REMOVAL OF FINE SUSPENDED PARTICLE DESIGN AND ANALYSIS

A DISSERTATION

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

MASTER OF TECHNOLOGY

in

CHEMICAL ENGINEERING

(With Specialization in Industrial Pollution Abatement)

By

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CANDIDATE'S DECLARATION

I hereby declare that the work, which is being presented in the dissertation entitled **"REMOVAL OF FINE SUSPENDED PARTICLE DESIGN AND ANALYSIS"** in the partial fulfillment of the requirements for 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 June 2005 to June 2006 under supervision of **Dr. I. M. Mishra**, Professor, Department of Chemical Engineering, Indian Institute of Technology, Roorkee, Roorkee and **Mr. Jayant K. Joshi,** Senior Manager (Environment), Engineers India Limited.

The matter, embodied in this dissertation has not been submitted by me for the award of any other degree.

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

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ABSTRACT

Treatment of sea water requires great attention for the removal of present impurities and pollutant. During the treatment of sea water one the basic problem is removal of suspended particle. There are several processes present for the removal of suspended particle such as reverse osmosis, membrane filtration, ion exchange process etc. All these process contain the porous polymer membrane for the removal of fine suspended particle range 0.2μm to 20μm. But high concentration of fine suspended particle in process sea water is result in membrane fouling which require high maintenance cost of the membrane as well as low cycle time of the process.

The objective of this current experimental work to discusses fine suspended particle removal from water by filtration through low cost coarse media. A laboratory scale technique was used to remove fine suspended particle from fine suspended particle bearing water in the case of Calcite, Anthracite and Garnet and a filtration column is design on high specific flow rate of 30 $m^3/m^2.h$ .The result was obtained that for very low size media (0.3mm to 0.45mm) the removal for Calcite is marginally high compare to Anthracite and quite high compare to Garnet. It can be observed from the result that the removal mechanism were due to the effect of rough solid surface and the porous structure of Calcite, Anthracite and Garnet.

Filtration result indicates that at an input with fine suspended particle concentration of 20 mg/dm^3 and particle size in the range of 0.5micron to 20 micron, a good removal was observed with the Calcite as compare to the Anthracite and Garnet at high specific flow rate of 30 $m^3/m^2.h$. 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 the removal efficiency.

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Vipender Singh

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INTRODUCTION

1.1 GENERAL

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 Ultrafiltration 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.

To ensure trouble free-operation of ion exchange units, resin manufacturers recommend removal of suspended solids to negligible levels, since any solids not removed 'up-front' will generally accumulate within the ion exchange resin beds, which are themselves excellent filters. In fact, the high electrostatic surface charge or zeta potential present on ion exchange resins enhances the ability of ion exchange resins to remove even extremely fine, charged colloidal particles[20]. The adverse effects of solids accumulation in resin beds include the distribution of water and regenerants within the resin bed and ultimately the pressure loss across the bed causing deterioration in the quality and quantity of the treated water. Even small amounts of solids adhering to the surface of the resin beads may adversely affect exchange kinetics, long before pressure drop increases are observed.

With conventional co-current ion exchange systems, a small amount of solids accumulating within the resin bed can often be removed by regular back-washing. There has been a trend over the past decade or so to replace conventional co-flow ion exchange demineralizers with countercurrent packed-bed. While the advantages of these newer technologies cannot be denied, their pre-treatment requirements are more stringent than traditional technology[17]. This is because packed-bed systems are not backwashed on a regular basis. To do so would obviate the advantages offered by counter-current regeneration. In one case where a packed bed ion exchange design was retrofitted to an existing conventional co-flow system, problems due to iron fouling occurred as often as every 3 months instead of every 6 months with the co-flow system[23] until the pre-treatment system was upgraded Although various methods have been developed to help alleviate this situation[18,19], the general consensus is that packed bed ion exchange systems are more prone to fouling with suspended solids than conventional systems.

The past two decades have seen a dramatic increase in the use of reverse osmosis demineralizers. RO systems are even less forgiving in terms of pre-treatment than ion exchange6. The close spacing of spiral wound membranes results in trapping of suspended solids inside the modules. This is exacerbated by the fact that, like ion exchange resins, RO membranes bear a surface charge which may cause fine solids to be attracted to the membrane surface. Once fouling begins, cleaning of the membranes becomes very difficult and the system may not return to original performance levels once fouling has occurred. According to one manufacturer, "membrane fouling in RO systems is as all pervasive and inevitable as the common cold" [20]. In fact, many of the failures experienced by these systems can be traced back to inadequate pre-filtration.

The level of suspended solids removal ahead of an ion exchange or reverse osmosis system depends on the concentration and nature of the contaminants as well as how the system is to be operated. Higher concentrations of suspended solids will obviously necessitate more frequent ion exchange bed backwashing or membrane cleanings. Although there is no such thing as a standard pretreatment process, the most prevalent IX/RO pretreatment has been direct media filtration using one or more layers of granular media such as sand and/or anthracite with coagulation. According to Cline[17], conventional depth media filtration is most efficient at removing materials down to 10-20 microns. If particle size distribution evaluations indicate a high proportion of solids smaller than 10 microns, supplemental filtration should be considered to deal with finer solids. For this reason, media filters are often supplemented with membrane cartridge filters.

Under conditions of optimal coagulation and system operation, multimedia filters provide excellent performance. Consistent operation is somewhat difficult to obtain in practice in many instances, however. Such filters typically undergo a ripening period when they are first put into service after a backwash, during which time water quality is inferior. Moreover, while the filtrate quality tends to improve as the cycle proceeds, termination of the cycle normally sees an increased pressure drop accompanied by a breakthrough or leakage of turbidity. One of the keys to successful operation is an effective backwash (i.e. cleaning) cycle. Long-term performance can deteriorate due to a slow buildup of solids in the media over many cycles, ultimately resulting in the creation of *mudballs*.

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.

Nevertheless, many in the industry are suggesting that the overall performance of media filters is inadequate for reliable packed-bed ion exchange or reverse osmosis performance. In order to provide improved TSS removal for ion exchange and reverse osmosis, microfiltration[21] and ultrafiltration[22] membranes have been 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. Moreover, in some cases, membrane pretreatment may just move the fouling problem upstream.

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. 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 densities, to high density inorganic particulates. Particulates ranging in size from the 2μm up to 100 μm can be captured by granular media, provided that surface conditions are suitable for attachment of the particulates to the media surface or to previously deposited particulates.

1.2 OBJECTIVES OF THE PRESENT 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 is emphasis manly on three media i.e. Calcite, Anthracite, and Garnet to ensure the fine suspended particle removal ability at different bed height as well as with time. Tap water with suspended solid (10mg/dm^3, size-2μm to 20μm) is taken as a flowing fluid at constant flow rate (60 dm^3/h) for study purpose. In the present study a comparisons are made among the three media on the basis of the fine suspended solid particle removal. To make this comparision all the analysis was done with the Particle Size Analyzer. The comparision is made on the basis of removal efficiency at different bed height of the media. Moreover the removal efficiency of each media is determined and compare 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 and Garnet.
- To study about change in fine suspended particle (below 20μm) removal efficiency of Calcite, Anthracite and Garnet at 200 mm, 300 mm, 400 mm and 500 mm bed height at constant flow rate.
- To study about change in fine suspended particle (below 20μm) removal efficiency of Calcite, Anthracite and Garnet with time (at 30, 60 and 90 min) at different bed height at constant flow rate.
- To compare the removal efficiency of Calcite, Anthracite and Garnet at different bed height at different time period (at 30, 60 and 90 min) at constant flow rate.

LITERATURE REVIEW

Filtration technologies are classified under two major categories, manly slow sand filtration and rapid sand filtration, depending on the mode of filtration. Slow sand filters, which include biological activity in addition to physical and chemical mechanisms for removing impurities from the raw water, are especially suitable for small community water supplies, because of their large areal requirements. The removal mechanism of particle in a slow sand filter includes mechanical straining, sedimentation, diffusion and chemical and biological oxidation. Coarse and fine particles of suspended matter are deposited at the surface of the filter bed by the action of mechanical straining and sedimentation respectively, while colloidal and dissolved impurities are removed by the action of diffusion. By chemical and biological oxidation, the deposited organic matter is converted into inorganic solids and discharged with filter effluent .microbial and biological process and hence the removal of impurities, take place mainly in the top layer of the filter bed.

The rapid sand filter, on the other hand, due to its lower areal requirement (25 – 150 times lass than the slow sand filter s), is used widely as a final clarification unit in water treatment. Rapid filtration is used as the clarifying step in municipal water treatment plants. If the raw water has water has turbidity in excess of 10-20 there is NTU, the rapid filter must be provided with efficient flocculation and sedimentation as pretreatment units. In rapid filters practically no biological action; at the most there is some nitrification in certain cases when the speed is limited , when the oxygen content is adequate, and when the nitrification bacteria find favorable nutritive condition in the water .

Comparison of Slow Sand Filter and Rapid Sand Filter [14]

Characteristic	Slow sand filter	Rapid sand filter
Filtration rate	2-5 $m^3/m^2.d$	120-360 $m^3/m^2.d$

Size of bed	Large (2000m^3)	Small (100m^3)
Depth of bed	300 mm gravel, 1 m sand	500 m gravel, 0.7-1.0m sand In some cases sand and anthracite are used as dual media
Effective size of sand	0.35 mm	0.6-1.2 mm
Head loss	Up to 1 m	Up to 3 m
Length of run	20-90 days	1-2 days
Method of cleaning	Scrape off top layer and Wash	Back wash with water and air + water scour

Three are several factors which have great impact on the removal of fine suspended particle present in the feed stream required great attention; such as flow rate of influent , concentration of suspended particle, type of media, time duration of filtration, media structure, attrition of media, backwashing process etc. To describe these effects and impact and to fill our goal there is a large literature present.

Within the scope of a research programme into wastewater treatment scenarios based on physical-chemical pretreatment, the application of direct influent filtration (DIF) as a first treatment step in a wastewater treatment system was investigated. The aim of the experimental research was to investigate 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 wastewater is possible, notwithstanding 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 solubles 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 *(Nieuwenhuijzen, Graaf and Mels, 2001)*.

Boller and Kavanaugh (1995) demonstrated that the rate of headloss build-up in a granular media filter, for a constant mass of solids being removed, is strongly dependent on the size of the particulates in suspension and the size of the granular media. It was postulated that the principal cause of the more rapid increase in headloss observed for smaller particles in many filtration studies is due to the nature of the aggregation or deposition process inside the pore spaces of the porous media, rather than a surface controlled phenomenon. The particulates deposit in a manner similar to the phenomenon occurring during floc formation in suspension. The larger the number of particles deposited, the lower the deposit density, and thus, the depletion of the available pore space for particle collection occurs more rapidly, manifested by a more rapid increase in hydraulic gradient in any given filter layer. Using data from several well controlled filtration studies, it was demonstrated that this hypothesis may be correct. Translation of the relationships for floc size-density developed for suspended particles to the deposition process in granular media filters was successful. In order to gain more insight into the hypothesized phenomena and to further prove the importance of the proposed parameters for headloss development in deep bed filters, more specific experimental studies will be necessary.

The analysis further demonstrated that deposit density is a strong function of the media size, with a relatively more rapid headloss development observed in larger size media, due to the lower relative density of deposits that occurs in the larger pore spaces in larger size media. Furthermore, the analysis also showed that filtration velocity appears to have little or no impact on the density of the particulate deposits. The results of this analysis have several implications for filtration operation (*Boller and Kavanaugh, 1995)*.

Filtration is an important solid–liquid separation technology employed widely in the mineral processing industries. The effectiveness of the filtration operation can be influenced by numerous variables, related to the particulate phase, the slurry rheology and the equipment. Continuous filtration of fine particles involves filter cake formation and removal of surface moisture by drawing air through the pore structure network. In order to gain a better understanding of the complex transport phenomena that occur in the filter cake, analysis of the effect of the three-dimensional pore geometry on the effective transport properties of the filter cake is necessary. In this regard, analysis of the pore connectivity in a packed bed of particles should allow for a detailed description of fluid flow and transport in the filter cake structure. Two interrelated approaches, namely computer simulation and experimental measurement, can be used to gain knowledge of pore microstructure and its correlation to macroscopic cake properties. In this regard, a three-dimensional Monte Carlo method was used in this work to simulate cake structure. As the resolution and the techniques for three-dimensional geometric analysis have advanced in the last decade, experimental measurements are now possible to specify in detail the pore structure in three-dimensional digital space using high-resolution X-ray microtomography. Thus in addition to computer simulation, this paper presents preliminary experimental findings of pore structure in three-dimensional using X-ray microtomographic techniques (*Lin and Miller, 2000)*.

Accurate assessment of the transport properties of porous media (in our case, filter cake) is of major importance in the development of improved filtration processes. Implications from these studies are important in the design and operation of filtration equipment in order to enhance the efficiency of this important solid–liquid separation process. The microstructure and connectivity of pore space are important features to describe detailed fluid flow phenomena in filter cake during fine particle filtration. In this regard, 3D characterization of pore structure is essential. The pore structure has to be described by parameters which are of special relevance to the interpretation of fluid transport phenomena. These parameters should be based on directly measured variables of the pore system and not inferred from indirect variables (such as those determined empirically from transport processes) valid only for a particular pore structure. In this way, fundamental relationships between pore structure and fluid transport at the microstructure level can be described. In order to achieve this level of sophistication, the three-dimensional interconnected pore structure of filter cake must be determined. Study of fluid transport phenomena in filter cake using X-ray microtomography (XMT) to characterize the complex three-dimensional pore geometry is discussed. On this basis, the lattice Boltzmann (LB) method is used to simulate fluid flow and to begin to establish a fundamental relationship between pore microstructure and effective transport coefficients. For example, can network analysis using skeletonization procedures be used to confirm fundamental relationships that might be developed? Eventually then, with this detailed model, design and operating variables could be optimized for improved filtration efficiency *(Lin and Miller, 2003)*.

Browen and Pickering (2002) introduce a concept of spectrum filter; by employing a fine media in the lower layer of the filter it is therefore possible to dramatically reduce the TSS of the filter. The Spectrum filter is basically a two layer depth media filter. There are a number of features which depart significantly from the conventional design, however. Its main design features are as follows:

Coarse upper layer: The top layer consists of approximately 30 inches (76 cm) of coarse anthracite. This material is similar to, but somewhat finer than that used in a conventional dual media filter. As with conventional dual layer filters, the top layer provides the bulk of the solids retention and therefore defines the run length.

Fine Micro-media lower layer: The lower layer of the Spectrum filter is a significant departure from the conventional design. Whereas dual media filters typically employ about 8 inches (20 cm) of silica sand with an effective size of about 0.35 mm, the Spectrum filter uses a layer of high density media with an effective size of less than 0.1 mm. The lower layer removes the residual quantity of fine suspended solids not retained in the upper layer and therefore effectively defines the filtration efficiency.

High service flow rate: Service flow rates are significantly higher with the Spectrum filter. The range is 12-18US gpm/ft^2 compared to normal maximum of about 8 gpm/ft^2.

Smaller diameter vessel: As a result of the higher flow rate, the diameter of the filter vessel required to treat a given flow of water can be reduced.

Simultaneous air scour/backwash cleaning cycle: Most media filters use a simple water backwash, wherein the filter is taken out of service and a flow of water is passed up through the filter bed to expand the media. This allows collected dirt to disengage and be flushed out the top of the filter.

- *Low backwash flow rate:* The smaller media particle diameter reduces the terminal settling velocity of the media and allows the use of lower backwash velocities- typically less than 50% of a conventional filter. Since the higher service flow allows the use of smaller vessels, the backwash flow rate can be as little as 25%of a conventional filter.

Somewhat higher pressure drop: The higher service flows and finer media result in higher initial (i.e. clean) pressure drop across the filter. Typical pressure drop is about 30 psi (0.2 Mpa) compared to about 5 psi (0.03 Mpa) for a conventional filter.

Shorter Service cycle: Although the dirt holding capacity of the Spectrum filter, which is defined primarily by the upper layer, is similar to a conventional filter, the service cycle is shorter because of the higher service flow rate.

Experimental data published by *Amirtharajah et. al. (1991)* on the conditions for washing filter media by simultaneous air and water have been re-examined, employing the classical relationships for pressure loss in granular media developed by Kozeny and Carman, making the assumption that the air and water move through the interstices at about the same velocity. It has been found that for sand Amirtharajah'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 (*Stevenson, 1997).*

Bai and Tien, (1996) examines 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 the 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 a short period of flow shock (i.e., to be subject to a higher flow rate during this period). Both sets of experiments reconfirmed the dependence of particle detachment on particle size collector gain size, and headloss gradient (degree of filter clogging) of the filter, as predicted by the analysis.

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. Experimental data on the development of suspension concentration profiles through laboratory scale filter beds during the backwash process were obtained. Previous attempts to obtain and record backwash profiles of this type have been unsuccessful due to the limited range of existing turbidimeters *(Hall and Fitzpatrick, 2000)*.

