STUDIES ON PARTICLE CHARACTERISTICS AND PRESSURE DROP IN GRANULAR MEDIA FILTRATION

A DISSERTATION

Submitted in partial fulfillment of the requirements for the award of the degree

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in

CHEMICAL ENGINEERING

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By

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CANDIDATE'S DECLARATION

I hereby declare that the work, which is being presented in the project report entitled "**Studies on Particle Characteristics and Pressure Drop in Granular Media Filtration**" in the partial fulfillment of the requirements of the award of the degree of **Master of Technology in Chemical Engineering** with specialization in **Industrial Pollution Abatement**, submitted in the Department of Chemical Engineering, Indian Institute of Technology Roorkee, Roorkee, is an authentic record of my own work carried out during the period from July 2006 to May 2007 under supervision of **Dr. B. Prasad**, Associate Professor, Department of Chemical Engineering, Indian Institute of Technology Roorkee, Roorkee.

I have not submitted the matter, embodied in this project report for the award of any other degree.

Date: 18 June, 2007
Place: Roorkee

18/6/07
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CERTIFICATE

This is to certify that the above statement made by the candidate is correct to the best of my knowledge and belief.

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ABSTRACT

Rapid gravity granular media filters are widely used in the water and wastewater treatment industries. Regular backwashing to clean the filters is a vital part of their efficient operation.

In the present study, emphasis was given on the removal of fine suspension particle by filtration through low cost coarse media. A set up was made in the laboratory and filtration process was proceeded. The media taken was the mixed bed of Calcite-Anthracite, Calcite-Garnet & Anthracite-Garnet. Clay was added in a definite amount (20 mg l^{-1}) in tap water and this suspension was used. Samples at regular intervals were collected and analyzed. Also, the effect of pressure drop was taken into consideration.

Results obtained indicated that the removal efficiency was best observed in the case of "Calcite-Garnet" as compared to "Anthracite-Garnet" and "Calcite-Anthracite" mixed beds. At 500 mm bed height, at a flow rate of 30 dm^3 h^{-1}, the removal efficiency of Calcite-Garnet was found to be 99.99%, of Anthracite-Garnet it was 99.85 % and of Calcite-Anthracite it was found 99.5% at time interval of 60 min. Studies on the effect of different parameter on the removal performance of fine suspended particle shows that a smaller particle size, a greater filter depth, shorter cycle time gave an advantage in removal efficiency.

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CHAPTER 1
INTRODUCTION

1.1. General

Waste water treatment is becoming more critical due to diminishing water resources, increasing wastewater disposal costs & stricter discharge regulations that have lowered permissible contaminant levels in waste streams. Physical, chemical and biological levels of contaminant removal, individual wastewater treatment procedures are combined into a variety of systems, classified as primary, secondary & tertiary wastewater treatment.

More rigorous treatment of wastewater includes the removal of specific contaminants as well as removal and control of nutrients. Natural systems are also used for the treatment of wastewater in land based applications. Sludge resulting from wastewater treatment operations is treated by various methods in order to reduce its water & organic content & make it suitable for final disposal and reuse. Wastewater treatment methods are broadly classifiable into physical, chemical & biological processes.

The removal of particulate matter from water by the use of granular media filters plays a critical role in water treatment of sea water, waste water treatment and industrial water and waste water applications. In order to provide improved TSS removal for ion exchange and reverse osmosis, micro filtration and ultra filtration membranes have proposed and are seeing increased application. Membranes provide a barrier to suspended solids and ensure consistently low levels of TSS to downstream processes. The downside is that these processes tend to have higher initial capital costs and operating costs for periodic membrane replacement are appreciable.

Granular media filters are used to remove particulate matter in water treatment processes. They require cleaning at regular intervals and this is usually achieved using water or combined air and water backwash. The medium commonly used in these filters, sand, processes high attrition resistance and therefore there has been little concern about media losses due to attrition during backwashing (Humby and Caroline, 1995). However, the use of granular activated carbon (GAC) and anthracite as filter media has been increasing and these media do not possess the attrition resistance of sand (Humby and Caroline, 1995). Excessive attrition would lead to a gradual loss of media and its size could reduce significantly (Ives, 1990). Therefore the relative abilities of all these media to withstand attrition need to be established so that the economic and operational effects of attrition can be predicted.

Filters have historically been cleaned with a water only backwash at 20-50% expansion (Humby and Caroline, 1995). It has been shown that this regime is fairly ineffective in cleaning due to the limited number of grain abrasions and impacts that occur (Amirtharajah, 1993). This was thought to be due to maximization of fluid shear and grain abrasions and impacts. A study using endoscopes and high speed video has confirmed this (Fitzpatrick, 1993).

Granular media filter provides an economical solid-liquid separation process for achieving a desired water quality level with respect to particulate parameters, and in some cases, with respect to specific contaminants (e.g. Metals, hydrophobic synthetic organic compounds) that is predominantly associated with particulate or colloidal phases. Granular media filters have been shown to be capable of efficiently removing a wide range of particle types from water, ranging from particulates of microbial origin with low density, to high density inorganic particulates. Granular media can capture particulates ranging in size from the 2μm to 100μm.

1.2 Objectives of current work

To study the nature of granular media a lab scale down flow media filter is designed and an experimental setup is made in lab. Our study emphasizes mainly on mixed media beds of 3 combinations i.e. Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite to ensure the fine suspended particle removal ability at different bed height as well as with time. Tap water with suspended solid (20mg/dm^3, size-2μm to 20μm) is taken as a flowing fluid at constant flow rate for study purpose. In the present study, comparisons are made among the three mixed bed media on the basis of the fine suspended solid particle removal. To make comparisons all the analysis was done with the Particle Size Analyzer. The comparisons are made on the basis of removal efficiency at different bed height of the media. Moreover the removal efficiency of each mixed media is determined and compared at different time period. Our main objectives of the experimental work are

- To study the fine suspended particle (below 20μm) removal efficiency of Calcite-Anthracite, Anthracite- Garnet and Garnet- Calcite mixed bed at different bed heights (200,300,400,500 mm) at different time intervals (30, 60, 90 min) at constant flow rate.
- To compare their removal efficiency at different bed height at different time period (at 30, 60 and 90 min) at constant flow rate.
- To study the effect of pressure drop along the bed height and time period.

CHAPTER 2
FILTRATION-TYPES AND METHODS

The separation of solids from a suspension in a liquid by means of a porous medium or screen which retains the solids and allows the liquid to pass is termed filtration. In general, the pores of the medium are larger than the particles which are to be removed, and the filter works efficiently only after an initial deposit has been trapped in the medium. Filtration is a well-studied process for drinking water treatment. Naturally, as groundwater migrates in the subsurface, contaminants are removed from the water due to ionic attraction as well as sieving based on size (Lundquist, 2006). Concurrently, contaminants such as iron and manganese may be dissolved into the groundwater and often remain in the dissolved form until pumped to the surface. Similarly, microorganisms are imparted to and extracted from the groundwater during subsurface movement. Surface water (e.g. Ponds, lakes, rivers) like groundwater, has ever changing quality with respect to micro organisms, particulates, chemistry, etc., but is more exposed to human activity, often degrading water quality. To reduce water contaminants and create potable water safe for human consumption, water treatment has included filtration to mimic and better the natural removal of water contaminant.

Filtration is used to

1. Dewater waste effluents, slurries and sludges generated from industrial treatment processes, and
2. Remove undissolved heavy metals present in suspended solids.

It does not reduce the toxicity of the waste. Sludge dewatering eliminates free liquids for landfill disposal, and reduces waste volume for more stable and economical transport and incineration. The major sludge dewatering processes include rotary drum vacuum filters, belt filter presses, and plate and frame filter presses. These methods use either negative or positive pressure to move water through filter media, leaving solids behind. Other methods of sludge dewatering include gravity thickening through sedimentation, flotation, and centrifugation.

Removal of heavy metals in suspended solids is usually done by granular media filtration. Granular media filtration uses gravity to pass fluid through a bed of granular material, removing solids from the fluid. Straining, physical adsorption or coagulation-flocculation removes the suspended solids. A wash water stream is used to unclog granular filter media and clean the operating parts of the vacuum filter or filter press.

2.1 Current use of Filtration for water treatment

The original slow sand filtration developed centuries ago has now been replaced with rapid sand filtration using multi-media beds, adsorption, utilizing electrochemical forces to attract contaminant to the media surface, natural and synthetic membranes engineered with distinct pore sizes, and ion exchange, where one ion is removed from the water and replaced with a less offensive ion (Lundquist, 2006).

2.2 Separation mechanisms

The mechanisms of separation during filtration vary depending on material and design. Overall, several mechanisms may be simultaneously rejecting contaminants. For example, during filtration primarily incorporating size exclusion, adsorption and depth filtration mechanisms are likely aiding in particle retention (Lundquist, 2006).

2.2.1 Straining

Straining entails the removal of particles by size exclusion when particles are larger than the void spaces in the filter (Lundquist, 2006). Straining is a removal mechanism for virtually all filtration technologies with the importance of this mechanism related to raw water quality and size of particulate matter in reference to pore size. For spherical granular media, close-packed arrangement will remove particles when the ratio of particle diameter to grain diameter is greater than 0.15 (Crittenden, et. al., 2005). For typical slow sand filters, this equates to the removal of particles down to about 15 μm, increasing to 30-80 μm for rapid sand filtration (Lundquist, 2006). It should be noted that other mechanisms aid in the removal of smaller particles for these filtration techniques. Specifically, for slow sand filtration a thin slimy layer of particulate sludge forms, termed smutzdecke, effective in trapping particulates and microorganism at the surface. When particulates form a layer during granular media filtration it may also be termed a cake. Cake filtration is often used to describe straining out particles, often smaller than the media pore size, by this top layer, or build-up, when evaluating granular carbon filtration.

2.2.2 Rejection by osmotic membranes

Two solutions in contact with one another with varying solute concentrations naturally try to equilibrate. In water treatment we can use this driving force to equilibrate, by placing a semi-permeable membrane between the two solutions. By engineering the membrane to allow passage of the water molecules through the membrane, yet reject the solutes, the two solutions will naturally equilibrate as the water dilutes the more concentrated side (Lundquist, 2006). Flux through the membrane will vary based on solute gradient, temperature, and membrane properties. Common practice in water treatment is to reverse the natural osmotic tendency by pressurizing the influent side, forcing water molecules through the membrane, and rejecting the solutes, termed reverse osmosis (RO). Despite use in water treatment for many years, the exact mechanism of water transport and solute rejection is still debated. The underlying question is whether these membranes are non-porous and diffusion driven or whether they contain very small pores for preferential (size exclusion) convective transport of the solvent.

2.2.3 Adsorption

Adsorption is a mass transfer operation in which contaminants present in liquid phase are accumulated on a solid phase, thereby being removed from the liquid. The constituent being adsorbed is referred to as the adsorbate and the solid onto which constituent adsorbs is the adsorbent. The degree of adsorption is affected by attraction of the three following interfaces: adsorbate/adsorbent, adsorbate/water, water/adsorbent (Lundquist, 2006). The strength of the adsorbate/adsorbent interface as compared to the others will determine adsorption efficacy. Dissolved species are concentrated onto the surface by physical attraction or chemical reaction. Physical adsorption is by nonspecific binding mechanisms such as Van der Waals forces. This binding is reversible, where absorbates may desorb in response to a decrease in solution concentration. Chemisorption entails specific attraction where chemical binding transfers electrons between the adsorbent and adsorbate.

2.2.4 Ion exchange

Ion exchange for drinking water is a process in which ions within the water stream are adsorbed to the surface of resins and exchanged for a less offensive ion that is then imparted into the finished water. A generic representation of softening using a sodium resin is shown below, with R representing the exchange resin (Lundquist, 2006).

$$R\text{-}(Na^{+})_4 + Ca^{+2} \rightarrow R\text{-}(Ca^{+2}) + (Na^{+})_4$$

Similar to adsorption, ion exchange is powered by electrostatic/electrochemical attraction in which ions of opposite charge attract, however, with ion exchange: the presaturate ions cannot be present in the bulk fluid. Natural tendency equilibrate will favor ions both in enough time (Owens, 1985). Resin beads are usually 0.04 to 1.0 mm in diameter and reversible, and once all exchange sites are exhausted they can be restored through regeneration, although eventually irreversible fouling will occur.

2.2.5 Membrane Fouling

A membrane is a thin layer of semi-permeable material that is capable of separating materials when a driving force is applied across the surface. This separation into two membranes is based on the physical and chemical properties of the materials being separated. Membranes are not considered to be passive materials but are termed functional materials whose performance characteristics are based on the nature of the elements to be separated and the driving force. Membranes are classified based on the size or molecular weight cutoff (MWCO) of the solutes they are capable of rejecting (Lundquist, 2006). Membranes used in water treatment, in order of decreasing pore size/MWCO, are micro filters, ultra filters, nano filters, and osmotic membranes.

2.3 Applicability

Filtration can be used on:

- secondary biological sludge
- water treatment alum sludge
- metal hydroxide sludge
- oily sludges (i.e. from API separators and dissolved air flotation units)
- brine sludge

Granular media filtration is generally used after gravity separation. It removes additional suspended solids and oils before other treatment processes. It is also a polishing step that lowers the levels of suspended solids and associated contaminants in treated wastes.

2.4 Granular media filters

The removal of particulate matter from water by the use of granular media filters plays a critical role in water treatment for potable use, waste water treatment and industrial water and waste water applications. In each case, a granular media filter provides an economical solids-liquid separation process for achieving a desired water quality level with respect to particulate parameters, and in some cases, with respect to specific contaminants (e.g. metals, hydrophobic synthetic organic compounds) that are predominantly associated with particulate or colloidal phases (Boller and Kavanaugh, 1994). Granular media filters have been shown to be capable of efficiently removing a wide range of particle types from water, ranging in size from particulates of microbial origin with low densities, to high density inorganic particulates such as ferric oxide scale or titanium oxide solids. Particulates ranging in size from the sub-colloidal (less than 0.01 μm) up to 100 μm can be easily captured by granular media, provided that surface chemical conditions are suitable for attachment of the particulates to the media surface or to previously deposited particulates (Boller and Kavanaugh, 1994). The one key limitation in the use of granular media filters in solids-liquid separation, however, is the capacity of the system for retention of solids within the pore space of the media. This constraint limits the solids loading that can be economically treated by granular filters. One of the major advantages of depth media filtration is that the cost of media replacement is extremely low, since the life is long and the media itself is inexpensive. It is also not prone to fouling and is generally considered very robust. Together with a reasonable initial capital cost, this explains the prominent place that this pretreatment technology has held for so many years in the industry.

Although granular media filtration has been the subject of many years of research and development, there are still no accepted analytical models of the process that permit non empirical optimization of process parameters. This is due in part to the relatively low cost of the process compared to alternatives that can achieve equivalent water quality levels, such as ultra filtration using membranes, and in part, due to the extremely complex physical and chemical sub processes controlling particulate parameters (size and shape of the granular media, depth of media, superficial liquid velocity, clean-bed porosity), and physical characteristics of the particulate suspensions (particle size, size distribution, number concentration, shape, and density). In addition, the effectiveness of a filter to capture particulates depends strongly on the surface chemistry of both the media and the particulates, and this chemistry is strongly influenced by water chemistry (pH, ionic strength, concentration of specific ions), the particular surface properties of the media and the particulates (Boller and Kavanaugh, 1994). As a result of the lack of analytical models of the process, most engineers design granular media filters based on extrapolation of previous experience. In many cases, pilot scale studies have been undertaken prior to design to validate design assumptions or to optimize the filter design. In the industrial environment, equipment vendors, who tend to avoid changes in equipment specifications due to cost considerations, normally control filter design. The goal of optimum design and operation is to minimize construction and operating costs. Operating costs are controlled by chemical costs, sludge disposal costs, pumping costs and costs associated with the regeneration of the granular media. In theoretical terms, it has been shown that the operating costs are at a minimum when the time at which a limiting head loss is reached coincides with the time at which the water quality objective in the filtered water is exceeded (Boller and Kavanaugh, 1994).

However, because of reliability requirements, granular media filters are normally designed and operated to insure that the time to reach breakthrough is considerably longer than the time to reach the limiting headloss. Thus, the rate of headloss increase is the dominant design constraint, provided that the particulates are adequately destabilized to achieve efficient capture in the granular media.

2.5 Filtration theory

2.5.1 General

The separation of solids from a suspension in a liquid by means of a porous medium or screen which retains the solids and allows the liquid to pass is termed filtration. In general, the pores of the medium are larger than the particles which are to be removed, and the filter works efficiently only after an initial deposit has been trapped in the medium. Two basic types of filtration processes may be identified, although there are cases where the two types appear to merge.

In the first, frequently referred to as cake filtration, the particles from the suspension, which usually has a high proportion of solids, are deposited on the surface of a porous septum which should ideally offer only a small resistance to flow. As the solids build up on the septum, the initial layers from the effective filter medium, preventing the particles from embedding themselves in the cloth, and ensuring that a particle- frees filtrate is obtained.

In the second type of filtration, depth or deep bed filtration, the particles penetrate into the pores of the filter medium, where impacts between the particles and the surface of the medium, are largely responsible for their removal and retention. This configuration is commonly used for the removal of fine particles from very dilute suspensions, where the recovery of the particles is not of primary importance. Typical examples here include air and water filtration. The filter bed gradually becomes clogged with particles, and its resistance to flow eventually reaches an unacceptably high level. For continued operation, it is therefore necessary to remove the accumulated solids, and it is important that this can be readily achieved. For this reason, the filter commonly consists of a bed of particulate solids, such as sand, which can be cleaned by back flushing often accompanied by fluidization.

There are two principal modes under which deep bed filtration may be carried out. In the first dead-end filtration, the slurry is filtered in such a way that is fed perpendicularly to the filter medium and there is little flow parallel to the surface of the medium. In the second termed cross flow filtration which is used particularly for very dilute suspension, the slurry is continuously recirculated so that it flows essentially across the surface of the filter medium at a rate considerably in excess of the flow rate through the filter cake.

For a Filtration at constant rate (Richardson et. al., 1991)

$$\frac{dV}{dt} = \frac{V}{t} = \text{Constant}$$

So that

$$\frac{V}{t} = \frac{(-\Delta P)A^2}{(r\mu Vv)}$$

Or:

$$\frac{t}{V} = \frac{r\mu v}{A^2(-\Delta P)}$$

For a Filtration at constant pressure difference (Richardson et. al., 1991)

$$\frac{V^2}{2} = \frac{A^2(-\Delta P)t}{r\mu v}$$

Or

$$\frac{t}{V} = \frac{r\mu v}{2V^2(-\Delta P)}V$$

Thus for a constant pressure filtration, there is a liner relation between V^2 and t or between t/V and V.

2.5.2 The Filter Medium

The function of the filter medium is generally to act as support for the filter cake, and the initial layers of cake provide the true filter. The filter medium should be mechanically strong resistant to the corrosive action of the fluid, and offer as little resistance as possible to the flow of filtrate. Woven materials are commonly used, through granular materials and porous solids are useful for filtration of corrosive liquids in batch units.

2.5.3 Backwash Theory

Rapid gravity filtration removes particles from suspension by attaching them to media or to previously retained particles (Ison and Ives, 1969). Because of the latter mode of attachment, the removal efficiency of filters improves over time after backwashing; this improvement is known as filter ripening. The period of high particulate passage, or ripening, has become a key issue within the water industry. This well-documented period of low filtrate quality is actually caused by a combination of the initial stages of filtration, i.e. the interface with the influent and the presence of backwash water remnants in filtration system. Amirtharajah and Wetstein (1980) reported that the backwash water remnants in the system can be subdivided into three types:

(1) Clean backwash water in the under drain and connecting pipe work from the backwash water supply system up to the bottom of the media.

(2) Backwash water remnants within the pores of the media; and

(3) Backwash water remaining above the filter media up to the level of the backwash water overflow weir.

The duration of the "collapse-pulsing" backwash had a significant effect on filter performance, in terms of the number of 2-5μm particles in the filtrate during the ripening period. Generally, very short backwash durations (1 min) resulted in high levels of particulates in the filtrate during ripening due to inefficient cleaning, and longer durations (5 min) resulted in high levels in the filtrate due to over-efficient cleaning. However, this effect was found to be influenced by media size. The ripening period is the result of two components: the backwash remnants and the interface with the influent. It was the interface with the influent component that was affected by the collapse-pulsing backwash duration. This implies that the longer the duration of collapse pulsing backwash, the cleaner the media becomes and the fewer previously retained particles are left on the filter media to act as additional collectors for the influent.

Backwashing of filter media is a more critical process than filtration itself. Whereas some tolerance can be allowed in setting the filtration rates or media size without losing much in performance, an incorrect backwash rate, for example, can lead to loss of media from the filter if the rate is too high or a rapid deterioration in performance if the backwash rate is low. It is not widely appreciated that the necessary upwash rates are related to the square of the media size, and that for laminar flow with each percentage change in the voidage or porosity there may be a 9.5% change in the required rate (less in the case of transitional flow or combined air and water.

The reference point in washing with water alone or preceded by an air scour is the fluidization threshold which is the point at which the hydraulic gradient through the bed matches the submerged weight of the media at its rest state. At higher rates the bed expands to reduce the hydraulic gradient to maintain the same overall head loss. This reference point has also been used by Amirtharajah (1984) in empirical equations to predict conditions for washing with simultaneous air and water.

2.5.4 Pressure Drop due to the Backwash (Stevenson et al., 1996):

Taking the specific surface area S of a (spherical) grain to be 6/D the hydraulic gradient from the Carman (1937) given by the following set:

$$\mathrm{Re} = \frac{D u \rho}{6 \mu (1 - e)} \qquad \text{------ (3.2)}$$

$$\left(\frac{R'}{\rho u^2} \right) = 5\,\mathrm{Re} + 0.4\,\mathrm{Re}^{-0.1} \qquad \text{------- (3.3)}$$

$$\frac{\Delta p}{h} = \frac{\Delta H \rho g}{h} = \frac{6}{D} \frac{(1-e)^2}{e^3} \rho u^2 \left(\frac{R'}{\rho u^2} \right) \qquad \text{------ (3.4)}$$

The submerged weight of the bed exerts a limiting pressure loss defined by:

$$\frac{\Delta p}{h} = (1 - e)(\rho_s - \rho_L) g \qquad \text{------ (3.5)}$$

The simpler Kozeny (1927) on for streamline flow which is equivalent to the first term only of equation (2) includes a constant, which for granular media is in the range of 4.8-5.0.

$$\frac{\Delta p}{h} = 5 \frac{(1 - e)^2}{e^3} \frac{36}{D^3} \mu u \qquad \text{------ (3.6)}$$

Amirtharajah (1984) has identified a "'collapse pulse" mechanism where small voids are created within the media as the air passes through. These voids collapse as the air migrates to the next level. The overall effect produces abrasion between the grains with negligible bed expansion, indeed often with a bed contraction from the post-fluidized condition.

The voidage of granular media varies with particle shape, size range or distribution and also the previous history of the bed. Materials that have been used for filtration have voidages in the "packed" or vibrated condition ranging from 33 up to 50%. In the working condition after fluidization the voidage has been measured at 43%. If washed in that condition without a preliminary air scour then fluidization will start at that voidage. On the other hand, if an air scour is applied first the bed compacts to about 38% voidage and expansion will start at a lower threshold. Thus, the voidage cannot be taken for granted. Anthracite has a packed voidage of about 42% and displays a similar variation with previous history (Amirtharajah et al., 1991).

2.6 Media Theory

2.6.1 General

Practical filter media is not a single size but contains grains in a size range from 1.5:1 to 3:1. The literature contains various proposals and little justification for the grain size representing the mixture to be used in such equations. However, no single size derived from the size grading plots can really characterize an arbitrary mixture. In view of the basis behind the Carman-Kozeny equation the correct parameter would seem to be the area mean size which, in the recent British Water Standard (Anon, 1993) is designated as the "hydraulic size". The specific area of a sphere is $6/D$; hence the hydraulic size is derived by summing the contribution of each size fraction to the area. (Divide the weight of the fraction by the retaining sieve size, convert the resultant sum back to a size by taking the reciprocal, and then add 10%, which is half the incremental step between the standard sieves normally used, to produce a centerline size rather than the retained size.) This gives the characteristic size that would have the same specific area as the disperse sample. This parameter provides data on the pressure or headloss of the clean media in filtration.

Media with an intrinsic higher voidage offers more capacity for solids in filtration as well as a lower headloss. On the other hand, such media will require a higher upwash flow to achieve the fluidization threshold or a given expansion. The statement, which appears in some older textbooks, that one should use rounded media for filtration, appear to have no foundation. It may be that plants designed for a rounded media have been unable to provide an adequate wash rate for angular media and hence the media has been blamed. Anthracite, a very angular material, has long been accepted as a very effective material.

2.6.2 Media Loss

Granular media are used to remove particulate matter in water treatment processes. They require cleaning at regular intervals and this is usually achieved using water or combined air and water backwash. The medium commonly used in these filters, sand, possesses high attrition resistance and therefore there has been little concern about media losses due to attrition during backwashing. However the use of granular activated carbon (GAC) and anthracite as filter media has been increasing and these media do not possess the attrition resistance of sand. Excessive attrition would lead to a gradual loss of media and its size could reduce significantly (Ives, 1990). Therefore the relative abilities of all these media to withstand attrition need to be established so that the economic and operational effects of attrition can be predicted.

As stated previously, attrition during backwashing is mainly due to abrasion. It can be said that for all media the attrition is very high initially but reduces almost exponentially with time. This high initial burst is probably due to the removal of sharp corners and edges. In the case of sand it may be due to dirt on the grains that was not washed off during the initial water only backwash. Some of the reduction is attributable to the fact that the amount of media in the column is reducing due to particles in the size range being lifted by the air and washed out (Humby and Fitzpatrick, 1995).

Attrition model:

Attrition in fluidized beds was found by Gwyn, 1969 to be represented by an equation of the form

$$W = k\,t^{m} \qquad \text{---------} (3.7)$$

Where: W= weight fraction abraded, t = time and k, m = empirical constants.

The maximum value is the difference between the initial and final amounts of media; the final amount was obtained by adding the media washed out during the test to the amount remaining in the column at the end of the test. Assuming that the amount remaining is all in the size range (a reasonable assumption) then the difference is the amount of media lost due to undersize particles being washed out and to attrition. The difference also includes losses that might have occurred during handling, i.e. during removal, drying, etc.

CHAPTER 3
LITERATURE REVIEW

Baker and Taras (1981) reported that filtration for water treatment dates backs to 2000 b.c.e., where crude sand and charcoal filters were used to provide better testing water. Centuries later Hipocrated designed a cloth bag known as Hippocrates Sleeve, used to remove sediments from water after boiling. In the mid 19th century the spread of Cholera was noticeably decreased where sand filtration was utilized. The benefits of water filtration for not only increasing water aesthetics, but decreasing the spread of disease, lead to the widespread use of filtration seen today when purifying water for potable use.

Boller et al. (1994) investigated buried filters in pilot and full scale and were operated by intermittent flushing which causes the water and the pollutant transport through the unsaturated media to be of highly dynamic nature. Various schemes of hydraulic flushing frequencies were found to be inversely proportional to loading. These findings were confirmed in a full scale plant through monitoring of the dynamic washout of inoxidised matter under different hydraulic loads. The moisture retention capacity of the filter media correlated to the grain size distribution was found to be an important parameter. COD removal and nitrification rates depend strongly on the oxygen supply to the media. In general, oxygen diffusion into the media and the air exchange, induced by intermittent flushing, is sufficient. However, when applying relatively large hydraulic loads to coarse filter grains, especially in the range above 1 mm, buried filters tend to larger breakthroughs of inoxidised matter due to short retention times and instantaneous lack of oxygen. Experiments on average treatment performance have showed that under optimized conditions even wastewaters containing relatively high ammonia contents can fully be nitrified when limestone type filter material is used. Full scale operation revealed further that careful pretreatment (e.g. Septic tank) for the removal of most of the suspended solids is necessary to guarantee safe operation.

