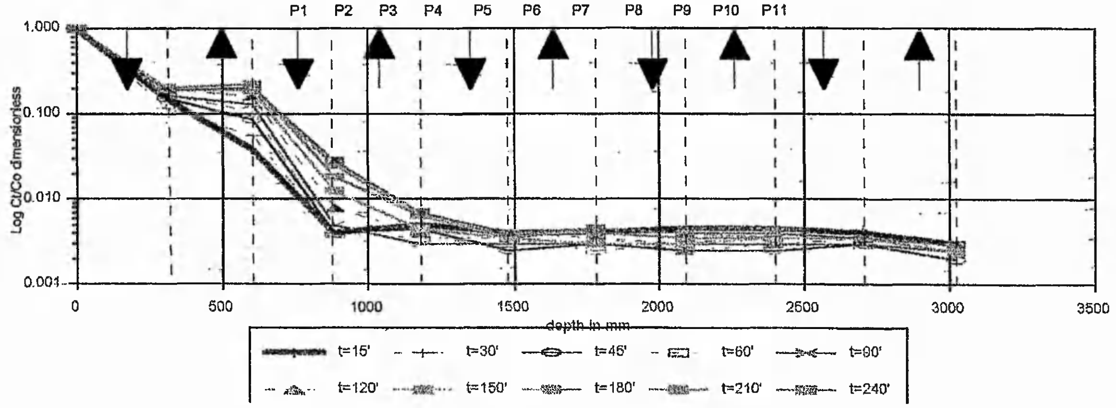
RUN 62



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

RUN 63 Co=200 mg/l Q=0.6 l/min

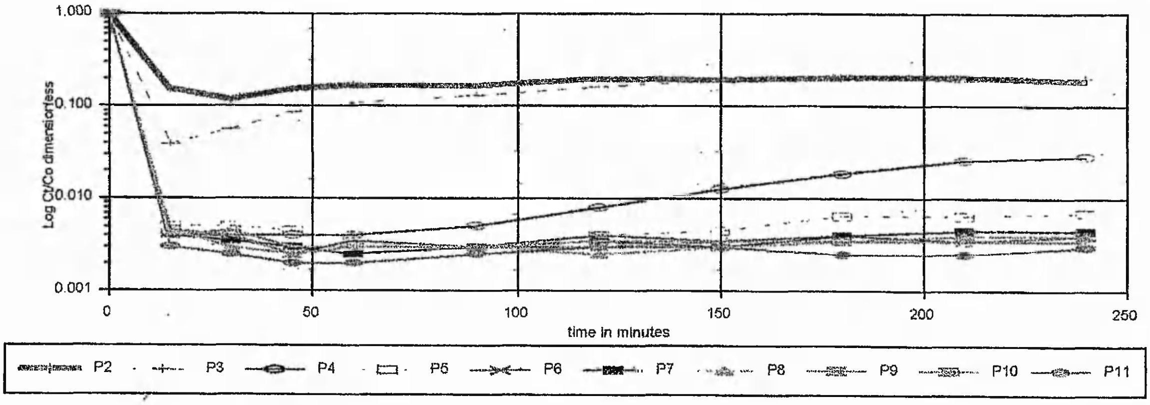


CONCENTRATION RATIO Vs TIME

LYCOPODIUM

RUN 63 Co=200 mg/l Q=0.6 l/min

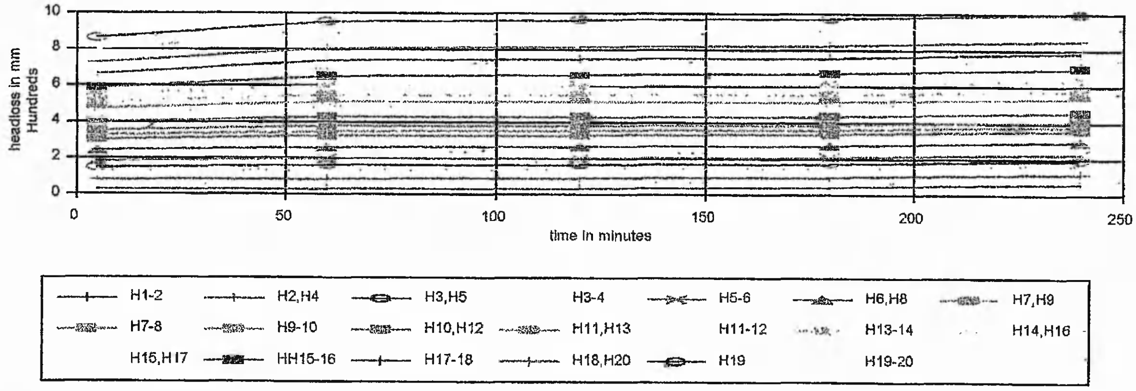
LYCOPODIUM



HEADLOSS Vs TIME

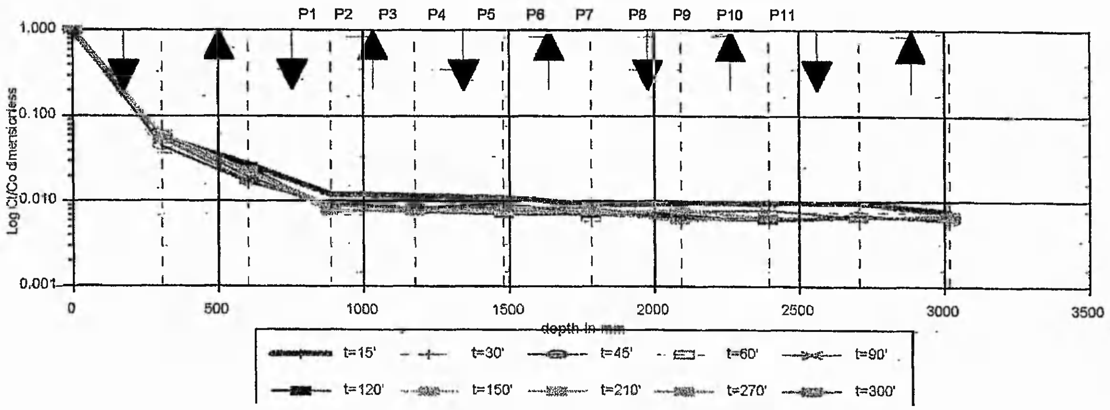
RUN 63 Co=200 mg/l Q=0.6 l/min

LYCOPODIUM



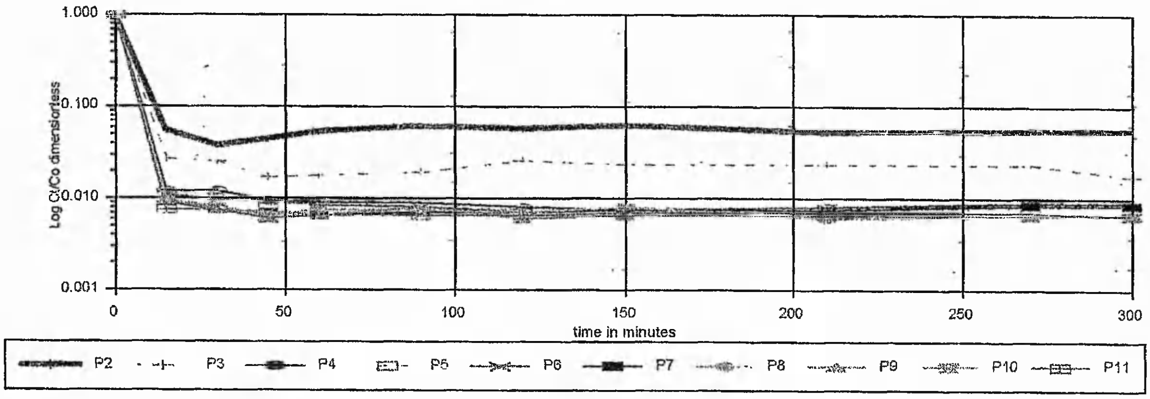
CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM RUN 64 Co=200 mg/l Q=0.4 l/min



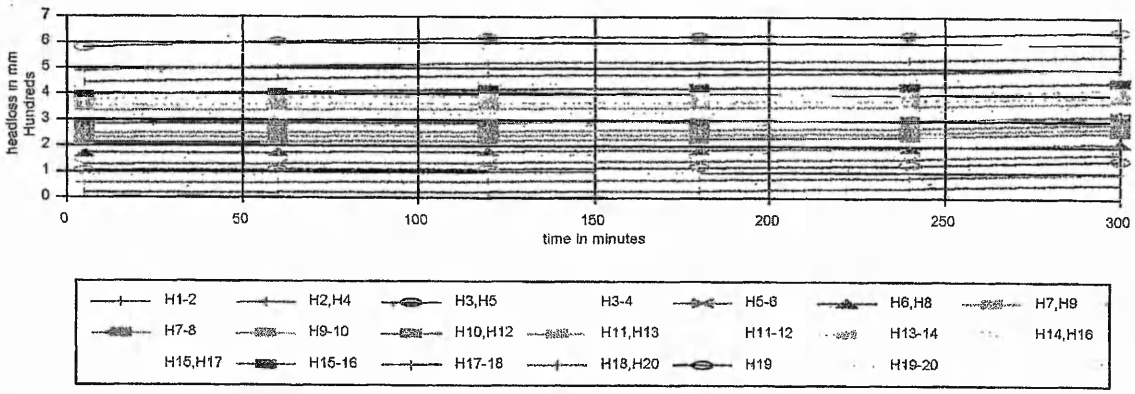
CONCENTRATION RATIO Vs TIME

RUN 64 Co=200 mg/l Q=0.4 l/min LYCOPODIUM



HEADLOSS Vs TIME

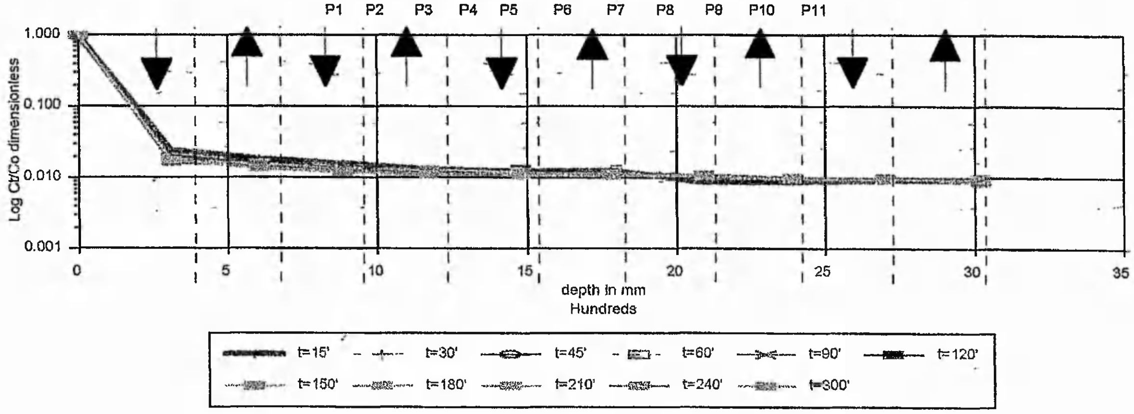
RUN 64 Co=200 mg/l Q=0.4 l/min LYCOPODIUM



