# Investigation of oily wastewater treatment processes.

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# Investigation of Oily Wastewater Treatment Processes

by

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## Dedication

**In loving memory of my great mother, Mrs Livinah Okam.**
**I love you mama.**
**May 29, 2007**

This work is dedicated to those who have dreams and who hear a voice which says they can't achieve what they want to but they go on to achieve their dreams despite all odds.

# Table of Contents

**Appendices**

# List of Tables

# List of Figures

# Summary of Abbreviations

| | |
|---|---|
| API | American Petroleum Institute |
| BOD | Biological Oxygen Demand |
| COD | Chemical Oxygen Demand |
| CPI | Corrugated Plate Interceptor |
| DO | Dissolved Oxygen |
| ICB | Immobilized Cell Bioreactor |
| MBR | Membrane Bioreactor |
| MTC | Mass Transfer Coefficient |
| O&M | Operation and Maintenance |
| PFCs | Perfluorocarbons |
| UF | Ultrafiltration |

## Nomenclature

| | |
|---|---|
| $a$ | Specific interfacial area ($m^2/m^3$) |
| $A$ | Interfacial area normal to mass transfer ($m^2$) |
| $A_P$ | Projected plate area of a plate ($m^2$) |
| $A_s$ | Surface area required for separation ($m^2$) |
| $BP$ | Brake Power (kW) |
| C | Concentration of oxygen in the bulk liquid phase (% saturation) |
| $C_c$ | Cost constant |
| $C^*$ | The saturation value of DO at the gas-liquid interface (% saturation) |
| $C_e$ | Purchased cost (£) |
| $C_L$ | Dissolved oxygen concentration in liquid (mg/l) |
| $C_s$ | Saturation dissolved oxygen concentration (mg/l) |
| $d$ | Oil droplet diameter (m) |
| $D$ | Diffusion coefficient of the oxygen in liquid ($m^2/s$) |
| $F_B$ | Buoyancy force in Newton (N) |
| $F_d$ | Drag force in Newton (N) |
| $F_{dp}$ | Design factor for corrugated plate interceptor (%) |
| g | Acceleration due to gravity ($m/s^2$) |
| $h_o$ | Height travelled by any droplet (m) |
| $H_s$ | Separator depth (m) |
| $V_{ICB}$ | Volume of Immobilized Cell Bioreactor (l) |
| $k_L$ | Mass transfer coefficient in the liquid phase (m/s) |
| $k_{La}$ | Volumetric liquid phase mass transfer coefficient ($s^{-1}$) |
| $l_{dp}$ | Distance between plates (m) |
| $L_P$ | Plate length (m) |

| | |
|---|---|
| $L_s$ | Separator length (m) |
| $m$ | Mass of oxygen (mg) |
| $n$ | Number of droplets in a size range or interval |
| $N_P$ | Number of plates required to achieve separation |
| $OD$ | Oxygen demand (mg/s) |
| $O_R$ | Overflow rate (velocity of rise of the waste stream) (m/s) |
| $OT$ | Oxygen actually transferred/air volume (kg/m$^3$) |
| $OTR$ | Oxygen transfer rate (mg/l.s) |
| $P$ | Blower discharge pressure (psig) |
| $Q$ | Volumetric flow rate (m$^3$/s) |
| $Q_A$ | Volumetric flow rate of air (m$^3$/s) |
| $R_1$ | Ratio of horizontal flow velocity ($v_h$) to rise velocity of critical droplet ($v_t$) |
| $R_2$ | Ratio of separator depth ($H_s$) to width ($w_s$) |
| $s$ | Average frequency at which any particular vertical element is mixed or rate of surface renewal (s$^{-1}$) |
| $S$ | Characteristic size parameter in units given in Appendix I |
| $t_t$ | Rise time of critical droplet (s) |
| $t_w$ | Residence time of oil water mixture in separator (s) |
| $v_h$ | Horizontal velocity (m/s) |
| $V_{ICB}$ | Volume of Immobilized Cell Bioreactor (m$^3$) |
| $v_o$ | Terminal velocity of oil droplet or oil droplet velocity (m/s) |
| $V_{oil}$ | Volume of oil in a droplet (m$^3$) |
| $v_p$ | Flow velocity along plates (m/s) |
| $v_t$ | Rise rate of the design droplet or critical droplet (m/s) |
| $V_t$ | Total volume of oil contributed by a droplet of a particular size (m$^3$) |
| $w_P$ | Plate width (m) |

| | |
|---|---|
| $w_s$ | Separator width (m) |
| $x$ | Droplet size of interest (μm) |
| $x_m$ | Mean droplet size (μm) |
| $X$ | Fraction of droplets of a particular size recovered |
| $Y$ | Fraction of droplets of a particle size not recovered or captured |
| $\theta$ | The time of exposure of the liquid element (s) |
| $\alpha$ | Angle of inclination of the plates to the horizontal (degrees) |
| $\delta$ | Film thickness (m) |
| $\mu$ | Dynamic viscosity of the continuous phase (Pa.s) |
| $\mu_w$ | Water viscosity (Pa.s) |
| $\eta$ | Mechanical efficiency (%) |
| $\rho_o$ | Density of the oil (kg/m$^3$) |
| $\rho_w$ | Density of water (kg/m$^3$) |
| $\sigma$ | Standard deviation from the mean |

# Abstract

With the ever-increasing price of oil and increasingly stringent environmental regulations, effective treatment of oily wastewater has become very important. Industry currently seeks to find cost-effective solutions to meeting discharge limits by integration of oil recovery units with biological treatment.

Two integrated processes were investigated so as to minimize the cost of treatment: (1) integration of a horizontal flow separator and an Immobilized Cell Bioreactor (ICB), and (2) integration of a Corrugated Plate Interceptor (CPI) and an ICB.

Theoretical models of each process were developed to investigate and compare the factors that affect oil recovery and the integrated process. These models were used to generate oil droplet size distributions, size and predict the performance of the separators, size the bioreactor, and estimate the cost of the integrated process. Experimental work was undertaken to investigate aeration in a laboratory-scale model of an ICB using support particles (Pall rings) of various sizes (16 mm, 25 mm, 38 mm, and 50 mm).

For the first process reducing the horizontal flow velocities best improved the separation efficiency and hence overall cost of treatment. A 33% change in droplet size resulted in an increase in separator efficiency of 55%, whilst a 33% change in horizontal velocity resulted in an increase of 68%. For the second process the effect of number of plates and plate dimensions, was investigated. For a 25% increase in design variable, the number of plates increased separator volume by 2%, plate length by 43% and plate width by 17%. Increasing the number of plates to improve interceptor efficiency resulted in the most cost-effective overall process.

The experimental analysis found that an intermediate size of support particle (1½") resulted in the best volumetric mass transfer coefficient ($k_{La}$). Aeration rate in the ICB column was varied between 9 and 56 ml/s. Using 38 mm Pall rings, $k_{La}$ increased from 0.0030 to 0.0097 $s^{-1}$; using 50 mm Pall rings, $k_{La}$ increased from 0.0015 to 0.0058 $s^{-1}$.

This work concludes that enhancing oil recovery is a matter of separator design, rather than manipulation of wastewater characteristics as achieved by the use of chemicals. Effective aeration, and therefore size, of an ICB is dependent on the characteristics of the support particles used. The solution to cost-effective integration lies in the proper design of the units involved.

# Chapter 1: Introduction

## 1.1 Introduction

This chapter introduces the research and its primary motivations. It presents the research problem, justifies why the research is important, and presents the scope of the research.

## 1.2 Background

With the increasing value of oil and increasingly stringent environmental regulations, finding cost effective solutions for oily wastewater treatment has become very important.

An estimated US$40 billion is spent worldwide in treatment of oily wastewater produced in the upstream oil and gas industry (http://www.torrcanada.com, 2007); the large capital outlay, increasing value of oil and environmental regulation necessitates the need for process integration in oily wastewater treatment. Recovered oil could be put back into the process, thereby reducing energy cost, or can be sold. Using the recovered oil for any of these purposes ultimately reduces the overall cost of treatment. Economic factors aside, the treated water needs to be suitable for discharge to the environment.

Several technologies have been developed for oily wastewater treatment such as horizontal flow separators, Corrugated Plate Interceptors (CPI), Dissolved Air Flotation (DAF) units, hydrocyclones, bioreactors, and the membrane processes, which are the most recent advancement in oily wastewater treatment (Cheryan and Rajagopalan 1998, Shui Li *et al.* 2006). To meet current discharge limits it has become necessary to integrate two or more of these units since all treatment technologies are limited in application (Tin and Cheng 1994, Mueller *et al.* 1997).

In the industry, the practice is to integrate oil recovery units and biological treatment units. This reduces the cost of treatment by recovering oil which can be sold to offset the cost of treatment associated with aerobically operated biological units.

Reducing the cost of aeration is paramount because it is the major cost in operating aerobic bioreactors (Atkinson 1986). The membrane bioreactor, another recent advancement in oily wastewater technology, was developed to reduce the cost of aeration but this treatment technology has not attained general commercial success due to the high cost of replacing membranes. The Immobilized Cell Bioreactor (ICB) is a relatively new technology in oily wastewater treatment. In this technology, cells are immobilized onto support particles; these support particles serve two purposes: they ensure cells are not washed away, therefore improving the bioreactor effectiveness and reducing cost of operation, and they also enhance oxygen transfer and further reduce cost due to aeration. Application of this technology in oily wastewater treatment is becoming popular (Park *et al.* 1998, Zhao *et al.* 2006 and Martin *et al.* 2000). The performance of the ICB unit in terms of oxygen transfer then depends on the physical components in the ICB unit such as spargers, sparger design, sparger location, and support particle type and size. To further reduce the cost of aeration, investigation of aeration in the various designs of ICB unit becomes necessary.

The two most widely used oil recovery units in the industry are the horizontal flow separator and the Corrugated Plate Interceptor. To optimize oil recovery and reduce the cost of an integrated process, it becomes necessary to investigate the effectiveness of these units to recover oil.

This project therefore sets out to investigate the treatment of oily wastewater by the effective integration of oil recovery units and biological treatment so as to maximize the cost effectiveness of the overall process.

## 1.3 The Research problem

Figure 1.1 outlines a simple integrated process. The aerobic treatment is intimately linked to the use of oil water separation process(es) to recover oil from the system. As more oil is recovered from the system, the less intense the aerobic treatment has to be. In fact the cost of the system can be significantly improved by a good oil-water separation as recovered oil can be sold; also the bioreactor becomes smaller and the output cleaner.

However, the ICB unit downstream of the oil recovery units contributes to the cost of the process. The cost of biological treatment with the ICB unit varies with oil concentration. Increasing oil concentration implies increasing cost of treatment for an aerobically operated ICB unit. Approximately 50% of energy cost in the bioprocess goes to aeration. This implies aeration is a key factor in improving cost of biological treatment (eutchinst.com).

Therefore to improve the efficiency and cost of the integrated process, the effectiveness of the oil recovery units to recover oil will be investigated and aeration in ICB units will be investigated.

This project therefore sets out to investigate two integrated processes:

1. The integration of a horizontal flow separator and an ICB.
2. The integration of a Corrugated Plate Interceptor (CPI) and an ICB.

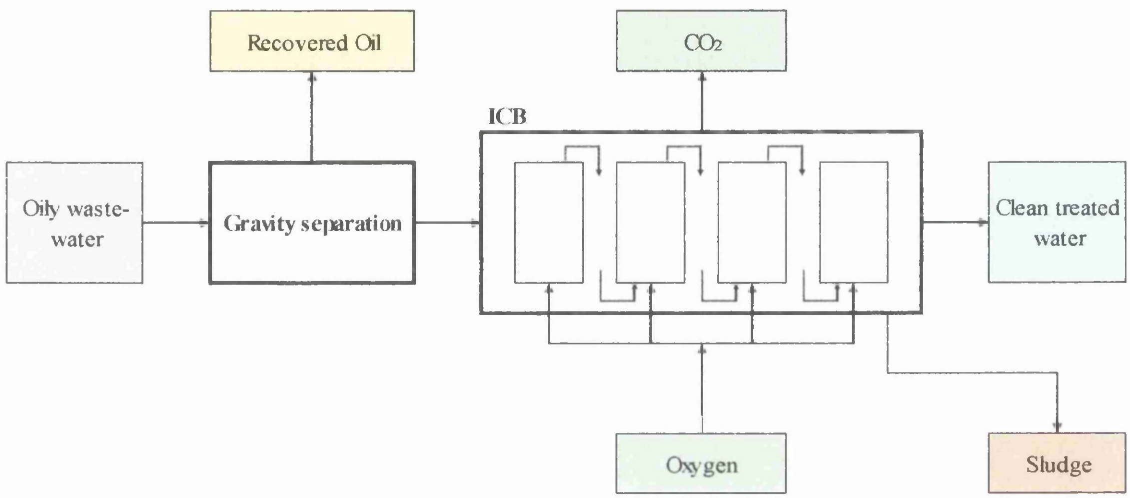


**Figure 1.1: Integrated system of gravity separation and ICB**

## 1.4 Research Methodology

Two methods were used in this research to obtain results: theoretical and experimental.

The theoretical aspect involved creating scenarios to investigate the effectiveness of the oil recovery units and their effect on the integrated process. It also involved developing spreadsheet models to generate theoretical oil droplet distributions in various oily wastewater streams, to design separators and predict their performance, and to size the Immobilized Cell Bioreactors required to treat the remaining oil in the effluent of the oil recovery units. Models were also developed in Excel to calculate the blower power required to meet the various oxygen demands of the ICB unit.

The experimental aspect involved a practical investigation of aeration in a laboratory scale Immobilized Cell Bioreactor (ICB) unit. Different ICB unit designs were investigated by using various sizes of Plastic Pall rings (support particles). The mass transfer characteristics of the various sizes of support particle were investigated to predict oxygen mass transfer in the various designs of ICB. Data obtained from the experimental investigation was used as part of the theoretical investigation (refer to Chapter 3).

For the costing of the various units, cost estimation models were employed (refer to Chapter 2). Calculation procedures for the various models developed are presented in Chapter 3.

## 1.5 Justification for research

- Considering the large number of industries utilizing oil and oil related products, there is a vast amount of revenue that can be generated if oil treatment units can be effectively optimized to recover more oil and treat unrecovered oil at minimum cost.
- A study of this nature may necessitate a change in the current practice of using chemicals to enhance separator effectiveness, a practice which adds cost to treatment and affects the environment.
- To serve as a guide for profitable integration of units for oil water separation and treatment and hence reduce cost of treatment.
- Most reported works (Moosai and Dawe 2003, Al-Shamrani *et al.* 2002, Lundh *et al.* 2002) on oily wastewater treatment methods have focused on investigating the factors that affect the effectiveness of each treatment or separation method studied. A smaller proportion has reported the benefits of integrating units (Karakulski *et al.* 1995, Zhong *et al.* 2003, Gryta *et al.* 2001) but no study has yet outlined the window for cost-effective integration of oily wastewater separation and treatment processes.

## 1.6 Research presentation

This thesis is presented in six chapters. The literature review in Chapter 2 justifies the investigation and provides the background information for the remaining chapters. The first section of Chapter 3 describes the materials and methods used for the experimental and aspect of the research and the second section explains the spreadsheet calculation procedures used for the theoretical aspect of the research. Chapter 4 presents and discusses the results from the experimental investigation of aeration in various models of

ICB units. Chapter 5 presents the results from the theoretical investigations; it reports the effectiveness of the oil recovery units in various scenarios and the effects on the integrated process. Chapter 6 reports the conclusions and recommendations for further work.

## 1.7 Scope of the research and key assumptions

The scope of this research is limited to the study of the two most common oil recovery units: the horizontal flow separator and the Corrugated Plate Interceptor. Dissolved Air Flotation, Induced Gas Flotation, and Hydrocyclones are other oil water separation technologies that were not studied because they are not widely used in the industry.

A theoretical waste water stream was used in this research. Though an actual oily wastewater stream could have been characterized to use as a sample, this was not done mostly because the method of using a microscope to count the number of droplets with a particular size is prone to errors. Nevertheless, an actual oily wastewater stream may have the same distribution as a theoretical stream.

Oil droplets are not solid particles but for this analysis they were assumed to be spherically shaped so as to determine the volume of oil contained in each oil droplet. The cost of treating sludge was not taken into account during costing because actual oily waste streams were not used. However, it is important to note that the amount of sludge generated would vary with the source of the oily waste stream.

# Chapter 2: Literature Review

## 2.1 Introduction

This chapter is divided into two broad sections. The first section (2.2) justifies the investigation and the second section (2.3) provides the background information for the rest of the thesis.

## 2.2 Research justification

### 2.2.1 Sources of oily wastewater and quantity

The volume of oily wastewater (oil-water mixture) generated continuously around the world is very large. This is due to the extensive and continuous utilization of oil and oil related products in most industries.

Major industrial sources of oily wastewater (Smith 1989, Villas-Bôas and Barreto 1996, Warhurst and Bridge 1996, Busca 2004) include oil fields, petroleum refineries, metal manufacturing and machining, minerals and mining. Oil fields and petroleum refineries produce large quantities of oil-water mixtures. In the metal industries, two major sources of oily water mixtures (oily wastewater) are steel manufacture and metal working (Wolfe 1992). Metal manufacturing and machining plants generally utilize machining coolants in their metal cutting, forming, rolling, and finishing operations (Mysore *et al.* 2005). These coolants often consist of chemically stabilized emulsions of oil in water with an oil concentration of 2-5%. Weintraub *et al.* 1983 reported that wastewater discharged from industrial washers may have an emulsified oil concentration of 300-7000 mg/l and free floating oil content of 30,000 mg/l. Nemerow 1978 reported that the concentration of oil in effluents from different industrial sources is found to vary widely from several mg/l to as high as 40,000 mg/l.

These large volumes of oily wastewater call for proper treatment before disposal. Also the amount of revenue which can be generated, or expense saved, from recovering these oils calls for a cost effective method of treatment which not only improves the efficiency of the units but also improves the cost of the process, decreases the number of stages involved and has a small footprint.

### 2.2.2 Review of oily wastewater separation and treatment units

Various technologies have been developed for oil recovery from oily wastewater, including membrane processes, flotation processes, and hydrocyclones.

The membrane process is the most recent advancement in oil recovery technologies and is a pressure driven process which relies on the pore size of the membrane to separate the oil-water stream into the two streams: the concentrate and the permeate. The concentrate contains the oil phase while the permeate contains the purified water. The flotation process is a concept widely used in the mineral processing industries (Kitchener 1985) but has also been applied for oily wastewater separation. The philosophy behind the flotation processes in the oily wastewater separation application is to increase the density difference between the oil and water in order to increase the rise rate of the oil droplets. The increase in density difference is achieved by attaching gas bubbles to the oil droplets. The hydrocyclone is a unit which operates on the principle that if the natural gravitational field is increased, separation is enhanced and the time required for separation is reduced.

However, these units have not gained general commercial success in the industry. Even though the membrane process is the most effective and has a small footprint, it has not gained general commercial success due to the high cost of replacing membranes (Cheryan and Rajagopalan 1998, Shui Li *et al.* 2006). The membranes form a synthetic barrier which allows the transport of water (the permeate) through the membrane whilst oil (the concentrate) accumulates on the membrane surface. As a result of oil droplets accumulating on the membrane, the permeating flux decreases with time as more oil accumulates on the membrane surface causing a relatively low flux limit (Li, Yu Shui *et al.* 2006). This phenomenon, which is known as concentration polarization and membrane fouling, has a devastating effect on the efficiency of the membrane process and limits the application of the membrane process in the industry due to the high cost of replacing membranes. Current research on membrane technology is based on tackling fouling (Li, Yu Shui *et al.* 2006, Li *et al.* 2006, and Wang *et al.* 2006). Several approaches to mitigate this problem have been attempted; Li *et al.* 2006 developed membranes to tackle fouling, Kim *et al.* 2007 reported reduced fouling when an electric field was used in the membrane process. Zhong *et al.* 2003 reported reduced fouling when membranes were coupled with flocculation. Srijaroonrat *et al.* 1999 and Ma *et al.* 2000 reported reduced membrane fouling with back pulsing. Kobayashi, *et al.* 2003 investigated the use of ultrasound for cleaning fouled membranes to increase membrane life. Karakulski *et al.* 1995 reported increased membrane efficiency by coupling membranes. At present there is no outstanding solution to mitigating the effects of membrane fouling, hence the limited use of membranes.

Much effort has been made to improve the efficiency of the flotation process. This has resulted in the development of various techniques to introduce gas to the oil wastewater systems. The most common techniques are Induced Gas Flotation (IGF) and Dissolved Air Flotation (DAF). The major difference between these flotation techniques is the manner in which the gas or air is introduced. However, the flotation process is not common in the industry for many reasons. Smaller air bubbles are required to improve the efficiency of the flotation process, but since smaller bubbles have a low rate of rise, they necessitate the construction of a large tank, and the industry is interested in maintaining a small footprint. Other issues which may have limited the application of flotation processes are the many factors which affect its effectiveness. Such factors are the ability of the gas bubbles to collect oil droplets from the water, the size of the gas bubbles (Moosai and Dawe 2003), the saturation pressure of the unit, the use of chemicals upstream of the flotation process (Al-Shamrani *et al.* 2002), contact zone configuration in dissolved air flotation (Lundh *et al.* 2002), type of valve (Bratby and Marais 1975), fluid dynamics (Maddock and Timlinson 1980), nozzle geometry (Takahashi *et al.* 1979), and concentration of air bubbles (Al-Shamrani *et al.* 2002). The process seems to be too complicated and its operation remains a black box.

Also there seem to be conflicting reports on how to improve the effectiveness of these processes. Al-Shamrani *et al.* 2002 reported improved performance of dissolved air flotation when integrated with chemicals; Zouboulis and Avranas 2000 reported that not all chemicals (coagulants) are effective in the dissolved air flotation process.

Another factor which may have limited the use of the flotation process may be the requirement for additional units upstream and downstream of the DAF units since the flotation process cannot recover droplets smaller than 20 microns or larger than 150 microns (Moosai and Dawe 2003).

The major limitation of the hydrocyclone unit in the industry is that it requires considerable pumping power, which involves significant cost (Tin and Cheng 1994). Hashmi *et al.* 2004 reported the various factors which affect the efficiency of hydrocyclones as orifice size, feed pressure, oil concentration, and conditioning of the water. They concluded that hydrocyclones cannot operate effectively with high concentrations of oil. The inability to handle large oil concentrations may also have resulted in hydrocyclones not being common in the industry. The technology is reported to be efficient in recovering droplets in the size range of 10-30 microns (Tin and Cheng 1994).

The most commonly used oil recovery units are the horizontal flow separator and the Corrugated Plate Interceptor.

The horizontal flow separator is the oldest technology developed and applied in the separation of oil from water; it is more effective in recovering bigger sized oil droplets than smaller sized oil droplets. In the past, the philosophy of design was to oversize these separation units, accepting a large footprint in order to ensure capture of the smaller oil droplets. This practice was adequate until rapid industrial developments and new environmental regulations emerged; the horizontal flow separator then became incapable of meeting more stringent discharge limits whilst maintaining an acceptable footprint (Tin and Cheng 1994; Mueller *et al.* 1997).

The difficulty associated with recovery of smaller sized oil droplets led to the use of chemicals in gravity separators to enhance oil recovery. This was desirable because it meant there was no need to increase separator size in order to improve separator efficiency. Other technologies, such as the Corrugated Plate Interceptor, were also developed due to the inability of horizontal flow separators to recover small oil droplets (Li, Yu Shui *et al.* 2006).

Smaller sized oil droplets pose problems to the industry. Smaller sized droplets are formed as a result of turbulent mixing of the oil and water phases, resulting in the oil droplets being uniformly dispersed in the water (Hong *et al.* 2003). When this mixing is done in the absence of chemicals the resulting mixture is called an unstable emulsion; it is unstable because oil has a natural repulsion with water and will break down through a slow process.

Unstable emulsions break down by the following mechanisms:

- Creaming,
- Flocculation, and
- Agglomeration.

Creaming is a gravitational separation process where the lighter phase oil droplets rise through the system with time; this rise is initiated by the density difference between the continuous and dispersed phases. The rate of creaming is governed by the size of the oil droplets. Creaming in itself does not destabilize an emulsion, but the high concentration of oil droplets in the creamed layer is what promotes interactions that lead to flocculation, aggregation, or coalescence (Capek 2004) and hence the total breakdown of an emulsion.

Flocculation occurs as a result of agitation or any other motion or chemical which causes two or more oil droplets to lump together. In flocculation each oil droplet retains its identity but looses its kinetic independence since the aggregate moves as a unit.

Agglomeration occurs as a result of flocculation. In this mechanism droplets loose their individual identity to form larger droplets which have faster rate of rise.

The problem posed by this unstable mixture is that it takes a long time for the oil droplets to rise to the surface. Consequently, the industry has resorted to the use of chemicals to enhance droplet rise rates.

Smaller sized droplets formed as a result of turbulent mixing in the presence of surfactants are called stabilized emulsions (Capek 2004). Surfactants stabilize the oil droplets by developing surfaces charges on them; this lowers the interfacial tension between water and oil and prevents the droplets from coalescing (Capek 2004). The overall effect is a very slow droplet rising time, such that a horizontal flow separator is unable to recover oil from these emulsions.

To recover these droplets, chemicals called coagulants and flocculants are employed (Beisinger *et al.* 1974, Gardner 1972, Lash and Kominek 1975). Coagulants act to overcome the effect of surfactants by destroying the protective action of the surfactants, i.e. by overcoming the repulsive effects of the electrical double layers to allow the smaller sized oil droplets to form larger droplets through coalescence; flocculants act to promote agglomeration of the destabilized droplets, therefore increasing the rise rates. This method of recovering oil is the norm in the industry but it poses additional problems to the environment and additional cost to the oily wastewater producers. It produces hazardous sludge which must be treated and requires further treatment of the effluent stream which all adds to the cost of treatment and the number of stages involved in treatment.

Heating of oil-water mixtures to reduce the water viscosity and hence increase the rise rate of oil droplets is another method used to reduce separator size. This thermal process involves considerable cost due to the high cost of energy (Gryta *et al.* 2001, Cheryan and Rajagopalan 1998).

The Corrugated Plate Interceptor was developed by the Royal Dutch Shell group in the 1950's (Tin and Cheng 1994). The design of this interceptor was based on two concepts:

1. If the distance the smaller droplets need to rise is reduced then the long residence time or large tank required by a horizontal flow separator can be avoided.
2. If the smaller droplets can increase in size then the rise rates will be increased (This is discussed in Section 2.3.1).

These conditions were achieved by introducing plates into the separation chamber (Iggleden 1978). The plates reduced the distance for a droplet to rise and also acted as a surface for droplets to coalesce on. This concept led to development of several types of plate separators (the various types of plate separators are discussed in Appendix I), all aimed at increasing efficiency. The physical process which takes place in Corrugated Plate Interceptors has been described by Rommel *et al.* 1992. They explained that when a dispersion is fed to a plate interceptor, the lighter phase ascends to the upper plate and forms a trickling film which flows along the plates following the hydrostatic pressure to the principal interface (that is between the lighter phase and heavy phase). Subsequent droplets coalesce on the trickling film. At this point of coalescence any surfactants present are swept away by the trickling film. This phenomenon eliminates the use of chemicals to improve recovery.

However, the horizontal flow separator and CPI are limited in application. They cannot effectively recover smaller sized oil droplets (less than 60 microns). The practice in the industry is to integrate these units with a biological process. This necessary is to meet environmental standards, but the cost of integration is high. Also, with the increasing value of oil, the industry seeks methods to improve oil recovery and reduce the cost of integrated processes.

To address the need to improve oil recovery and to reduce the cost of treatment, this project investigates the effectiveness of these oil separation units in various scenarios.

### 2.2.3 Biological reactors and the Immobilized Cell Bioreactor (ICB)

Biological reactors are most effective in treating unrecovered smaller sized oil droplets. The biological process actually entails the transformation of oil and other dissolved organics into biomass and gases ($CO_2$, $CH_4$, $N_2$, and $SO_2$) (Low and Chase 1999) but its efficiency depends on a number of factors.

One of the critical factors which affect the performance of any bioreactor is the concentration of microbes in the reactor. Different configurations of bioreactors have since been developed, all in an effort to improve the concentration of microbes and hence efficiency. The initial configuration involved suspending the microbes freely in the reactor. Many problems were associated with this configuration (conventional bioreactor) such as loss of microbes, the need to operate within a certain range of flow rates or dilution rates, short retention time and high sensitivity to fluctuations in flow, concentration and temperature, high cost of operation (Gray 1989), and high sludge production. The treatment of sludge accounts for 60% of the total cost of treatment plants (Wei *et al.* 2003). These limitations led to the development of various configurations for bioreactors, for example the fixed film technologies and membrane bioreactor. Table 2.1 compares the various bioreactor configurations; the major advantage of the submerged biological contactor over the rotating biological contactor is its compact design.

The membrane bioreactor (MBR) is the most recent innovation in bio-processing. It is an integrated system of membranes and bioreactor. Water passes through the membranes whilst oil is retained as the concentrate; any oil passing into the permeate is oxidized by microbes. The MBR is currently the only single treatment method which recovers valuable oil as well as degrading what remains to meet discharge limits; but it has some major disadvantages: membrane fouling and the high cost of membranes (Zhi-Guo *et al.* 2005, Wang *et al.* 2006). Whilst the biological and chemical oxygen demands of the process are reduced by the initial stage of oil recovery, high aeration rates are still needed to suppress fouling. Le-Clech *et al.* 2003 reviewed the literature and listed the various factors that suppress the onset of fouling in MBRs as increasing aeration rate and decreasing channel width. The problems with MBRs can then be said to be two-fold: first, the high cost associated with membrane replacement, and second, the high cost associated with aeration. These issues associated with membrane fouling have restricted the commercial utilization of MBRs (Zhi-Guo *et al.* 2005 and Li *et al.* 2005).

| Evaluation parameter | Trickling filter | Rotating biological contactor | Submerged biological contactor | Membrane bioreactor |
|---|---|---|---|---|
| Effective bioassay/toxicity control | | | | ✓ |
| Effective BOD removal efficiency | | | | ✓ |
| Effective COD removal efficiency | | | | ✓ |
| Low O&M | | ✓ | ✓ | ✓ |
| Low sludge production | ✓ | ✓ | ✓ | ✓ |
| Low sludge disposal cost | ✓ | ✓ | ✓ | ✓ |
| Minimum operator attention | | ✓ | ✓ | |
| Quick upset recovery | ✓ | ✓ | ✓ | |
| Energy efficient | ✓ | ✓ | ✓ | |
| Minimum space requirements | | | | ✓ |

**Table 2.1: Comparison of fixed film and membrane technologies (Schultz 2005)**

The Immobilized Cell Bioreactor (ICB) is another variation of the fixed film technologies. It offers a larger surface area for microbes compared to the other fixed film technologies and hence is more efficient; due to the presence of support particles it is also well mixed with respect to aeration - this allows for better mass transfer and control of biofilm thickness. Other advantages of the ICB are:

- Smaller reactor size due to immobilization of cells, which allows for higher degradation rates.
- Better protection against toxic or inhibitory compounds than free-cell systems due to the protective effect of concentration gradient that occurs within the biofilm.
- Operation at higher dilution rates without fear of washout and a much lower operating cost compared to the membrane bioreactor.

Immobilized Cell Bioreactors (ICB) are gaining ground in oily wastewater treatment (Park *et al.* 1998, Zhao *et al.* 2006 and Martin *et al.* 2000) but oxygen transfer remains crucial to their performance (Atkinson 1986). Many efforts have gone into overcoming the problem of oxygen transfer. Holst *et al.* (1982) reported generating oxygen in-situ by decomposing hydrogen peroxide using the high catalytic activity present in a micro-organism called *Glucunobacter oxydans*. Adlercreutz *et al.* 1982, Brodelius *et al.* 1981 and Szwajcer *et al.* 1982 reported generating oxygen by co-immobilizing oxygen producing micro-organisms and oxygen consuming organisms and co-immobilizing hydrogen peroxide degrading agent, such as activated carbon, and microbial cells. Elibol and Mavituna 1995 reported generating oxygen in immobilized cells using perfluorocarbons (PFCs) or silicon compounds. The problems associated with these methods are:

- Inherent toxicity associated with the carriers.
- Carriers are expensive e.g. Perfluorocarbons (Elibol and Mavituna 1995).
- There is no inherent mixing which is necessary for transfer of other nutrients.

All of these efforts directed to overcoming the problem of oxygen transfer are due to the high cost of aeration. If aeration problems can be eliminated then operation of bioreactors becomes less expensive and the cost of an integrated process involving an aerobically operated bioreactor becomes greatly reduced and cost of treatment improved. However, this has not been achieved so there is room for other solutions; one solution is to investigate other ways of reducing the cost of aeration.

This project therefore investigates aeration in various ICB units so as to reduce the cost of aeration.

A review of literature on oily wastewater treatment reveals that much study on oily wastewater has centred on developing models for design and for predicting separator performance. Nassif and Hansard (2003) developed a new approach to the design of oil water separators; they claimed Stokes' Law does not adequately describe the rise rate of a smaller sized droplet. Nakhla *et al.* 2006 developed a kinetic model for aerobic degradation of oil and Mohammadi *et al.* 2006 developed a model to predict membrane fouling. Though these studies have provided better understanding of these technologies they have failed to identify critical factors that optimize each unit so as to optimize the unit or an integrated process.

Other studies have been centred on integrated processes but these studies have only reported the benefits of integration. Zhong *et al.* 2003 reported improved performance of micro-filtration when coupled with chemical flocculation as a pre-treatment. Gryta *et al.* 2000 reported that the permeate obtained from an ultrafiltration (UF) process generally contains less than 5 ppm of oil, but when coupled with membrane distillation complete recovery of oil from water was achieved. Karakulski *et al.* 1995 investigated purification of oily wastewater using ultrafiltration and reported a permeate of acceptable quality was obtained for direct sewage discharge but reported further treatment (such as reverse osmosis) will permit the re-use of the water as process water. Although these studies are commendable they fail to provide a guide to the integration of oil recovery units and biological treatment units so as to minimize the overall cost of treatment.

### 2.2.4 Aims/Objectives

The industry is currently interested in maximizing treatment at minimum cost.

This project therefore sets out to investigate the treatment of oily wastewater by the effective integration of oil-water separation and biological treatment so as to maximize the cost effectiveness of the treatment. It is hoped this study will serve as a guide to profitable integration of these units.

The objectives of this study are therefore as follows:

- To investigate aeration in various ICB laboratory models.
- To develop models to design and size units in the integrated processes and predict their performances.
- To investigate the effectiveness of oil recovery units to recover oil in various scenarios and the effect on the cost of integrated processes.
- To investigate regions of economic operation and economic designs.

## 2.3 Theoretical aspects used in the research

This section introduces the theoretical concepts used in the research.

### 2.3.1 Oily wastewater

Oily wastewater is defined by its oil concentration, oil density, water temperature, average droplet size, and viscosity. All of these wastewater characteristics affect the performance of separation or treatment technologies in various ways. For example, droplet size affects the performance of gravity separators, flotation processes, membranes, and bioreactors. Specifically, large oil droplets coat and kill microbes in bioreactors and therefore reduce efficiency and increase costs.

The state in which oil exists in water governs the choice of treatment technology. Large oil droplets are most economically removed by gravity separation technologies while smaller droplets are best removed by biological treatment or membranes. Generally, oily wastewater is classified according to the states in which oil may exist in water; these states are: as free-floating droplets, unstable or dispersed emulsions, and stable emulsions (Rhee *et al.* 1987). Table 2.2 outlines oil-water classifications.

| | Oil droplet size range (μm) |
|---|---|
| Free floating oil | >150 |
| Unstable emulsions | 20-150 |
| Stable emulsions | <20 |

**Table 2.2: Oil-water system classification**

While the droplets are not actually particles the convention is to treat each discrete oil droplet suspended in water as a spherical droplet. Oily wastewater is modelled depending on the droplet diameters (Table 2.2); sampling techniques such as the dynamic light scattering method (Poulsen *et al.* 2007) and video recording techniques (Angeli and Hewitt 2000) are used in determining the number of droplets of a particular size in a given volume of oil water mixture. A wide variety of mathematical and empirical distribution functions, such as the normal distribution, log-normal distribution, Rosin-Rammler equation and Nukiyama-Tanasawa equation have commonly been used to simulate the droplet size distributions (Lefebvre 1989). The normal distribution does not usually show a good fit to experimental data, since the distributions are rarely symmetrical. To fit experimental data, Rosin–Rammler (Angeli and Hewitt 2000) (Mugele and Evans 1951) and log-normal distributions are usually employed (Sood and Awasthi 2003).

The oil droplet distribution shows the size distribution of droplets in a given sample of oily wastewater. The volume of oil contributed by droplets of a particular size depends on the size of the droplets, since the volume of oil contributed by a droplet is a function of the droplet size (Equation 2.0).

$$V_{oil} = \frac{\pi}{6} d^3 \qquad (2.0)$$

where,

$V_{oil}$ - Volume of oil in a droplet ($m^3$)

$d$ - Oil droplet diameter (m)

The total volume of oil contributed by droplets of a particular size in a distribution is therefore:

$$V_t = n \times \frac{\pi}{6} d^3 \qquad (2.1)$$

where,

$V_t$ - total volume of oil contributed by all droplets of a particular size ($m^3$)

$n$ - Number of droplets in a size range or interval

In this research, droplet distribution was generated using theoretical functions (Equation 2.2). A normal distribution was assumed. Based on an assumed mean droplet size and standard deviation, the normal distribution was obtained.

$$F(x, x_m, \sigma) = \frac{1}{\sigma\sqrt{2\pi}} e^{-\left[\frac{(x-x_m)^2}{2\sigma^2}\right]} \qquad (2.2)$$

where,

$x$ - Droplet size of interest (μm)

$x_m$ - Mean droplet size (μm)

$\sigma$ - Standard deviation from the mean

### 2.3.2 General review of Immobilized Cell Bioreactors

Immobilized cell bioreactors are bioreactors in which cells are immobilized on support particles (packing). Cell immobilization offers a lot of advantages:

- It facilitates cell separation.
- It leads to a high cell concentration within the reactor.
- It allows continuous operation beyond washout flow rates.
- Protects against contamination.
- Permits manipulation of growth rate in continuous systems independently of dilution rate.
- Provides possibilities for spatial location of different microbial populations within reactors.

Atkinson 1986 identified that the performance of any reactor containing immobilized cells depends upon:

- the number of support particles,
- the average biomass hold-up per support particle
- the average specific rate of reaction of the immobilized biomass and
- the average overall yield coefficients of the immobilized biomass

The yield coefficient and specific rate of reaction of the immobilized biomass depend, amongst other factors, on the ability of the cells to obtain oxygen, which in turns depends on the support particle size (Atkinson 1986, Da Fonseca 1983). The average biomass hold up per particle is subject to diffusional limitations on the rates of reaction, which are also affected by the support particle size. It can then be concluded that the support particle (packing) is the key feature of an ICB. It acts as the surface on which cells are immobilized and it also acts as a static mixer for the bioreactor, eliminating the need for mechanical stirrers which can disrupt immobilized aggregates. The support packing also affects pressure drop in a packed bed reactor. Da Fonseca *et al.* (1986) conclude that particle size may be considered as a design parameter in bioreactor design. For this reason, the effect of various sizes of plastic Pall support particle on mass transfer coefficient was investigated so as to minimize the cost due to aeration in the ICB.

Equation 2.3 represents the biodegradation process that takes place in an ICB.

$$C_nH_{2n} + O_2 \rightarrow CO_2 + H_2O + Cells \qquad (2.3)$$

Theoretically, for every 3.5 grams of oil oxidized, 3.5 grams of oxygen is required but the rate at which it is transferred depends on the mass transfer coefficient, which is affected by support particle size.

#### 2.3.2.1 Mass transfer theories

Oxygen transfer is the primary use of energy in the bioreactor and represents a significant capital cost as well as a running cost. The microbes are living cells which need oxygen for growth, product formation, and maintenance. The volumetric mass transfer co-efficient indicates the rate at which oxygen is transferred. The higher the rate of oxygen transfer, the better the performance of the bioreactor and the lower the cost of aeration. Correct determination of the volumetric mass transfer coefficient is important in bioreactor design.

Various theories have been developed to describe oxygen mass transfer from the gaseous phase to the liquid phase. The most widely used theory is the Lewis Whitman theory. This theory states that when two fluid phases are brought in contact with each other, a thin layer of stagnant fluid exists on each side of the phase boundary (Lewis and Whitman 1924). Mass transfer by convection is assumed to be insignificant in this unstirred layer and the transport is achieved by steady state diffusion. This theory postulates that the gas concentration in the liquid film is in equilibrium with the gas phase while the gas concentration in the bulk liquid is kept uniform by turbulent mixing. The main resistance to mass transfer lies in the films. For a gas which is sparingly soluble in water, for example oxygen, Lewis theory postulates that the main resistance to transfer of oxygen lies in the liquid film and the mass transfer coefficient $k_L$ is expressed as:

$$k_L = \frac{D}{\delta} \qquad (2.4)$$

where,

$k_L$ - mass transfer coefficient in the liquid phase (m/s)

$D$ - Diffusion coefficient of the oxygen in liquid ($m^2/s$)

$\delta$ - film thickness (m)

and the rate of mass transfer is expressed as

$$\frac{\partial m}{\partial t} = k_L A\,(C_s\text{-}C_L) \qquad (2.5)$$

where,

$m$ - Mass of oxygen (mg)

$A$ - Interfacial area normal to mass transfer ($m^2$)

$C_s$ - Saturation dissolved oxygen concentration (mg/l)

$C_L$ - dissolved oxygen concentration in liquid (mg/l)

Equation 2.5 divided by the liquid volume, V, gives

$$\frac{\partial C}{\partial t} = = k_L\, a\,(Cs\text{-}C_L) \qquad (2.6)$$

where,

$\frac{\partial C}{\partial t}$ - is the concentration change with respect to time

$a$ - specific interfacial area, (A/V) ($m^2/m^3$)

Other mass transfer theories are Higbie's and Danckwerts' theories. Higbie's theory refutes the interface as being unstirred but as being made up of small liquid elements, which are continuously brought up to the surface from the bulk of the liquid by the motion of the liquid itself (Higbie 1935). Higbie's theory postulates that the rate at which mass transfers to the liquid film is greater than the rate at which the mass leaves the film, so when a fresh element of liquid comes into the interface, unsteady state diffusion readily occurs since there is no gaseous component in the stream. Higbie's theory assumes that these fresh elements of liquid all stay the same time at the surface before being replaced by another stream of fresh liquid elements. Higbie's theory mass transfer coefficient is expressed as:

$$k_L = 2\sqrt{\frac{D}{\pi\theta}} \qquad (2.\ 7)$$

where,

$\theta$ - is the time of exposure of the liquid element (s)

Danckwerts' theory also opposes the concept of a stagnant film as suggested by Lewis theory and builds on Higbie's theory. While Higbie proposed that liquid elements are continuously brought up and stay the same time at the surface, Danckwerts' theory, known as surface renewal theory, proposes that the liquid elements do not stay the same time at the surface but are renewed at a rate known as the surface renewal rate (Danckwerts 1951). Danckwerts' theory assumes that if turbulent mixing occurs in the liquid in the proximity of a bubble, it is likely that unsteady state molecular diffusion will take place in individual eddies.

Danckwerts' mass transfer coefficient is expressed as:

$$k_L = \sqrt{D.s} \qquad (2.8)$$

where,

$s$ - Average frequency at which any particular vertical element is mixed or rate of surface renewal ($s^{-1}$)

The surface renewal rate increases with turbulence in Equation 2.8.

Higbie's and Danckwerts' theories were not used in this investigation because of the difficulty of determining both the liquid film coefficient and the interfacial area. The Lewis Whitman theory combines these parameters together. Determination of the interfacial area requires measurements of the bubble size, their lifetime in air or water, and the statistical knowledge of their distribution. The two-film theory on the other hand uses both $k_L$ and $A/V$ together as a lumped parameter, $k_{La}$; the combining of the two parameters essentially eliminates the difficulty in determining the interfacial area $a$, which strategically makes it more favourable than the two other theories. In the determination of the mass transfer characteristics of the various sizes of plastic Pall rings, the Lewis Whitman theory was employed.

#### 2.3.2.2 Bioreactor design

The characteristics of a well-designed bioreactor have been spelt out by various authors. Williams 2002 reported that a well-designed bioreactor is one which takes reactor parameters for the biological, chemical, and physical system (macro-kinetic) and bioreaction parameters into consideration during design. Freeman and Lilly 1998 define a well-designed immobilized cell bioreactor as one which takes a set of processing parameters into consideration during design. These include: the immobilization method, mode of operation (e.g., repeated batch vs. continuous), aeration and mixing, bioreactor configuration, medium composition (including feeding of substrates, precursors, or additional nutrients), temperature, pH and, whenever required, *in situ* product and/or excess biomass removal. In conclusion, a well-designed bioreactor is one which can specifically influence the metabolic pathways of the immobilized cells.

As stated previously, oxygen supply is a key issue in the design of an aerobic immobilized cell bioreactor (Da Fonseca *et al.* 1986). This problem is particularly acute in immobilized cells where oxygen has to diffuse through a solid phase. Immobilized cells depend on transport of oxygen with molecular diffusion or convective flow through the support materials used. The cells also depend on the diffusion of oxygen into the micro-colonies which have been formed by their growth. Diffusion limitations do exist in such colonies. Da Fonseca *et al.* 1986 compared the activity of immobilized cells at maximum cell hold-up using two aerating media. When aerating with air, the immobilized cell activity was estimated as 0.025, compared to 0.06 when aerating with pure oxygen. This confirmed the severity of oxygen diffusion through the solid phase.

Therefore to ensure complete degradation of oil in the bioreactor the following assumptions were made:

- No diffusion limitations exist.
- Sufficient cells are present to carry out the transformation.
- Immobilized cells are held to high concentration.
- Material of construction: carbon steel, coated to prevent corrosion or contamination.

To obtain complete degradation, bioreactors were sized to match oxygen demand for a given oxygen transfer rate (see Equation 2.9).

$$V_{ICB} = \frac{OD}{OTR} \qquad (2.9)$$

where,

$V_{ICB}$ - Immobilized cell bioreactor volume (l)

$OD$ - Oxygen demand (mg/s)

$OTR$ - oxygen transfer rate (mg/l.s)

where,

$$OTR = k_{La}\,(Cs - C) \qquad (2.9a)$$

### 2.3.2.3 Bioreactor scale-up

The volumetric mass transfer coefficient is the most widely used index for bioreactor design and scale-up. Maintaining a constant volumetric mass-transfer coefficient ($k_{La}$) is a common scale-up strategy in bioreactor scale-up. In this bioreactor design, to meet the oxygen demand, the laboratory immobilized cell bioreactor (ICB) was scaled up by expanding the capacity length-wise and width-wise whilst keeping the height (depth) constant to maintain the same hydrodynamics conditions in the model ICB.

### 2.3.2.4 Aeration system design

Oxygen transfer in immobilized cell bioreactors is achieved by aeration. Aeration accounts for the largest portion of operating costs (Atkinson 1986).