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 are *(Hall and Fitzpatrick, 2000):*

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

Aziz and Smith (1996) discuss 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 at 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. Studies on the effect of different parameters on the removal performance of manganese have shown that a smaller particle size, a greater filter depth, and a lower flow rate gave advantages in the removal efficiency.

This paper discusses manganese removal from water by filtration through low coast media. Four parameters were studied in the filtration technique to see each effect on the removal efficiency of manganese.

(i) Type of filtration media (limestone and gravel);

(ii) Filtration media depth (250 and 500 mm);

(iii) Particle size (2 and 4 ram); and

(iv) Flow rate (10, 20, 40 and 80 ml/min).

The percentage removal of manganese increases marginally with a decrease in the particle size of limestone the difference is due to a higher surface area of smaller particles, giving a better surface contact between the particle and the manganese solution. It is also observed that the manganese removal decreases with an increase in the flow rates. At high flows, there is a reduced period of surface contact between the particle and the manganese solution as well as a higher velocity of flow through the media increasing the sloughing of precipitate from the media.

By increasing the filter depth the removal performance of manganese increases significantly. A good removal at a higher filter depth is due to a greater detention time (i.e. the time taken for the influent to move from the inlet point, passing the column's packed material and going out to the outlet) offered by a higher filter depth. The filtration results confirm the previous studies that limestone particle gives a better manganese removal performance as compared to gravel. This removal can be achieved without increasing the water hardness exceeding standards. A good removal of Manganese was also observed with smaller particles, high filter depth, and at low flow rates *(Aziz and Smith, 1996).*

A significant consideration in forward planning for water treatment works design and operation concerns the effectiveness of a filtration plant in providing a barrier to particulates in the low micrometer size range, including Cryptosporidium oocysts. The performance of rapid gravity filtration plants is believed to be dependent on backwash and start-up regimes. It was the aim of this study to optimize direct sand filtration by identifying optimum filter backwash and start-up conditions which minimized the passage of particulates into the filtrate. 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 h run, do so in the first hour of operation. Optimum combined air water "collapse-pulsing" backwash durations were identified that reduced the number of 2-5μm particles entering the filtrate, especially during the ripening period. Slow start-up was also found to reduce the number of 2-5μm particles in the filtrate during the ripening period. The reductions in particulate passage resulting from a slow start was found to be media dependent, with smaller media requiring a longer slow start duration than coarser media. This experimental work examines the effects on particulate passage during filter ripening of *(Colton, Hillis and Fitzpatrick, 1996)*:

(1) "collapse-pulsing" backwash duration and

(2) Slow start.

Attrition resistance of granular filter media is becoming increasingly important as materials such as GAC and anthracite are being more frequently used as filter media. The purpose of this 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 at combinations that gave the condition known 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%). It can be seen 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 (*Humby and Fitzpatrick, 1996)*.

Size density relationships for aggregated particulates in suspension are transferred 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 (*Boller and Kavanaugh, 1994).*

The limited efficacy of disinfectants, other than ultraviolet irradiation and ozonation, as a barrier against Cryptosporidium parvum in drinking water treatment has underscored the increased importance of oocyst removal by filtration. Currently, no reliable surrogates have been identified for C. parvum removal by filtration. As a result, evaluations of the Cryptosporidium removal by treatment operations have been performed using oocysts. It has typically been assumed that chemically inactivated oocysts are suitable surrogates for viable oocysts. Measurements of electrophoretic mobility, however, have shown that chemical inactivation changes the surface charge of Cryptosporidium oocysts. The present bench-scale research indicated that formalin-inactivated oocysts are reliable surrogates for viable oocysts during both stable filter operation and periods where filtration processes are challenged, such as coagulation failure. This finding is important because of the practical difficulties associated with using viable oocysts in filtration investigations. Poor coagulation conditions severely compromised removal of viable and inactivated oocysts by dual- and tri-media filters compared to stable operating conditions and filter ripening, emphasizing the importance of optimized chemical pre-treatment (coagulation) for the successful removal of oocysts during filtration. The treatment optimization experiments also indicated that tri-media filters offered only marginally higher oocyst removals than dual-media filters *(Emelko, 2003).*

Biological aerated filters (BAFs) are an attractive process option, particularly when low land usage is required. They can combine BOD, solids and ammoniacal 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 $CODm^{-3}d$ 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^{-1}, the BAF using the larger medium produced an effluent containing more than 20mg L^{-1} 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 2.3m 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 2.1-0.3m was used by the 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 (*Moore, Quarmby and Stephenson, 2000).*

GMM Ochieng, FAO Otieno, TPM Ogada, SM Shitote and DM Menzwa study was aimed at introducing multistage filtration (MSF) (a combination of slow-sand filtration (SSF) and pretreatment system - horizontal flow roughing filter (HRF)) as an alternative water treatment technology to the conventional one. A pilot- plant study was undertaken to achieve this goal. Evaluating the MSF performance vs. the existing conventional system in removing selected physical and chemical drinking water quality parameters together with the biological water quality improvement by the MSF without chemical use was done. Evaluation of the effectiveness of the MSF system utilizing locally available material, i.e. gravel, improved agricultural waste (charcoal maize cobs) and broken burnt bricks as pretreatment filter material was also done The benchmark was the Kenya Bureau of Standards (KEBS) values for the selected parameters. Results showed that with proper design specifications, MSF systems perform better than conventional systems under similar conditions of raw water quality and environmental conditions. The tested locally available materials can also be effectively used as pretreatment media with each allowing a filter run greater than 82 d and therefore could serve as alternatives where natural gravel is not readily available. With special reference to the bacteriological quality improvement, the MSF greatly improved the bacteriological quality of the water recording removal efficiencies of over 99% and 98% respectively for *E. coli* and total coliforms. Despite the observed performance, MSF should be complemented with chlorination as a final buffer against water-borne diseases. However, in this case, the dosing will be greatly reduced when compared to the conventional system.

Slow sand filters are used in rural regions where source water may be subjected to antimicrobial contaminant loads from waste discharges and diffuse pollution. A numerical model (LETA) was derived to calculate aqueous antimicrobial concentrations through time and depth of a slow sand filter and estimate accumulating contaminant mass in the schmutzdecke. Input parameters include water quality variables easily quantified by water system personnel and published adsorption, partitioning, and degradation coefficients. Simulation results for the tetracycline, quinolone, and macrolide classes of antimicrobials suggested greater than 3-log removal from 1 mg/dm^3 influent concentrations within the top 40 cm of the sand column, with schmutzdecke antimicrobial concentrations comparable to other land-applied waste biosolids. A 60-day challenge experiment injecting 1 mg/dm^3 tylosin to a pilot slow sand filter showed an average 0.1 mg/kg of the antimicrobial remaining in the schmutzdecke layer normally removed during filter maintenance, and this value was the same order of magnitude as the sorbed concentration predicted by the LETA model *(Rooklidge and Ketchum, 2002).*

Chapter-3

FUNDAMENTAL OF FILTRATION

3.1 GENERAL

The term solid-liquid filtration covers all process where a liquid containing suspended solids is freed of some or the entire solid when suspension is drawn through porous medium. Filtration process are widely used in different industrial applications including water treatment and reclamation, water purification for the pharmaceutical, micro-electronic, chemical and mineral industries, separation of hazardous solids from chemical wastes and removal of deleterious components. Description of filtration process include the physical features of granular medium depth filter, filter media characteristic, the filtration process in which suspended material is removed from the liquid, and the backwash process, in which the material that has been retained within the filter is removed.

3.2 FILTRATION THEORY

3.2.1 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 porous 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 recalculated 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[28]

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

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[28]

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

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.

3.2.2 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.

3.2.3 Effect of the Sedimentation on Filtration

There are two important effects due to particle sedimentation which may affect the rate of filtration. First, if the sediments particle are all setting at approximately the same rate, as is frequently the case in a concentrated suspension in which the particle site distribution is not very wide, a more rapid build-up of the particle size distribution is not very wide, a more rapid build –up of particle will occur on an upward-facing surface and a correspondingly reduced rate of build-up will take place if the filter surface is facing downwards. Thus, there will be a tendency for accelerated filtration with downwards-facing filter surfaces and reduced filtration rates for upward-facing surfaces. On the other hand, if the suspension is relatively dilute, so that the large particles are settling at a higher-rate than the small ones, there will be a preferential deposition of large particles on an upwards-facing surface during the initial stages of filtration, giving rise to a low resistance cake. Conversely, for downward-facing surface, fine particles will initially be deposited preferentially and the cake resistance will be correspondingly increased. It is thus seen that there can be complex interactions where sedimentation is occurring at an appreciable rate, and the orientation of the filter surface is an important factor.

3.2.4 Bed Filters

For the purification of water supplies and for waste water treatment where the solid content is about 10 g/m^3 or less, granular bed filters have largely replaced the former very slow and sand filters. The beds are formed from granular material of grain size 0.6-1.2mm in beds 0.6-1.8m deep. The very fine particle of solids is removed by mechanical action although the particles finally adhere as a result of surface electric forces or adsorption. This operation has been analyzed by *IWASAKI (25)* (*Coulson and Richardson, Volume-2*) who propose the following equation

$$-\frac{\partial C}{\partial l} = \lambda C$$

on integration

$$\frac{C}{C_o} = e^{-\lambda l}$$

Where C is the volume concentration of solids in suspension in the filter,

C_0 is the value of C at the surface of the filter,

l is the depth of the filter,

λ is the filter coefficient.

3.3 BACKWASH THEORY

3.3.1 General

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 pipework 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.

3.3.2 Head Loss due to Backwash

To expand the filter bed, comprised of uniform medium hydraulically, the head lost must equal the buoyant mass of granular media. Mathematically, this relationship can be expressed as:

$$h = L_e(1-\alpha_e)\left[\frac{\rho_s-\rho_w}{\rho_w}\right] \qquad \text{------ (3.1)}$$

3.3.3 Pressure Drop due to the Backwash [6]:

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:

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

$$\left(\frac{R'}{\rho u^2}\right) = 5\,\mathrm{Re}^{-1} + 0.4\,\mathrm{Re}^{-0.1} \qquad \text{------- (3.3)}$$

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

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*).

3.3.4 Mechanism

The retention of particles in a filter bed requires that the adhesion force between a retained particle and a collector (i.e., one of the filter grains) is at least in equilibrium with the hydrodynamic force which tends to detach the particle. Particle detachment was the cause of Particle detachment in filtration therefore occurs only when the hydrodynamic force overcomes the adhesive force. The mechanisms for particle detachment from filter grains include rolling, sliding, and lifting. The sliding mechanism implies that a deposited particle may slide because of the hydrodynamic tangential drag force acting on it. The rolling mechanism depicts that a deposited particle may roll due to the torque resulting from the asymmetry of the shear force acting on the deposited particle. As particle rolling can occur over a collector surface and does not necessarily lead to particle detachment, the following analysis is therefore based on the sliding and lifting mechanisms. In deep bed filtration, the diameter of filter grains in most cases is two orders larger than those of the particles to be removed. Thus, attachment of particles onto filter grains may therefore be treated as that of a spherical particle onto a collector plane.

3.4 MEDIA THEORY

3.4.1 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.

3.4.2 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.

Other new develop filter system which meet the present filtration requirements are:

- Disc Tube Filter
- Spin Klin Filter
- Valveless Autowash Gravity Filter
- Travelling Band Screening
- Rotary Drum Screens
- Automatic Self Cleaning Strainers
- Rotary Micro Drum Screens
- Horizontal Vacuum Belt Filter
- High Pressure Filter etc.

Chapter-4

EXPERIMENTAL PROCEDURE

4.1 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 Fluid	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

4.2 EXPERIMENTAL SETUP

The experimental work was carried out on a direct filtration setup using in line fine powder. A down flow media filter was built from a 1200mm tall plexiglass tube with a 50mm internal diameter (Fig.1), to which a pump is attached for the continuous flow of the slurry 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 slurry 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 inlet and outlet of the flow. A $0.2m^3$ capacity holdup tank was used to maintain the continuous flow of slurry.

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.
- 5 globe valves and 1 gate valve for controlling the flow rate.
- A rotameter to measure the flow rate of water in the range of 10 dm^3/h to 100 dm^3/h.

Backwash arrangement

4.3 MATERIAL AND METHOD

The experimental setup was operated as a down-flow filter. The solid-water slurry 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 three parameter were studied on the filtration and the removal efficiency of the fine suspended particle, viz.

(1) the effect of the filtration media ; the effect of the
(2) filtration media bed height ; and the
(3) cycle time (at 30min, 60min, and 90min)

4.3.1 Procedure of filtration

Three types of filter media were used, viz. calcite, anthracite and garnet. 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 in to the column. The media was supported on a 75 mesh stainless steal 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, 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 Institute Instrumentation Analysis center, 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 and wait for 15 min.
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.

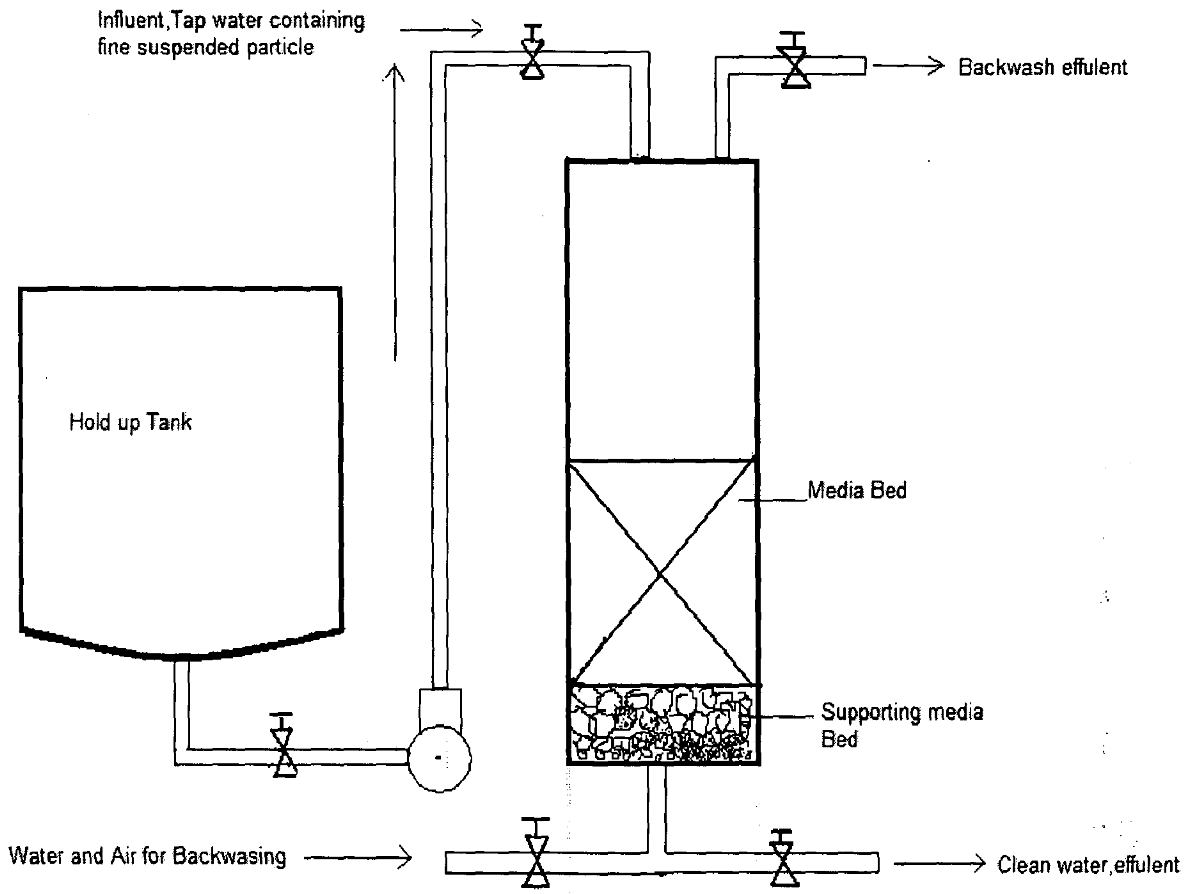


Figure 4.1 Experimental Setup for the Removal of Fine Suspended Particle

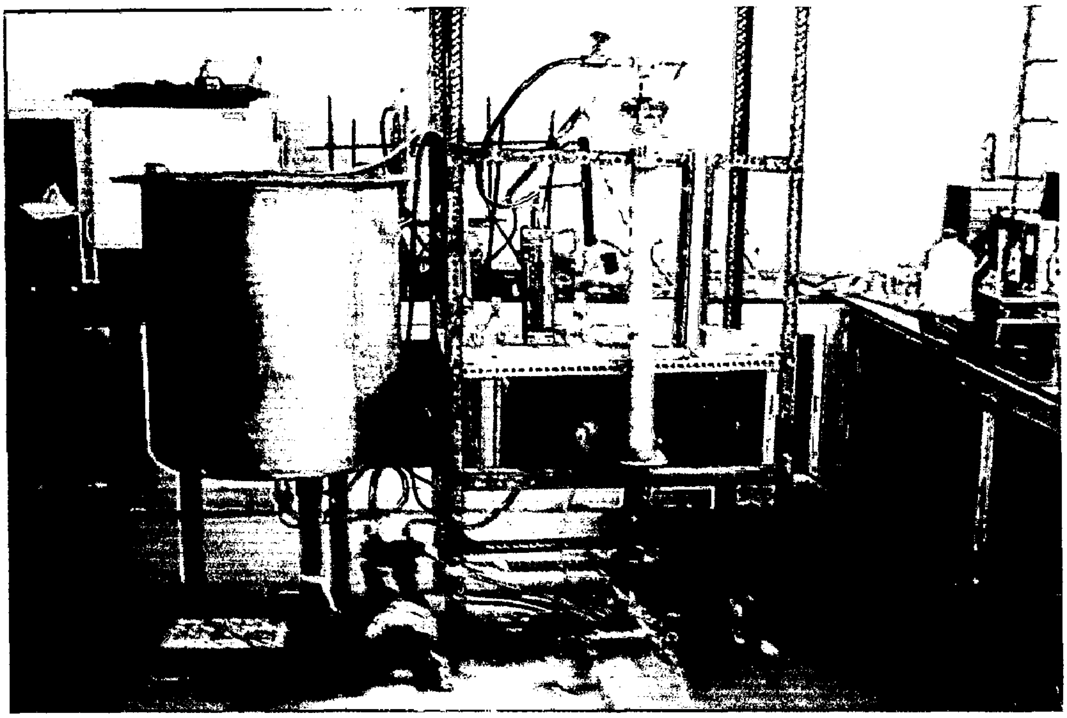


Figure 4.2: Actual Setup of Filter Column in Mass Transfer Lab, IIT Roorkee.