Stevenson (1995) reexamined the experimental data published by Amritharajah et al. on the conditions for washing filter media by simultaneous air and water, employing the classical relationships for the pressure loss in granular media developed by Kozeny and Carman, making the assumption that the air and water move through the interstices and about the same velocity. It has been found that for sand Amritharajah's experimental data, as correlated by his empirical equation, can be predicted fairly accurately direct from the hydraulic size and voidage of the media. For low density materials where the buoyancy of the air is much greater than the submerged density of the bed it is suggested, with some support from direct observation, that bubbles are able to disrupt the structure and in some cases churn the bed at zero water flow.

Boller and Kavanaugh (1995) transferred the size density relationships for aggregated particulates in suspension into a model describing the accumulation of particulate deposits in the pore space of granular media filters. Using data from several shallow filter layer experiments. The deposit density and the actual pore volume occupied by the captured particulates were estimated for solids of different characteristics. Based on extension of existing headloss models, the effects of particulate size, particulate density, filtration rate, and media grain size on headloss development during particle deposition were evaluated.

Humby and Fitzpatrick (1995) found that attrition resistance of granular filter media was becoming increasingly important as materials such as granular activated carbon (GAC) and anthracite are becoming more frequently used as filter media. The purpose of their study was to evaluate the attrition experienced by various media during backwashing by performing accelerated backwash tests in a pilot column, using a combined water and air backwash combinations that gave the condition as collapse-pulsing. Since the dominant mode of attrition was assumed to be abrasion, the effluent was sampled at a number of intervals to determine the amount of fine material in the effluent. Coal based GAC exhibited the highest weight loss (~7%) and sand the least (~2%).

Aziz and Smith (1996) discussed the manganese removal from water by filtration through low cost coarse media. A laboratory scale filtration technique was used to remove manganese from manganese bearing water to prove previous batch studies which showed that the removal of manganese was better in the case of limestone particle as compared to the gravel, crushed brick or with no media addition, and the conclusion made that removal mechanisms were due to the effect of rough solid surfaces and the presence of carbonate in the limestone particle. Filtration results indicated that an input pH of 7 with manganese concentration of 1 Mn/l, a good removal was observed in the limestone media as compared to the gravel media, which validates the batch results. Results also show that water hardness did not significantly increase in this filtration technique.

Colton et al. (1996) aimed to optimize direct sand filtration by identifying optimum filter backwash and start-up conditions which minimize the passage of particulates into the filtrate. The performance of rapid gravity filtration plants is believed to be dependent on backwash and start-up regimes. ~~The filter ripening~~ period has long been identified as a cause for concern with respect to particulate passage into the filtrate : this work has shown that up to 40 % of all particles that pass into supply during a 48 hr run, do so in the first hour of operation. Optimum combined air water "collapse pulsing" backwash durations were identified that reduced the number of 2-5 μm particles entering the filtrate, especially during the ripening period. Slow start-up was also found to reduce the number of 2-5 μm particles in the filtrate during the ripening period. The reduction in particulate passage resulting from a slow start was found to be media dependent with smaller media requiring longer slow start duration than coarse media.

Bai and Tien (1997) examined the behavior of particle detachment in deep bed filtration and the dependence of particle detachment on several filtration operating parameters. Theoretical analysis was carried out by considering the balance of various relevant interaction forces. Based on the results, experiments were conducted to study the effect of particle size, filter grain size, and headloss gradient on particle detachment in deep bed filtration. In one set of experiments, the filter was operated at constant filtration rate in order to build up the headloss across the filter bed continuously. In the other set of experiments, the filter in operation was allowed to experience a short period of flow shock (i.e. to be subject to a higher flow rate during this period). Both sets of experiments confirmed the dependence of particle detachment on particle size, collector grain size, and headloss gradient (degree of filter clogging) of the filter, as predicted by the analysis. The study showed that flow shock may cause significant particle detachment and the penetration of small particles into the filter. It is normal in practice that filters in operation often receive a sudden flow rate increase, i.e. a flow shock when some filters are taken out of work or for backwashing. As bacteria or pathogens may be regarded as small size particles, their penetration into filter and their poor retention require further attention for future investigations.

Willemse and Brekvoort (1999) designed a full scale dead-end membrane filtration installation and built at the Eindhoven production plant of the Water Utility Company East Brabant in The Netherlands. The design was based on a flux of 170 l/m^2 h bar and a maximum capacity, investigation into its performance during a longer period shows that the used method proved beyond any doubt to be very appropriate in terms of purification performance, energy and chemical consumption, productivity, recovery and costs. After more than one year, the full scale plant still operates at a flux of 160 l/m^2 h bar. The recovery was found to be 93%. The exploitation costs amounts to 0.15 NLG/m^3. The UF system performance appears to be a stable process and the performance of the full-scale installation demonstrated the technical feasibility and reliability of using UF treatment systems for recovery of backwash water of sand filters from drinking water plants. The productivity amounts to approximately 160 l/m^2 h bar and a recovery of 93% could be achieved. The energy consumption appears to be less than expected and amounts to 0.15 KWH/m^3. The expected quality meets the drinking water standard and very stable process performance was consistently demonstrated during the first year of operation. The operation costs regarding the membrane process are less than f 0.43/ m^3 which made the system very economical. Recently, full scale backwash water recovery by dead-end membrane filtration has been successfully implemented at other locations in the Netherlands.

Hall and Fitzpatrick (2000) obtained experimental data on the development of suspension concentration profiles through laboratory scale filter beds during backwash process. Previous attempts to obtain and record backwash profiles of this type have been unsuccessful due to the limited range of existing turbidity meters. The results have been used to validate a new model developed by the authors. This paper describes experimental work undertaken as part of a larger project attempting to model the backwash process in terms of the volume of clogging deposit to be removed and the time required to remove it. The main objectives of the experimental work were

1. To investigate the development of a suspension concentration profiles at different depths within the filter bed using different backwash regimes using a simple concentration meter, and
2. Compare experimental suspension concentration profiles with those generated from the backwash model for fluidizing water wash only.

The content of the paper was limited to describing the experimental set-up and methods and to also examine some of the results obtained for a fluidizing water wash.

From this they concluded that:

1. Backwash suspension concentration profiles have been measured at different depths within the filter bed and emerging from the bed for a fluidizing water backwash for four backwash water velocities to give different bed expansions up to a maximum of 40%.
2. The shape of the concentration profiles broadly follow those predicted by the model. The model does not however, predict the mixing phenomena as characterized by the dual peaks shown on the measured concentration profiles.

Moore et al. (2001) evaluated Biological aerated filters (BAFs) as an attractive process option, particularly when low land usage is required. They can combine BOD, solids and ammonical nitrogen removal and can be utilized at both secondary and tertiary stages of wastewater treatment. Media selection is critical in the design and operation of BAFs to achieve effluent quality requirements. Two size ranges, 1.5-3.5 and 2.5-4.5 mm, of foamed clay called StarLight C were used in pilot-scale reactors. Both performed well as BAF media, with reactor loads up to 12 kg COD $m^{-3}d$ and 4 kg suspended solids $m^{-3}d$ (based on working volumes). The most consistent effluent was obtained using the smaller medium since, at flow rates above 0.4 l/min, the BAF using the larger medium produced an effluent containing more than 20 mg/l of suspended solids for over 30 min after backwashing. Up to 70% longer run times, as determined by reaching a set head loss, were recorded for the BAF containing the larger rather than the smaller medium. Additionally, the development of pressure above the smaller medium filter bed tended to be logarithmic rather than linear. Reactor profiles indicated that suspended solids removal did not occur over the full 2.3 m depth of the columns. The BAF containing the smaller medium utilized a mean depth of 1.7-0.3 m, whereas a mean depth of 2.1-0.3 m was used by larger medium BAF. Both the head loss development data and the suspended solids removal profiles indicated that the smaller medium BAF was underperforming as a filter. The tracer study results for large StarLight highlight that poor performance especially immediately after backwashing, may result from preferential channeling. Meanwhile, the reactor containing the smaller StarLight showed better plug flow conditions, which improves nitrification and solids removal.

Bourgeous et al. (2001) investigated the effects that wastewater quality and mode of operation have on the performance on an asymmetric, hollow fiber, polysulfone, ultrafiltration (UF) membrane with a molecular weight cutoff of 100,000 Daltons. Performance was assessed through monitoring membrane flux, transmembrane pressure, effluent biochemical oxygen demand, and operational cost of the experimental system while treating filtered secondary, secondary and filtered primary effluents. Fluxes achieved for filtered secondary (129 ±173 l/m^2h), secondary (101 ± 158 l/m^2 h), and filtered primary (20±41 l/m^2 h) effluents were compared to those obtained at three other locations. A conceptual model of the impact of an insufficient backwash and of operating the UF system at constant flux on membrane performance is presented to explain the differences in fluxes. Employing pre-membrane granular filtration to remove a protion of the problematic particles in secondary effluent prior to UF led to optimal operational conditions.

Although the use of recirculation could increase maintainable flux when treating a concentrated feed (eg. Filtered primary effluent), the associated costs were high. Improved UF performance was found to result from allowing flux to decline naturally, rather than using a constant flux mode of operation. The effluents produced when filtered secondary and secondary effluents were the feeds would be equivalent to an oxidized, coagulated, clarified, and filtered wastewater.

Nieuwenhuijzen et al. (2001) investigated the feasibility of DIF as a pretreatment step for advanced particle removal. With a large scale pilot-plant filter at WWTP Leiden-Noord, The Netherlands the removal characteristics for suspended and colloidal material were investigated as well as operational conditions of eight different filter configurations. From the experimental research it was concluded that filtration of raw waste water is possible, not withstanding the relatively short runtimes due to clogging. In general, the filters produced a filtrate with a constant quality with low concentrations of solids and low turbidity. Without addition of chemicals hardly any colloids and soluble were retained, but only suspended particulates were removed. After dosage of iron or polymer, it was possible to remove more suspended matter and a high proportion of colloidal material. Finally it was concluded that DIF could be applied as a compact treatment system to produce a high quality primary effluent with a constant composition, but for practical application further research has to be done.

Rooklidge and Ketchum (2002) evaluated that the associated decrease in pH in slow sand filters, due to CO_2 conversion and biological activity, may produce effluent that is slightly corrosive to downstream distribution pipe material. This pilot study examined the use of a 3-cm crushed dolomite limestone media layer placed within the filter column of a slow sand filter to enhance effluent corrosion control by the introduction of beneficial dolomite, dissolution products, without impacting turbidity removal efficiencies. Turbidity removal, calcium concentration, pH, conductivity, total hardness and alkalinity changes were calculated for the filter during a 60 day pilot study, and water chemistry values were used to estimate the changes of the saturation index (SI) throughout the filter run. Total hardness change through the filter was compared to change calculated by a derived equation for hardness using calcium concentrations to determine if the media was dissolving in stoichiometric proportions, and mineral service life in the filter was estimated using an assumption of stoichiometric dissolution at a constant flow rate. Effluent SI was raised an average of 30%, alkalinity was increased by 195 and effluent pH averaged 7.7. Filter effluent compiled with current turbidity regulatory requirements for the provision of potable water, and mineral service life was estimated between 7.5 and 9.5 years. It was concluded that microbial activity within slow sand filters treating raw water of moderated pH and low alkalinity may reduce effluent pH to the extent of being slightly acidic. This reduction may be compensated by the addition of a dolomite limestone layer within the sand media bed, which increases effluent pH due to mineral dissolution.

For the raw water source used in this study during the 60-day period, the dolomite-amended slow sand filter produced treated effluent that adhered to current turbidity regulatory requirements.

The service life of dolomite media was estimated to be within an acceptable range for use in water treatment facilities ~~experiences raw water chemistry similar~~ to that encountered during pilot study. The effects of biological growth on dolomite media dissolution were not known for the pilot raw water source, and continued study of this treatment technology is encouraged.

Kim et al. (2002) used membrane filtration to treat a secondary effluent emanating from a sewage treatment works that treats a combined industrial and municipal wastewater. Three feed pretreatments for a spiral wound reverse osmosis (RO) membrane filtration were evaluated, including:

(i) membrane ultrafiltration (system1);

(ii) dual media filtration and granular activated carbon (GAC) adsorption(system 2);

(iii) dual media filtration with dosage of organic flocculent and GAC adsorbion(system 3);

It is shown that system 1 yielded the best turbidity removal, with turbidity below 1.15 NTU. The combination of system 1 and RO showed the least flux decline between cleans. In addition, flux recovery was easily achieved with mechanical clean without chemicals. The overall total dissolved solids (TDS) rejection was well maintained at 81-89%. The dual media filter and GAC did not provide adequate pretreatment; this led to rapid fouling of RO membrane. The impact on RO performance was a greater flux decline coupled with TDS rejection decrease from 78 to 66%. The addition of an organic flocculant (dosed at 15 mg/l to form filterable flocs) did not significantly improve the performance of the dual media filtration and the GAC. It was also observed that inadequate pretreatment had an adverse impact on the membrane flux recovery by cleaning. Simple mechanical cleaning was insufficient in recovering when systems 2 and 3 were employed as pretreatment. Furthermore, longer chemical cleaning duration was required to recover membrane flux.

Freitas et al. (2006) evaluated in his work performance of double-layered ceramic filters of aerosol filtration at high temperature. The filtering structure was composed of two layers; a thin granular membrane deposited on the reticulate ceramic support of high porosity. The goal was to minimize the high pressure drop inherent of granular structures, without decreasing their high collection efficiency of small particles. The reticular support was developed using the technique of ceramic replication of polyurethane foam substrates of 45 and 75 pores per inch (ppi). The filtering membrane was prepared by depositing a thin layer of granular alumina-clay paste on one face of the support. Filters had their permeability and fractional collection efficiency decreased with the gas temperature and was enhanced with fractional time. Also, the support layer influenced the collection efficiency: the 75 ppi support was more effective than 45 ppi. Particle collection efficiency dropped considerably for particles below 2μm in diameter. The maximum collection occurred for particle diameters of approximately 3μm, decreased again for diameters between 4 and 8μm. Such trend was successfully represented by the proposed correlation. Inertial impaction seems to be the predominant collection mechanism, with particle bouncing/re-entrainment acting as detachment mechanisms.

A double-layered ceramic filter was developed and tested for use for gas filtration at high temperature. Tests included measurement of porosity, permeability and filtration performance in temperatures ranging from ambient to 700°C. Results showed that the filters presented high collection efficiency, comparable to other ceramic filters reported in the literature. Fractional collection efficiency was sensitive to the gas temperature, to the structural (foam) layer and to filtration time. For similar experimental conditions, it is decreased with the increase in temperature. Also, the efficiency was lower for particles below 2μm and in the range 4-8μm and explanation is based in the influence of different dust particle collection mechanisms. A correlation that estimates the fractional efficiency was proposed, being based on semi-empirical expressions available in the literature. Results showed satisfactory agreement between experimental data and the proposed correlation. They also revealed the particle inertia as the predominant collection mechanism within the studied range, with particle bouncing/re-entrainment acting as detachment mechanisms.

Tanaka et al. (2006), developed a novel granular medium consisting (1.5 5 mm in diameter) of inert perlite particles as nuclei and an effective surface layer containing sulfur, $CaCO_3$ and $Mg(OH)_2$ for advanced treatment of agro-industrial wastewater. The performance of the medium was examined with a laboratory-scale down flow fixed-bed column reactor using piggery wastewater, which had been treated by an upflow anaerobic sludge blanket reactor and trickling filter.

The novel medium developed in this study consisted of porous perlite particles as inert nuclei and a sulfur matrix outer layer in which powdered $CaCO_3$ and $Mg(OH)_2$ were embedded. The purpose of adding $Mg(OH)_2$ in addition to $CaCO_3$ was to add phosphate removal ability to the medium. It was hypothesized that this medium would have the following advantages: -

1. Since $CaCO_3$ and $Mg(OH)_2$ were homogenously dispersed in the sulfur matrix, their dissociation would proceed simultaneously with sulfur oxidation.
2. The spherical shape of the medium would be an advantageous characteristic for filtration process since it might homogenize flow-velocity distribution in the medium bed and head loss.

3. Expansion and fluidization of the medium bed would be easy during back washing due to the lower density and smaller diameter (1.5 - 5mm) of the medium; and
4. $Mg(OH)_2$ may contribute to PO_4^{3-} -P removal by establishing a weak alkaline condition that would the crystalline of phosphate with calcium and/or magnesium.

The study demonstrated that a medium consisted of perlite, sulfur, $CaCO_3$ and $Mg(OH)_2$ was useful for the removal of nitrate and phosphate. The loading rate of nitrate should not exceed 0.3 kg NO_3^- -N m^{-3} d^{-1} to ensure high nitrate removal efficiency. The effluent pH should be above 7.9 to remove phosphate. Evidently $Mg(OH)_2$ contributed strongly to the phosphate removal, and that raising the $Mg(OH)_2$ content in the medium effectively extended the life of the medium for phosphate removal.

Machin et al. (2006) conducted a study on new granular material (Lapilli) for hot gas filtration. We have characterized this material by means of set of instrumental methods involving: surface area (S_{bet}), scanning electron microscopy (SEM), X-ray diffractometry (XRD), thermogravimetry (TG), differential thermal analysis (DTA) and X-ray flouroscence (XRF). Also, they used a fixed bed heat exchanger-filter (FHEC) for removing fine dust particles from gases. The influence of a number of variables was examined, including gas velocities, filter height and gas temperature. Also, a model was proposed to describe the granular filtration of the FHEF using Lapilli as a granular medium. A numerical solution of the filtration equations based on the finite element method was presented.

CHAPTER-4
EXPERIMENTAL PROGRAMME

4.1 EXPERIMENTAL SETUP

The experimental work was carried out on a direct filtration setup. A down flow media filter was built from a 1200mm tall plexiglass tube with a 50mm internal diameter (Fig.4.1), to which a centrifugal pump is attached for the continuous flow of the liquid through the column. The 250mm crushed rock support media was installed in layer of decreasing size from 2 mm at the bottom to 20mm at the gravel/media interface having 250mm depth. The depth of media in the filter was varied from 200mm to 500mm with effective size of 0.30 mm. The flow rate of liquid was controlled by two globe valves. The change in the pressure drop was measured with the help of a manometer. The two limbs of the manometer are connected to the taps in the column between the lowest and highest point of the media bed. A $0.2m^3$ capacity holdup tank was used to maintain the continuous flow of liquid.

So the main parts of setup are

- A 1200mm Plexiglass column
- A $0.2m^3$ hold up tank
- A maximum capacity of $0.120m^3/h$ Centrifugal Pump.
- Valves for controlling the flow rate.
- A rotameter to measure the flow rate of water in the range of 10 dm^3/h to 120 dm^3/h.
- Backwash arrangement

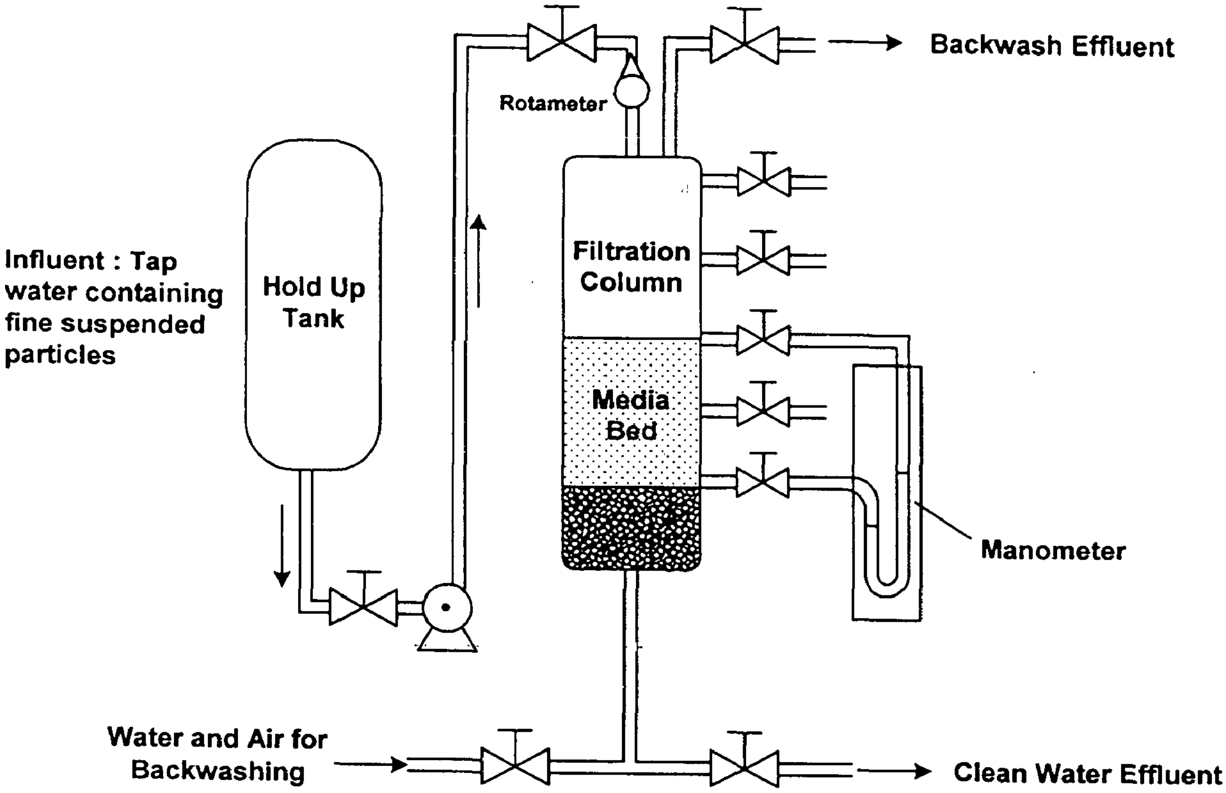

Fig. 4.1: Experimental Setup for Filtration through Coarse Media

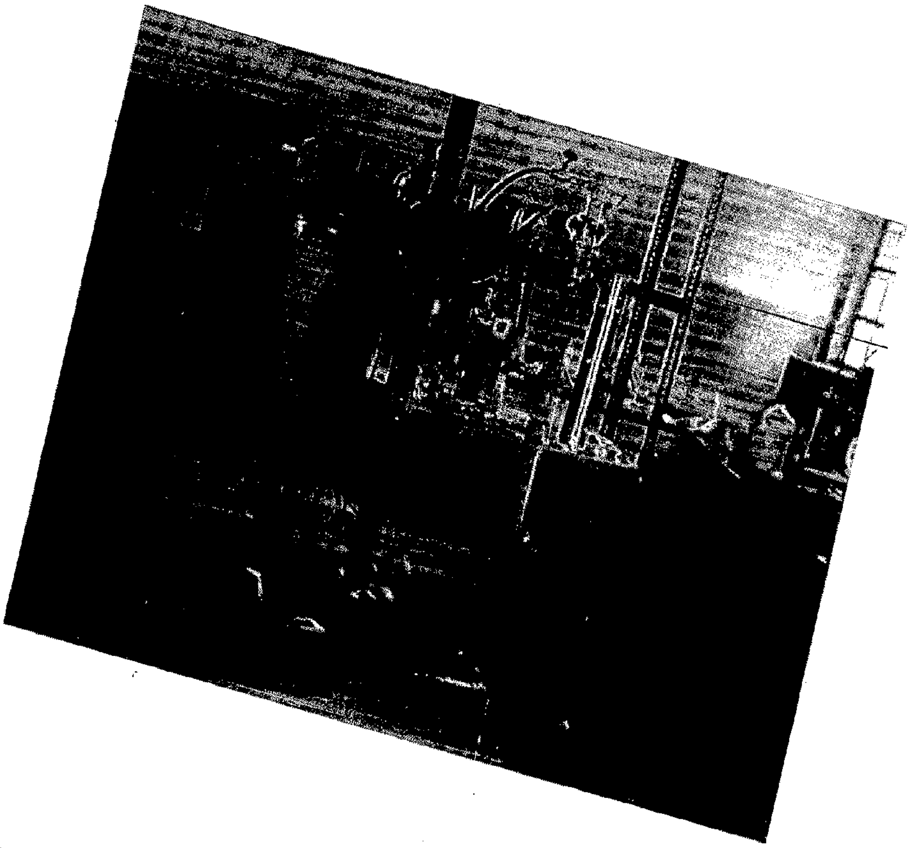

Figure 4.2: Actual Setup of Filteration Column in Bio-Chemical Lab, IIT Roorkee

4.2 DESIGN PARAMETER

The choice of filter design for a given process depend on many factors such as the particle size and shape distribution, the state of aggregation, the properties of the filtrate, the quantity of the material handled, media size, etc.

Design Parameters for the designing of media bed filter are

• Flow Rate (dm^3/h)	60 – 100
• Specific Velocity ($m^3/m^2.h$)	28 – 32
• Media Size (mm)	0.3 – 0.5
• Supporting Media size (mm)	2 - 20
• Flowing Liquid	Tap water
• Material of construction of filter tube	Plexiglass

On the basis of design parameter experimental setup has following dimensions

• Diameter of filter (mm)	50
• Total height of filter (mm)	1200
• Supporting media bed height (mm)	250
• Space for backwashing	more than 50% from bed height.
• Media bed height (mm)	200 - 500

4.3 MATERIAL AND METHOD

The experimental setup was operated as a down-flow filter. The solid-water suspension was pumped up to an inlet pipe and was introduced above the filter bed. With a rising water level in the filter column the flow through the filter bed was kept at a constant velocity. The water was passed through the filter bed and the filter bottom into a filtrate overflow. The nozzles in the filter floor were protected against clogging by the gravel support layer. At the end of the filtration process either because of the clogging or breakthrough of the bed, the filter bed was cleaned with up-flow backwashing with air and water or a combination of both. The effects of four parameters were studied on the filtration and the removal efficiency of the fine suspended particle, viz.

(1) The effect of the filtration media ;

(2) The effect of the filtration media bed height ;

(3) The effect of the cycle time (at 30min, 60min, and 90min)

(4) The effect of the pressure changes.

4.3.1 Procedure of filtration

Three types of filter media were used, viz. calcite, anthracite and garnet. Then their combinations were prepared in the ratio of 1:1. The bed height was varied in the range of 200 mm to 500 mm and the cycle time was varied in the range of 30 min to 90 min. The slurry is dispersed with a shower into the filter from the top of the media filter. The column was operated at a filtration rate 60 – 100 dm^3/h with a concentration of solid particles of 20 mg/dm^3. The particle size was in the range of 0.5 to 20μm. The media particles were soaked in water for 24 h before being transferred into the column. The media was supported on a 75 mesh stainless steel screen while allowed the filtration of fine suspended particles from tap water to ensure that all the air trapped in the media particle pores was removed. After the run of 90 min for each media at particular bed height the bed was backwashed. At an interval of 30 min 3 samples of water was taken from the column outlet for the analyzing of particle concentration and the particle size distribution.

4.3.2 Procedure of Backwashing

The procedure for backwash was that the media was fluidized with water to a 50 % bed expansion (based on dry length). The combined water and air flow rates that gave the collapse–pulsing condition were determined using the procedure given by Amirtharajah et al. (1991*)*. At collapse – pulsing air cavities form and collapse within the media. The Collapse of an air cavity result in rapid downward movement of a section of the bed due to the downward movement of section of the bed due to the media moving to fill the space left by the air cavity. The procedure consisted of setting the flow rate to a particular value and gradually increasing the water flow rate until collapse-pulsing was observed at the end of backwash the test was terminate and the media was removed from the column, dried and weight.