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

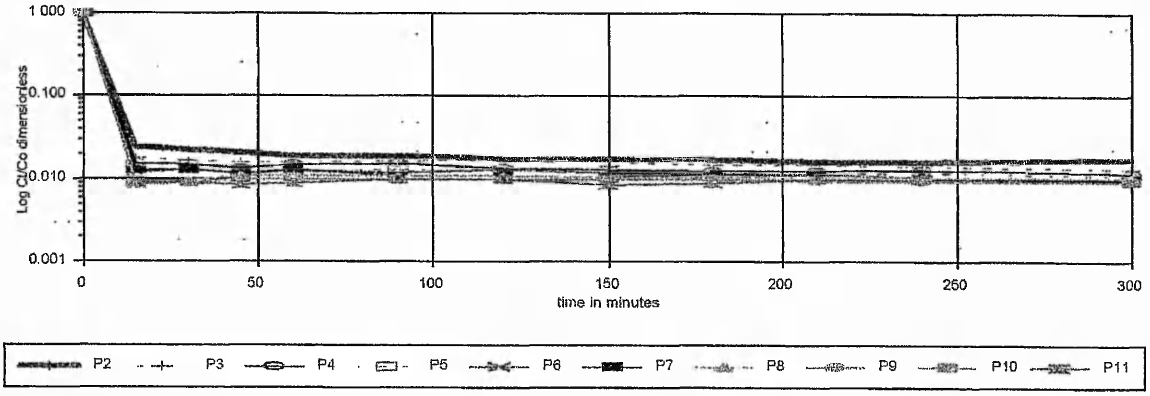
RUN 65 Co=200 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

RUN 65 Co=200 mg/l Q=0.2 l/min

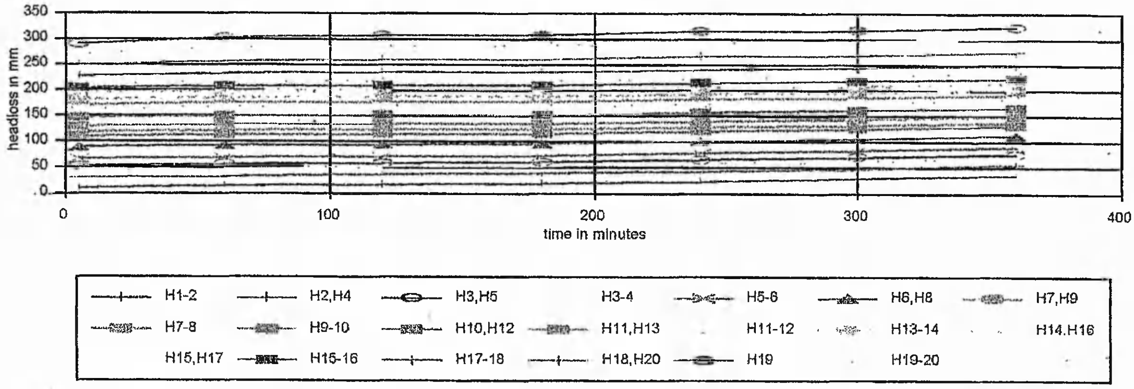
LYCOPODIUM



HEADLOSS Vs TIME

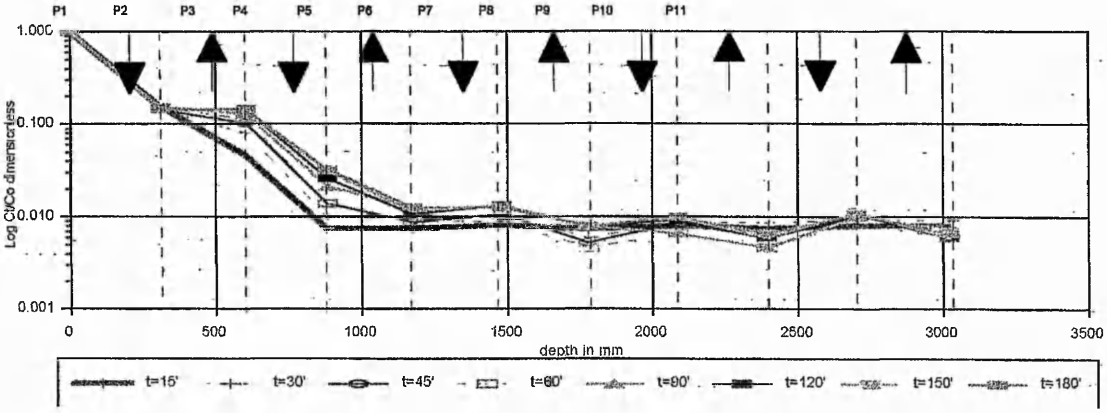
RUN 65 Co=200 mg/l Q=0.2 l/min

LYCOPODIUM



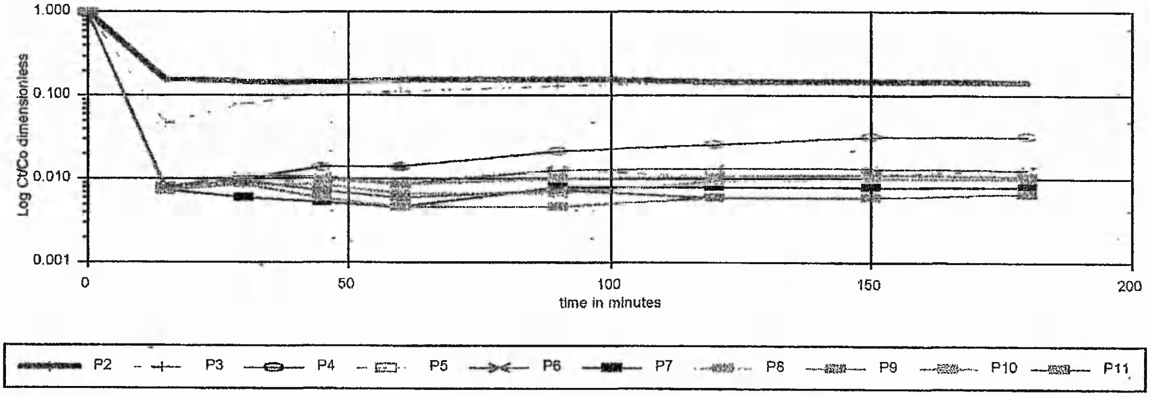
CONCENTRATION RATIO Vs DEPTH

RUN 66 LYCOPODIUM Co=150 mg/l Q=1.0 l/min



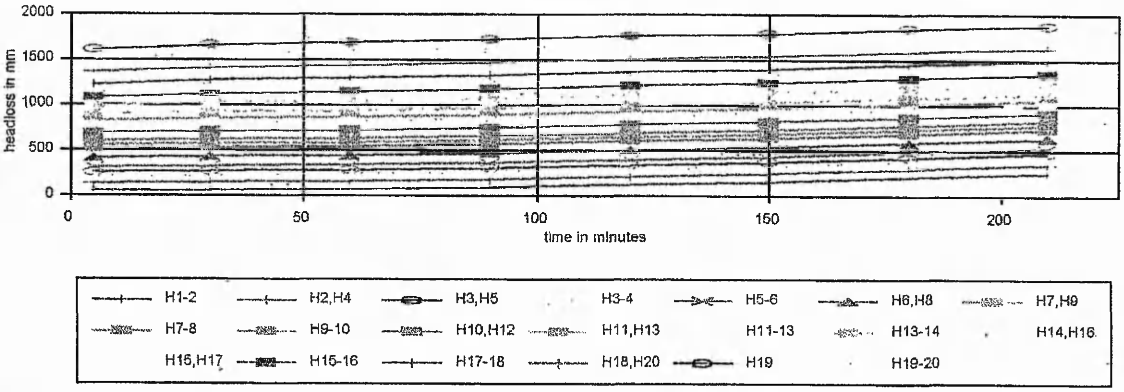
CONCENTRATION RATIO Vs TIME

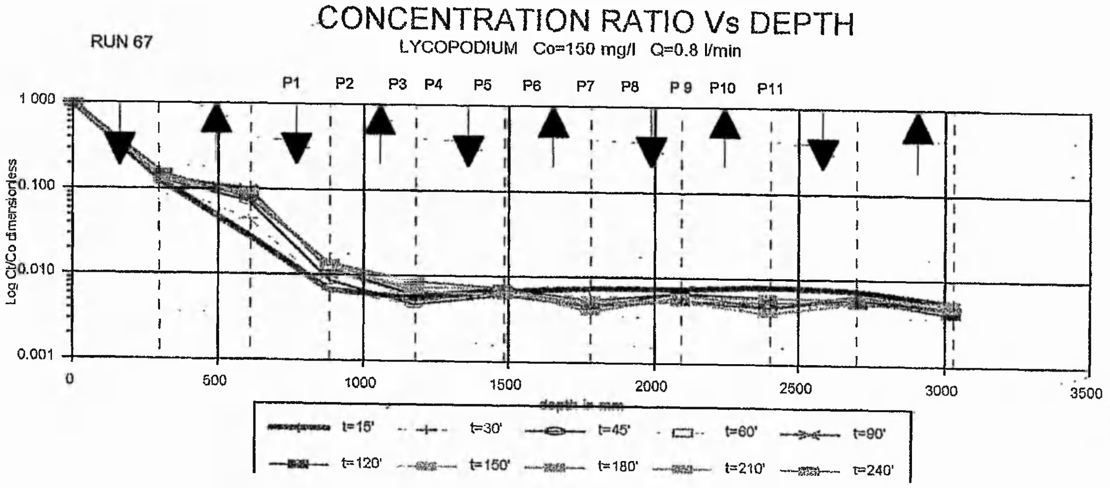
LYCOPODIUM Co=150 mg/l Q=1.0 l/min RUN 66