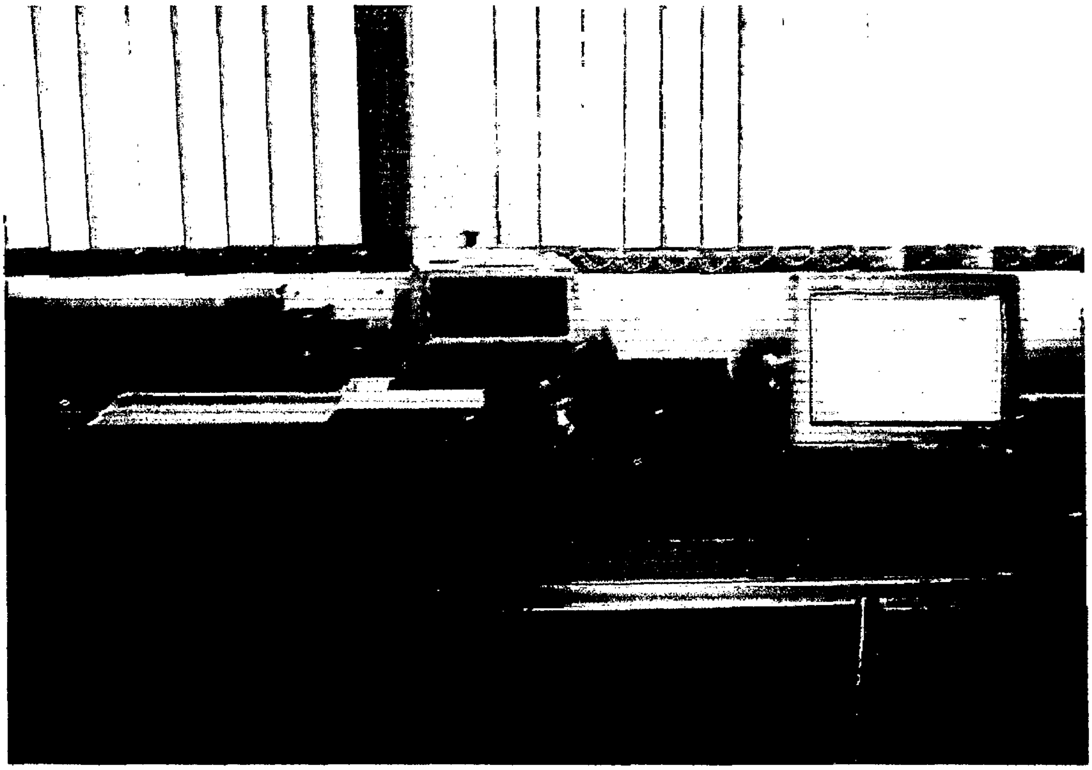
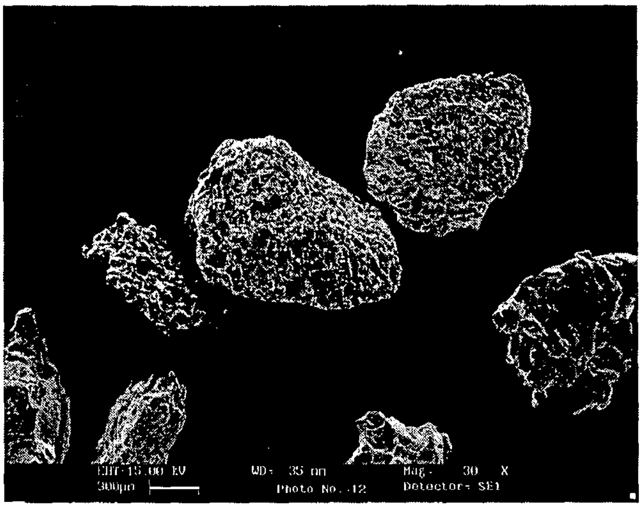
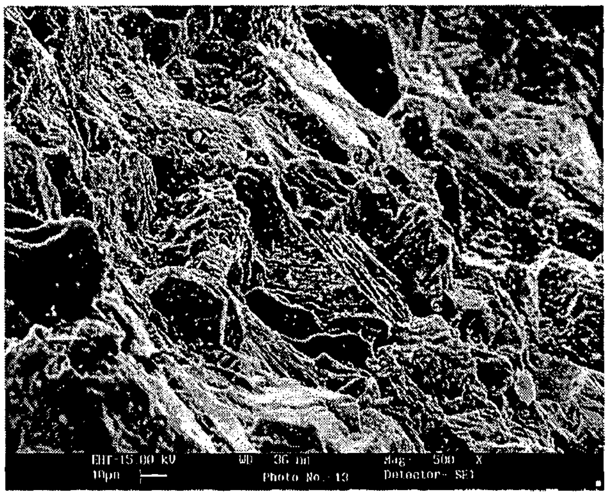


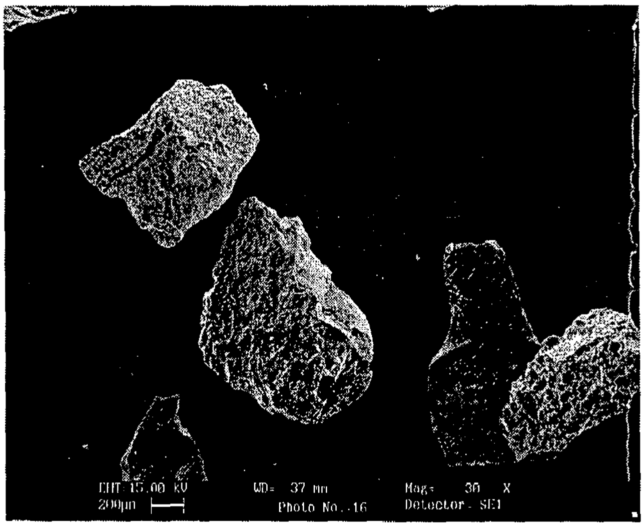
Figure 4.3: Particle Size Analyzer in Instrumental Analysis Lab, in IIT Roorkee.

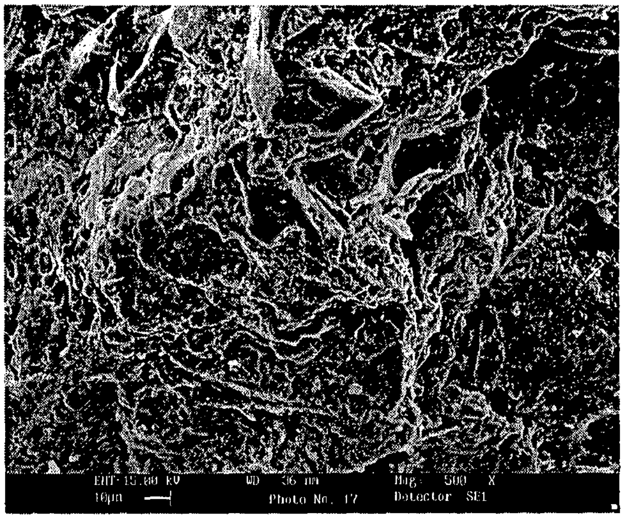




SEM of Calcite at 30 X

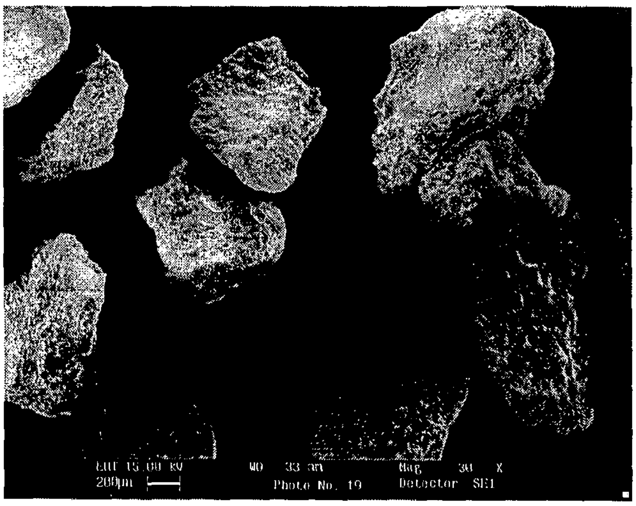
at 500X

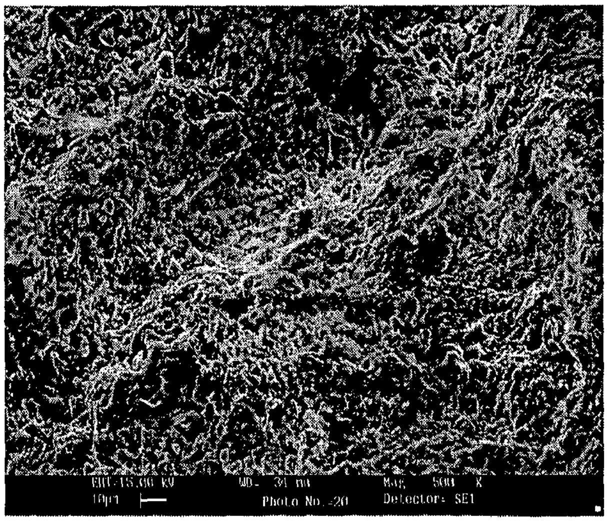




SEM of Anthracite at 30 X

at 500 X



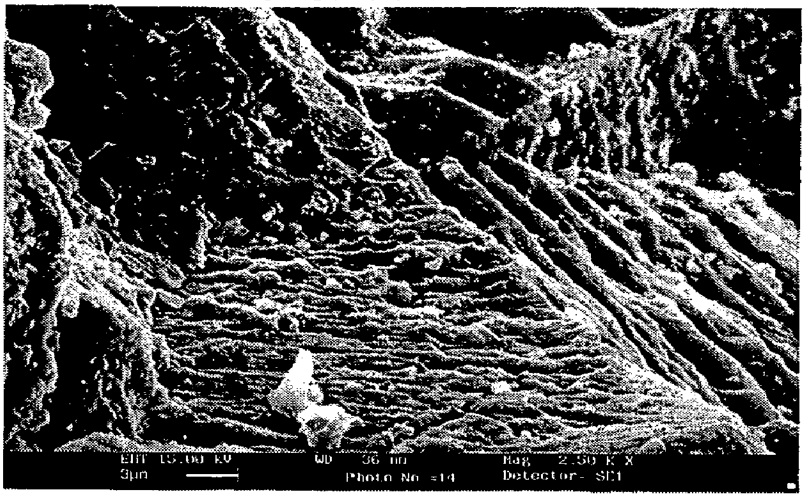


SEM of Garnet at 30 X

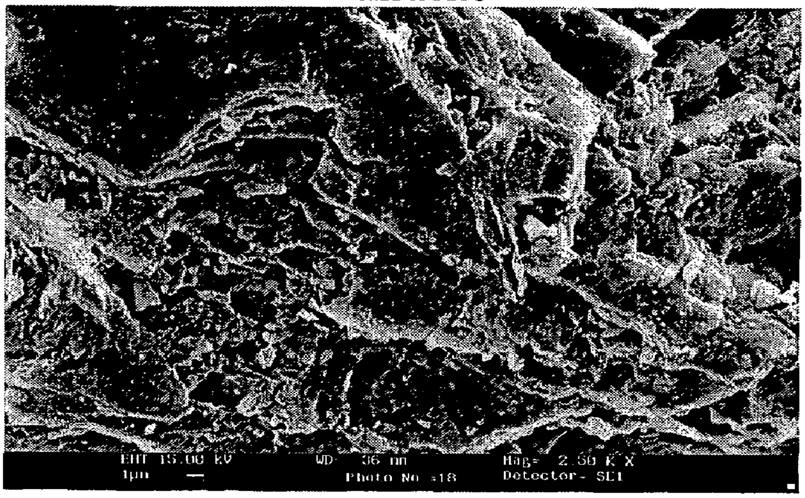
at 500 X

Figure- 4.4 SEM of Calcite, Anthracite and Garnet at 30 X and 500 X

SEM of Calcite



SEM of Anthracite



SEM of Garnet

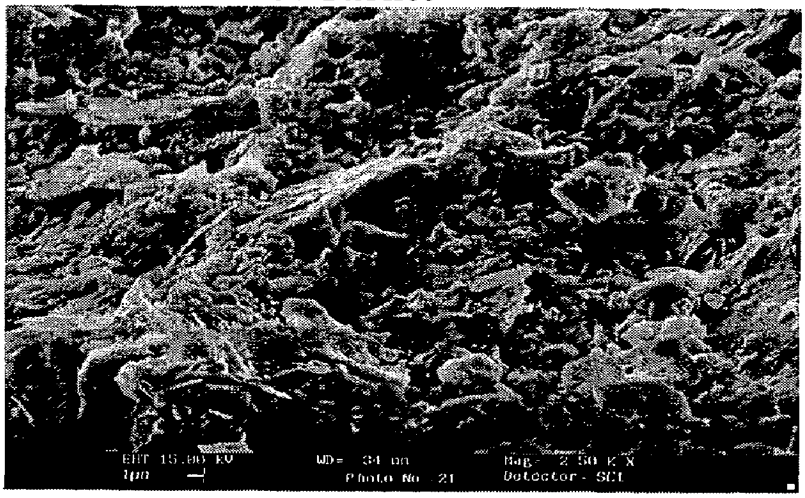


Figure- 4.5 SEM of Calcite, Anthracite and Garnet at 2.5 KX

Chapter-5

RESULTS AND DISCUSSION

The performance of different filtration media in removing of fine suspended particle is given in Fig.5.1 to 5.12 in term of Percentage Cumulative Number present in sample, taken at definite time interval of 30 min in the 90 minute run time of experiment at particular bed height, for the removal of different particle size. The samples called main sample, sample-1, sample-2 and sample-3 are taken at t=0 from influent at t=30 min, t=60 min and t=90 min from effluent respectively for each run.

The results shown are at different bed height for each media. Fig. 5.1, 5.2, 5.3 and 5.4 shows the percentage cumulative number of different size suspended particle in effluent sample at 200 mm, 300 mm, 400 mm and 500 mm bed height for calcite respectively and comparison is made at 30, 60 and 90min in each figure. While Fig. 5.5, 5.6, 5.7, and 5.8 shows the result for Anthracite and Fig. 5.9, 5.10, 5.11, and 5.12 shows the result for Garnet.

Fig.5.1 to Fig.5.12 shows that mean particle size diameter is shifting from higher particle mean diameter to lower particle mean diameter, which intrpretate that the higher size particles concentration is reducing. Results shows that mean particle size are reducing during first 60 min and than start increases again. It can be seen that in Fig.5.1 the mean sample size for the main sample is 2.19μm which is decreases to1.16 μm for first 30 min and further decreases to 0.82 μm after 60 min and than start increases to 1.92 μm for sample taken after 90 min.

The statistic data gives the concentration of particle (Total Number of Particle/ml in the measurement cell of Particle Size Analyzer) given in Table 5.1, percentage of solid particle in the sample and undersize Number Percentage given in Appendix-A, B, and C.