4.4 ANALYSIS

The particle size and the distribution of the slurry solid phase is measured with particle size analyzer (product of ANKERSMID) named CIS systems at the Instrumentation Analysis Lab, IIT Roorkee. The CIS systems employ laser and video measurement channels for particle size analysis. For the measurement of the particle size and the distribution of the slurry solid phase a magnetic stirring cell is used. It contains a cuvette with a magnetic stirrer that disperses the particles equally in the liquid medium in order to maintain a uniform suspension of particle.

Main step to analyze the sample are:

1. Switch on the particle size analyzer.
2. Take the sample volumes from 1.5 ml to 3.5 ml in the cuvette and place it into the magnetic stirring CIS system.
3. Set all the parameter required according to specification for analysis with CIS system.
4. Start the measurement.
5. Collect and save the data.
6. Remove the cuvette from system, wash it properly and put another sample in the cuvette for another measurement.

Scanning electron microscopy (SEM) of active biomass sample was also taken by using LEO Electron Micrograph at the Institute Instrumentation Centre, IIT Roorkee

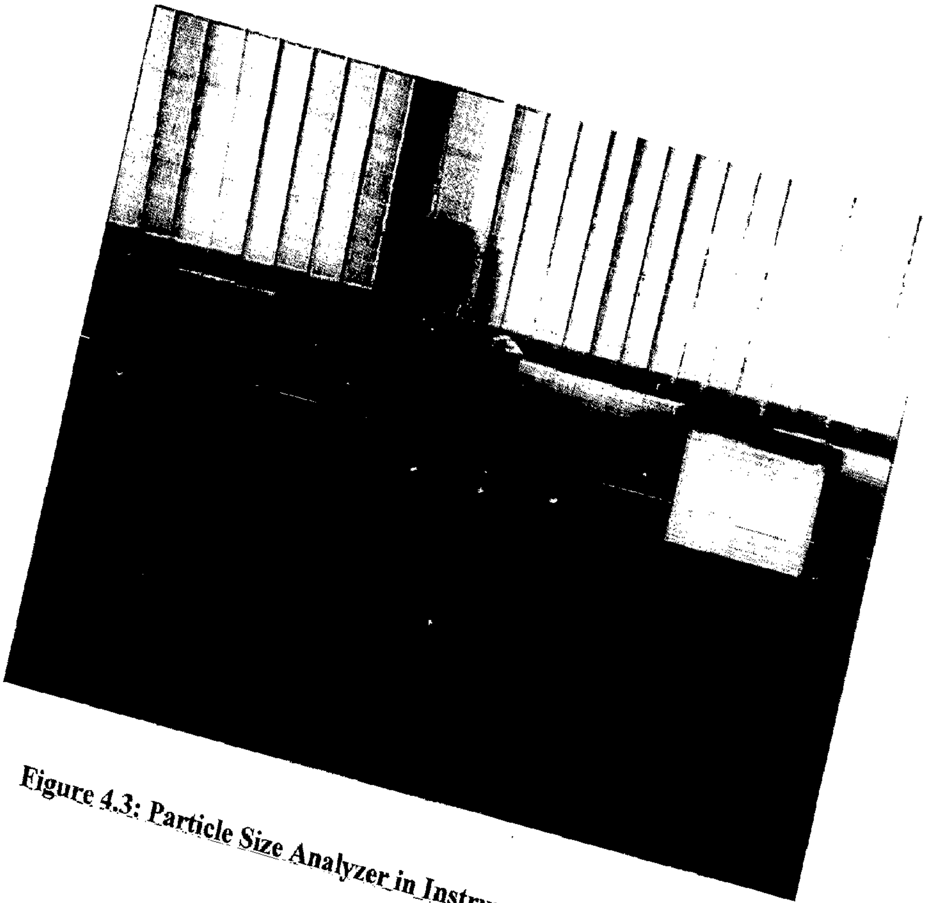

Figure 4.3: Particle Size Analyzer in Instrumental Analysis Lab, IIT Roorkee.

CHAPTER 5
RESULTS AND DISCUSSION

5.1 GENERAL

A detailed study of particle characteristics removal efficiencies and effect of pressure drop in granular media filtration is presented in this chapter.

5.2 SEM OF GRANULAR MEDIA

Scanning electron microscopy of the Granular Media used in this study (Calcite, Anthracite, and Garnet) was done. This was done basically to study the physical structure of media. It shows clearly the pores and crevices within the media at higher magnifications. It was clear from the figures obtained that Garnet is more porous as compared to Anthracite and Calcite. Fig. 5.A shows the SEM of Anthracite at different magnifications (500X, 600X, 700X). Fig 5.B shows the SEM of calcite at different magnifications (400X, 500X, 600X). Fig 5.C shows the SEM of Garnet at different magnifications (300X, 400X, 500X).

5.3 REMOVAL EFFICIENCY AND PRESSURE DROP CALCULATIONS

The removal efficiency of suspended particles depends on many factors like the filter bed height, media characteristics etc. Mathematically, it is calculated as

$$\eta = 100 * (N_i - N_o) / N_i$$

Where η is the removal efficiency

N_i is the number of Suspended Particles present in Inlet Stream.

N_o is the number of Suspended Particles present in Outlet Stream.

Percentage Removal of Suspended particles is calculated as above and Pressure Drop are presented in tabular forms in Table 1, 2, 3, and 4. Table 1 represents removal efficiency and pressure drop of Calcite-Anthracite, Anthracite-Garnet, and Garnet-Calcite mixed bed at a bed height of 200 mm. The flow rate was kept constant for a particular cycle. Each cycle was carried on for duration of 90 minute. Samples were collected and analyzed at an interval of 30 min. Four cycles were carried on at a particular bed height. The first cycle at a constant flow rate of 30 dm^3/hr, the second cycle at a constant flow rate of 60 dm^3/hr , third cycle at the flow rate of 90 dm^3/hr and the fourth cycle at the flow rate of 120 dm^3/hr.

Similarly, Table 2 represents removal efficiency and pressure drop calculations of Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite mixed bed at a bed height of 300 mm. Other conditions were kept the same as in Table 1.

Table 3 represents removal efficiency and pressure drop calculations of Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite mixed bed at a bed height of 400 mm. ~~Other conditions being the same as kept in Table 1 and 2~~.

Table 4 represents removal efficiency and pressure drop calculations of Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite mixed bed at a bed height of 500 mm. Other conditions being the same as kept in Table 1, 2 and 3.

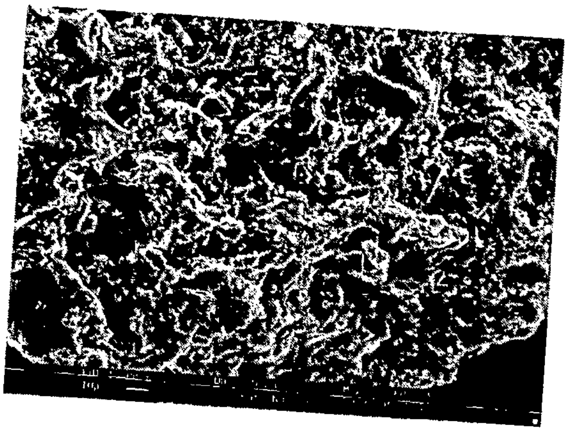

Fig. 5.A1 SEM of Anthracite at 500 X

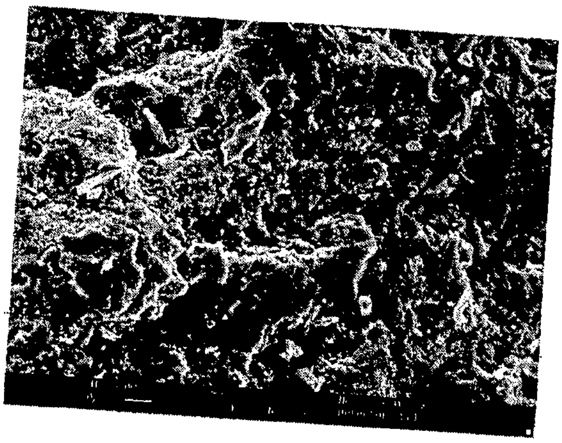

Fig. 5.A2 SEM of Anthracite at 600X

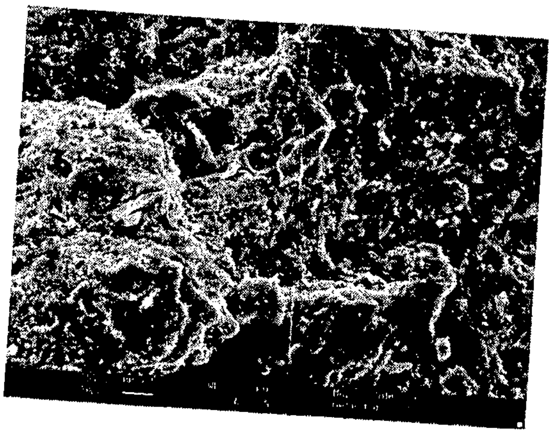

Fig 5.A3 SEM of Anthracite at 700 X

Fig. 5.A: SEM of Anthracite

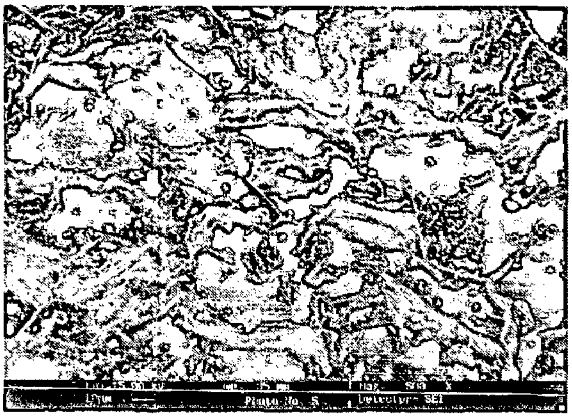

Fig 5.B1 SEM of Calcite at 500 X

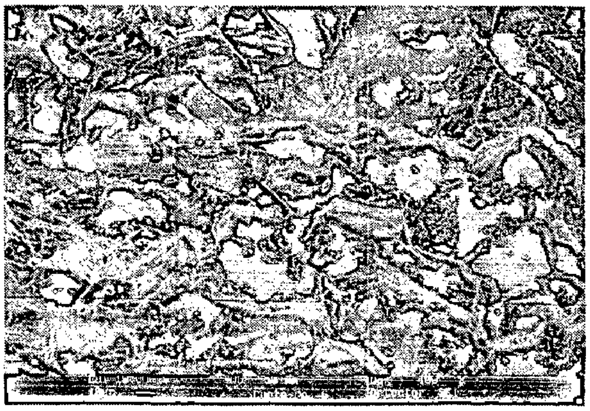

Fig. 5.B2 SEM of Calcite at 400 X

Fig 5.B3 SEM of Calcite at 600X

Fig. 5.B: SEM of Calcite

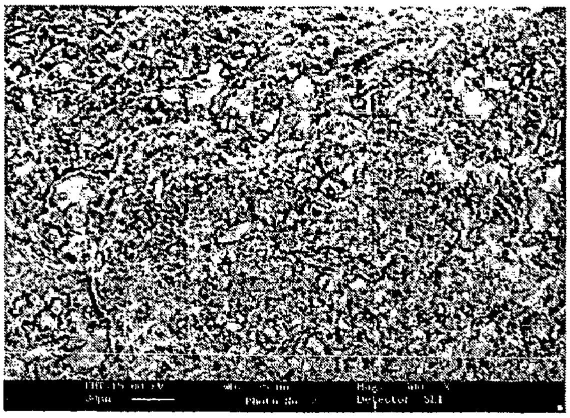

Fig 5.C1 SEM of Garnet at 300 X

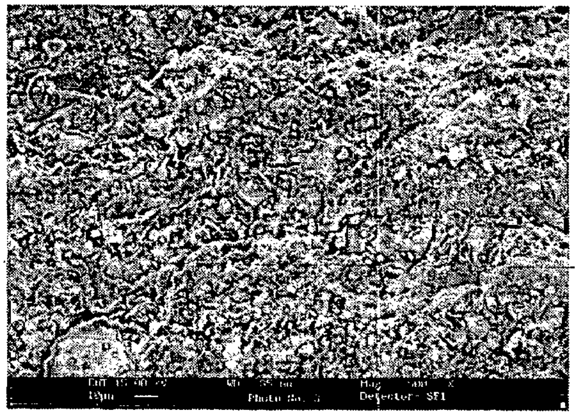

Fig 5.C2 SEM of Garnet at 500 X

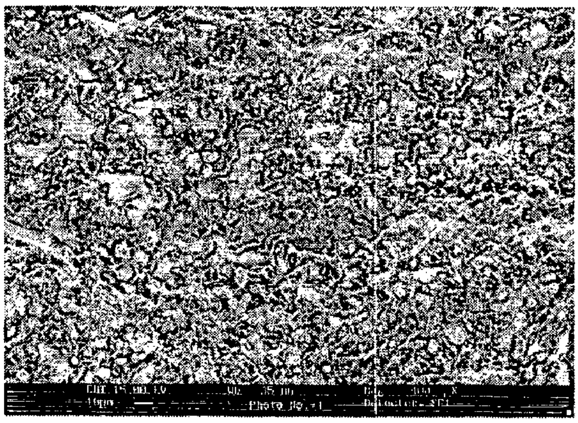

Fig 5.3C SEM of Garnet at 400 X

Fig. 5.C: SEM of Garnet

Table 1 : Percentage reduction and observed pressure drop of various media at bed height 200 mm

Media			Calcite-Anthracite			Anthracite-Garnet			Garnet-Calcite		
Bed height (mm)	Flow rate (dm^3hr^{-1})	Sample	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop
200	30	S-0 feed sample	1253	-	-	73	-	-	2193	-	-
		S-1 at t=30min	103	92	3708	11	85	1236	0.4	99.98	23486
		S-2at t=60min	20	98	8653	2	97	5562	0.13	99.99	10507
		S-3 at t=90min	5	99.6	3708	1	99	2472	0.2	99.99	28183
200	60	S-0 feed sample	1253	-	-	1593	-	-	2193	-	-
		S-1 at t=30min	365	71	2843	53	97	1236	3.0	99.86	22867
		S-2at t=60min	254	80	5810	148	91	1236	1.0	99.95	25339
		S-3 at t=90min	270	79	6551	126	93	3090	0.1	99.99	36464
200	90	S-0 feed sample	1253	-	-	2850	-	-	2193	-	-
		S-1 at t=30min	349	72	44498	463	62	2472	11	99.5	8653
		S-2at t=60min	719	43	44498	535	88	1607	3	99.86	18541
		S-3 at t=90min	625	50	24721	735	57	9889	99	95.5	1236
200	120	S-0 feed sample	1253	-	-	713	-		2193	-	-
		S-1 at t=30min	353	72	1236	469	34		3.0	99.86	4944
		S-2at t=60min	283	77	6057	77	89		0.5	99.97	45734
		S-3 at t=90min	173	86	8529	3	99.6		2.0	99.9	53151

Table 2 : Percentage reduction and observed pressure drop of various media at bed height 300 mm

Media			Calcite-Anthracite		Anthracite -Garnet	Garnet-Calcite			Garnet-Calcite		
Bed height (mm)	Flow rate (dm^3hr^{-1})	Sample	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop
300	30	S-0 feed sample	1253	-	-	93	-	-	2193	-	-
		S-1 at t=30min	143	89	1236	1	98.9	1731	1.0	99.95	1236
		S-2at t=60min	8	99	3585	0.2	99.8	247	0.2	99.99	2472
		S-3 at t=90min	18	98.6	4449	0	100	2472	0.13	99.99	3708
300	60	S-0 feed sample	1253	-	-	794	-	-	2193	-	-
		S-1 at t=30min	214	83	7664	72	91	618	2.0	99.91	2472
		S-2at t=60min	9	99	5439	0.5	99.94	1483	0.2	99.99	6180
		S-3 at t=90min	23	98	12114	0.0	100	1236	1.0	99.95	3708
300	90	S-0 feed sample	1253	-	-	165	-	-	2193	-	-
		S-1 at t=30min	421	66	8776	25	85	2349	1.0	99.95	2225
		S-2at t=60min	153	88	10012	6	96	1236	0.1	99.99	1731
		S-3 at t=90min	7	99	9517	0.1	99.9	2472	0.3	99.98	3708
300	120	S-0 feed sample	1253	-	-	469	-	-	2193	-	-
		S-1 at t=30min	241	81	13968	50	89	6180	10	99.54	8652
		S-2at t=60min	249	80	15080	9	98	4944	1.0	99.95	3956
		S-3 at t=90min	54	96	15080	19	96	7416	2.0	99.9	11125

Table 3 : Percentage reduction and observed pressure drop of various media at bed height 400 mm

Media			Calcite-Anthracite			Anthracite-Garnet			Garnet-Calcite		
Bed height (mm)	Flow rate (dm^3hr^{-1})	Sample	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop
400	30	S-0 feed sample	1253	-		815	-	-	2193	-	-
		S-1 at t=30min	14	99	6675	71	91.3	1236	0.2	99.99	2472
		S-2at t=60min	2	99.8	5686	0.3	99.96	4449	19	99.13	4449
		S-3 at t=90min	0.1	99.99	4203	0.7	99.92	1236	1609	26.6	1236
400	60	S-0 feed sample	1253	-	-	549	-	-	2193	-	-
		S-1 at t=30min	140	89	5315	7	99	3708	0.01	99.99	3708
		S-2at t=60min	6	99.5	7664	11	98	2967	0.2	99.99	8652
		S-3 at t=90min	3	99.8	7416	36	93	6181	0.015	100	6180
400	90	S-0 feed sample	1253	-	-	1498	-	-	2193	-	-
		S-1 at t=30min	44	96	9147	14	99	8282	3	99.86	7911
		S-2at t=60min	52	95.8	7911	3	99.8	7416	2	99.9	6922
		S-3 at t=90min	2	99.8	8899	9	99.4	7416	0.03	99.99	6675
400	120	S-0 feed sample	1253	-		65	-	-	2193	-	-
		S-1 at t=30min	23	98	10754	0.49	99.25	14586	2	99.9	12361
		S-2at t=60min	61	95	11619	0.23	99.65	11866	0.2	99.99	11867
		S-3 at t=90min	59	95.2	12114	No particles found	100	14091	0.4	99.98	13844

Table 4 : Percentage reduction and observed pressure drop of various media at bed height 500 mm

Media			Calcite-Anthracite			Anthracite-Garnet			Garnet-Calcite		
Bed height (mm)	Flow rate (dm^3hr^{-1})	Sample	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop	Number of particle (ml^{-1})	Percent reduction %	Pressure drop
500	30	S-0 feed sample	1253	-	-	171	-	-	2193.0	-	-
		S-1 at t=30min	23	98	1236	0.8	99.53	3708	15.0	99%	5191
		S-2at t=60min	6	99.5	2967	0.26	99.85	3214	2.0	99.9%	6675
		S-3 at t=90min	59	95	1237	0.0008	99.99	4944	0.1	99.99%	8652
500	60	S-0 feed sample	1253	-	-	98	-	-	2193.0	-	-
		S-1 at t=30min	262	79	1236	0.1	99.89	6675	73.0	97%	4450
		S-2at t=60min	11	99	7416	0.14	99.86	10754	1.0	99.95%	495
		S-3 at t=90min	0.8	99.9	7169	0.096	99.9	10630	0.01	99.99%	3214
500	90	S-0 feed sample	1253	-	-	4.6	-	-	2193	-	-
		S-1 at t=30min	2	99.8	6675	0.1	97.82	13597	51	98	16069
		S-2at t=60min	0.01	99.99	6675	0.322	93	13102	10	99.5	15079
		S-3 at t=90min	0.2	99.98	6180	0.013	99.71	12360	19	99.13	13597
500	120	S-0 feed sample	1253	-	-	73	-	-	2193	-	-
		S-1 at t=30min	34	97.3	19035	66	9.59	15822	422	81	19530
		S-2at t=60min	0.17	99.99	18047	0.87	98.8	17552	153	93	21013
		S-3 at t=90min	0.14	99.98	18294	0.0034	99.9	20148	7	99.7	25957

5.4 EFFECT OF TIME ON THE REMOVAL EFFICIENCY AT A FIXED BED HEIGHT AT DIFFERENT FLOW RATES

Fig. 5.1 represents the removal of fine suspended particle for Calcite-Anthracite as Media at 200 mm bed height. This graphical representation shows that as the time increases the removal efficiency increases. But, it is also clear that the removal efficiencies at the time period of 60 min and 90 min are almost the same. They don't vary much as compared to the removal efficiency at 30 min. Also, the removal efficiency is highest at the flow rate of 30 dm^3/hr, then comes 60 dm^3/hr, then 120 dm^3/hr and finally 90 dm^3/hr. The removal efficiency at 30 dm^3/hr is 92% at 30 min, 98% at 60min and 99.6% at 90 min. Thus, it is seen that there is not much difference in the removal efficiencies at 60 and 90 min.

Also, the removal efficiency at 120 dm^3/hr is 72% at 30 min, 77% at 60 min & 86 % at 90 min. From comparing the data at 30 dm^3/hr and at 120 dm^3/hr, it is clear that as the flow rate is increasing removal efficiency is decreasing

Fig 5.2 represents the removal of fine suspended particle for Calcite-Anthracite as media at 300 mm bed height at various flow rates. In these graphical representations, the removal efficiency at 30 dm^3/hr is quite low as compared to 60 dm^3/hr initially, but as the cycle proceeds, the removal efficiency increases. It is lowest at 30 min at all the flow rates and highest at 90 min for all the flow rates. The removal efficiency at 30 dm^3/hr is 89.5% at 30 min, 99% at 60 min and 98.6 % at 90 min. At 120 dm^3/hr, the removal efficiencies are 81% at 30 min, 80 % at 60 min and 96% at 90 min. It is showing the same trend as Fig. 5.1 i.e. as the time is increasing and flow rate is decreasing, removal efficiency is decreasing.

Fig 5.3 shows the removal of fine suspended particle for Calcite-Anthracite as media at 400 mm bed height. From this graphical representation, it is clear that the removal efficiency is highest at 30 dm^3/hr at all times. The removal efficiency at 120 dm^3/hr is quite high at 30 min (98%) but then it constantly decreased. It was 95.5% at the end of 60 min and 95.2% at the end of 90 min. At 60 dm^3/hr, the removal efficiency was very low initially (89% at 30 min) but then it increased to 99.5 % at 60 min and 99.8% at 90 min. The removal efficiency at 90 dm^3/hr also showed same trend. It was low (96% at 30 min) and then increased to 99.8% at the end of 90 min. Fig. 5.3 also showed the same trend as Fig 5.1 and Fig 5.2.

Fig 5.4 shows the removal of fine suspended particles for Calcite-Anthracite as media at 500 mm bed height. In this graphical representation, removal efficiency is initially high for the flow rate of 90 dm^3/hr but as the time proceeded the removal efficiencies become almost same at 30 dm^3/hr and 90 dm^3/hr. The removal efficiency at 30 dm^3/hr is 98% at 30 min, 99.5 % at 60 min and 95% at 90 min. Also, the removal efficiencies at 120 dm^3/hr is 97.3 % at 30 min, 99.99 % at 60 min and 99.98% at 90 min.

Fig 5.5 shows the removal of fine suspended particle with Anthracite and Garnet as Media Bed at 200 mm bed height. This graphical representation shows that the removal efficiency initially increases with time but then after approximately 60 min, it becomes almost constant. Also, as the flow rate is increasing, removal efficiency is deteriorating i.e. it is 97.5% at 60 min at 30 dm^3/hr while it is 89% at 60 min at 120 dm^3/hr.

Fig 5.6 shows the removal of fine suspended particle with Anthracite-Garnet as media bed at 300 mm bed height. It again shows the same trend as the previous graphs that as the time is proceeding removal efficiency is first increasing then becoming constant. Also, removal efficiency is best at low flow rates i.e. at 30 dm^3/hr, removal efficiency is 100% at 90 min and 99.8 % at 60 min whereas at 60 dm^3/hr removal efficiency is 96% at 90 min and 98% at 60 min.

Fig 5.7 shows the removal of fine suspended particle with Anthracite-Garnet as media bed at 400 mm bed height. The graph shows that the removal efficiency is lowest at 30 dm^3/hr initially (91.3% at 30 min) but then it increases and becomes 99.96% at 60 min and 99.92 % at 90 min. The removal efficiency is high initially at 60 dm^3/hr (99% at 30 min) but then it degrades to 93% at 90 min.

Fig 5.8 shows the removal of fine suspended particles with Anthracite-Garnet as Media Bed at 500 mm bed height, it is clear from this graph that the removal efficiency is almost same at flow rates of 30 dm^3/hr and 60 dm^3/hr at all times. But the removal efficiency at 90 dm^3/hr first decreases (93% at 60 min) and then increases again(99.71% at 90 min).

Fig 5.9 shows the removal of fine suspended particles with Calcite-Garnet as media bed at 200 mm bed height. Here, also as in previous graphs, the removal efficiency is highest at low flow rates (30 dm^3/hr) and at time intervals of 60 and 90 min.

Fig 5.10 shows the removal of fine suspended particle with Calcite-Garnet mixed bed at 300 mm bed height. The removal efficiency is almost same at 30 and 60 dm^3/hr flow rate (99.13% at 30 dm^3/hr and 99.99% at 60 dm^3/hr at 60 min).

Fig 5.11 shows the removal of fine suspended particle for Calcite Garnet as media at 400 mm bed height. Here the removal efficiencies are quite high and almost equal at the end of 90 min for 60 dm^3/hr, 90 dm^3/hr and 120 dm^3/hr, but for 30 dm^3/hr, it decreases continuously as time progresses.

Fig 5.12 shows the removal of fine suspended particle with Calcite-Garnet mixed bed at 500 mm bed height. Here again same trend follows as in the previous graphs. At flow rate of 30 dm^3/hr, removal efficiency is highest (99.99% at 60 min) while at 120 dm^3/hr, removal efficiency is 93% at 60 min.