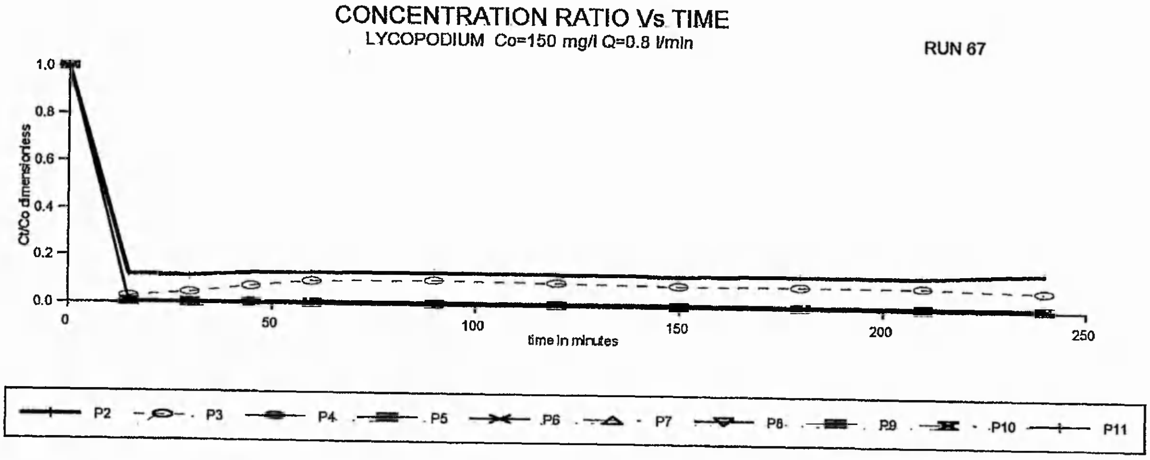
HEADLOSS Vs TIME

LYCOPODIUM Co=150 mg/l Q=1.0 l/min RUN 66

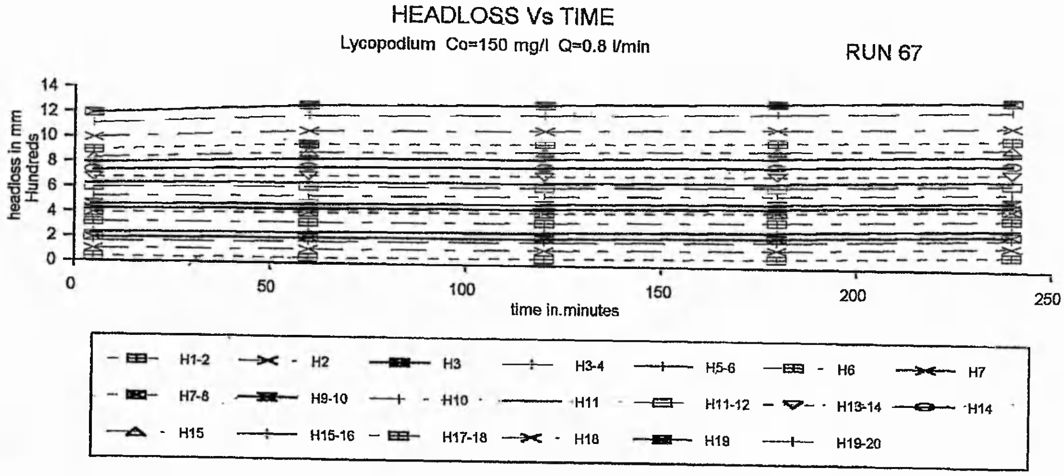



CONCENTRATION RATIO Vs DEPTH

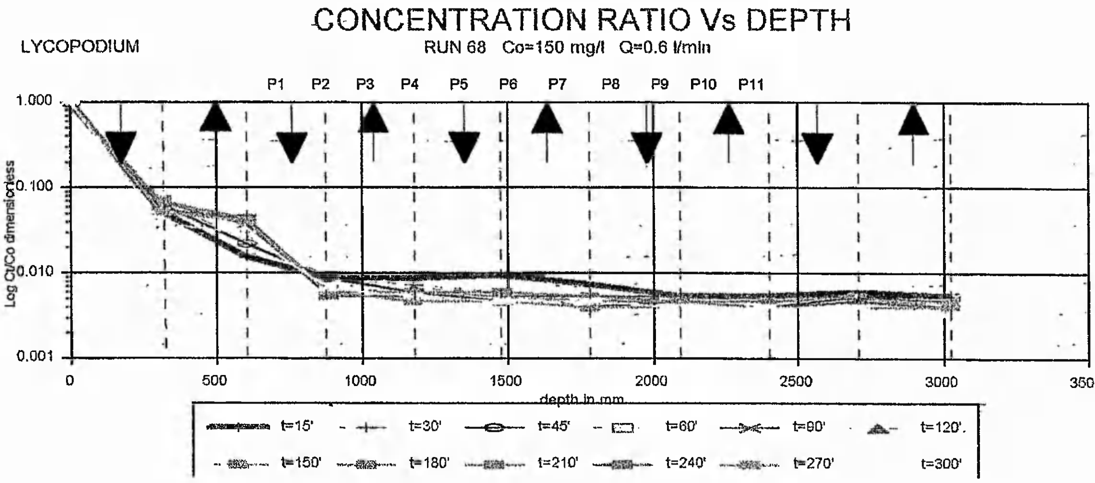
LYCOPODIUM Co=150 mg/l Q=0.8 l/min


CONCENTRATION RATIO Vs TIME

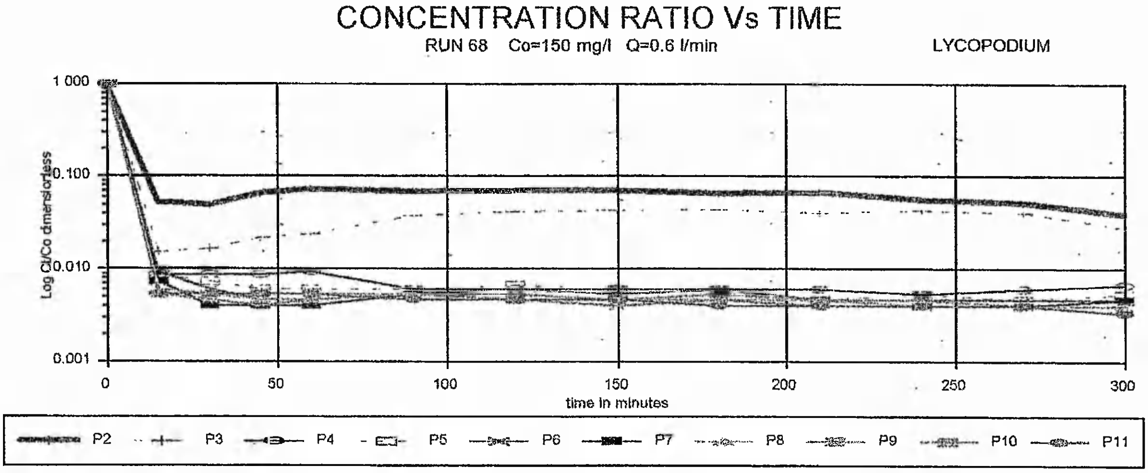
LYCOPODIUM Co=150 mg/l Q=0.8 l/min


HEADLOSS Vs TIME

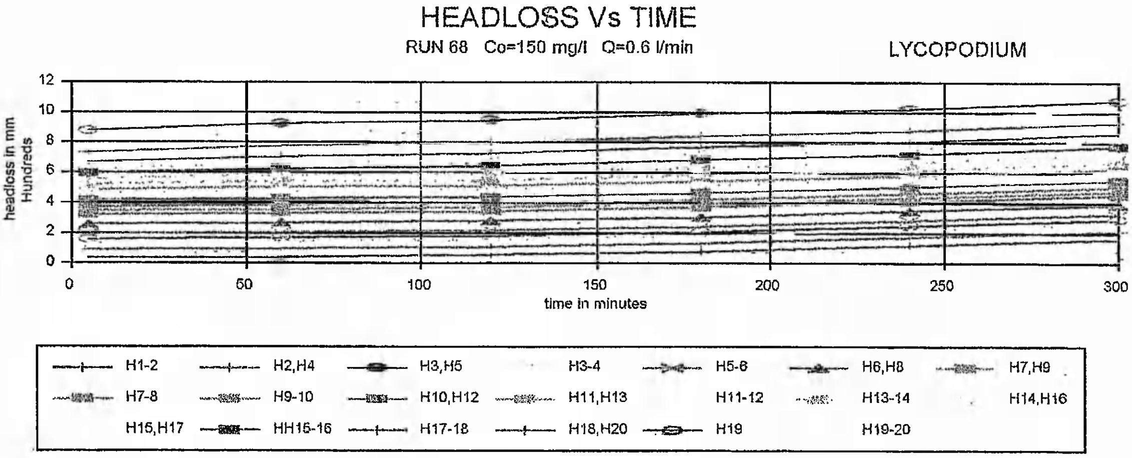
Lycopodium Co=150 mg/l Q=0.8 l/min


CONCENTRATION RATIO Vs DEPTH

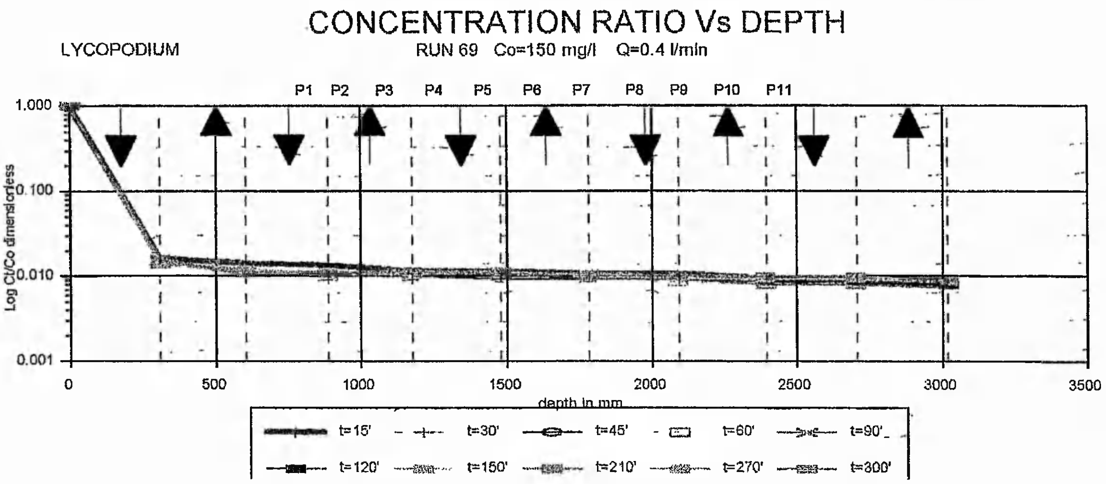
RUN 68 Co=150 mg/l Q=0.6 l/min

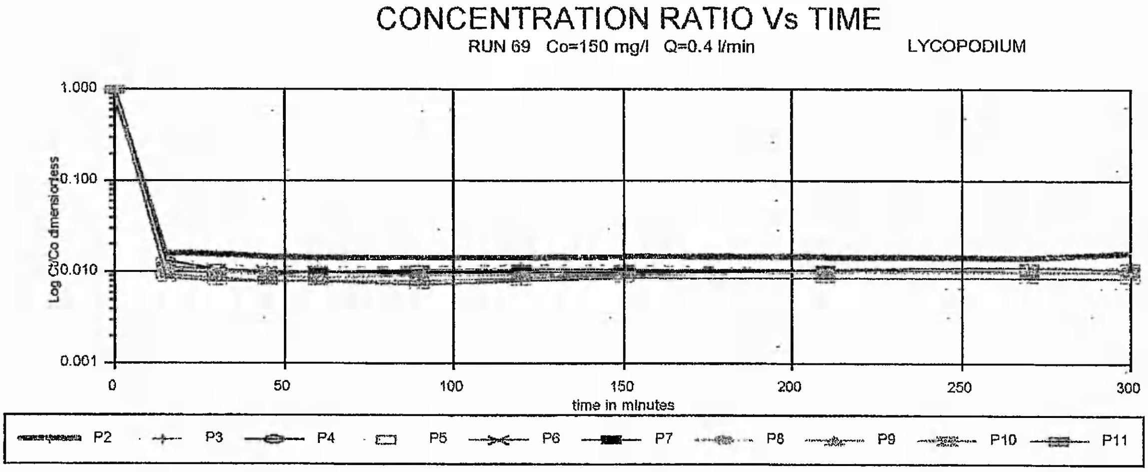