The value of removal efficiency and total number of suspended solid particle can be calculated with the help of percentage cumulative number undersize (Appendix-A) and total number of particle in the samples for each sample. All the data are taken from the statistics of each sample analyze by Particle Size Analyzer.

Removal efficiency of suspended particle is calculated as

$$100\times\left[N_i - N_0\right]/N_i \qquad \text{---------------- (5.1)}$$

Where N_i is the number of suspended particle present in main sample (taken from influent) and N_0 is number of particle present in the outlet sample.

Fig.5.13 shows the effect of time on the removal of fine particle with Calcite, Fig. 5.14 shows the effect of time on the removal of fine particle with Anthracite and Fig. 5.15 shows the effect of time on the removal of fine particle with Garnet used as media.

The removal efficiency for Calcite, Anthracite and Garnet at different bed height is given in Fig. 5.16, 5.17 and 5.18 respectively. The comparisons of different media in removing fine suspended particle are given in Fig. 5.19, 5.20 and 5.21 at different bed height after 30 min, 60 min and 90 min respectively.

The performance of the Calcite and Anthracite is higher during starting of the experiment and continuously decreases as time of filtration increases in removing fine suspended particle shown in Table 5.1. It can be seen from Fig. 5.1 that at t=30 min mean and t=60 min is lower than the mean particle size distribution. The result is followed by the Fig.5.2, 5.3, and 5.4 for Calcite and Fig.5.5, 5.6, 5.7 and 5.8 for Anthracite. It can be seen from Fig.5.13 the removal efficiency is decreases with time for Calcite at all bed height. The same results are shown in Fig. 5.14 for Anthracite and Fig.5.15 for Garnet.

This behavior of media is due to the physical structure of media as shown in Fig. 4.4 and 4.5. In case of Anthracite and Calcite the pore structure and grain is filling by the fine particle in early stage of experiment which result in low removing efficiency in later stage. While the SEM of Garnet Fig. 4.4 and 4.5, is quite different from Anthracite and Calcite. To remove the fine particle attached to the media particle and regain the removal efficiency of media the backwashing of media is required. Fig. 5.13 and 5.14 for Calcite and Anthracite respectively shows that removal efficiency is 94% at t=30 min, 92% at t=60 min, and 88% at t=90min for Calcite and 93% at t=30 min, 93.5% at t=60 min and 86% at 200 mm bed height for Anthracite at 30 $m^3/m^2.h$ specific velocity of influent (60 dm^3/h for the experiment).

The performance, in terms of fine suspended particle removing, of media increase as the bed height increases. It can be seen from Fig. 5.16 for Calcite, Fig. 5.17 for Anthracite and Fig.5.18 for Garnet that as the bed height increase from 200 mm to 500 mm the removal efficiency increases. For example at 200, 300, 400 and 500 mm bed height the removal efficiency are 92%, 96%, 97% and 99% for Calcite ; 93%, 94%, 95.7% and 96% for Anthracite and 83%, 86%, 92% and 90% for Garnet.

A good removal at a higher filter depth is due to a greater detention time (i.e. time taken for the influent to move from the inlet point; passing the column packed material and going out to the outlet) offered by the higher filter depth.

Finally a comparison is made among the Calcite, Anthracite and Garnet at different bed height in Fig. 5.19, 5.20 and 5.21 at different time which shown that at all bed height the removal efficiency is higher for Calcite than Anthracite and Garnet. For example, it can be seen from the Fig. 5.19 that the removal efficiency is more than 99% for Calcite, 97% for Anthracite and around 94 % for Garnet at 500 mm bed height.

Table 5.1: Total Number of Particle Present in the Sample & Percent Reduction

Media		Calcite		Anthracite		Garnet	
Bed Height (mm)	Sample	Number of Particle / ml	Percent Reduction %	Number of Particle / ml	Percent Reduction %	Number of Particle / ml	Percent Reduction %
200	S-0 feed sample	18605	-	21924		18605	
	S-1 at t=30min	1116	94	1534	93	2976	84
	S-2 at t=60min	1488	92	1540	93	3162	83
	S-3 at t=90min	2232	88	1447	93.4	3721	80
300	S-0 feed sample	18605		21924		18605	
	S-1 at t=30min	558	97	1052	95.2	2046	89
	S-2 at t=60min	744	96	1315	94	2604	86
	S-3 at t=90min	930	95	1622	92.6	2790	85
400	S-0 feed sample	18605		21924		18605	
	S-1 at t=30min	372	98	811	96.3	1674	91
	S-2 at t=60min	550	97	942	95.7	1489	92
	S-3 at t=90min	556	97	1381	93.7	1860	90
500	S-0 feed sample	18605	-	21924		18605	
	S-1 at t=30min	130	99.3	658	97	1120	94
	S-2 at t=60min	186	99	876	96	1864	90
	S-3 at t=90min	378	98	1316	94	1824	90.2