In this research, oxygen transfer is achieved by sparging air into the bioreactor. The aeration system consists of a blower and a coarse bubble diffuser. The blower was sized based on the oxygen demand of the process. The oxygen transfer efficiency of the coarse bubble diffuser was assumed to be 29.53%/m diffuser submergence (29.53% of the oxygen contained in 0.00508 $m^3$/s of air at standard conditions of 25°C and 100 kPa is transferred per metre of diffuser submergence). The blower horsepower requirements for each application were then estimated using the adiabatic compression formula:

$$BP = \frac{653.5Q_A\left[\left(\frac{100+P}{100}\right)^{0.283} - 1.0\right]}{\eta} \qquad (2.10)$$

where,

$BP$ - Brake Power (kW)

$Q_A$ - Volumetric flow rate of air ($m^3/min$)

$\eta$ - Mechanical efficiency (%)

$P$ - Blower discharge pressure (kPa)

A mechanical efficiency of 70% was assumed in estimating the horsepower requirements of the blower.

Blower discharge pressure was estimated based on static submergence, inlet losses, and equipment losses and piping losses. An estimate of 13.8 kPa was used in all designs.

The air volume required based on the coarse bubble oxygen transfer coefficient can be calculated from:

$$Q_A = \frac{OD}{OT} \qquad (2.11)$$

where,

$OT$ - Oxygen actually transferred / air volume ($kg/m^3$)

### 2.3.3 Oil water separation design theory

The theory which governs the design of gravity separators (horizontal flow and corrugated plate interceptors) is based on the relative velocities of the oil droplets and the waste stream. An oil droplet in a separator has two velocity components (http://www.megator.com/pdf/Parallel%20plate%20USA.pdf ):

- Horizontal velocity due to the forward movement of the waste stream ($v_h$) and
- The rising velocity of the oil droplet or terminal velocity ($v_o$).

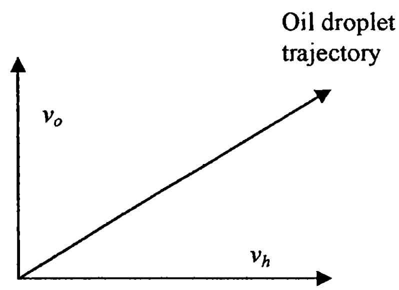


**Figure 2.1: Velocity components of an oil droplet in a flowing stream**

The resultant of the two velocity components describes the trajectory of the oil droplet (Figure 2.1). An oil droplet flowing in a separator will be removed or recovered if the resultant trajectory of the oil droplet will allow it rise out of the effluent flow path before the bulk flow reaches the separator exit (Figure 2.2).

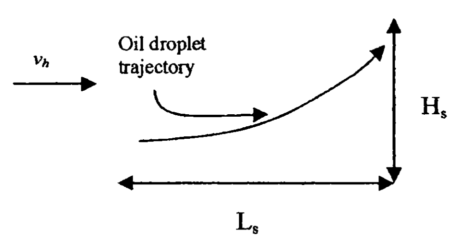


**Figure 2.2: Trajectory of an oil droplet in a moving stream**

Usually, in separator design, the objective is to completely remove droplets of a certain size: this droplet is called the design droplet or critical droplet. Separator dimensions are calculated based on the rise velocity of the critical droplet ($v_t$).

To ensure separation the following must be achieved:

- The time it takes the critical droplet to rise to the surface where it will be captured or recovered must be less than the time it takes the waste stream to flow to the exit.
- The rate of rise of the oily waste stream (overflow rate) must be less than the rate of rise of the droplet velocity.

The overflow rate is defined as:

$$O_R = \frac{Q}{A_s} \qquad (2.12)$$

where,

$O_R$ - Overflow rate (velocity of rise of the waste stream) (m/s)

$A_s$ - Surface area required for separation ($m^2$)

$Q$ - Volumetric flow rate of waste stream ($m^3/s$)

The rise rate of a droplet as described by Stokes' Law (Nordvik *et al.* 1996) is expressed as:

$$v_o = \frac{\pi d^2(\rho_w - \rho_o)g}{18\mu} \qquad (2.13)$$

where,

$v_o$ - is the terminal velocity of the oil droplet (m/s)

$d$ - is the diameter of the oil droplet (m)

$\rho_w$ - is the density of water ($kg/m^3$)

$\rho_o$ - is the density of the oil ($kg/m^3$)

g - is the acceleration due to gravity ($m/s^2$)

$\mu$ - is the dynamic viscosity of the continuous phase (Pa.s)

The derivation of this law follows from the postulation that two forces act on an oil droplet moving in a fluid. These forces are the buoyancy and drag forces. The buoyancy force is related to the density differential between oil and water and is responsible for the upward movement of any droplet, while the drag force is related to the velocity of the droplet and opposes the upward movement of the oil droplet. Stokes postulated that the oil droplet reaches its terminal velocity when the two forces are equal. The buoyancy force is expressed as:

$$F_B = \frac{\pi}{6} d^3(\rho_w - \rho_o)g \qquad (2.14)$$

where,

$F_B$ - the buoyancy force in Newton (N)

$d$ - is the diameter of the oil droplet (m)

The drag force is expressed as

$$F_d = 3\pi\mu v_o d \qquad (2.15)$$

where,

$F_d$ - is the drag force in Newton (N)

$\mu$ - is the dynamic viscosity of the continuous phase (Pa.s)

$d$ - is the diameter of the oil droplet (m)

when

$$F_B = F_d \qquad (2.16)$$

then the terminal velocity $v_o$ is attained (see Equation 2.13).

Stokes' Law adequately describes the rise rate of droplets only if the following conditions exist:

- Flow is laminar.
- Droplets are spherical.
- There is no interaction between oil droplets (ideal distribution).

### 2.3.4 General review of horizontal flow separators

The horizontal flow separator is basically a settling tank divided into sections:

- Inlet zone,
- Retention zone, and
- Outlet zone.

The baffles in the separator allow for smooth, slow, undisturbed flow along the length of the separator. This separator is an improvement of the settling tank which was only suitable for batch operations. The horizontal flow separator is designed for continuous operation.

Vapour losses from horizontal flow separators are minimized by use of fixed or floating covers. Various materials are used in the construction of a horizontal flow separator, they include: stainless steel, reinforced concrete and carbon steel.

Advantages of the horizontal flow separator

- Easy to construct.
- Reliable and simple.
- Inexpensive - little or no maintenance.

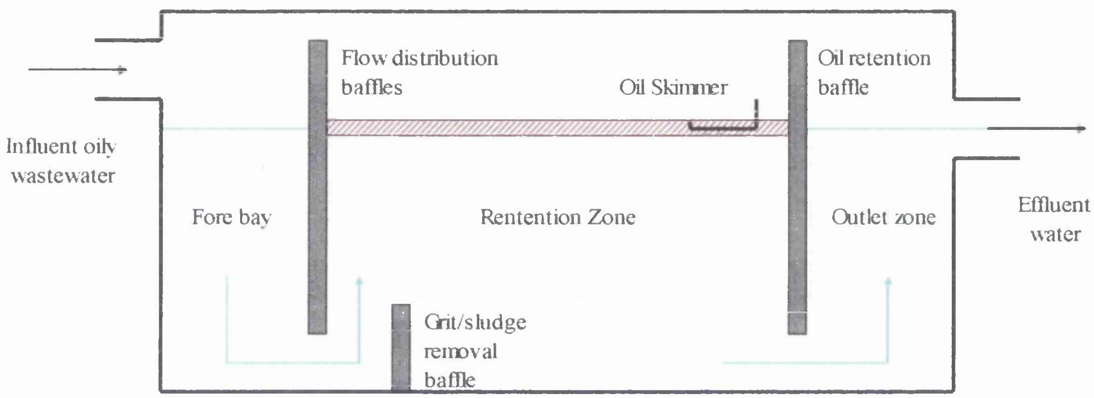


**Figure 2.3: A typical horizontal flow gravity separator**

### 2.3.4.1 Design of horizontal flow separators

The critical design droplet size for a horizontal flow separator is typically 150 microns. Using API specifications (American Petroleum Institute, February 1990), the separator is designed so that the two conditions stated below are satisfied.

- The rise rate of the wastewater must be less than or equal to the critical droplet rise rate.

$$O_R = v_t \qquad (2.17)$$

- The rise time of the critical droplet must be less than the residence time of the flow stream.

$$t_t < t_w \qquad (2.18)$$

where,

$v_t$ - rise rate of the design droplet or critical droplet (m/s)

$t_t$ - rise time of critical droplet (s)

$t_w$ - Residence time of oil-water stream (s)

Separator dimensions are governed by Equation 2.18a.

$$\frac{(v_h \times H_s)}{L_s} < v_t \qquad (2.18a)$$

where,

$H_s$ - Separator depth (m)

$L_s$ - Separator length (m)

The American Petroleum Institute (API) has developed a set of guidelines for the design of horizontal flow separators; these guidelines were used in this research.

The principles governing separator design as laid down by the API are based on four relations:

- A minimum horizontal area.
- Horizontal velocity.
- A minimum vertical cross section area.
- Depth to width ratio.

Minimum horizontal area

The API stipulates that the minimum surface area required for separation has to be corrected by a factor F (see Equation 2.19). This factor allows for turbulence and short-circuiting (a lower than ideal performance).

$$A_s = F\left(\frac{Q}{v_t}\right) \qquad (2.19)$$

$F$ - Design factor as a result of turbulence ($F_t$) and short circuiting ($F_s$)

The $F_s$ or short-circuiting factor is a function of ($R_1$) the ratio of the horizontal flow ($v_h$) velocity to the rise rate of the critical droplet ($v_t$). Typical values of $F$ are displayed in Appendix I.

Horizontal velocity

To limit the extent of turbulence which affects separator operation the API specifies that the horizontal velocity of the waste stream is:

$$v_h = 15 \times v_t \qquad (2.20a)$$

or,

$$R_1 \times v_t = v_h = 0.015 \qquad (2.20b)$$

where,

$R_1$ - is ratio of horizontal flow velocity ($v_h$) to rise velocity of critical droplet ($v_t$).

The smaller of the two values from Equations 2.20a or 2.20b is usually the desired $v_h$.

Minimum vertical cross section area

The minimum vertical cross-sectional area of the separator is then calculated from:

$$A_h = \frac{Q}{v_h} \qquad (2.21)$$

Based on conditions specified by API the width of the separator is calculated, using Equation 2.21a.

$$w_s = \left(\frac{Q}{R_2 \times v_h}\right)^{0.5} \qquad (2.21a)$$

To reduce turbulence the following ranges for separator dimensions are desirable:

$$0.9144\ m \leq H_s \leq 2.4384\ m \qquad (2.21b)$$

$$1.8288\ m \leq w_s \leq 6.096\ m \qquad (2.21c)$$

where,

$w_s$ - Separator width (m)

Depth to width ratio

Desirable depth to width ratio is given in Equation 2.21d.

$$R_2 = \frac{H_s}{w_s} = 0.3 \text{ to } 0.5 \qquad (2.21d)$$

where,

$R_2$ - is ratio of separator depth ($H_s$) to width ($w_s$)

From Equation 2.21d the separator depth becomes:

$$H_s = R_2 \times w_s \qquad (2.21e)$$

Based on the values of $F$, $R_1$ and $H_s$, the length of the separator can be calculated using Equation 2.21f.

$$L_s = F \times R_1 \times H_s \qquad (2.21f)$$

and,

$$L_s > H_s \qquad (2.21g)$$

### 2.3.4.2 Analysis of droplet capture in horizontal flow separator

Theoretically, the droplets that will be 100% recovered in a horizontal flow separator are those droplets with a rise rate equal to or greater than the overflow rate.

Any droplet with rise velocity less than the critical droplet velocity, or overflow rate, will be removed in the following proportions:

$$X = \frac{h_o}{H_s} \qquad (2.22)$$

where,

$X$ - Fraction of droplets of a particular size recovered

$h_o$ - height travelled by any droplet (m)

The height travelled by any droplet size is a function of the residence time of the wastewater.

$$h_o = v_o \times t_w \qquad (2.22a)$$

where,

$t_w$ - residence time of oil-water mixture in separator (s)

and,

$$t_w = \frac{L_s}{v_h} \qquad (2.22b)$$

When $X$=1, then all droplets of that size are recovered and when X is less than 1 then the fraction of droplets of that size captured is in the proportions defined by Equation 2.22. The fraction of droplets not recovered is:

$$Y = 1 - X \qquad (2.23)$$

where,

$Y$ - is fraction of droplets of a particle size not recovered or captured

The un-recovered fraction flows to the separator exit.

This analysis is based on the principle of unhindered settling (Metcalf and Eddy 1991) which assumes the following:

- Oil droplets are spherical droplets.
- Ideal distribution (no interaction of oil droplets).
- Particles with the same rising velocity move in parallel paths.
- Even distribution of droplets across the inlet cross-sectional area.
- Laminar flow.

### 2.3.5 General review of Corrugated Plate Interceptors

A corrugated plate interceptor is basically a separation tank with coalescing plates inclined at an angle. The inclination of the plates allows for removal of solids, as well as oil, from the waste stream.

The plate interceptor improves on the horizontal flow gravity separator by providing a much smaller distance for oil droplets to rise and an increased surface area. The shorter distance travelled by the oil droplets is accomplished by having coalescing plates in the separation chamber. The oil droplets only have to rise to the next higher plate where they attach to other droplets. The coalesced oil droplets then rise alongside the underside of each plate. At the top of the plate, large droplets break away and rise to the surface where they are skimmed off (Figure 2.4).

The increased surface area of the corrugated plate separator is as a result of the inclusion of the plates. The combined area of all the plates constitutes the total surface area of the plate separator. Various materials are used in the construction of CPI units: stainless steel, reinforced concrete, and carbon steel.

Flow into a corrugated plate pack can be either down flow or cross-flow.

In down flow separators, the wastewater flows down between the plates; sludge deposits on the top of the plates and flows to the bottom of the separator, whilst oil accumulates beneath the plates and flows counter currently to the top of the separator.

In cross flow separators, flow enters the plate pack from the side and flows horizontally between the plates. Oil and sludge accumulate on the plate surfaces above and below the wastewater. As the oil and sludge build up the oil globules rise to the separator surface and sludge gravitates towards the bottom of the separator.

Advantages of the Corrugated Plate Interceptor (CPI)

- Absence of moving parts.
- It is compact compared to horizontal flow separators.
- An inexpensive method of enhancing separation efficiency.
- Allows for higher wastewater flow compared to horizontal flow separators.

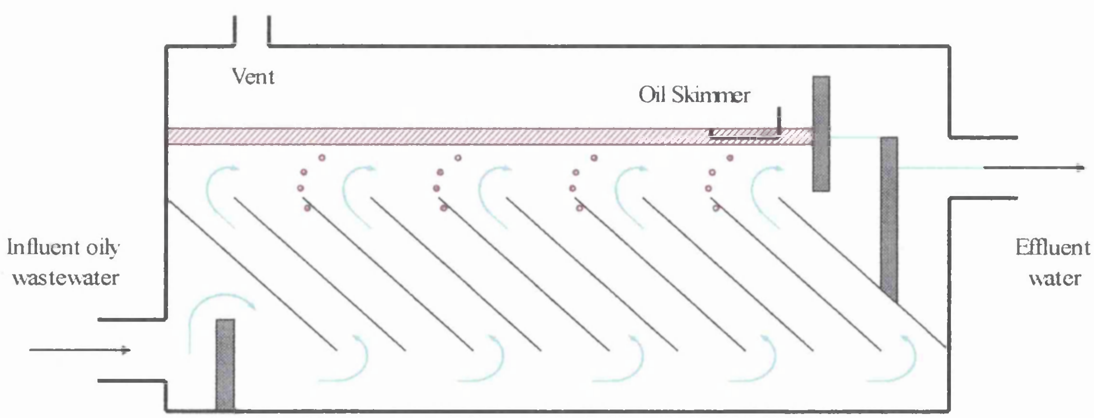


**Figure 2.4: A cross flow Corrugated Plate Interceptor**

#### 2.3.5.1 Design of corrugated plate interceptors

The corrugated plate separator is usually sized to remove droplets of size 60 microns.

The effective surface area required for separation for any given flow rate is obtained from Equation 2.24.

$$A_s = F_{dp} \frac{Q}{v_t} \qquad (2.24)$$

where,

$F_{dp}$ -is a design factor for corrugated plate interceptors (%)

Since plates are inclined, only the projected horizontal area is effective for separation (see Figure 2.5).

The projected area available for separation for given plate dimensions is:

$$A_P = L_P \times w_P \times \cos\alpha \qquad (2.25)$$

where,

$A_P$ - projected plate area of a plate ($m^2$)

$L_P$ - plate length (m)

$w_P$ - plate width (m)

$\alpha$ - angle of inclination of the plates to the horizontal (degrees)

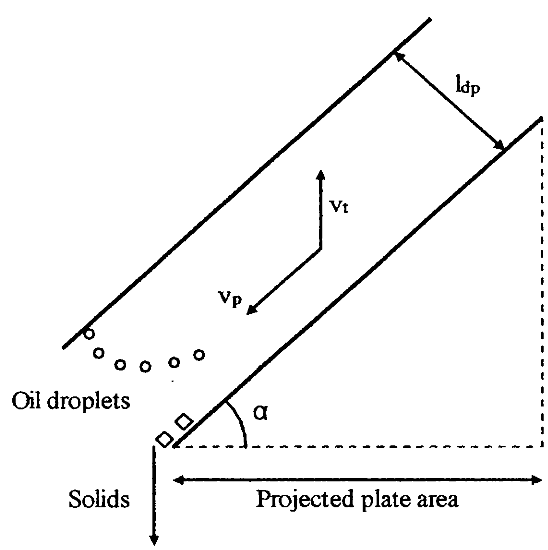


**Figure 2.5: Effective (projected) area of an inclined plate**

The total number of plates required for a given plate dimension and for a given flow capacity is obtained from Equation 2.26.

$$N_P = \frac{A_s}{A_P} \qquad (2.26)$$

where,

$N_P$ - Number of plates required to achieve separation

The effective settling length is the plate length.

There are no set design guidelines for calculation of the tank dimensions; tanks are generally sized to contain the plates. The separators are usually designed to minimize turbulence, short circuiting, and channelling of the inflow especially through and around the plate packs. To avoid plugging, the plates are spaced apart. The spacing of the plates affects the velocity of flow through the separator. Usually the plates are spaced to ensure that laminar flow conditions exist in the separation bay. The separator is usually designed to have a Reynolds number of less than 500 (laminar flow). In this research, the separator was sized to contain the plate packs and in addition to the calculated tank height 0.45 m was added to allow for sludge settling and headspace above the plate packs.

$$L_s = (N_P - 1) \times l_{dp} + 2 \times L_P \times \cos\alpha \qquad (2.27)$$

$$w_s = w_P \qquad (2.28)$$

$$H_s = L_P \times \sin\alpha + 0.1524 + 0.3048 \qquad (2.29)$$

To verify laminar flow conditions exist, the Reynolds number through the bay was calculated using Equation 2.30.

$$Re = l_{dp} \times v_p \times \rho_w / \mu_w \qquad (2.30)$$

where,

$l_{dp}$ - is the distance between plates (m)

$v_p$ - Flow velocity along plates (m/s)

$\rho_w$ - water density (kg/m$^3$)

$\mu_w$ - water viscosity (Pa.s)

The flow velocity is given by:

$$v_p = \frac{Q}{\left(l_{dp} \times w_P\right)N_P} \qquad (2.30a)$$

### 2.3.5.2 Analysis of droplet capture or recovery in corrugated plate interceptor

Yao 1973 predicted that the velocity of the oil droplet size 100% theoretically removed or recovered in a corrugated plate interceptor with a flow velocity along plates ($v_p$) can be found by using the equation below:

$$v_t = v_p \times d\,(d \sin \alpha + w_P \cos \alpha) \qquad (2.31)$$

The derivation of Equation 2.31 is given in Appendix I.

Theoretically, the droplets that will be 100% recovered in a corrugated plate interceptor are those droplets with a rise rate equal or greater than the rise rate of the critical droplet. The droplets with a lower rise rate than the overflow rate will only be removed in certain proportion.

Any droplet with a rise velocity less than the critical droplet velocity predicted by Yao will be removed in the following proportions:

$$X = \frac{v_o}{v_t} \qquad (2.32)$$

where,

$X$ - Fraction of droplets of a particular size recovered

$v_o$ - rise rate of any droplet (m/s)

$v_t$ - rise rate of the critical droplet, obtained from Equation 2.31

If $X = 1$ then all droplets of that size are recovered and if X is less than 1, then only the proportion defined by Equation 2.32 is captured. The fraction of droplets not recovered is:

$$Y = 1 - X \qquad (2.33)$$

### 2.3.6 Cost modelling

A cost factor model was used in estimating the cost of the process units (Equation 2.34).

$$C_e = CS^n \qquad (2.34)$$

where,

$C_e$ - purchased cost (£)

$S$ - Characteristic size parameter in units given in Appendix I

$C$ - Cost constant (see Appendix I)

Costing of the process was based on the following:

- Cost of sludge disposal from pre-treatment units was not taken into consideration and a mixed culture is assumed in the ICB to prevent sludge formation.
- For the bioprocess, only two main costs were considered: cost of the bioreactor unit and cost of electricity for aeration.
- In the costing of the pre-treatment units cost of skimmers was not taken into consideration. But cost considered includes cost of unit and cost of recovered oil.
- Cost of recovered oil is assumed to be £0.62/$m^3$.
- Electricity rate was assumed to be £0.05/kWh.
- The life span of each unit was assumed to be 25 years.
- The cost of energy was not taken into consideration in investigating the effect of viscosity on the cost of integrated process.
- The cost of dissolving air to increase density difference was not taken into consideration.
- The cost analysis was done on a yearly basis.

### 2.3.7 Theory of Process Integration

The process integration technique involves adopting a holistic approach to achieve certain objectives such as technical, environmental, economical and safety goals. To achieve these objectives, the approach in process integration involves understanding the effect of changes in a unit and how it affects the process as a whole. Based on this knowledge, decisions can be made.

To this end, to reduce the cost of treatment in the integrated processes, a holistic approach was employed; this method involved investigating the effect of change in a unit on the cost of the overall process.

# Chapter 3: Materials, Equipment and Methods

## 3.1 Introduction

This chapter describes the materials, methods, and experimental procedures used in the practical aspect of this research, and also describe the spreadsheet calculation procedures used for the theoretical aspect of the research.

The practical part of this research involved the investigation of aeration in various designs of a laboratory scale Immobilized Cell Bioreactor (ICB) unit. The designs were created by using one size of plastic Pall ring at a time in the ICB unit. The investigation involved determination of the mass transfer characteristics of the various sizes of plastic Pall support particles. The importance of aeration and the underlying theory are given in the Chapter 2.

The theoretical aspect of this research involved:

- Generating the various oil droplet size distributions in oil-water systems,
- Designing separation units and predicting their performance,
- Sizing the bioreactor and associated system (blower) required in each scenario
- Costing individual units and the integrated process

The exercise was done to help make judgements in seeking the most cost effective method of improving oil recovery and improving the cost effectiveness of the integrated processes. The algorithms used and calculation procedures are presented and explained.

## 3.2 Materials

The fluids used in the experimental aspect of this research were nitrogen gas, compressed air, and water. Nitrogen was used to strip oxygen in solution. Plastic Pall rings were used as support particles or packing. The physical data of the plastic Pall rings are given in Table 3.1.

| Size (mm) | Weight ($kg/m^3$) | Surface Area $m^2/m^3$ |
|---|---|---|
| 16 | 115 | 340 |
| 25 | 80 | 196 |
| 38 | 60 | 150 |
| 50 | 60 | 106 |

**Table 3.1: Physical data of plastic Pall rings**

The equipment used in the experiment were an oxygen probe/meter, a stop watch, and a rotameter.

### 3.2.1 Oxygen probe and meter

A calibrated polarographic automatic temperature-compensated oxygen electrode (Mettler Toledo probe/Mettler Model IP67) and meter (Mettler Toledo MO128) was used to monitor and read the variation of the dissolved oxygen (DO) concentration with time in the system.

#### 3.2.1.1 Oxygen probe calibration

To ensure accuracy in probe readings, the oxygen probe was calibrated using the 2 point method.

The 2-point method involves calibrating the probe at 2 points, 100% saturation, and 0% saturation of oxygen. For the 100% calibration, the probe was calibrated in air and the output current was set when the meter read 100%. For the second calibration point (0%), the probe was dipped in a magnetically stirred beaker of nitrogen-saturated water; the output current was set when the meter read 0%. The 0% reading corresponded to zero oxygen concentration in the solution.

Once the probe was calibrated, the response time of the probe was determined by quickly moving the probe from an oxygen-rich solution to a nitrogen-rich solution and noting the time it took to response to change. The experiment was repeated in the reverse direction, moving the probe from a nitrogen-rich solution to an oxygen-rich solution. A short response time was observed, less than 1 second, and hence the probe response was not taken into consideration in $k_{La}$ estimation.

### 3.2.2 Rotameter

A rotameter equipped with a control valve was used in measuring the airflow into the reactor. The rotameter was calibrated using the water displacement method.

The water displacement method involves using air collected from the column to displace a known volume of water and noting the time it takes to displace the water.

To obtain the equivalent flow rate for each observed scale on the rotameter, the air flow was set to each observed scale by adjusting the control valve; compressed air passed into the column at each scale reading was collected via tubing from the tightly sealed column. The air was then used in displacing a known volume of water, noting the time it took to achieve this. These values were then converted to volumetric air flow rates.

This procedure was repeated for each observed scale reading on the rotameter and a calibration graph was plotted (see Appendix II).

## 3.3 Experimental set-up

The experimental set-up is represented in Figure 3.1. The experiments were carried out in a glass cylindrical column with a tapered bottom. The total height of the column was 0.83 m, inner diameter 0.17 m and outer diameter 0.19 m. For each ICB unit design, the column was packed randomly with plastic Pall rings of a particular size; these plastic Pall rings were fixed in place by wire meshes at the top and bottom (Figure 3.1). Mains tap water was used to fill the column. The column was operated in a semi-batch mode: with respect to water it was operated in a batch mode and with respect to air it was operated in a continuous mode. The probe was located well above the packing height at an angle. This was done for two reasons. Firstly, the probe requires a high liquid velocity for proper operation as the output current depends on liquid velocity (Silver 1965); the second reason is to prevent air bubbles resting on the membrane. Votruba *et al.* 1976 and Votruba *et al.* 1977 reported that when air bubbles rest on the probe membrane, the probe measures a concentration between the equilibrium value and the bulk value, rather than the bulk concentration. The gas sparger was located below the mesh to ensure satisfactory gas distribution. All measurements were carried out at room temperature and atmospheric pressure.

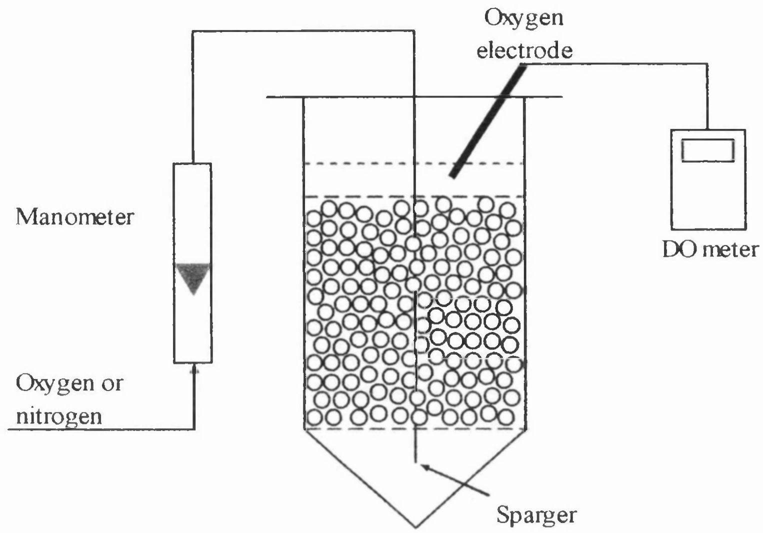


**Figure 3.1: The experimental set-up for mass transfer measurements**

## 3.4 Method

### 3.4.1 Determination of $k_{La}$

The static gas method was used to measure the change in dissolved oxygen with time. This involved stripping the water of oxygen by sparging with nitrogen and then sparging with oxygen at a steady flowrate while taking measurements of the dissolved oxygen concentration every ten seconds until the solution became saturated. The change in dissolved oxygen concentration was measured with and without packing. Oxygen concentration measurements were taken well above the height of the support particles, an ideal mixing of the liquid phase was assumed in that location.

In order to observe the effect of aeration on $k_{La}$ in the various ICB designs (each filled with a support particle of a particular size), the aeration rate was varied between 8 ml/s and 57 ml/s for each ICB design. It was ensured that the total measurement time was long enough to obtain the oxygen saturation concentration, $C^*$. The value of C (oxygen concentration in the solution) was read off from the oxygen probe meter at equal intervals. Values for $k_{La}$ were then determined for each aeration rate investigated in each ICB design using Equation 3.0.

$$dC/dt = k_{La}(C^*-C) \qquad (3.0)$$

where,

$C^*$ - the saturation value of DO at the gas-liquid interface (% saturation)

C - Concentration of oxygen in the bulk liquid phase (% saturation)

$k_{La}$ - volumetric liquid phase mass transfer coefficient ($s^{-1}$)

## 3.5 Calculation procedure for generating oil droplet distribution

To generate the oil droplet distribution, a spreadsheet using Equation 2.2 was employed. The spreadsheet calculation procedure is presented in Figure 3.2.

The parameters specified in the spreadsheet calculation are flow rate, oil concentration, average oil density, droplet sizes, mean droplet size of the oil-water mixture, the standard deviation from the mean and an assumed number of droplets in the first instance. The spreadsheet uses these parameters to further define the influent oil water mixture characteristics, determine the number of droplets less than a particular size, and calculate the volume of oil contained by all droplets of a certain size.

### 3.5.1 Influent oil characteristics

To calculate the mass flow rate of the oil in the influent the spreadsheet uses the specified oily wastewater flow rate and oil concentration. This calculation is necessary to determine the volume of oil in the oil-water mixture. The spreadsheet uses the calculated mass flow rate and the specified average oil density to calculate the volume of oil in the mixture. Table 3.2 is an example of values generated as a result of specifying the required input data. These results are used in generating the oil droplet size distribution and the volume of oil in each size interval.

**Oily Wastewater oil droplet size distribution**

Influent characteristics

| | | |
|---|---|---|
| Water flowrate | **10** | $m^3/h$ |
| Oil concentration | **1000** | mg/l |
| Average oil density | **800** | $kg/m^3$ |
| Average droplet dia | **60** | μm |
| Standard deviation | **15** | - |

Calculations

| | | |
|---|---|---|
| Mass of oil in influe | 10 | kg/h |
| Volume of oil in influ | 0.0125 | $m^3/h$ |
| No. of droplets in in | 9.25E+10 | - Solver variable |

| Droplet size interval (1) | Minimum droplet size (μm) (2) | Maximum droplet size (μm) (3) | Mean droplet size (μm) (4) | Cumulative distribution function (5) | Fraction of droplets in interval (6) | Number of droplets in interval (7) | Volume of droplets in interval interval (m3) (8) |
|---|---|---|---|---|---|---|---|
| 1 | 0 | 10 | 5 | 0.00043 | 0.00043 | 3.970E+07 | 2.598E-09 |
| 2 | 10 | 20 | 15 | 0.00383 | 0.00340 | 3.147E+08 | 5.561E-07 |
| 3 | 20 | 30 | 25 | 0.02275 | 0.01892 | 1.751E+09 | 1.432E-05 |
| 4 | 30 | 40 | 35 | 0.09121 | 0.06846 | 6.334E+09 | 0.0001422 |
| 5 | 40 | 50 | 45 | 0.25249 | 0.16128 | 1.492E+10 | 0.000712 |
| 6 | 50 | 60 | 55 | 0.50000 | 0.24751 | 2.290E+10 | 0.0019949 |
| 7 | 60 | 70 | 65 | 0.74751 | 0.24751 | 2.290E+10 | 0.0032929 |
| 8 | 70 | 80 | 75 | 0.90879 | 0.16128 | 1.492E+10 | 0.0032963 |
| 9 | 80 | 90 | 85 | 0.97725 | 0.06846 | 6.334E+09 | 0.0020368 |
| 10 | 90 | 100 | 95 | 0.99617 | 0.01892 | 1.751E+09 | 0.0007859 |
| 11 | 100 | 110 | 105 | 0.99957 | 0.00340 | 3.147E+08 | 0.0001908 |
| 12 | 110 | 120 | 115 | 0.99997 | 0.00040 | 3.677E+07 | 2.928E-05 |
| 13 | 120 | 130 | 125 | 1.00000 | 0.00003 | 2.789E+06 | 2.852E-06 |
| 14 | 130 | 140 | 135 | 1.00000 | 0.00000 | 1.372E+05 | 1.767E-07 |
| 15 | 140 | 150 | 145 | 1.00000 | 0.00000 | 4.370E+03 | 6.975E-09 |
| 16 | 150 | 160 | 155 | 1.00000 | 0.00000 | 9.007E+01 | 1.756E-10 |
| 17 | 160 | 170 | 165 | 1.00000 | 0.00000 | 1.200E+00 | 2.823E-12 |
| 18 | 170 | 180 | 175 | 1.00000 | 0.00000 | 1.032E-02 | 2.897E-14 |
| 19 | 180 | 190 | 185 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 20 | 190 | 200 | 195 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 21 | 200 | 210 | 205 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 22 | 210 | 220 | 215 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 23 | 220 | 230 | 225 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 24 | 230 | 240 | 235 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 25 | 240 | 250 | 245 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 26 | 250 | 260 | 255 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 27 | 260 | 270 | 265 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| Totals | | | | | 1.00000 | 9.252E+10 | 0.012499 |

Difference between calculated oil volumes 1E-06 - use Solver to make it zero.

**Table 3.2: Oil droplet size distribution**

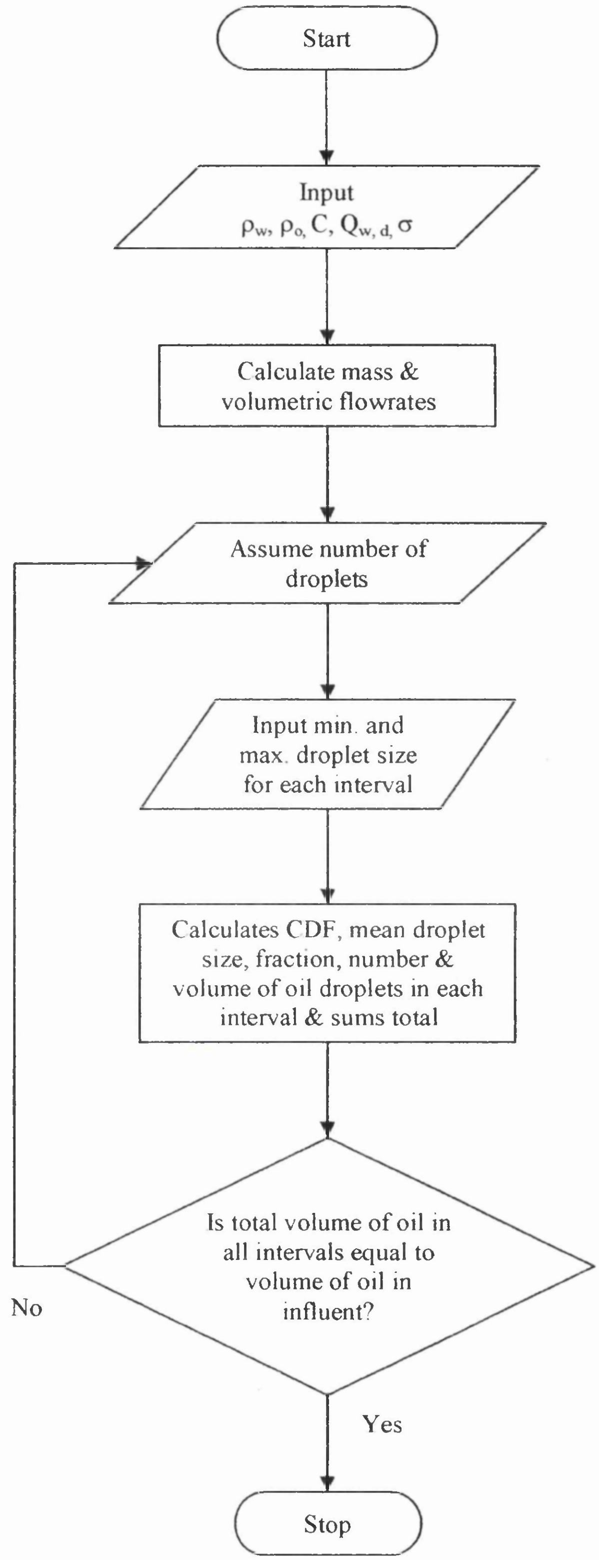


**Figure 3.2: Calculation procedure for generation of oil droplet distribution**

### 3.5.2 Droplet size distribution

By specifying a minimum (column 2) and maximum (column 3) droplet size, the droplet sizes are classified into equal intervals (see Table 3.2). The spreadsheet first calculates the mean droplet size (4) of each interval, which is then used to represent the droplets sizes in that interval and also for calculating the volume of oil contributed by each interval (8).

Based on a specified mean droplet size for the oil-water mixture and its standard deviation, the spreadsheet calculates the fraction of droplets less than the maximum droplet size (5) in each interval using the cumulative distribution function in Excel. This calculation aids in determining the fraction of droplets in each interval (6). The fraction of droplets in each interval in column six is calculated by subtracting the cumulative distribution fraction of the previous interval from the current cumulative distribution fraction, except in the first interval where the fraction of droplets in the interval is equal to the cumulative distribution function.

To obtain the number of droplets in an interval in column (7), the spreadsheet multiplies the assumed number of oil droplets in the oil-water mixture by the fraction of droplets in each interval in column (6). Figure 3.2 shows the spreadsheet calculation procedure.

### 3.5.3 Calculation of oil volume

To calculate the volume of oil contributed by droplets in each interval (column 8) the spreadsheet multiplies the calculated number of oil droplets in each interval (column 7) by the volume of oil contributed by the mean droplet size of each interval (see Equation 2.1 and column 4 in Table 3.2).

The spreadsheet compares the total volume of oil contributed by each interval (sum of column 8) with the volume of oil in the influent to check if the assumed total number of droplets was correct. If the summed oil volume is not equal to the calculated influent oil volume, the Excel solver function is used to reduce to zero the difference between the calculated oil volume and the summed oil volume by varying the assumed number of oil droplets in the oil-water mixture (see Table 3.2 and Figure 3.2). When these operations have been carried out the spreadsheet then presents the droplet size distribution of the oil-water mixture. The droplet size distribution can then be plotted using the probability mass function in Excel.

## 3.6 Calculation procedure for design of gravity separators

The spreadsheet uses specified values of flow rates, design droplet velocity, and assumed values of $R_1$ (see Equation 2.20b) and $R_2$ (see Equation 2.21d) to obtain the effective separation area, horizontal flow velocity, and dimensions of the separator. The spreadsheet calculation procedure is presented in Figure 3.3.

In the first step, the spreadsheet calculates the separation area required for separation of the oil-water mixture, based on the design droplet velocity and the specified flow rate of the oil water mixture.

Using the assumed $R_1$ and the specified flow rate, the spreadsheet calculates the horizontal flow velocity using Equation 2.20b. The spreadsheet then compares the calculated horizontal flow velocity (Figure 3.3) with the guidelines set by the American Petroleum Institute (API) (see Equations 2.20a and 2.20b) and chooses the minimum of the two values.

The spreadsheet then uses the calculated horizontal velocity and the specified flow rate to calculate the cross sectional area of the separator using Equation 2.21.

The spreadsheet then calculates the dimensions (depth and width) of the cross sectional area of the separator using Equations 2.21e and 2.21a and compares the calculated width with the constraints set up by the API (Equation 2.21c). The parameters used in the calculation of the width are the calculated horizontal flow velocity, specified flow rate and an assumed $R_2$ (Equation 2.21d). $R_2$ is the variable that has to be changed if the calculated width is not within the limits of the constraints. The spreadsheet then calculates the separator length using Equation 2.21f.

The calculated separator dimensions are then compared to the constraints defined by API (Equations 2.21b, 2.21c, and 2.21g). If the calculated dimensions do not satisfy the imposed constraints then another value of $R_2$ must be entered and the process repeated again until the separator dimensions satisfy the constraints.

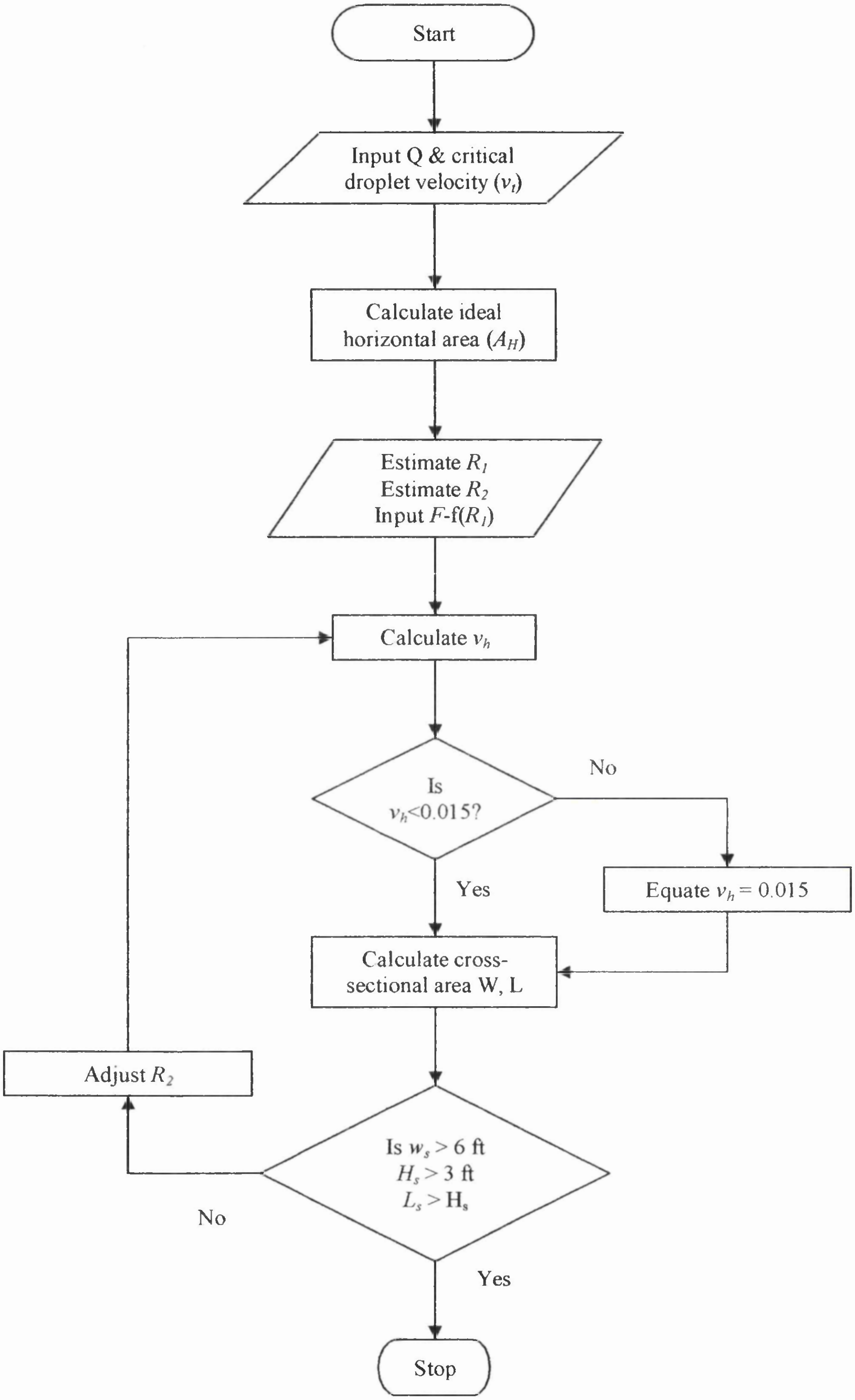


**Figure 3.3: Calculation procedure for design of gravity separators**

### 3.6.1 Predicting separator performance

The spreadsheet calculation procedure for oil droplet removal is presented in Figure 3.4. To analyze the droplets removed or recovered, separator dimensions, residence time of the oily wastewater stream in the separator, droplet distribution in the influent stream, and oil-water influent characteristics were specified. Using these values, the spreadsheet analyzes which droplets can be recovered from the oily wastewater stream before it reaches the exit of the separator. This analysis was done using the principle of unhindered settling (see section 2.3.4.2). The droplet size removal analysis was based on Equation 2.22. The spreadsheet calculation procedure is described below.

The spreadsheet first calculates the terminal velocity of each droplet size using Equation 2.13 and then calculates the height travelled by each droplet size based on the residence time of the oil-water mixture in the separator (see Equation 2.22a). The height travelled by each droplet size is then compared to the separator depth. If the height travelled by any droplet of that size is greater than the separator depth then the analysis assumes all droplets of that size are recovered or removed and the spreadsheet displays a value of 1; if the height travelled by any droplet in the given time is less than the separator depth, the spreadsheet then assumes only a fraction of that droplet size can be recovered. The fraction recovered is calculated using Equation 2.22 and the un-recovered fraction is calculated using Equation 2.23. The volume of oil recovered and unrecovered is then calculated using Equation 2.1.