CONCENTRATION RATIO Vs TIME

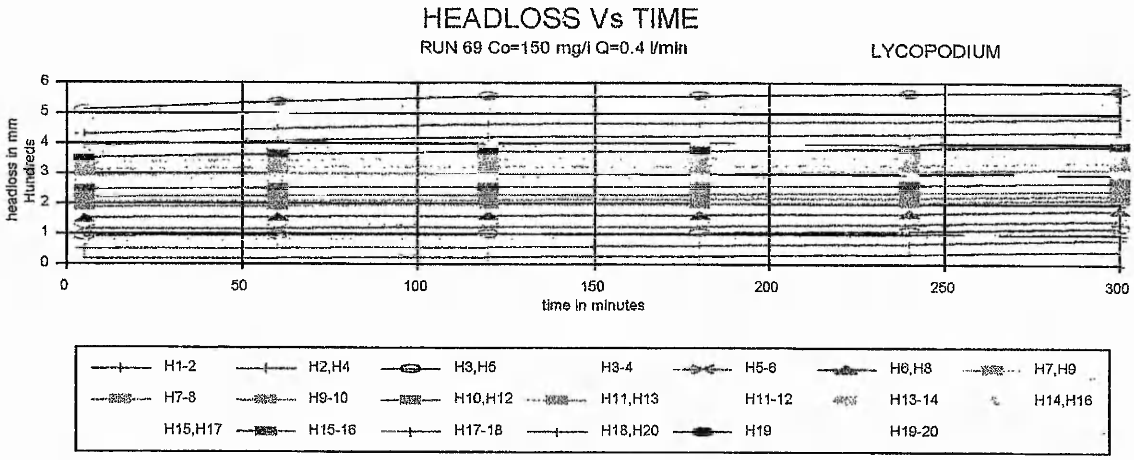
RUN 68 Co=150 mg/l Q=0.6 l/min

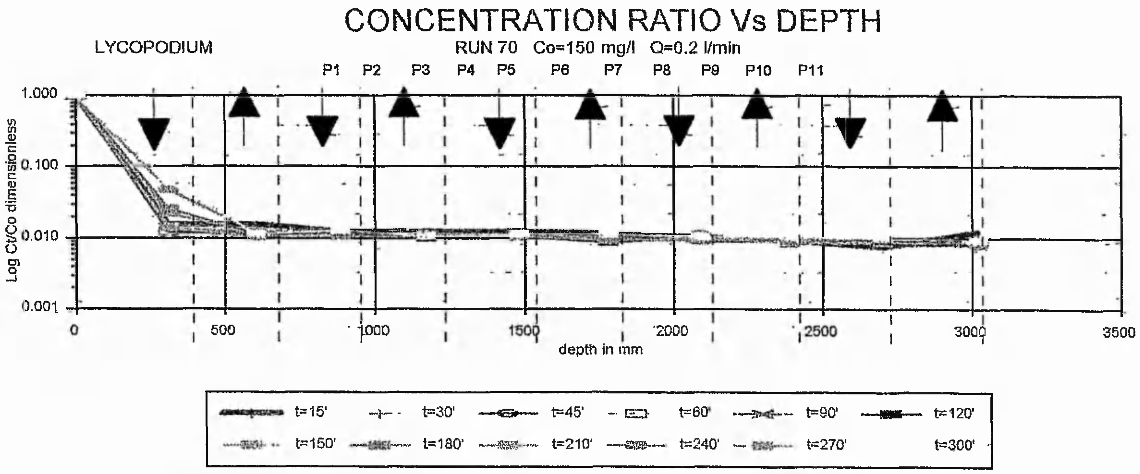

HEADLOSS Vs TIME

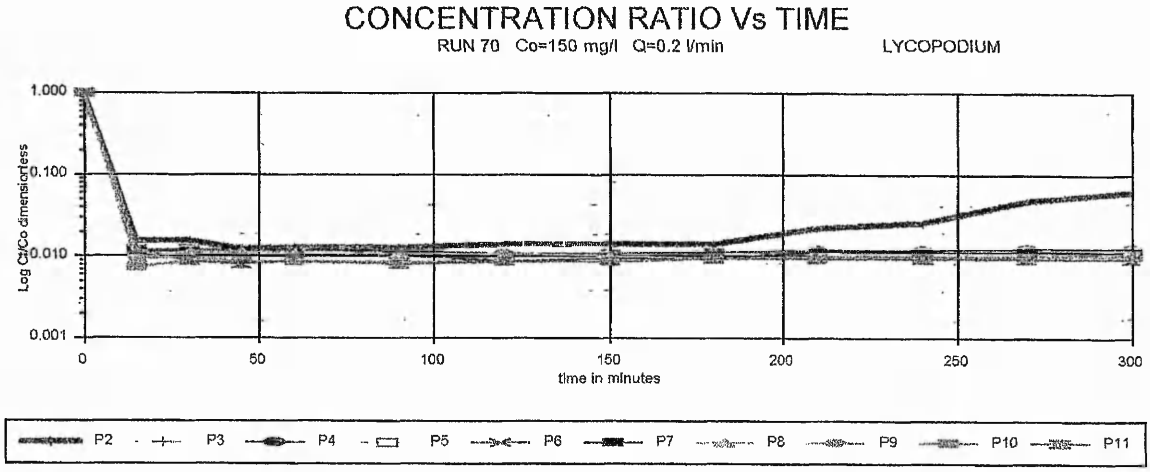
RUN 68 Co=150 mg/l Q=0.6 l/min

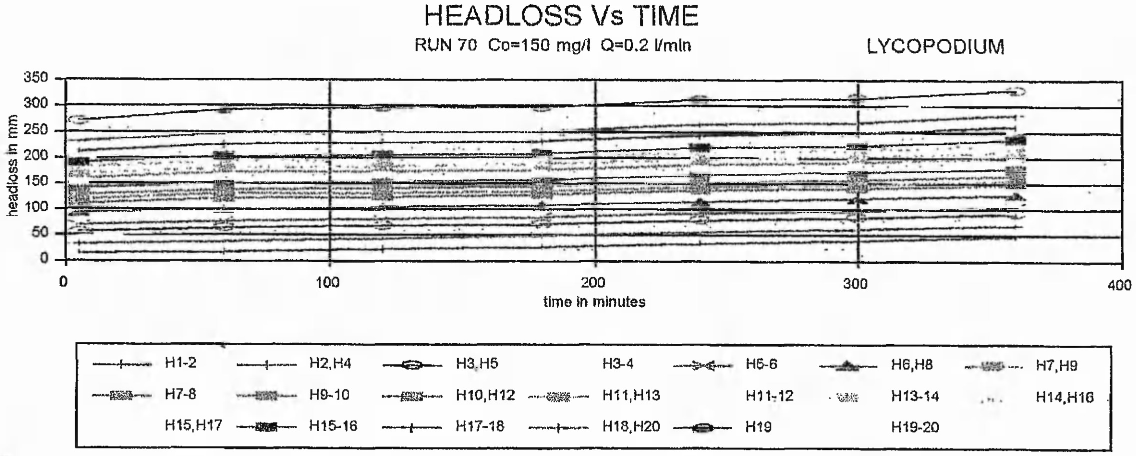

CONCENTRATION RATIO Vs DEPTH
LYCOPODIUM
RUN 69 Co=150 mg/l Q=0.4 l/min
P1 P2 P3 P4 P5 P6 P7 P8 P9 P10 P11
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
500
1000
1500
2000
2500
3000
3500
depth in mm
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
t=150'
t=210'
t=270'
t=300'