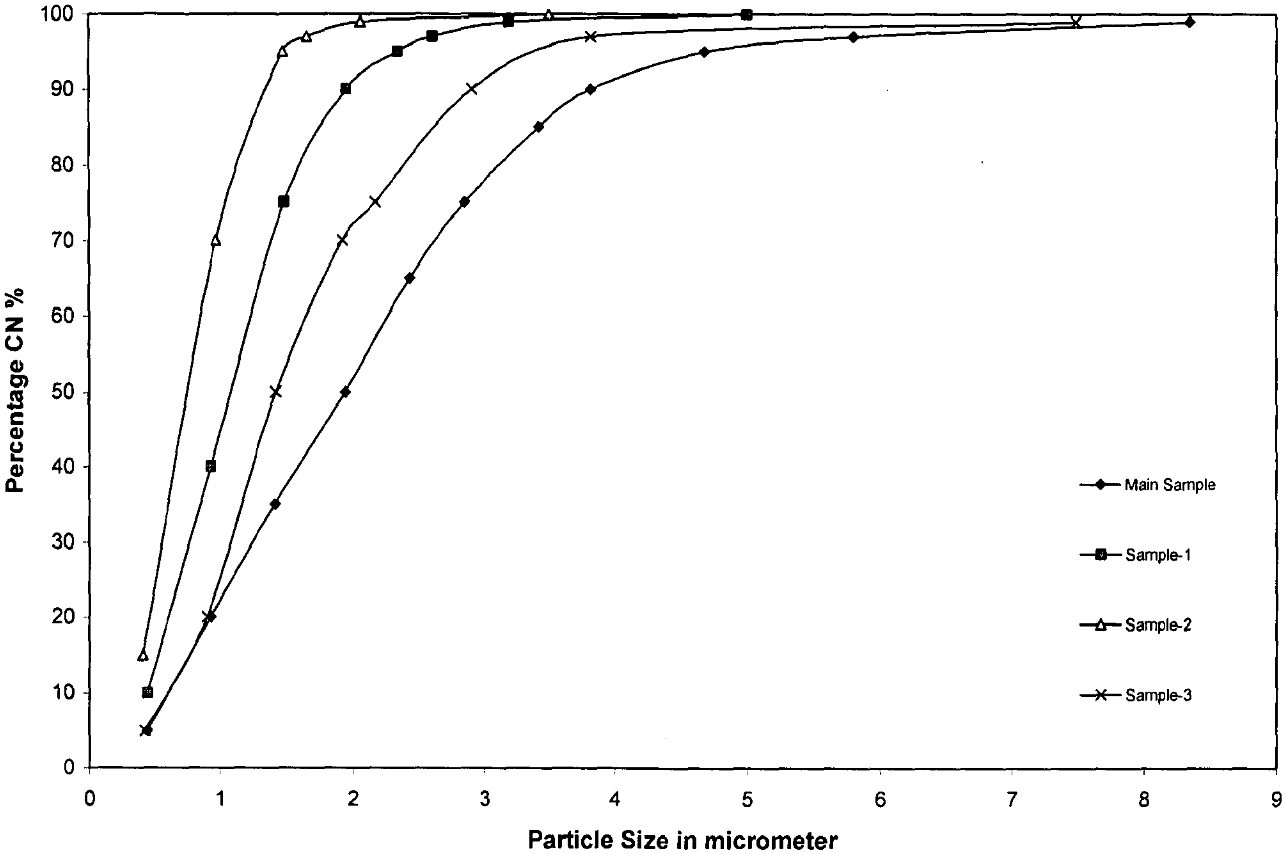


Figure-5.1 Removal at 200 mm Bed Height of Calcite

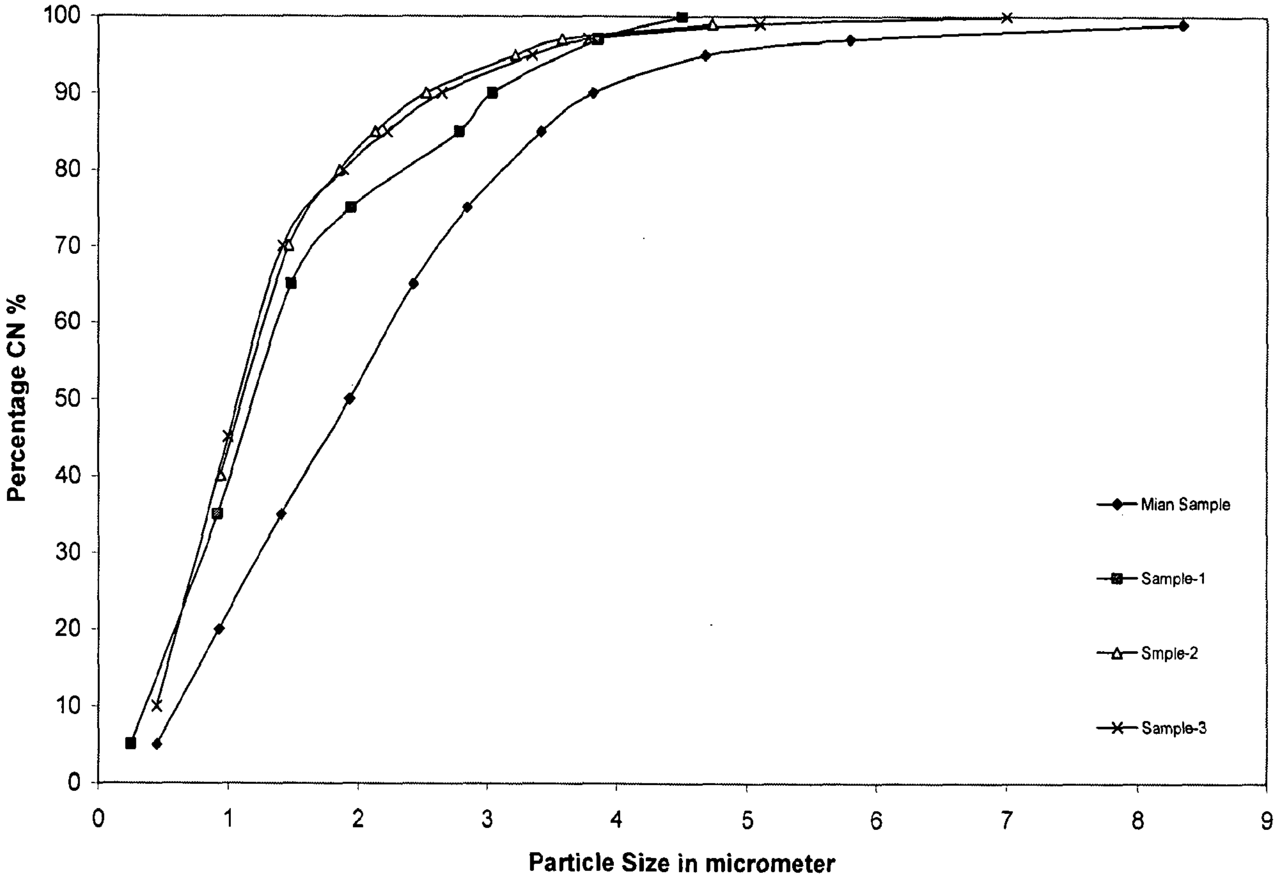


Figure-5.2 Removal at 300 mm Bed Height of Calcite

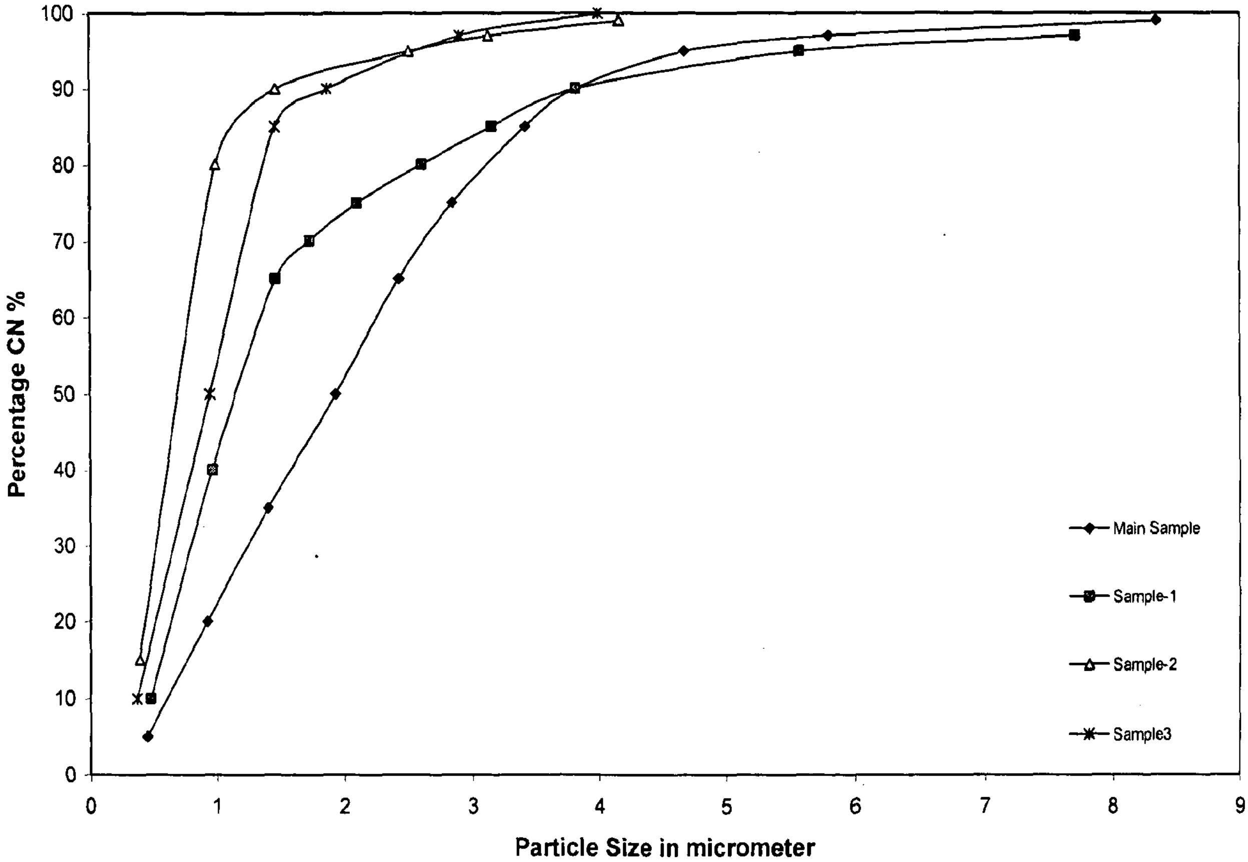


Figure-5.3 Removal at 400 mm Bed Height of Calcite

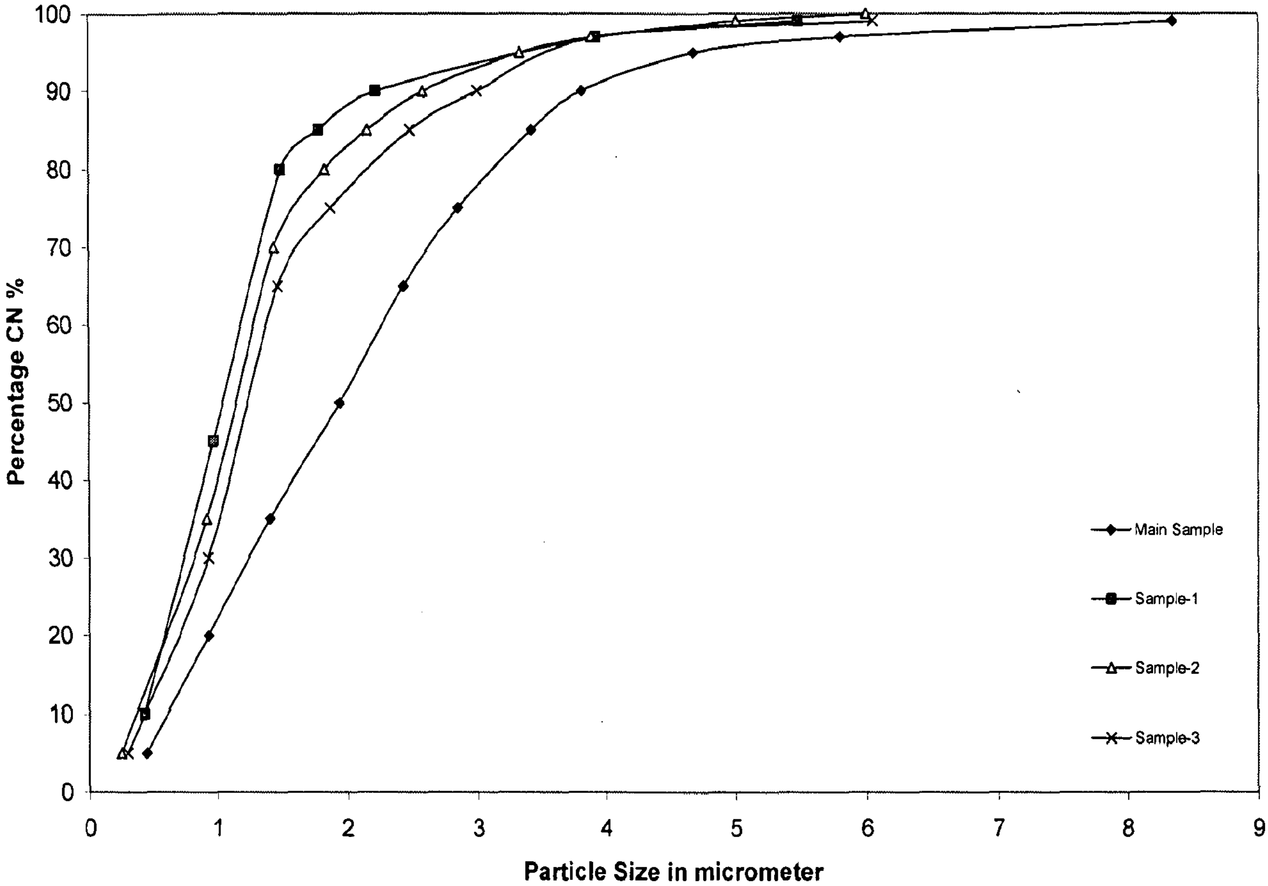


Figure-5.4 Removal at 500 mm Bed Hight of Calcite

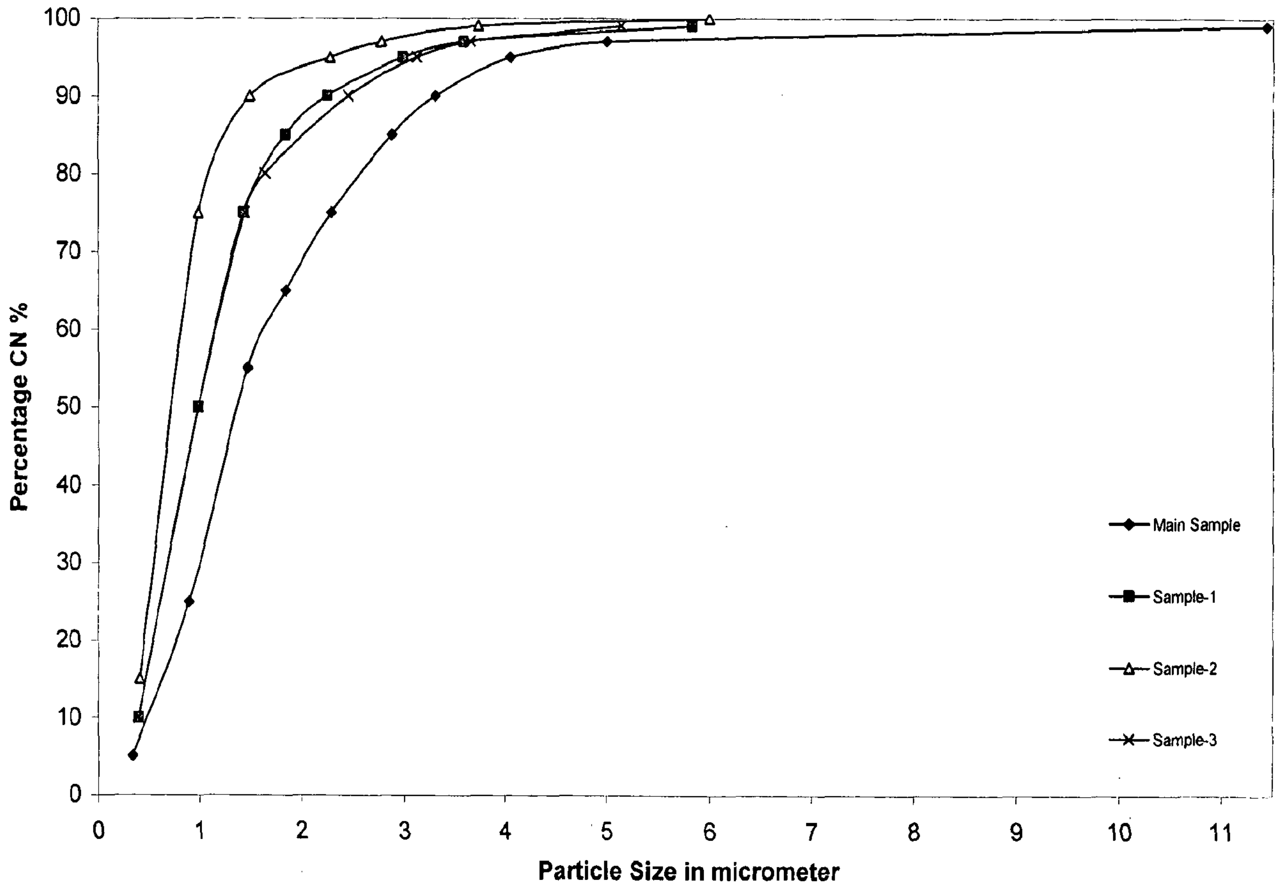


Figure-5.5 Removal at 200 mm Bed Height of Anthracite

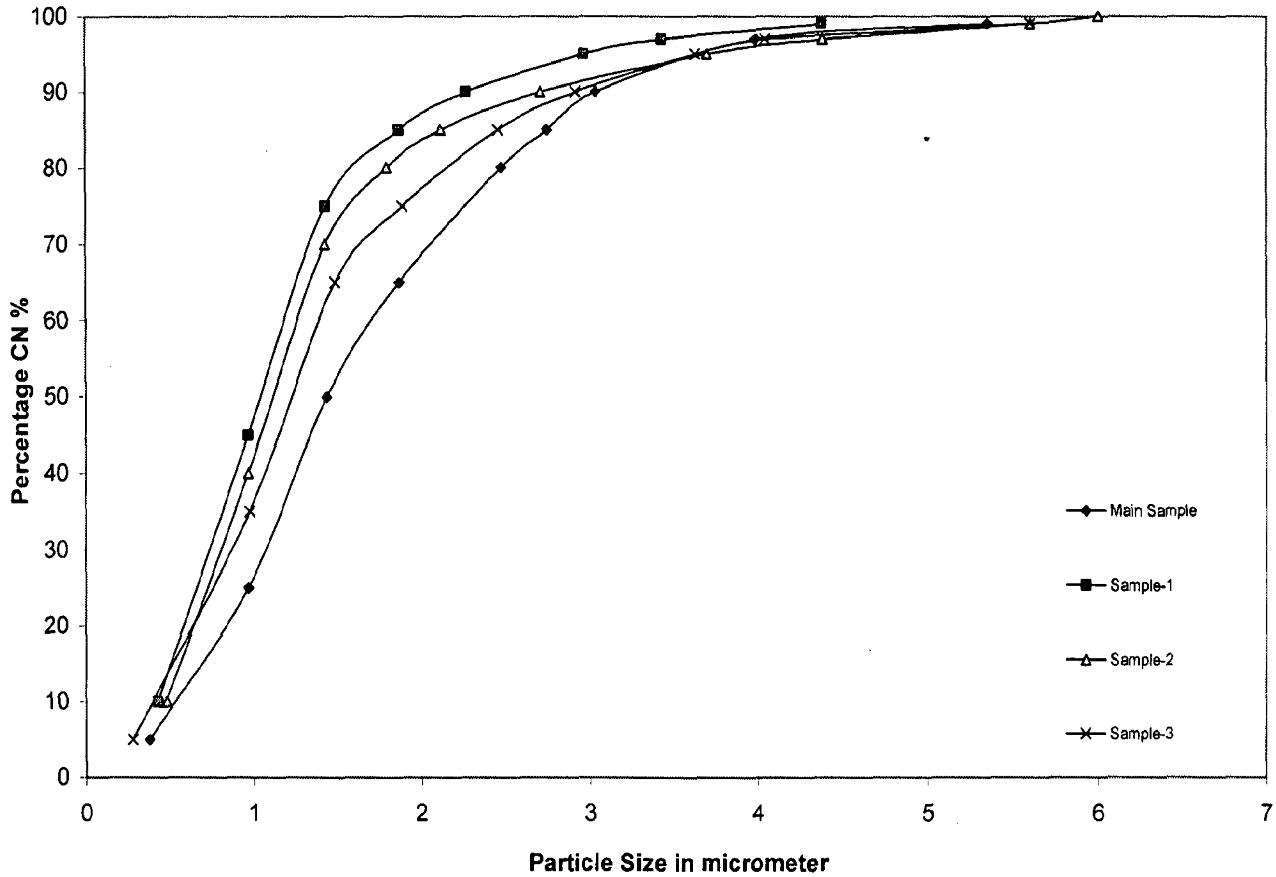


Figure-5.6 Removal at 300 mm Bed Height of Anthracite