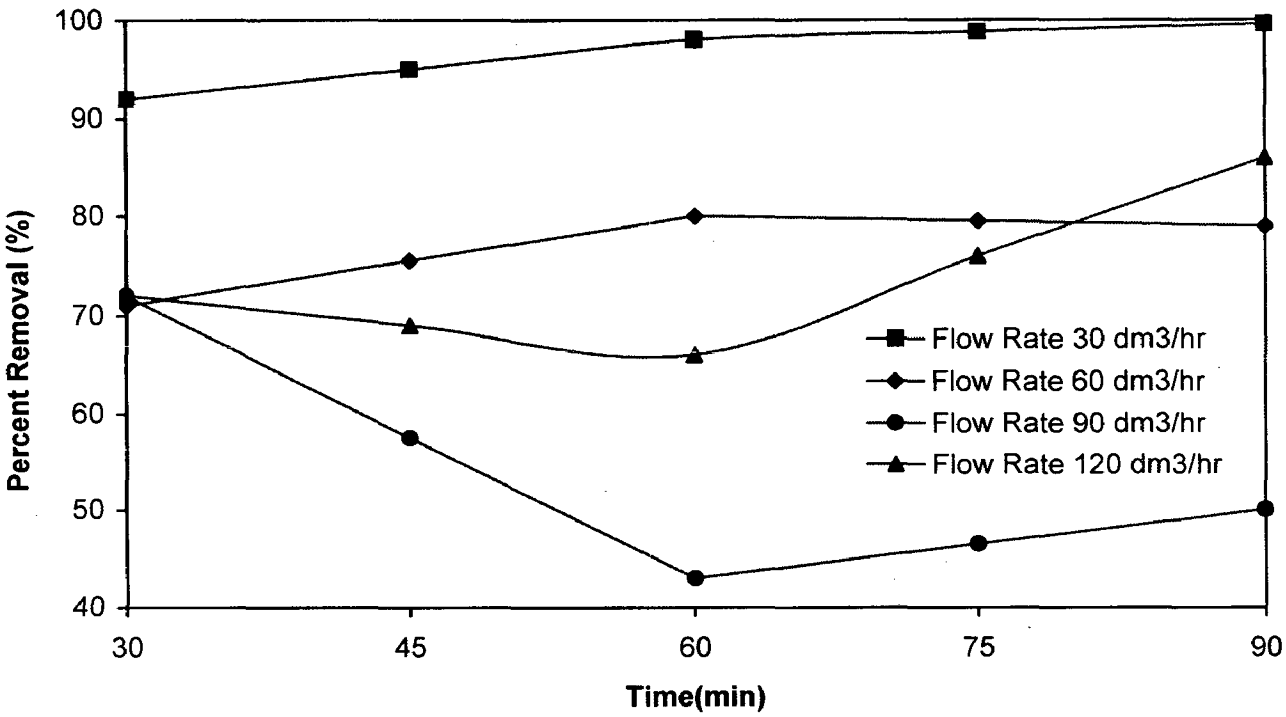

Fig. 5.1 Removal of Fine Suspended Particle for Calcite-Anthracite as Media 200mm Bed Height

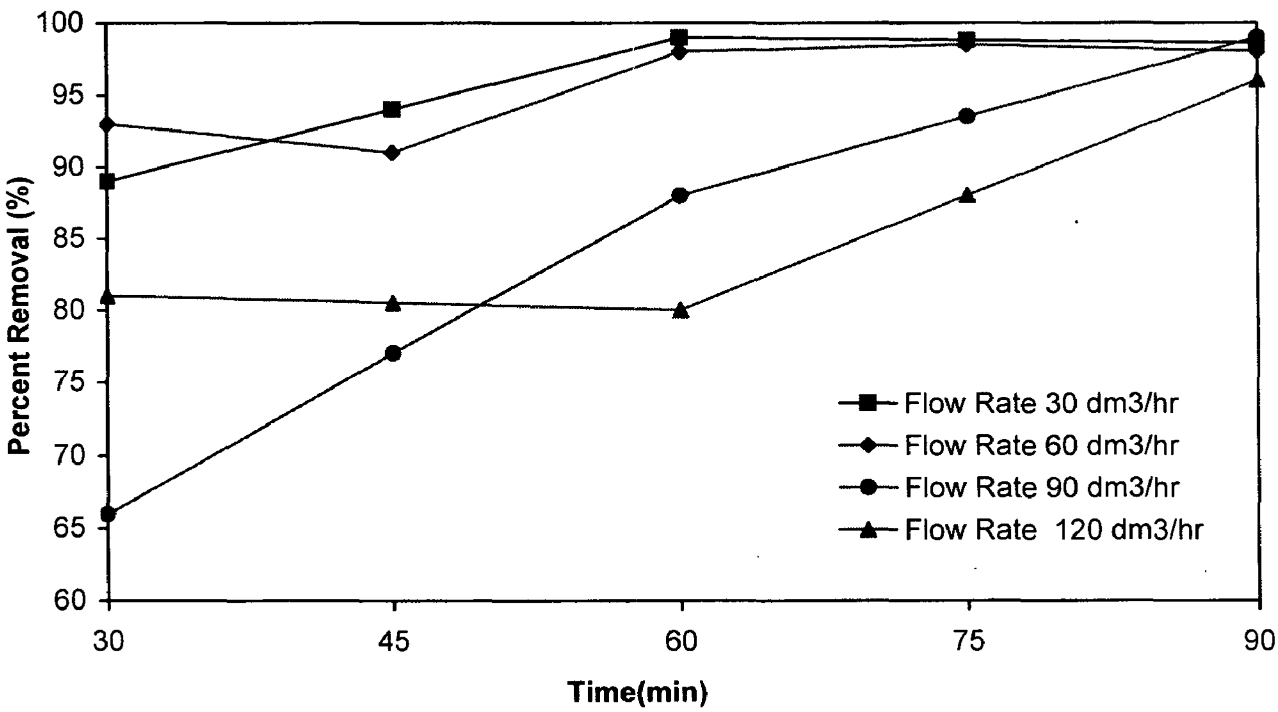

Fig. 5.2 Removal of Fine Suspended Particle for Calcite-Anthracite as Media at 300mm Bed Height

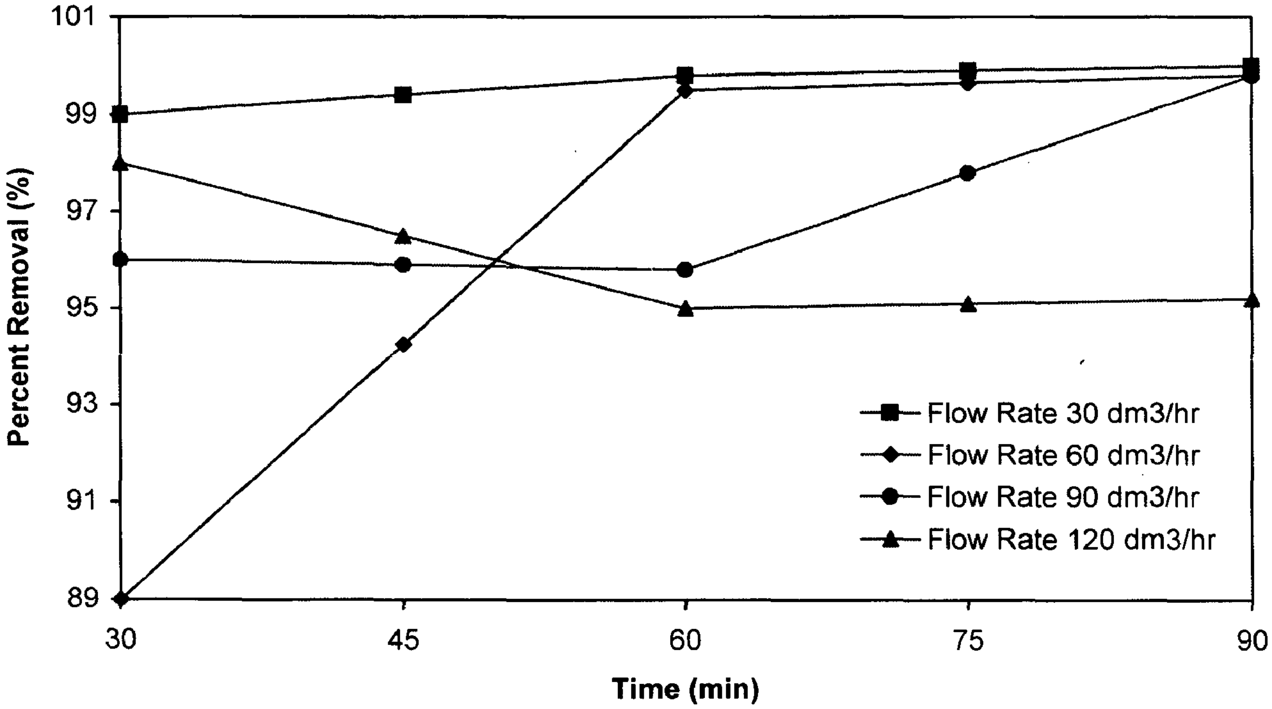

Fig 5.3 Removal of Fine Suspended Particle for Calcite-Anthracite as Media at 400mm Bed Height

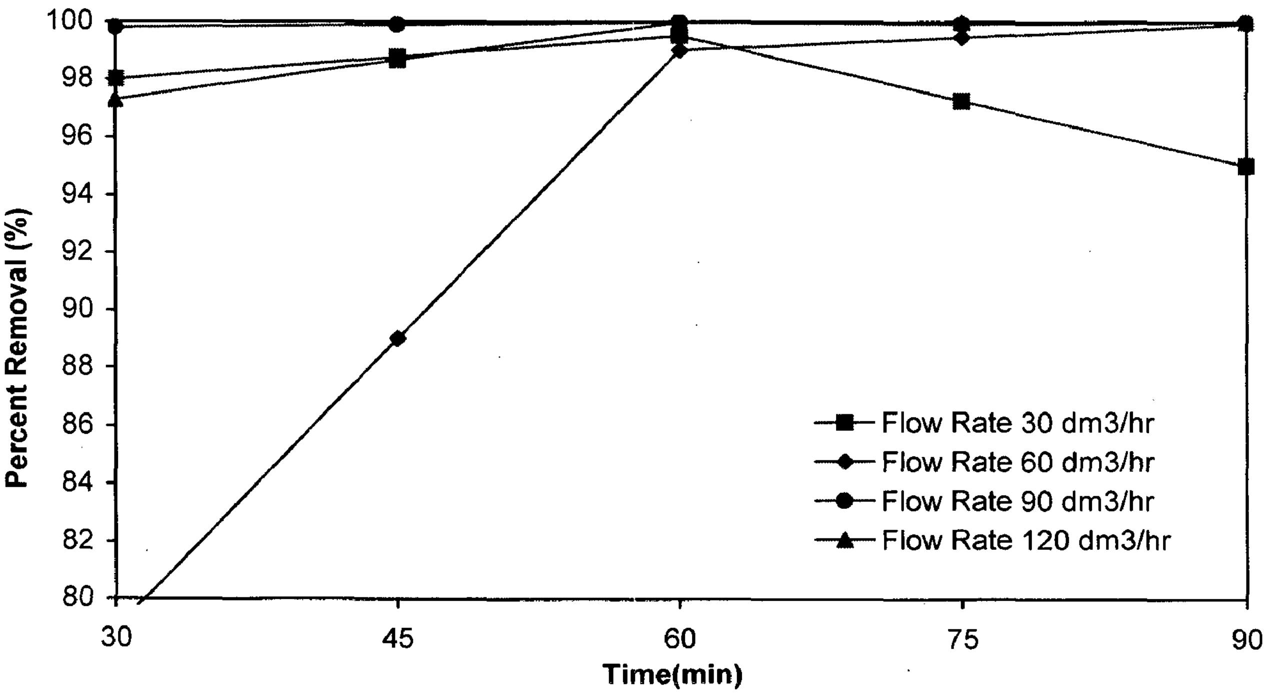

Fig. 5.4 Removal of Fine Suspended Particle for Calcite-Anthracite as Media at 500mm Bed Height

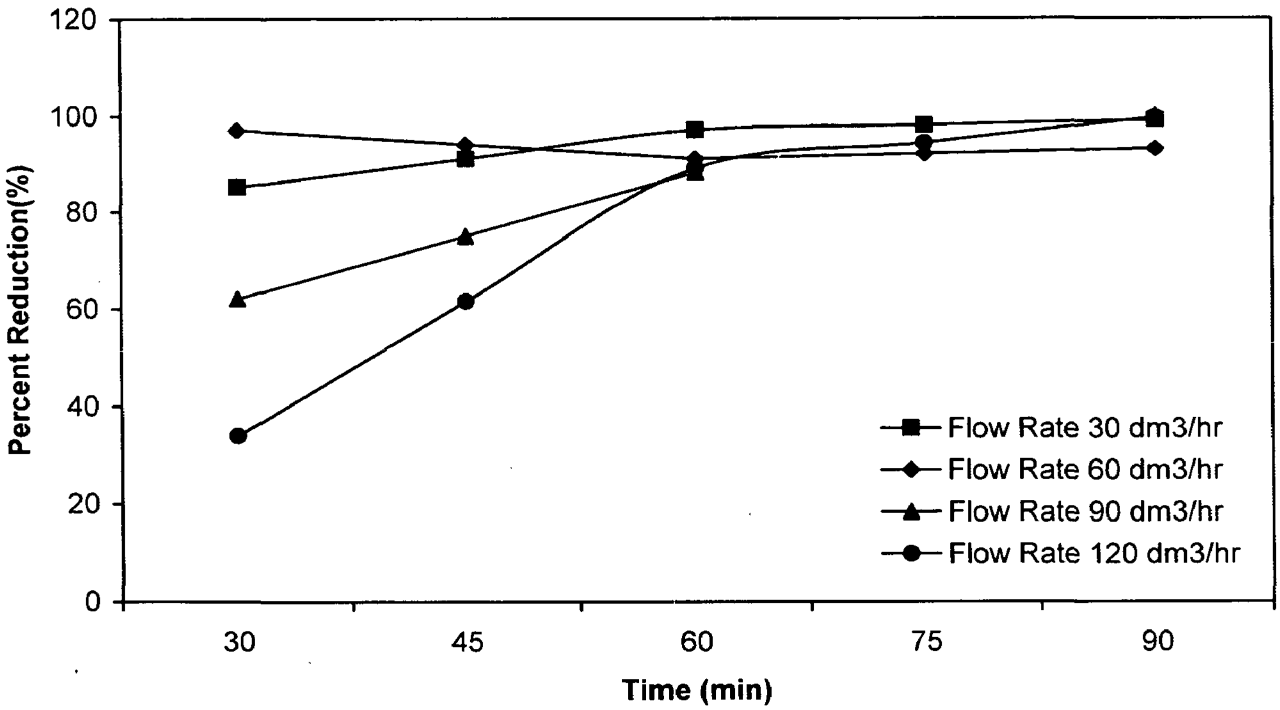

Fig 5.5 Removal of Fine suspended Particle with Anthracite-Garnet as Media Bed at 200 mm bed height

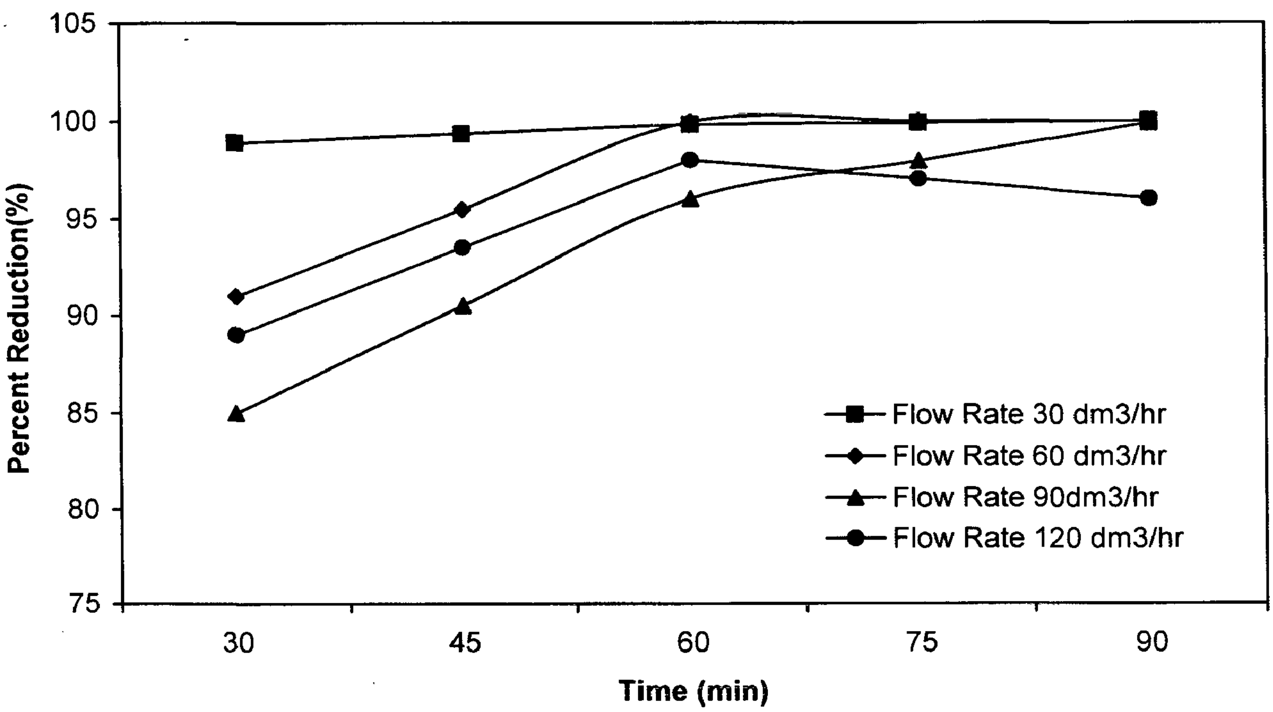

Fig. 5.6 Removal of Fine suspended Particle with Anthracite-Garnet as Media Bed at 300 mm bed height

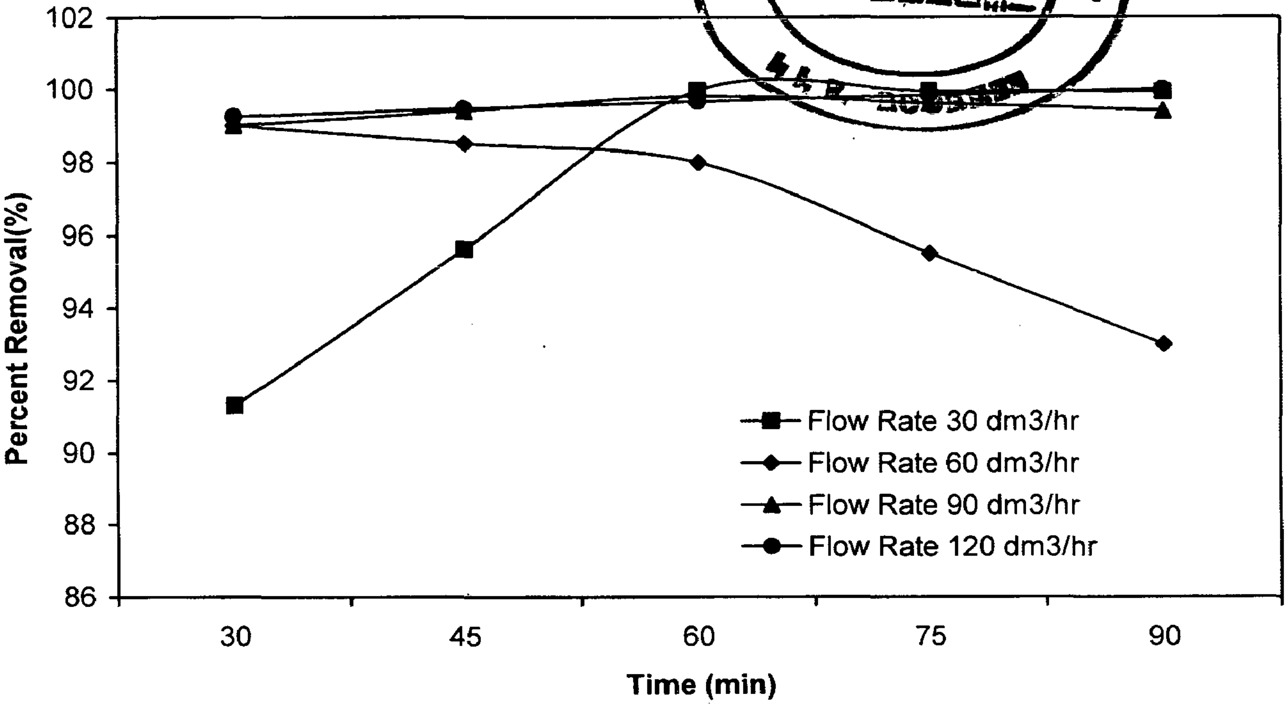

Fig. 5.7 Removal of Fine suspended Particle with Anthracite-Garnet as Media Bed at 400 mm bed height

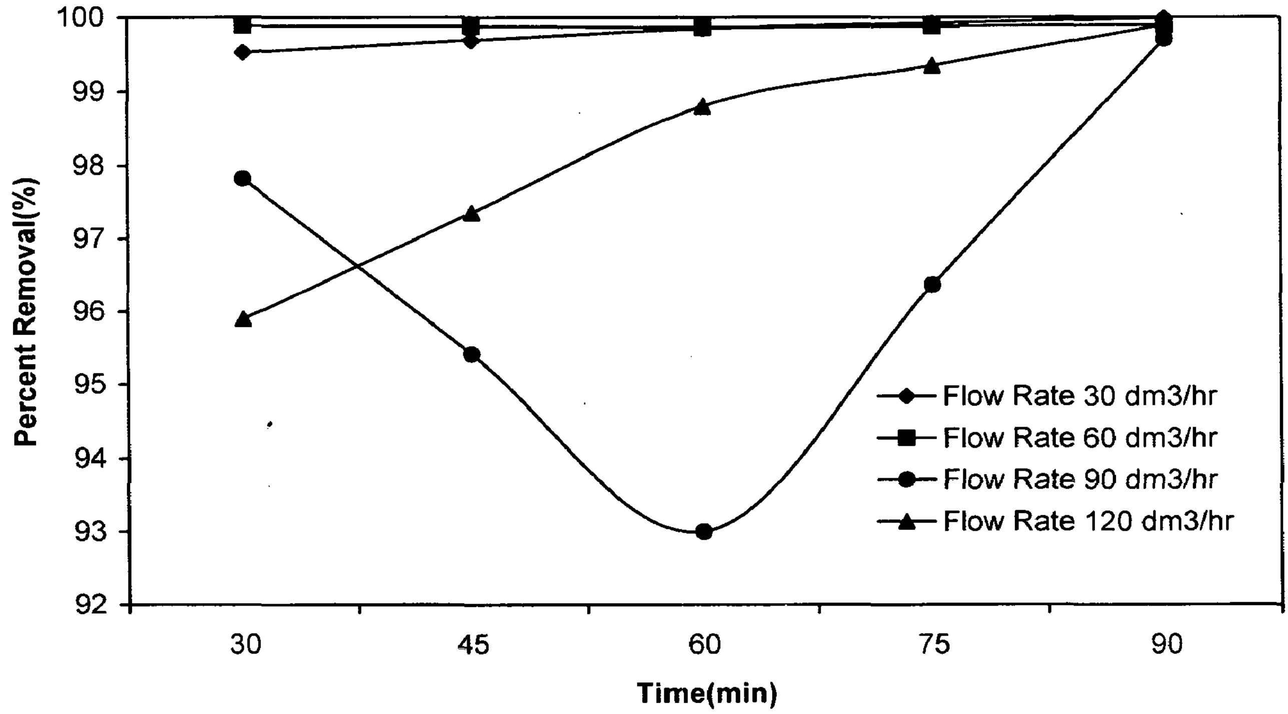

Fig. 5.8 Removal of Fine suspended Particle with Anthracite-Garnet as Media Bed at 500 mm bed height

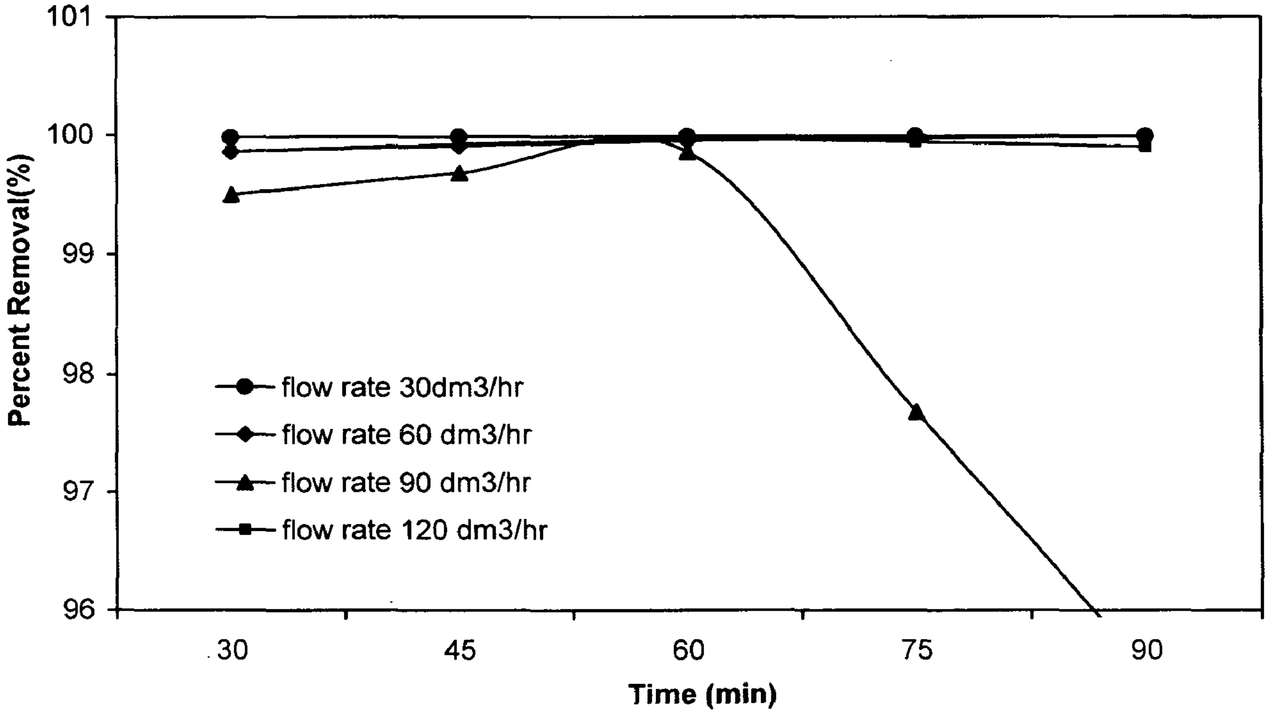

Fig. 5.9 Removal of Fine Suspended Particle with Calcite-Garnet as Media at 200 mm Bed Height

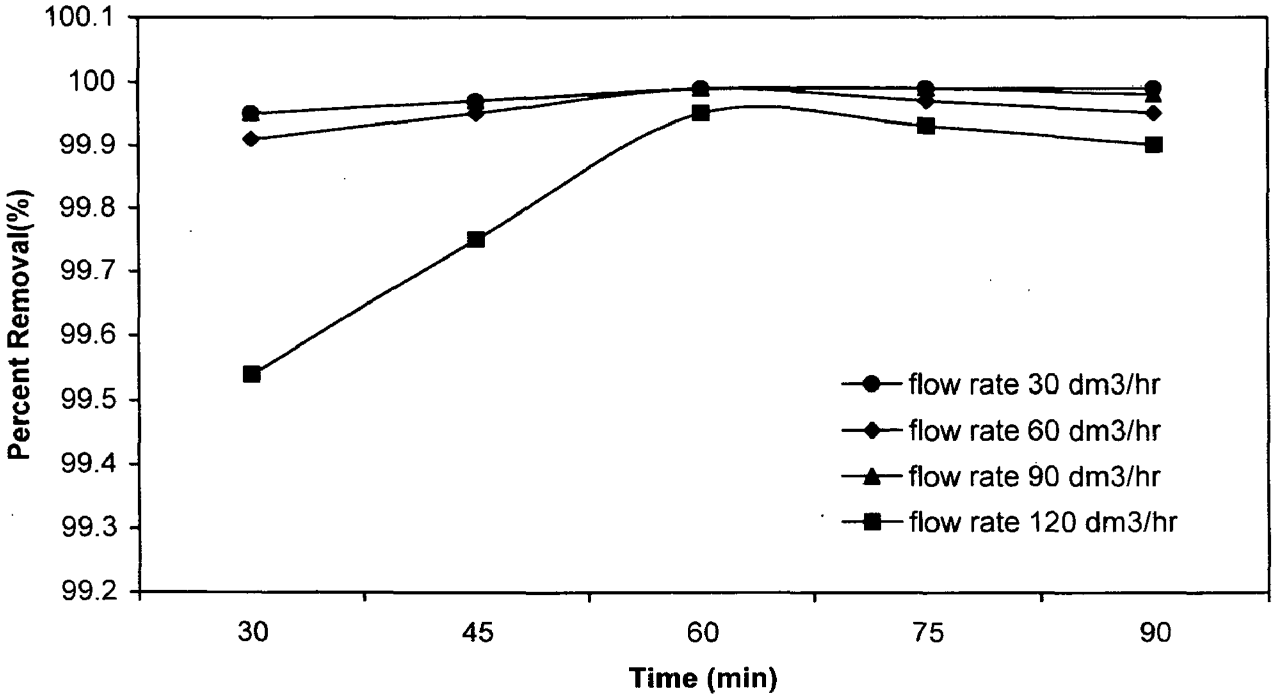

Fig. 5.10 Removal of Fine Suspended Particle with Calcite and Garnet mixed bed at 300 mm Bed Height