CONCENTRATION RATIO Vs TIME
RUN 69 Co=150 mg/l Q=0.4 l/min
LYCOPODIUM
1.000
0.100
0.010
0.001
Log Ct/Co dimensionless
0
50
100
150
200
250
300
time in minutes
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11


HEADLOSS Vs TIME
RUN 69 Co=150 mg/l Q=0.4 l/min
LYCOPODIUM
6
5
4
3
2
1
0
headloss in mm
Hundreds
0
50
100
150
200
250
300
time in minutes
H1-2
H2,H4
H3,H5
H3-4
H5-6
H6,H8
H7,H9
H7-8
H9-10
H10,H12
H11,H13
H11-12
H13-14
H14,H16
H15,H17
H15-16
H17-18
H18,H20
H19
H19-20

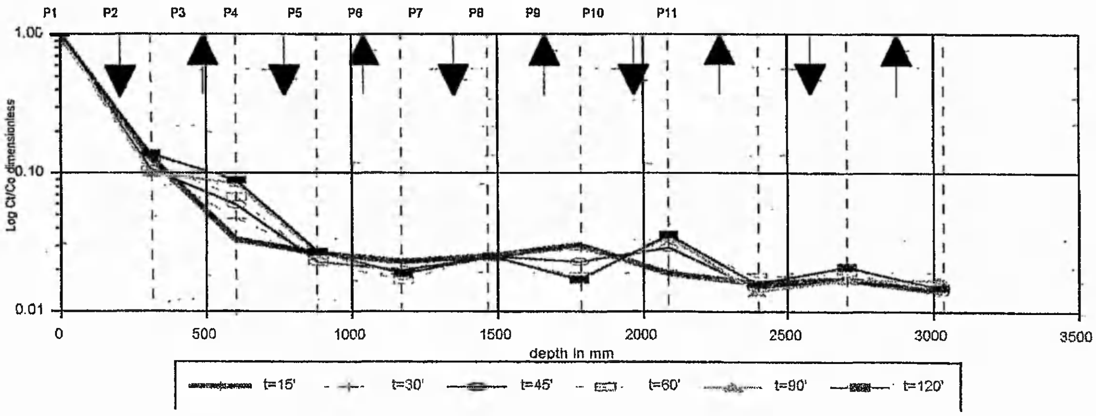

CONCENTRATION RATIO Vs DEPTH
LYCOPODIUM
RUN 70 Co=150 mg/l Q=0.2 l/min
P1 P2 P3 P4 P5 P6 P7 P8 P9 P10 P11
Log Ct/Co dimensionless
1.000
0.100
0.010
0.001
0
500
1000
1500
2000
2500
3000
3500
depth in mm
t=15'
t=30'
t=45'
t=60'
t=90'
t=120'
t=150'
t=180'
t=210'
t=240'
t=270'
t=300'


CONCENTRATION RATIO Vs TIME
RUN 70 Co=150 mg/l Q=0.2 l/min
LYCOPODIUM
Log Ct/Co dimensionless
1.000
0.100
0.010
0.001
0
50
100
150
200
250
300
time in minutes
P2
P3
P4
P5
P6
P7
P8
P9
P10
P11


HEADLOSS Vs TIME
RUN 70 Co=150 mg/l Q=0.2 l/min
LYCOPODIUM
headloss in mm
350
300
250
200
150
100
50
0
0
100
200
300
400
time in minutes
H1-2
H2,H4
H3,H5
H3-4
H5-6
H6,H8
H7,H9
H7-8
H9-10
H10,H12
H11,H13
H11-12
H13-14
H14,H16
H15,H17
H15-16
H17-18
H18,H20
H19
H19-20

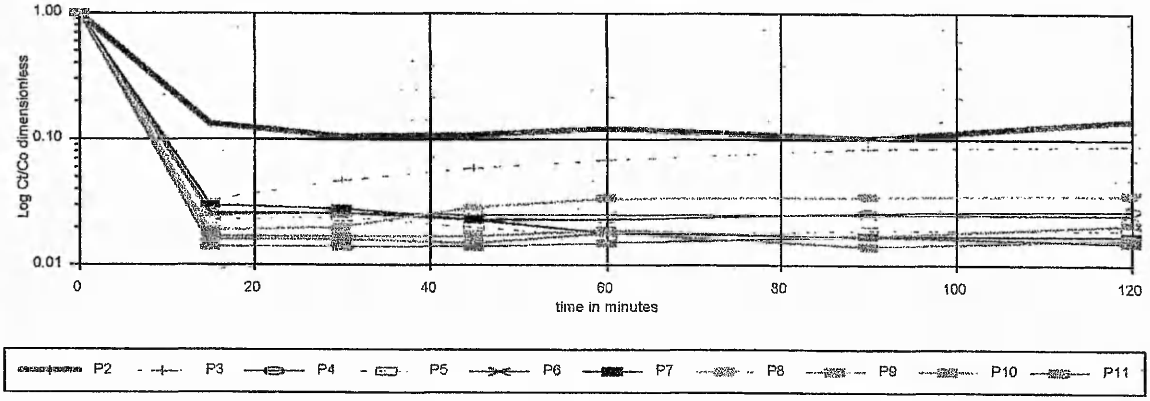
CONCENTRATION RATIO Vs DEPTH

RUN 71 LYCOPODIUM Co=100 mg/l Q=1.0 l/min



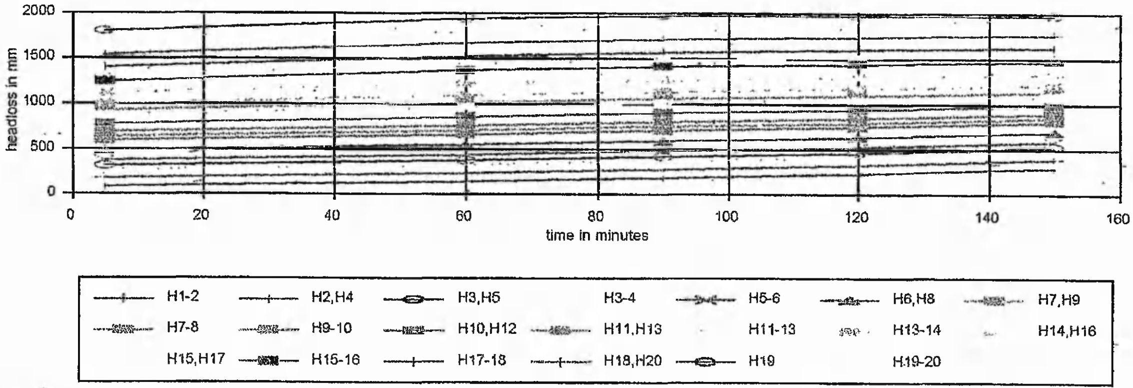
CONCENTRATION RATIO Vs TIME

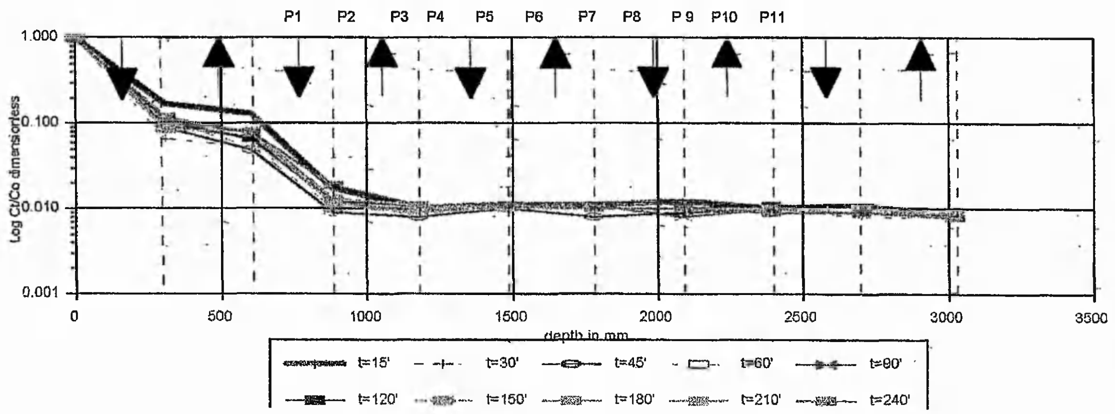
LYCOPODIUM Co=100 mg/l Q=1.0 l/min RUN 71