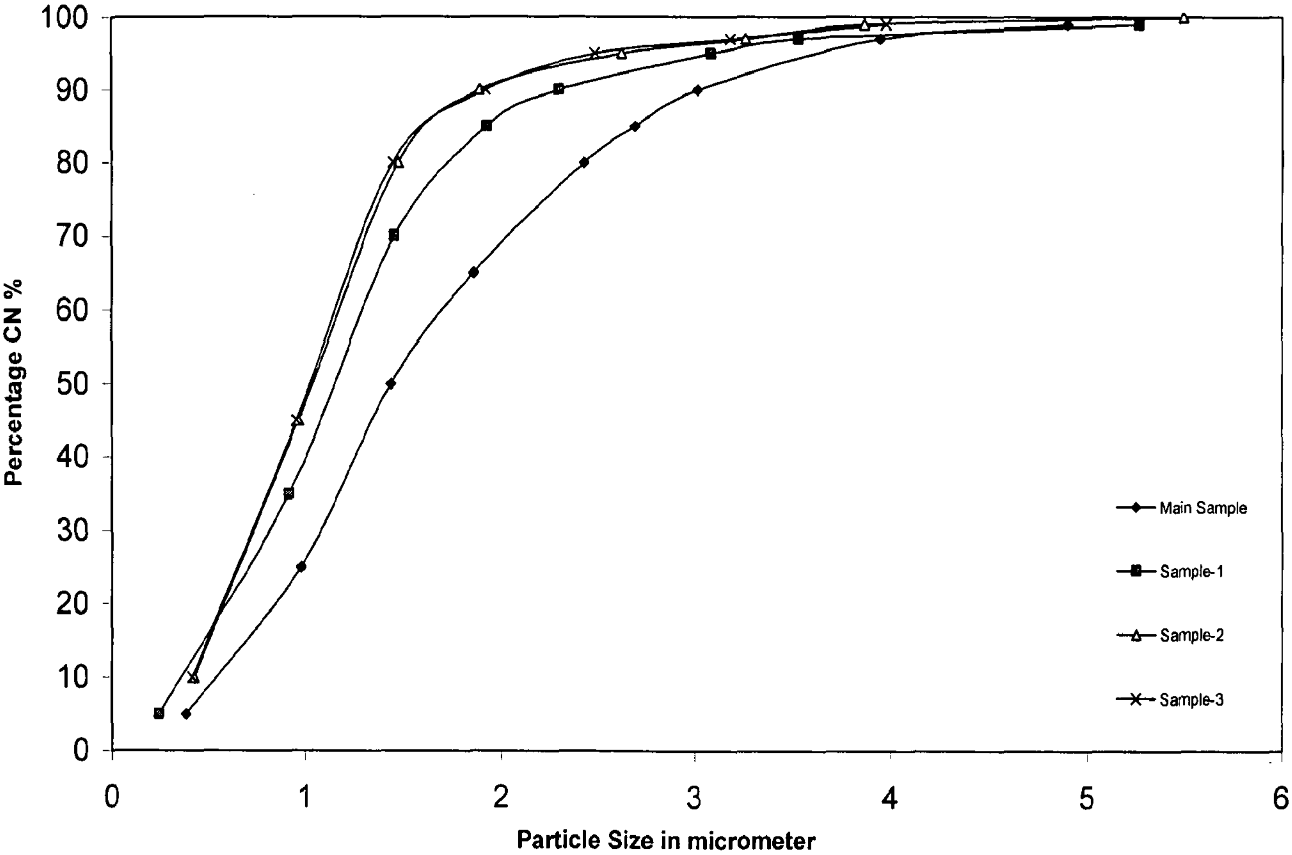


Figure-5.7 Removal at 400 mm Bed Height of Anthracite

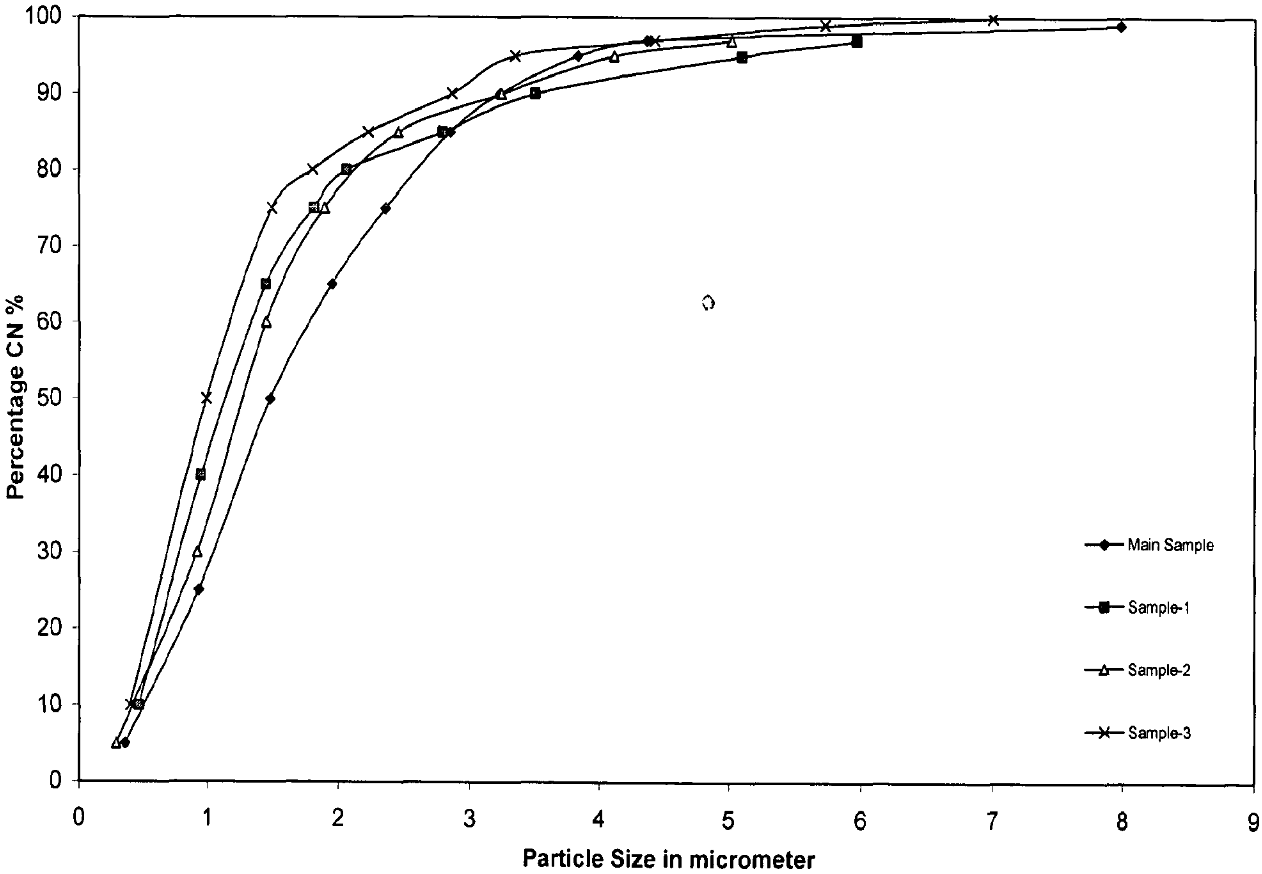


Figure-5.8 Removal at 500 mm Bed Height of Anthracite

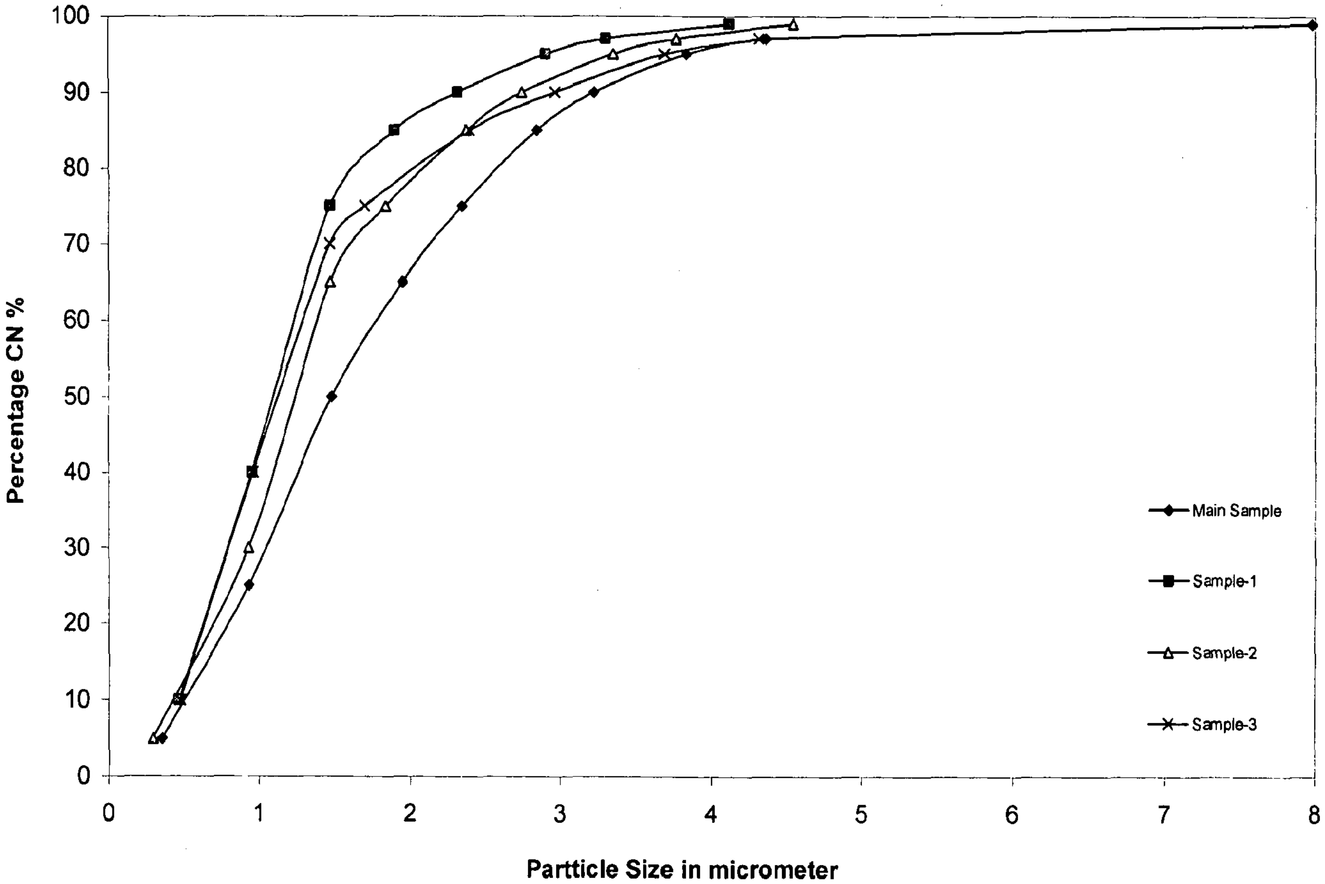


Figure- 5.9 Removal at 200 mm Bed Height of Garnet

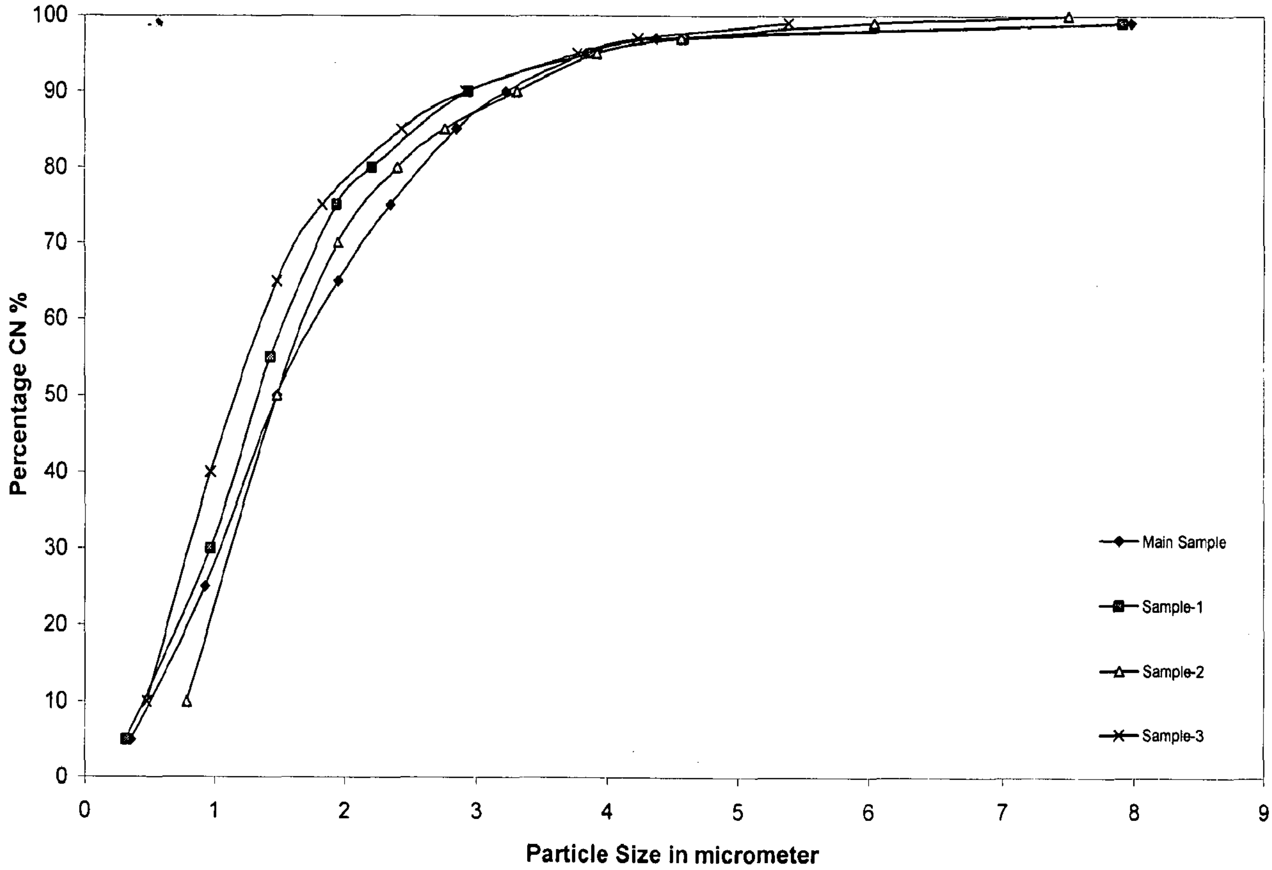


Figure-5.10 Removal at 300 mm Bed Height of Garnet

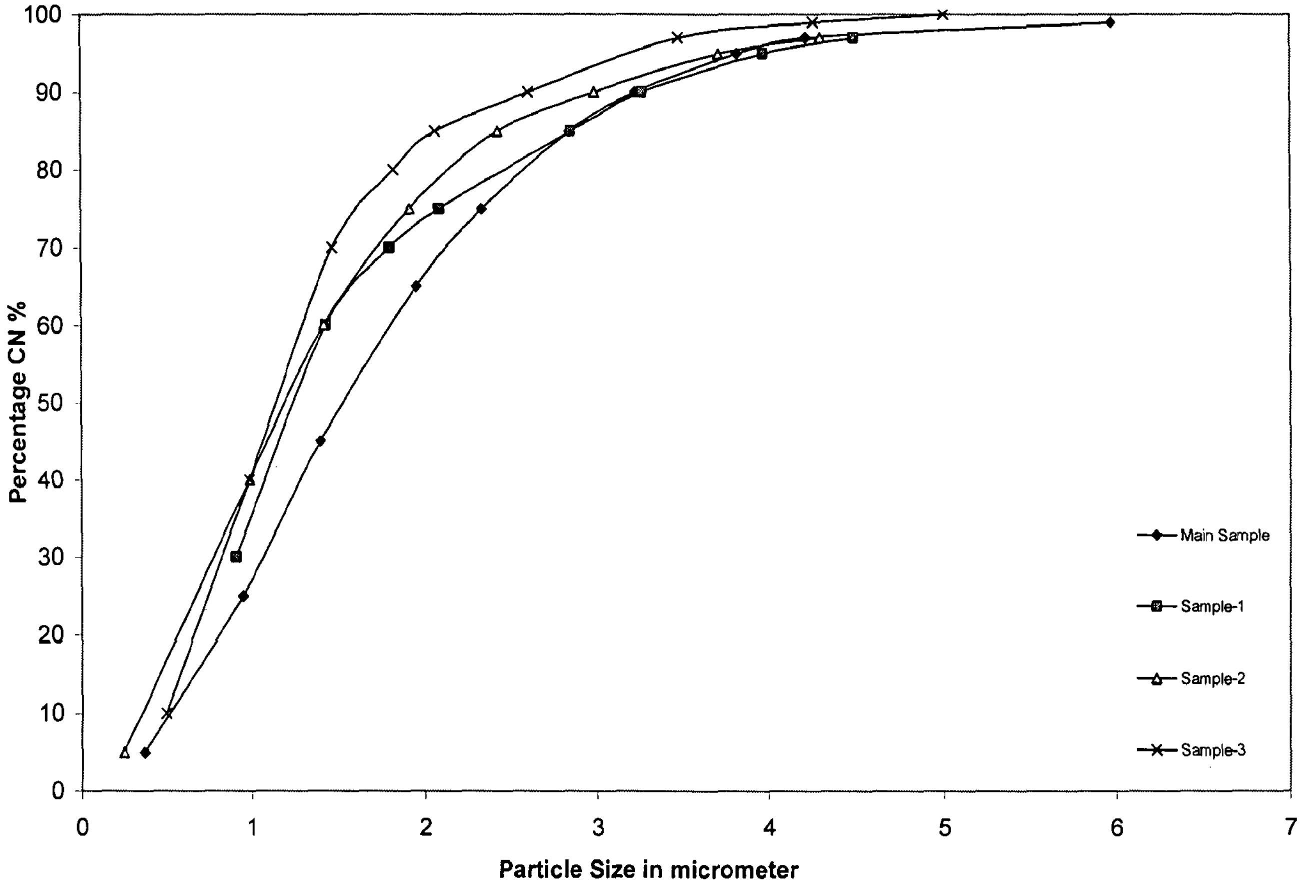


Figure-5.11 Removal at 400 mm Bed Height of Garnet

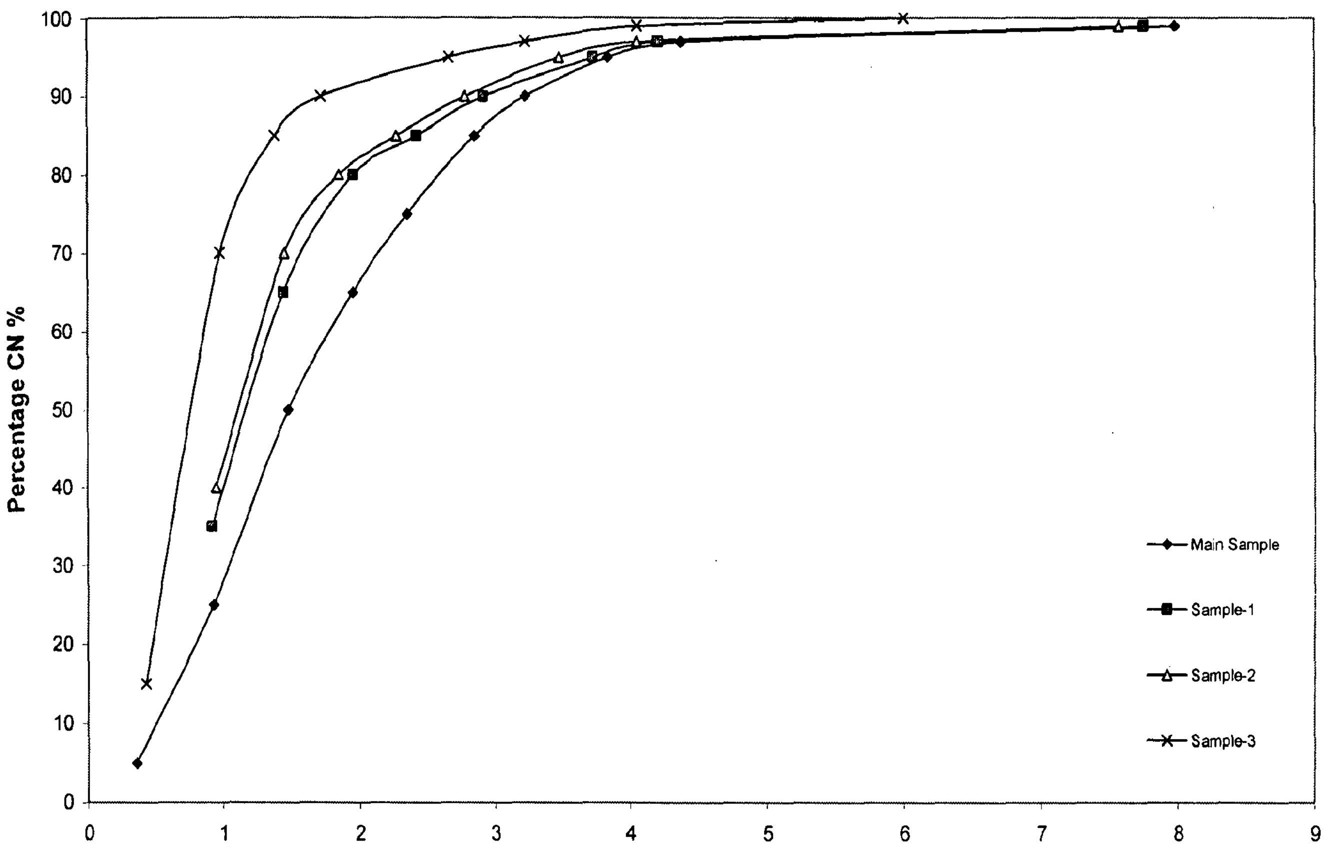


Figure-5.12 Removal at 500 mm Bed Height of Garnet