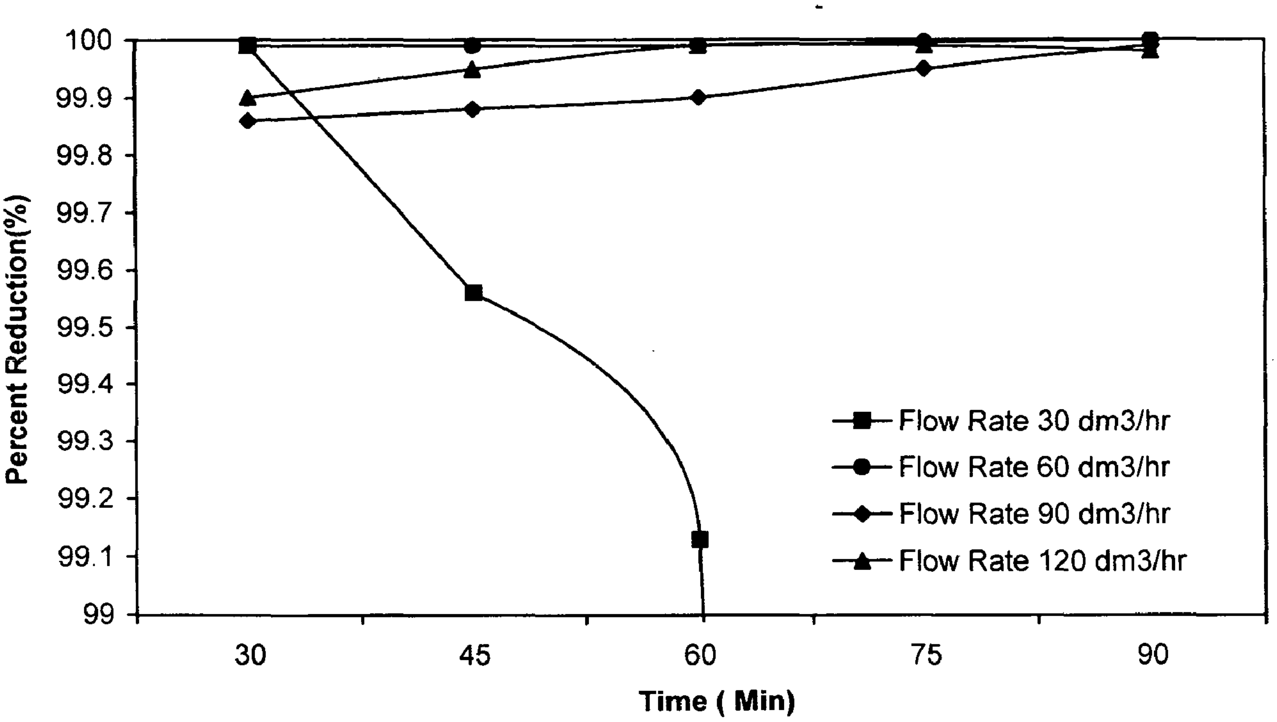

Fig. 5.11 Removal of Fine Suspended Particle for Calcite and Garnet as Media at 400 mm bed height

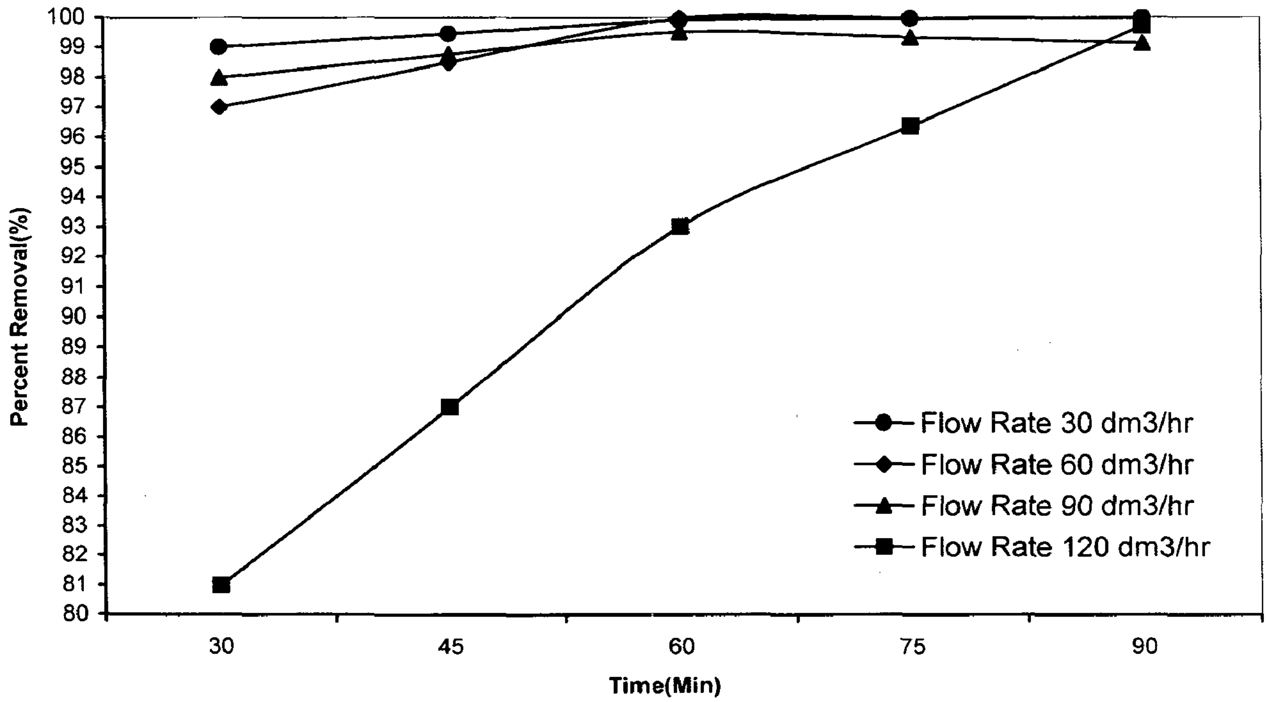

Fig. 5.12 Removal of Fine Suspended Particle with Calcite and Garnet mixed bed at 500 mm Bed Height

5.5 EFFECT OF BED HEIGHT ON REMOVAL EFFICIENCY AT A FIXED FLOW RATE AT VARIOUS TIME PERIODS

Fig 5.13 shows the effect of removal of fine suspended particle at different bed height of Calcite-Garnet mixed bed at 30 dm^3/hr flow rate. In this graphical representation, it is found that at 30 min, at 200 mm bed height, the removal efficiency is 99.38%.It remains approximately the same till 400 mm bed height then it suddenly decreases. While at 60 min, initially the efficiency is high at 200-300 mm bed height then it decreases at 400 mm bed height and then increases again. While at 90 min, the efficiency continuously increases as the bed height increases.

Fig 5.14 shows the removal of fine suspended particle at different bed height of Calcite-Garnet mixed bed at 60 dm^3/hr flow rate. The removal efficiency at 30 min is lowest while it is approx the same at 60 min and 90 min (99.95% at 60 min and 99.99% at 90 min at the bed height of 500 mm). It implies that as the time is increasing the removal efficiency first decreases then remains constant. Also, as the bed height increases, removal efficiency increases.

Fig 5.15 shows the removal of fine suspended particle at different bed height of Calcite-Garnet mixed bed at 90 dm^3/hr flow rate. Here as the bed height is increasing, the removal efficiency first increases and then decreases.

Fig 5.16 shows the removal of fine suspended particles at different bed height of Calcite-Garnet mixed bed at 120 dm^3/hr flow rate. In this plot as the bed height is increasing, removal efficiency is increasing at 30 and 90 min but at 60 min it increases till 400 mm and then decreases.

Fig 5.19 represents the removal of fine suspended particle at different bed height of Anthracite-Garnet at 90 dm^3/hr flow rate. Here in this plot as the bed height is increasing from 200 mm to 500 mm removal efficiency is highest at 90 min at 200 mm bed height and is approximately the same as 60 min at higher bed heights.

Fig 5.20 shows the removal of fine suspended particle at different bed height of Anthracite-Garnet at 120 dm^3/hr. Here also, it can be seen that as the bed height increases and as the time increases, removal efficiency increases.

Fig 5.21 shows the removal of fine suspended particles at different bed height of Anthracite-Calcite mixed bed at 30 dm^3/hr flow rate. Initially, at 30 min, at 200 mm bed height, poor removal efficiency is observed (85%) which increases as the bed height increases (98% in 500 mm) at 60 and 90 min. Also efficiency increases marginally as the bed height increases.

Fig 5.22 shows removal of fine suspended particle at different bed height of Anthracite-Calcite mixed bed at 60 dm^3/hr flow rate. Here at 60 min as the bed height ~~increases from 200 mm to 500 mm,~~ efficiency initially increases (80% at 200 mm, 99% at 300 mm), it then decreases (93% at 400 mm) and then again increases 99% at 500 mm.

Similar patterns are shown by Fig 5.23 and Fig 5.24, that with the increase of bed height, the removal efficiency increases. Also, the increase in time is favored.

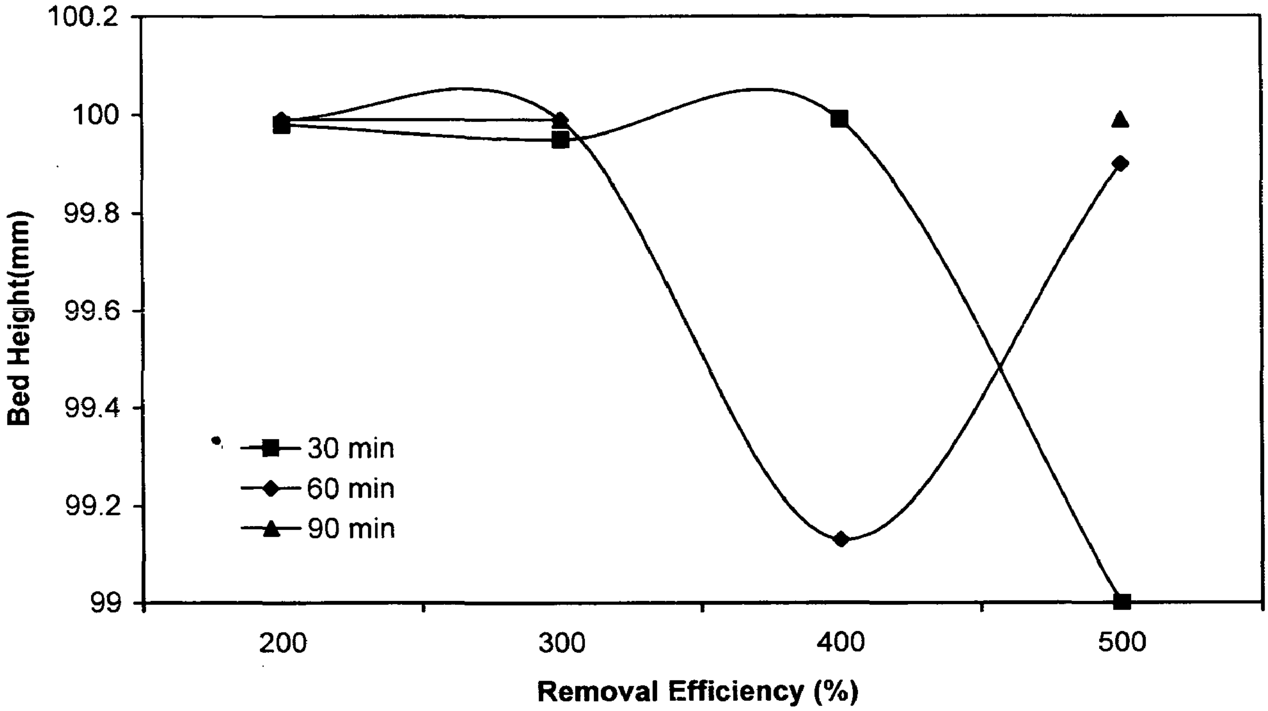

Fig. 5.13 Removal of Fine Suspended Particle at different Bed Height of Calcite and Garnet mixed bed at 30 dm^3/hr flow rate

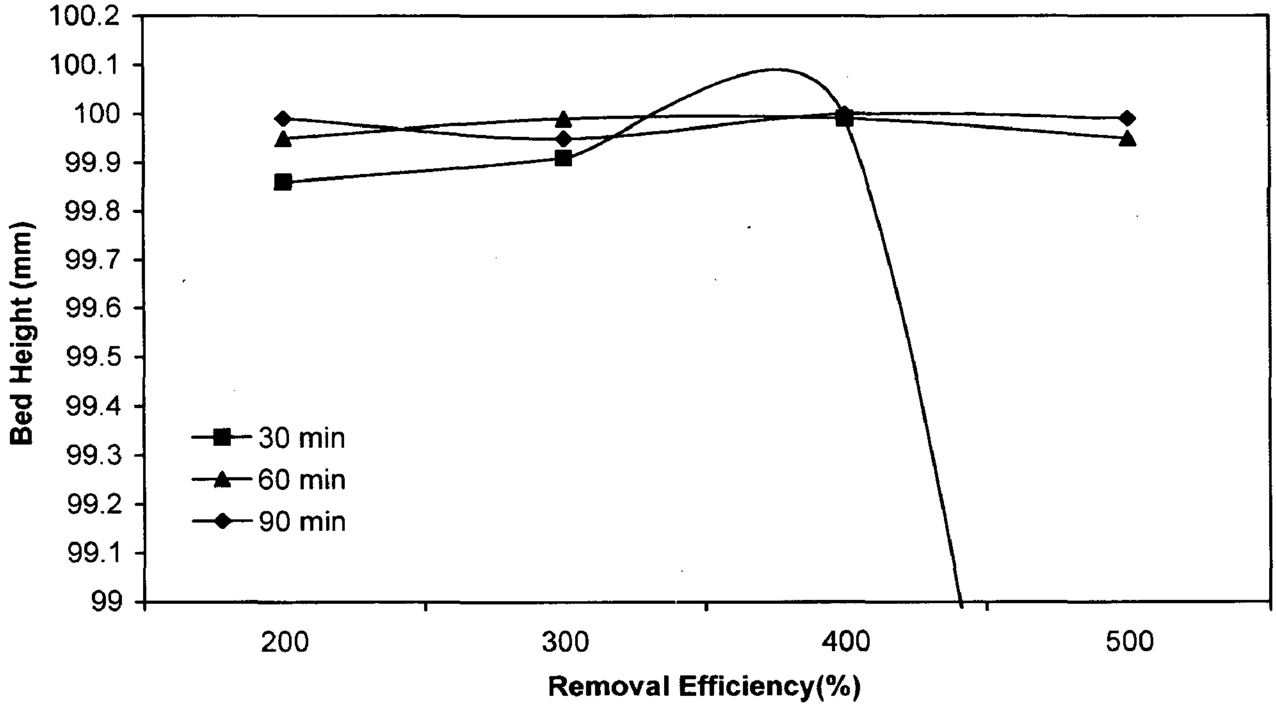

Fig. 5.14 Removal of Fine Suspended Particle at different Bed Height of Calcite and Garnet mixed bed at 60 dm^3/hr flow rate

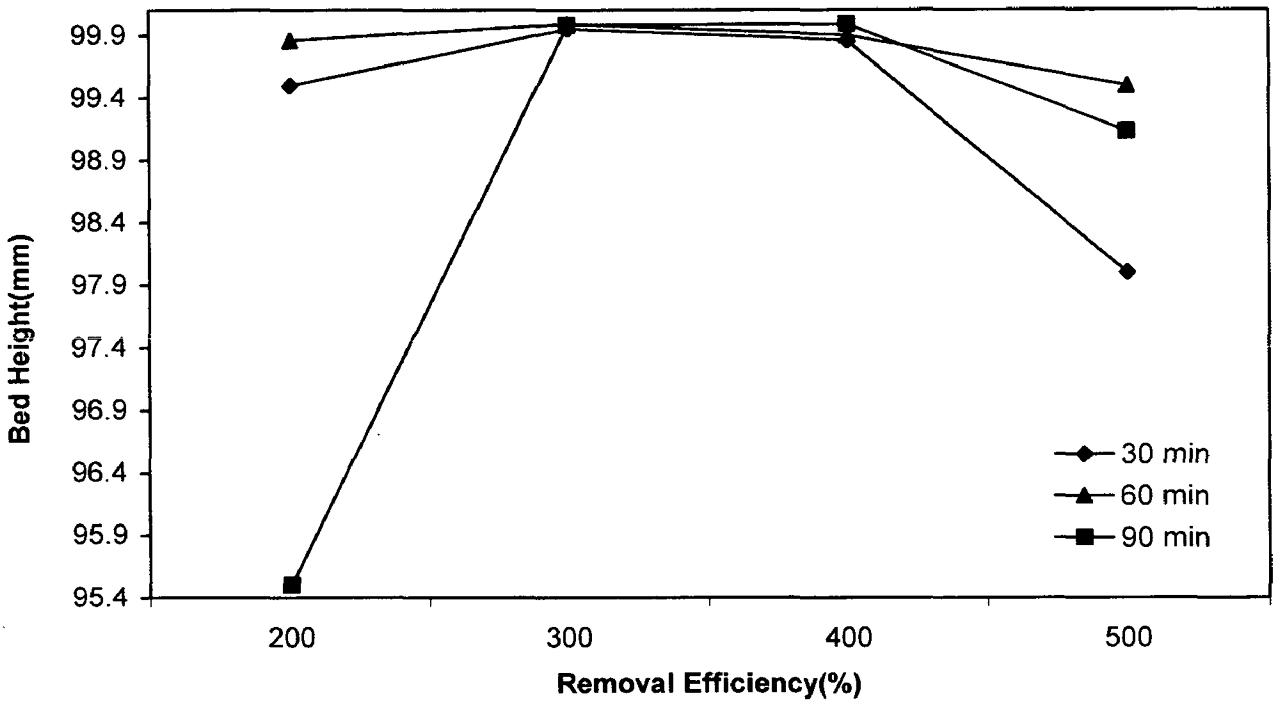

Fig. 5.15 Removal of Fine Suspended Particle at different Bed Height of Calcite and Garnet mixed bed at 90 dm^3/hr flow rate

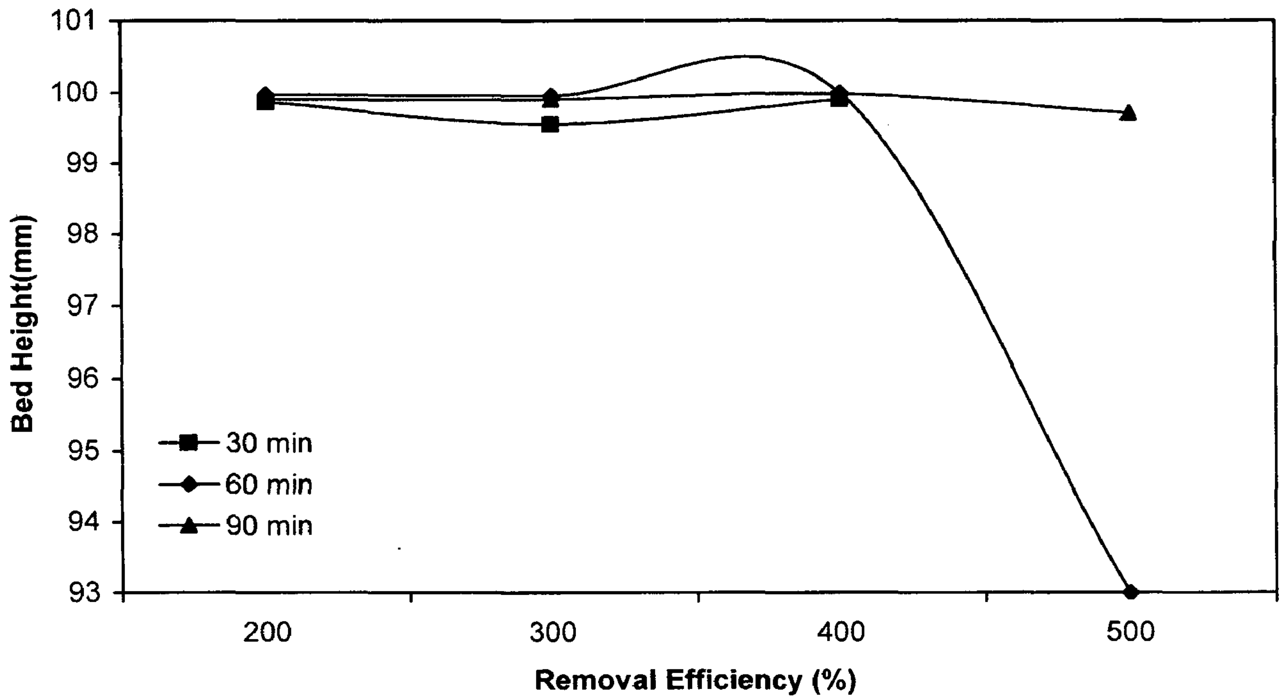

Fig. 5.16 Removal of Fine Suspended Particle at different Bed Height of Calcite and Garnet mixed bed at 120 dm^3/hr flow rate

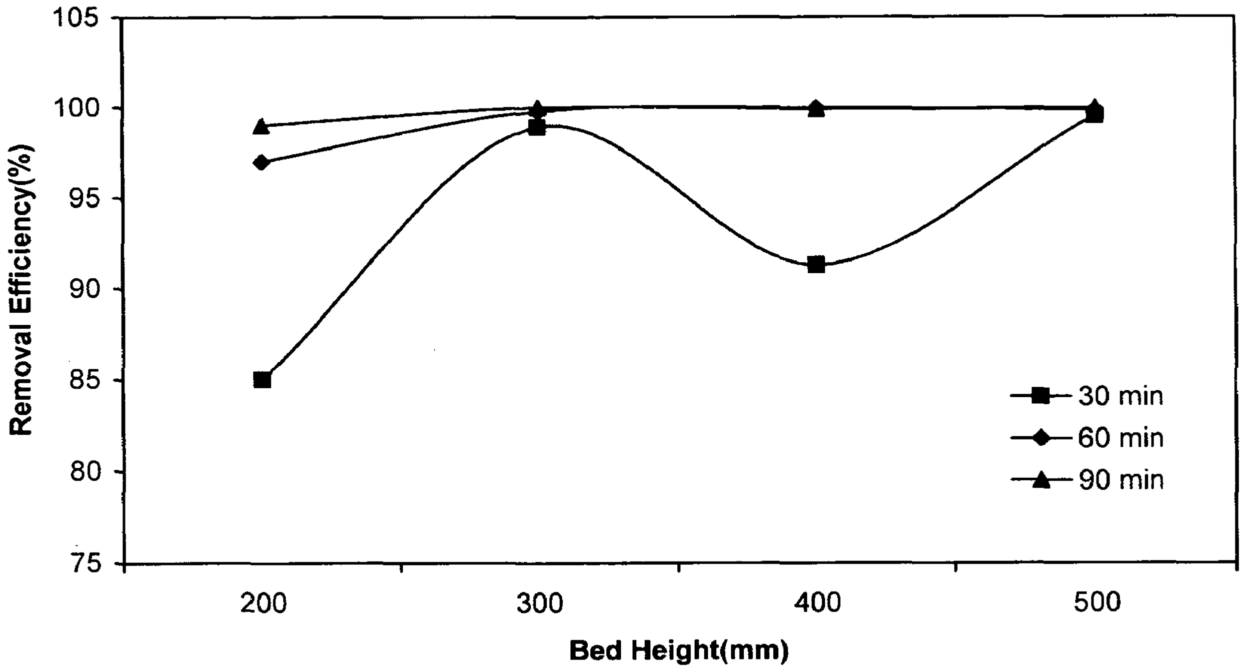

Fig. 5.17 Removal of Fine Suspended Particle at different Bed Height of Anthracite and Garnet at 30 dm^3/hr flow rate

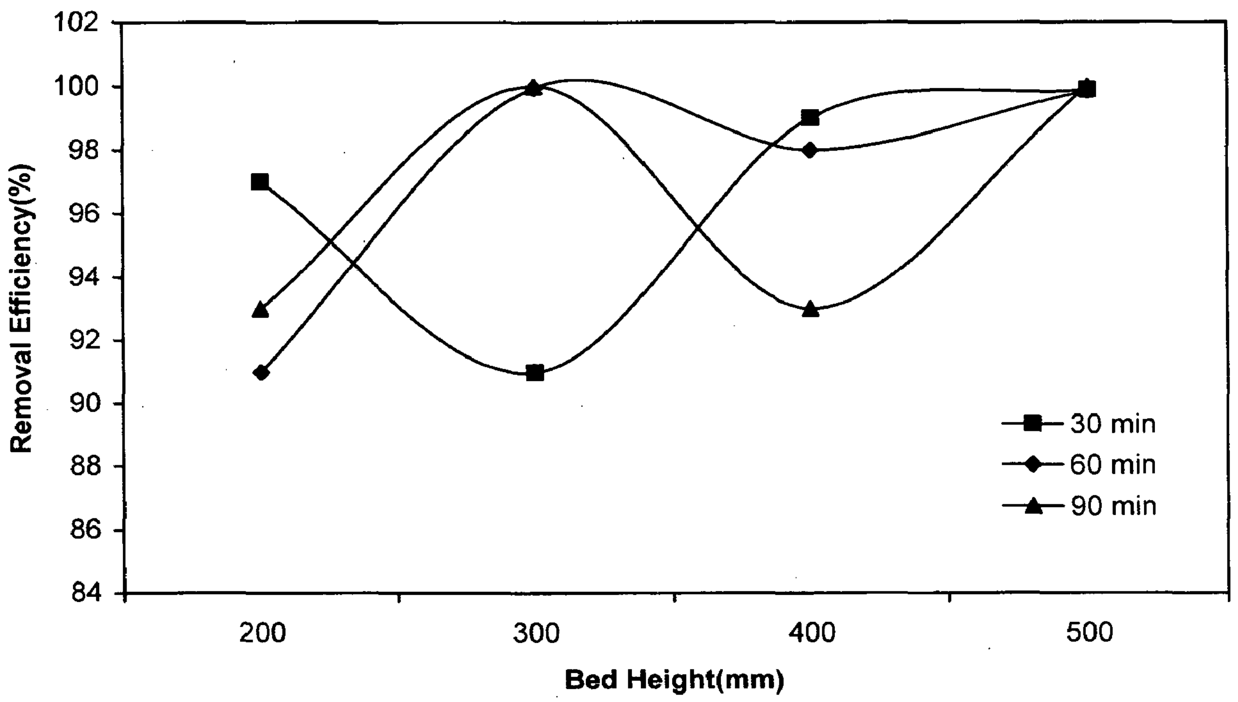

Fig. 5.18 Removal of Fine Suspended Particle at Different Bed Height of Anthracite and Garnet at 60 dm^3/hr