HEADLOSS Vs TIME

LYCOPODIUM Co=100 mg/l Q=1.0 l/min RUN 71

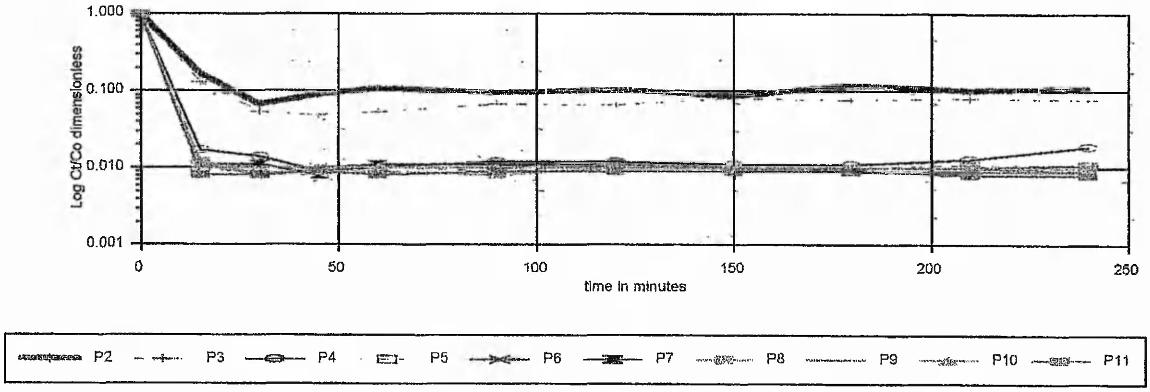




CONCENTRATION RATIO Vs DEPTH

RUN 72

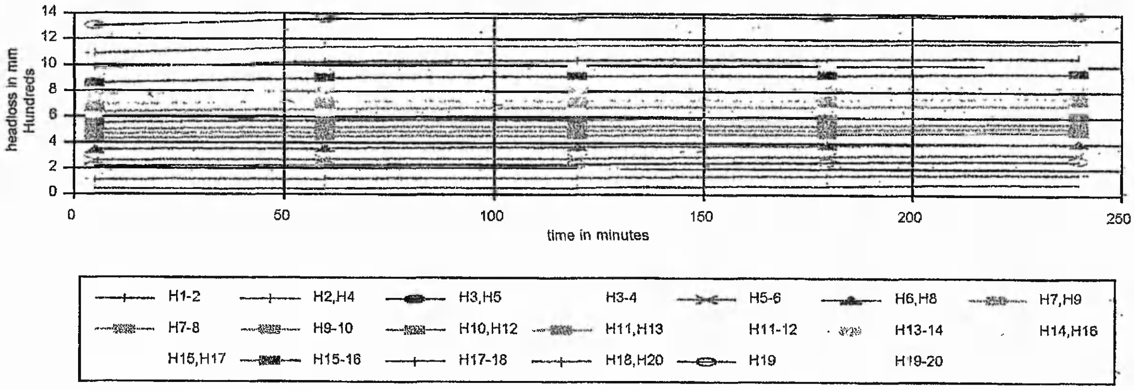
LYCOPODIUM Co=100 mg/l Q=0.8 l/min



CONCENTRATION RATIO Vs TIME

LYCOPODIUM Co=100 mg/l Q=0.8 l/min

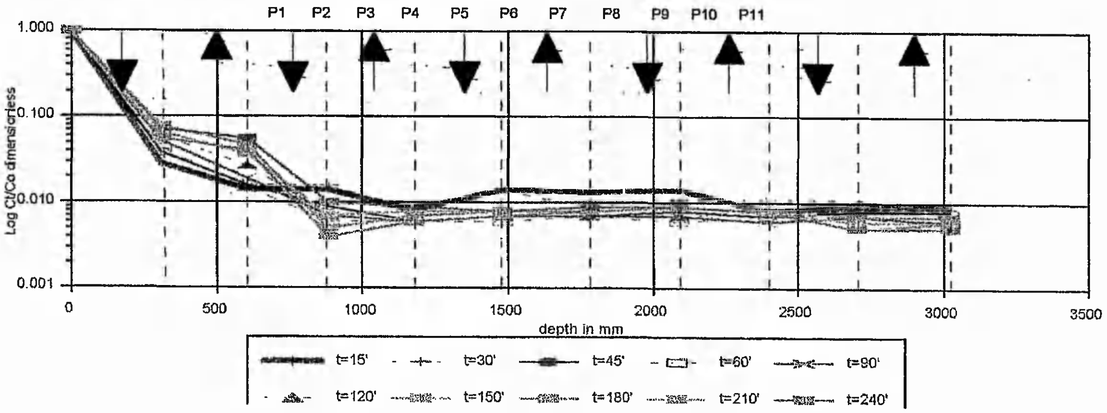
RUN 72



HEADLOSS Vs TIME

LYCOPODIUM Co=100 mg/l Q=0.8 l/min

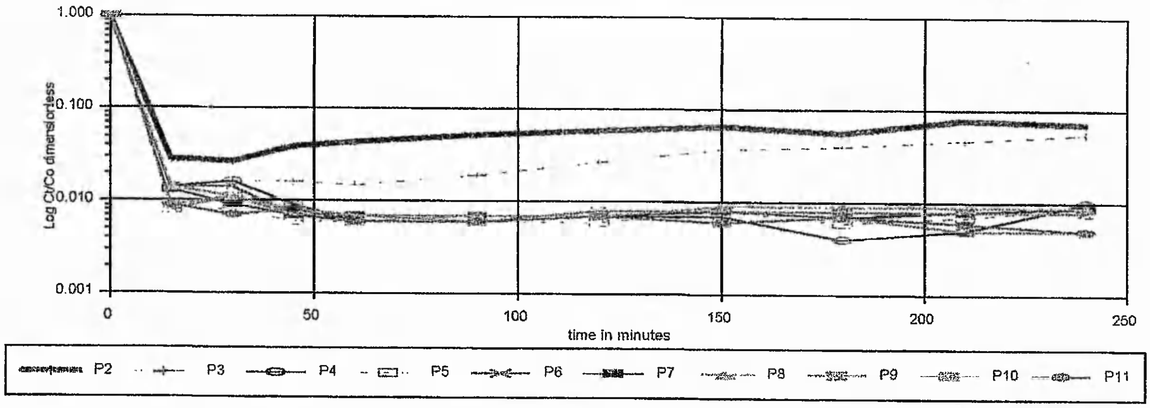
RUN 72



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

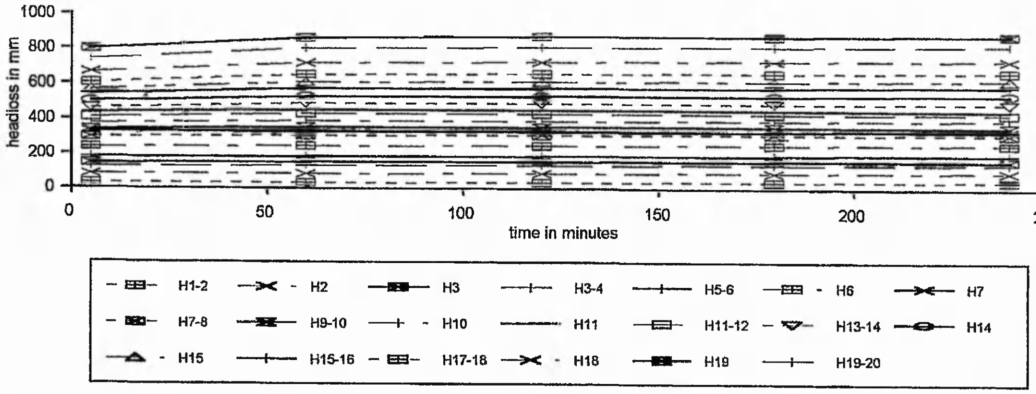
Co=100 mg/l Q=0.6 l/min RUN 73



CONCENTRATION RATIO Vs TIME

LYCOPODIUM Co=100 mg/l Q=0.6 l/min

RUN 73



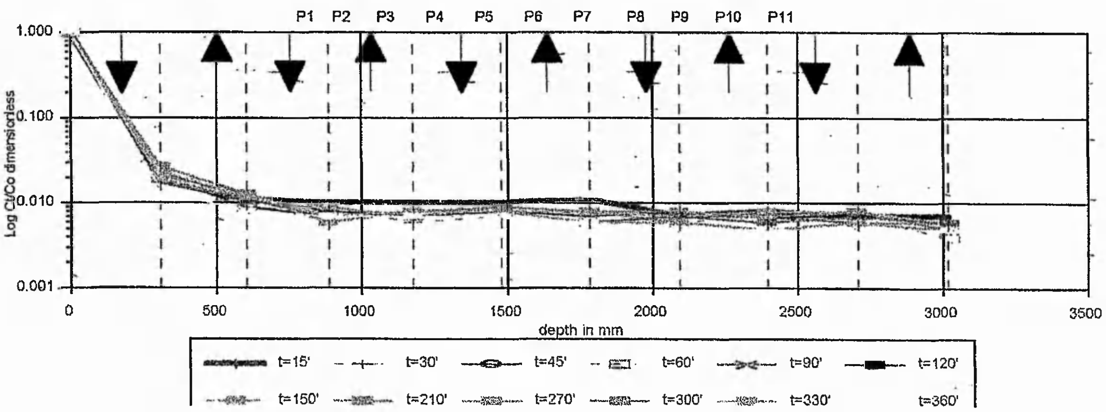
HEADLOSS Vs TIME

LYCOPODIUM Co=100 mg/l Q=0.6 l/min

RUN 73

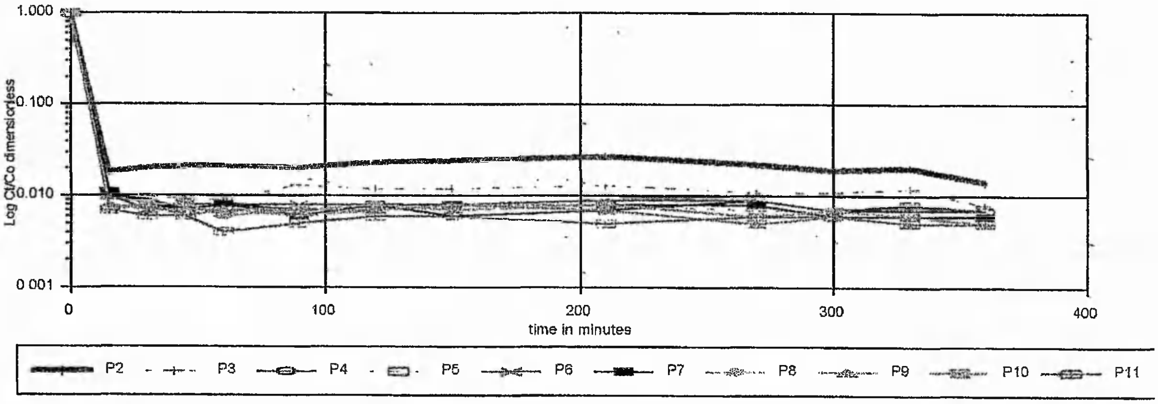
CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM Co=100 mg/l Q=0.4 l/min RUN 74