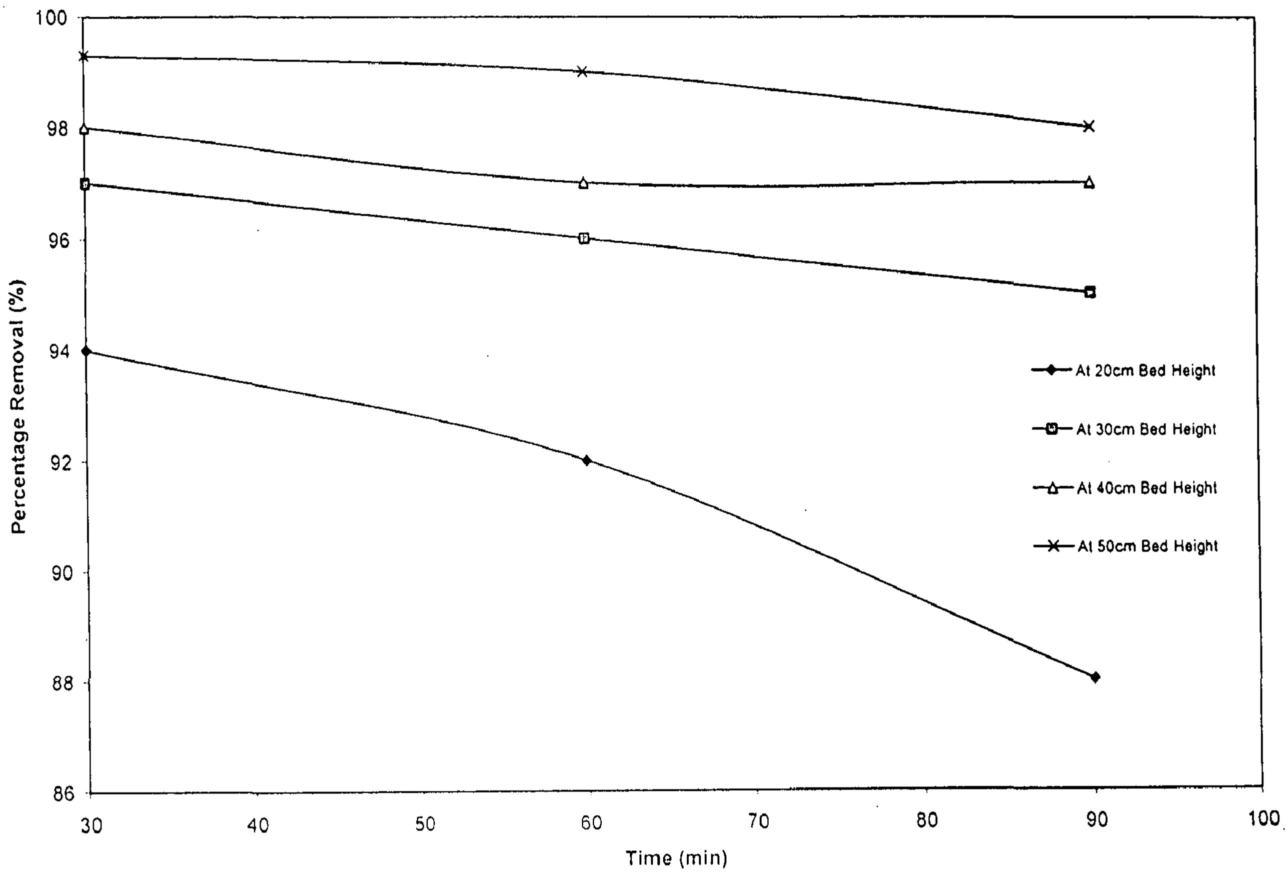


Figure-5.13 Removal of Fine Suspended Particle for Calcite as a Media

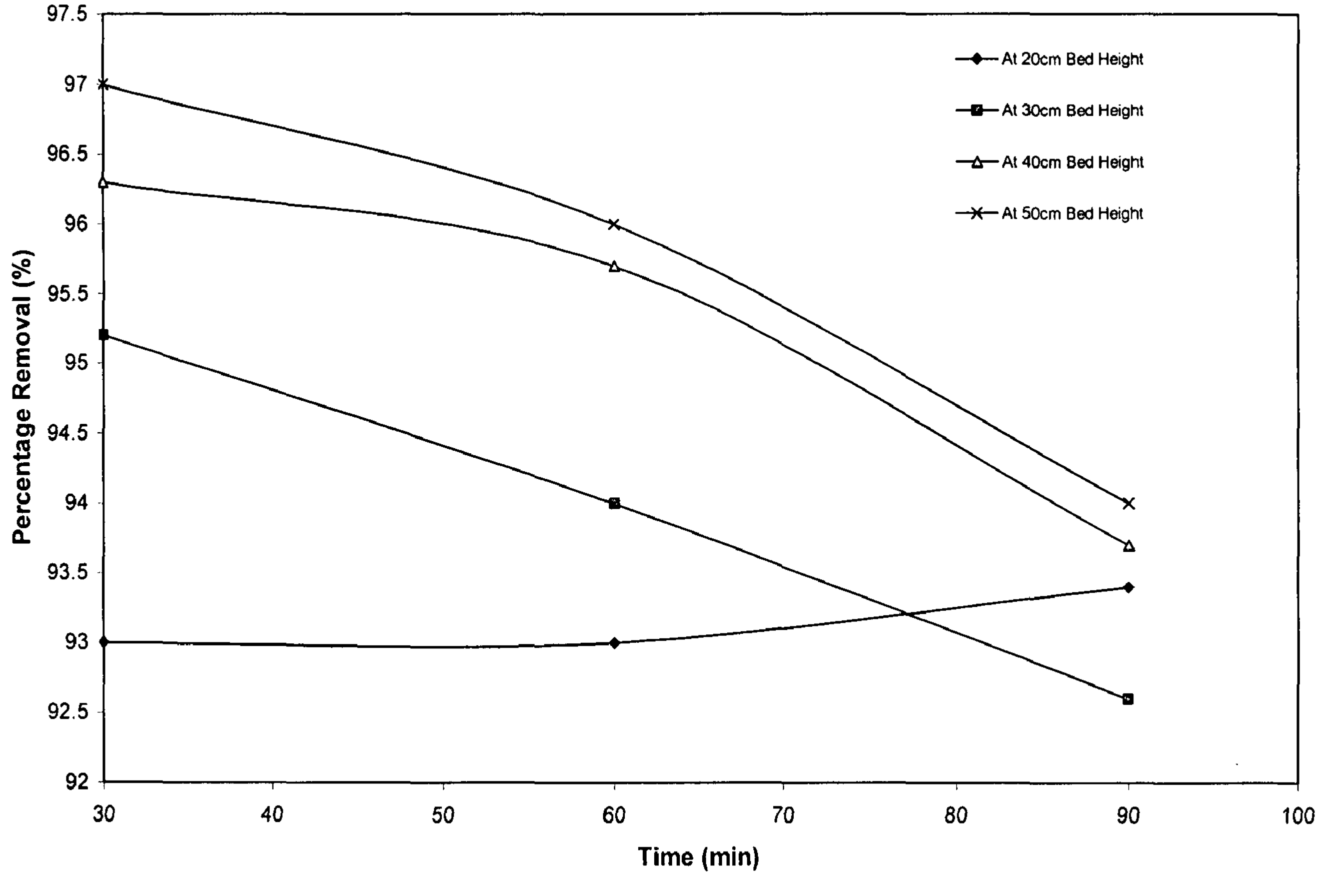


Figure-5.14 Removal of Fine Suspended Particle for Anthracite as a Media

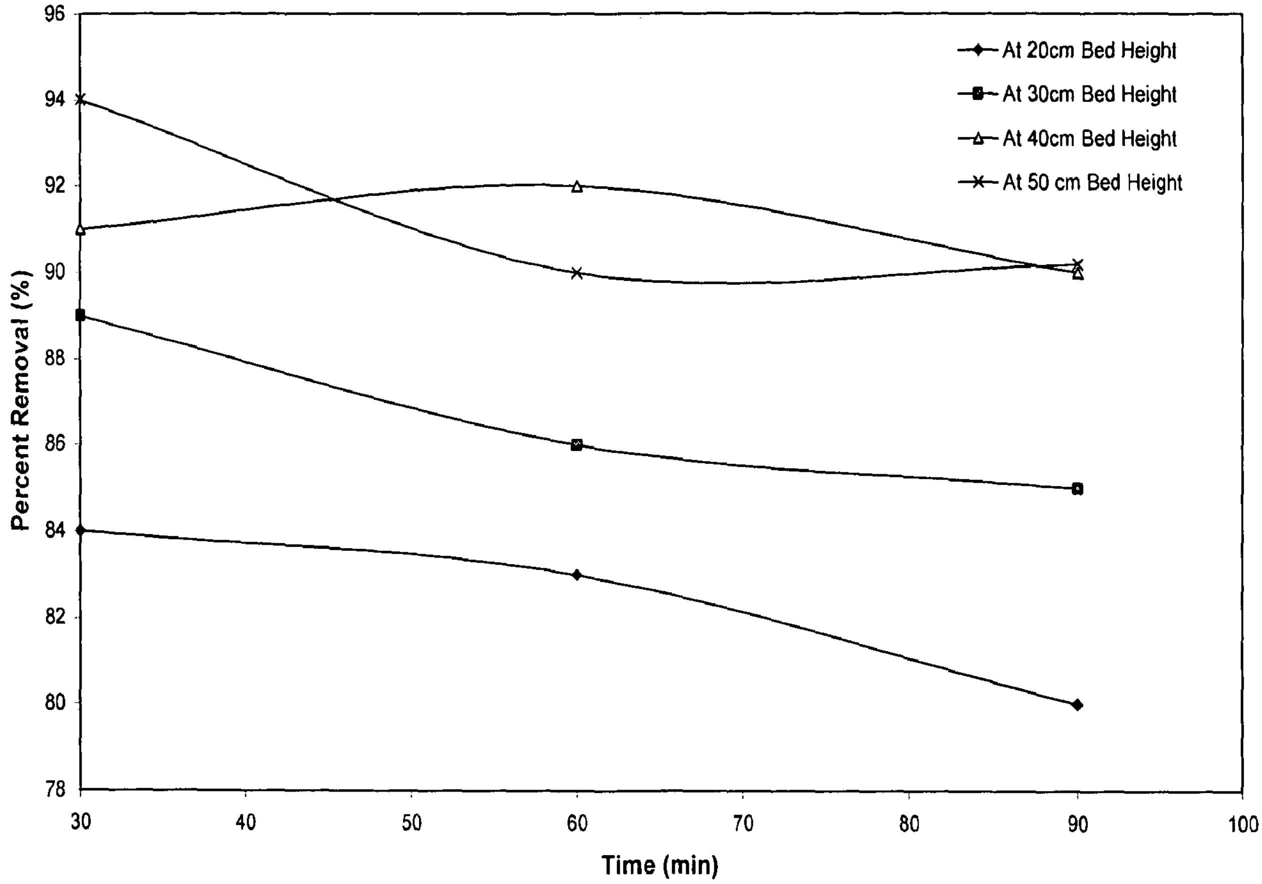


Figure-5.15 Removal of Fine Suspeded Particle for Garnet as a Media

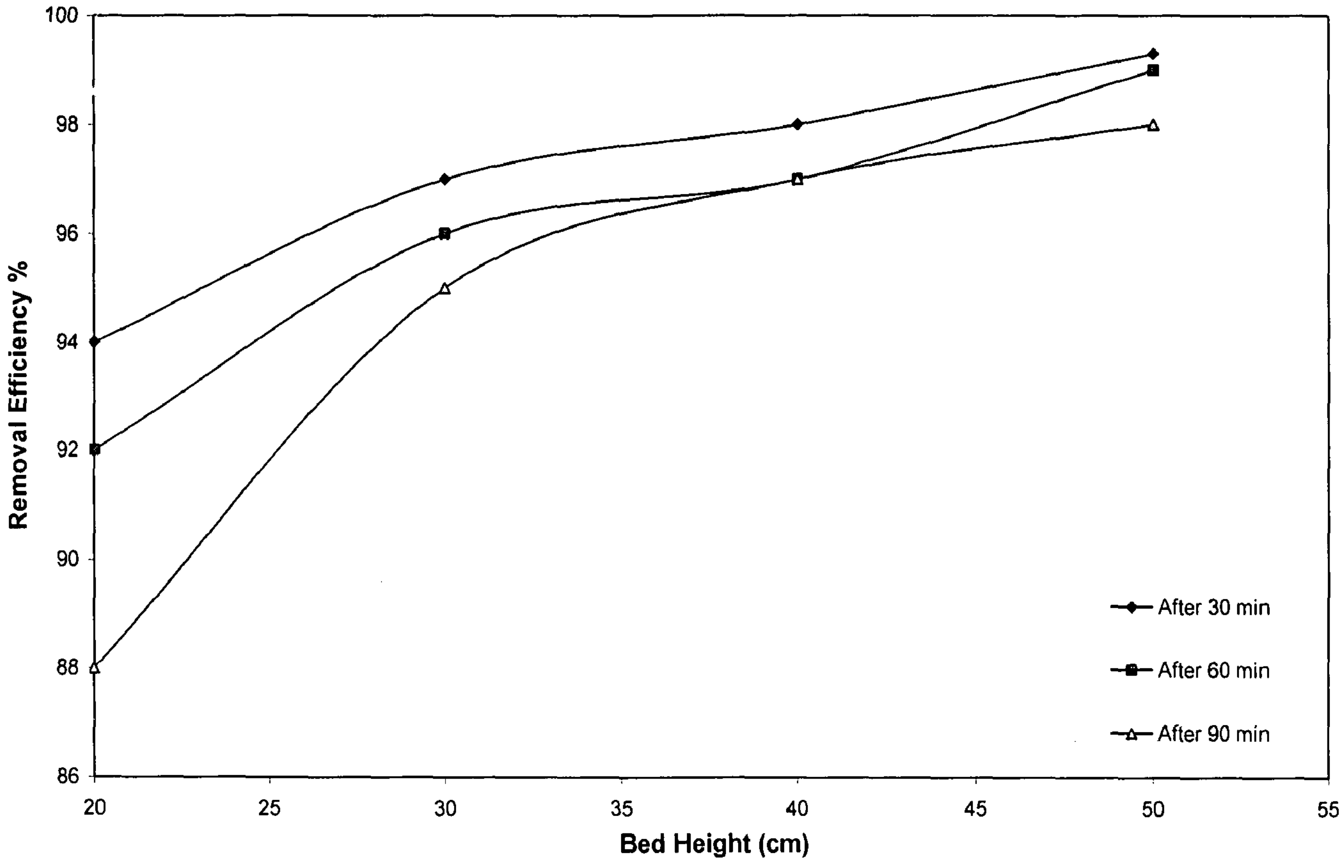


Figure-5.16 Removal of Fine Suspended Particle at different Bed Height of Calcite

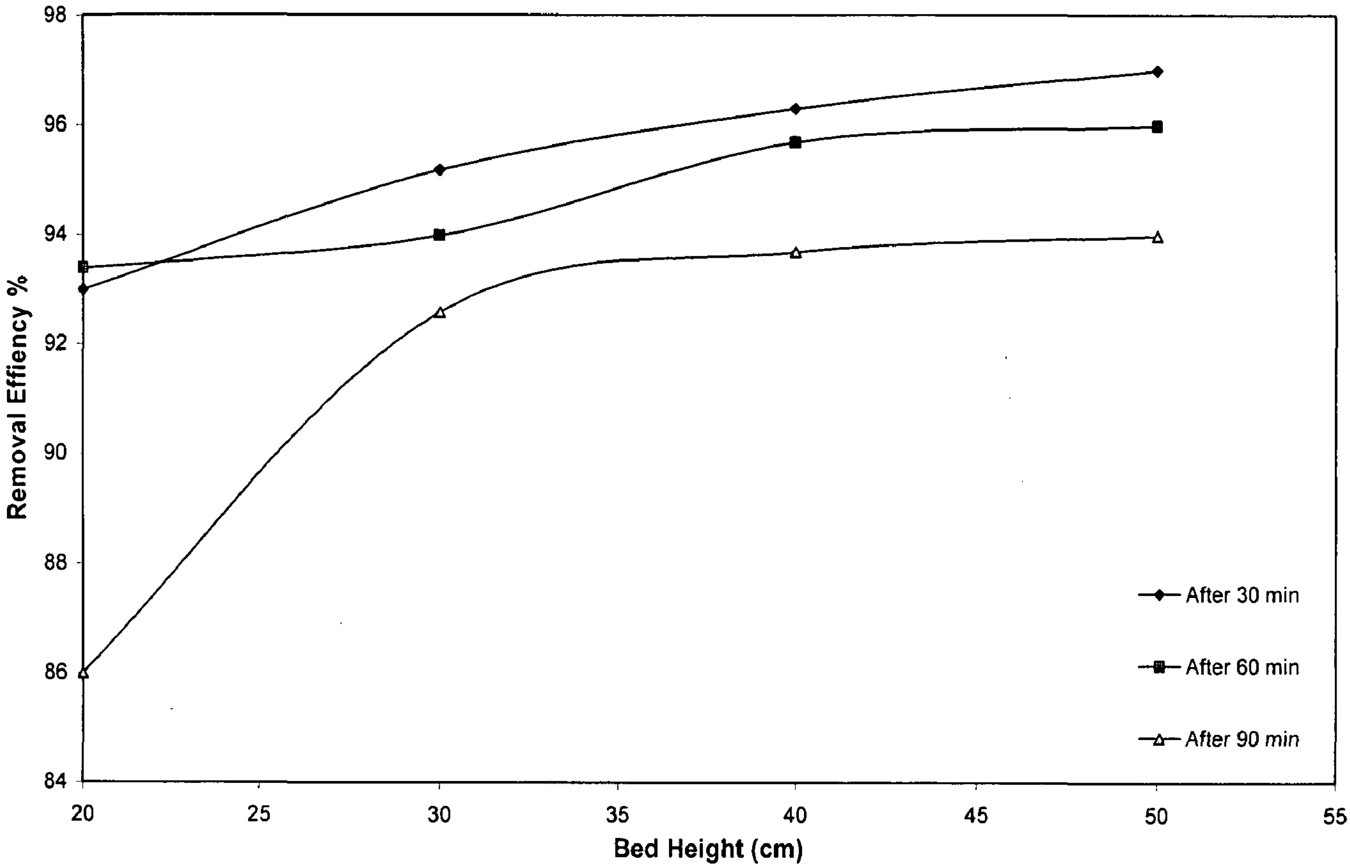
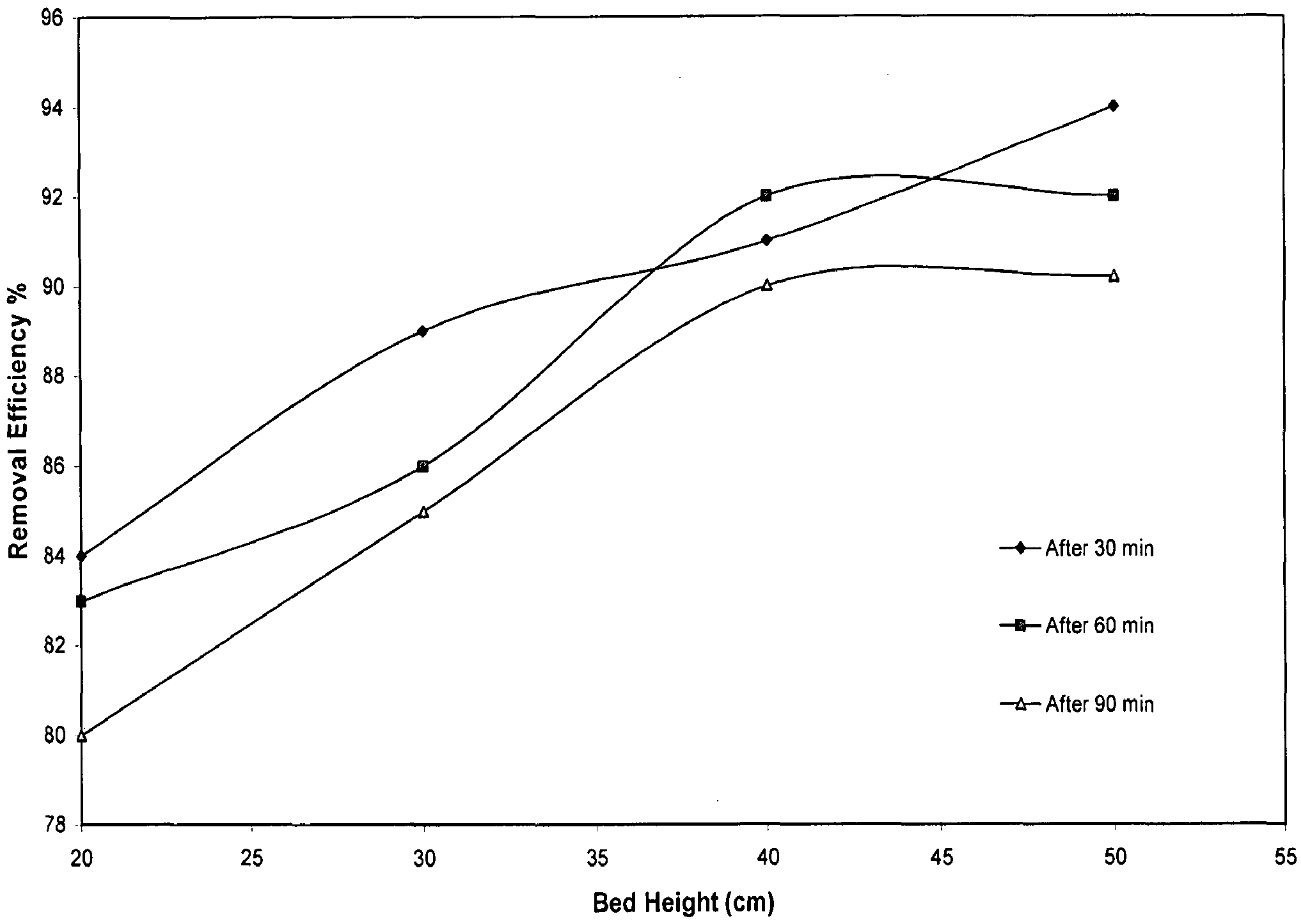