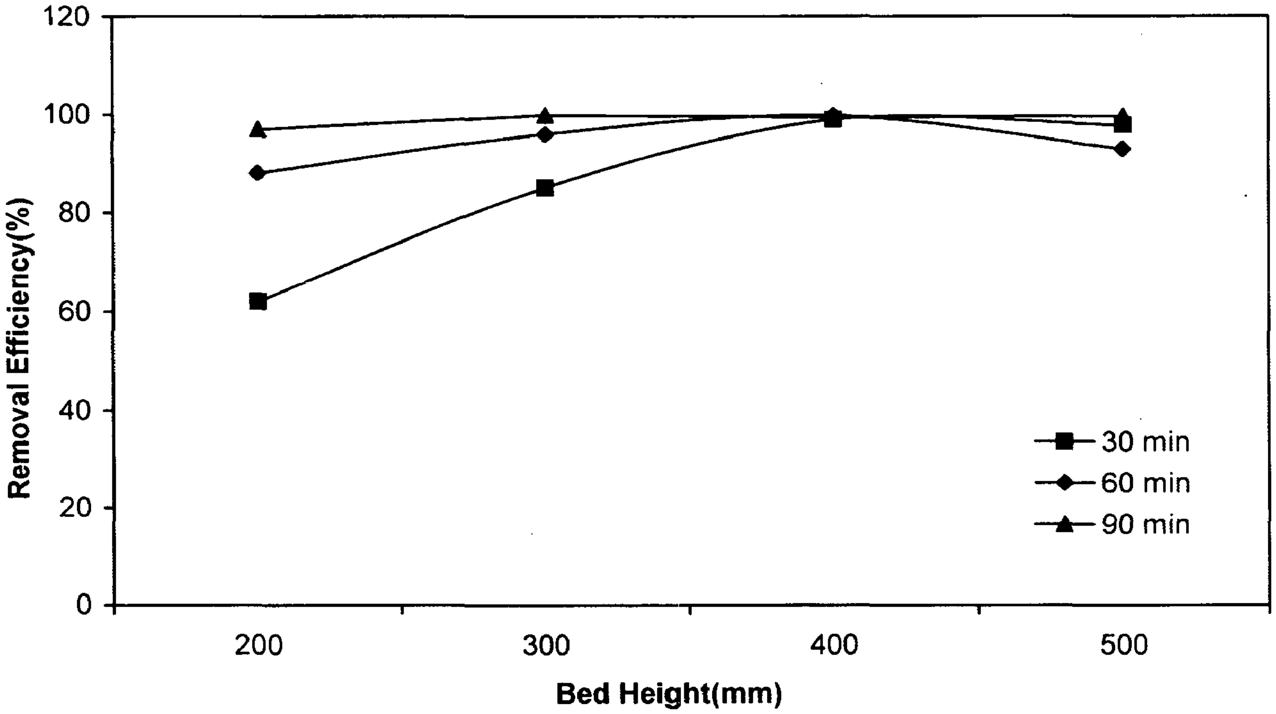

Fig. 5.19 Removal of Fine Suspended Particle at different Bed Height of Anthracite and Garnet at 90 dm^3/hr flow rate

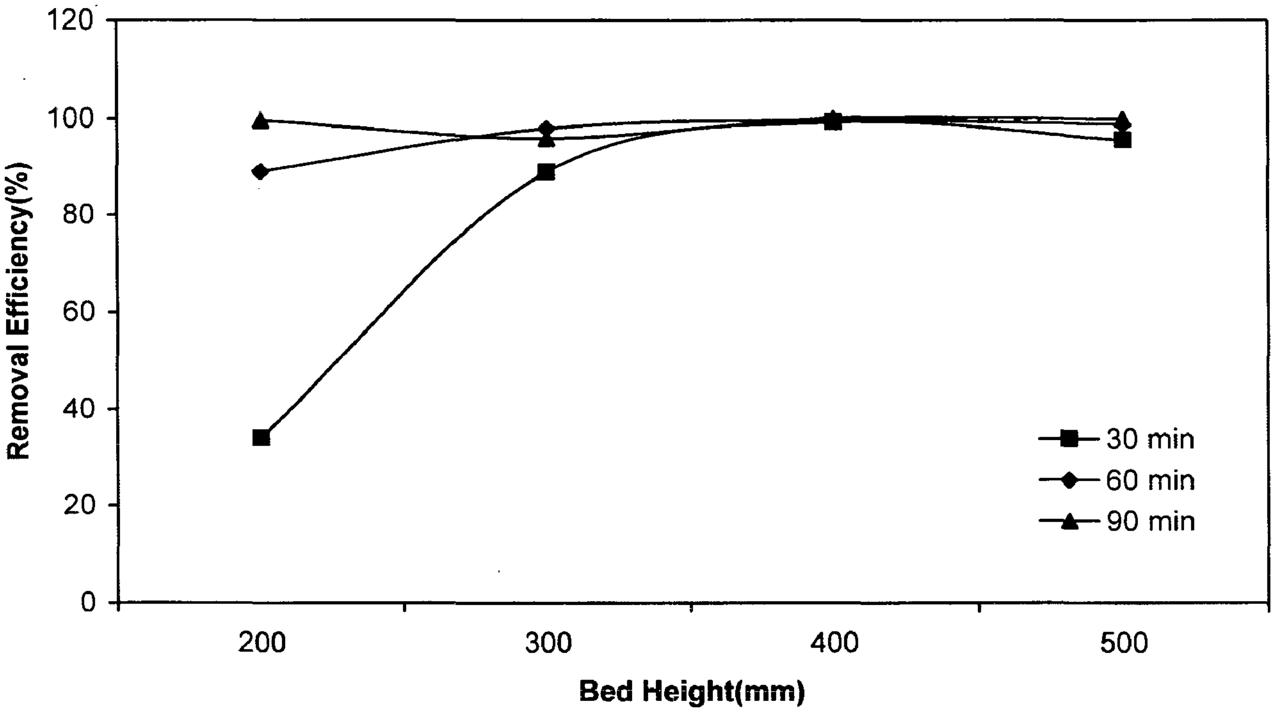

Fig. 5.20 Removal of Fine Suspended Particle at Different Bed Height of Anthracite and Garnet at 120 dm^3/hr

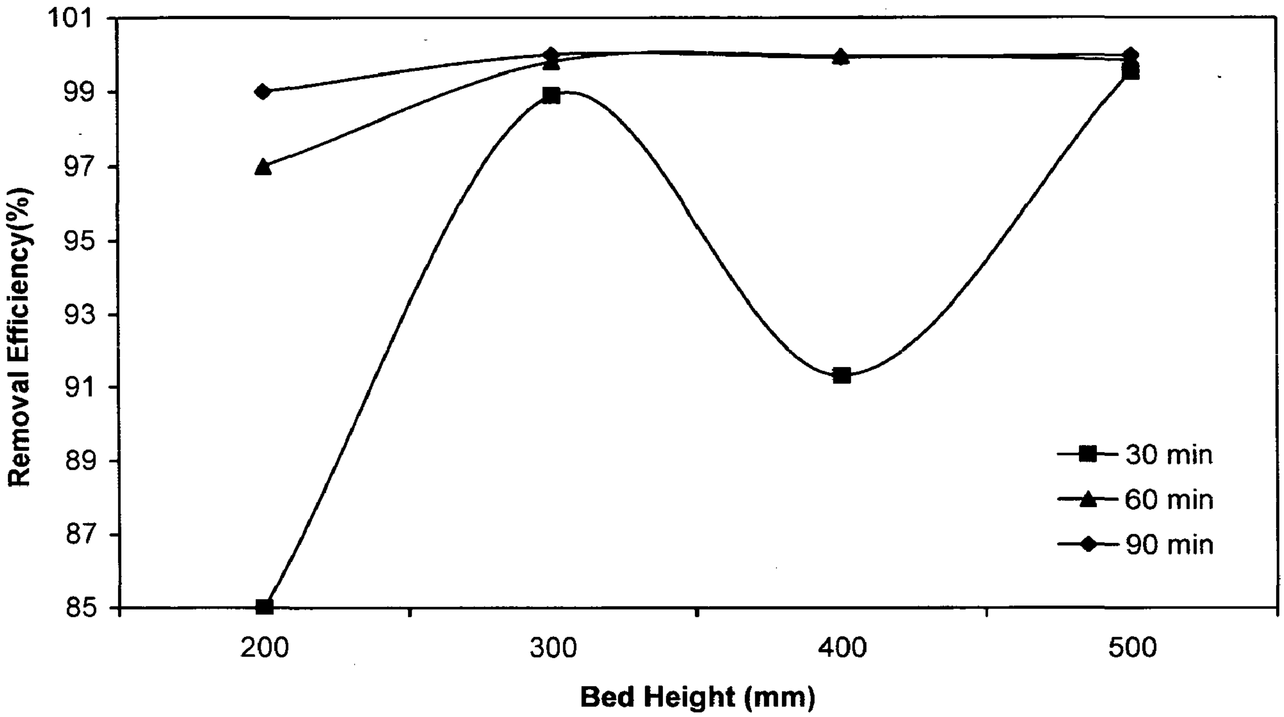

Fig. 5.21 Removal of Fine Suspended Particle at different bed height of Anthracite and Calcite mixed bed at 30 dm^3/hr flow rate

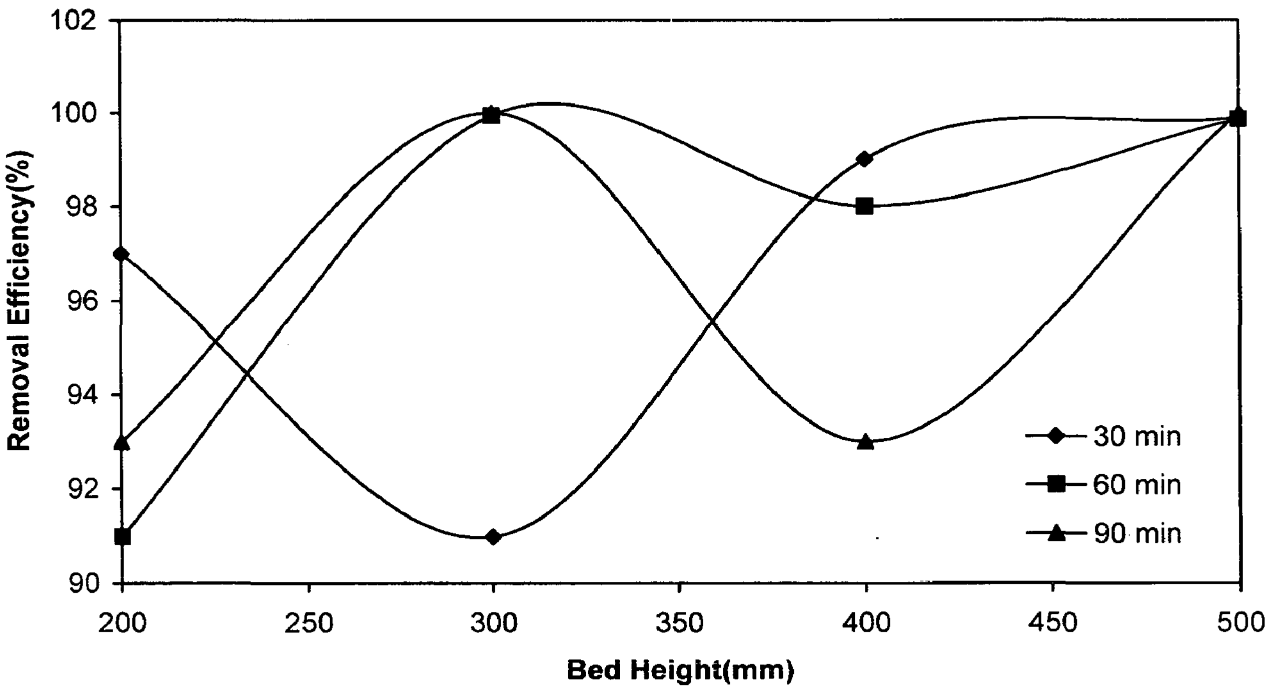

Fig. 5.22 Removal of Fine Suspended Particle at different Bed Height of Anthracite and Calcite mixed bed at 60 dm^3/hr flow rate

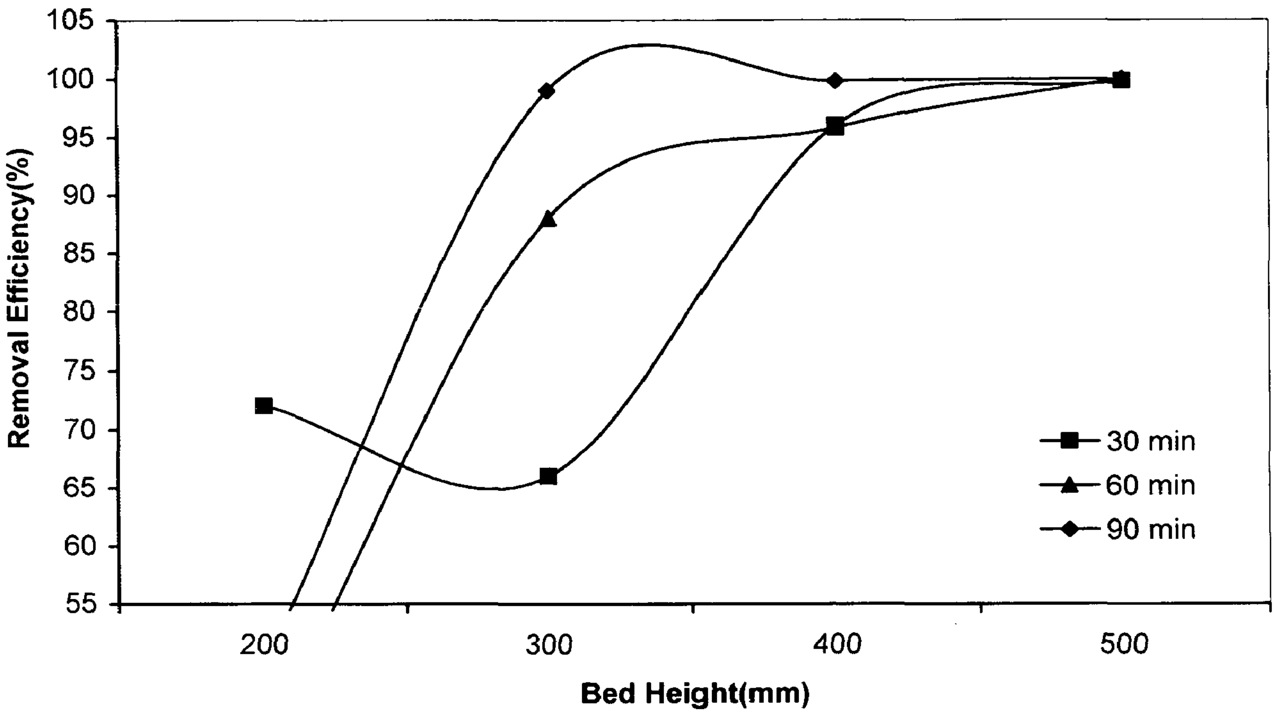

Fig. 5.23 Removal of Fine Suspended Particle at different Bed Height of Calcite and Anthracite at 90 dm^3/hr

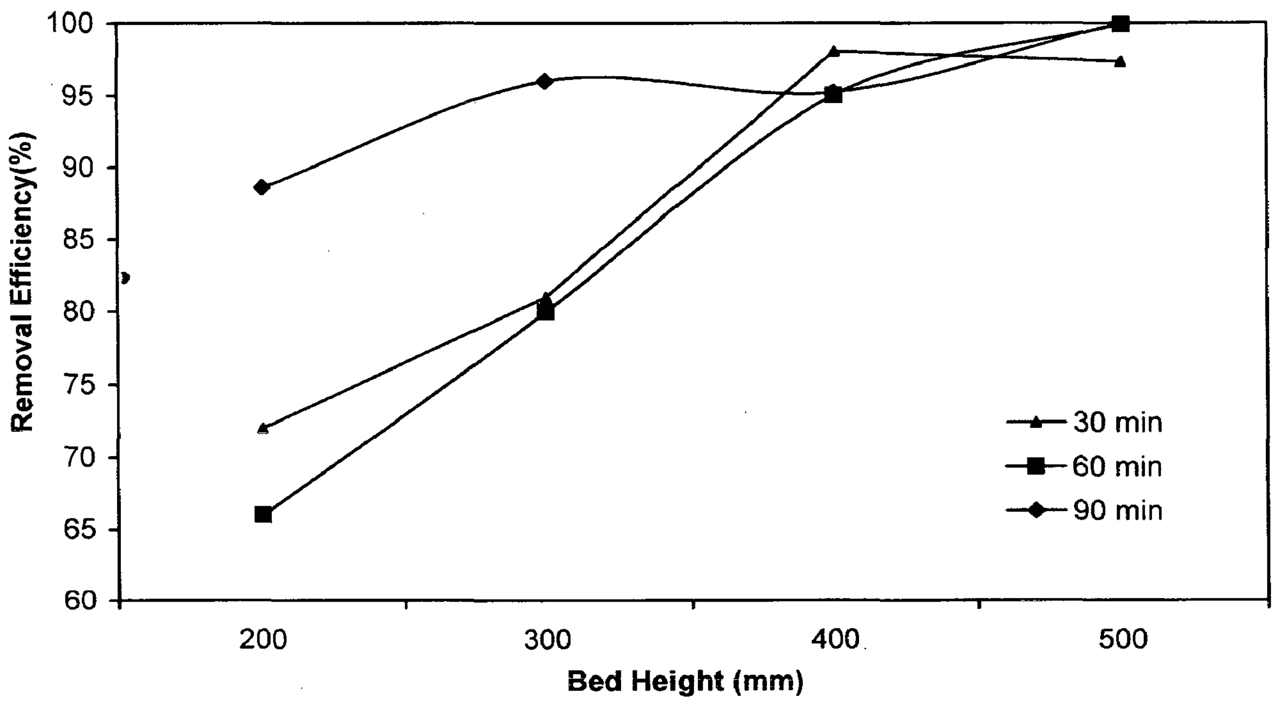

Fig. 5.24 Removal of Fine Suspended Particle at different Bed Height of Calcite and Anthracite at 120 dm^3/hr

5.6 EFFECT OF PRESSURE DROP

In table 1, at a bed height of 200 mm at various flow rates for various media (Calcite-Anthracite, Anthracite-Garnet, and Calcite-Garnet), the effect of pressure drop was studied. For Calcite-Anthracite mixed bed at 60 dm^3/hr at 30 min, pressure drop was 2843 N mm^{-2} which increased to 5810 N mm^{-2} at 60 min which further increased to 6551 N mm^{-2} at 90 min. Similarly, at 120 dm^3/hr, Δ P was 1236 N mm^{-2} at 30 min which increased to 6057 N mm^{-2} at 60 min which further raised to 8529 N mm^{-2} at 90 min.

In case of Anthracite-Garnet, pressure drop was 1236 N mm^{-2} at 30 min at 30 dm^3/hr which increased to 2472 N mm^{-2} at 90 min whereas, at a flow rate of 90 dm^3/hr, it increased from 2472 N mm^{-2} (at 30 min) to 9889 N mm^{-2} (at 90 min).

In case of Garnet-Calcite, it increased from 23486 N mm^{-2} (at 30 min) to 28183(at 90 min) N mm^{-2} at 30 dm^3/hr while at 120 dm^3/hr it increased from 4944 N mm^{-2} (at 30 min) to 53151 N mm^{-2} (at 90 min) N mm^{-2}.

In table 2, the effect of pressure drop at a bed height of 300 mm at various flow rate (30,60,90,120 dm^3/hr) of different media (Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite) was studied. In case of Calcite Anthracite at 30 dm^3/hr, pressure drop was 1236 N mm^{-2} at 30 min which increased to 4449 N mm^{-2} at 90 min. Similarly, at 120 dm^3/hr, pressure drop was 13968 N mm^{-2} at 30 min which increased to 15080 N mm^{-2} at 90 min.

In case of Anthracite-Garnet, pressure drop was 1731 N mm^{-2} at 30 dm^3/hr at 30 min which increased to 2472 N mm^{-2} at 90 min and at 120 dm^3/hr. It was 6180 N mm^{-2} at 30 min which increased to 7416 N mm^{-2} at 90 min.

In case of Garnet-Calcite, pressure drop was 1236 N mm^{-2} at 30 dm^3/hr at 30 min which increased to 3708 N mm^{-2} at 90 min. While at 120 dm^3/hr, pressure drop was 8652 N mm^{-2} at 30 min which increased to 11125 N mm^{-2} at 120 min.

In table 5, the effect of pressure drop at a bed height of 400 mm at various flow rate (30, 60, 90,120 dm^3/hr) on different media was studied. In case of Calcite-Anthracite, pressure drop increased from 5315 N mm^{-2} to 7416 N mm^{-2} at 60 dm^3/hr from 30 min to 90 min while at 120 dm^3/hr it increased from 10754 N mm^{-2} to 12114 N mm^{-2}.

In case of Anthracite-Garnet, pressure drop increased from 3708 N mm^{-2} to 6181 N mm^{-2} from 30 to 90 min at 60 dm^3/hr while at 120 dm^3/hr it increased from 14586 N mm^{-2} to 14091 N mm^{-2}.

In case of Garnet-Calcite, pressure drop increased from 3708 N mm^{-2} to 6180 N mm^{-2} at 60 dm^3/hr from 30 to 90 min while at 120 dm^3/hr it increased from 12361 N mm^{-2} to 13844 N mm^{-2}.

In table 4, the effect of pressure drop at a bed height of 500 mm was studied. In case of Calcite Anthracite, pressure drop increased from 1236 to 7169 at 60 dm^3/hr from 30 min to 90 min while at 120 dm^3/hr it increased from 18035 N mm^{-2} to 18294 N mm^{-2}.

In case of Anthracite- Garnet, at 30 dm^3/hr, pressure drop increased from 6675 N mm^{-2} to 10630 N mm^{-2} from 30 to 90 min.

In case of Garnet-Calcite, pressure drop increased from 5191 N mm^{-2} to 8652 N mm^{-2} at 30 dm^3/hr while at 120 dm^3/hr, it increased from 19539 to 25957 N mm^{-2}.

From the various data, when a comparison is made, it is clear that pressure drop is increasing as the time increases also; it increases along the bed height.

5.7 COMPARISON OF REMOVAL EFFICIENCIES OF VARIOUS MEDIA

A Comparison of Removal Efficiency of fine suspended particle at different bed height of Calcite-Anthracite, Anthracite-Garnet & Garnet-Calcite mixed bed is done. Fig. 5.25, 5.26, 5.27 and 5.28 shows the graphical representation of this comparison after 30 min of run time at flow rates of 30 dm^3 hr^{-1}, 60 dm^3 hr^{-1}, 90 dm^3 hr^{-1} and 120 dm^3 hr^{-1}. It is clear from Fig. 5.25 that the removal efficiency of Calcite-Garnet is higher at all the bed heights as compared to the other two after 30 min at flow rates of 30 dm^3 hr^{-1} . At the flow rate of 60 dm^3 hr^{-1} after 30 min (Fig. 5.26) the removal efficiency of Calcite-Garnet and Anthracite-Garnet are almost equal and higher as compared to Calcite-Anthracite. Similarly in Fig 5.27 and Fig 5.28, the removal efficiency of Calcite-Garnet is found to be more than Anthracite-Garnet which is indeed higher than Anthracite-Calcite.

Fig 5.29, 5.30, 5.31, 5.32 shows the comparison of Removal Efficiency of fine suspended particle at different bed height of Calcite-Anthracite, Anthracite-Garnet and Garnet- ~~Calcite mixed bed after~~ 60 min at different flow rates. Fig 5.29 represents the comparison after 60 min of run time. It is shown that removal efficiency of Calcite-Garnet is highest at the beginning, but then it falls but then again at the bed height of 500 mm it becomes the highest. The removal efficiency of Anthracite-Garnet was low at the beginning but it increases continuously with the increase in bed height. The removal efficiency of Calcite-Anthracite also increases with the bed height but is quite low in comparison with the other two media beds. Fig 5.30 shows that the removal efficiency of Calcite-Garnet is almost 100% at all the bed heights, the removal efficiencies of Anthracite-Garnet and Calcite-Anthracite increases as the bed height increases but not to the extend of Calcite-Garnet. Similar trends are shown by Fig. 5.31 and Fig 5.32.

Fig. 5.33, 5.34, 5.35, 5.36 shows the comparison of Removal Efficiency of fine suspended particle at different bed height of Calcite-Anthracite, Anthracite-Garnet and Garnet-Calcite mixed bed after 90 min at different flow rates: As the bed height is increasing, the removal efficiency of Calcite-Garnet is almost constant (near 100%) while the removal efficiency of Calcite-Anthracite is increasing continuously (Fig. 5.34).

Thus by observing the pattern followed by various materials, it can be concluded that the removal efficiency of Calcite-Garnet is higher than Anthracite-Garnet and

Calcite-Anthracite. The removal efficiency of Calcite-Anthracite being the lowest among the three media bed.

5.8 DISCUSSIONS

As discussed in previous sections, the removal efficiency increases with the increase in bed height. This may be because at a higher filter depth a greater detention time (i.e. the time taken for the influent to move from the inlet point, passing the column packed material and going out to outlet) is offered by the higher filter depth.

Also, as the flow rate increases, the removal efficiency decreases. This may be due to the fact that sufficient time for filtration is not provided. Also, the pressure drop increases with increase in flow rate. Thus, it can be concluded that low flow rates results in better removal efficiencies with less pressure drops.

Also, as the time interval increases, the removal efficiency improves. But it is seen that it is almost same at intervals of 60 and 90 min i.e. after certain duration of time, removal efficiency do not increase with time. This is the time when pores of the filter media are packed with suspended particle and it needs to be cleaned. So, backwashing of filters is needed at this time. Also, although the efficiency remains same, pressure drop continuously increases. Thus, for the same efficiency, 60 min duration is quite better from the economical aspects.