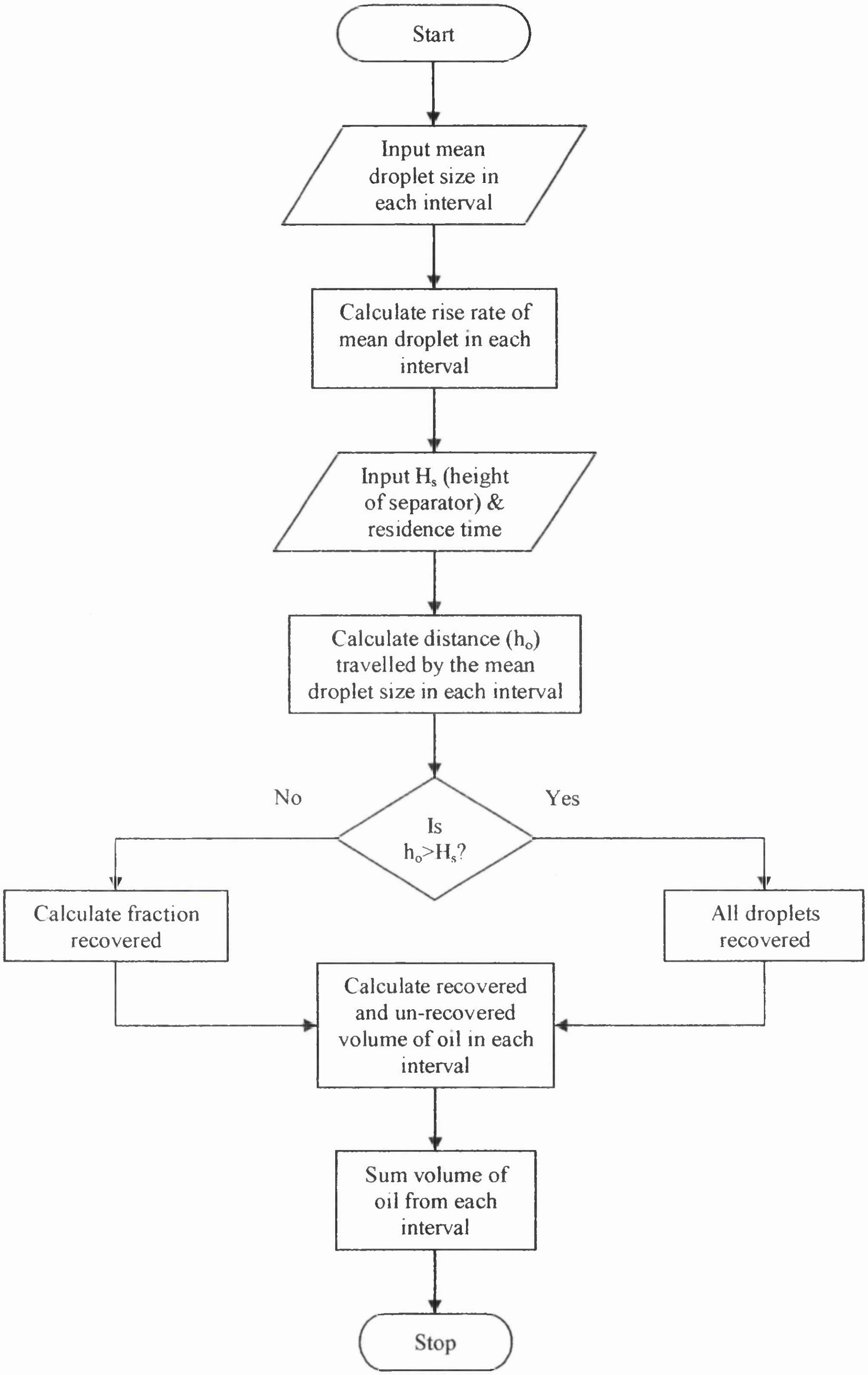


**Figure 3.4: Calculation procedure for oil droplet removal**

## 3.7 Calculation procedure for design of CPI

For the design of a Corrugated Plate Interceptor, the spreadsheet uses specified values of oil influent characteristics, design droplet velocity, and plate characteristics such as plate length or settling length, plate width, distance between plates, and angle of inclination of plates (Figure 3.5).

The first step in designing the Corrugated Plate Interceptor involves calculating the required separation area to achieve a given separation. The spreadsheet calculates this area using Equation 2.24 and an assumed design factor of 12%.

To obtain the number of plates required to achieve the effective separation area the spreadsheet uses Equation 2.25 to calculate the projected area of one plate and then uses Equation 2.26 to determine the number of plates required.

The dimensions of the separator are then calculated using Equations 2.27, 2.28 and 2.29. The flow velocity along the plates is also calculated using 2.30a.

The Reynolds Number is then calculated using Equation 2.30 to ensure laminar flow conditions are maintained in the interceptor.

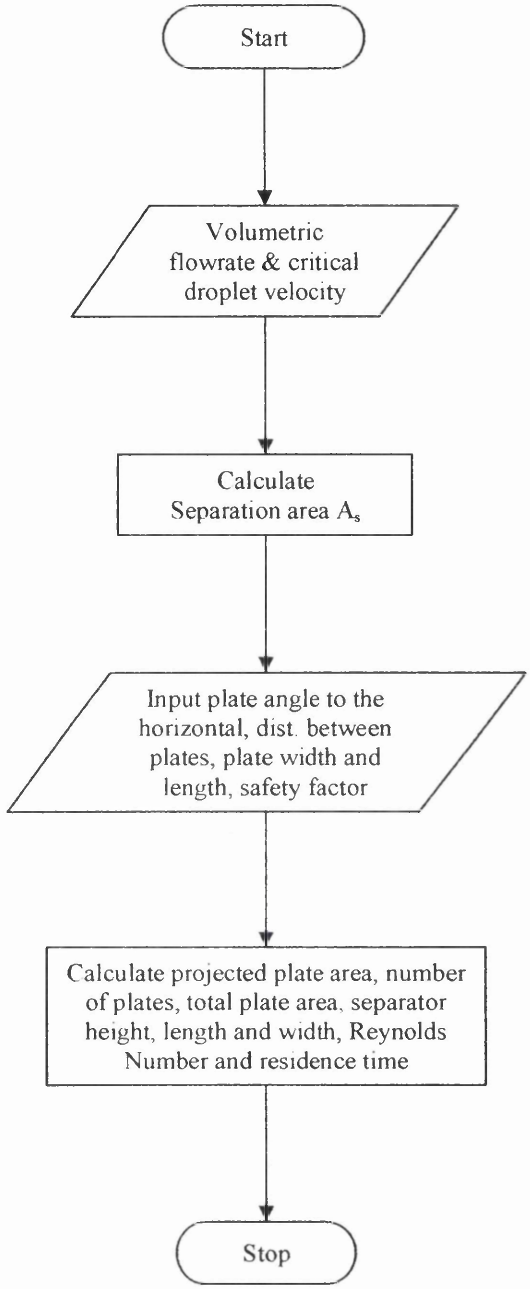


**Figure 3.5: Calculation procedure for design of plate interceptors**

### 3.7.1 Predicting CPI performance

To analyze droplet capture in the Corrugated Plate Interceptor, separator dimensions, droplet sizes, plate characteristics, oil-water influent characteristics are specified for the spreadsheet calculations (Figure 3.6). The droplet removal analysis is based on Equation 2.31. Only droplets with a velocity greater than the rise rate of the critical oil droplet defined by Equation 2.31 are completely removed. Oil droplets with rise rates less than the rise rate of the critical droplet are removed in certain proportions. The fraction of droplets recovered is calculated by the spreadsheet using Equation 2.32. The spreadsheet calculation procedure is described below.

The spreadsheet first calculates the terminal velocity of the mean droplet size in each droplet size interval using Equation 2.13; it then compares the terminal velocity of the mean droplet in each interval with the velocity of the (critical) droplet which is 100% theoretically recovered (Equation 2.31). If velocity of rise of any oil droplet is greater or equal to the critical droplet velocity as defined by Equation 2.31, then the spreadsheet displays a value of 1; if any oil droplet's velocity is less than the critical droplet's velocity then only a fraction of the droplets of that size are recovered. The spreadsheet calculates the recovered fraction using Equation 2.32 and the unrecovered fraction is calculated using Equation 2.33.

The volume of oil recovered from each size interval is then summed to obtain the total volume of oil recovered per hour from each volume of oily wastewater treated. The spreadsheet calculation procedure is presented in Figure 3.6.

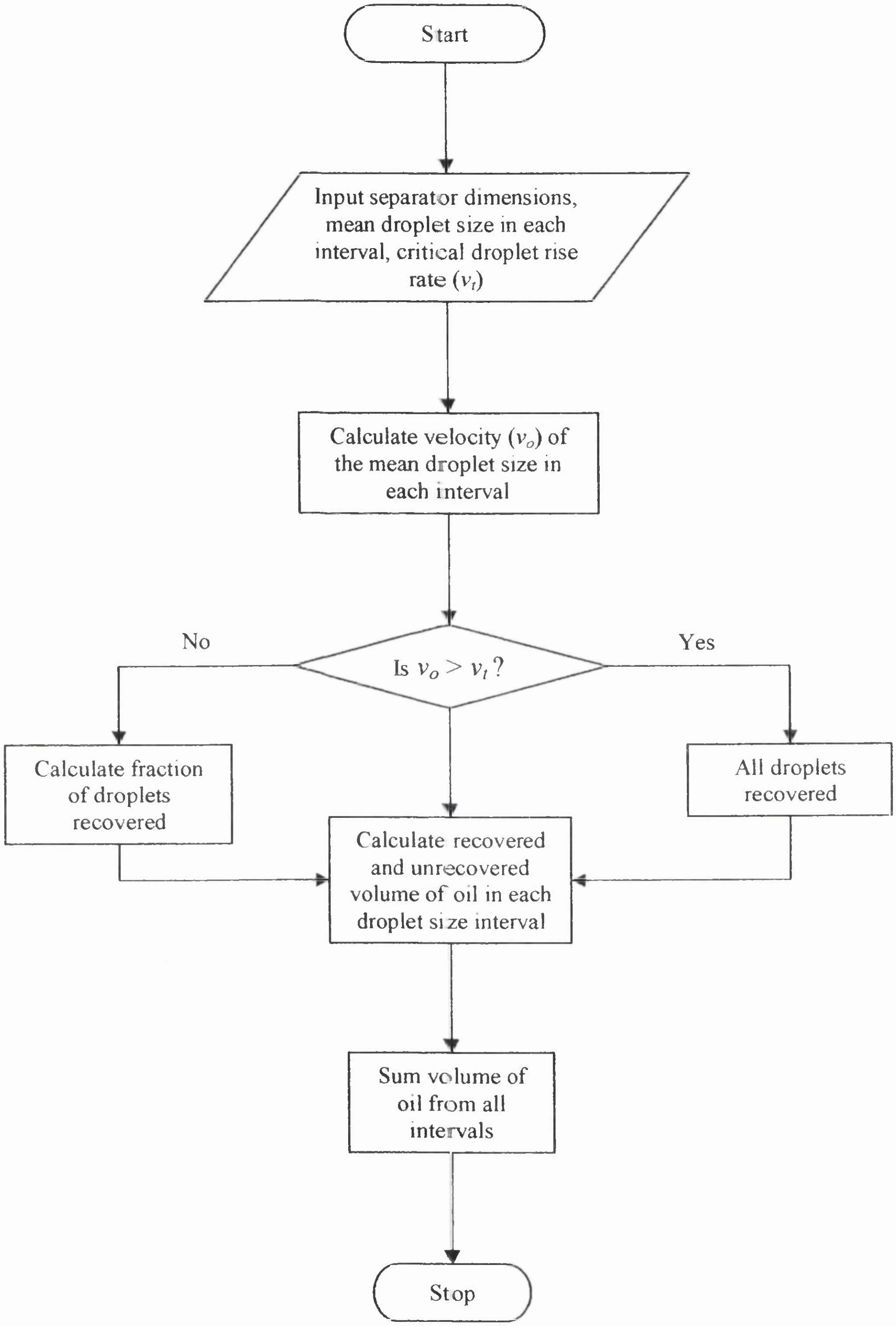


**Figure 3.6: Calculation procedure for oil droplet removal in a CPI**

## 3.8 Bioreactor design procedure

To calculate the bioreactor size required to treat a particular oil concentration in an oily wastewater stream, the spreadsheet requires specified values of wastewater flowrate, inlet oil concentration, the molecular formula of the oil, water density, average oil density, and mass transfer coefficient (Figure 3.7).

The spreadsheet first determines the oxygen requirement of the oil-water mixture using Equation 2.3. The specified parameter for this calculation is the oil concentration of the oil-water mixture.

The next set of calculation procedures after the oxygen requirement has been determined is the calculation of the oxygen transfer rate (Equation 2.9a) and the bioreactor size (2.9). For the calculation of the oxygen transfer rate, the specified inputs are the mass transfer coefficient, maximum oxygen solubility in water at 25°C, and the percentage saturation of oxygen in the bioreactor. Based on the oxygen transfer rate the spreadsheet calculates the bioreactor size to match oxygen supply to oxygen demand.

### 3.8.1 Power requirements of the blower

The spreadsheet estimates the blower horsepower requirements based on a calculated air volume value and an estimated discharge pressure (Figure 3.8).

In this research, the blower discharge pressure was estimated based on an assumed pressure loss in the aeration equipment, main piping, and diffuser.

First the spreadsheet calculates the air volume required to meet oxygen demand using Equation 2.11. Using the specified discharge pressure, mechanical efficiency, and calculated air volume the spreadsheet calculates the power requirements using Equation 2.10.

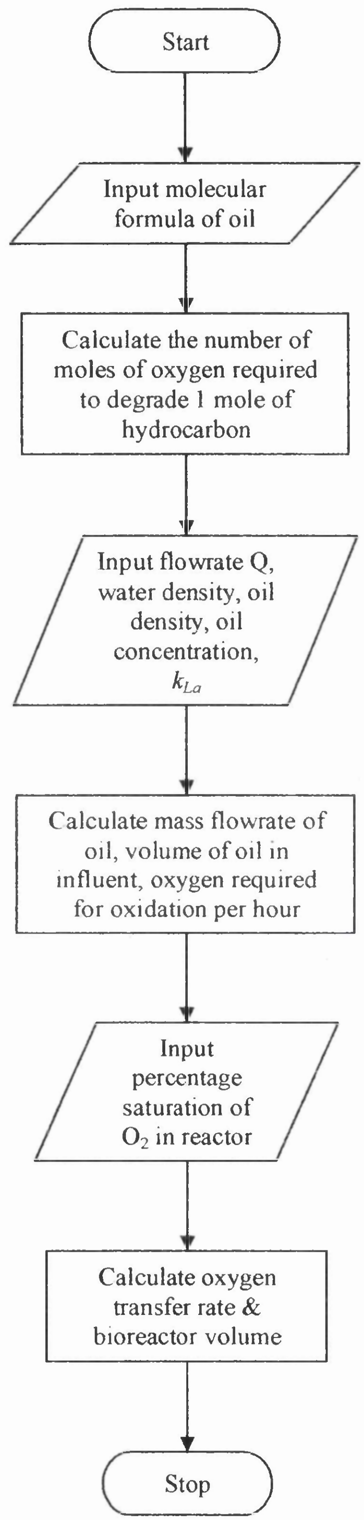


**Figure 3.7: Bioreactor design calculation procedure**

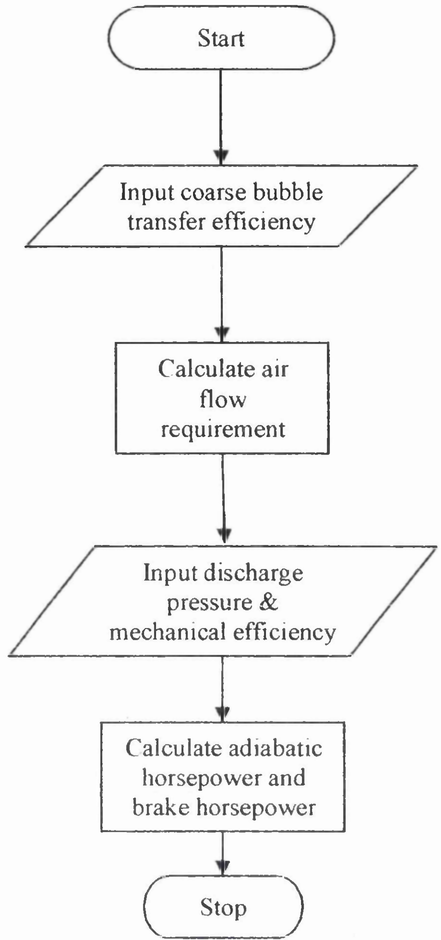


**Figure 3.8: Blower size calculation procedure**

# Chapter 4: Results and discussion - Aeration in various ICB models

## 4.1 Introduction

This chapter reports the results of aeration in various laboratory scale ICB models. Using the column described in Chapter 3, various models of laboratory scale ICB unit were designed by using various sizes of plastic Pall support particles. The mass transfer characteristics of various sizes of plastic Pall support particles in the column were investigated using the methods previously described in Chapter 3 and are presented in Figures 4.1- 4.4. The data used for the various charts are presented in Appendix II. This investigation was necessary to determine the effectiveness of various designs of ICB units in terms of aeration.

## 4.2 Results

The $k_{La}$ values used in Figures 4.1 to 4.4 were determined using methods described in Section 3.4.

Figures 4.1 to 4.4 show the effect of aeration rate on volumetric mass transfer coefficient for the various support particle sizes. For each size of support particle investigated, the volumetric mass transfer coefficient increased as aeration rate increased. Figure 4.1 shows the effect of aeration rate on volumetric mass transfer coefficient for the column filled with 16 mm plastic Pall support particles. The volumetric mass transfer coefficient increased from 0.002 to 0.004 $s^{-1}$ as aeration rate increased from 9 to 48 ml/s. For the ICB unit filled with 25 mm plastic Pall support particles, the volumetric mass transfer coefficient increased from 0.002 to 0.0043 $s^{-1}$ as aeration rate increased from 13 to 56 ml/s (Figure 4.2). The volumetric mass transfer coefficient of the ICB unit filled with 38 mm plastic Pall support particles, increased from 0.003 to 0.0097 $s^{-1}$ as aeration rate increased from 9 to 47.6 ml/s (Figure 4.3). For the ICB unit filled with 50 mm plastic Pall support particles (Figure 4.4), the volumetric mass transfer coefficient increased from 0.0015 to 0.0058 $s^{-1}$ as aeration rate increased from 10.5 to 50 ml/s.

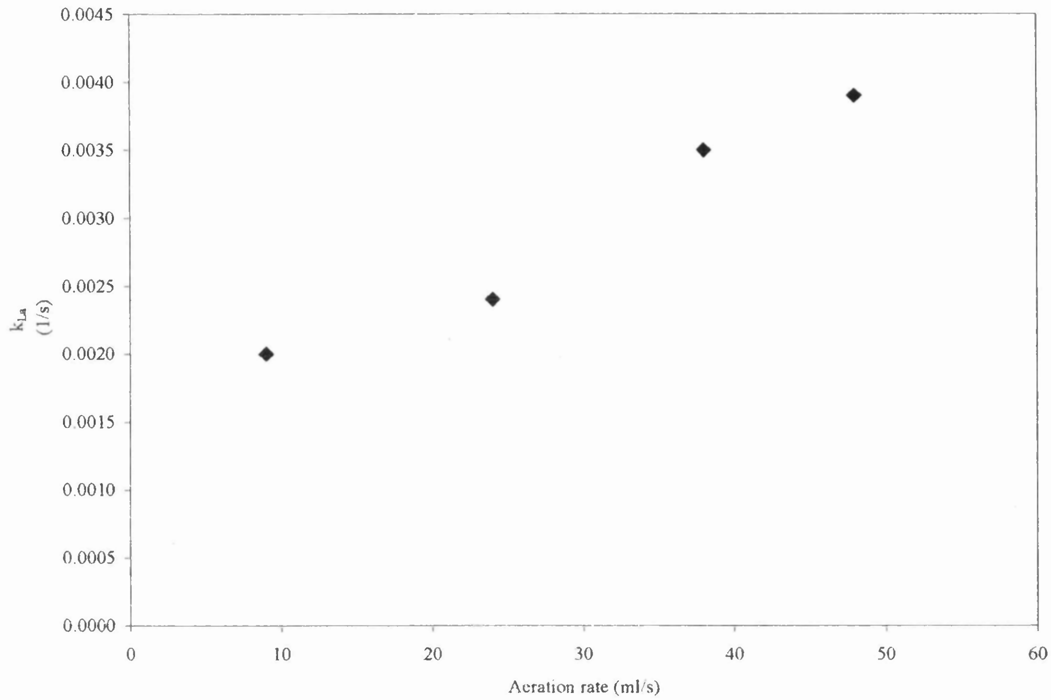


**Figure 4.1: Oxygen MTC vs. aeration rate in ICB model with 16 mm Pall rings**

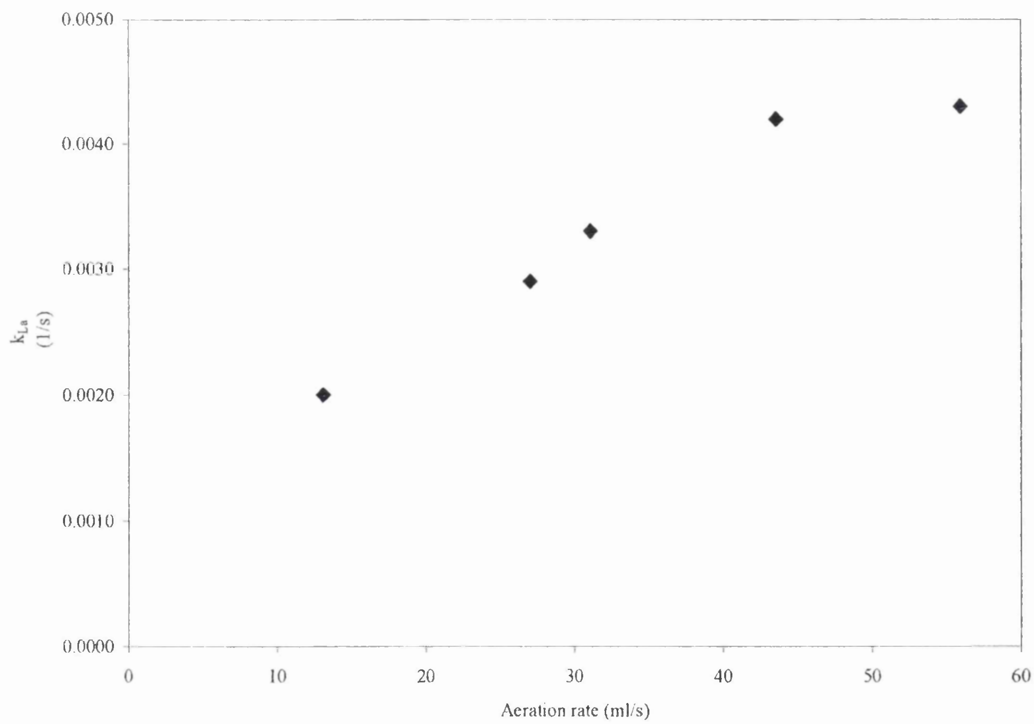


**Figure 4.2: Oxygen MTC vs. aeration rate in ICB model with 25 mm Pall rings**

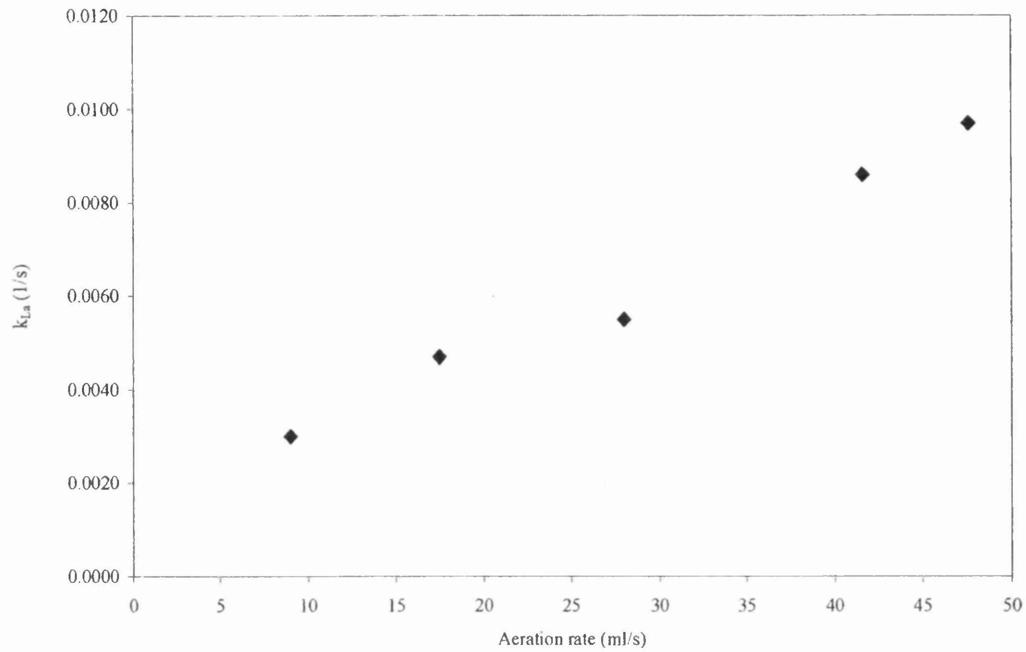


**Figure 4.3: Oxygen MTC vs. aeration rate in ICB model with 38 mm Pall rings**

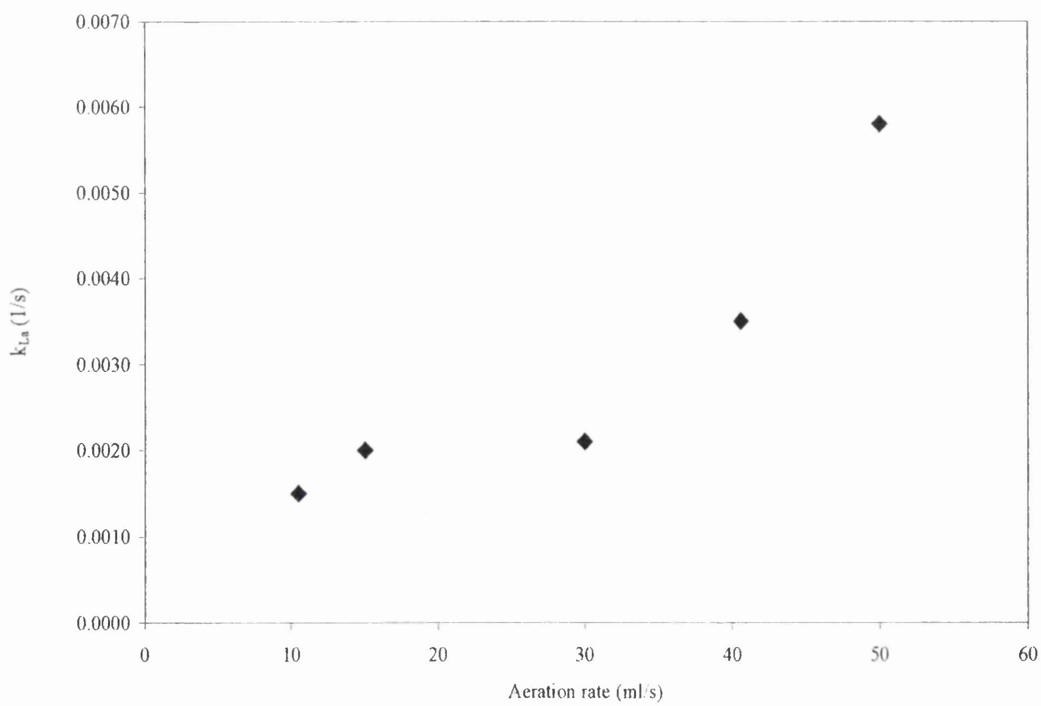


**Figure 4.4: Oxygen MTC vs. aeration rate in ICB model with 50 mm Pall rings**

## 4.3 Discussion

For each of the ICB models, volumetric mass transfer coefficient increased with aeration rate. This was as expected. Nikakhtari and Hill (2005) investigated the mass transfer characteristics of a column with support particles and one without support particles and reported an increase in mass transfer coefficient as superficial air velocity increased in both columns and also observed a higher oxygen mass transfer coefficient in the column with support particles.

The support particles improve the mass transfer characteristics of the column in two ways. Firstly, they break up large air bubbles to smaller bubbles and hence increase the surface area available for mass transfer. Secondly, they prevent larger gas bubbles from escaping rapidly from the column by breaking them into smaller bubbles which have slower rise velocities (Stokes' Law); in other words, the support particles increase the residence time of the air bubbles and hence allow for more oxygen transfer. Nikakhtari and Hill (2005) also reported a partial drop in interstitial velocity of the air bubbles in the presence of support particles.

Increasing aeration rate increases the frequency of bubble formation, therefore increasing gas hold-up and consequently increasing oxygen mass transfer rate. Gas hold-up measurements were not carried out in these investigations but various studies have confirmed this phenomenon (Nikakhtari and Hill 2005, Bhatia *et al.* 2004).

The improvement in mass transfer characteristics as aeration rates increased for each support particle size may also be attributed to increased liquid circulation. Increasing liquid circulation decreases the liquid film thickness and hence improves mass transfer. It can be explained that as aeration rate increased the liquid film thickness reduced therefore improving mass transfer.

Mixing studies were not performed to ascertain the liquid circulation rate. The mixing time for each support particle size would have been indicative of the liquid circulation rate.

### 4.3.1 Comparing aeration in various ICB models

The 38 mm packing had the best mass transfer characteristics compared to the other packing sizes. This was concluded because of the faster rate of rise of its curve compared to the other curves. The 50 mm support particle had better mass transfer characteristics than the 25 mm support particle at higher aeration rates (at about 40 ml/s).

The 16 mm support particles showed the least mass transfer capabilities. Ostergaard (1978) reported better mass transfer characteristics with 6 mm glass ballotini when compared to 1 and 3 mm glasses. This trend was not expected in this investigation considering the internal structure of the support particles (void space) (Figure 4.5 and Figure 4.6).

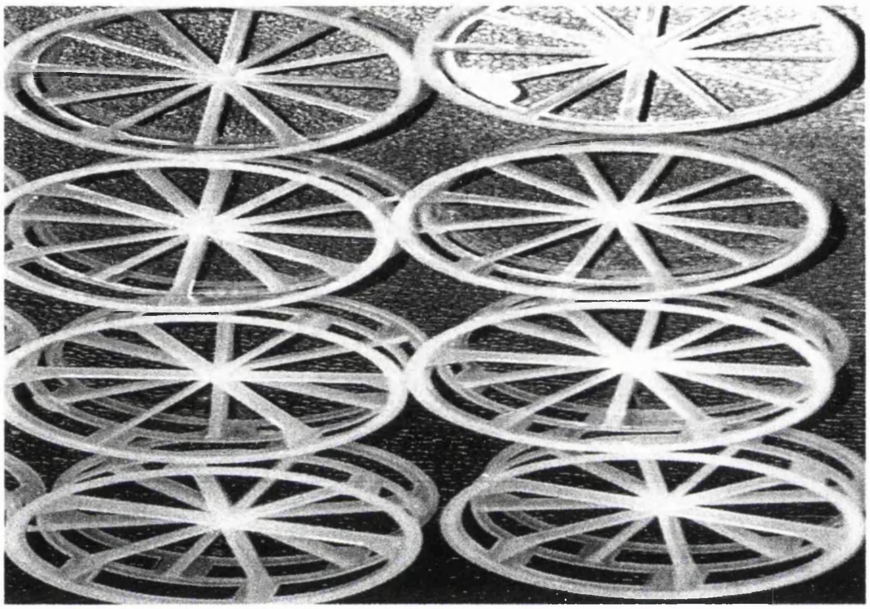

**Figure 4.5: Internal structure of the 50 mm support particle**

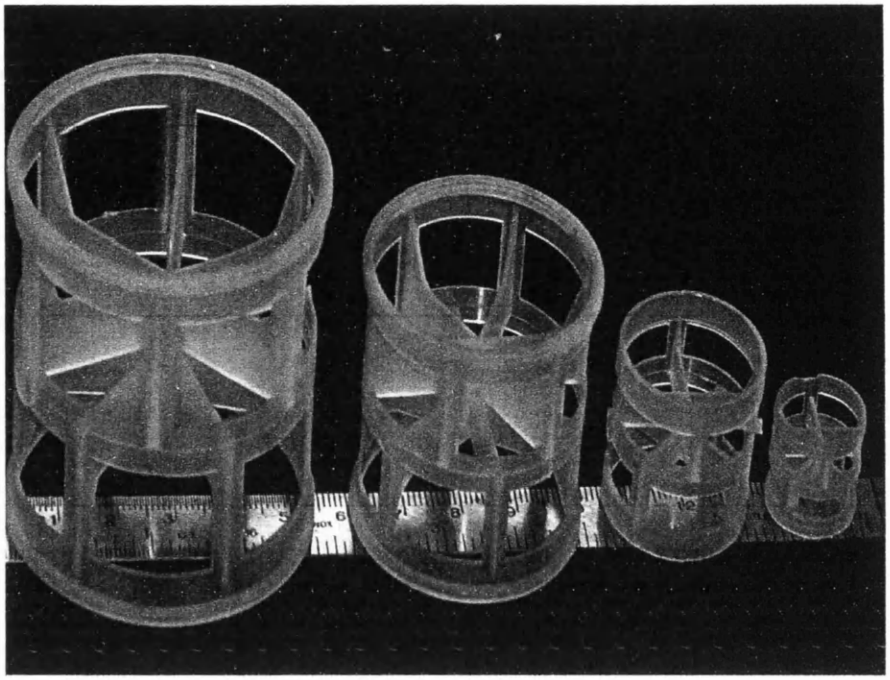

**Figure 4.6: The various sizes of packing (mm): 50, 38, 25, 16 (from left to right)**

The smallest support particle has the smallest void area (Figure 4.5 and Figure 4.6), so it was expected to have a higher probability of disintegrating large bubbles than the bigger support packing (38 mm and 50 mm), and hence better mass transfer capabilities were expected. However, this was not so. The probable reason for the poor mass transfer characteristics of the smaller support particles may be as a result of poor gas circulation due to low bed voidage. Considering that the various ICB models were operated in a semi-batch mode (liquid movement was solely due to gas flow) it follows that liquid circulation was poor when the column was filled with smaller particles (16 mm) due to poor gas circulation or distribution. For better mass transfer, good liquid circulation is needed to reduce the liquid film thickness.

Daraktschiev (1984) characterized the gas flow distribution in a packed bed and reported that particle arrangement affects gas distribution. In this investigation, it was suspected that not only had the arrangement affected the mass transfer but also the number of support particles in the bed. Since the height of the support particles in the column was fixed by the position of the wire mesh the bigger support particles resulted in a higher bed voidage (due to their internal structure and the smaller number of particles in the volume provided for the particles) while the smaller support particles had a low bed voidage due to a higher number of support particles. It can be concluded that the size and number of particles in the fixed bed height are the reasons for the poor mass transfer characteristics of the smaller support particles.

Comparing the mass transfer capabilities of the 50 mm support particle and the 38 mm support particle, the expected trend was observed. The 38 mm support particles had better mass transfer characteristics than the 50 mm support particles. This trend was expected due to the internal structure of the support particles. The void spaces in the 50 mm support particle were very large, so the probability of disintegrating gas bubbles were less as a result a poor mass transfer characteristics when compared to the 38 mm support particle. However, good liquid circulation was observed with both support particle sizes.

Comparing the mass transfer characteristics of the smaller 16 mm support particle and the 25 mm support particle, the 25 mm support particle showed better mass transfer characteristics. The probable reasons may be as discussed earlier, poor liquid circulation in the ICB column filled with 16 mm support particle.

## 4.4 Conclusions

In order to reduce cost due to aeration, aeration in different designs of ICB models was investigated. These results show that design affects aeration. Design of the ICB includes the type, arrangement, and number of support particles in the unit. The 38 mm plastic Pall ring had the best mass transfer characteristics and may be said to be the particle size which best improves the cost of aeration; other factors also contribute to it, such as arrangement and internal structure. These results do not agree with those of Ostergaard 1978. Ostergaard 1978 investigated mass transfer characteristics of various sizes of glass ballotini and reported that the largest glass size had the best mass transfer characteristics. Lee and Buckley (1981) also reported that large diameters such as 4 mm and 6 mm are beneficial to aeration. One thing remains clear: the support particle size affects aeration.

In conclusion, these studies have shown that the physical structure of the ICB (or design) affects aeration and the solution to reducing cost due to aeration may lie in the design of the ICB unit rather than the use of oxygen *in situ*.

# Chapter 5: Results and Discussion – Optimizing Oil Recovery

## 5.1 Introduction

This chapter presents the results obtained from various scenarios where the effectiveness of oil recovery units and their effects on the integrated process were investigated.

The overall aim of this chapter is to provide a solution to improving oil recovery and reducing the cost of treatment in the integrated processes studied.

Section 5.2 presents the results of the various scenarios studied to investigate the effectiveness of the horizontal flow separator to recover oil and its effect on the integrated process, followed by an analysis and a discussion sub-section.

Section 5.3 presents the results of the various scenarios studied to investigate the effectiveness of the Corrugated Plate Interceptor to recover oil and its effect on the integrated process, followed by an analysis and a discussion sub-section.

In the presentation of these results, "bioprocess" refers to the bio-treatment and aeration, which in this case is achieved with an ICB and a blower; "separation process" refers to the gravity separation unit used in each integrated process and the skimming of the oil recovered. The overall cost of the plant is the cost of the plant less the revenue generated from recovered oil. The unit cost of oil was an estimated value (see Chapter 2).

## 5.2 Integrated Process of Horizontal Flow Separator and ICB

A design exercise was carried out using the models previously described (see Sections 2.3.1, 2.3.2, 2.3.4, 2.3.5, 3.6, and 3.7). Using various simulated oily wastewater systems, the effects of varying influent characteristics, such as oil concentration and average oil droplet size, on the cost per year of a horizontal separator and oil recovery per hour were studied. The effects of other factors associated with design of the separator were also investigated. For each factor investigated, a volume of oily water was fed into the separator and the performance predicted (see Section 3.6.1). For each case, the size of Immobilized Cell Bioreactor (ICB) required to treat the effluent from the separator was calculated based on the oxygen demand of the effluent (see Section 2.3.2.2). The costs of constructing and operating the separator and the ICB were then calculated in each case.

The separator dimensions, influent characteristics, value of oil recovered, and cost of each unit are presented in Appendix III.

### 5.2.1 Effect of varying influent oil concentration

The effectiveness of the horizontal flow separator to recover oil from influent streams with various oil concentrations was investigated using the models previously described in Chapters 2 and 3. Several values of oil concentration were used whilst holding other influent characteristics and separator dimensions constant. Separator dimensions, influent characteristics, and results are presented in Appendix III. The cost of the treatment is presented below.

Figure 5.1 shows the effect of influent oil concentration on the cost of separation. As the influent oil concentration is increased from 1000 to 1500 mg/l, the amount of oil treated increases from 40 to 60 tonnes per year and the cost of separation decreases from £35.53 to £31.01 per year. This decrease in cost is as a result of the increased revenue due to the greater volume of oil recovered. The volume of oil recovered increases from 0.0036 to 0.0055 $m^3/h$ (Figure 5.2), and the revenue from recovered oil increases from £9.03 to £13.55 per year (Figure 5.3) as the inlet oil concentration increases.

Figure 5.4 shows the relationship between oil concentration and separator efficiency. Increasing the influent oil concentration by does not affect separation efficiency; efficiency remains constant at 29%.

The relationships between oil concentration and oxygen demand, bioreactor volume and cost of the bioprocess are presented in Figures 5.5, 5.6 and 5.7. As oil concentration in the separator influent is increased, the amount of oxygen required to oxidize the oil arriving at the bioreactor increases from 2.46 to 3.69 g/l.h; consequently the bioreactor size required for complete oxidation increases from 113 to 170 $m^3$, (Figures 5.5 and 5.6). The effect of oil concentration on the cost of the integrated process is shown in Figure 5.8. As oil concentration is increased, the cost of the integrated process increases from £5189.25 to £7648.56 per year.

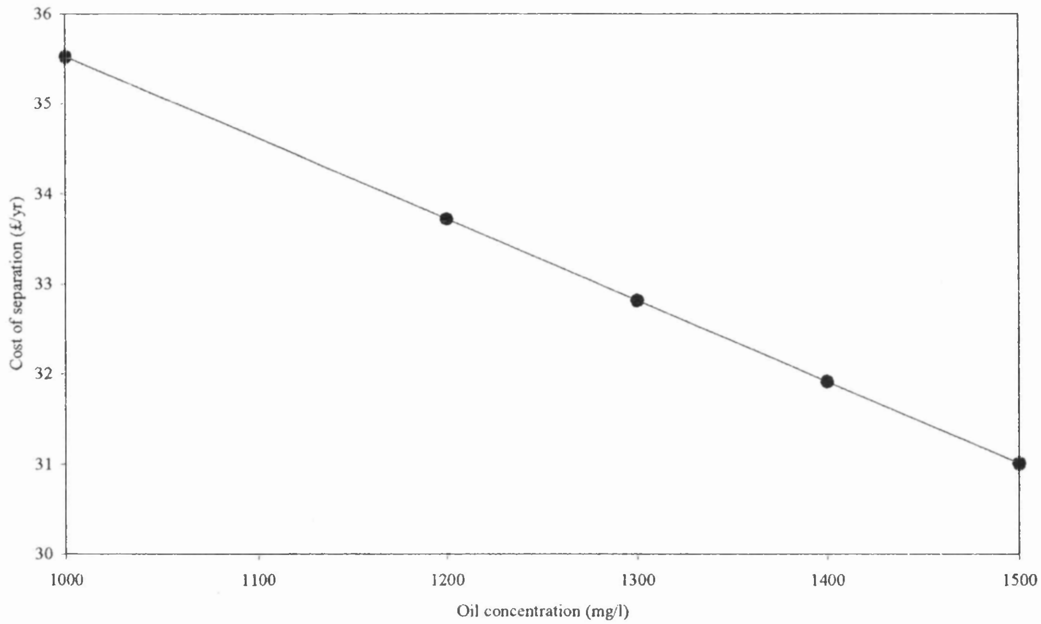


**Figure 5.1: Effect of oil concentration on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

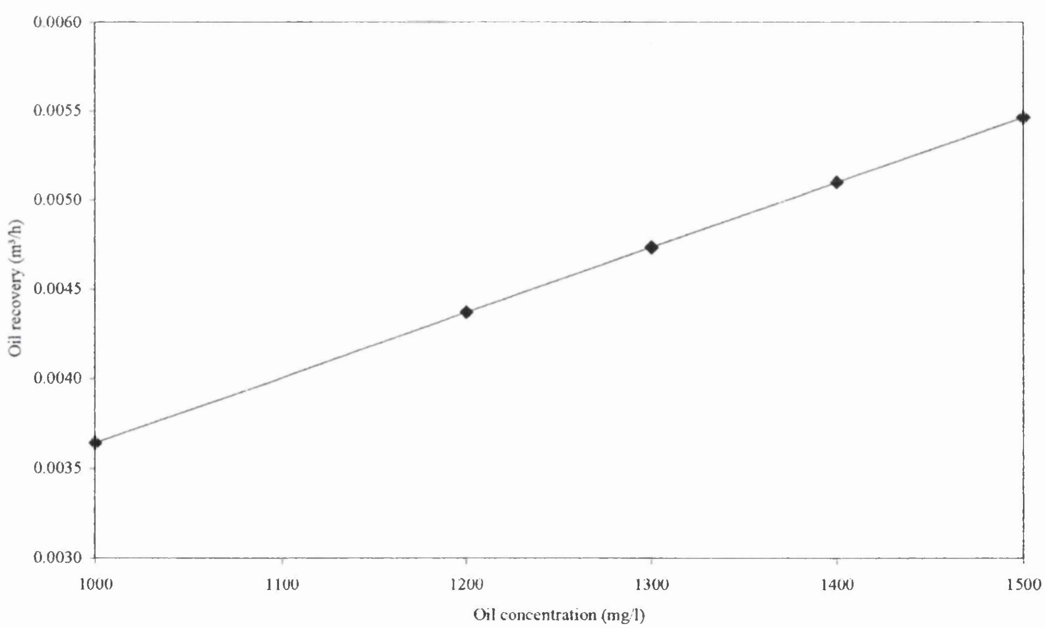


**Figure 5.2: Effect of oil concentration on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

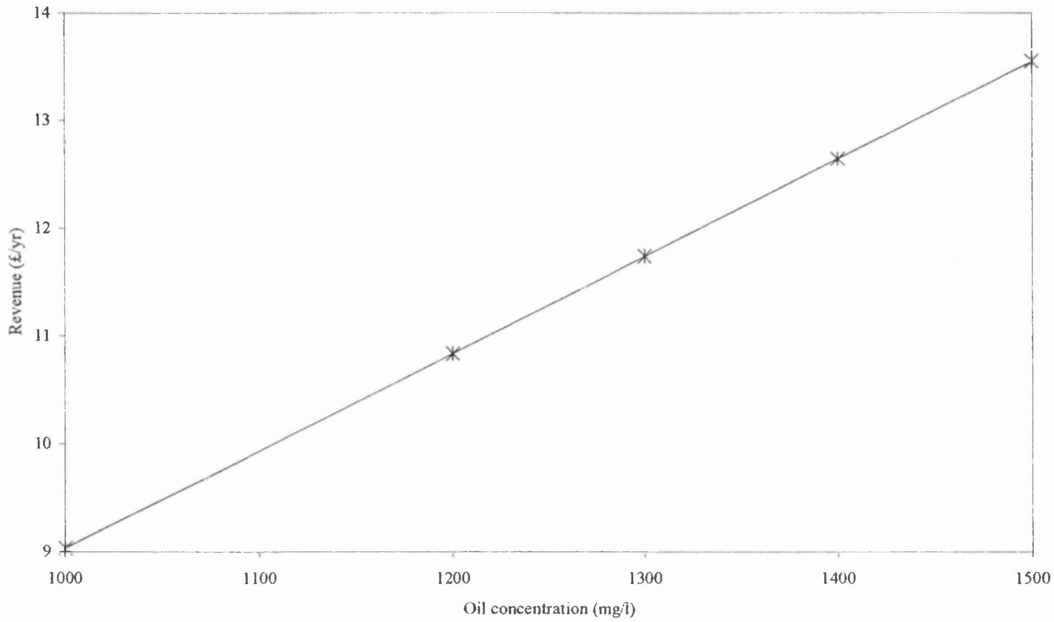


**Figure 5.3: Effect of oil concentration on revenue generated from recovered oil**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