Figure-5.17 Removal of Fine Suspended Particle at different Bed Height of Anthracite



Figuer-5.18 Removal of Fine Suspended Particle at different Bed Height of Garnet

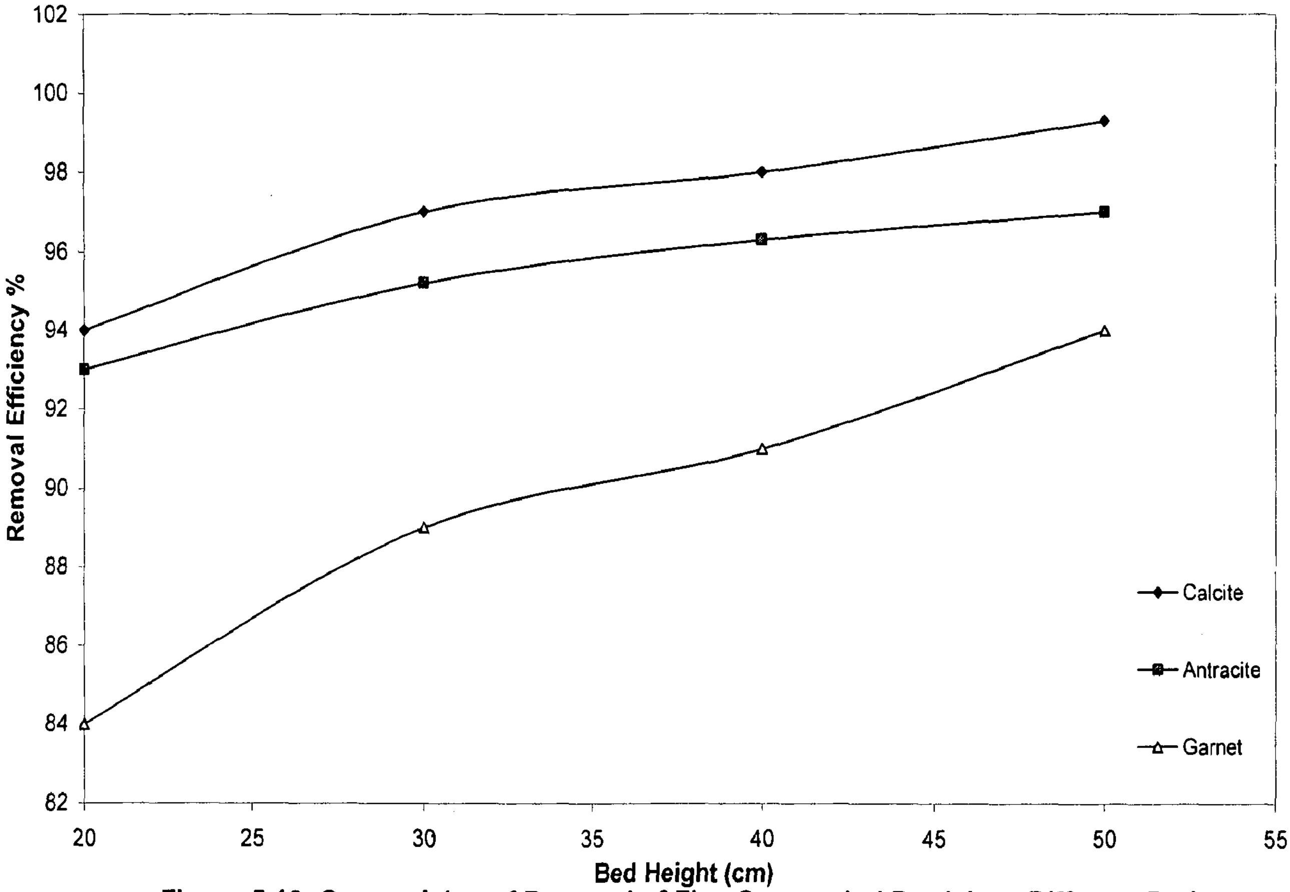


Figure-5.19 Comparision of Removal of Fine Suspended Particle at Different Bed Height of Calcite, Anthracite and Garnet after 30 min

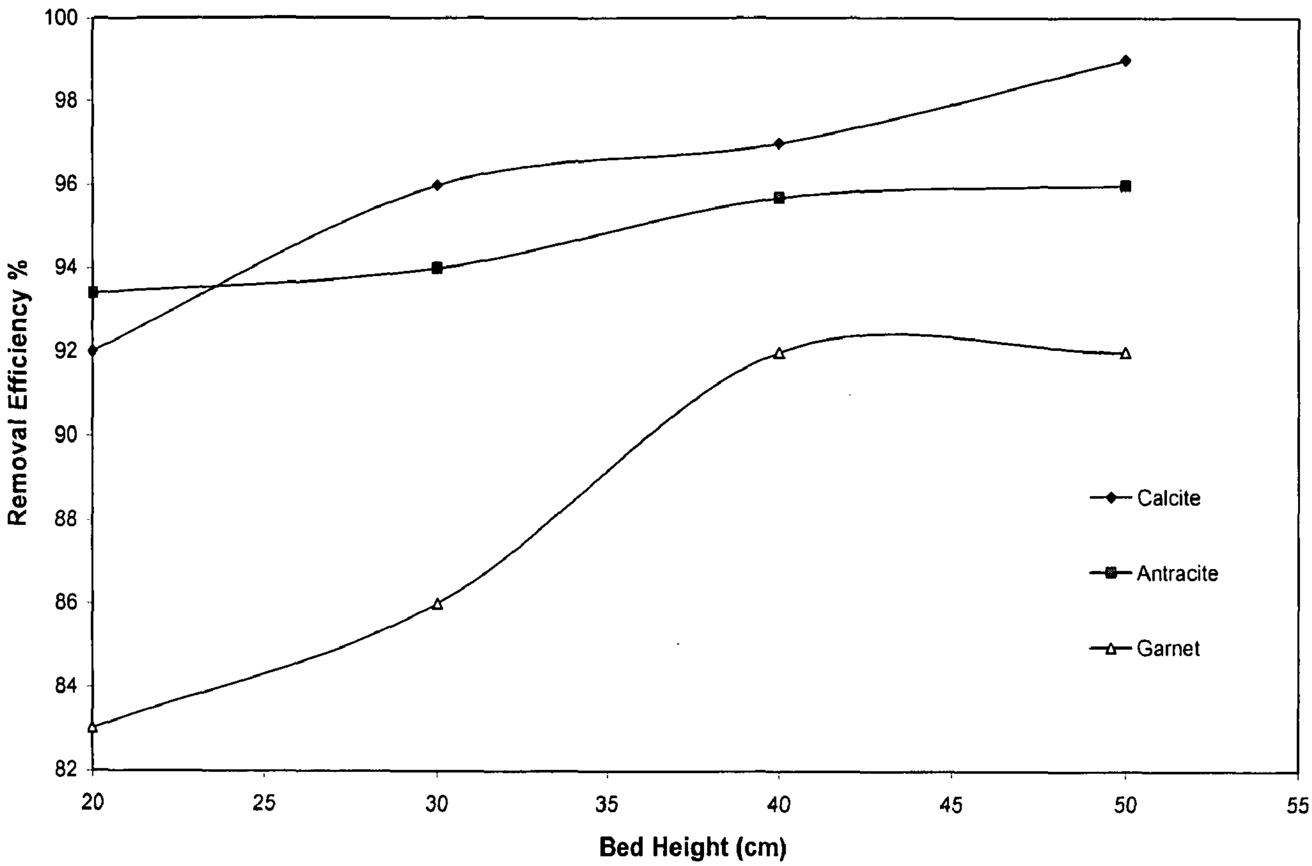


Figure-5.20 Comparision of Removal of Fine Suspended Particle at Different Bed Height of Calcite, Anthracite and Garnet after 60 min

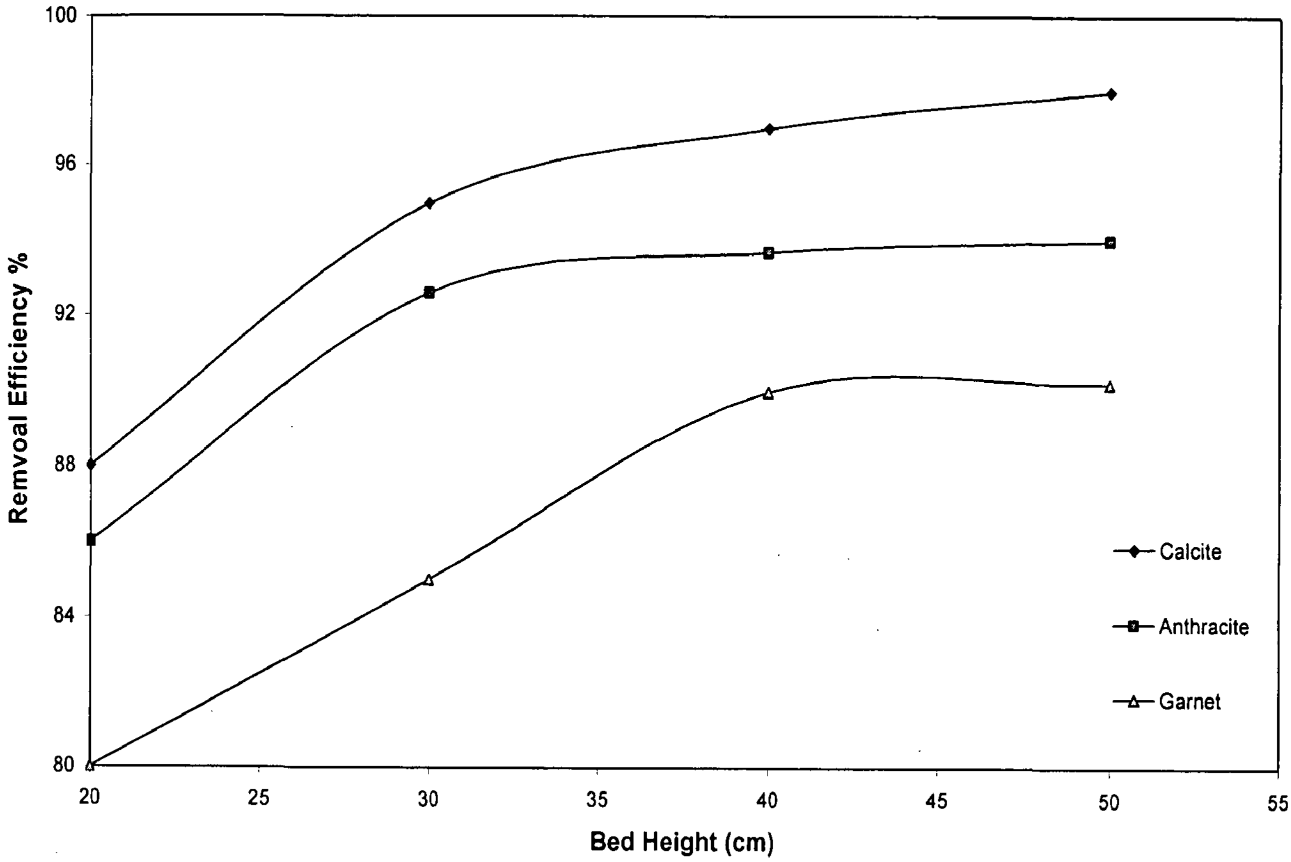


Figure-5.21 Comparision of Removal of Fine Suspended Particle at Different Bed Height of Calcite, Anthracite and Garnet after 90 min

Chapter-6

CONCLUSIONS AND RECOMMENDATIONS

6.1 CONCLUSIONS

Under the operation condition used in the experiment and from the result obtained, the following conclusion 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 i.e. for Anthracite at 500 mm bed height the removal efficiency decreases from 97% to 94 % when the sample of filtrate analyzes after 30 min and 90 min.
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 from 94% to 99.3% for Calcite and 93% to 97% for Anthracite.
4. From the observed result the removal efficiency for Calcite is higher than the Anthracite and Garnet in term of the removal of 2-20 μm particles.

6.2 RECOMMENDATIONS

The scope of the present study is limited to the study about removal efficiency of the fine suspended particle using single media as a filter medium at varied bed height of media. As the bed height increases for filtration operation the pressure drop also increases. Furthermore the experiment is studied at designed flow rate while the flow rate changes the removal efficiency also changes. So the optimization among the flow rate , pressure drop and removal efficiency is required. Further study can be done using multimedia filter bed for the removal of fine suspended particle.

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APPENDIX-A

Table A: Undersize Number Percentage (%) for Calcite

For 20cm Bed Height of Calcite						30cm Bed Height of Calcite					
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	2.53	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
				100	12			99	4.73	99	5.1
								100	9	100	7

For 40cm Bed Height of Calcite						For 50cm Bed Height of Calcite					
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		

APPENDIX-B

Table B: Undersize Number Percentage (%) for Anthracite

For 20cm Bed Height of Anthracite						For 30cm Bed Height of Anthracite					
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 40cm Bed Height of Anthracite						For 50cm Bed Height of Anthracite					
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

APPENDIX-C

Table C: Undersize Number Percentage (%) for Garnet

For 20cm Bed Height of Gar net						For 30cm Bed Height of Gar net					
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
40cm Bed Height of Anthracite						50cm Bed Height of Anthracite					
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										