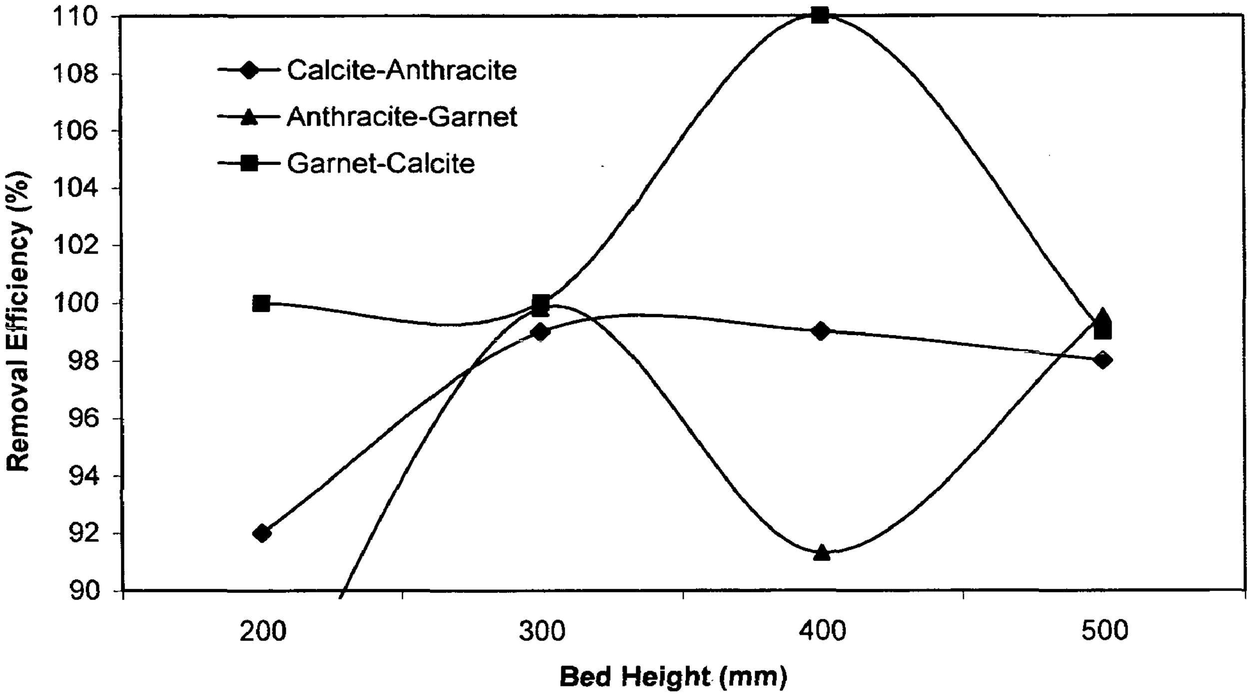

Fig. 5.25 Comparison of Removal Efficiency of fine suspended particle at different bed height of Calcite-Anthracite, Anthracite-Garnet & Garnet-Calcite mixed bed after 30 min at 30 dm3/hr

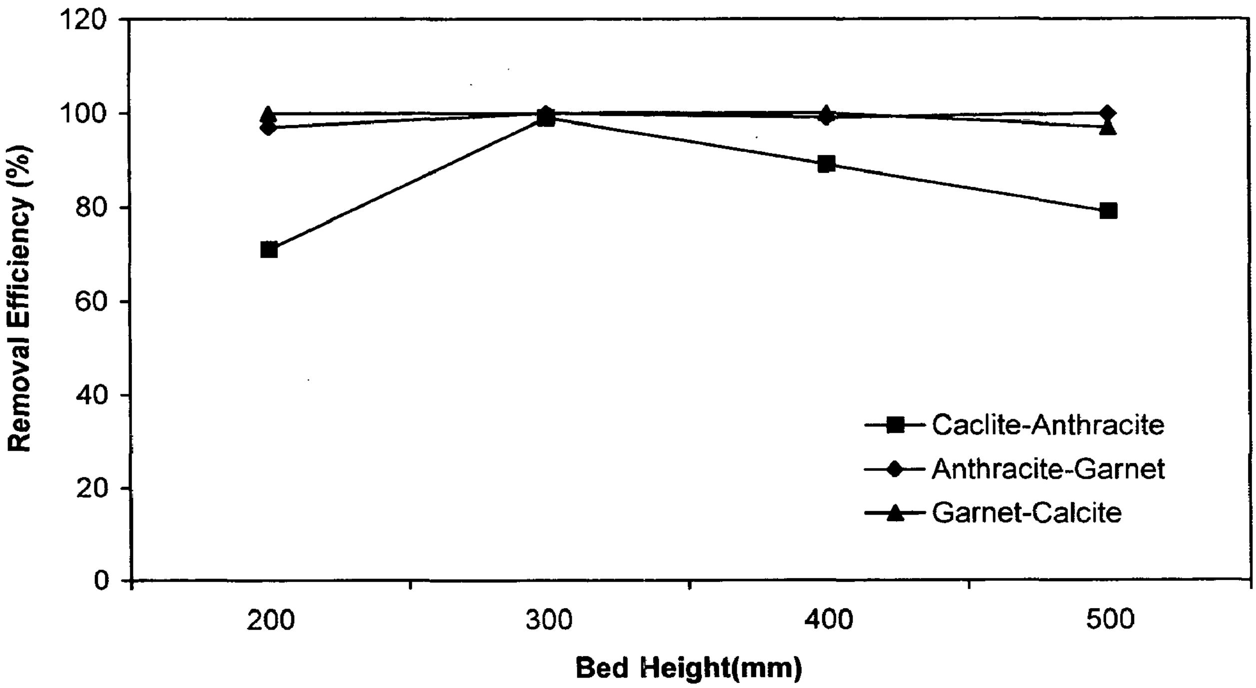

Fig. 5.26 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 30 min at 60 dm^3/hr

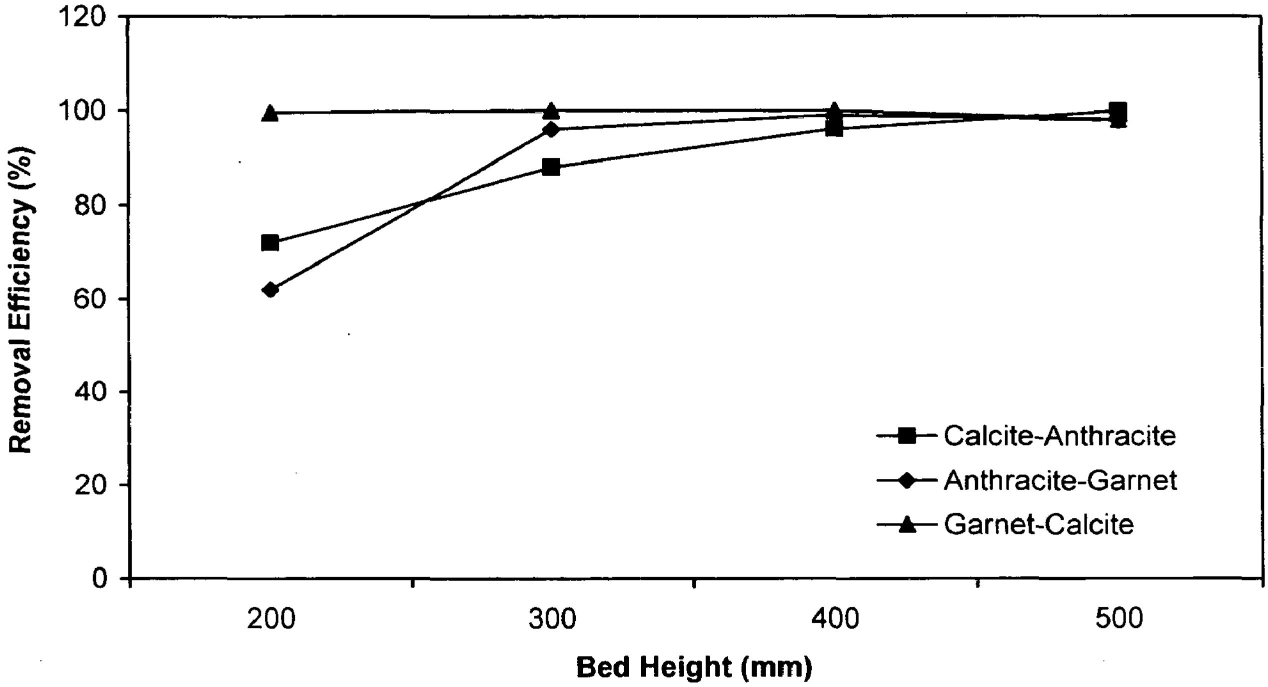

Fig. 5.27 Comparison of Removal of fine suspended particle at different bed height of Caclite-Anthracite,Anthracite-Garnet and Garnet-Calcite mixed bed after 30 min at 90 dm^3/hr

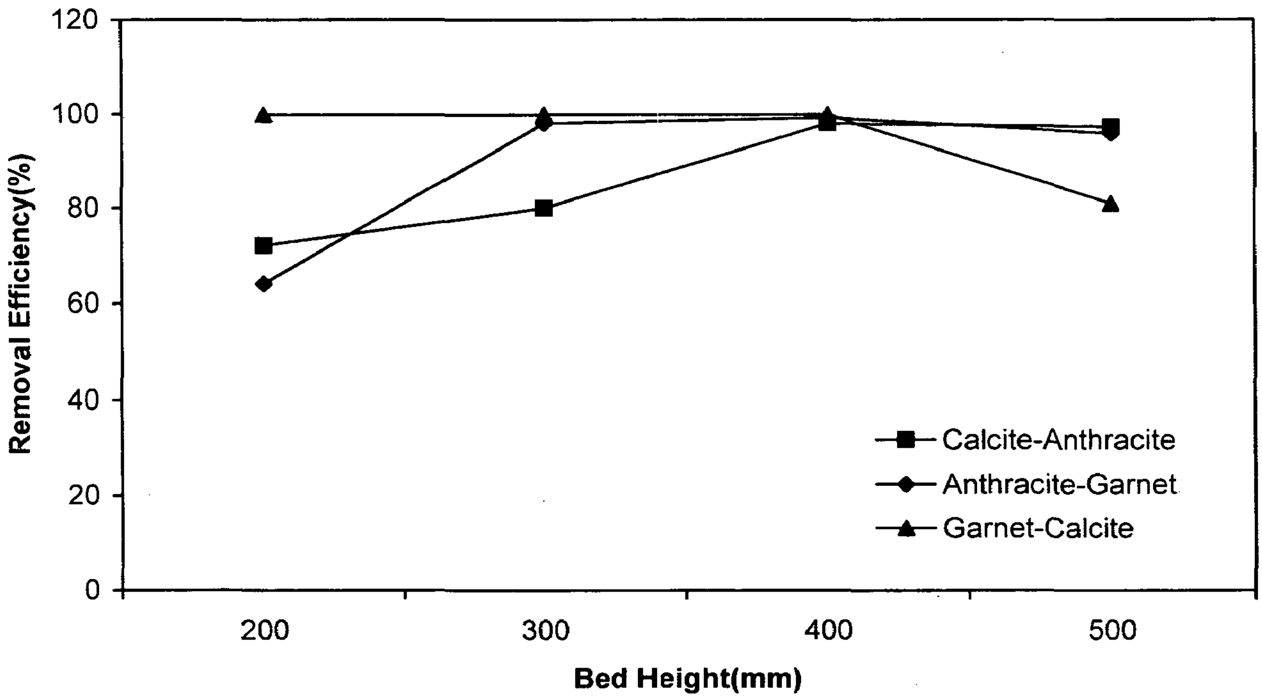

Fig. 5.28 Comparison of Removal of fine suspended particle at different bed height of Caclite-Anthracite,Anthracite-Garnet and Garnet-Calcite mixed bed after 30 min at 120 dm^3/hr

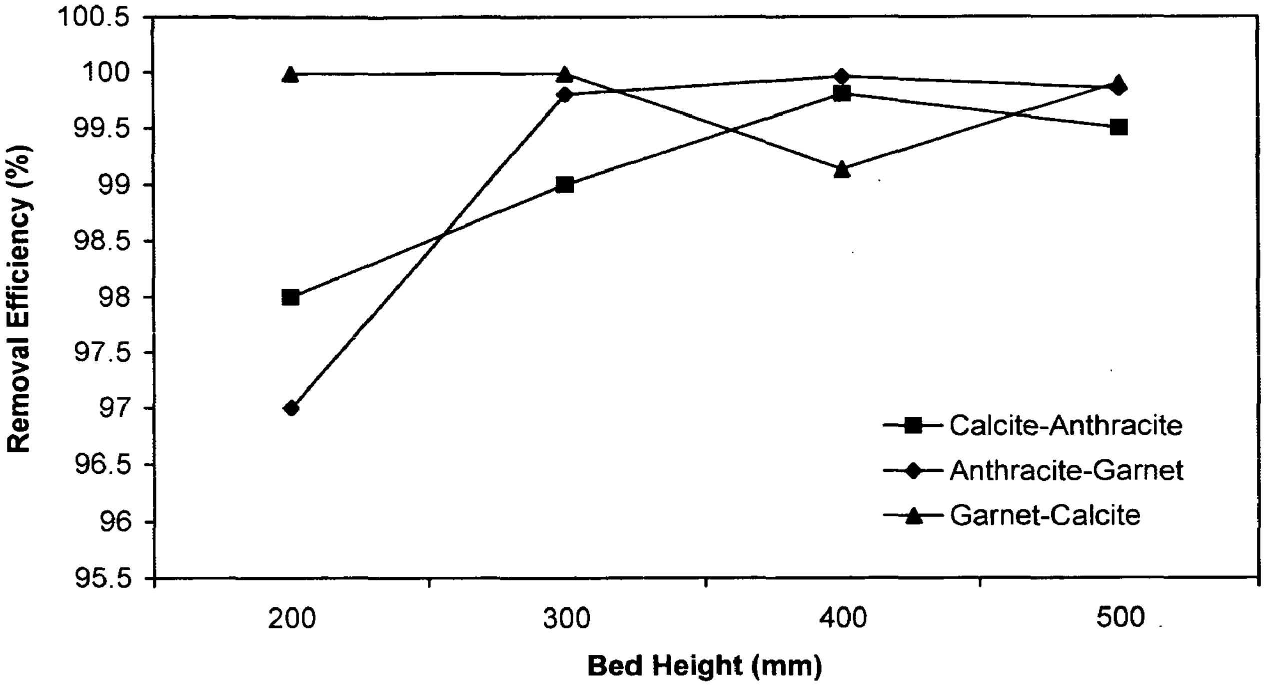

Fig. 5.29 Comparison of Removal of fine suspended particle at different bed height after 60 min at 30 dm^3/hr

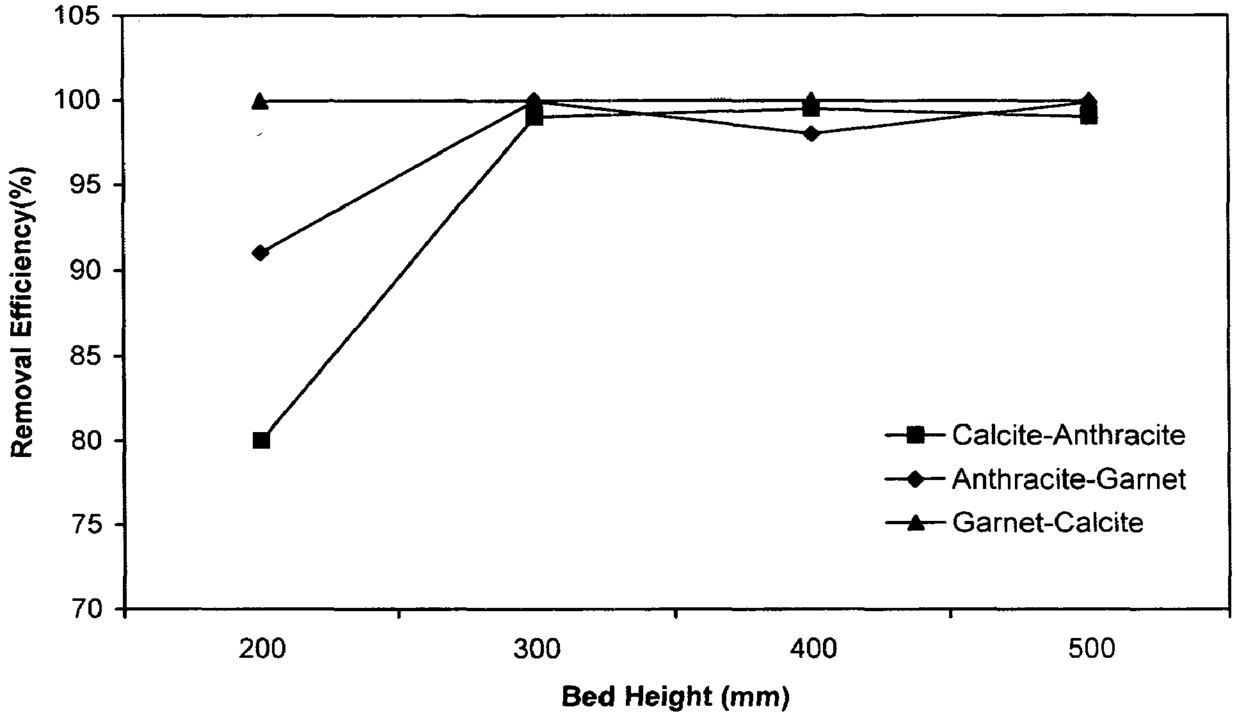

Fig. 5.30 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 60 min at 60 dm3/hr

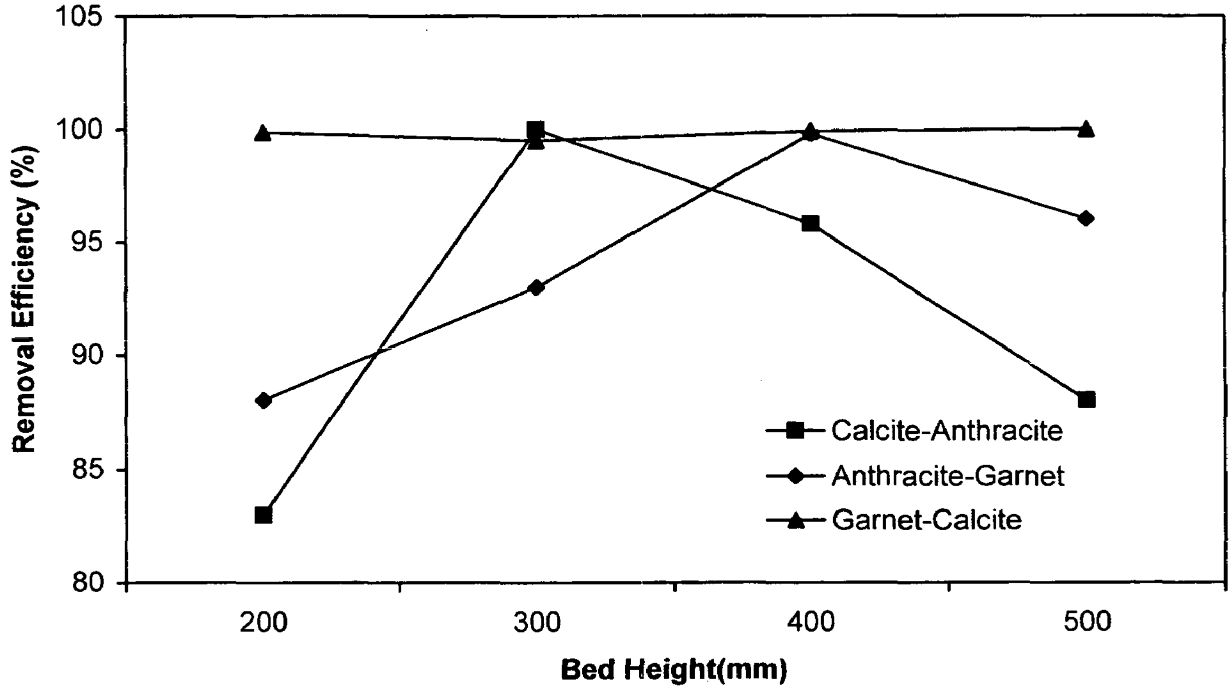

Fig. 5.31 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet and Garnet-Calcite mixed bed after 60 min at 90 dm^3/hr

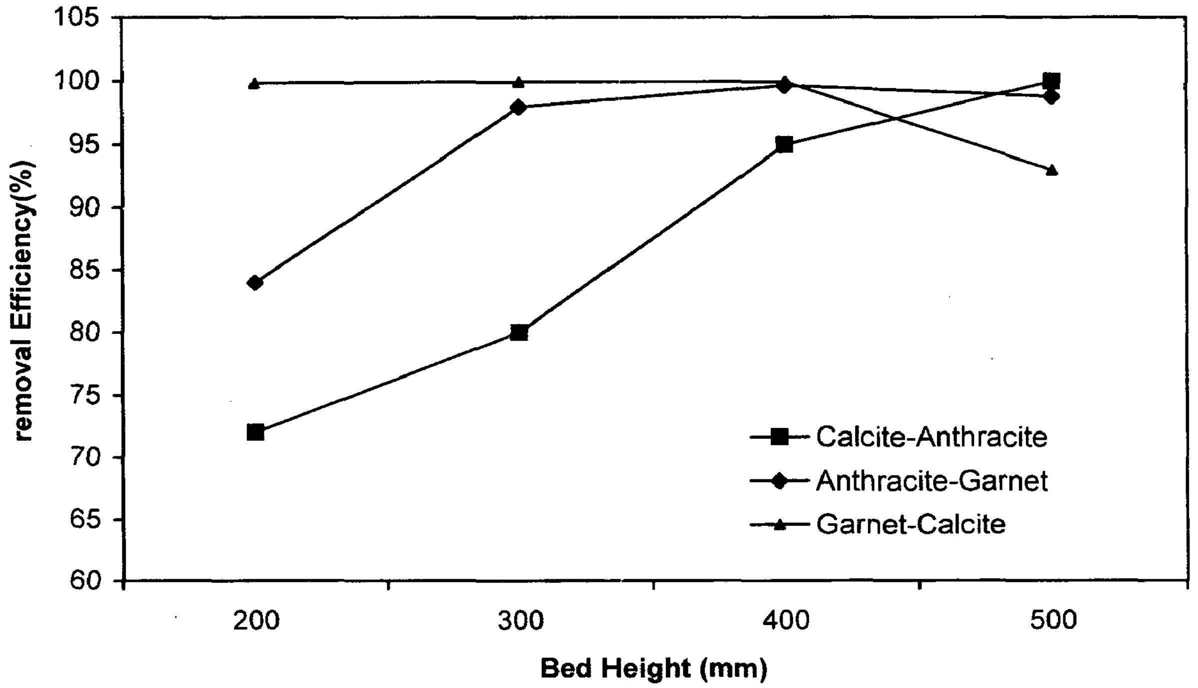

Fig. 5.32 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet and Garnet-Calcite mixed bed after 60 min at 120 dm^3/hr

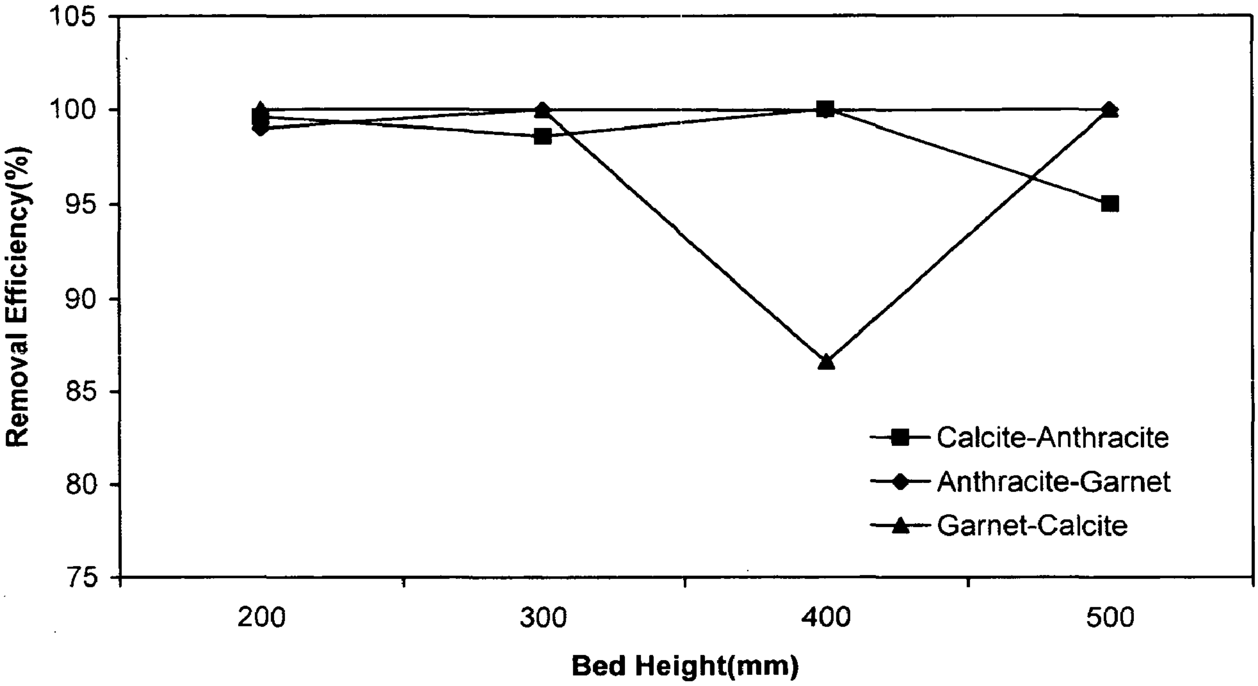

Fig. 5.33 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 90 min at 30 dm^3/hr

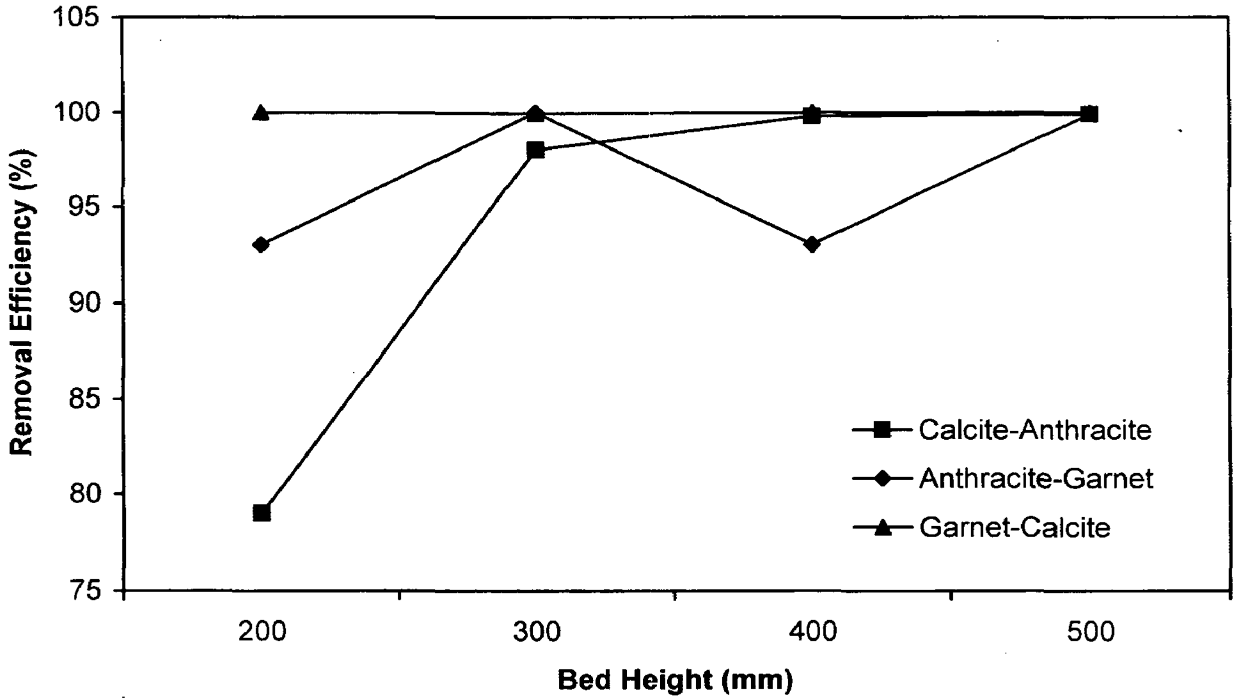

Fig. 5.34 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 90 min at 60 dm^3/hr