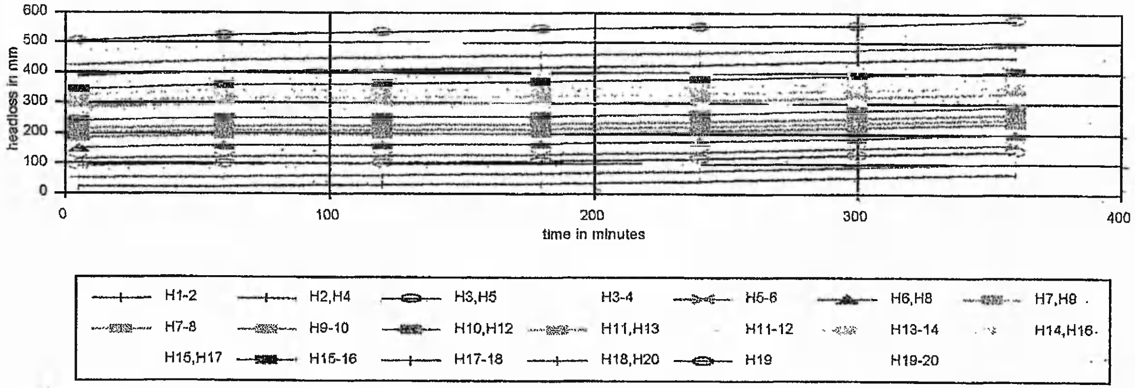
CONCENTRATION RATIO Vs TIME

RUN 74 Co=100 mg/l Q=0.4 l/min LYCOPODIUM



HEADLOSS Vs TIME

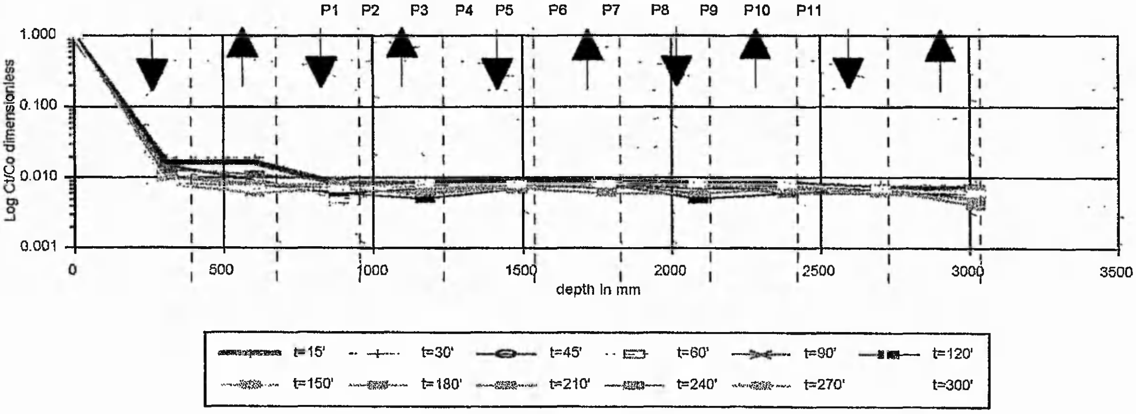
RUN 74 Co=100 mg/l Q=0.4 l/min LYCOPODIUM



CONCENTRATION RATIO Vs DEPTH

LYCOPODIUM

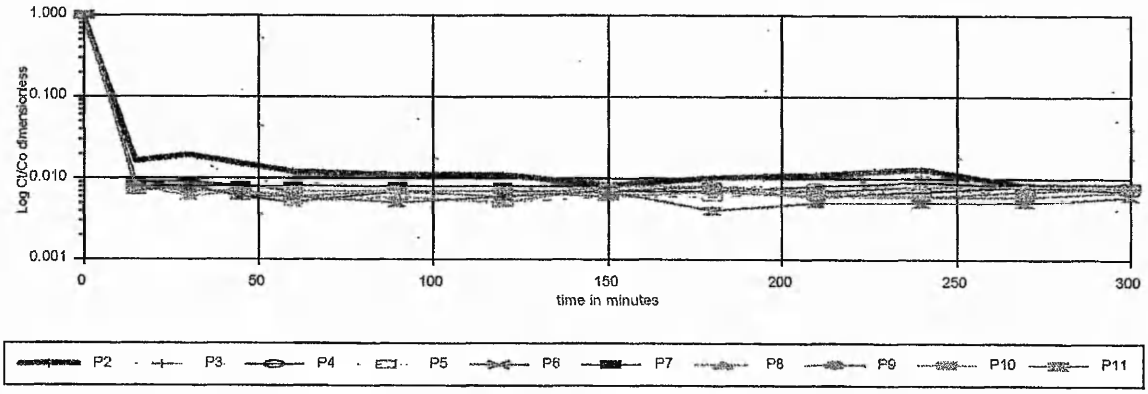
RUN 75 Co=100 mg/l Q=0.2 l/min



CONCENTRATION RATIO Vs TIME

RUN 75 Co=100 mg/l Q=0.2 l/min

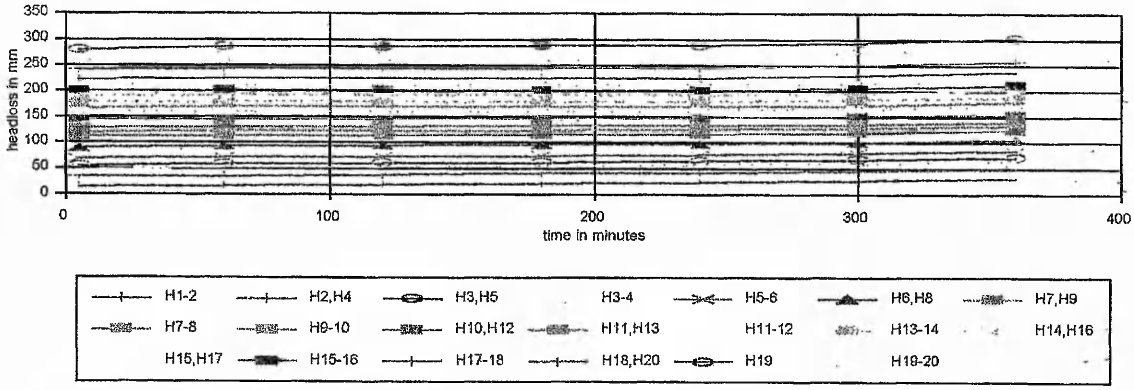
LYCOPODIUM



HEADLOSS Vs TIME

RUN 75 Co=100 mg/l Q=0.2 l/min

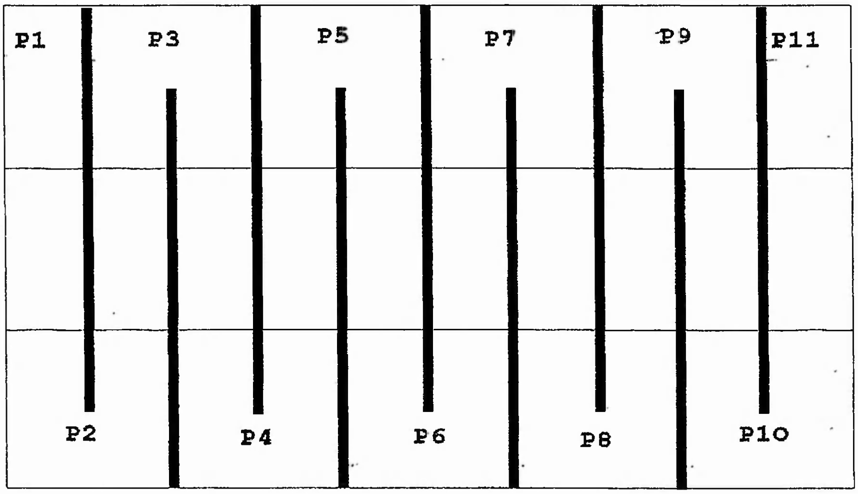
LYCOPODIUM



RUN NUMBER: 17 **DATE OF EXPERIMENT:** 12 JANUARY 1996 **TYPE OF TURBIDITY:** PVC
INITIAL CONCENTRATION: 150 mg / l **RATE OF FLOW:** 0.8 l / min **TEMPERATURE:** 11 C

GRAIN SIZE (μm) D_G	1086		922		777		653		548	
DEPTH OF BED (mm)	310	295	281	290	305	300	309	304	309	320
POROSITY (e)	0.38	0.35	0.32	0.34	0.37	0.36	0.38	0.37	0.44	0.46

INLET **OUTLET**

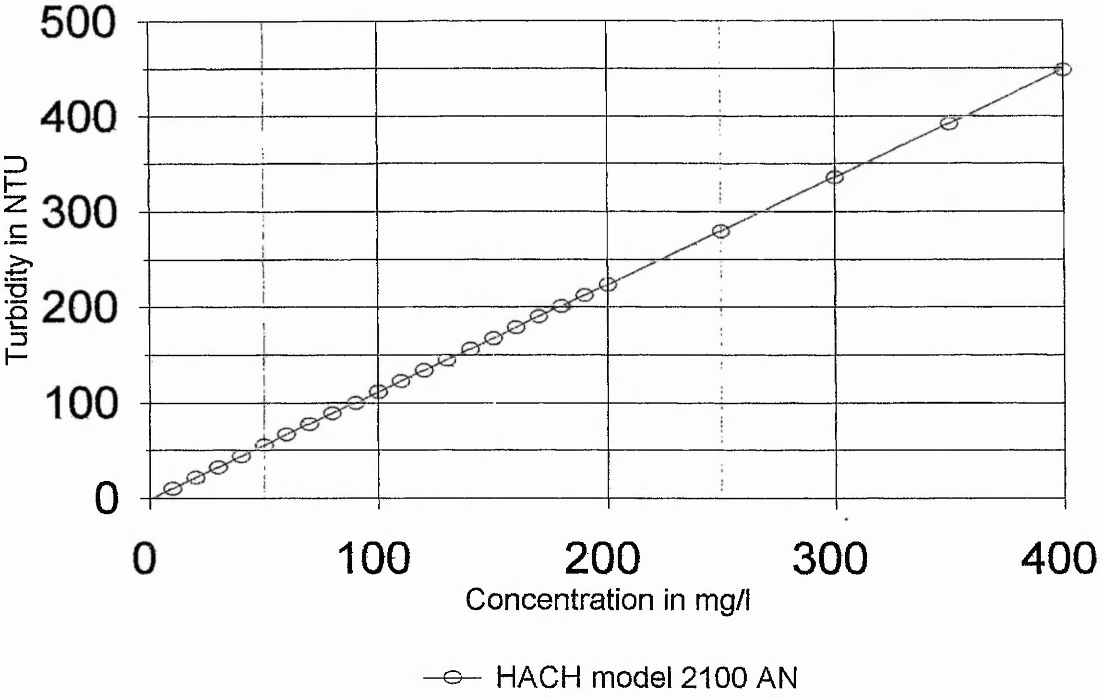


TIME (min)	SAMPLE NUMBER	P1	P2	P3	P4	P5	P6	P7	P8	P9	P10	P11
15	TURBIDITY (mg/ l)	150.0	28.9	20.8	9.8	8.7	5.2	3.9	1.9	1.9	0.9	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.19	0.14	0.07	0.06	0.03	0.03	0.01	0.01	0.01	0.01
30	TURBIDITY (mg/ l)	150.0	24.7	17.7	9.4	8.2	3.9	3.5	1.9	1.8	0.8	1.1
	TURBIDITY RATIO (C / C_o)	1.00	0.16	0.12	0.06	0.05	0.03	0.02	0.01	0.01	0.01	0.01
45	TURBIDITY (mg/ l)	150.0	21.6	16.2	8.9	7.5	4.5	3.7	2.1	1.7	1.0	1.0
	TURBIDITY RATIO (C / C_o)	1.00	0.14	0.11	0.06	0.05	0.03	0.02	0.01	0.01	0.01	0.01
60	TURBIDITY (mg/ l)	150.0	19.0	14.1	8.6	7.4	4.9	3.9	2.1	2.0	1.1	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.13	0.09	0.06	0.05	0.03	0.03	0.01	0.01	0.01	0.01