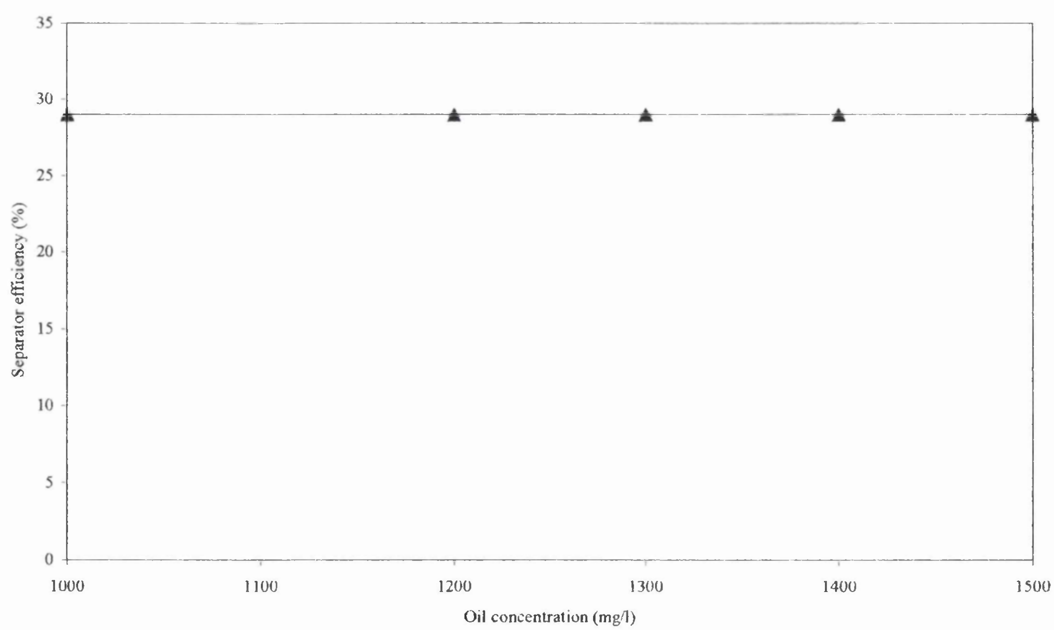


**Figure 5.4: Effect of oil concentration on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

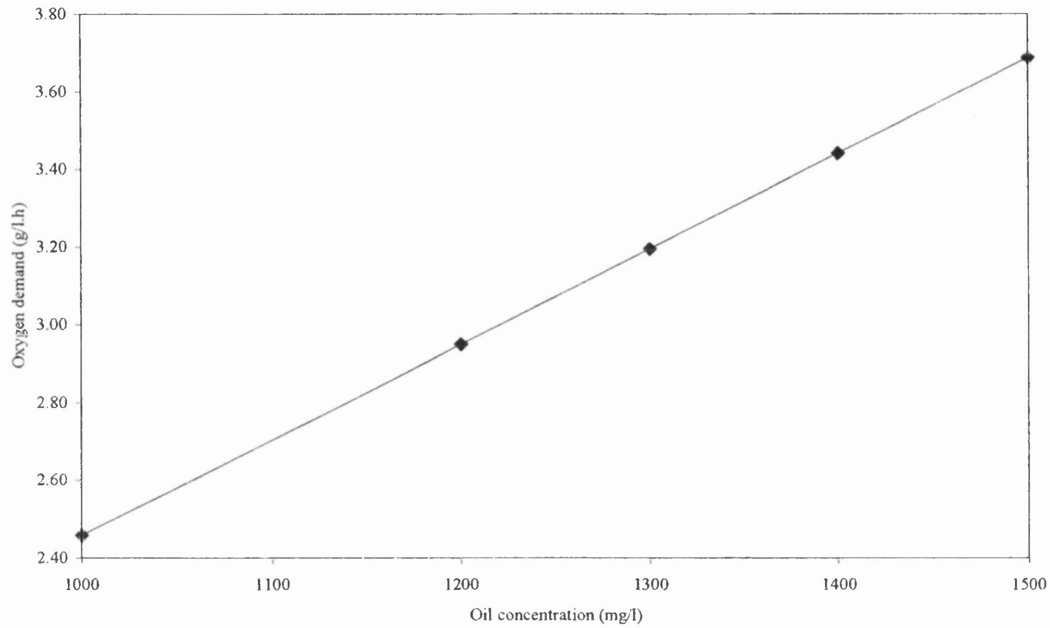


**Figure 5.5: Effect of oil concentration on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

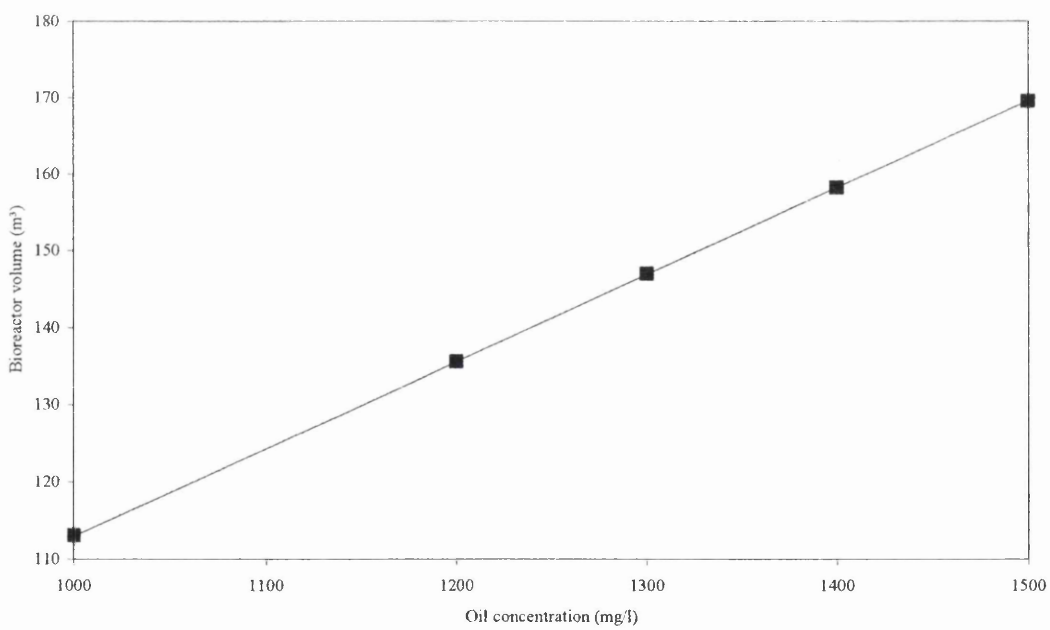


**Figure 5.6: Effect of oil concentration on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

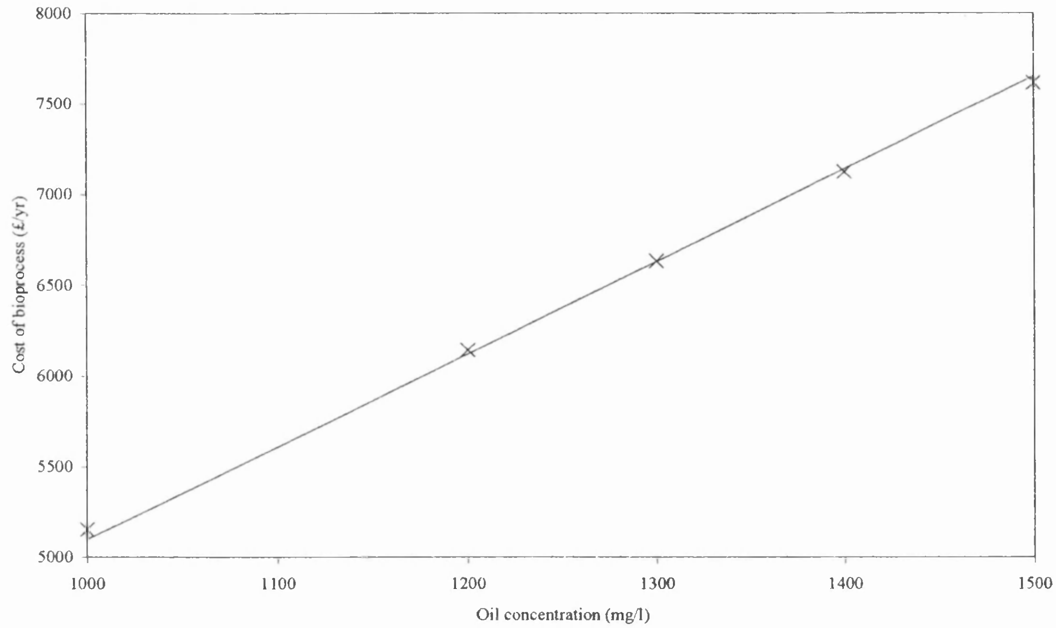


**Figure 5.7: Effect of oil concentration on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

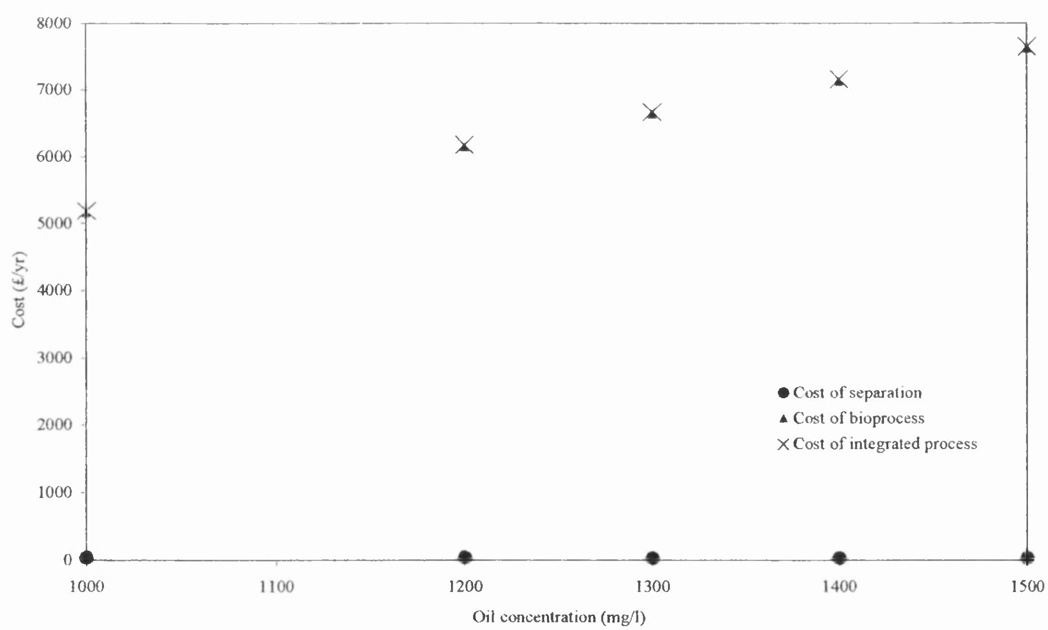


**Figure 5.8: Effect of oil concentration on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.2 Effect of varying separator horizontal velocity

The effectiveness of using a horizontal flow separator operating at various horizontal velocities was investigated using the models previously described in Chapters 2 and 3. Horizontal flow velocity was varied whilst holding separator dimensions and influent characteristics constant. Separator dimensions, influent characteristics, and results are presented in Appendix III. The cost of treatment is presented below.

Figure 5.9 shows the effect of horizontal flow velocity of the separator on the cost of separation. As the horizontal flow velocity is reduced from 0.0049 to 0.0005 m/s, the cost of separation decreases from £35.53 to £13.69 per year. Figures 5.10 and 5.11 show that as horizontal velocity is reduced, the volume of oil recovered increases from 0.004 to 0.012 $m^3$/h and hence the revenue generated from recovered oil increases from £9.03 to £30.87 per year. Figure 5.12 shows the relationship between horizontal velocity in the separator and separator efficiency. A 90% decrease in horizontal velocity improves the efficiency of the separator by 242% while a 50% decrease in horizontal velocity improves the efficiency by 98%.

Relationships between horizontal velocity and oxygen demand, bioreactor volume and cost of bioprocess are presented in Figures 5.13, 5.14 and 5.15. As horizontal velocity in the upstream separator is reduced, oxygen demand decreases from 2.46 to 0.01 g/l.h and bioreactor volume decreases from 113 to 1 $m^3$ (see Figures 5.13 and 5.14). Decreasing horizontal velocity therefore reduces both the cost of the separator and the bioprocess. In Figure 5.15, as horizontal velocity is reduced, the cost of the bioprocess decreases from £5153.72 to £43.70 per year. The overall effect of decreasing horizontal velocity on the integrated process is shown in Figure 5.16, where reducing horizontal velocity decreases the cost of the integrated process from £5189.25 to £57.39 per year.

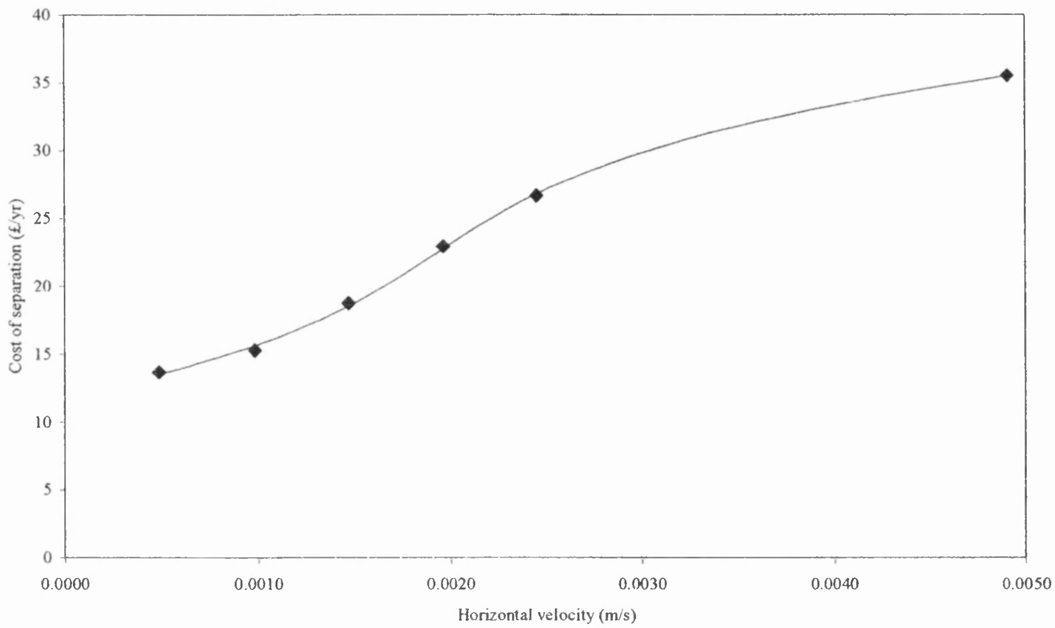


**Figure 5.9: Effect of horizontal velocity on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

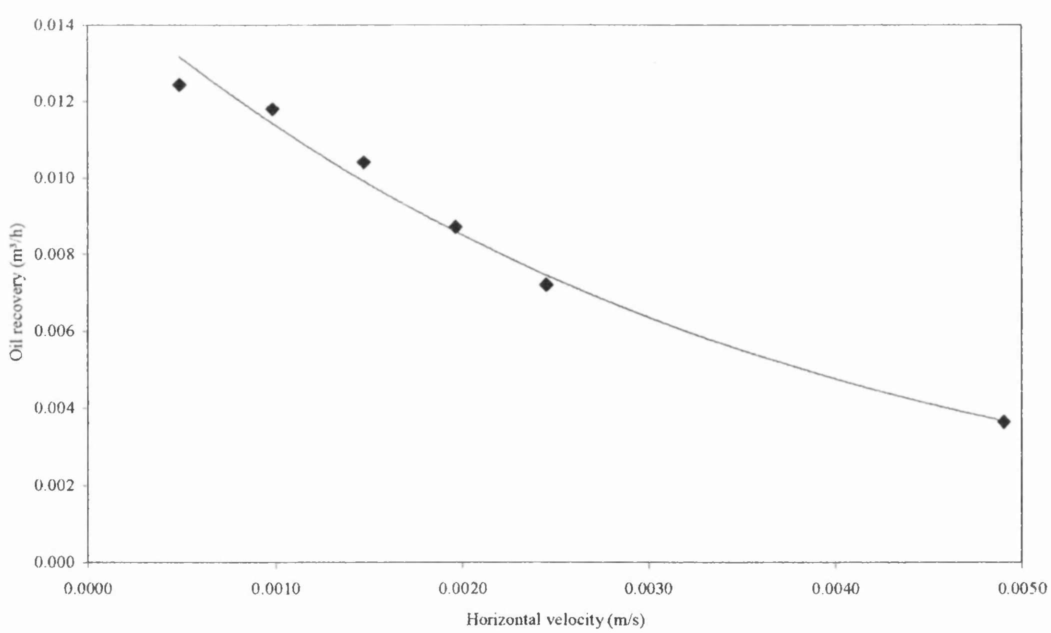


**Figure 5.10: Effect of horizontal velocity on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

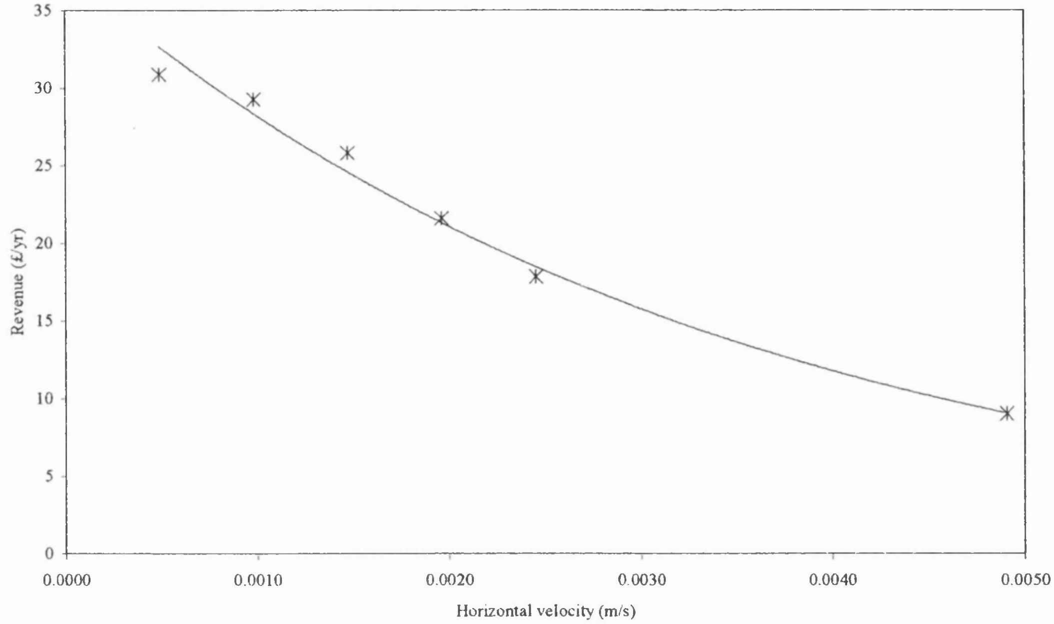


**Figure 5.11: Effect of horizontal velocity on revenue generated from recovered oil**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

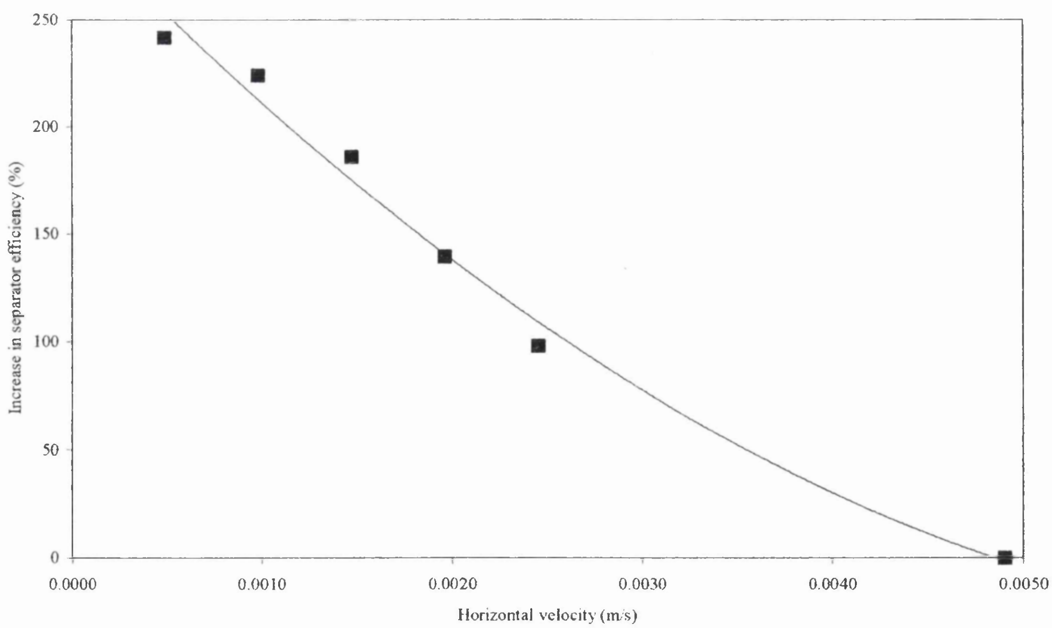


**Figure 5.12: Effect of horizontal velocity on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

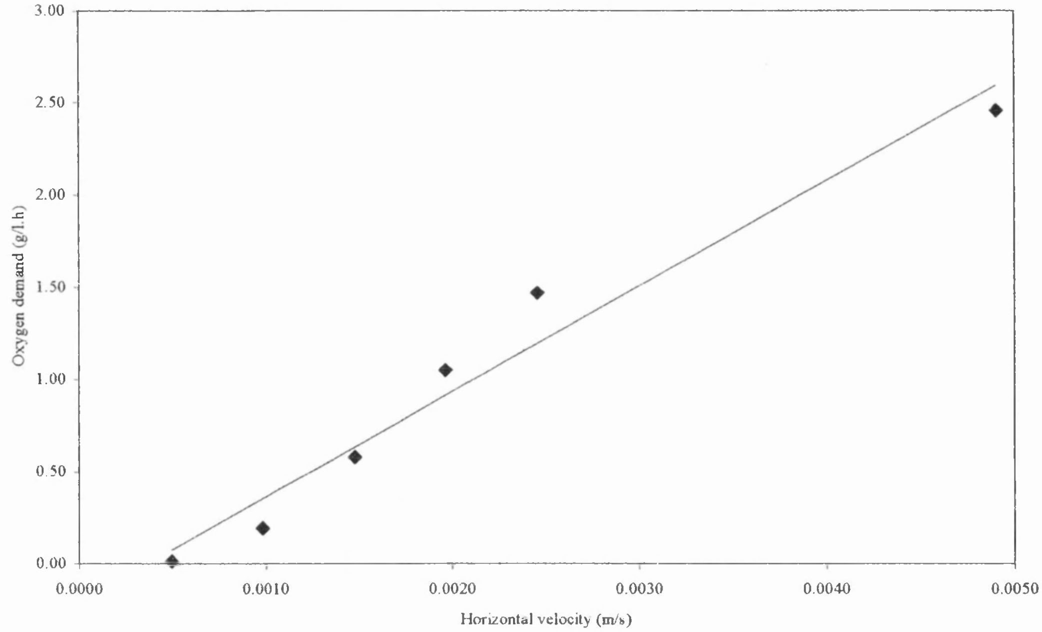


**Figure 5.13: Effect of horizontal velocity on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

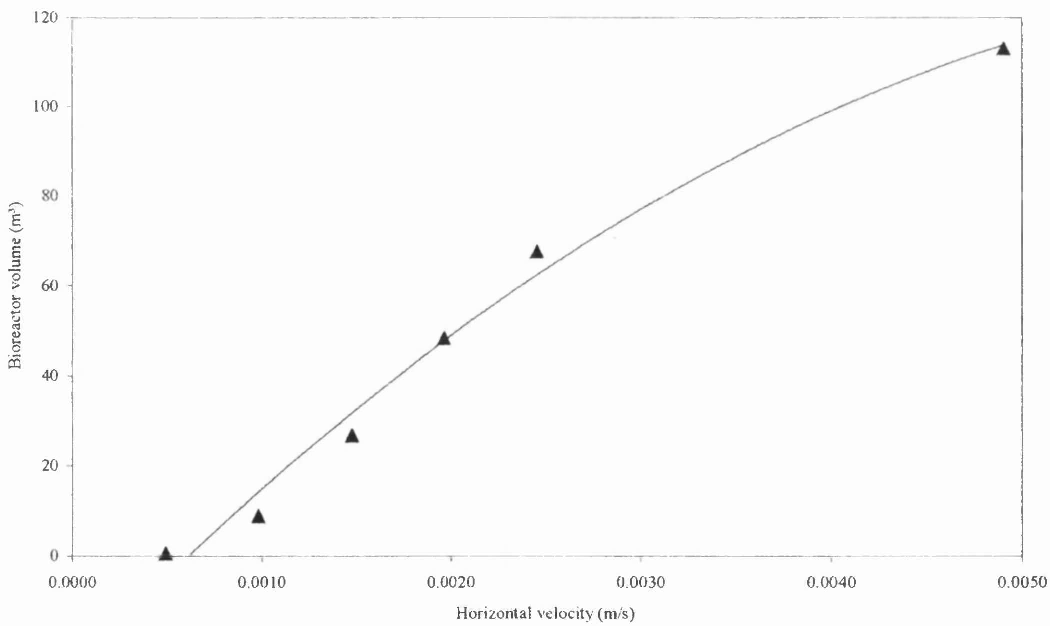


**Figure 5.14: Effect of horizontal velocity on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

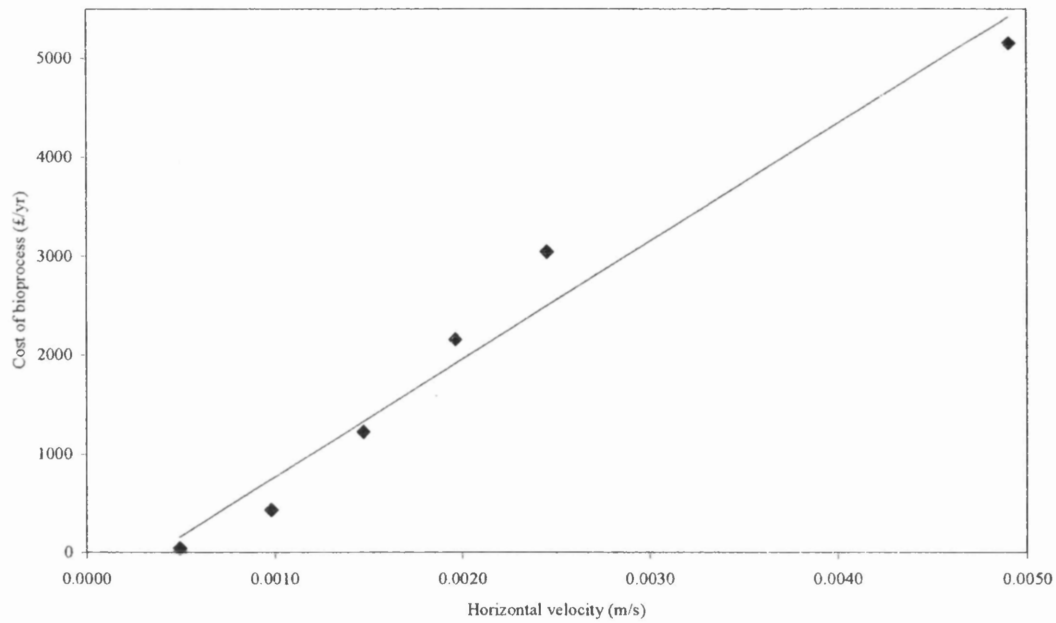


**Figure 5.15: Effect of horizontal velocity on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

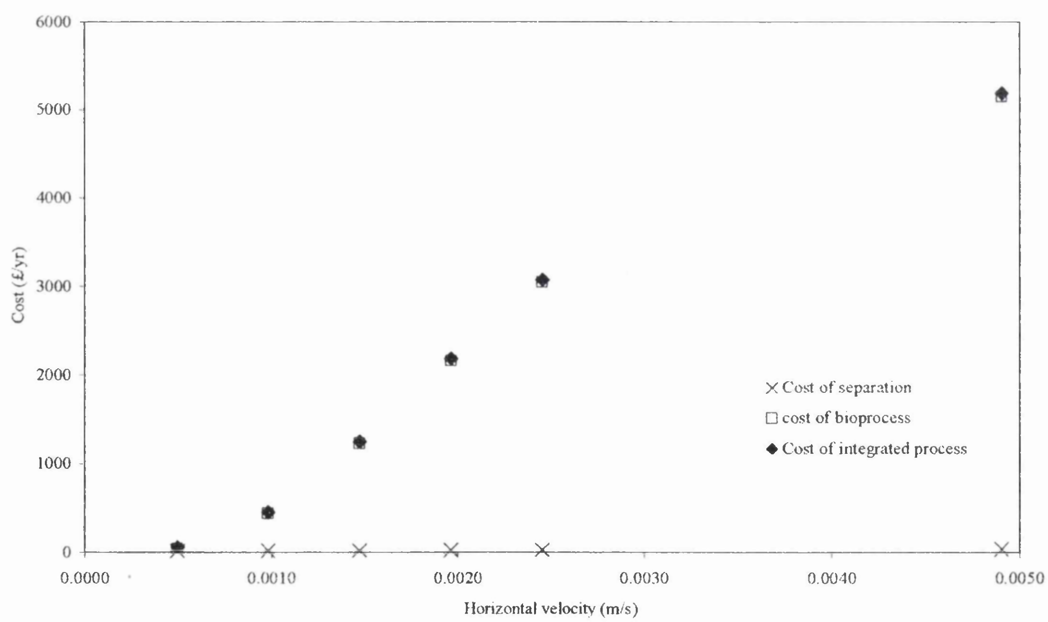


**Figure 5.16: Effect of horizontal velocity on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.3 Effect of varying separator volume

The effectiveness of using horizontal flow separators of various volumes to treat an oily wastewater stream was investigated using the models previously described in Chapters 2 and 3. Separator length was varied whilst holding other dimensions and influent characteristics constant. Separator dimensions, influent characteristics, and results are given in Appendix III. The cost of treatment is presented.

Figure 5.17 shows the effect of separator volume on the cost of separation. As separator volume is increased from 4 to 47 $m^3$, the cost of separation increases from £35.53 to £162.13 per year, despite the fact that oil recovery increases from 0.004 to 0.012 $m^3/h$ (Figure 5.18) and the revenue generated increases from £9.03 to £30.92 per year, see Figure 5.19. Figure 5.18 shows that beyond a certain size, increase in separator volume does not result in a significant improvement in oil recovery; this may be attributed to the presence of very small oil droplets which have very slow rise rates.

Figure 5.20 shows the relationship between separator volume and separator efficiency. Increasing separator volume results in a significant improvement in efficiency but beyond a volume increase of about 400%, the increase in separator efficiency, and hence improvement in oil recovery, is not significant.

Figures 5.21, 5.22, 5.23 show the relationships between the volume of the upstream separator and oxygen demand, bioreactor volume and cost of the bioprocess. As separator volume is increased, oxygen demand decreases from 2.46 to 0.01g/l.h (Figure 5.21) and bioreactor volume decreases from 113 $m^3$ to 400 litres (Figure 5.22); the cost of the bioprocess decreases from £5153.72 to £28.97 per year, Figure 5.23. From Figure 5.23, increasing separator volume beyond a certain size (about 23 $m^3$) does not result in a significant reduction in the cost of the bioprocess; optimization therefore becomes very necessary.

The effect of separator size on the cost of the integrated process is shown in Figure 5.24. As separator volume is increased, the cost of the separation process increases but the cost of the integrated process decreases from £5189.25 to £191.09 per year.

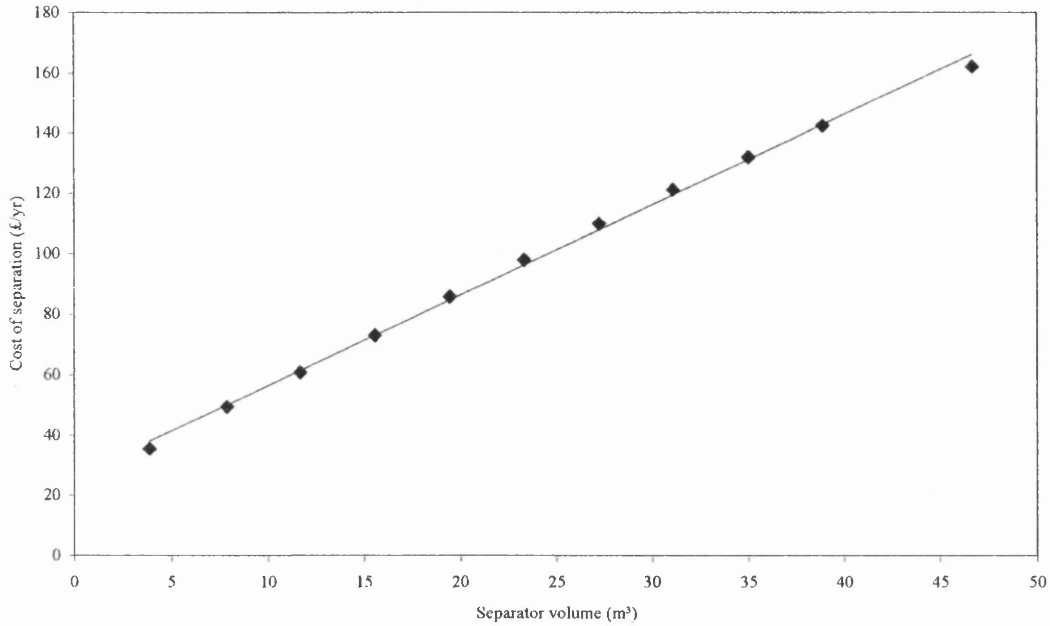


**Figure 5.17: Effect of separator volume on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

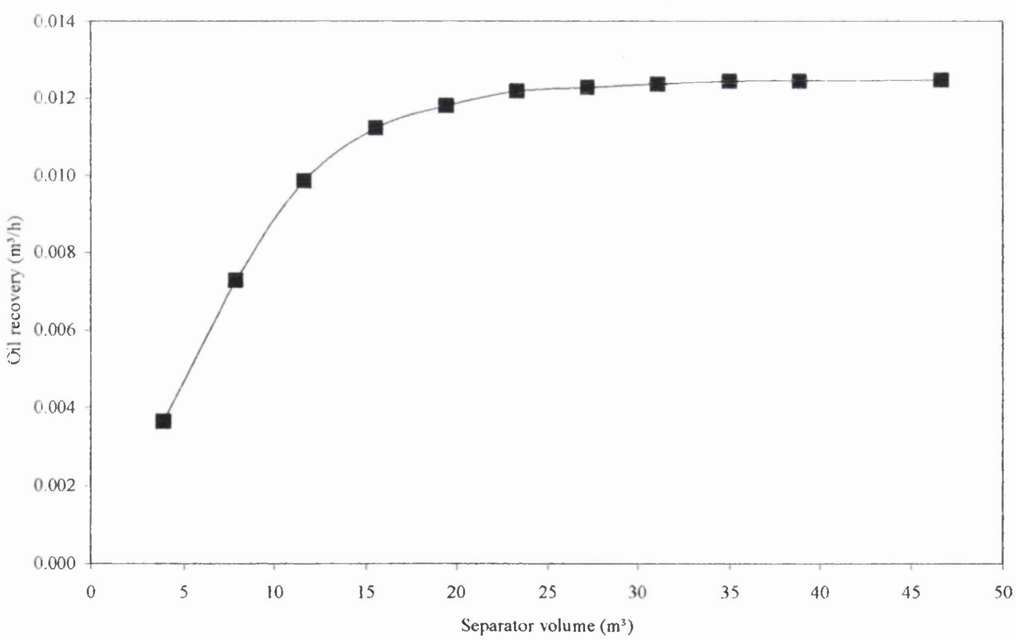


**Figure 5.18: Effect of separator volume on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

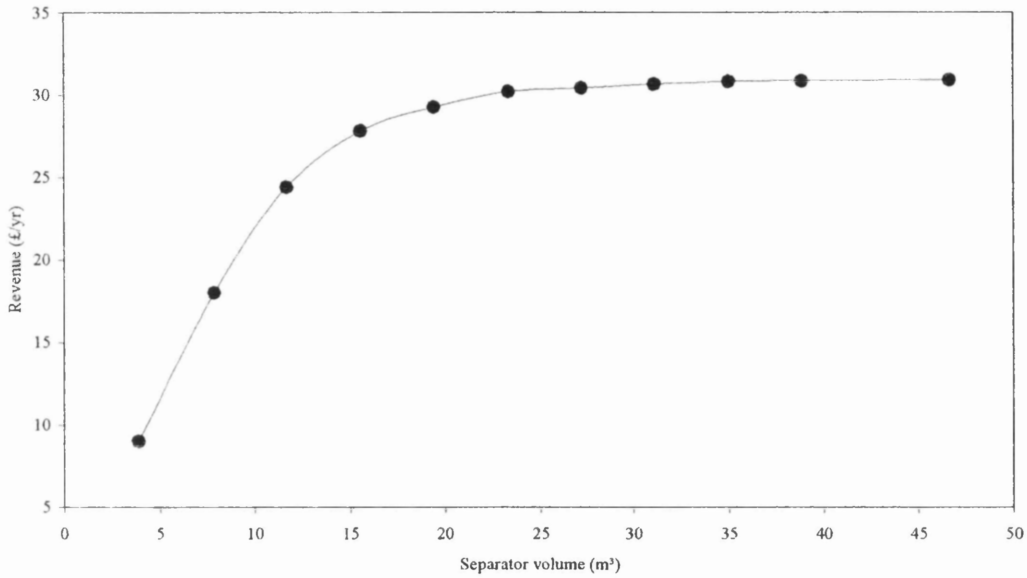


**Figure 5.19: Effect of separator volume on revenue generated from recovered oil**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

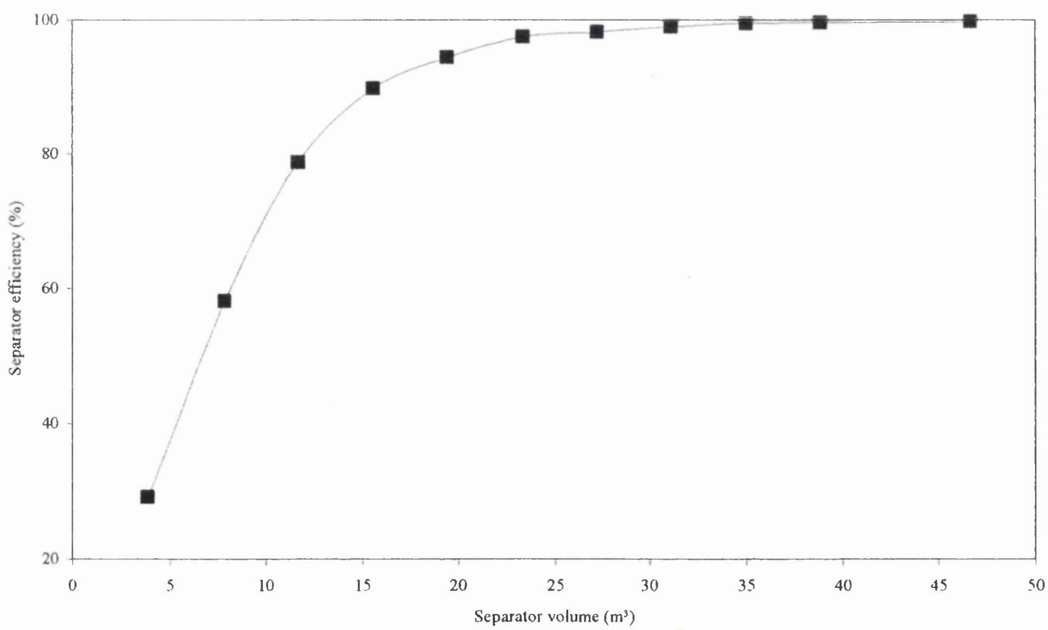


**Figure 5.20: Effect of separator volume on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

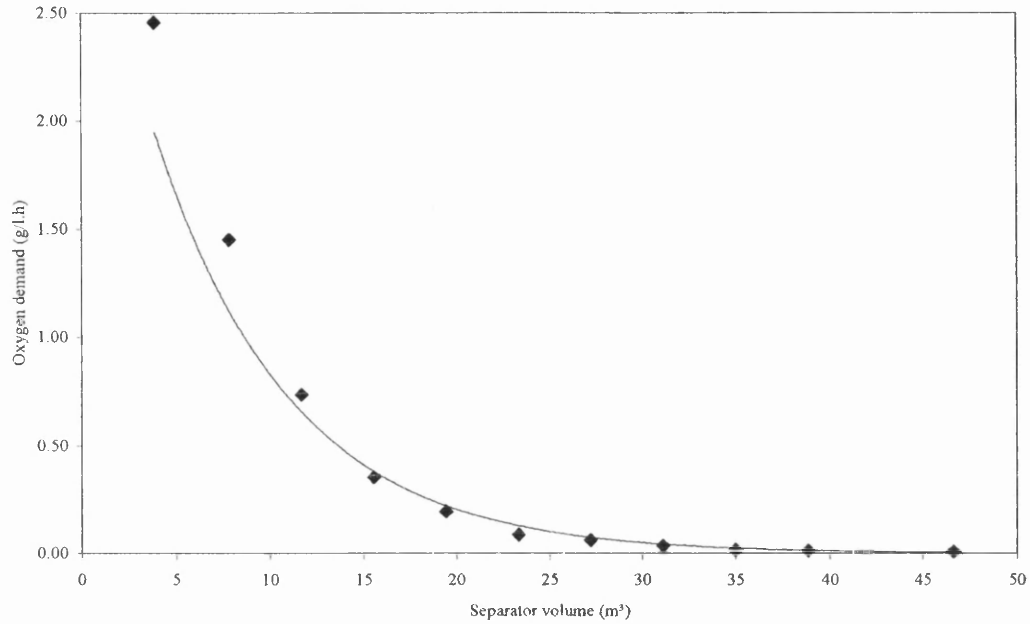


**Figure 5.21: Effect of separator volume on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

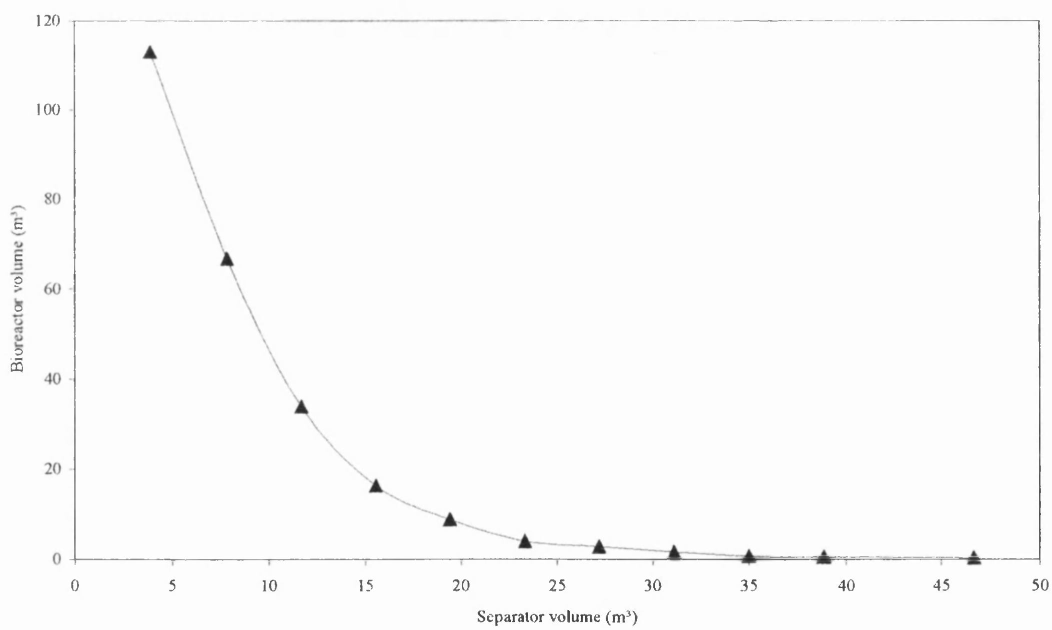


**Figure 5.22: Effect of separator volume on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

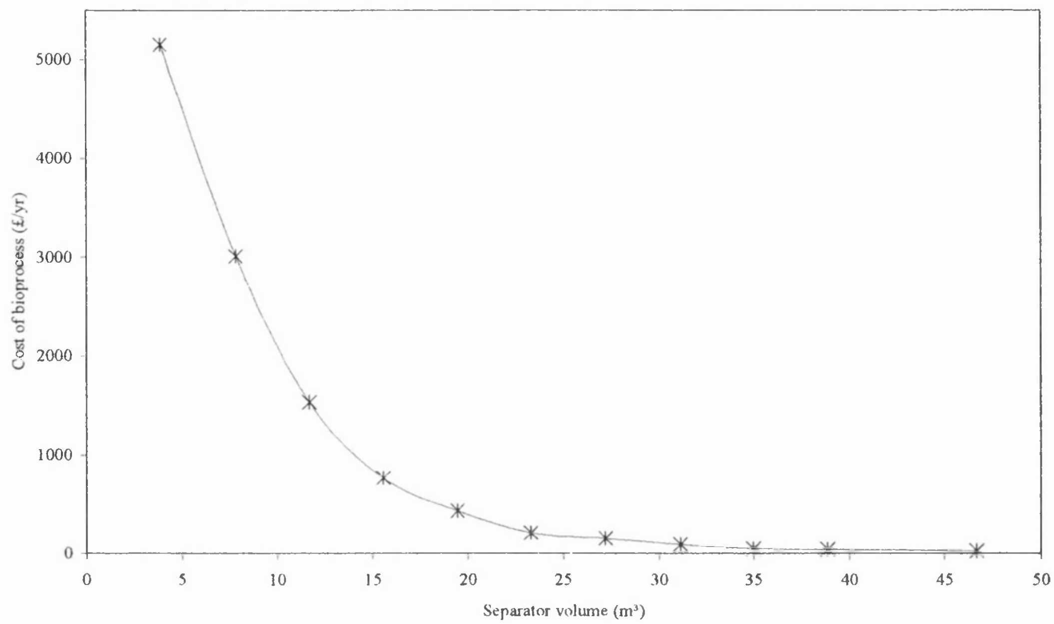


**Figure 5.23: Effect of separator volume on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

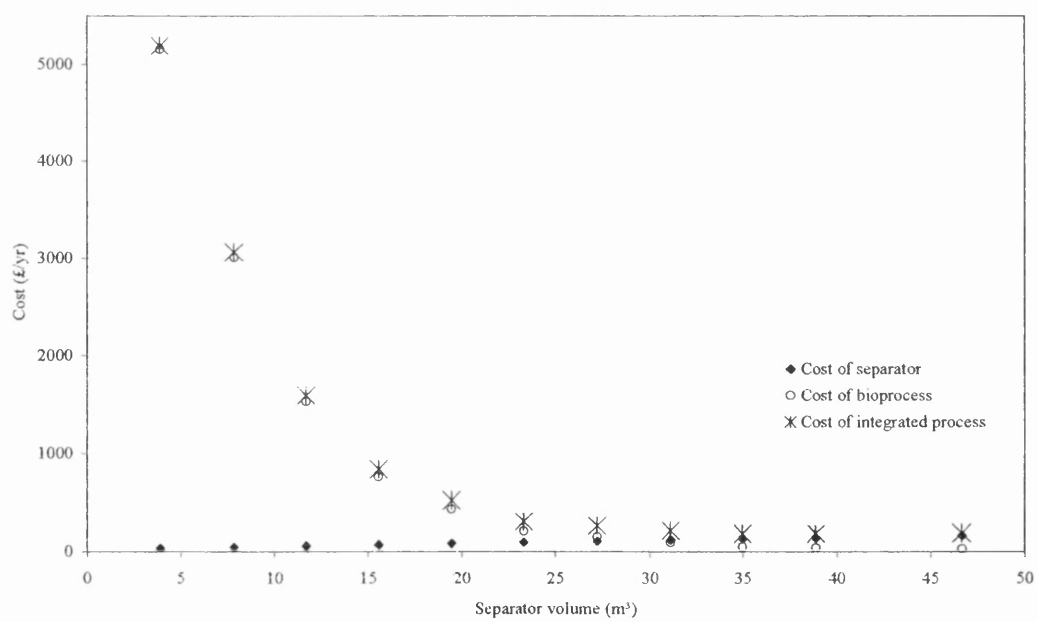


**Figure 5.24: Effect of separator volume on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.4 Effect of varying average droplet size

The effectiveness of using a horizontal flow separator to recover oil from oily wastewater streams with varying average droplet sizes was investigated using the models previously described in Chapters 2 and 3. The average oil droplet size was varied whilst holding other influent characteristics and separator dimensions constant. Separator dimensions, influent characteristics, and results are presented in Appendix III. The cost of treatment is presented below.