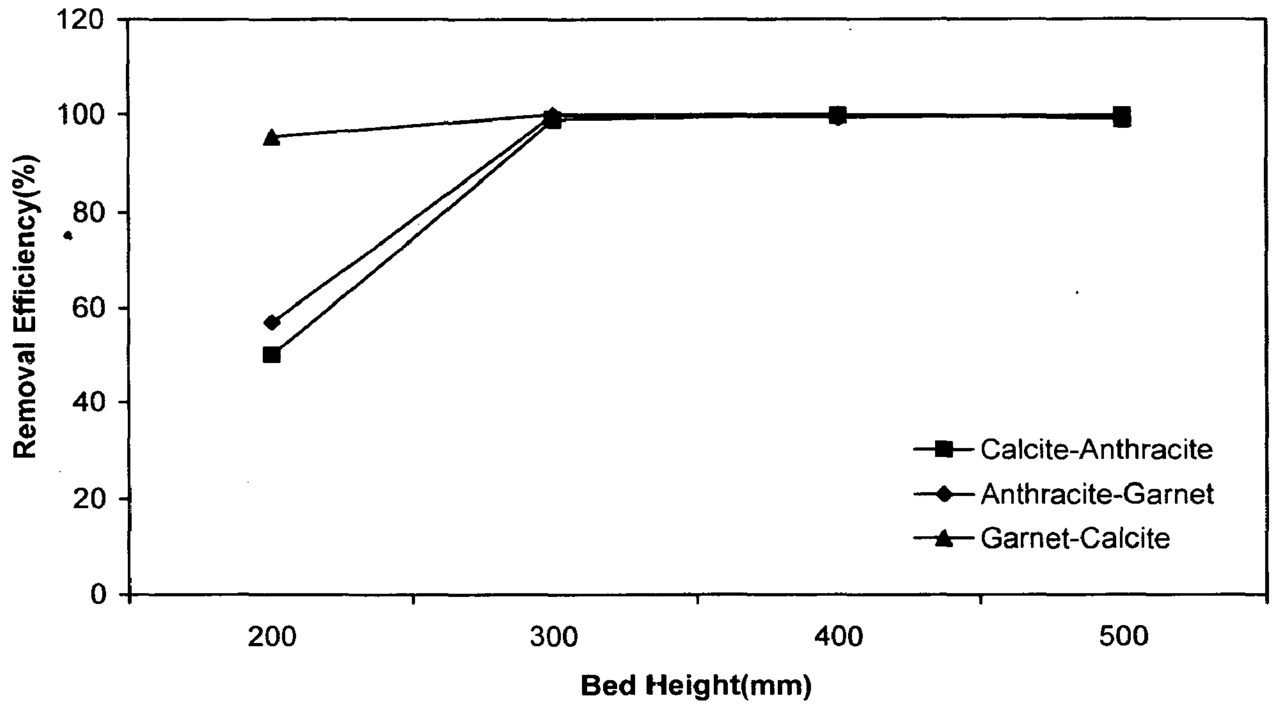

Fig. 5.35 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 90 min at 90 dm^3/hr

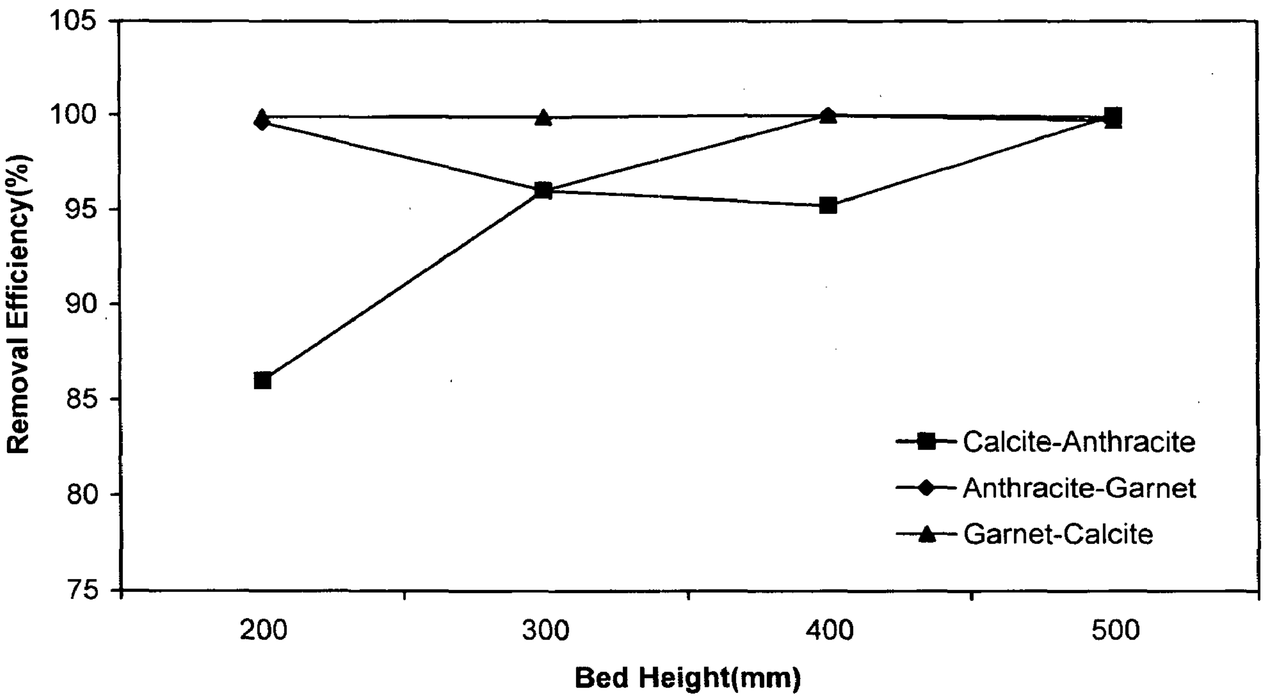

Fig. 5.36 Comparison of Removal of fine suspended particle at different bed height of Calcite-Anthracite,Anthracite-Garnet,Garnet-Calcite mixed bed after 90 min at 120 dm^3/hr

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 CONCLUSIONS

Under the operating conditions used in the experiment and from the result obtained, the following conclusions can be made:

1. During the filtration the media type and size, cycle time, and bed height of media in filter column had significant effect on removing of 2-20 μm suspended particle.
2. During the filtration operation high cycle time result in low removal efficiency and a high level of suspended particle in the filtration during filtration.
3. A good removal of suspended particle was also obtained when the bed height increases i.e. as the bed height increases from 200 mm to 500 mm the removal efficiency increases.
4. As the bed height is increasing, Pressure Drop is increasing. Also, it increases along the flow rate. A deep and detailed study of the results show that low flow rate and highest bed height gives the best optimum results.
5. From the observed result the removal efficiency for Calcite-Garnet is higher than Anthracite-Garnet & Calcite-Anthracite in term of removal of 2-20 μm particles.

6.2 RECOMMENDATIONS

The scope of the present study is limited to the study of removal efficiency of fine suspended particle using multimedia filter bed for the removal of fine suspended particle. Studies were done on the designed flow rate and allowed pressure drop. Further studies can be done using multimedia filter beds in various combinations as well as using the combination of more than two media. Also, changes can be made in the filter design to allow larger flow rates and pressure drops. Also, mathematical models can be prepared and simulation can be done to know the exact flow rate and pressure range.

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APPENDIX-A

Table A1: Undersize Number Percentage (%) for Calcite-Anthracite

For 200 mm bed height											
For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.5	10	0.36	10	0.37	10	0.45	10	0.45	10	0.45
40	0.99	55	0.99	50	0.95	45	1.0	40	0.93	40	0.94
50	1.37	70	1.4	65	1.36	55	1.34	55	1.36	55	1.39
60	1.97	80	1.93	75	1.94	65	1.90	65	1.87	65	1.96
65	2.27	85	2.44	80	2.45	70	2.22	70	2.2	70	2.28
75	2.87	90	3.0	85	2.89	80	2.87	80	2.87	80	2.92
85	3.49	95	3.84	90	3.51	85	3.32	85	3.29	85	3.39
90	3.91	97	4.43	95	4.73	90	3.89	90	3.78	90	3.86
95	4.92	99	7.32	97	6.31	95	4.78	95	4.71	95	4.73
97	5.89	100	60.5	99	11.58	97	6.19	97	5.96	97	5.88
99	14.06			100	54.50	99	11.35	99	9.42	99	10.71
100	72.0					100	35.00	100	29.0	100	29.00
For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.38	15	0.40	20	0.49	10	0.44	10	0.42	15	0.48
80	0.97	75	0.97	85	1.0	45	0.98	45	0.95	60	0.96
90	1.48	85	1.44	90	1.42	55	1.33	65	1.44	75	1.39
95	2.67	90	2.26	95	2.62	65	1.84	70	1.71	80	1.64
97	3.33	95	3.34	97	3.24	75	2.48	80	2.37	85	2.06
99	4.87	97	3.92	99	4.4	80	2.80	85	2.74	90	2.67
100	30.0	99	6.31	100	27.0	85	3.18	90	3.21	95	3.45
		100	28.5			90	3.65	95	3.97	97	3.87
						95	4.55	97	4.69	99	5.75
						97	5.68	99	7.09	100	24.5
						99	10.03	100	22.0		
						100	34.0				

Table A2: Undersize Number Percentage (%) for Calcite-Anthracite

For 300 mm bed height											
For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.35	15	0.46	15	0.40	15	0.38	15	0.38	15	0.38
55	0.97	65	0.99	75	0.97	80	0.97	80	0.97	80	0.98
70	1.41	80	1.49	85	1.44	85	1.14	90	1.45	85	1.19
75	1.72	85	1.97	90	2.29	90	1.58	95	2.73	90	1.61
80	2.22	90	2.64	95	3.56	95	2.87	97	3.44	95	2.80
85	2.74	95	3.7	97	4.23	97	3.67	99	5.35	97	3.49
90	3.41	97	4.32	99	6.67	99	6.38	100	24.5	99	4.67
95	4.48	99	7.97	100	27.0	100	30.5			100	29.0
97	5.79	100	34.5								
99	10.94										
100	45.0										
For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.38	15	0.41	20	0.5	10	0.38	10	0.34	10	0.39
80	0.98	75	0.98	80	0.96	50	0.95	55	0.95	50	0.97
90	1.46	85	1.34	90	1.32	70	1.49	75	1.46	70	1.43
95	2.63	90	1.64	95	1.92	75	1.85	80	1.83	80	1.90
97	3.29	95	2.47	97	2.68	80	2.24	85	2.32	85	2.28
99	4.75	97	3.08	99	3.96	85	2.64	90	3.01	90	2.81
100	26.5	99	4.31	100	25.0	90	3.06	95	3.82	95	3.47
		100	23.5			95	3.76	97	4.57	97	4.15
						97	4.35	99	8.21	99	5.45
						99	6.26	100	23.0	100	17.0
						100	14.0				

Table A3: Undersize Number Percentage (%) for Calcite-Anthracite

For 400 mm bed height											
For 30 $dm^3\ hr^{-1}$						For 60 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.4	5	0.26	15	0.39	10	0.42	10	0.34	15	0.5
50	0.99	35	0.94	75	0.96	45	0.94	55	0.95	60	0.99
65	1.41	55	1.39	85	1.28	60	1.35	75	1.50	70	1.36
75	1.93	65	1.77	90	1.76	70	1.83	80	2.09	75	1.66
80	2.31	75	2.33	95	3.12	75	2.15	85	2.54	80	2.27
85	2.77	85	2.97	97	3.62	85	2.94	90	3.22	85	2.84
90	3.36	90	3.42	99	4.80	90	3.37	95	3.95	90	3.39
95	4.10	95	4.17	100	26.5	95	4.05	97	4.80	95	4.05
97	5.02	97	4.98			97	4.93	99	7.13	97	5.31
99	18.36	99	6.99			99	8.11	100	26.5	99	8.93
100	64.0	100	35.0			100	58.0			100	39.5
For 90 $dm^3\ hr^{-1}$						For 120 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.49	15	0.42	15	0.39	10	0.37	10	0.39	10	0.35
60	0.96	75	1.0	75	0.96	50	0.94	50	0.97	55	0.98
75	1.40	80	1.28	85	1.33	70	1.44	65	1.41	70	1.43
80	1.68	85	1.69	90	1.90	75	1.82	75	1.97	75	1.78
85	2.15	90	2.57	95	3.02	80	2.27	80	2.31	80	2.25
90	2.79	95	3.41	97	3.68	85	2.71	85	2.70	85	2.72
95	3.62	97	4.03	99	5.46	90	3.23	90	3.17	90	3.33
97	4.24	99	6.18	100	16.5	95	4.10	95	3.86	95	4.04
99	6.19	100	39.5			97	5.28	97	4.32	97	4.76
100	28.5					99	8.54	99	6.76	99	7.93
						100	17.5	100	18.5	100	25.0

Table A4: Undersize Number Percentage (%) for Calcite-Anthracite

For 500 mm bed height											
For 30 $dm^3\ hr^{-1}$						For 60 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
20	0.5	15	0.45	15	0.38	5	0.25	10	0.39	15	0.46
80	0.96	65	0.96	80	0.98	35	0.92	50	0.97	65	0.99
90	1.32	80	1.36	90	1.48	60	1.46	65	1.39	80	1.42
95	2.24	90	1.98	95	2.58	65	1.70	75	1.93	85	1.80
97	3.22	95	2.82	97	3.29	75	2.42	80	2.33	90	2.62
99	4.21	97	3.40	99	4.49	80	2.83	85	2.76	95	3.58
100	26.5	99	4.22	100	24.0	85	3.30	90	3.27	97	4.33
		100	6.00			90	3.91	95	3.9	99	7.63
						95	4.99	97	4.55	100	30.00
						97	6.42	99	8.26		
						99	9.81	100	38.5		
						100	20.50				
For 90 $dm^3\ hr^{-1}$						For 120 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.39	15	0.39	5	0.34	10	0.37	5	0.42	10	0.38
80	1.0	80	0.99	25	0.90	50	0.94	20	0.89	50	0.95
85	1.35	85	1.27	55	1.47	65	1.40	45	1.50	70	1.47
90	2.15	90	1.95	65	1.90	75	1.98	55	1.85	75	1.77
95	3.14	95	3.32	75	2.39	80	2.44	70	2.42	80	2.13
97	3.92	97	3.89	80	2.90	85	2.93	75	2.72	85	2.53
99	7.94	99	8.15	85	3.29	90	3.5	85	3.41	90	3.12
100	71.0	100	12.50	90	3.73	95	4.29	90	3.82	95	3.74
				95	4.95	97	5.0	95	4.84	97	4.07
				97	5.29	99	9.83	97	6.37	99	5.92
				99	5.84	100	43.0	99	9.13	100	10.0
				100	7.50			100	22.5		

APPENDIX-B

Table B1: Undersize Number Percentage (%) for Anthracite-Garnet

For 200 mm bed height											
For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.38	10	0.34	15	0.44	10	0.40	10	0.48	10	0.37
80	0.98	55	0.95	70	0.99	50	0.99	40	0.96	50	0.94
90	1.38	80	1.47	80	1.40	65	1.39	60	1.41	70	1.46
95	2.11	85	1.85	85	1.78	75	1.85	70	1.85	75	1.78
97	2.76	90	2.55	90	2.35	80	2.22	75	2.25	80	2.18
99	4.08	95	3.39	95	3.24	85	2.70	85	2.96	85	2.63
100	56.0	97	4.08	97	3.85	90	3.21	90	3.43	90	3.26
		99	6.67	99	6.24	95	3.86	95	4.25	95	4.02
		100	25.0	100	28.50	97	4.31	97	4.9	97	4.68
						99	6.45	99	7.89	99	8.69
						100	16.50	100	17.5	100	39.0
For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.40	15	0.44	15	0.40	15	0.40	15	0.41	15	0.39
75	0.96	70	1.0	75	0.97	75	0.97	75	0.99	80	1.00
85	1.28	85	1.48	90	1.48	85	1.33	85	1.46	90	1.45
90	1.57	90	1.98	95	2.37	90	1.73	90	2.05	95	2.35
95	2.47	95	2.86	97	3.00	95	2.66	95	3.25	97	2.94
97	3.02	97	3.39	99	4.46	97	3.26	97	3.84	99	4.14
99	4.33	99	4.45	100	26.00	99	4.48	99	5.57	100	15.0
100	17.0	100	14.50			100	31.0	100	45.0		

Table B2: Undersize Number Percentage (%) for Anthracite-Garnet

For 300 mm bed height											
For 30 dm^3 hr^{-1}						For 60 dm^3 hr^{-1}					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.42	15	0.39	20	0.48	15	0.39	15	0.41	20	0.47
70	0.97	80	0.99	85	0.98	80	1.00	75	0.99	85	0.96
85	1.41	90	1.45	90	1.18	85	1.30	85	1.36	95	1.41
90	1.85	95	2.35	95	1.70	90	1.88	90	1.75	97	1.80
95	2.85	97	3.00	97	2.25	95	2.99	95	2.79	99	2.47
97	3.57	99	3.91	99	3.01	97	3.74	97	3.42	100	3.50
99	4.50	100	5.50	100	3.50	99	7.94	99	5.27		
100	6.50					100	52.0	100	17.0		
For 90 dm^3 hr^{-1}						For 120 dm^3 hr^{-1}					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.34	10	0.38	10	0.45	15	0.39	20	0.48	20	0.50
55	0.95	50	0.97	40	0.93	80	0.99	85	0.98	80	0.96
70	1.35	75	1.46	75	1.45	85	1.21	95	1.49	90	1.37
~~80~~	~~1.99~~	~~80~~	~~1.77~~	85	1.95	90	1.51	97	2.29	95	2.39
85	2.54	85	2.26	90	2.47	95	2.64	99	3.83	97	3.20
90	3.10	90	2.85	95	3.21	97	3.40	100	9.0	99	4.39
95	3.99	95	3.68	97	3,71	99	4.78			100	25.0
97	5.20	97	4.09	99	4.63	100	41.00				
99	8.31	99	5.82	100	6.50						
100	49.0	100	21.5								

Table B3: Undersize Number Percentage (%) for Anthracite-Garnet

For 400 mm bed height											
For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.40	20	0.49	15	0.48	10	0.39	10	0.45	5	0.26
75	0.97	85	0.99	60	0.97	50	0.97	40	0.93	35	0.93
90	1.49	90	1.25	80	1.39	75	1.49	70	1.48	50	1.34
95	2.28	95	1.67	85	1.54	80	1.82	75	1.86	60	1.85
97	2.86	97	2.21	90	2.05	85	2.39	80	2.40	65	2.20
99	3.91	99	3.67	95	3.36	90	2.98	85	2.98	75	2.78
100	28.0	100	12.50	97	3.94	95	4.0	90	3.56	80	3.09
				99	5.69	97	5.11	95	4.51	85	3.72
				100	110.50	99	7.93	97	5.77	90	4.21
						100	21.5	99	8.67	95	4.93
								100	24.5	97	5.68
										99	7.27
										100	8.50
For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.45	15	0.42	10	0.43	10	0.34	15	0.38	5	0.25
65	0.96	70	0.97	45	0.98	55	0.96	80	0.97	35	0.92
75	1.37	80	1.35	60	1.42	75	1.41	90	1.42	65	1.43
80	1.86	85	1.85	70	1.97	80	1.63	95	2.69	75	1.84
85	2.40	90	2.65	75	2.39	85	2.12	97	3.41	80	2.16
90	2.92	97	3.98	80	2.79	90	2.60	99	4.82	85	2.51
95	3.75	99	5.44	85	3.24	95	3.49	100	7.50	90	3.09
97	4.31	100	10.0	90	3.72	97	3.92			97	3.99
99	6.11			95	4.58	99	5.43			99	5.13
100	23.50			97	5.42	100	15.00			100	11.0
				99	7.17						
				100	29.00						

Table B4: Undersize Number Percentage (%) for Anthracite-Garnet

For 500 mm bed height											
For 30 $dm^3\ hr^{-1}$						For 60 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
5	0.25	10	0.38	10	0.49	10	0.45	15	0.50	10	0.46
35	0.92	50	0.96	40	0.99	40	0.93	60	0.99	40	0.94
65	1.43	70	1.43	75	1.49	75	1.45	95	1.48	70	1.49
75	1.84	80	1.97	97	2.49	90	1.96	97	2.55	80	1.85
80	2.16	85	2.42	100	3.0	97	3.30	99	6.24	85	2.10
85	2.51	90	3.08			99	3.79	100	21.50	90	2.19
90	3.09	95	3.95			100	7.50			95	3.55
97	3.99	97	4.46							97	4.09
99	5.13	99	8.54							99	5.65
100	11.00	100	23.50							100	28.0
For 90 $dm^3\ hr^{-1}$						For 120 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
15	0.50	5	0.27	15	0.41	10	0.45	10	0.45	10	0.35
60	0.99	35	0.96	75	0.99	40	0.93	45	1.00	55	0.97
85	1.47	65	1.44	85	1.34	65	1.48	65	1.44	80	1.46
90	1.84	80	1.93	90	2.63	75	1.97	75	1.98	85	1.81
95	2.71	85	2.21	95	3.96	85	2.43	80	2.37	90	2.28
97	3.39	90	2.57	97	4.30	90	2.83	85	2.88	97	2.85
99	3.96	95	3.18	100	5.0	95	3.72	90	3.38	100	3.50
100	5.50	97	3.65			97	6.11	95	4.18		
		99	4.67			100	9.00	97	5.19		
		100	7.00					99	6.46		
								100	28.50		

APPENDIX-C

Table C1: Undersize Number Percentage (%) for Garnet-Calcite

For 200 mm bed height											
For 30 $dm^3\ hr^{-1}$						For 60 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.4	15	0.41	10	0.4	10	0.43	10	0.48	5	0.28
50	0.99	75	0.99	50	0.99	45	0.97	40	0.97	35	0.98
75	1.43	90	1.49	75	1.44	75	1.43	70	1.43	65	1.49
85	1.84	95	2.28	80	1.64	85	1.87	80	1.8	75	1.89
90	2.25	97	2.78	90	2.45	90	2.27	85	2.12	85	2.46
95	2.99	99	3.74	95	3.13	95	2.97	90	2.71	90	2.92
97	3.59	100	6	97	3.66	97	3.43	95	3.7	95	3.63
99	5.83			99	5.15	99	4.38	97	4.39	97	4.05
100	30.5			100	18	100	15	99	5.6	99	5.6
								100	6	100	11
For 90 $dm^3\ hr^{-1}$						For 120 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
5	0.25	10	0.43	10	0.42	10	0.47	5	0.29	10	0.4
35	0.92	45	0.97	45	0.96	40	0.95	30	0.92	50	0.99
70	1.46	80	1.48	80	1.45	65	1.45	60	1.45	75	1.49
85	1.93	90	1.89	90	1.92	75	1.81	75	1.89	80	1.8
90	2.3	95	2.63	95	2.49	80	2.06	85	2.45	85	2.22
95	3.09	97	3.27	97	3.19	85	2.79	90	3.24	90	2.86
97	3.54	99	3.87	99	3.98	90	3.5	95	4.12	95	3.35
99	5.27	100	5.5	100	10	95	5.1	97	5.02	97	4.43
100	9					97	5.97	99	9.22	99	5.73
						99	10.94	100	18.5	100	7

TableC2: Undersize Number Percentage (%) for Garnet-Calcite

For 300 mm bed height											
For 30 $dm^3\ hr^{-1}$						For 60 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.47	5	0.3	10	0.48	5	0.32	10	0.79	10	0.48
40	0.95	30	0.93	40	0.96	30	0.97	50	1.48	40	0.97
75	1.47	65	1.47	70	1.47	55	1.43	70	1.95	65	1.48
85	1.9	75	1.84	75	1.7	75	1.94	80	2.4	75	1.83
90	2.32	85	2.38	85	2.4	80	2.21	85	2.76	85	2.43
95	2.91	90	2.75	90	2.97	90	2.94	90	3.31	90	2.92
97	3.31	95	3.36	95	3.7	95	3.88	95	3.92	95	3.77
99	4.12	97	3.77	97	4.32	97	4.58	97	4.56	97	4.23
100	14.5	99	4.55	99	8.42	99	7.91	99	6.04	99	5.39
		100	8	100	30	100	18	100	7.5	100	11
For 90 $dm^3\ hr^{-1}$						For 120 $dm^3\ hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
5	0.29	5	0.25	10	0.5	5	0.26	10	0.47	15	0.43
30	0.91	40	0.99	40	0.99	35	0.92	40	0.95	70	0.98
60	1.43	60	1.42	70	1.47	65	1.44	70	1.45	85	1.38
70	1.8	75	1.91	80	1.82	80	1.96	80	1.85	90	1.72
75	2.08	85	2.42	85	2.06	85	2.42	85	2.27	95	2.66
85	2.85	90	2.99	90	2.6	90	2.92	90	2.78	97	3.23
90	3.27	95	3.71	97	3.48	95	3.73	95	3.48	99	4.06
95	3.97	97	4.3	99	4.26	97	4.21	97	4.06	100	6
97	4.49	99	6.66	100	5	99	7.75	99	7.57		
99	7.8	100	7.5			100	19.5	100	16.5		
100	10										

TableC3: Undersize Number Percentage (%) for Garnet-Calcite

For 400 mm bed height											
For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.45	15	0.41	5	0.43	5	0.25	10	0.46	10	0.45
40	0.93	70	0.97	20	0.9	35	0.92	40	0.94	45	1
75	1.48	95	1.47	50	1.42	65	1.49	70	1.47	70	1.42
90	1.95	97	1.65	70	1.92	75	1.95	80	1.86	80	1.89
95	2.34	99	2.06	75	2.17	85	2.79	85	2.14	85	2.23
97	2.6	100	3.5	90	2.9	90	3.04	90	22.53	90	2.65
99	3.19			97	3.82	97	3.86	95	3.22	95	3.35
100	10			99	7.49	100	4.5	97	3.58	97	3.78
								99	4.73	99	5.1
								100	9	100	7
For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.48	15	0.39	10	0.37	10	0.43	5	0.25	5	0.3
40	0.97	80	1	50	0.95	45	0.97	35	0.92	30	0.93
65	1.47	90	1.47	85	1.47	80	1.49	70	1.44	65	1.47
70	1.74	95	2.51	90	1.87	85	1.78	80	1.83	75	1.87
75	2.11	97	3.13	97	2.9	90	2.22	85	2.15	85	2.48
80	2.61	99	4.16	99	8.0	97	3.93	90	2.58	90	3
85	3.16					99	5.48	95	3.33	97	3.93
90	3.82					100	4	97	3.89	99	6.05
95	5.58							99	5.01	100	14
97	7.71							100	6		

Table C4: Undersize Number Percentage (%) for Garnet-Calcite

For 500 mm bed height

For 30 $dm^3 hr^{-1}$						For 60 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
10	0.43	10	0.48	5	0.28	5	0.32	10	0.79	10	0.48
45	0.97	40	0.97	35	0.98	30	0.97	50	1.48	40	0.97
75	1.43	70	1.43	65	1.49	55	1.43	70	1.95	65	1.48
85	1.87	80	1.8	75	1.89	75	1.94	80	2.4	75	1.83
90	2.27	85	2.12	85	2.46	80	2.21	85	2.76	85	2.43
95	2.97	90	2.71	90	2.92	90	2.94	90	3.31	90	2.92
97	3.43	95	3.7	95	3.63	95	3.88	95	3.92	95	3.77
99	4.38	97	4.39	97	4.05	97	4.58	97	4.56	97	4.23
100	15	99	5.6	99	5.6	99	7.91	99	6.04	99	5.39
		100	6	100	11	100	18	100	7.5	100	11

For 90 $dm^3 hr^{-1}$						For 120 $dm^3 hr^{-1}$					
Sample-1		Sample-2		Sample-3		Sample-1		Sample-2		Sample-3	
%	Size	%	Size	%	Size	%	Size	%	Size	%	Size
5	0.29	5	0.25	10	0.5	10	0.43	5	0.25	5	0.3
30	0.91	40	0.99	40	0.99	45	0.97	35	0.92	30	0.93
60	1.43	60	1.42	70	1.47	80	1.49	70	1.44	65	1.47
70	1.8	75	1.91	80	1.82	85	1.78	80	1.83	75	1.87
75	2.08	85	2.42	85	2.06	90	2.22	85	2.15	85	2.48
85	2.85	90	2.99	90	2.6	97	3.93	90	2.58	90	3
90	3.27	95	3.71	97	3.48	99	5.48	95	3.33	97	3.93
95	3.97	97	4.3	99	4.26	100	4	97	3.89	99	6.05
97	4.49	99	6.66	100	5			99	5.01	100	14
99	7.8	100	7.5					100	6		
100	10										