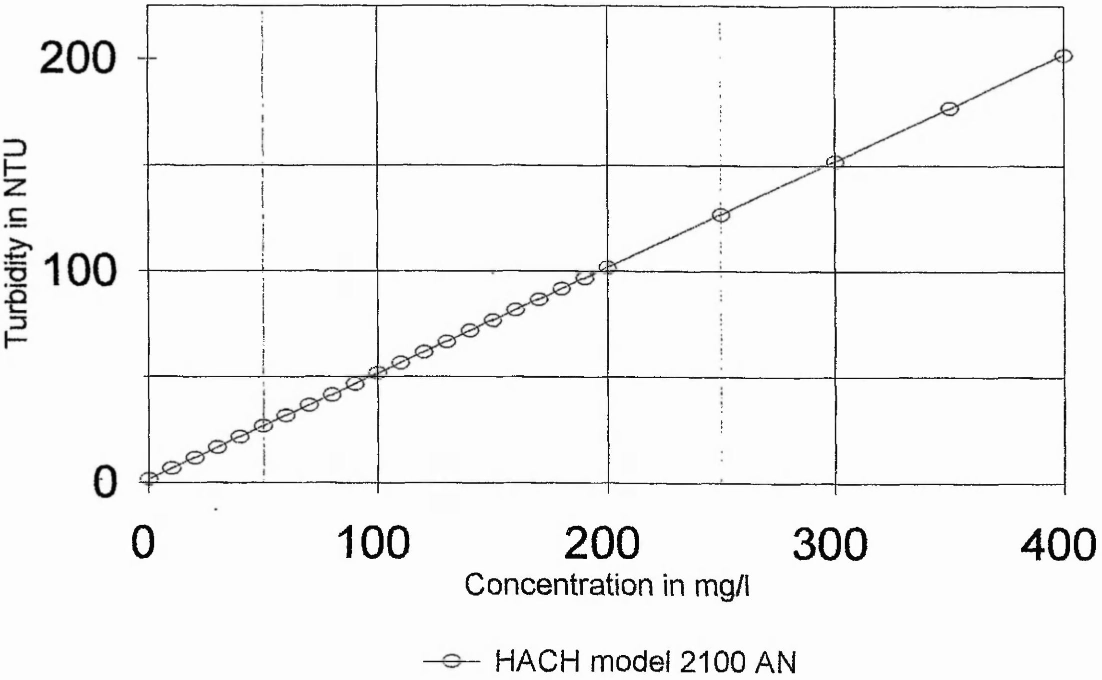
90	**TURBIDITY (mg/ l)**	150.0	14.8	12.3	7.2	6.9	4.1	3.8	2.5	1.6	1.0	0.8
	TURBIDITY RATIO (C / C_o)	1.00	0.10	0.08	0.05	0.05	0.03	0.03	0.02	0.01	0.01	0.01
120	**TURBIDITY (mg/ l)**	150.0	12.0	9.3	6.2	5.7	3.5	3.2	2.0	1.5	1.1	0.8
	TURBIDITY RATIO (C / C_o)	1.00	0.08	0.06	0.04	0.04	0.02	0.02	0.01	0.01	0.01	0.01
150	**TURBIDITY (mg/ l)**	150.0	9.5	7.8	5.3	5.2	3.5	3.1	2.1	1.4	1.0	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.06	0.05	0.04	0.03	0.02	0.02	0.01	0.01	0.01	0.01
180	**TURBIDITY (mg/ l)**	150.0	8.2	8.3	4.8	4.6	3.0	2.8	2.1	1.5	1.0	0.8
	TURBIDITY RATIO (C / C_o)	1.00	0.05	0.06	0.03	0.03	0.02	0.02	0.01	0.01	0.01	0.01
210	**TURBIDITY (mg/ l)**	150.0	7.2	7.2	4.2	3.9	2.6	2.5	1.8	1.5	1.0	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.05	0.05	0.03	0.03	0.02	0.02	0.01	0.01	0.01	0.01
240	**TURBIDITY (mg/ l)**	150.0	6.3	6.0	3.9	3.9	2.5	2.4	1.5	1.3	0.9	0.8
	TURBIDITY RATIO (C / C_o)	1.00	0.04	0.04	0.03	0.03	0.02	0.02	0.01	0.01	0.01	0.01
270	**TURBIDITY (mg/ l)**	150.0	5.7	5.6	3.9	3.7	2.3	2.4	1.5	1.2	0.8	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.04	0.04	0.03	0.02	0.02	0.02	0.01	0.01	0.01	0.01
300	**TURBIDITY (mg/ l)**	150.0	5.5	5.4	3.1	3.0	1.9	2.3	1.4	1.2	0.9	0.8
	TURBIDITY RATIO (C / C_o)	1.00	0.04	0.04	0.02	0.02	0.01	0.02	0.01	0.01	0.01	0.01
330	**TURBIDITY (mg/ l)**	150.0	4.5	4.1	2.9	2.5	1.8	2.2	1.3	1.2	0.9	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.03	0.03	0.02	0.02	0.01	0.01	0.01	0.01	0.01	0.01
360	**TURBIDITY (mg/ l)**	150.0	4.1	4.2	2.8	2.4	1.6	1.9	1.2	1.1	1.0	0.9
	TURBIDITY RATIO (C / C_o)	1.00	0.03	0.03	0.02	0.02	0.01	0.01	0.01	0.01	0.01	0.01
390	**TURBIDITY (mg/ l)**											
	TURBIDITY RATIO (C / C_o)											

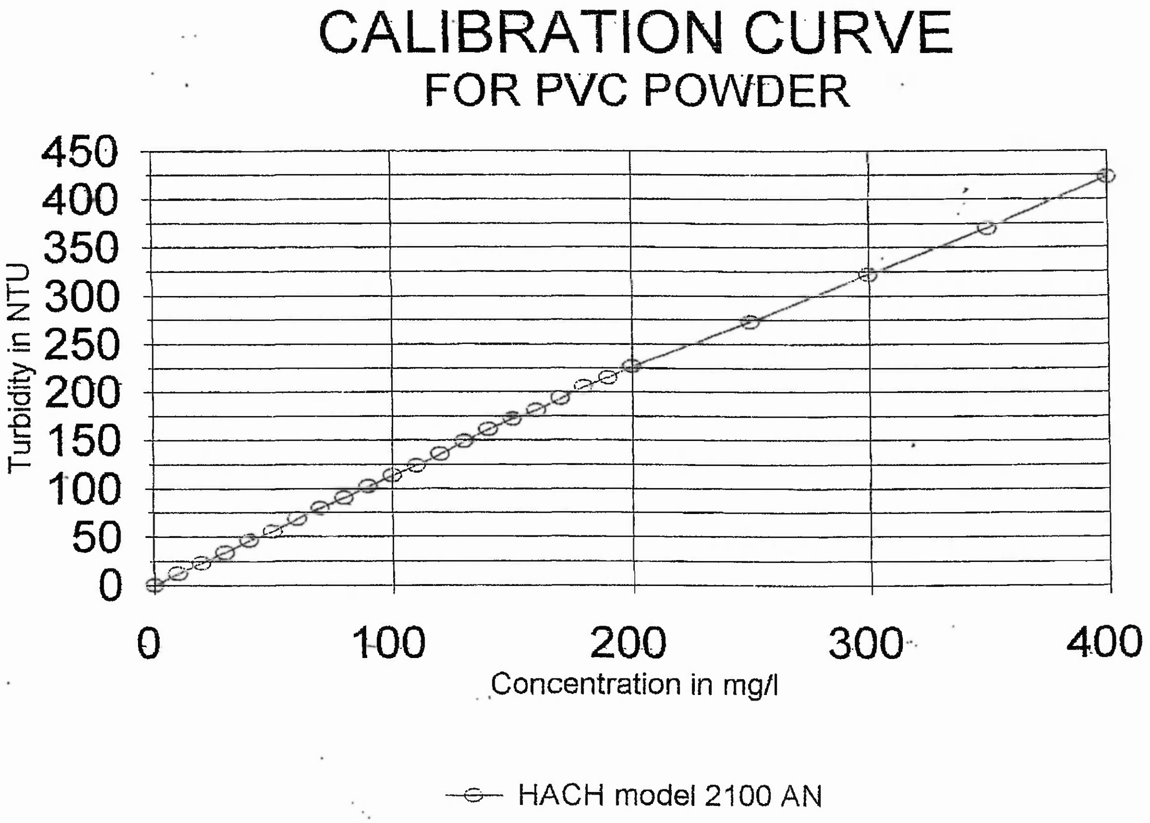
Appendix 3 Calibration Curves

CALIBRATION CURVE
FOR KAOLIN POWDER



CALIBRATION CURVE
FOR LYCOPODIUM POWDER




CALIBRATION CURVE
FOR PVC POWDER
Turbidity in NTU
450
400
350
300
250
200
150
100
50
0
0
100
200
300
400
Concentration in mg/l
HACH model 2100 AN