Figure 5.25 shows the effect of average droplet size on cost of the separation. As average droplet size is increased from 60 to 150 microns the cost of separation decreases from £35.53 to £13.78 per year; oil recovery increases from 0.0036 to 0.0124 $m^3/h$ (Figure 5.26) and hence the revenue generated from recovered oil increases from £9.03 to £30.78 per year (Figure 5.27). Figure 5.28 shows that as average droplet size is increased from 60 to 150 microns the base separator efficiency increases from 29 to 99%.

The effect of droplet size on the oil oxidation part of the process is presented in Figures 5.29, 5.30 and 5.31. In Figure 5.29, as average droplet size is increased oxygen demand decreases from 2.46 to 0.02 g/l.h; the volume of the bioreactor reduces from 113 $m^3$ to 1 $m^3$ (Figure 5.30) and hence the cost of the bioprocess decreases from £5153.72 to £67.25 per year (Figure 5.31). Figure 5.32 shows the effect of average droplet size on the cost of the integrated process. As average droplet size is increased, the cost of the integrated process decreases from £5189.25 to £81.03 per year.

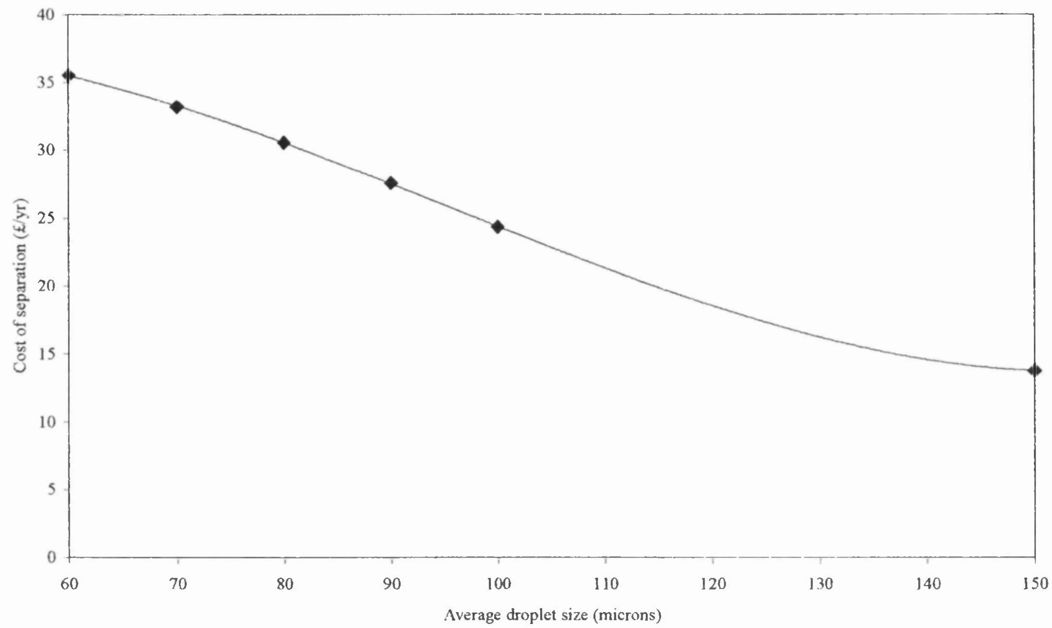


**Figure 5.25: Effect of average droplet size on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

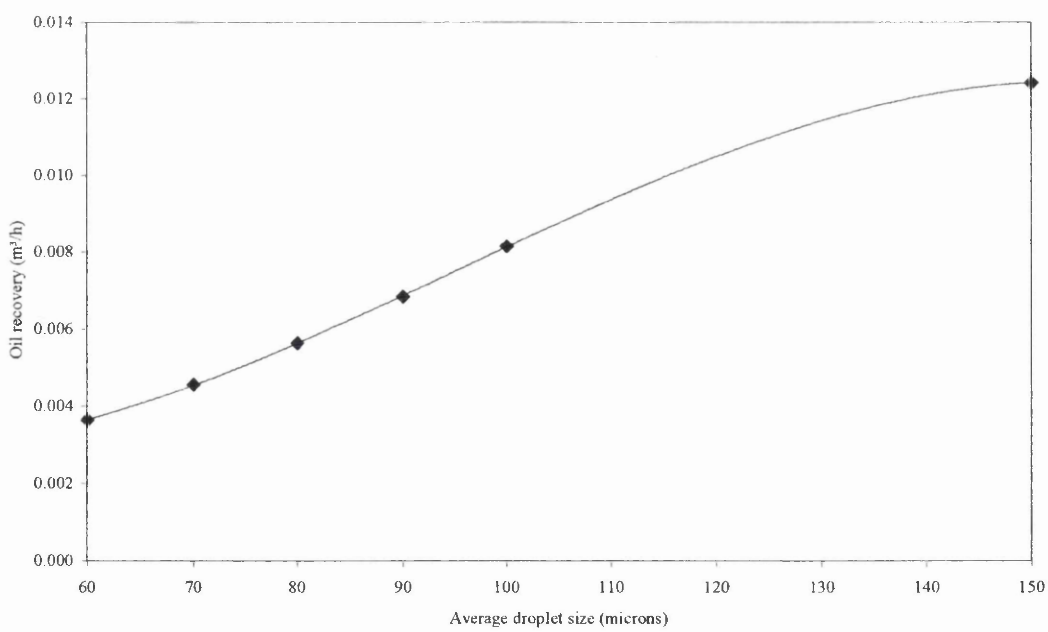


**Figure 5.26: Effect of average droplet size on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

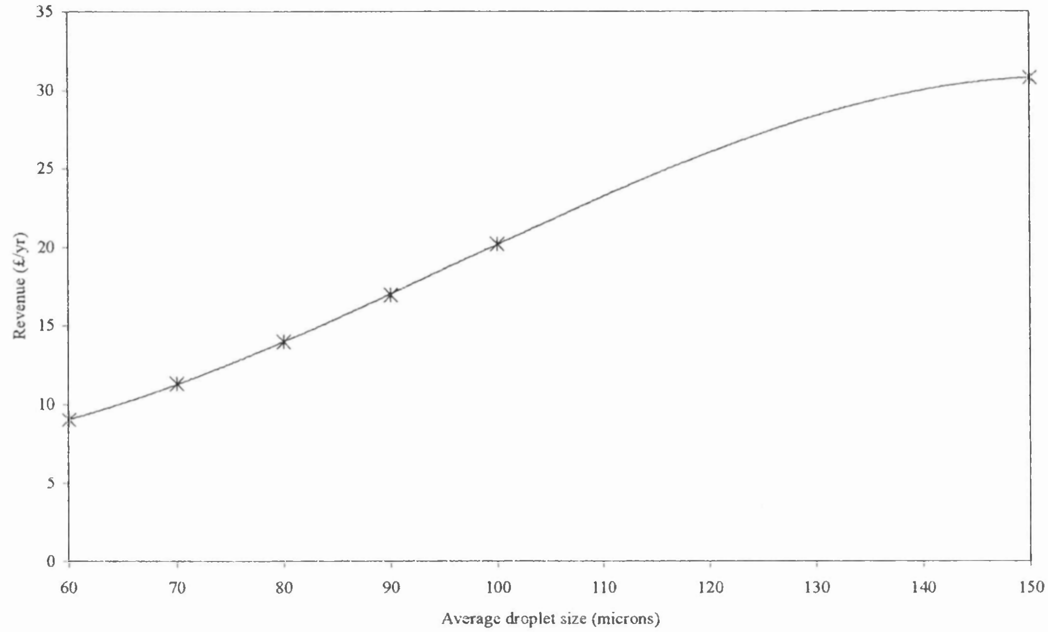


**Figure 5.27: Effect of average droplet size on revenue generated from recovered oil**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

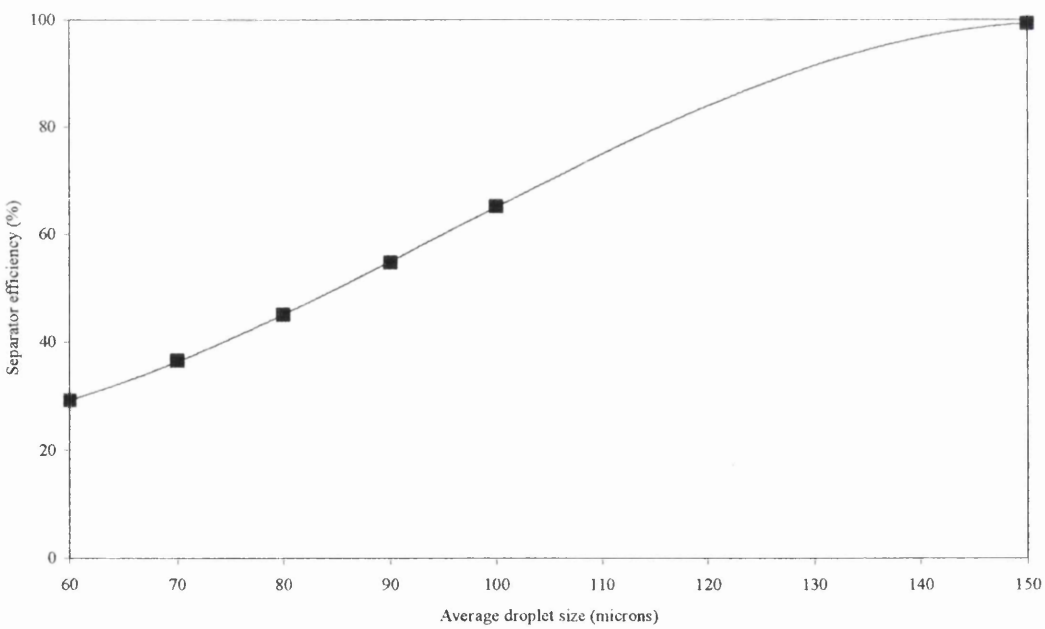


**Figure 5.28: Effect of average droplet size on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

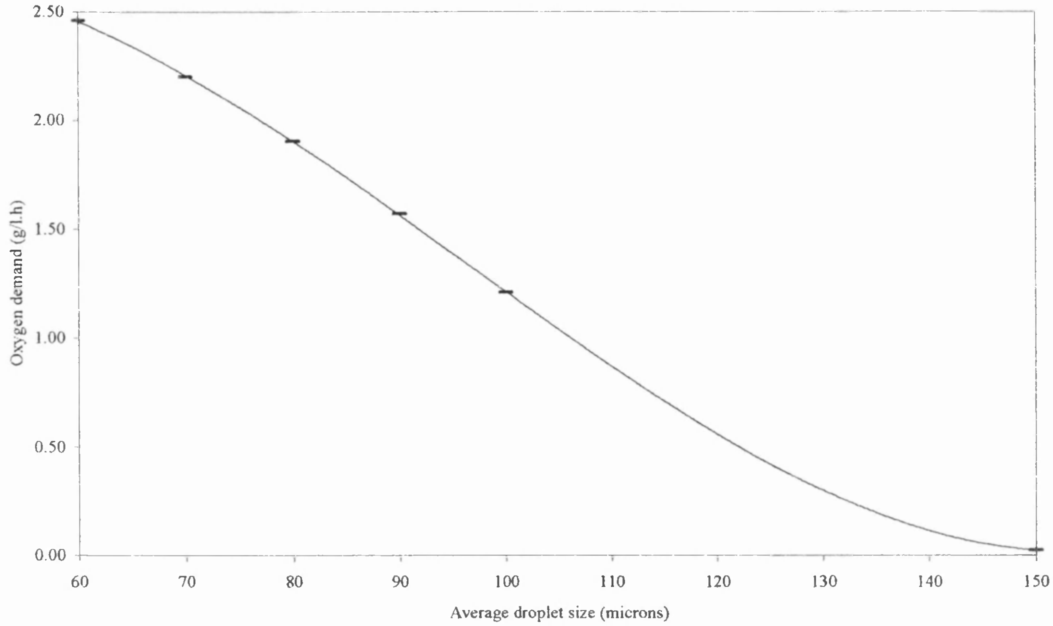


**Figure 5.29: Effect of average droplet size on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

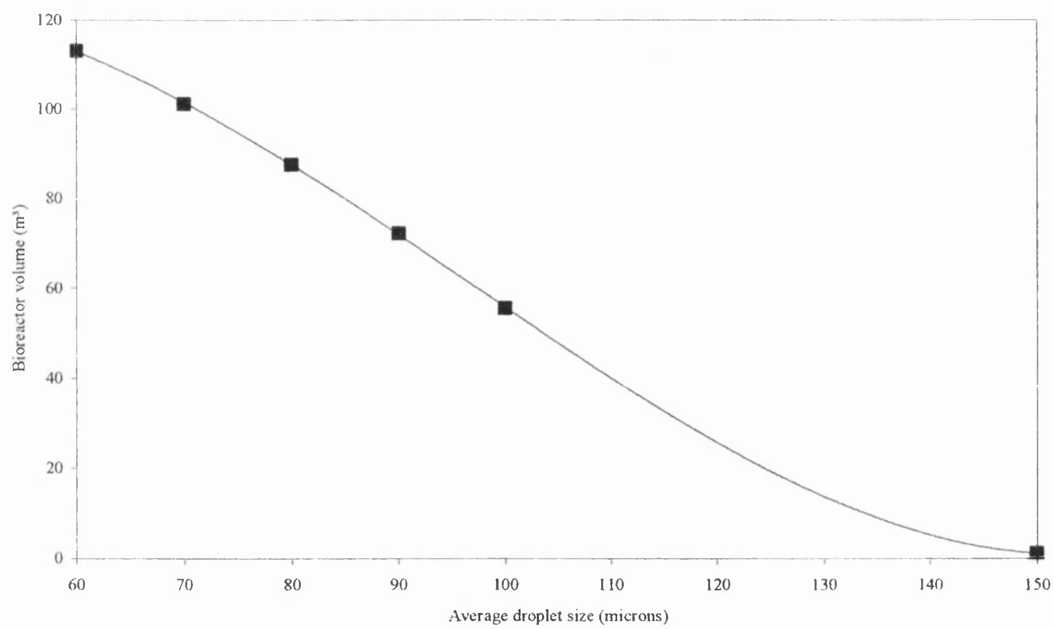


**Figure 5.30: Effect of average droplet size on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

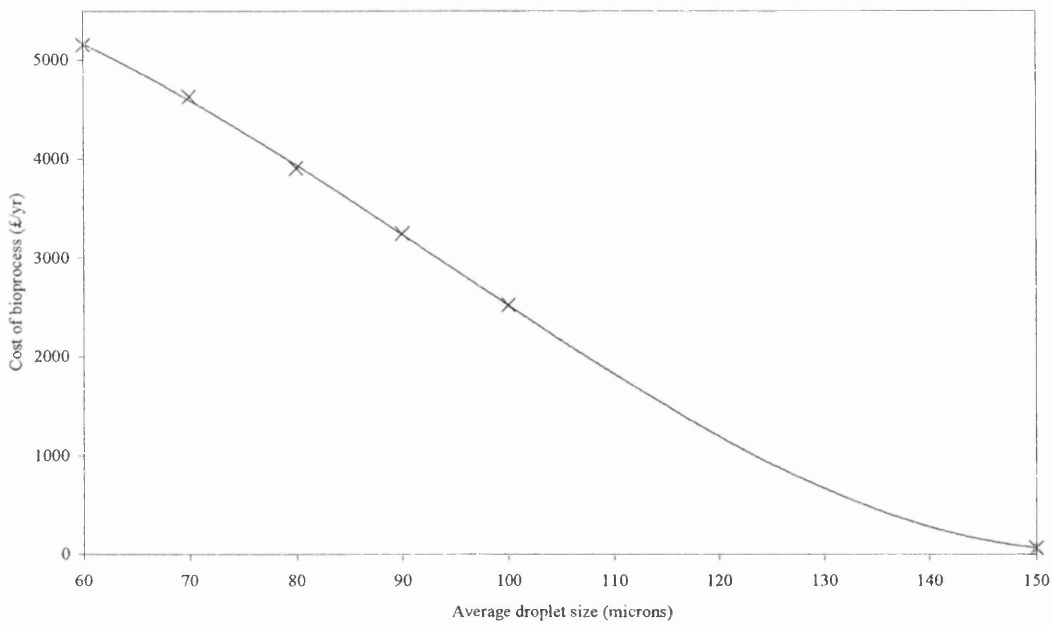


**Figure 5.31: Effect of average droplet size on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

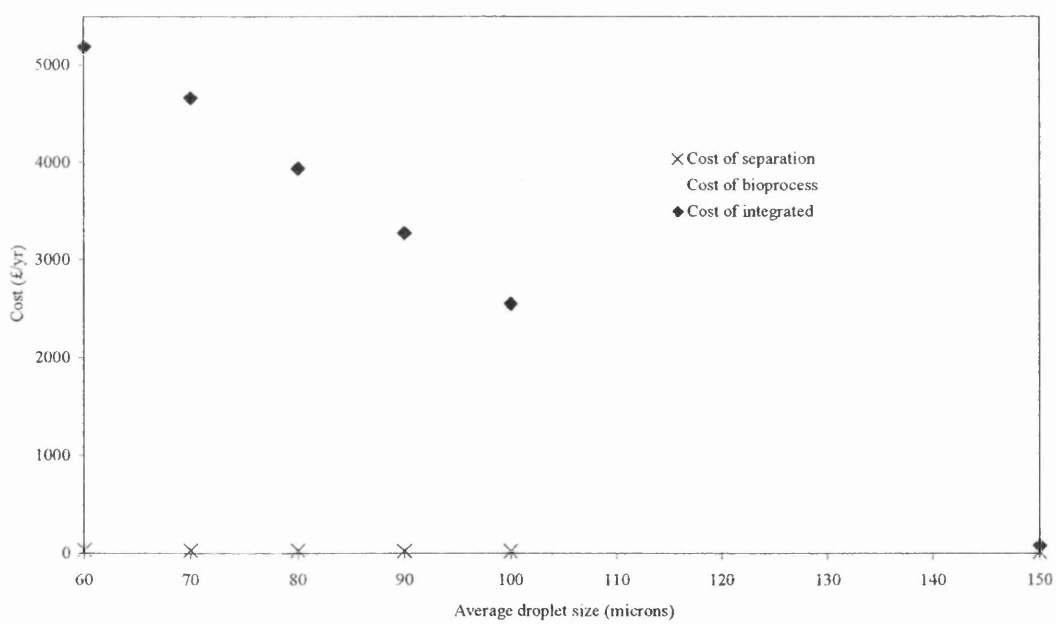


**Figure 5.32: Effect of average droplet size on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.5 Effect of varying wastewater viscosity

The effectiveness of using a horizontal flow separator to recover oil from oily wastewater streams with varying water viscosities was investigated using the models previously described in Chapters 2 and 3. The average droplet size was varied whilst holding other influent characteristics and separator dimensions constant. Separator dimensions, influent characteristics, and results are presented in Appendix III. The cost of treatment is presented below.

Figure 5.33 shows the effect of oily wastewater viscosity on the cost of the separation. As viscosity is decreased from 0.0010 to 0.0003 Pa.s, the cost of separation decreases from £35.53 to £18.11 per year; oil recovery increases from 0.0036 to 0.0106 $m^3/h$ (Figure 5.34). Figure 5.35 shows that the value of recovered oil increases from £9.03 to £26.45 per year as viscosity is reduced. The effect of oily wastewater viscosity on separator efficiency is shown in Figure 5.36; separator efficiency increases from 29% to 85% as oily wastewater viscosity is decreased.

The effect of varying oily wastewater viscosity on the oil oxidation process is shown in Figures 5.37, 5.38 and 5.39. In Figure 5.37, as viscosity is decreased, oxygen demand decreases from 2.46 to 0.51 g/l.h and hence bioreactor volume and cost of the bioprocess decrease from 113 to 23 $m^3$ and from £5153.72 to £1079.17 per year in Figures 5.38 and 5.39 respectively.

Figure 5.40 shows the effect of varying oily wastewater viscosity on the cost of the integrated process. As viscosity is decreased, the cost of the integrated process decreases from £5189.25 to £1097.17 per year. It therefore follows that reducing water viscosity minimizes the cost of the integrated process.

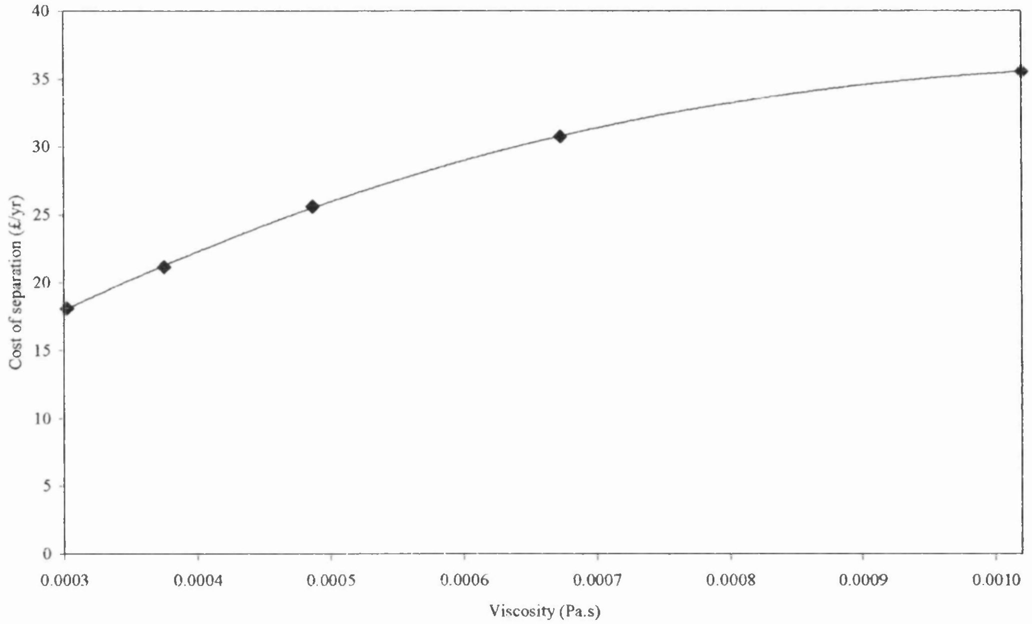


**Figure 5.33: Effect of wastewater viscosity on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

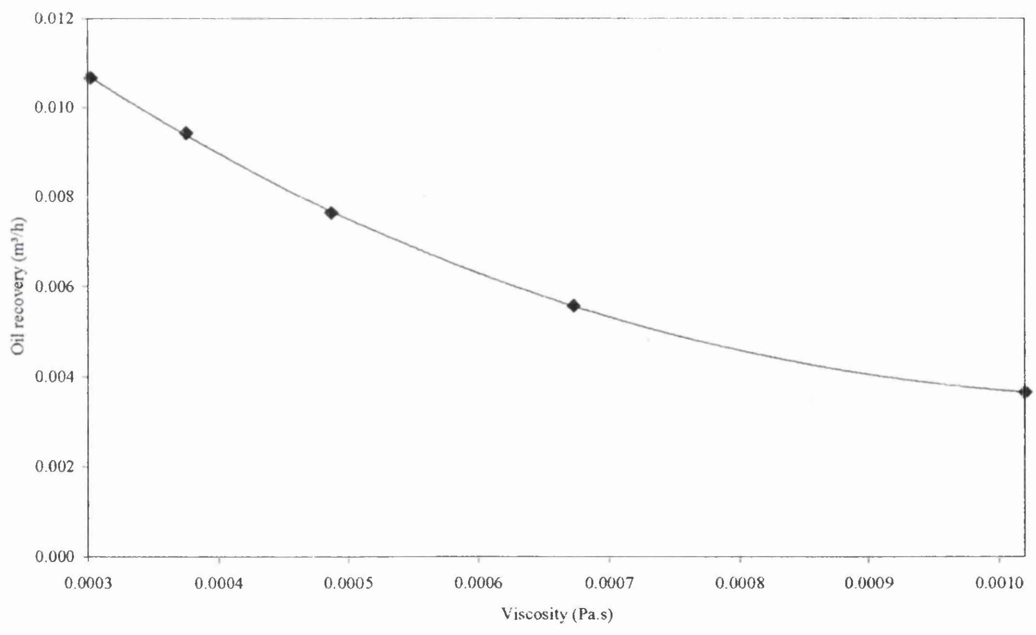


**Figure 5.34: Effect of wastewater viscosity on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

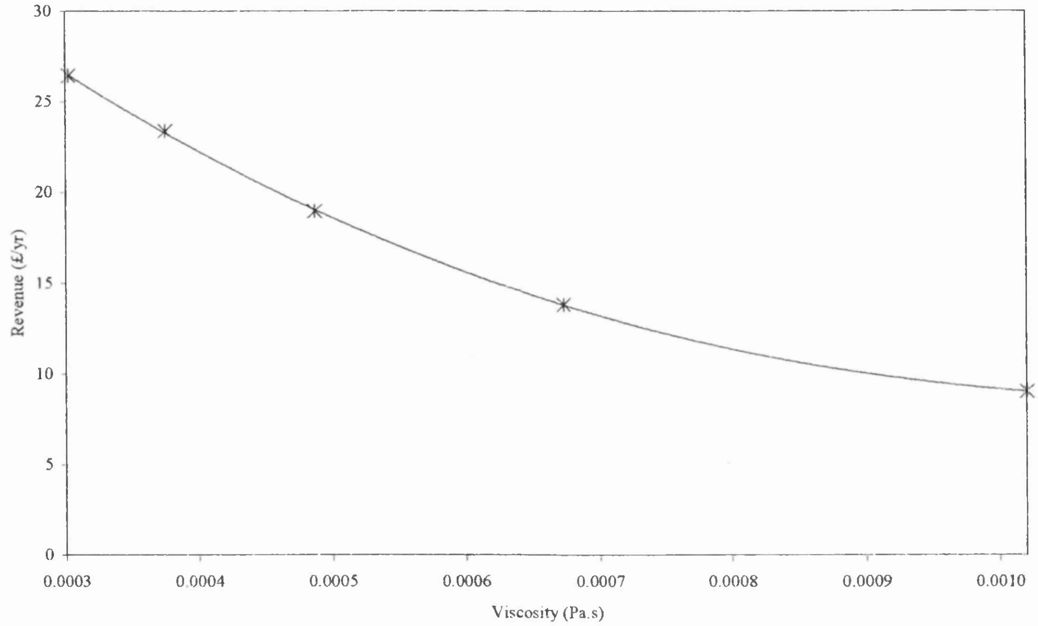


**Figure 5.35: Effect of wastewater viscosity on revenue generated from recovered oil**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

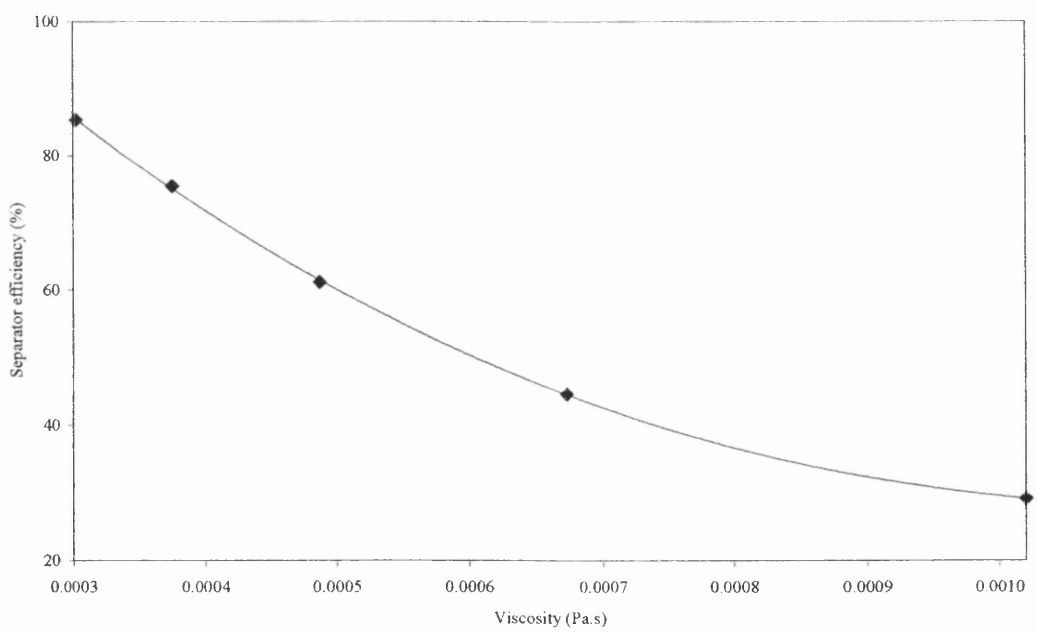


**Figure 5.36: Effect of wastewater viscosity on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

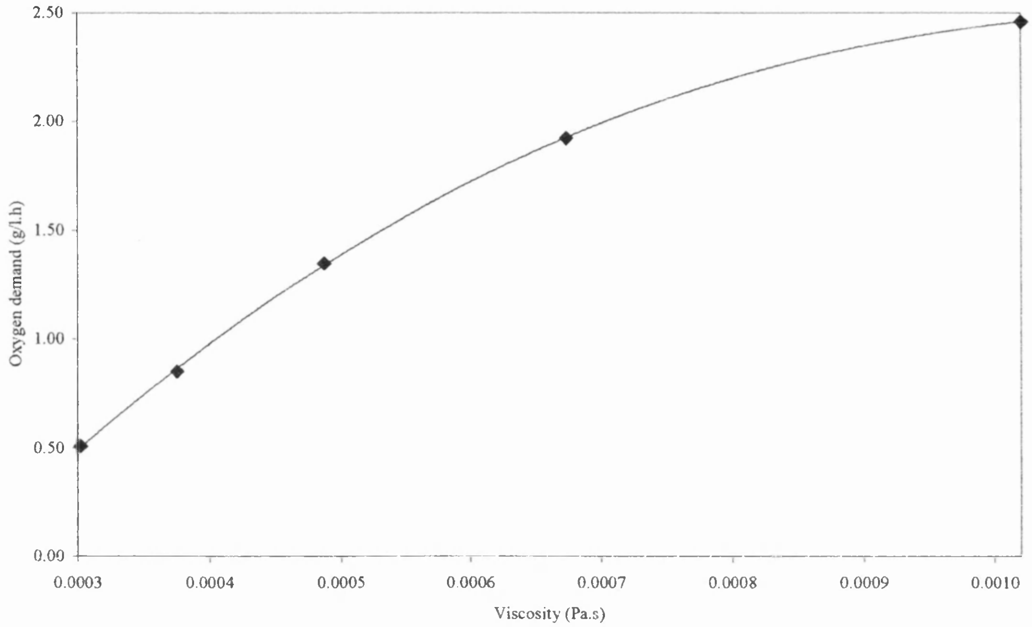


**Figure 5.37: Effect of wastewater viscosity on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

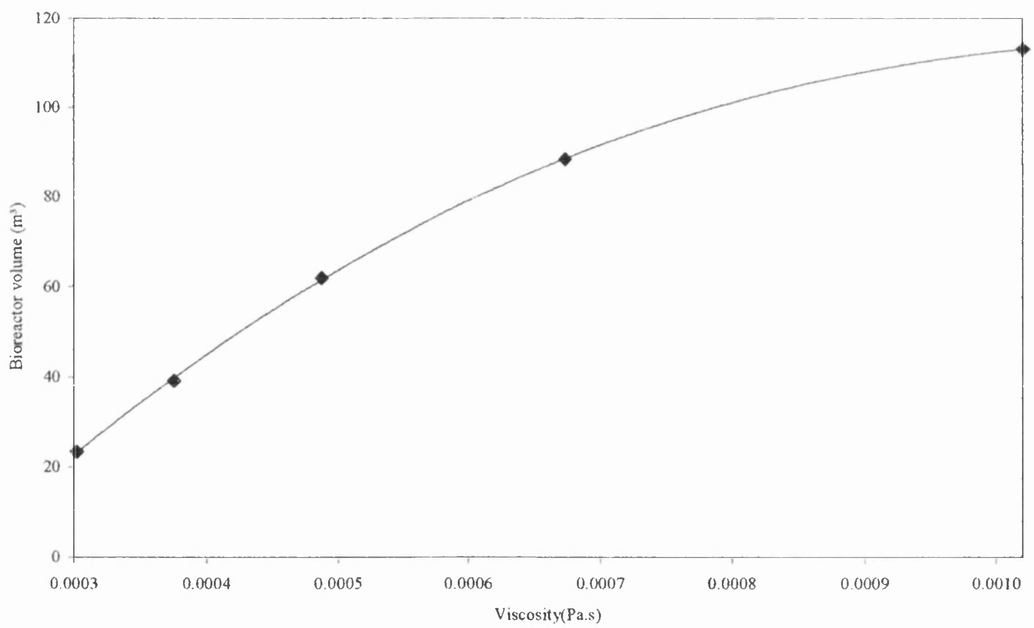


**Figure 5.38: Effect of wastewater viscosity on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

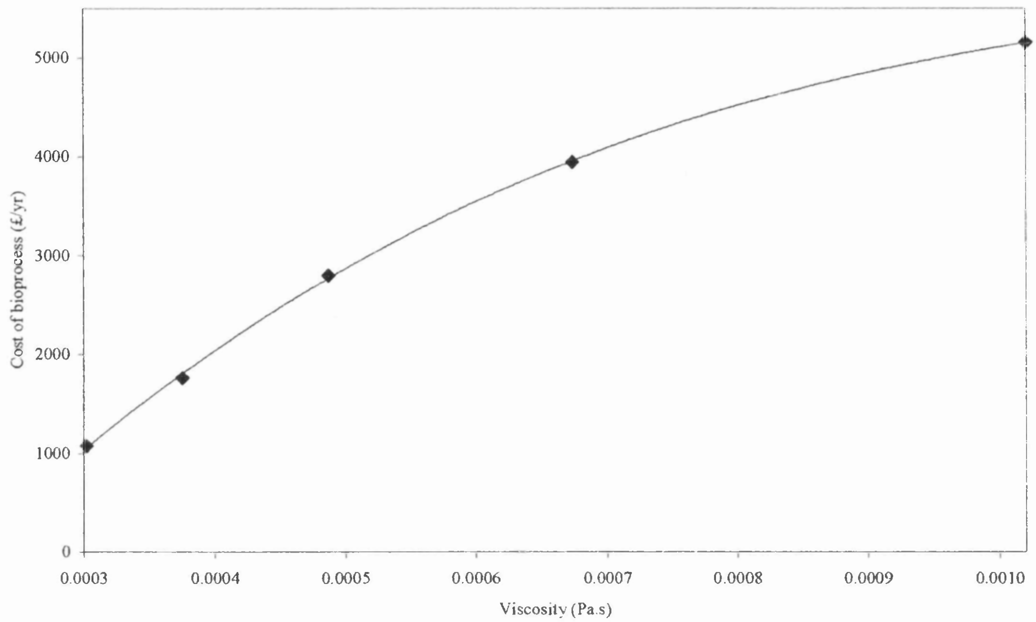


**Figure 5.39: Effect of wastewater viscosity on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

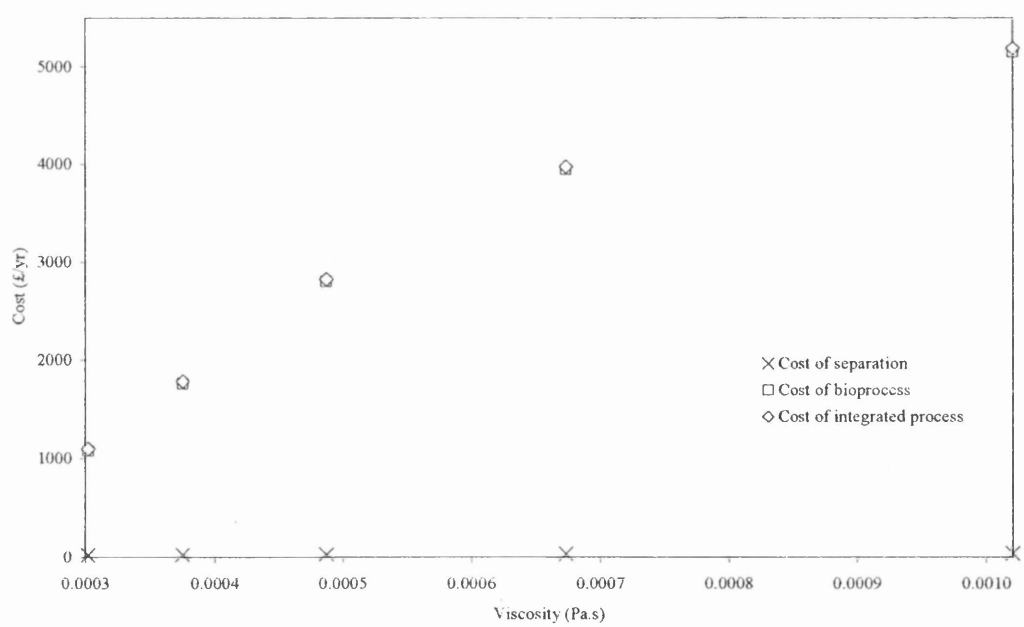


**Figure 5.40: Effect of wastewater viscosity on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.6 Effect of varying oil-water density difference

The effectiveness of using a horizontal flow separator to recover oil from oily wastewater streams with various density differences between the oil and water phases was investigated using the models previously described in Chapters 2 and 3. The density difference was varied whilst holding other influent characteristics and separator dimensions constant. Separator dimensions, influent characteristics, and results are presented in Appendix III. The cost of treatment is presented below.

Figure 5.41 shows the effect of density difference on the cost per year of separation. As density difference is increased from 200 to 235 kg/m$^3$, the cost of separation decreases from £35.53 to £33.95 per year. The decrease in cost of separation is due to the increased oil recovery. In Figure 5.42, oil recovery increases from 0.0036 to 0.0043 m$^3$/h and hence the cost of separation is reduced by revenue generated from the recovered oil; the revenue generated increases from £9.03 to £10.61 per year (Figure 5.43). Figure 5.44 shows the effect of density difference on the efficiency of the separator. Efficiency increases from 29% to about 34% as density difference is increased.

The effect of density difference on the cost of the oil oxidation process is presented in Figures 5.45, 5.46 and 5.47. As density difference is increased, oxygen demand decreases from 2.46 to 2.28 g/l.h; bioreactor volume decreases from 113 to 105 m$^3$ and the cost of the bioprocess decreases from £5153.72 to £4796.47 per year. The effect on the cost of the integrated process is shown in Figure 5.48. The cost decreases from £5189.25 to £4830.42 per year. Increasing the density difference therefore improves the cost of the separator and minimizes the cost of the integrated process.

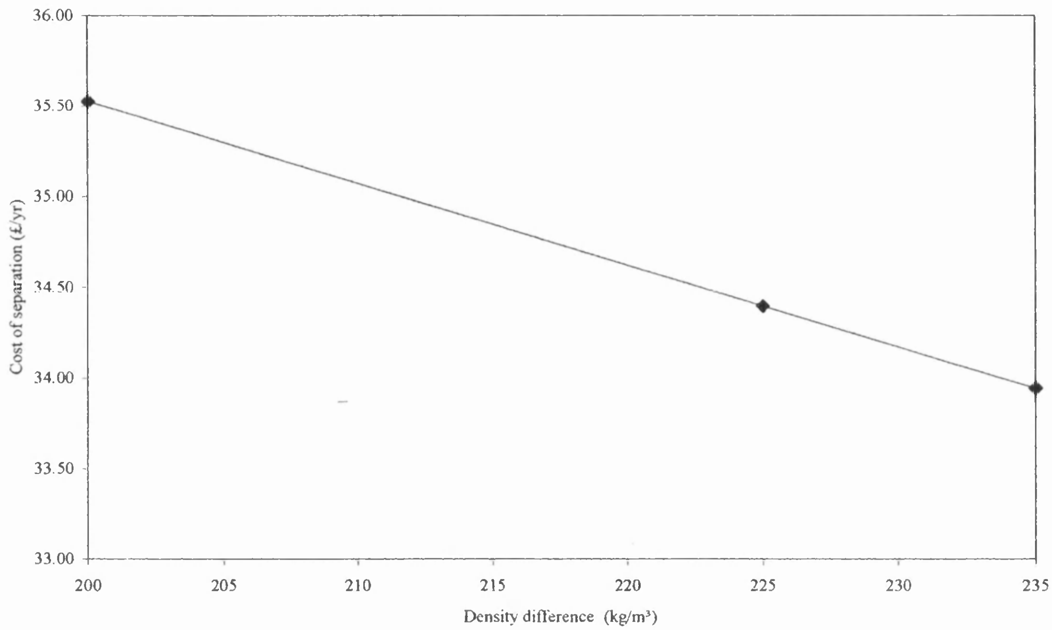


**Figure 5.41: Effect of density difference on cost of separation**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

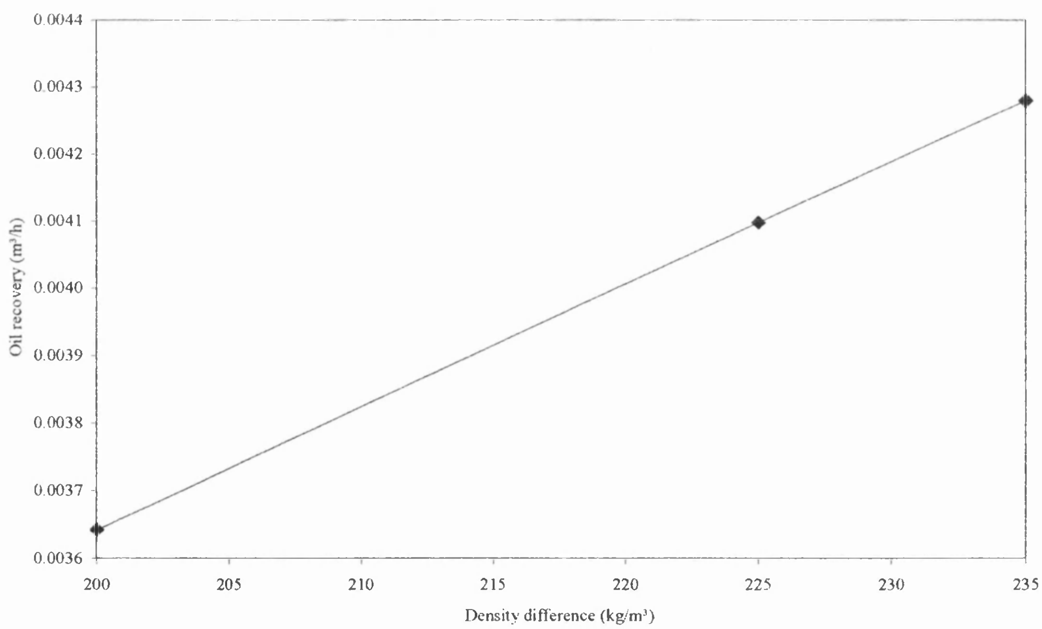


**Figure 5.42: Effect of density difference on oil recovery**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

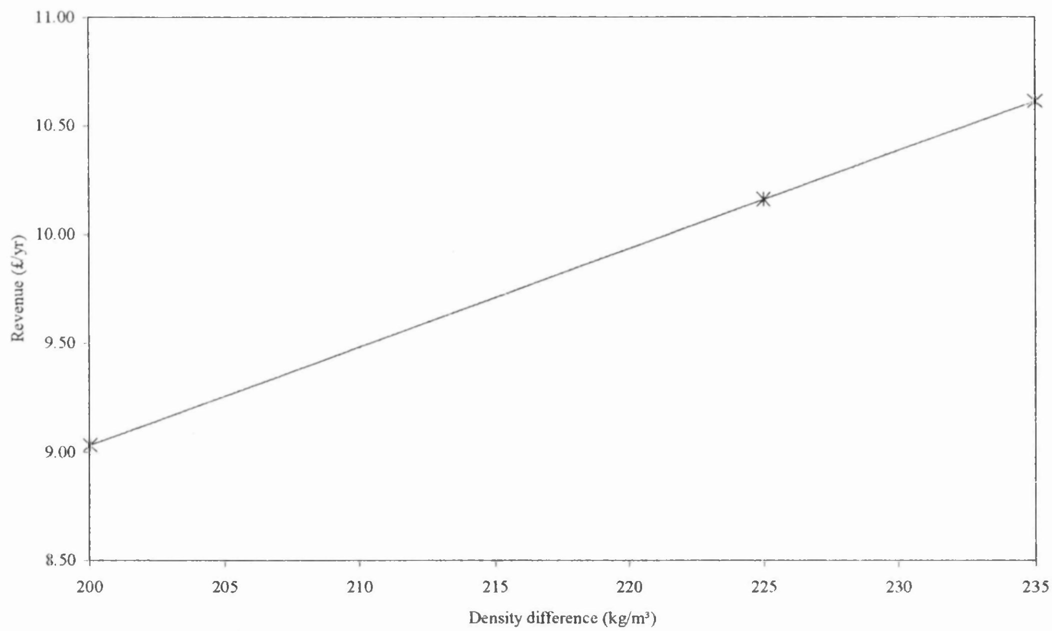


**Figure 5.43: Effect of density difference on revenue generated from recovered oil**

(Influent characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

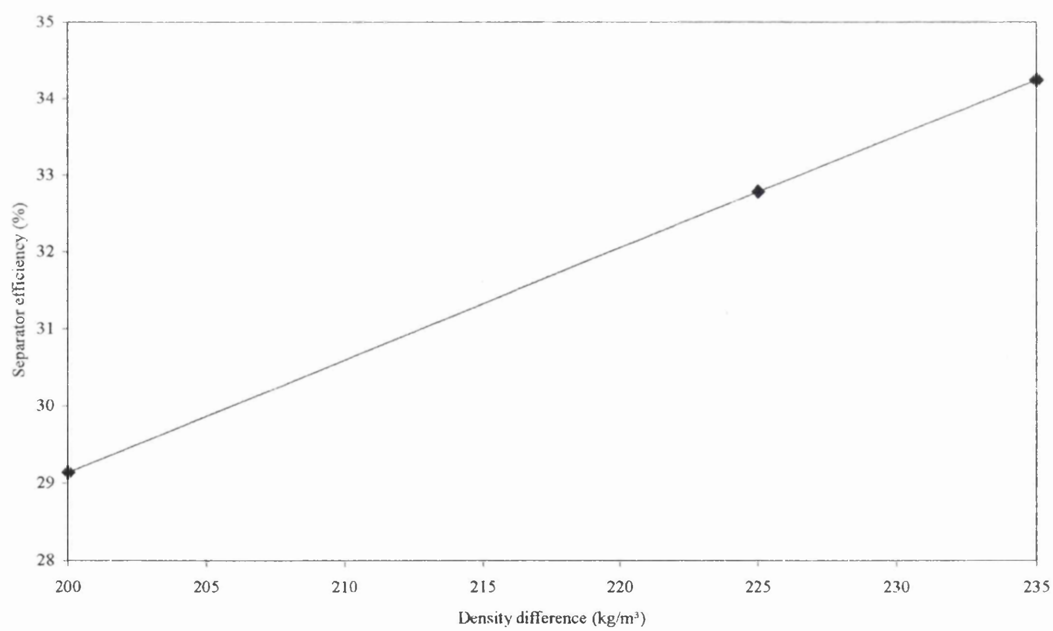


**Figure 5.44: Effect of density difference on separator efficiency**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

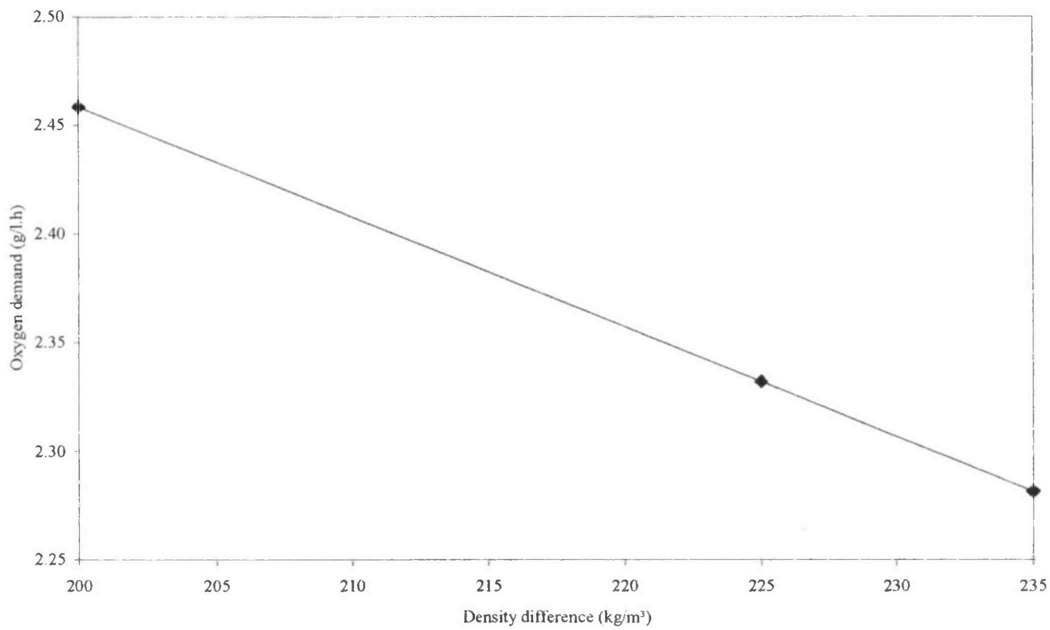


**Figure 5.45: Effect of density difference on oxygen demand**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

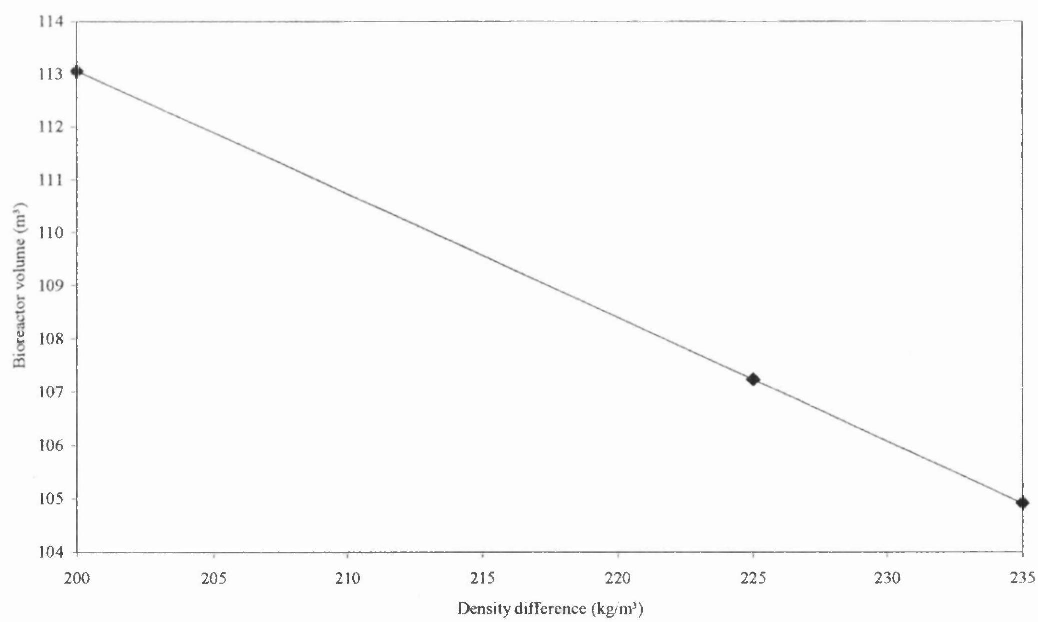


**Figure 5.46: Effect of density difference on bioreactor volume**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

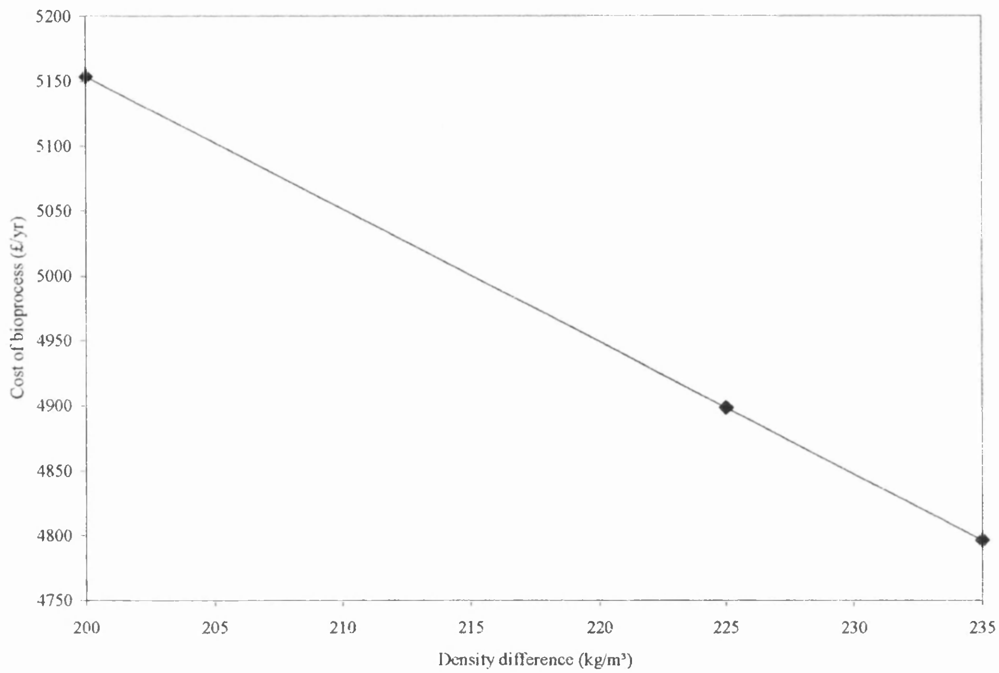


**Figure 5.47: Effect of density difference on cost of bioprocess**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

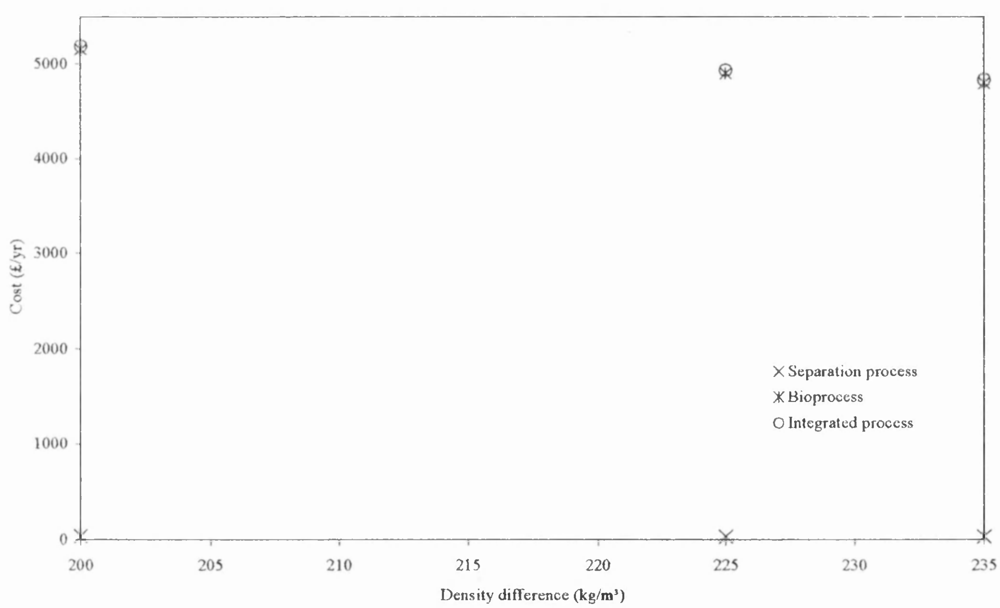


**Figure 5.48: Effect of density difference on cost of integrated process**

(Influent oil characteristics, separator dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix III).

### 5.2.7 Discussion of results and analysis

The various scenarios in which the horizontal flow separator can be operated have been investigated and their effects on the integrated process presented. Generally, the scenarios can be divided into two subsets; one subset considers operating the separator with varying influent conditions and the other considers varying the design of the separators.

The results revealed the following: operating the horizontal flow separator with oily wastewater streams with a high density difference between the oil droplets and the water stream improves the efficiency of the separator and also the cost of separation. Operating the separator with increased oil concentration does not improve separator efficiency but reduces the cost of separation (due to increased revenue from recovered oil). Operating the separator to recover oil from influent oily wastewater streams with a large average oil droplet size improves the efficiency of the separator and the cost of separation. Using the horizontal flow separator to treat oily wastewater streams of low viscosities increases the efficiency of the separator and improves the cost of separation.

Increasing the size of the separator to treat a particular oily wastewater stream improves the efficiency up to a point, beyond which further increase in separator volume does not result in significant increase in efficiency. However, increasing separator volume increases the cost of treatment. Reducing separator horizontal velocity increases efficiency of the separator and reduces the cost of separation.

It was observed that all the factors that increased separator efficiency without any limiting effects on efficiency improved the cost of integrated process.

#### 5.2.7.1 Effect of influent oil concentration on the integrated process

Operating the separator with increasing influent oil concentration reduces the cost of the separator (Figure 5.1). This decrease in cost of separation was as a result of increasing revenue generated from recovered oil. As influent oil concentration increased, influent oil volume increased. From Figure 5.4, it can be seen that increasing oil concentration does not affect the efficiency of the separator: efficiency remains the same at 29%. Therefore, whatever the volume of oil that is fed into the separator, 29% of that volume will be recovered, so the greater the volume of oil in the influent, the greater the volume of oil recovered, the greater the revenue generated (Figure 5.3), and the smaller the overall cost of the separator (Figure 5.1).

It also followed that as influent oil concentration increased the effluent oil concentration increased and consequently the cost of the bioprocess increased (Figure 5.7). Oxygen demand is a function of oil concentration; Figure 5.5 shows the relationship between oxygen demand and oil concentration. Increasing oil concentration results in increasing bioreactor volume (Figure 5.6), and hence increased cost of the bioprocess.

Comparing Figures 5.1 and 5.8 it can be seen that while the cost of separation decreases with increasing oil concentration, the costs of the bioprocess and the integrated process increase. It has been seen that increasing oil concentration does not improve the cost of the integrated process, but these oil concentration-cost relationships can be used to define economic areas of operation. From Figure 5.1 it can be seen it is more profitable to operate the horizontal flow separator when at high oil concentrations (about 1500 mg/l) than when the oil concentration is low. Definition of low and high oil concentrations and may differ; in this exercise the low oil concentration was taken to be 1000 mg/l. Using the horizontal flow separator to recover oil when oil concentrations are low does not improve the cost of treatment: this is because much of the oil cannot be recovered and the bioprocess still needs to be employed to meet environmental regulations, which increases the cost of treatment. Using only the bioprocess decreases the cost of treatment at low oil concentrations. At very high oil concentrations, such as 40,000 mg/l, a large amount of revenue from recovered oil is expected; this would improve the cost of the separation since the relationship between oil concentration and revenue is directly proportional (Figure 5.3). This analysis assumes the droplet distributions for all influent oil concentrations are the same.

Increasing oil concentration does not affect the size of the separator and neither does it affect the efficiency of the separator. The efficiency of the separator remained constant at 29% (Figure 5.4) for various influent oil concentrations. Therefore oil concentration cannot be optimized to improve oil recovery in separator units.

#### 5.2.7.2 Effect of horizontal flow velocity on the integrated process and economic region of operation

Decreasing horizontal flow velocity reduces the cost of the separation unit (Figure 5.9) and also reduces the cost of the bioprocess (Figure 5.15). This improvement in the economics of both units is as a result of the increased residence time of the oily wastewater in the separator unit. Decreasing the velocity allows more time for smaller droplets to flow out of the effluent flow path, therefore increasing oil recovery; the effect is an increase in revenue (Figure 5.11) and the cost of separation decreases. As a consequence of the increased oil recovery in the separator, the effluent oil concentration decreases; the oxygen demand decreases (Figure 5.13), hence the size and cost of the bioprocess also decrease (Figures 5.14 and 5.15). The overall effect on the integrated process is a reduction in cost as horizontal flow velocity decreases.

The relationship between horizontal velocity and the various costs is inversely proportional. A similar trend is observed in the horizontal flow velocity versus cost relationship for both the separation unit (Figure 5.9) and the bioprocess (Figure 5.15), suggesting that horizontal velocity is a parameter which can be optimized to improve the profitability of the integrated process. Figure 5.16 shows that the most economical region to operate both units is at low horizontal velocities.

The size of the separator is of paramount importance to the industry. In fact, following the need to enhance separation, the issue of separator size was the genesis for the development of other technologies. Ways in which separator efficiency could be increased without increasing separator size were also sought, leading to the use of chemicals such as coagulants and flocculants.

Decreasing horizontal velocity does not affect the size of the separator. A reduction in horizontal velocity can be achieved by the addition of baffles. These baffles improve laminar conditions, therefore improving separator efficiency. Figure 5.12 shows that separator efficiency increased by 234% when the horizontal velocity was decreased by 90%.

As can be seen in Figure 5.12, decreasing the horizontal velocity has no limiting effect on the efficiency of the separator as observed in the case of increasing separator size in Figure 5.20. Therefore, since decreasing horizontal velocity improves the cost of the separator, bioreactor, and the integrated process, it seems to be a potential parameter in optimizing oil recovery in separator units.

#### 5.2.7.3 Effect of separator volume on the integrated process

Increasing the separator volume increases the cost of the separation but decreases the cost of the bioprocess (Figures 5.17 and 5.23). The separator volume was increased by making the separator longer; the increasing cost was due to the increasing quantity of construction materials. The decrease in the cost of the bioprocess is as a result of a reduced oil concentration in the effluent from the separator.

Increasing the separator volume increases the residence time that the oily wastewater remains in the separator and hence allows more droplets to rise out of the effluent path. The result is a decrease in effluent oil concentration and hence a decrease in oxygen demand, bioreactor size and hence cost of bioprocess (Figures 5.21, 5.22 and 5.23).

Comparing the contrasting relationship between separator size and cost of separation and the relationship between separator size and cost of the bioprocess (Figures 5.17 and 5.23), it follows therefore that increasing separator size cannot improve the cost of the integrated process: optimization is necessary.

Figure 5.20 also shows that as separator size increases, efficiency increases significantly but beyond 500% (about 23m$^3$ in volume), increase in separator size does not increase efficiency significantly. However, a different trend is observed in the relationship between separator volume and cost (Figure 5.17). As separator volume increases, the cost of separation increases in an almost linear fashion; there is no limiting effect on cost of separation as compared with the effect of separator volume on separator efficiency (Figure 5.20). The limiting effect is also observed in the relationship between separator volume and revenue generated (Figure 5.19); this is as a result of the limiting effect noticed in the separator volume–efficiency curve.

Figure 5.24 shows that as separator size increased, the cost of the integrated process decreased. However, by projection, continuous increase in separator size would result in an increase in separation cost and hence cost of the integrated process. It therefore follows that a minimum cost is expected in the cost curve for the integrated process. The minimum cost point defines the separator size which minimizes the cost of treatment; at that point the integrated process is operating economically in terms of separator volume.

#### 5.2.7.4 Effect of average droplet size on the integrated process

As average droplet size was increased, the costs of both the separation and the bioprocess decreased (Figures 5.25 and 5.31); this improvement in the cost of both units is as a result of increased oil recovery (Figure 5.26), increasing the revenue generated (Figure 5.27). Increasing oil recovery reduces effluent oil concentration, and therefore reduces oxygen demand (Figure 5.29) and hence the size (Figure 5.30) and the cost of the bioprocess (Figure 5.31).

Increasing droplet size improves oil recovery by increasing the rate of rise of the droplets, which results in more droplets rising out of the effluent flow path.

Since increasing droplet size improves both the cost of separation and cost of bioprocess, it seems to be a parameter for improving oil recovery. As previously mentioned, the industry uses this factor to improve oil recovery. This is achieved by the addition of chemicals that promote agglomeration and coalescence; this increases the buoyancy of the droplets and hence improves oil recovery.

Figure 5.32 shows that increasing droplet size improves the cost of the integrated process; therefore it seems profitable to improve oil recovery by increasing droplet size. However, this exercise did not take into account the cost of chemicals, so in reality this may increase the cost of separation. Other costs incurred due to the use of chemicals that were not considered in this analysis are the cost of treatment of the recovered oil, the cost of treatment of the effluent, and the cost of disposal of the sludge. All of these costs should increase cost of treatment.

As previously stated, the size of equipment is important to the industry. Increasing droplet size is achievable by use of chemicals and does not affect the size of the separator. Therefore it seems a suitable parameter for optimizing oil recovery.

As can be seen in Figure 5.28, increasing droplet size increases efficiency and has no limiting effect on the efficiency of the separator as observed in the case of increasing separator size (Figure 5.20). Therefore, since increasing droplet size improves the cost of the separation, bioprocess, and integrated process, it seems to be a suitable parameter in optimizing oil recovery in horizontal flow separators.

#### 5.2.7.5 Effect of viscosity on the integrated process

Oily wastewater streams with low water viscosity improve oil recovery (Figure 5.34) and revenue generated (Figure 5.35) and hence the costs of separation and the bioprocess are reduced (Figures 5.33 and 5.39).

Low viscosity implies the drag effect on the oil droplet is low; the droplets have a faster rise rate and hence oil recovery is increased. Increasing oil recovery reduces effluent oil concentration and hence oxygen demand (Figure 5.37). Consequently the bioreactor volume and cost of the bioprocess decreased with low viscosity oily wastewater streams (Figures 5.38 and 5.39).

The similar trend observed in the viscosity-cost relationships (Figures 5.33, 5.39 and Figure 5.40) suggests that viscosity is a suitable parameter to manipulate to improve oil recovery.

Currently, there is only one method of reducing viscosity and this is by heating.

Reducing the viscosity of the medium does not affect the size of the unit but it increases separator efficiency in an almost linear fashion (Figure 5.36). The relationship between viscosity and efficiency is inversely proportional. Also there is no limiting effect on separation efficiency as viscosity decreases. Therefore it may be used as a parameter in improving oil recovery.

#### 5.2.7.6 Effect of density difference on the integrated process

Increasing density difference increases the buoyancy of the oil droplets which results in more oil droplets rising rapidly to the surface, therefore improving oil recovery (Figure 5.42). The effect of increasing oil recovery is an increase in revenue generated (Figure 5.43) and hence a reduction in the cost of separation (Figure 5.41). As oil recovery increases, effluent oil concentration, and hence oxygen demand (Figure 5.45), decreases. Consequently the size of the bioreactor decreases and hence the cost of the bioprocess (Figures 5.46 and 5.47) is reduced.

The similar trends in the density difference-cost relationship on both units suggest density difference may be a suitable parameter for optimizing separator units and the integrated process.

In practice, to increase density difference, bubbles are dissolved into oil-water mixtures at high pressures: these bubbles attach to the oil droplets to form composite particles which are much less dense than water and hence rise rapidly to the surface where they are skimmed off. Density difference is not usually used to improve oil recovery in horizontal flow separators. This factor is used in flotation units.

In this exercise, increasing density difference, with a fixed separator volume, resulted in an increase in the efficiency of the separator (Figure 5.44). No limiting effect on the efficiency was observed as density difference was increased; therefore density difference seems to be a suitable parameter to improve oil recovery.

#### 5.2.7.7 Summary of results discussion

In summary, the factors which influence oil recovery and cost of the integrated process can be grouped into influent stream characteristics and separator designs or configurations. All influent stream characteristics except oil concentration can be manipulated to improve separation efficiency. Separator design or configuration can be manipulated to improve oil recovery. Increasing separator size and decreasing horizontal flow velocities are methods by which separator efficiency can be improved by manipulating the design or configuration. While the design variables increase the residence time of the wastewater in the separator, varying influent stream characteristics improves the rise rate of the oil droplets. While increase in separator size requires the construction of a larger tank to allow smaller droplets to rise out of the effluent flow path, decreasing horizontal velocity and influent stream characteristics may not necessitate the construction of a larger tank. The effects of manipulating the influent stream characteristics, except for oil concentration, are all predicted by Stokes' Law.

The factors which improve the cost of the integrated process are: viscosity, density difference, droplet size, and horizontal velocity. Separator size and oil concentration do not seem to improve the cost of the integrated process. It was observed that most factors which improved separator efficiency reduced the cost of the integrated process. So it can be said that the better the efficiency of the separator, the lower the cost of integrated process.

#### 5.2.7.8 Analysis of Stokes' Law variables

Based on an analysis of Stokes' Law, the most influential factors which improve the rise rate of an oil droplet are the viscosity and droplet size. Figure 5.49 compares the variables in Stokes' Law. This analysis is based on data given in Appendix III.

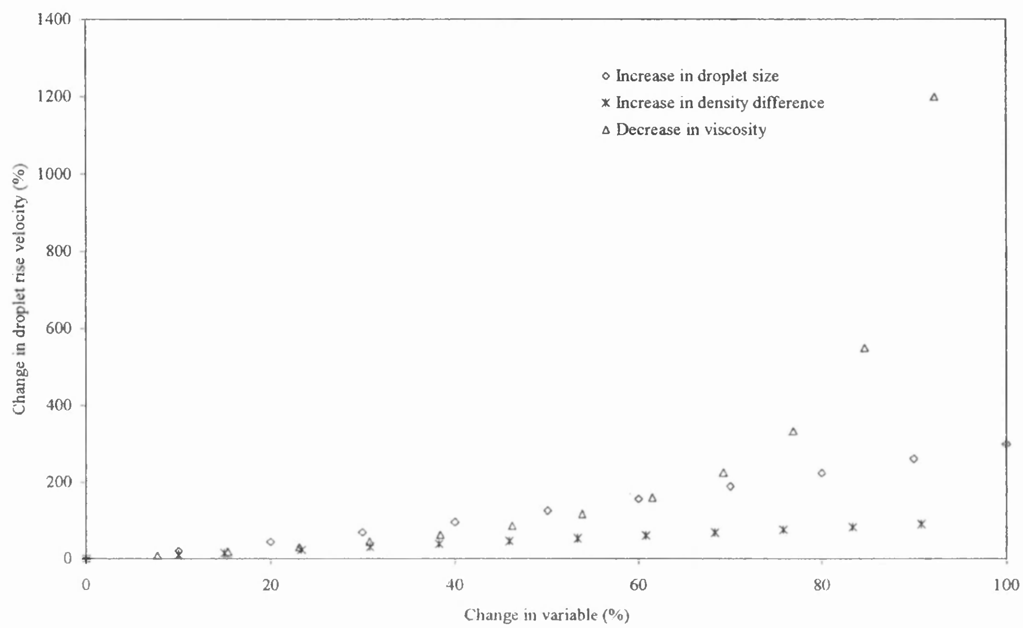


**Figure 5.49: Comparing variables in Stokes' Law**

From Figure 5.49, it can be seen that increasing droplet size has the greatest effect on rise rate when the percentage change in each variable is less than or equal to 60%; beyond a 60% change in variable, reducing the medium viscosity has the most effect on rise rate. These results seem to be the basis for the use of chemical additives to increase droplet size in the industry. However, from the analysis of Stokes' Law, it is seen that reducing the viscosity can have the greatest effect on rise rate; however, this requires the input of a lot of energy to heat the wastewater stream. Thermally-enhanced separation processes are not common in the industry, probably due to the high cost of energy.

The results presented in Section 5.1 were analyzed and presented in Tables 5.1, 5.2 and 5.3 to compare the effect of the various factors on separator efficiency.

| Increase in separator size (%) | Increase in separator efficiency as a result of increase in size (%) | Increase in average droplet size (%) | Increase in separator efficiency as a result of increase in average droplet size (%) |
|---|---|---|---|
| 0 | 0 | 0 | 0 |
| 17 | 21 | 17 | 26 |
| 33 | 39 | 33 | 55 |
| 50 | 57 | 50 | 88 |
| 67 | 73 | 67 | 124 |
| 150 | 139 | 150 | 241 |

**Table 5.1: Effect of separator size and droplet size on separator efficiency**

| Decrease in viscosity (%) | Increase in separator efficiency as a result of decreasing viscosity (%) | Increase in average droplet size (%) | Increase in separator efficiency as a result of increase in average droplet size (%) |
|---|---|---|---|
| 0 | 0 | 0 | 0 |
| 17 | 18 | 17 | 26 |
| 33 | 48 | 33 | 55 |
| 50 | 98 | 50 | 88 |
| 67 | 168 | 67 | 124 |
| 150 | 877 | 150 | 241 |

**Table 5.2: Effect of viscosity and droplet size on separator efficiency**

| %age decrease in horizontal velocity | %age increase in separator efficiency as a result of decreasing horizontal velocity | %age increase in average droplet size | %age increase in separator efficiency as a result of increase in average droplet size |
|---|---|---|---|
| 0 | 0 | 0 | 0 |
| 17 | 28 | 17 | 26 |
| 33 | 63 | 33 | 55 |
| 50 | 109 | 50 | 88 |
| 67 | 164 | 67 | 124 |
| 150 | 566 | 150 | 241 |

**Table 5.3: Effect of horizontal velocity and droplet size on separator efficiency**

### 5.2.7.9 Comparing increasing droplet size and increasing separator volume

Oil droplet size can be increased, to improved buoyancy, by the addition of chemicals. The effect of using chemicals to increase separator efficiency was simulated by using the various average droplet sizes in the streams investigated.

From Table 5.1, a 17% increase in droplet size increases the separator efficiency by 26% and a 17% increase in separator size increases the separator efficiency by 21%. It may be concluded that increasing average droplet size improves separator efficiency more than increasing separator size, hence the usual practice of using chemical additives.

**5.2.7.10 Comparing increasing droplet size and decreasing viscosity**

The effect of decreasing the viscosity of the wastewater stream, to improve separator efficiency, was simulated by assuming the various streams studied were heated to reduce their viscosities.

Table 5.2 compares the effect of decreasing the viscosity and increasing average droplet size on separator efficiency. From Table 5.2, it seems a reasonable amount of energy must be expended to achieve a separator performance better than that achieved by increasing droplet size. These results also tally with that obtained in the analysis of Stokes' Law. Another disadvantage of using the thermal process is that the oily wastewater has to be cooled before the effluent is fed to the bioprocess. This involves huge operating expenditures and time and may be the reason the industry currently prefers the use of chemicals.

**5.2.7.11 Comparing increasing droplet size and decreasing horizontal velocity**

The effect of decreasing the horizontal velocity in the separator was simulated by assuming that the horizontal flow velocity of a separator with a fixed volume of 4 $m^3$ was decreased by addition of baffles for each velocity investigated.

Table 5.3 compares the effect of decreasing horizontal velocity and increasing average droplet size on separator efficiency. From Table 5.3 it seems a 33% change in horizontal velocity increases the separator efficiency by 63% while a 33% change in average droplet size only increased separator efficiency by 55%. From Table 5.3, it may be said that decreasing the horizontal velocity has more effect on separator efficiency than increasing droplet size.

**5.2.7.12 Critical design factor**

From this analysis, it seems the factors or scenarios which best improve the effectiveness of the horizontal flow separator to recover oil are associated with the design of the separator rather than manipulation of the influent oily wastewater stream characteristics.

Reducing horizontal velocity seemed to improve separator effectiveness better than the use of chemicals in this analysis.

The issue of longer residence time due to decreasing horizontal velocity popularized the use of chemicals to increase droplet size and thereby enhance oil recovery.

#### 5.2.7.13 Comparing decreasing horizontal velocity with use of chemicals to improve oil recovery in a horizontal flow separator

- Decreasing horizontal velocity does not pose any hazard to the environment; use of chemicals has adverse effects on the environment
- Decreasing horizontal velocity does not involve any expense or continuous operating cost, the cost involved is a capital cost. Use of chemicals involves continuous operating cost which further increases cost of treatment
- Decreasing horizontal velocity involves long residence time. Use of chemicals involves long treatment times since recovered oil must be treated and effluent sent to the bioprocess must be treated
- Use of chemicals involves addition of more treatment units to treat both the recovered oil and the effluent. Decreasing horizontal velocity does not require additional treatment therefore does not require addition of more treatment units to treat effluent and recovered oil.
- Decreasing horizontal velocity in the design of separators does not result in additional formation of sludge which is harmful to the environment and adds further cost. Use of chemicals involves sludge formation.
- Decreasing horizontal velocity does not necessitate an increase in separator size - it is achieved by addition of baffles. Use of chemicals does not necessitate an increase in separator volume but may require large doses (Zeng *et al.* 2007).
- The chemicals may be corrosive, necessitating a more expensive material of construction of the separator. Decreasing horizontal velocity does not have such an effect.

## 5.3 Integrated treatment process using CPI and ICB

The Corrugated Plate Interceptor (CPI) is basically a horizontal flow separator as previously reported. The major difference between the horizontal flow separator and the CPI is the addition of plates. To improve oil recovery, horizontal flow separators are sometimes retrofitted by the addition of plates. This section investigates the effect of varying CPI designs so as to ascertain a cost effective design which improves oil recovery.

The underlying theory for the design and predicting the performance of the CPI has been previously described in Chapters 2 and 3. For each design investigated, a volume of oily water was fed into the interceptor and the performance predicted (see Section 3.7.1). For each design, the size of Immobilized Cell Bioreactor (ICB) required to treat the effluent from the interceptor was calculated based on the oxygen demand of the effluent (see Section 2.3.2.2). The costs of constructing and operating the interceptor and the ICB were then calculated in each case. The results of the investigations are presented below. Interceptor dimensions and influent characteristics are presented in Appendix IV.

### 5.3.1 Effect of varying interceptor plate width

To investigate the effect of plate width, a number of Corrugated Plate Interceptors were designed, using models previously described in Chapters 2 and 3, to recover oil from a fixed oily wastewater stream. The width of the plate pack was varied whilst holding influent characteristics and other design parameters constant. Interceptor dimensions, plate pack details, influent characteristics, and results are presented in Appendix IV. The cost of the treatment is presented below.

Figures 5.50 and 5.51 show that as plate width is increased from 1 to 5 m, oil recovery increases from 0.007 to 0.012 $m^3/h$ and the revenue generated increases from £16.14 to £30.84 per year. The effect of plate width on the cost of separation is shown in Figure 5.52: as plate width is increased, the cost of separation increases from £120.28 to £312.61 per year. Figure 5.53 shows the interceptor volume increases from 2.14 to 7.86 $m^3$ with the same increase in plate width. The effect on interceptor performance is shown in Figure 5.54: as plate width is increased, interceptor efficiency increases from 52 to 99%.

The effect of interceptor plate width on the bioprocess is shown in Figures 5.55, 5.56 and 5.57. As plate width is increased, oxygen demand decreases due to decreasing effluent oil concentration from the interceptor; bioreactor size decreases from 76 $m^3$ to 817 litres and the cost of bioprocess decreases from £3428.27 to £50.93 per year. Increasing plate width also decreases the cost of the integrated process from £3548.56 to £363.55 per year (Figure 5.57).

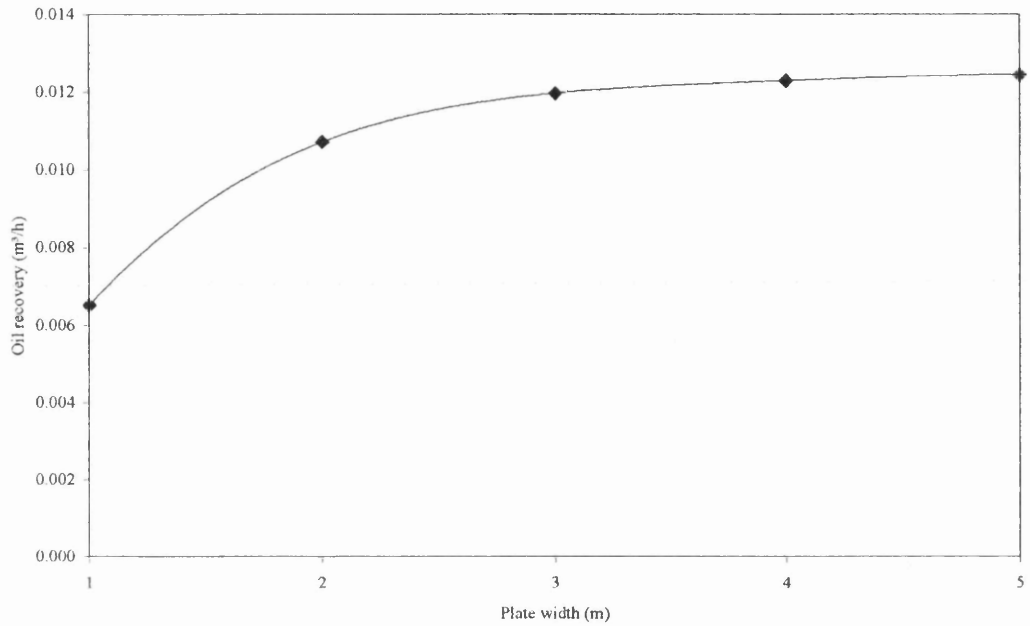


**Figure 5.50: Effect of plate width on oil recovery**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

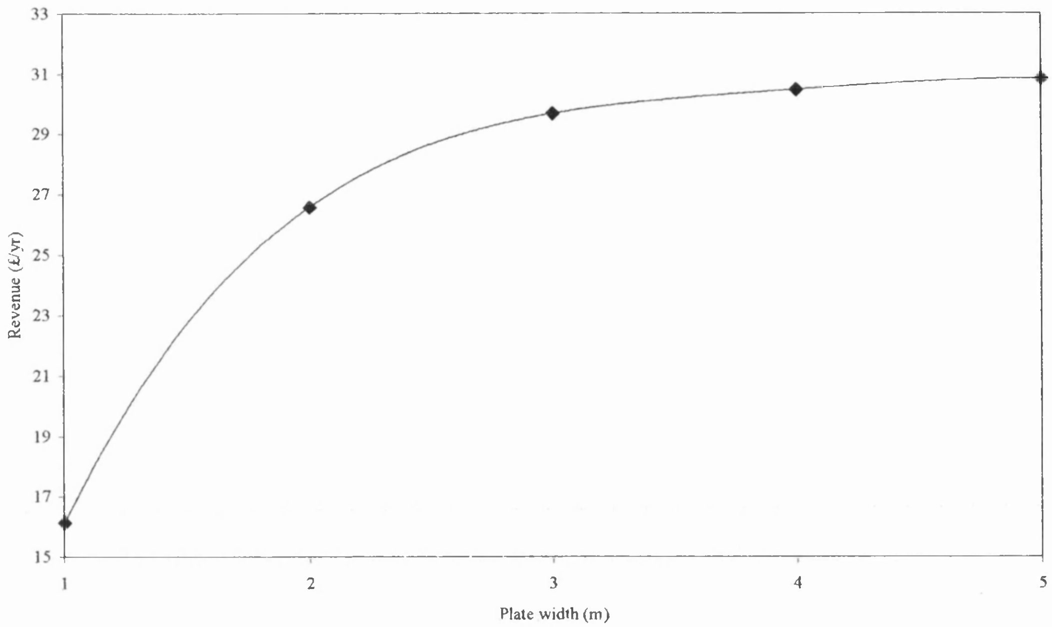


**Figure 5.51: Effect of plate width on revenue generated from recovered oil**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

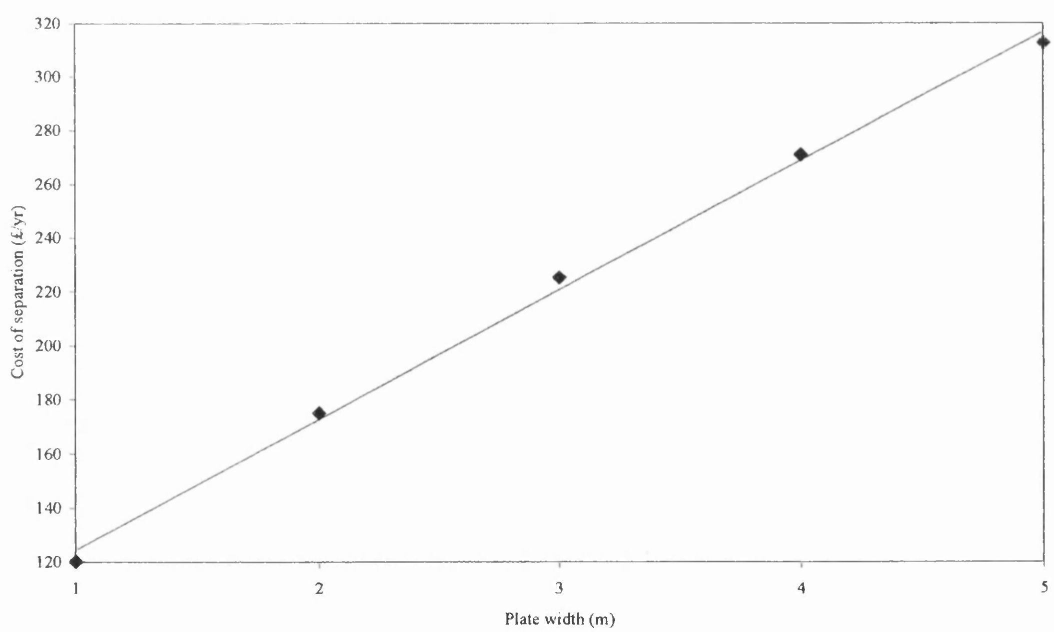


**Figure 5.52: Effect of plate width on cost of separation**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

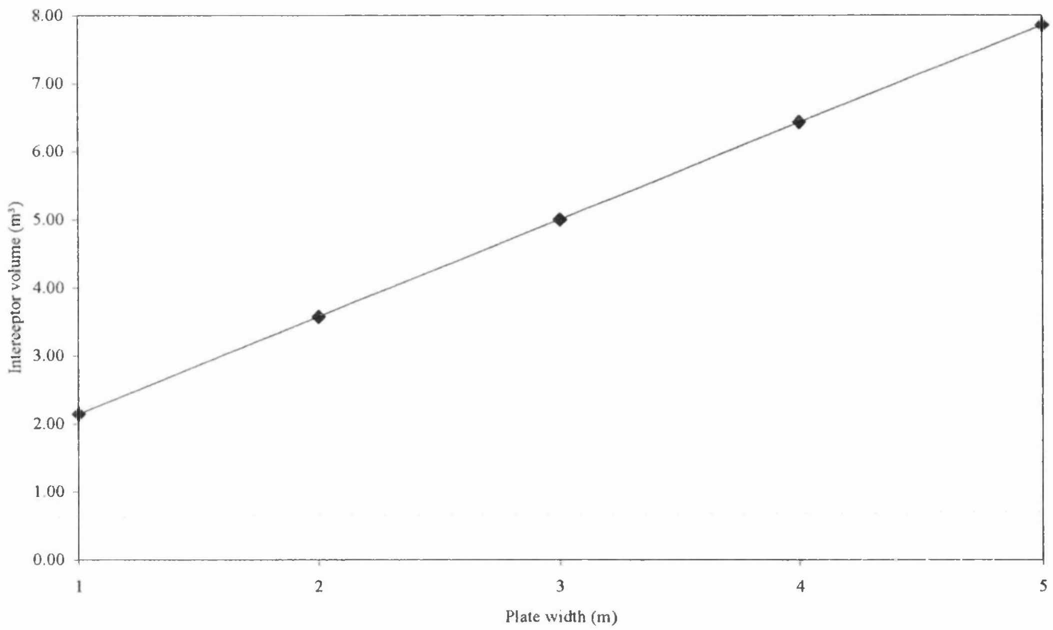


**Figure 5.53: Effect of plate width on interceptor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

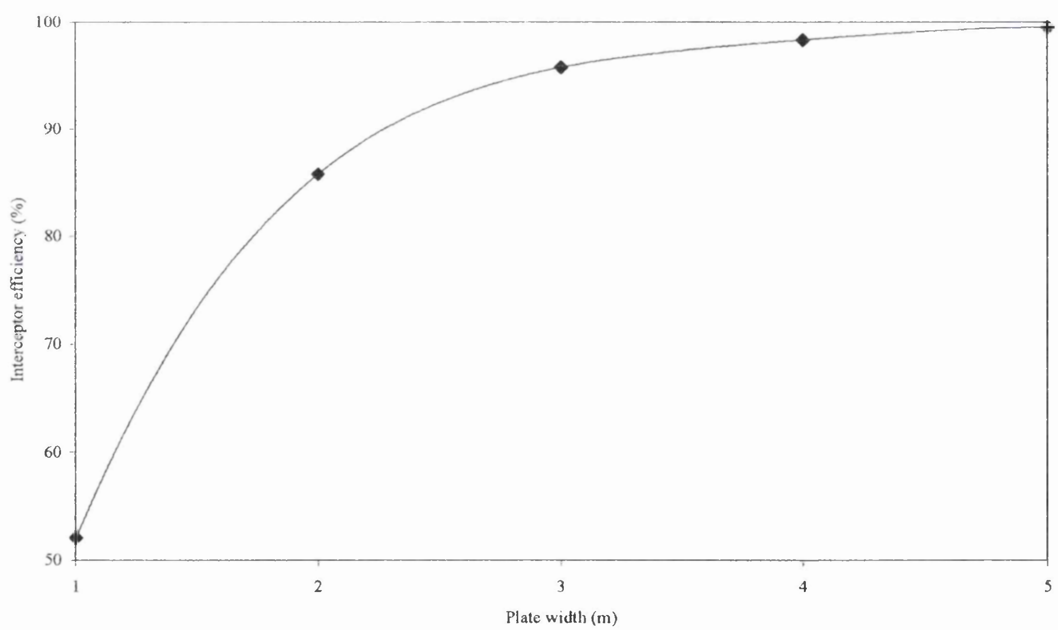


**Figure 5.54: Effect of plate width on interceptor efficiency**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

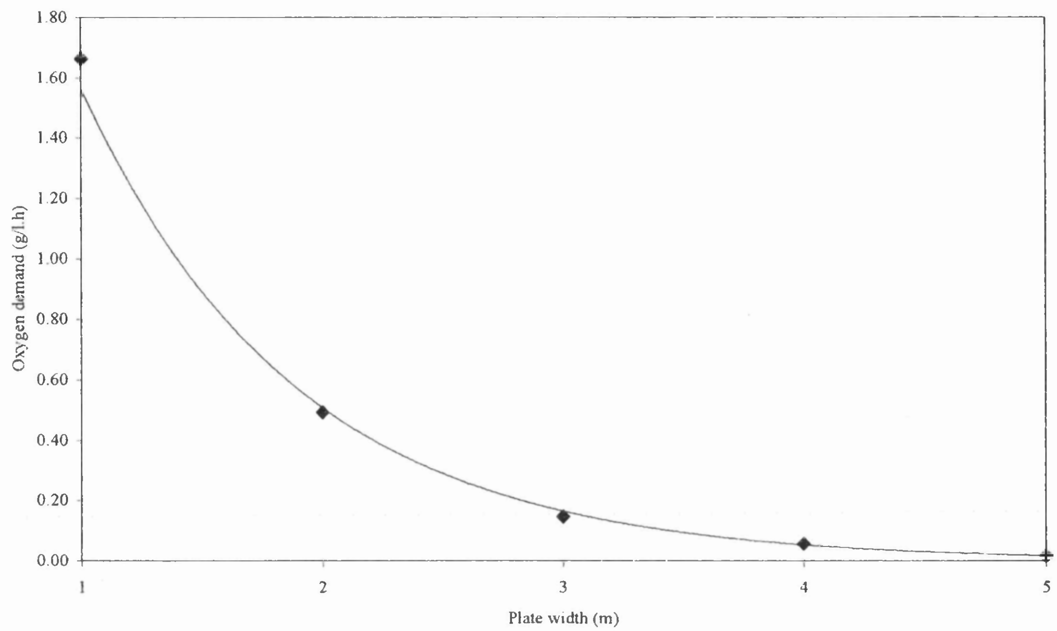


**Figure 5.55: Effect of plate width on oxygen demand**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

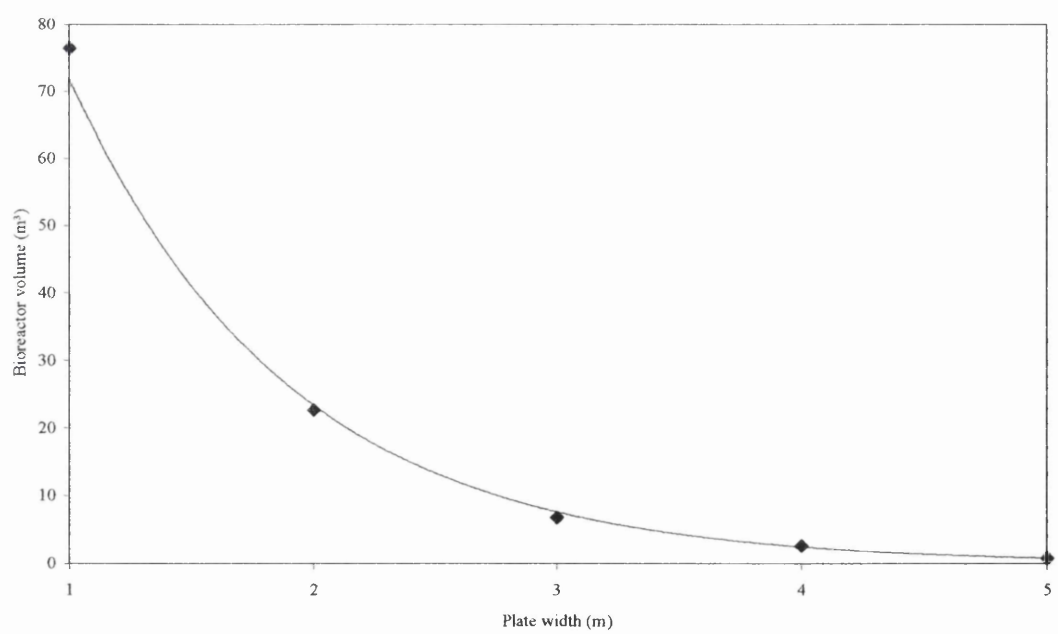


**Figure 5.56: Effect of plate width on bioreactor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

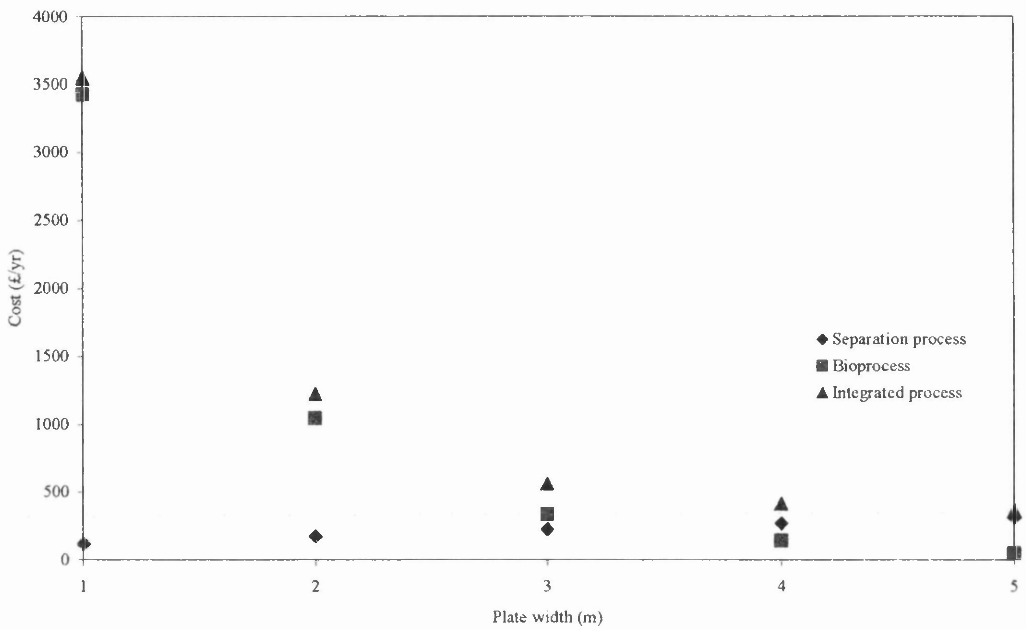


**Figure 5.57: Effect of plate width on cost of integrated process**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

### 5.3.2 Effect of varying interceptor plate angle

To investigate the effect of plate angle, a number of Corrugated Plate Interceptors were designed, using the models previously described in Chapters 2 and 3, to recover oil from a fixed oily wastewater stream. The plate angle was varied whilst holding influent characteristics and other design parameters constant. Interceptor dimensions, plate pack details, influent characteristics, and results are presented in Appendix IV. The cost of treatment is presented below.

Figure 5.58 shows the effect of the plates' angle of inclination to the horizontal on oil recovery in the Corrugated Plate Interceptor. As the angle of inclination is reduced from 60° to 10°, oil recovery increases from 0.0065 to 0.0105 $m^3$/h and the revenue generated increases from £16.14 to £26.10 per year (Figure 5.59). Figure 5.60 shows that decreasing plate angle reduces the cost of the separation; the cost of separation decreases from £120.28 to £108.52 per year. The effect of plate angle on interceptor volume is shown in Figure 5.61: as plate angle is decreased, interceptor volume increases to a maximum of 2.66 $m^3$ and afterwards decreases to a minimum value of 1.94 $m^3$ at 10° to the horizontal. The effect of plate angle on the efficiency of the Corrugated Plate Interceptor is shown in Figure 5.62: as plate angle is decreased, interceptor efficiency increases from 52 to 84%. The increase in interceptor efficiency is as a result of increasing effective separation area as plate angle decreases. Increasing separation area for a constant flowrate decreases the flow velocity along the plates.

The effect of plate angle on the bioprocess is shown in Figures 5.63 and 5.64: as plate angle is decreased, the influent oil concentration into the bioreactor decreases; consequently oxygen demand decreases from 1.66 to 0.55 g/l.h and bioreactor volume decreases from 76.4 to 25.2 $m^3$. In Figure 5.65, as plate angle is decreased, the cost of the bioprocess decreases from £3428.27 to £1156.72 per year; the cost of the integrated process decreases from £3548.56 to £1265.24 per year.

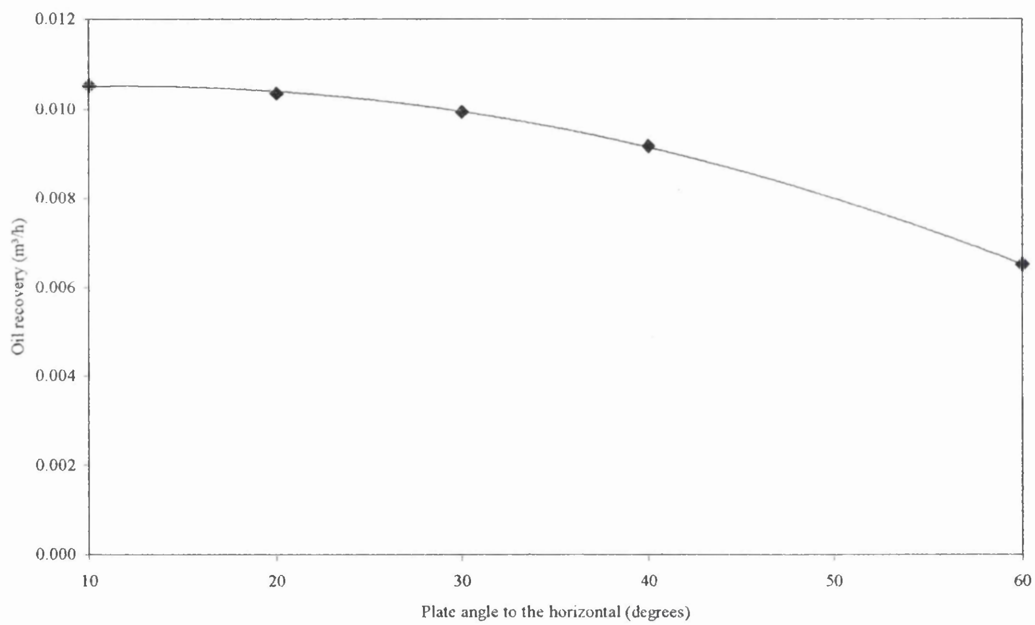


**Figure 5.58: Effect of plate angle on oil recovery**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

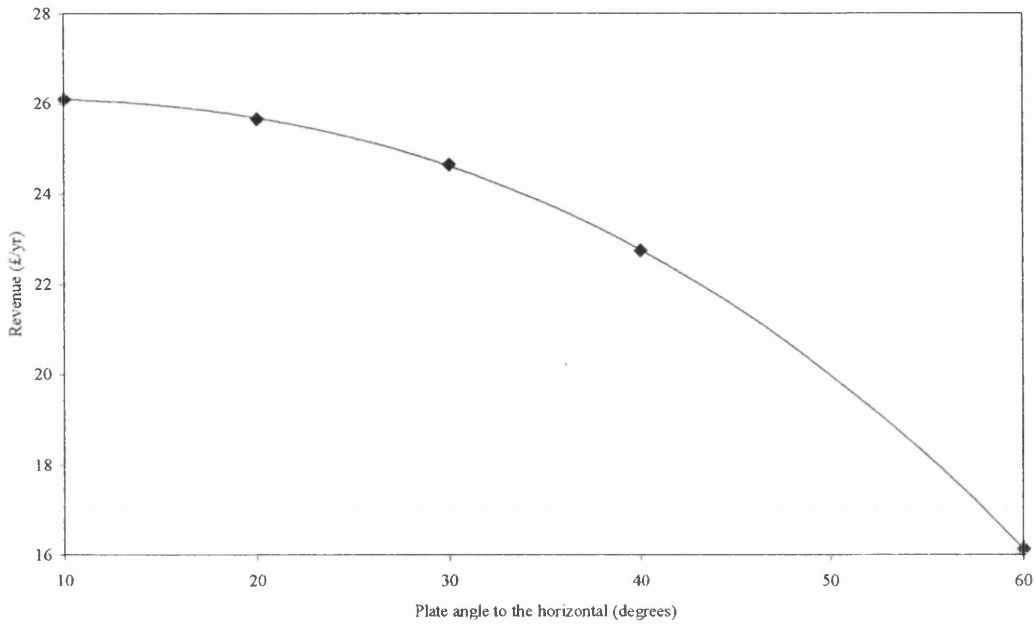


**Figure 5.59: Effect of plate angle on revenue generated from recovered oil**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

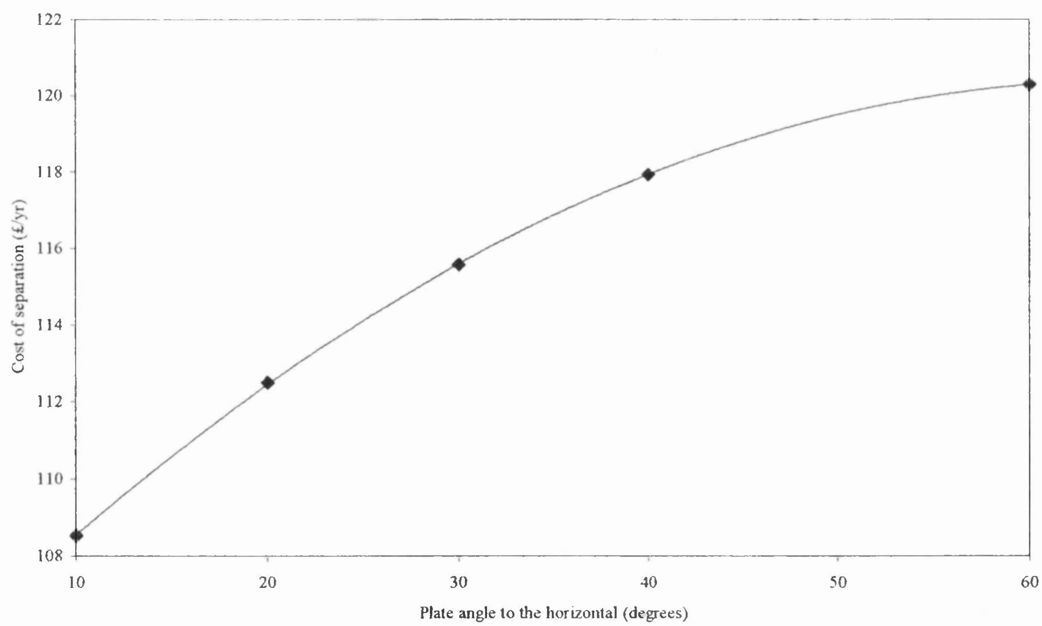


**Figure 5.60: Effect of plate angle on cost of separation**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

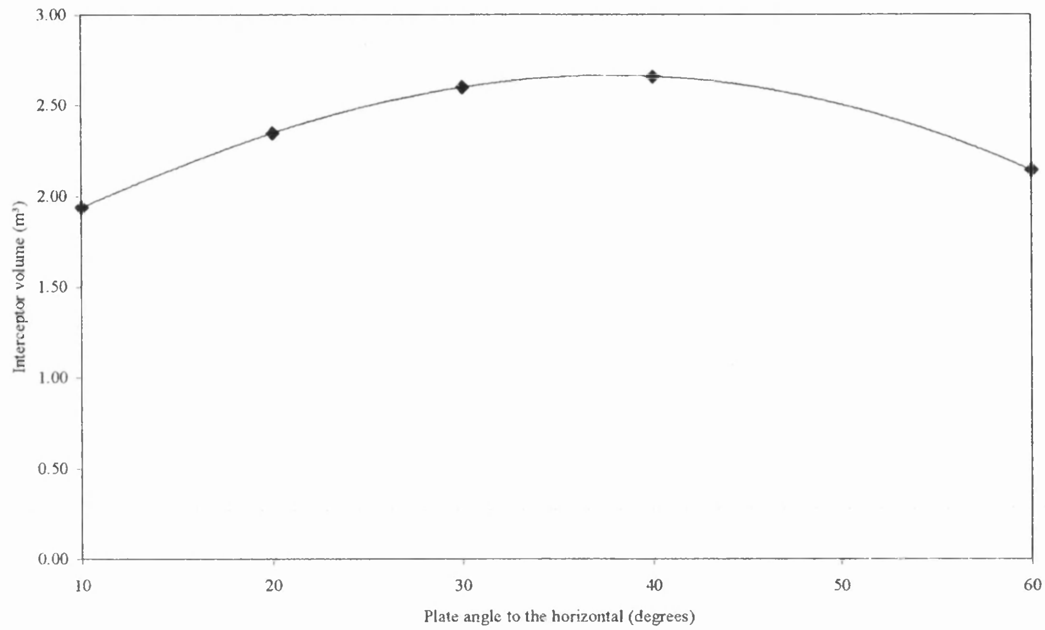


**Figure 5.61: Effect of plate angle on interceptor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

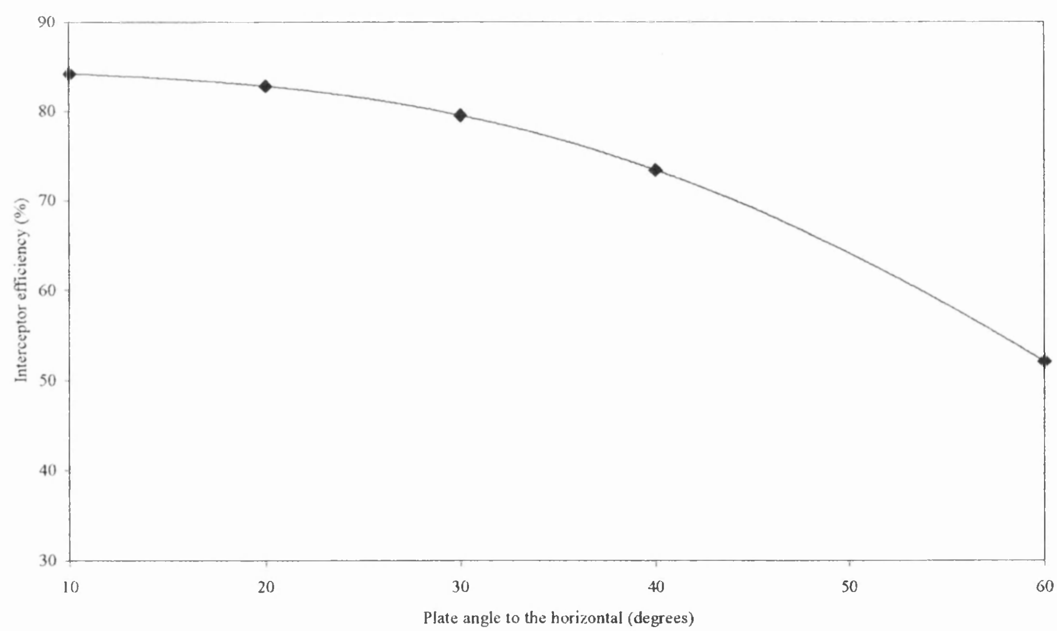


**Figure 5.62: Effect of plate angle on interceptor efficiency**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

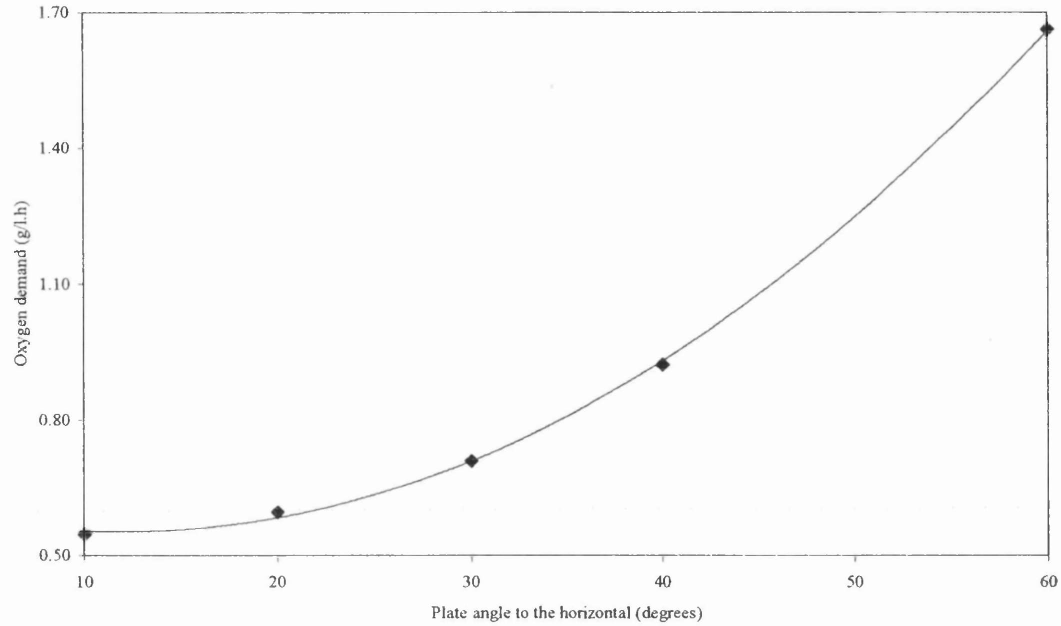


**Figure 5.63: Effect of plate angle on oxygen demand**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

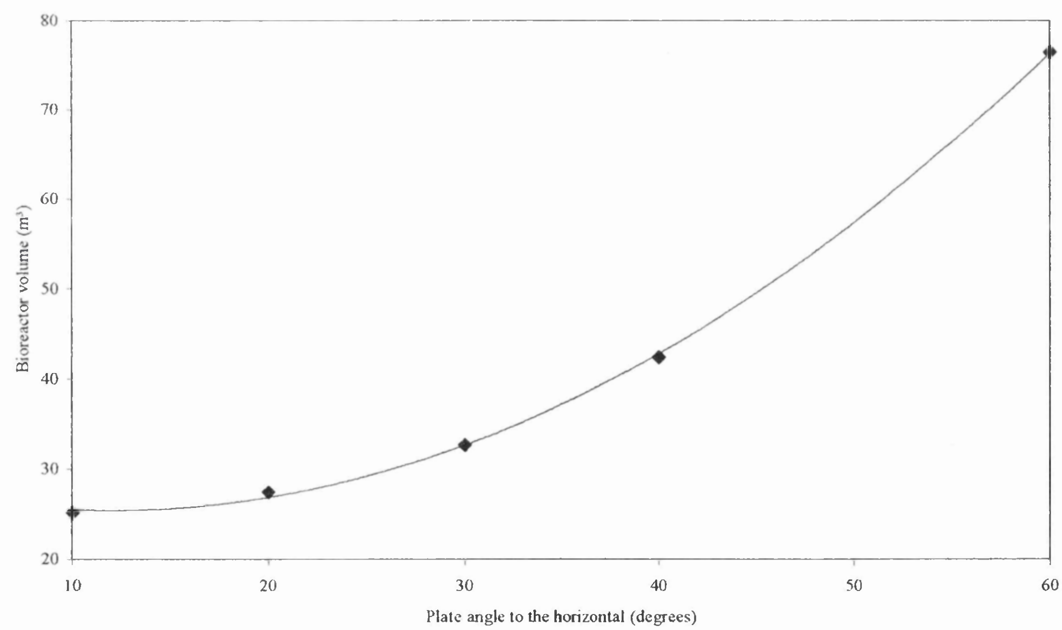


**Figure 5.64: Effect of plate angle on bioreactor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

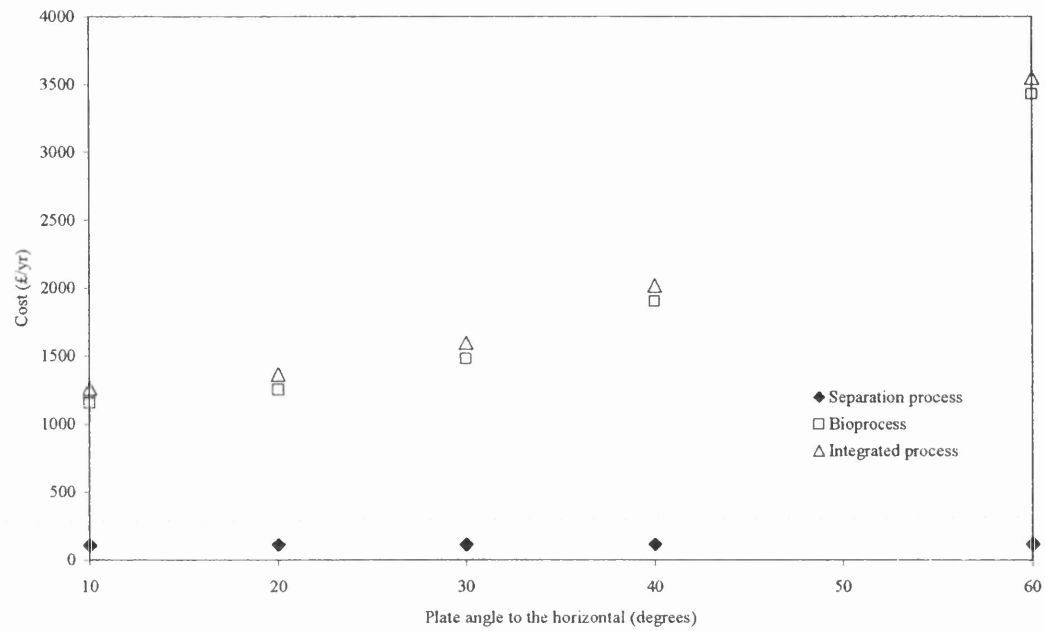


**Figure 5.65: Effect of plate angle on cost of integrated process**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

### 5.3.3 Effect of varying interceptor plate spacing

To investigate the effect of plate spacing, a number of Corrugated Plate Interceptors were designed, using the models previously described in Chapters 2 and 3, to recover oil from a fixed oily wastewater stream. The plate spacing was varied whilst holding influent characteristics and other design parameters constant. Interceptor dimensions, plate pack details, influent characteristics, and results are presented in Appendix IV. The cost of treatment is presented below.

Figure 5.66 shows that as plate spacing is increased from 0.02 to 0.05 m, oil recovery increases from 0.0065 to 0.0068 $m^3/h$ and interceptor volume increases from 2.14 to 2.38 $m^3$ (Figure 5.67), to accommodate the increased plate spacing. Figure 5.68 shows that as the plate spacing is increased, the cost of separation increases from £120.28 to £121.51 per year; the revenue generated from recovered oil increases from £16.14 to £16.92 per year (see Appendix IV). The increase in revenue is due to the increasing volume of oil recovered as plate spacing is increased.

The increase in oil recovery with increasing plate spacing for the same flowrate, influent oil concentration, number of plates, plate angle and plate dimensions, is due to the reduction in flow velocity along the plate. Increasing plate spacing means increasing the area the wastewater flows through. The greater the area for flow for the same flowrate, the lower the velocity of flow along the plates; the oil droplets have more time to rise out of the effluent flow path and hence the oil recovery is improved. Figure 5.69 shows that as plate spacing is increased the interceptor efficiency increases from 52.1 to 54.6%.

The effect of plate spacing on the bioprocess is shown in Figures 5.70 and 5.71. As plate spacing is increased, the oxygen demand of the bioprocess decreases from 1.66 g/l.h to 1.58 g/l.h and hence the bioreactor volume reduces from 76.4 to 72.4 $m^3$. The relationship between plate spacing and cost of the integrated process is shown in Figure 5.72 where the cost of the bioprocess decreases from £3428.27 to £3254.26 per year and the cost of the integrated process decreases from £3548.56 to £3375.77 per year.

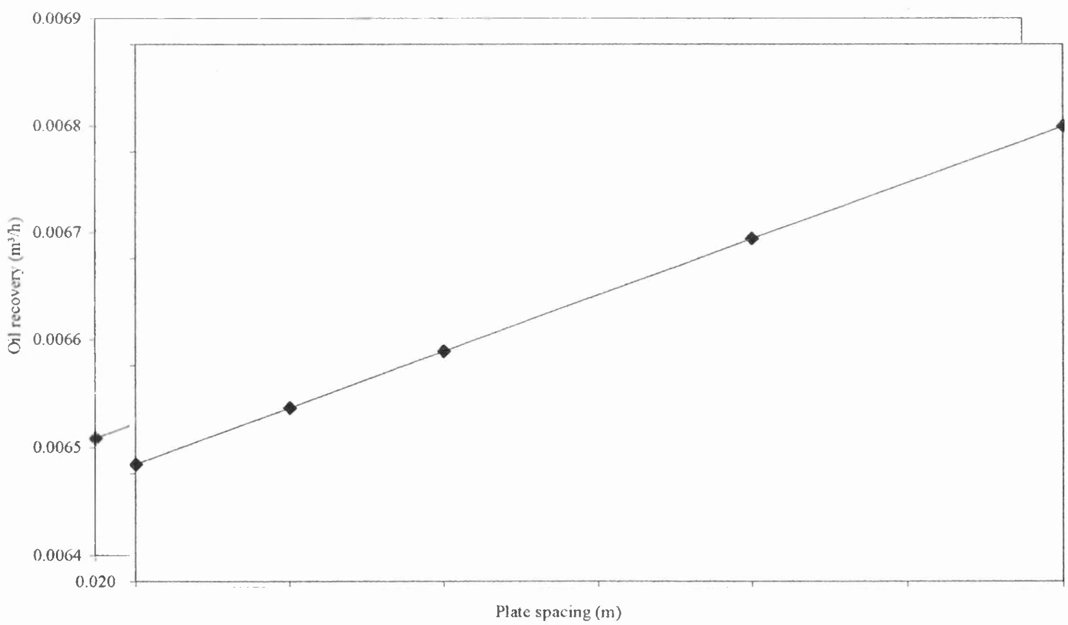


**Figure 5.66: Effect of plate spacing on oil recovery**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

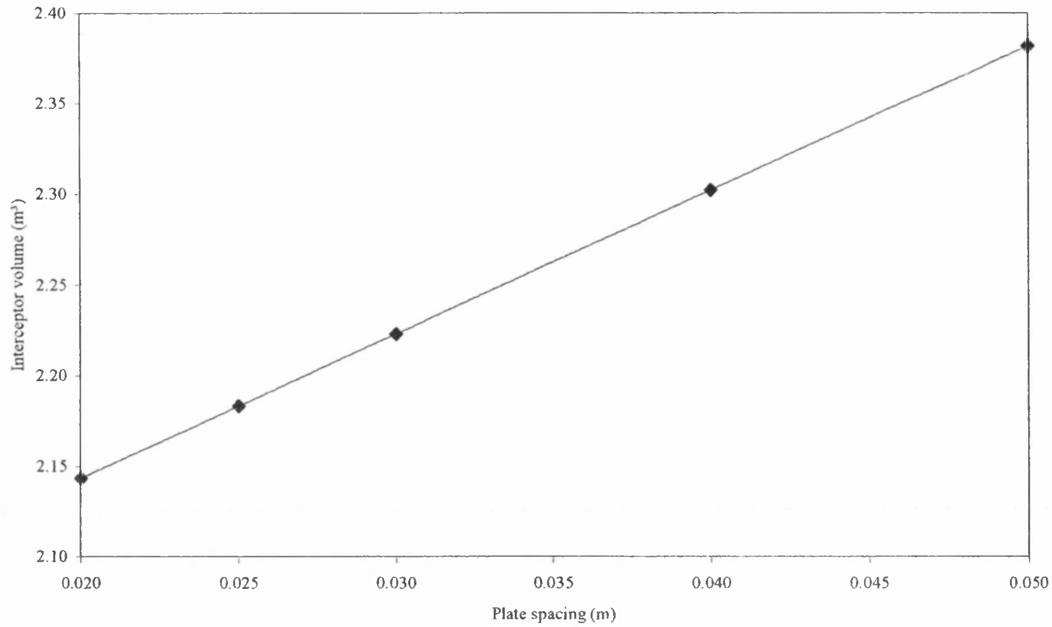


**Figure 5.67: Effect of plate spacing on interceptor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

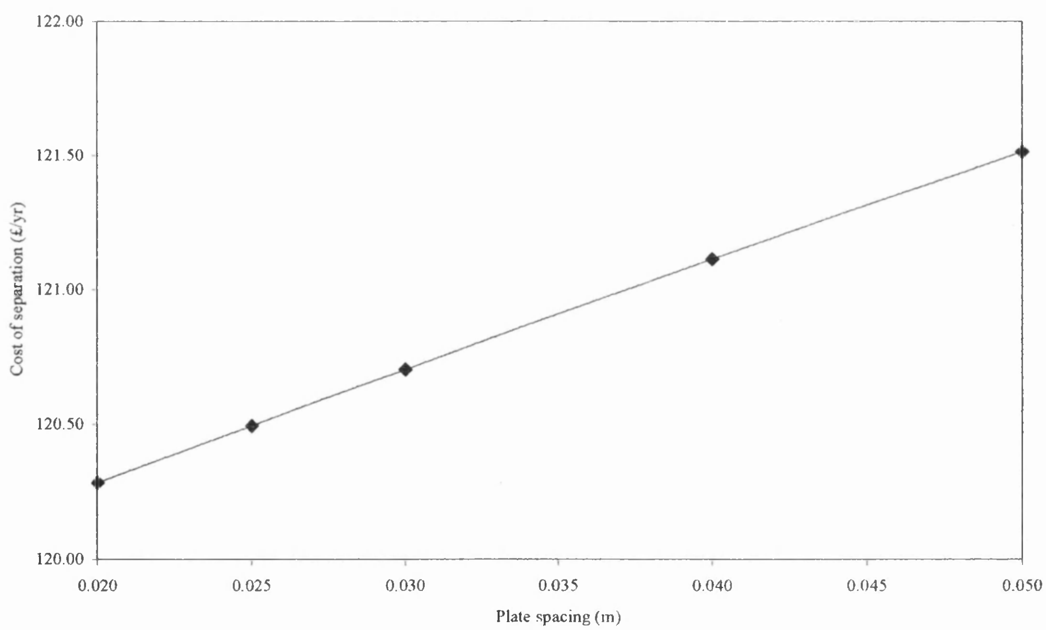


**Figure 5.68: Effect of plate spacing on cost of separation**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

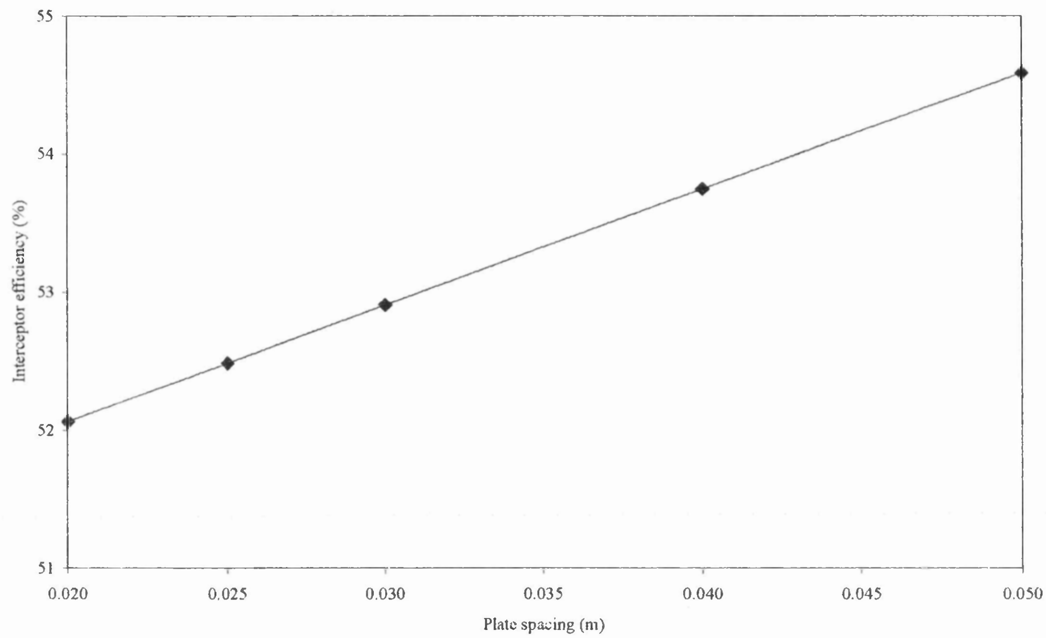


**Figure 5.69: Effect of plate spacing on interceptor efficiency**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

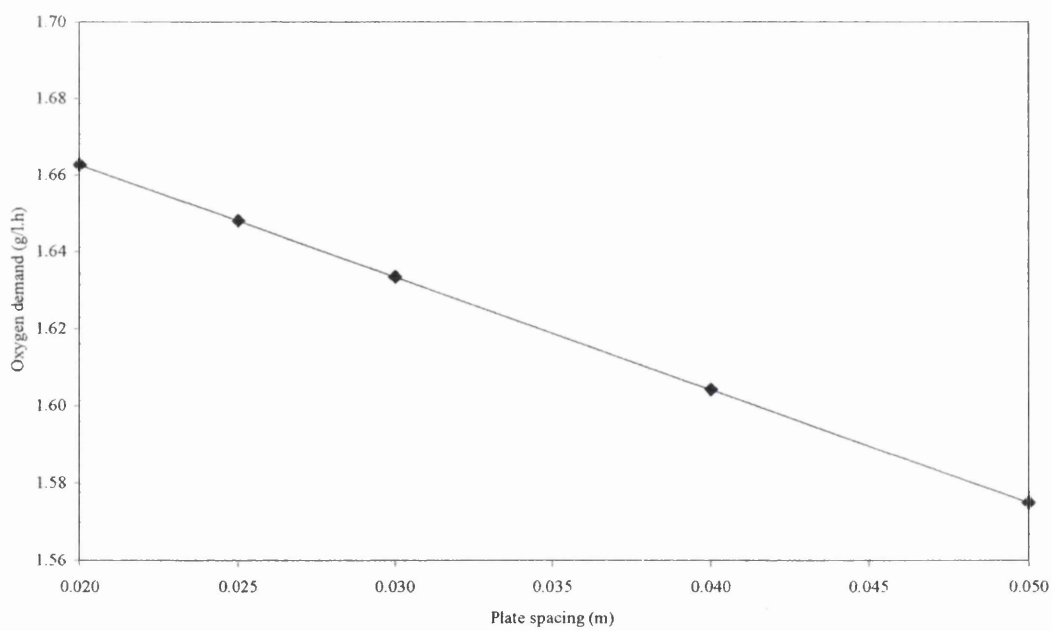


**Figure 5.70: Effect of plate spacing on oxygen demand**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

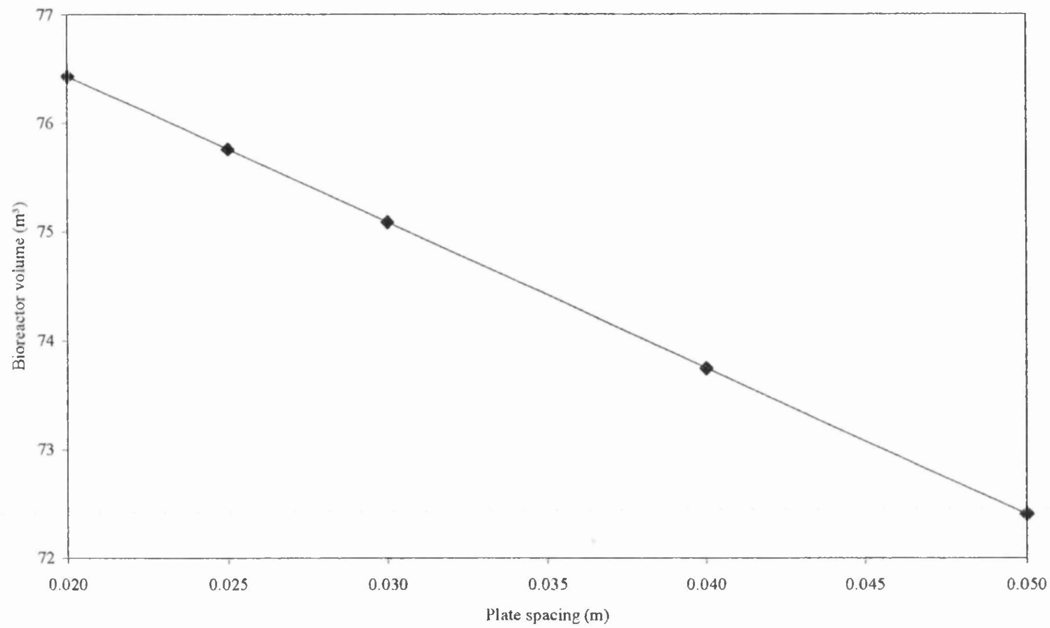


**Figure 5.71: Effect of plate spacing on bioreactor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

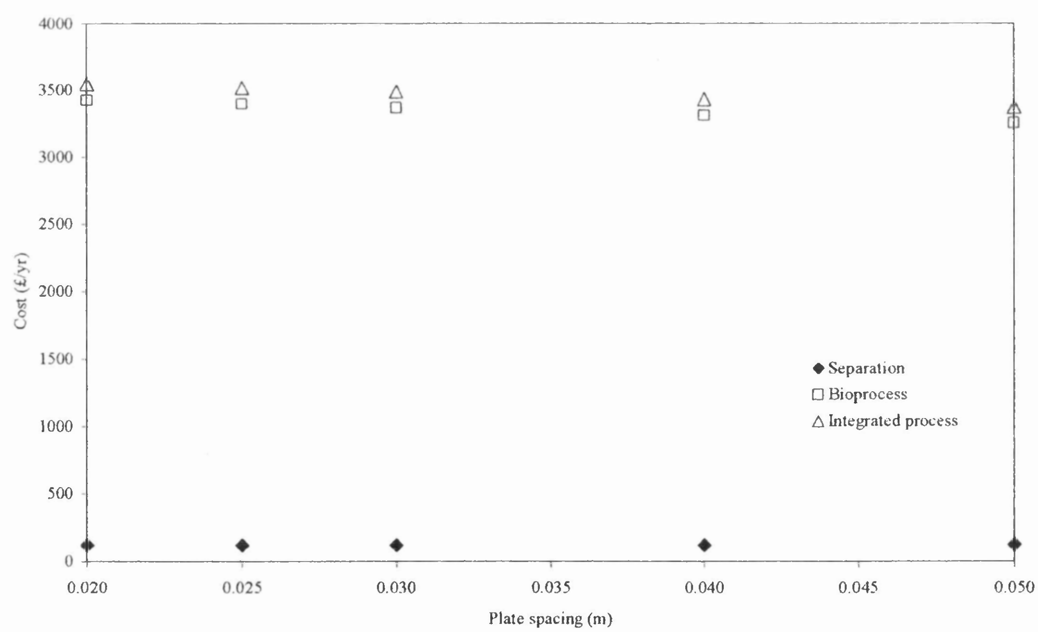


**Figure 5.72: Effect of plate spacing on cost of integrated process**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

### 5.3.4 Effect of varying interceptor plate length

To investigate the effect of plate length, a number of Corrugated Plate Interceptors were designed, using the models previously described in Chapters 2 and 3, to recover oil from a fixed oily wastewater stream. The plate length (parallel to flow) was varied whilst holding influent characteristics and other design parameters constant. Interceptor dimensions, plate pack details, influent characteristics, and results are presented in Appendix IV. The cost of treatment is presented below.

As plate length is increased from 1 to 5 m, oil recovery increases from 0.0065 to 0.0124 $m^3/h$ (Figure 5.73). The revenue generated then increases from £16.14 to £30.83 per year (Figure 5.74). Despite the increase in revenue, the cost of separation increases from £120.28 to £412.08 per year (Figure 5.79). The relationship between plate length and the size of the interceptor is shown in Figure 5.75: as plate length is increased, interceptor volume increases from 2.14 to 36.48 $m^3$. The relationship between plate length and efficiency is shown in Figure 5.76, as plate length is increased, efficiency increases from 52 to 99.5%. The increase in efficiency is as a result of increasing the effective separation area as plate length increases. Increasing the separation area for the same flow rate reduces the velocity of flow along the plates and therefore more droplets have enough time to rise out of the effluent path. The increase in plate length also reduces the velocity of flow, it does this in two ways; first, it increases the separation area for the same flow rate therefore reducing the velocity of flow along the plates. Secondly, it increases the frictional force along the plates, so for the same flow rate as plate length increases velocity of flow decreases.

The effects of interceptor plate length on the bioprocess are presented in Figures 5.77 and 5.78. As plate length is increased, influent oil concentration into the bioprocess decreases and hence oxygen demand decreases from 1.66 g/l.h to 0.018 g/l.h (Figure 5.77); bioreactor volume decreases from 76 $m^3$ to 849 litres (Figure 5.78). The effect of plate length on the integrated process is presented in Figure 5.79; the cost of the bioprocess reduces from £3428.27 to £52.63 per year and the cost of the integrated process reduces from £3548.56 to £464.71 per year.

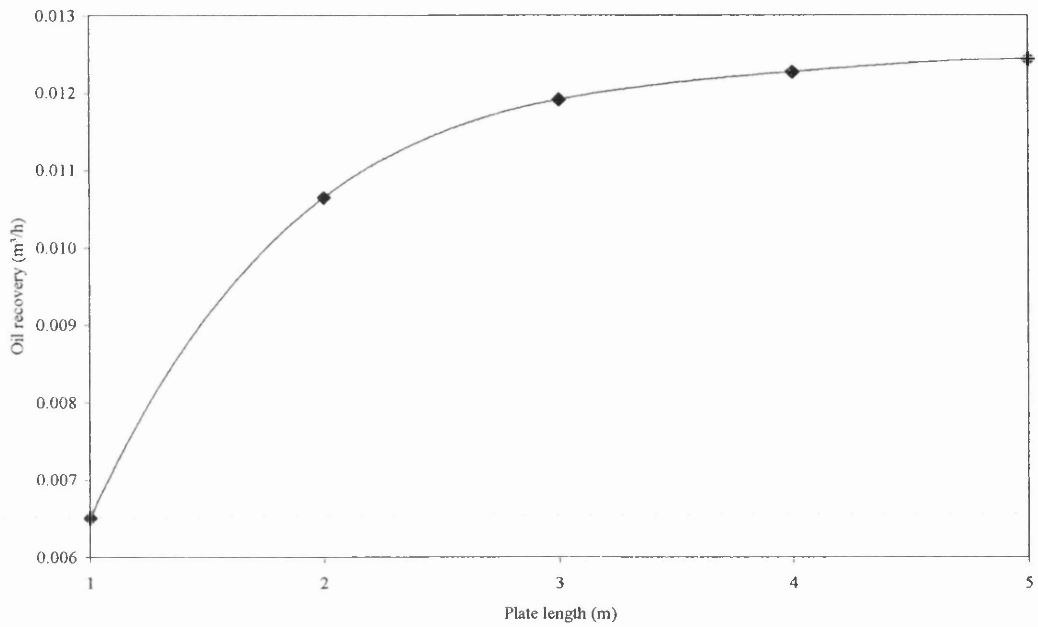


**Figure 5.73: Effect of plate length on oil recovery**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

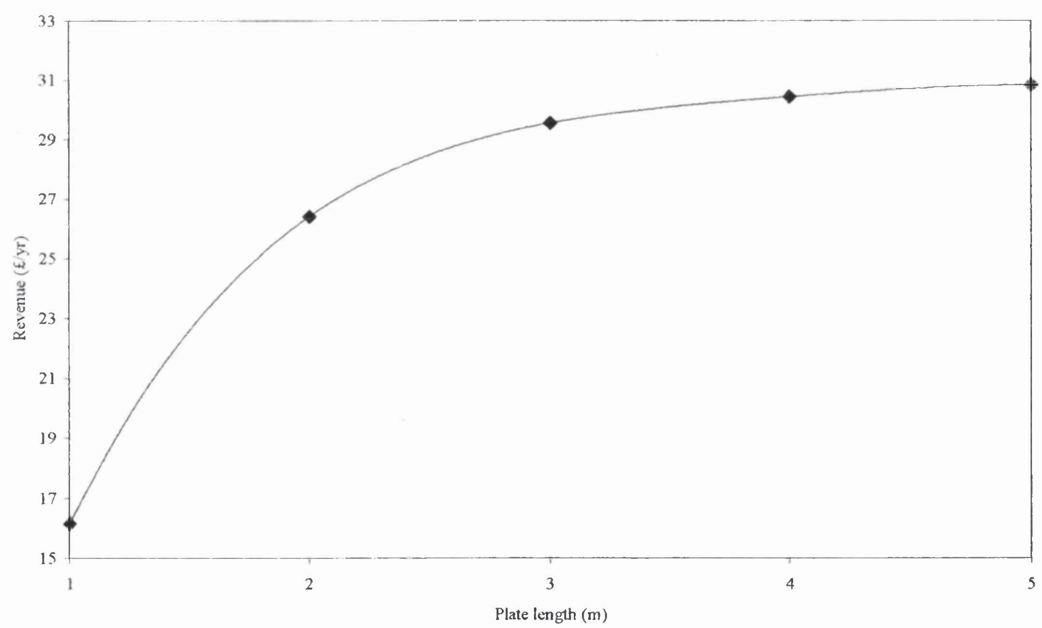


**Figure 5.74: Effect of plate length on revenue generated from recovered oil**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

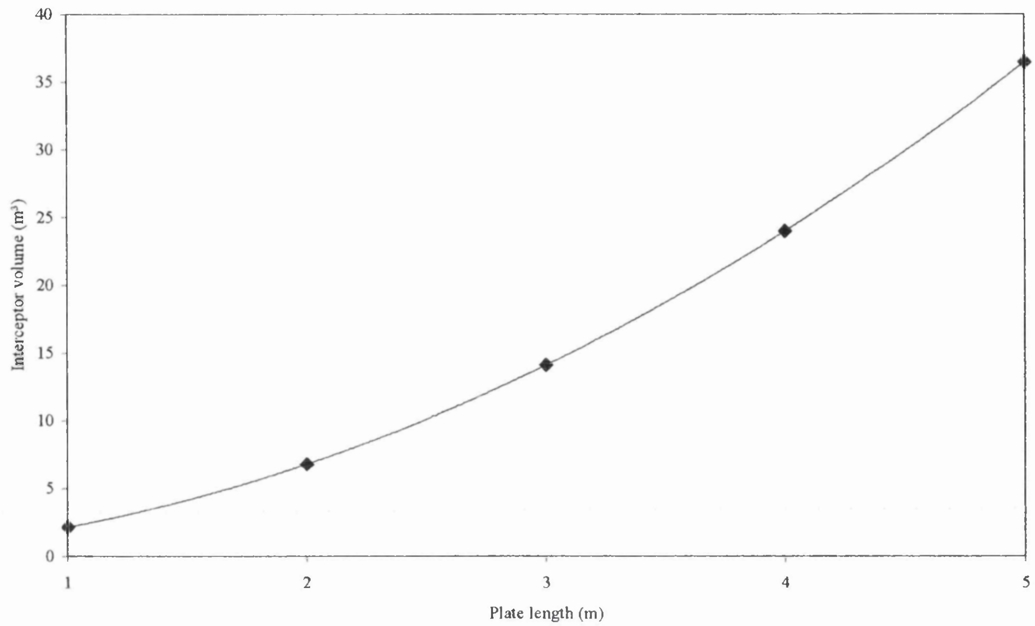


**Figure 5.75: Effect of plate length on interceptor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

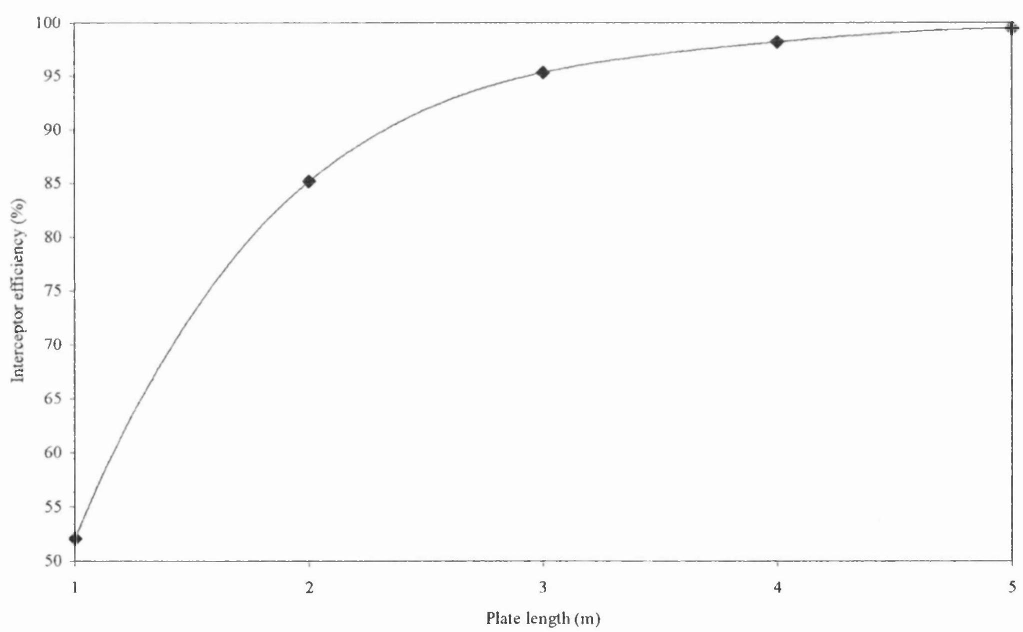


**Figure 5.76: Effect of plate length on interceptor efficiency**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

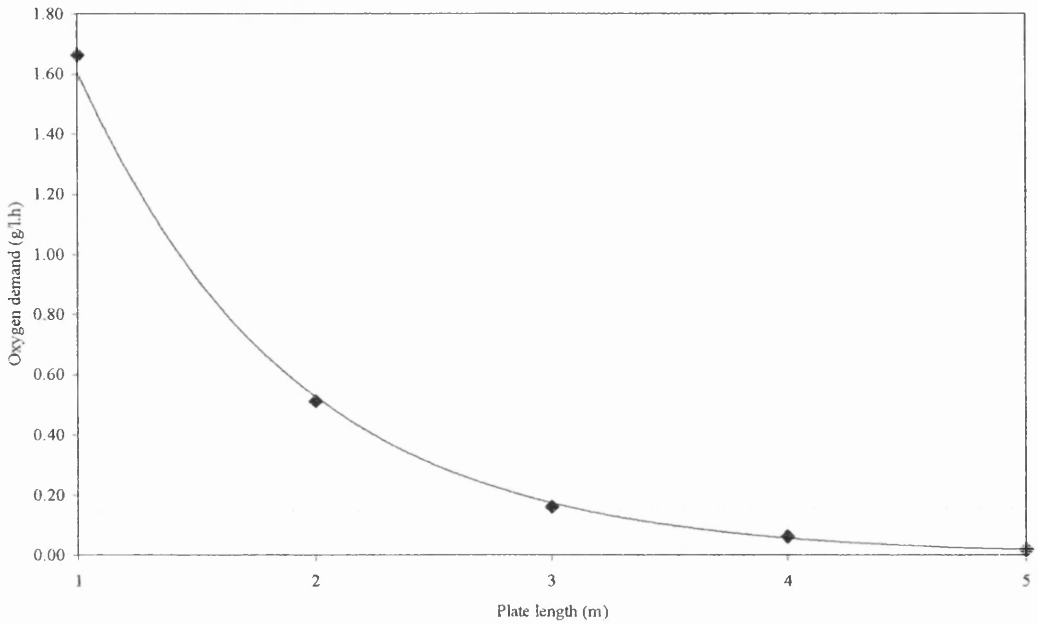


**Figure 5.77: Effect of plate length on oxygen demand**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

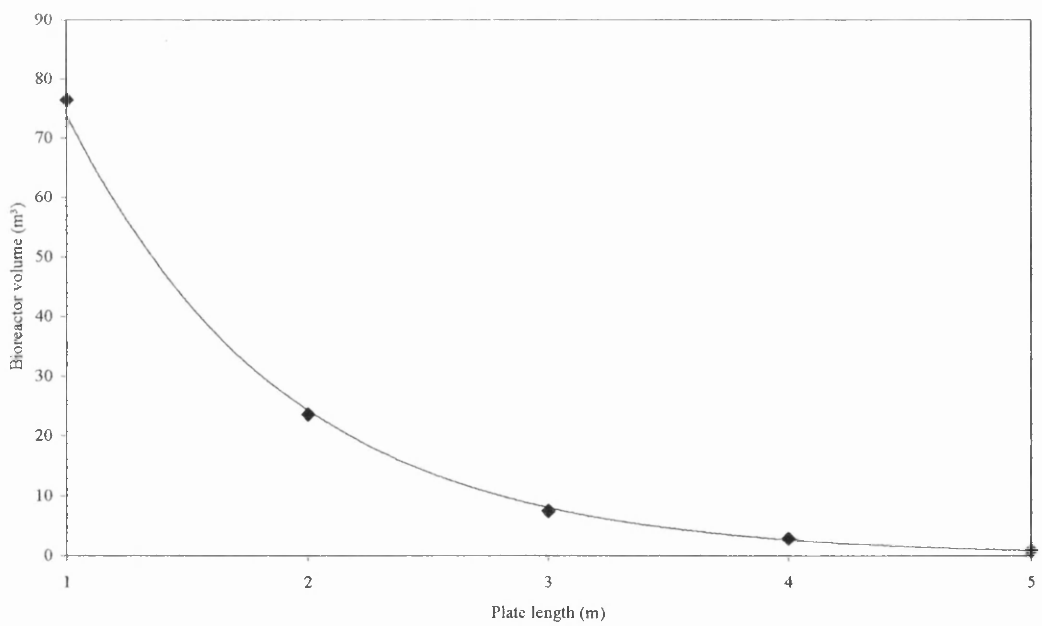


**Figure 5.78: Effect of plate length on bioreactor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

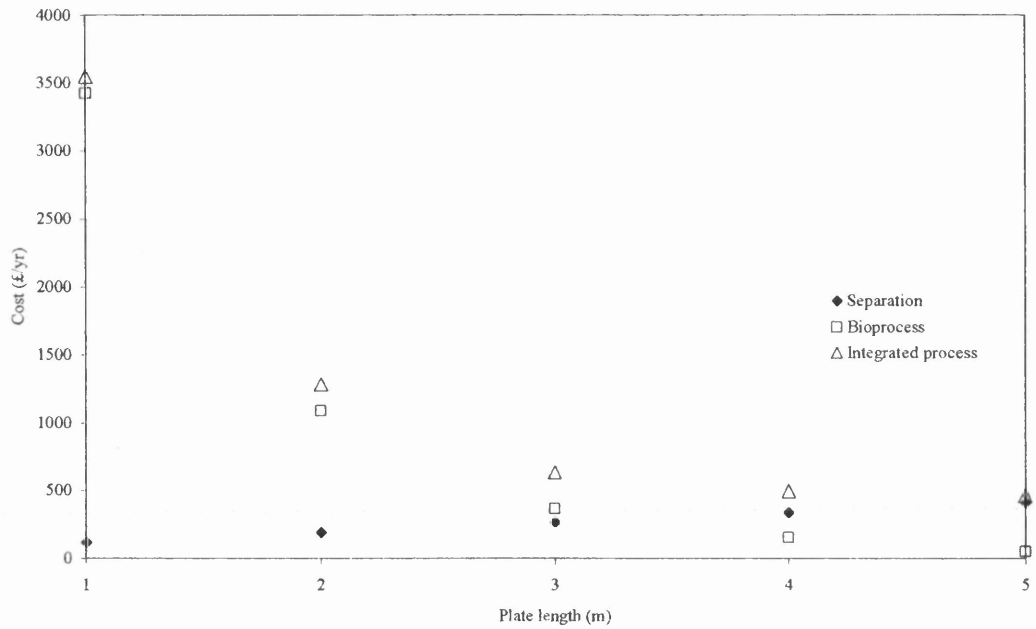


**Figure 5.79: Effect of plate length on cost of integrated process**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

### 5.3.5 Effect of varying number of plates in interceptor

To investigate the effect of the number of plates, a number of Corrugated Plate Interceptors were designed, using the models previously described in Chapters 2 and 3, to recover oil from a fixed oily wastewater stream. The number of plates was varied whilst holding influent characteristics and other design parameters constant. Interceptor dimensions, plate pack details, influent characteristics, and results are presented in Appendix IV. The cost of treatment is presented below.

As the number of plates is increased from 5 to 25, oil recovery increases from 0.0065 to 0.0124 $m^3/h$ (see Figure 5.80) and the revenue generated from recovered oil increases from £16.14 to £30.84 per year (see Figure 5.81). Figure 5.82 shows that interceptor volume increases from 2.14 to 2.94 $m^3$ as the number of plates is increased; this increase in interceptor volume is as a result of the fixed plate spacing considered in this investigation. It is possible that interceptor size may not be affected by increasing number of plates.

The number of plates also affects the efficiency of the separator. Figure 5.83 shows that efficiency increases from 52% to 99.5%. The increase in efficiency of the interceptor is as a result of increasing separation area. Increasing the number of plates in the interceptor increases the area available for separation and for coalescing of oil droplets. Increasing the number of plates for a given flow rate not only increases the area for separation and coalescing but also reduces the velocity of flow along the plates: the net effect therefore is an increase in oil recovery. Figure 5.84 presents the effect of number of plates on the cost of the interceptor. As the number of plates is increased, the cost of the interceptor increases from £120.28 to £282.88 per year.

The effects of the number of plates on the bioprocess are shown in Figures 5.85 and 5.86. As the number of plates is increased the oil concentration flowing into the bioreactor reduces and hence oxygen demand reduces from 1.66 to 0.018 g/l.h; bioreactor volume reduces from 76.4 $m^3$ to 817 litres.

The effect on the integrated process is presented in Figure 5.87, where the cost of the bioprocess reduces from £3428.27 to £50.93 per year, the cost of separation increases from £120.28 to £282.88 per year, and the cost of the integrated process reduces from £3548.56 to £333.81 per year.

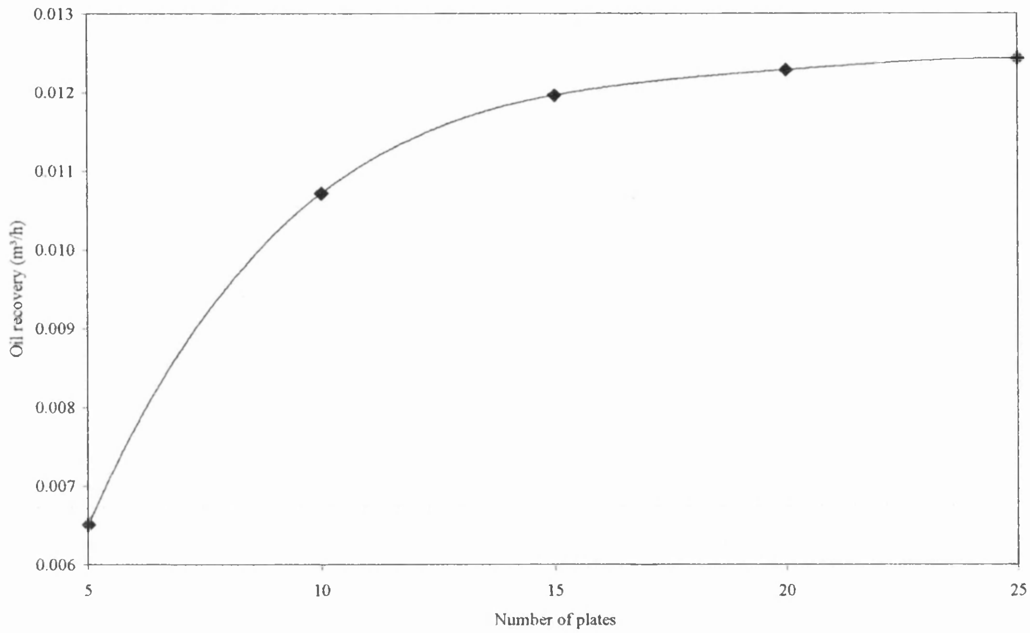


**Figure 5.80: Effect of number of plates on oil recovery**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

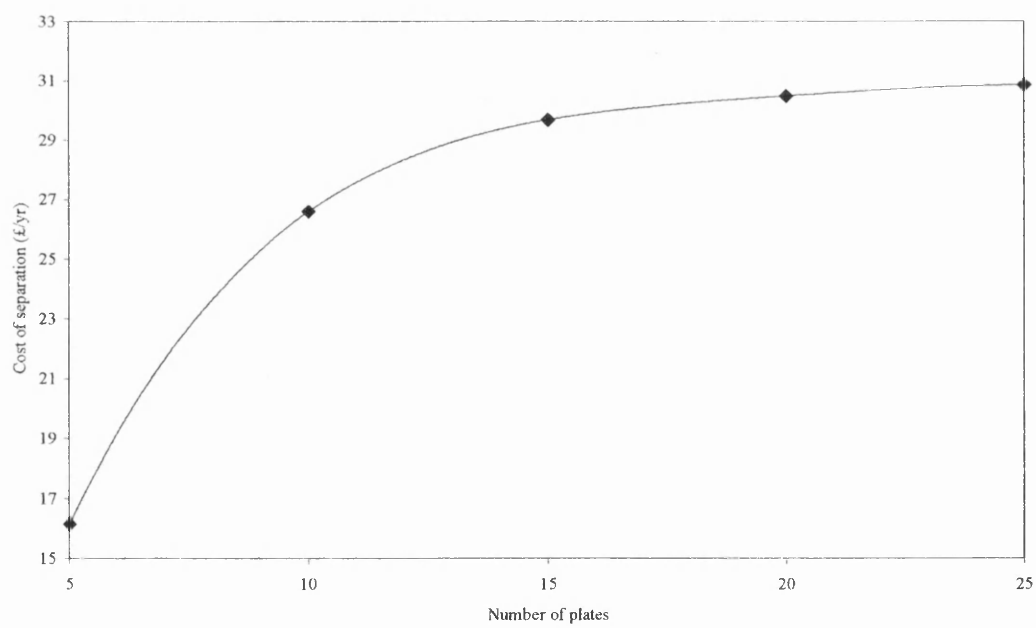


**Figure 5.81: Effect of number of plates on revenue generated from recovered oil**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

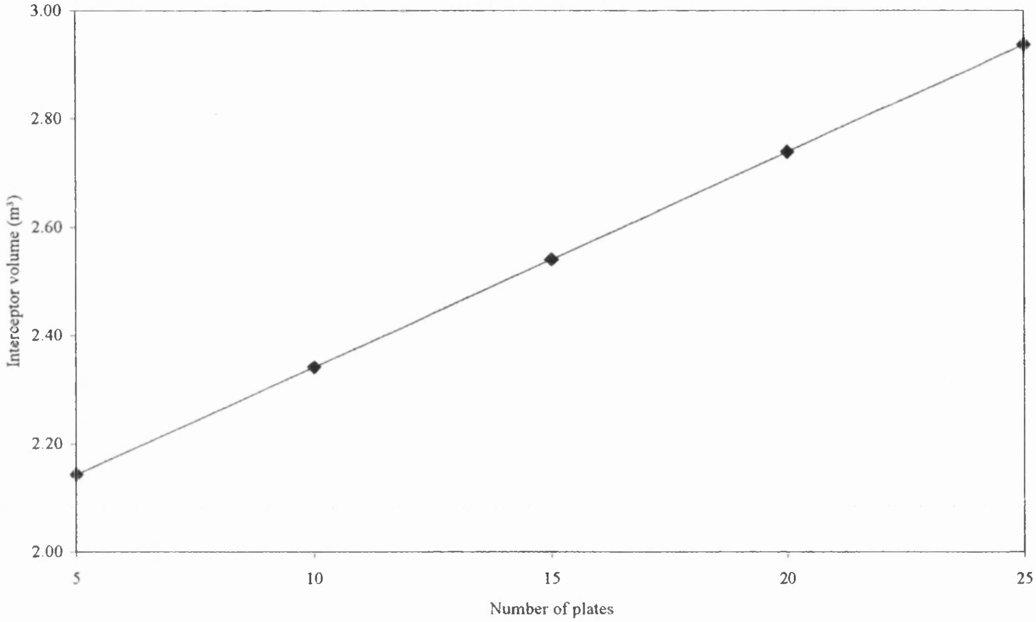


**Figure 5.82: Effect of number of plates on interceptor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

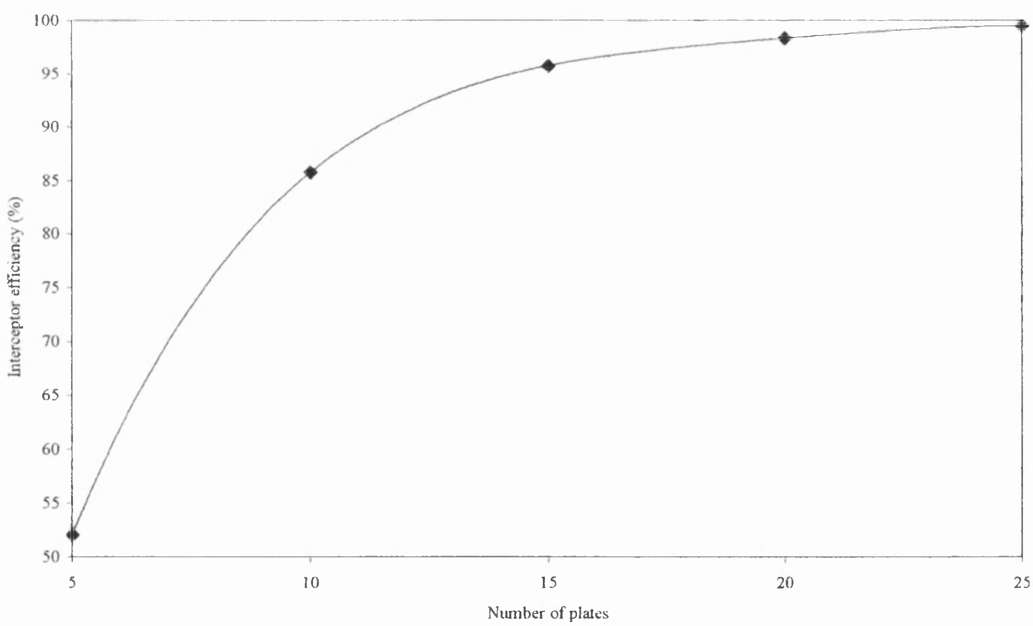


**Figure 5.83: Effect of number of plates on interceptor efficiency**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

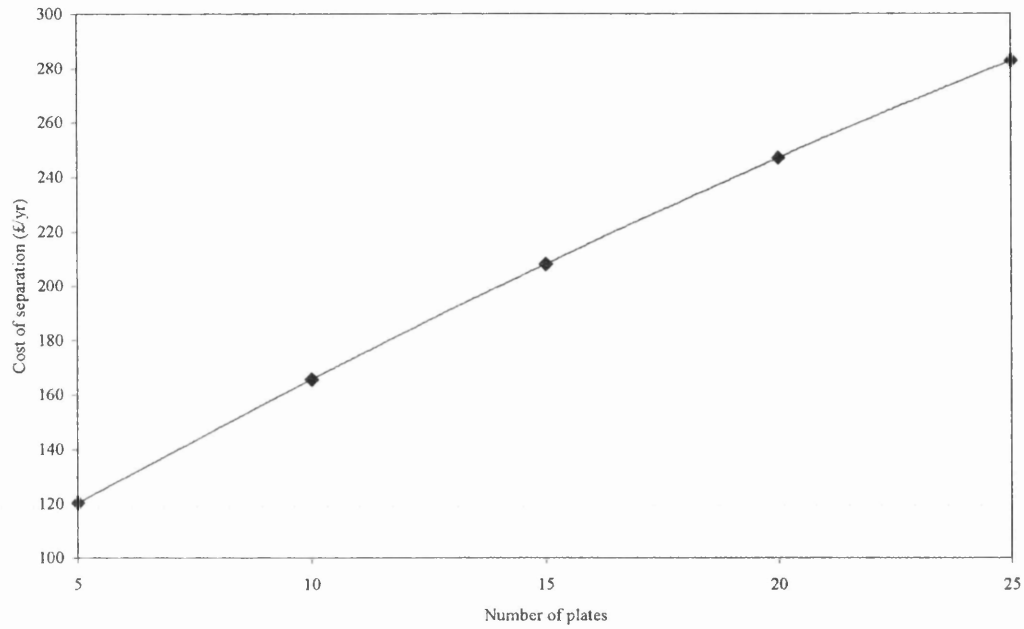


**Figure 5.84: Effect of number of plates on cost of separation**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

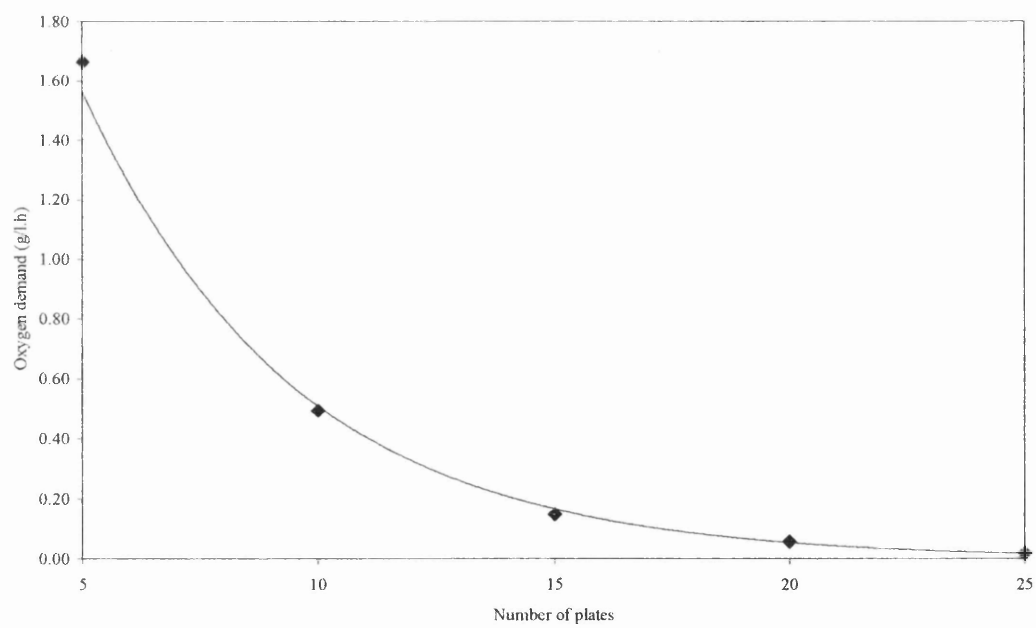


**Figure 5.85: Effect of number of plates on oxygen demand**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

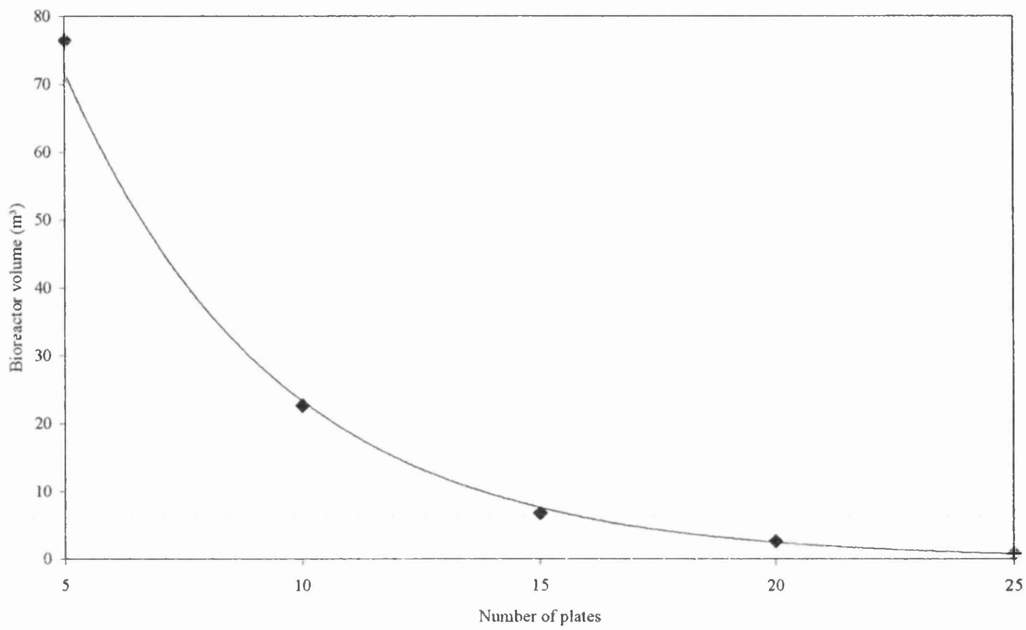


**Figure 5.86: Effect of number of plates on bioreactor volume**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

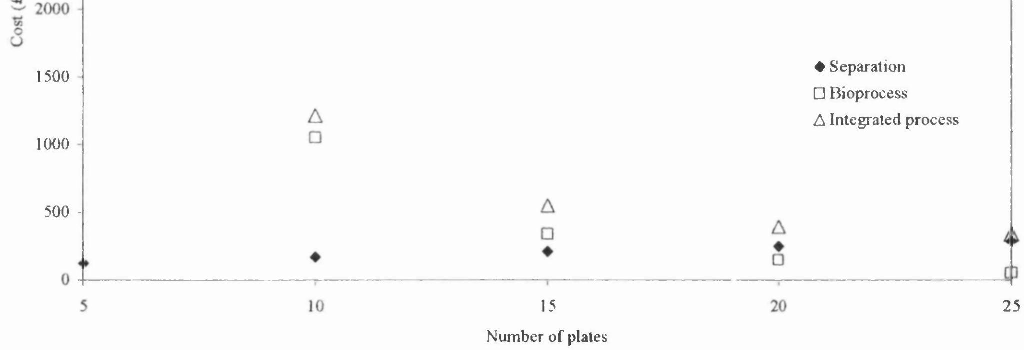


**Figure 5.87: Effect of number of plates on cost of integrated process**

(Influent oil characteristics, interceptor dimensions, and performance were obtained using models developed in Chapter 3. The details are presented in Appendix IV).

### 5.3.6 Discussion of results and analysis

#### 5.3.6.1 Effect of plate width on the integrated process

From Figure 5.50, it can be seen that increasing plate width to increase oil recovery improves oil recovery significantly in an almost linear fashion. Beyond a plate width of 3 m the trend changes and further increase in plate width does not result in a significant increase in oil recovery; however, there is an approximately linear relationship between plate width and cost of separation. As plate width increases, cost of separation increases significantly and there is no limiting effect.

Figure 5.57 shows that as plate width increases, cost of separation increases and cost of the bioprocess decreases; this implies there is a plate width which minimizes the cost of the integrated process. Therefore, in the integration of a corrugated plate interceptor and an immobilized cell bioreactor, the plate width which minimizes the cost of integration should be obtained.

#### 5.3.6.2 Effect of plate angle on the integrated process

From Figures 5.60 and 5.62, decreasing plate angle to the horizontal improves separation efficiency and reduces cost. However, in Figure 5.61, the effect of plate angle on interceptor volume follows a slightly different trend at a plate angle of 40°. This is because the interceptor designed as the base interceptor with a plate angle of 60° to the horizontal is inefficient. At the plate angle of 40° to the horizontal, the separator size increases and the efficiency of the base interceptor is improved, hence the abnormality in the trend. In general, for a well-designed interceptor, decreasing plate angle to the horizontal decreases the interceptor volume and increases efficiency. It can be seen from Figures 5.60, 5.62 and 5.65 that operating the interceptor at small angles to the horizontal increases efficiency and hence reduces the cost of the integrated process but in practice, this is not so. A plate angle between 45° and 60° is used in the industry. This is to enhance sludge removal, since smaller plate angles are less steep, sludge may take a longer time to slide down the plates.

#### 5.3.6.3 Effect of plate spacing on the integrated process

From Figure 5.66, it can be seen that increasing the spacing between plates increases oil recovery for the same flow rate. Figures 5.68 and 5.70 show that increasing the spacing between plates increases the cost of the interceptor but decreases the oxygen demand and hence the cost of the bioprocess. It follows therefore there is an optimal distance apart which minimizes the cost of the integrated process. No limiting effect was observed on the effect of plate spacing on interceptor efficiency (Figure 5.69).

### 5.3.6.4 Effect of plate length on the integrated process

From Figure 5.73, it can be seen that increasing the plate length (parallel to flow) in the corrugated plate interceptor does not result in significant increase in oil recovery beyond a plate length of 3 m; however, the cost of separation has an approximately linear relation with plate length (see Figure 5.79). It therefore follows that, beyond a certain length, increase in plate length does not significantly increase oil recovery but significantly increases cost of separation.

Figure 5.79 shows that while the cost of separation increases with increasing plate length, the cost of the bioprocess decreases, so there is expected to be an optimal plate length which minimizes the cost of an integrated process; this would be found if the curves are extrapolated to longer plate lengths.

### 5.3.6.5 Effect of number of plates

From Figure 5.80, increasing the number of plates in the interceptor increases oil recovery and like the other variables such as plate length and plate width, there is an optimal number of plates beyond which increase in number does not significantly increase oil recovery; in Figure 5.80 the optimal number of plates is about 15. Since the cost of the separation increases and cost of the bioprocess reduces with increasing number of plates, it follows that there will be a number of plates at which the cost of integration is minimized.

In summary, the number of plates, plate spacing, angle of inclination to the horizontal, and plate dimensions all affect oil recovery and hence the cost of an integrated process. Plate dimensions and number of plates have a limiting effect on oil recovery which means beyond a certain value, increase in any of these factors does not result in significant increase in oil recovery so there is a need for these variables to be optimized. This effect was also noticed in the horizontal flow separators where increase in separator volume has a limiting effect on efficiency. Number of plates, plate dimensions, and separator volume are all related to surface area. It follows therefore that all factors which improve oil recovery by increasing surface area have limiting effects on efficiency and should be optimized for profitable operations.

The spacing between plates and angle of inclination does not have a limiting effect on oil recovery but there is a need to optimize spacing between plates because there is significant increase in size of interceptor as a result of increasing plate spacing. Construction of large tanks to improve oil recovery is not appreciated in the industry. The practice of not reducing the plate angle to low values has already been discussed.

Analysis of these results shows that, unlike the horizontal flow separator, all design parameters that affect oil recovery in the Corrugated Plate Interceptor result in an increase in interceptor volume. The industry appreciates a small footprint and since efficiency entails an increase in separator size, a variable which has the least effect on interceptor volume and the most effect on increasing efficiency would be the first choice in optimizing the process. The next section therefore compares the various effects of these factors on interceptor volume and efficiency so as to determine the best configuration which entails a small change in separator size and big positive change in efficiency.

#### 5.3.6.6 Comparing the various parameters

To identify the critical factor for improving oil recovery in a corrugated plate interceptor, the following variables are compared: number of plates, plate angle, plate spacing, and plate dimensions. This section compares the effect of these factors on interceptor size (governed by the size of the plate pack) and efficiency of oil recovery.

| Change in any variable (%) | Change in interceptor volume due to decrease in plate angle (%) | Increase in interceptor volume due to increase in plate spacing (%) | Increase in interceptor volume due to increase in plate length (%) | Increase in interceptor volume due to increase in plate width (%) | Increase in interceptor volume due to increase in number of plates (%) |
|---|---|---|---|---|---|
| 0 | 0 | 0 | 0 | 0 | 0 |
| 25 | 22 | 2 | 43 | 17 | 2 |
| 50 | 21 | 3 | 94 | 33 | 5 |
| 100 | -38 | 7 | 219 | 66 | 9 |

**Table 5.4: Effect of various factors on interceptor volume**

| Change in any variable (%) | Increase in interceptor efficiency due to decrease in plate angle (%) | Increase in interceptor efficiency due to increase in plate length (%) | Increase in interceptor efficiency due to increase in plate spacing (%) | Increase in interceptor efficiency due to increase in plate width (%) | Increase in interceptor efficiency due to increase in number of plates (%) |
|---|---|---|---|---|---|
| 0 | 0 | 0 | 0 | 0 | 0 |
| 25 | 33 | 22 | 1 | 21 | 23 |
| 50 | 52 | 40 | 2 | 38 | 41 |
| 100 | 60 | 64 | 3 | 64 | 64 |

**Table 5.5: Effect of various factors on plate interceptor efficiency**

Table 5.4 compares effects the effects of decreasing plate angle, increasing plate spacing, increasing plate dimensions, and increasing number of plates on interceptor volume. From Table 5.4, a 25% decrease in plate angle resulted in a 22% increase in interceptor volume; as explained earlier this was an anomaly - decreasing plate angle decreases interceptor size as can be seen in the 100% decrease in plate angle.

Increasing the number of plates increases the volume of the interceptor necessary to accommodate the plate pack. From Table 5.4, a 25% increase number of plates resulted in a 2% increase in interceptor volume while a 25% increase in plate width and plate length resulted in a 17% and 43% increase in interceptor volume. From Table 5.4, increasing the plate length has the most effect on separator volume compared to all other factors. Comparing number of plates and plate dimensions, increasing the number of plates has the least effect on interceptor volume.

Decreasing plate angle reduces the size of the interceptor; the anomaly in trend in Table 5.4 has been explained previously. However, in practice plate angles of 45-60° are preferred for easy removal of sludge.

From Table 5.4, increasing plate spacing to improve recovery has the smallest effect on interceptor volume, compared to increasing number of plates and plate dimensions. For a 25% increase in plate spacing interceptor volume increased by only 2%.

From Table 5.5, increasing plate spacing has the least effect on efficiency when compared with all other factors. Increasing plate spacing does not increase the surface area available for separation but reduces the velocity of flow. For a 25% increase in plate spacing, efficiency increased only by 2%. Plate angle will not be analysed here, even though this has the greatest effect on efficiency, because in practice the plate angle must be between 45-60° for easy removal of sludge.

The effect of number of plates, plate dimensions on efficiency are comparable. For a 25 % increase in any of these factors, the increase in efficiency was between (21-23%). The effects are comparable because they increase surface area and reduce velocity.

The critical design factor which improves oil recovery and reduces cost of treatment should be based on the factor which has the least effect on interceptor volume but also improves the efficiency of the interceptor. The number of plates has the least effect on separator volume whilst the effect of increase in number of plates on oil recovery is comparable to that achieved by increasing plate dimensions. Increasing the number of plates is therefore better than increasing plate dimensions as the impact on the overall cost of the process is smaller.

#### 5.3.6.7 Comparing the use of baffles and plates to improve efficiency in horizontal flow separators

Oil recovery in horizontal flow separators can be enhanced by the addition of baffles inside the unit; these baffles do not increase the size of the separator but reduce the velocity of flow. In Section 5.2, it was shown that reducing the velocity of flow is more effective in improving separator efficiency than increasing droplet size; based on other considerations the use of baffles improves the cost of integration more than the use of chemicals.

The effect of addition of baffles or plates on oil recovery can be compared by analysing the effect of plate spacing and number of plates on oil recovery in the corrugated plate interceptor. Increasing plate spacing reduces flow velocity along the plates and increasing number of plates increases surface area. From Table 5.5, increasing number of plates improves oil recovery more than increasing plate spacing. This is because the plates not only increase surface area - they reduce the velocity of flow as well. Therefore, in improving oil separation in horizontal flow separators, addition of plates seems to be better than addition of baffles, though this may not improve the cost of treatment due to the high cost of plates compared to the use of baffles. Therefore improving oil recovery in horizontal flow separator should be a matter of design rather than manipulation of influent oil characteristics.

#### 5.3.6.8 Conclusions

Various means to improve oil recovery in a CPI were investigated by altering the design of a base-case CPI unit. The effectiveness of the CPI was improved by increasing the number of plates and increasing plate dimensions rather than increasing plate spacing, and reducing plate angle to the horizontal. The investigations revealed the following:

- Numbers of plates and plate dimensions have a limiting effect on improving separator efficiency but varying the plate angle and distance between plates did not produce such effects but rather had a linear relation with interceptor efficiency.
- Increasing number of plates, plate dimensions and distance between plates for a given flow rate improved the separation efficiency while decreasing the plate angle improved separator efficiency.

All the design factors which improve separator efficiency involve increasing interceptor volume. The only factor which does follow this trend is the decreasing the plate angle, it not only improves separator efficiency but decreases interceptor size (for lower plate angles). However, the plate angle is limited to between 45 and 60° for practical reasons.

Amongst the design factors which improve oil recovery, increasing the number of plates has the least effect on interceptor volume; therefore this becomes the critical factor for minimizing the cost of separation and integration.

# Chapter 6: Conclusions and Recommendations

## 6.1 Conclusions

Common industrial practice is to integrate processes for the separation and treatment of oily wastewater. Integration is necessary because, for the majority of wastewater streams, no single unit can economically meet the environmental discharge limits and recover valuable oil. With the increasing price of oil, and ever tighter discharge limits, the industry seeks ways of improving the cost-effectiveness of integrated processes for oily wastewater treatment. To this end, this research concerned itself with the study of two integrated processes for oily wastewater treatment:

- Integration of horizontal flow separator and Immobilized Cell Bioreactor
- Integration of Corrugated Plate Interceptor and Immobilized Cell Bioreactor

By studying the various factors which affect oil recovery in the integrated processes and their effect on the cost of the integrated process, a solution to improving the cost effectiveness of integrated processes was reached. To further improve the cost effectiveness of integrated treatment processes using an aerobically operated ICB, aeration in an ICB was studied, since aeration constitutes a major cost in ICB operation.

The results obtained from the studies suggest that the solution to improving the cost effectiveness of integrated processes lies in the design of the units used in the process, rather than manipulation of the wastewater characteristics. For the integrated process of horizontal flow separator and ICB, a 33% decrease in horizontal velocity increased separator efficiency by 63% in terms of oil recovery, while a 33% increase in droplet size, which could be achieved by the use of chemicals, increased separator efficiency by only 55%. The reduction in velocity increases the residence time of the oil in the separator therefore allowing more droplets to rise out of the effluent water stream before it reaches the exit. This residence time approach has been discarded by the industry, the argument being that increasing the residence time requires a longer processing time. However the cost benefits of this approach are twofold:

- It avoids the additional operating cost of buying chemicals (coagulants and flocculants) since reduction of horizontal velocity can be brought about by addition of baffles.

- It avoids the additional capital cost of the extra treatment units required to treat the recovered oil, effluent water, and sludge when chemicals are employed.

In the integrated process of Corrugated Plate interceptor and ICB, the results suggest that increasing the number of plates is the most cost effective method of improving oil recovery in plate interceptors. For a 25% increase in number of plates, interceptor efficiency increased by 23% while for a 25% increase in plate dimensions (length or width) interceptor efficiency increased by 22% or 21% respectively. These figures are close. The effect of increasing these design parameters on the interceptor volume was used as a deciding factor on choosing the most cost-effective solution to improving oil recovery. Increasing the plate length by 25% resulted in a 43% increase in interceptor volume; increasing plate width by 25% resulted in a 17% increase in interceptor volume, whilst increasing the number of plate by 25% resulted in only a 2% increase in interceptor volume. Increasing interceptor volume suggests increase in cost of construction and hence increased cost of separation.

The mass transfer characteristics in a laboratory scale ICB were studied for various different support particles. The 38 mm plastic Pall ring support particles were observed to have the best mass transfer characteristics when compared with the other sizes of 16 mm, 25 mm, and 50 mm. Though the smaller support particles were expected to produce better mass transfer characteristics, this was not observed. The reasons for this may be due to the arrangement of the support particles in the ICB unit and the internal structure of the support particles. However, these results show that optimization of the design plays a very important role in improving the cost of aeration.

Analysis of the various cost relationships in each of the integrated processes also revealed the following cost effective windows of operation in oily wastewater treatment or separation:

- It is more cost-effective to operate only an Immobilized Cell Bioreactor (ICB) when the average oil droplet size is small (less than 60 microns).
- It is more cost-effective to operate only an Immobilized Cell Bioreactor when only low oil concentrations of oil are present in the wastewater stream.

In conclusion, the solution to improving the cost effectiveness of integrated processes may lie in the design of the units rather than manipulation of influent oil characteristics as currently practiced in the industry.

## 6.2 Recommendations

The design and performance of the separators investigated in this research was based solely on theoretical models and considerations. The results obtained may not actually reflect what happens empirically. Therefore it is recommended that empirical studies be carried out to further investigate the claims of this study.

Also, in this research an actual oily wastewater stream was not characterized and used. Complex interactions occur between oil droplets, which may improve separator performance. This study did not consider interactions between droplets: the principle of unhindered settling was used in analyzing separator performance. A study which takes oil droplet interaction into consideration would present better results.

Another factor which could affect plate interceptor performance is the location of the plates in the interceptors. This was not studied and could have a considerable effect on separator efficiency and cost of separation. It is recommended further studies be carried out to investigate the effect of plate pack location on interceptor performance.

In the costing of the separation process, the cost of supporting equipment for the oil recovery units (such as oil skimmers) was not taken into consideration. These units are necessary to ensure that separators function properly; to make better judgements, the cost of these units should be taken into consideration.

In the study of aeration in ICB units, only a particular size of support particle was used in each design of ICB. Possibly, a mixture of various sizes of support particles may have resulted in better mass transfer characteristics but this was not investigated. It is recommended further work be done in this area. Also the effect of using support particles of various designs on aeration should be investigated. This research did not confirm the effect of arrangement of support particle on aeration. A random arrangement was used in the experimental studies; the effect of structured packing and random packing could be compared to investigate which improves aeration and correlations developed to aid in decision making. A mixing experiment should have been carried out to investigate the effect of support particle arrangement on liquid circulation rate. Also, in the aeration investigations, oily wastewater was not used as the medium. It is recommended that oily wastewater, with microbes attached to the support particles, be used for further studies.

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# Appendix I Design and cost data for oil-water separators

## 1 Factors used in the design of a horizontal flow separator

The product of the short-circuiting factor $F_s$ and turbulent factor $F_t$ yields the design factor $F$. A curve can be plotted to obtain the factor $F$ from Table 1.

| $v_h$ / $v_t$ | $F$ |
|---|---|
| 20 | 1.45 |
| 15 | 1.37 |
| 10 | 1.27 |
| 6 | 1.14 |
| 3 | 1.07 |

**Table 1: Typical values of design factor ($F$)**

## 2 Plate Separators

There are four basic types of plate separator. These are inclined plate, tilted plate, cross-flow, and matrix type plate separators.

### 2.1 Inclined Plate Separator

The plates within this separator are flat, parallel, and inclined at 45 degrees. This unit is capable of removing droplets of diameter 60 microns and larger. The majority of entrained solids fall out at the inlet of inclined plate separators; oily water flows though the plate packs in laminar flow conditions. Oil rises and adheres to the underside of the plates. Separated oil flows as a film up the plates, disengages from the plates and rise to the liquid surface in large drops where it is skimmed off; clean water underflows an oil retention baffle and leaves the unit.

### 2.2 Corrugated Plate Interceptor

The plates in this unit are corrugated. The corrugations in the plates do not increase the effective plate area but bring about two benefits to the separator. Firstly, they serve to segregate the oil from the solids being separated and enhance coalescence/agglomeration of each of the liquid phases. Secondly, the corrugations add rigidity to the plate assembly. As with the inclined plate separator, oil droplets rise to the underside of the plates. This unit is capable of removing droplets of diameter 60 microns and larger.

### 2.3 Cross-flow Separators

Cross-flow separators may have corrugated plates; the major difference between the corrugated plate separator and the crossflow separator is the water flow direction. In order to minimize separator size while maintaining throughput, the water flow is arranged across rather than along the corrugations. The oil and sludge flow at 90° to the water: oil flows upwards to the underside of the plates and sludge downwards to the upper surface of the plates. This unit is capable of removing droplets of diameter 60 microns and larger.

### 2.4 Matrix Plate Separators

The plates in this separator are located in a horizontal cylindrical vessel. The open design of the plates maintains a large contact area for oil separation whilst minimizing the possibility of oil and solids mutual attachment. This unit is also capable of removing droplets of diameter 60 microns and larger.

## 3 Analysis of droplet capture in a Corrugated Plate Interceptor

The actual path travelled by an oil droplet takes the form of a curve (EF) but Yao 1973 approximated this path to a straight-line (BD) (Figure 1).

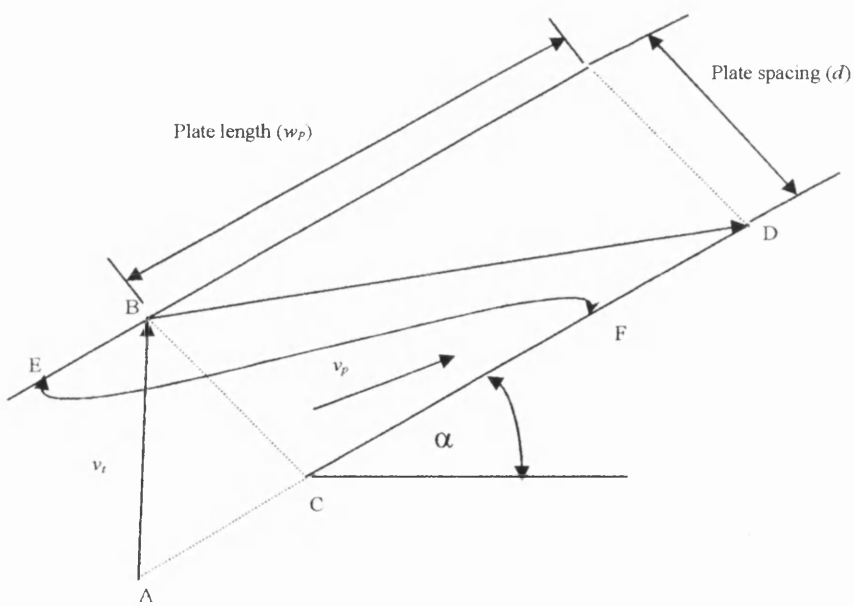



**Figure 1: Theory of plate settlers**

The path of the droplet is complex because of the effect of parabolic velocity gradient. However the straight line BD simplifies the path of the particle with the slowest rise rate.

Inspecting Figure 1,

$$v_t / v_p = AB/AD \qquad (1.0)$$

hence,

$$v_t = v_p * AB/AD \qquad (1.0a)$$

but,

$$AB = d / \cos \alpha \qquad (1.1)$$

and,

$$AD = w_P + d / \tan \alpha \qquad (1.2)$$

therefore the rise rate of the critical droplet is given by:

$$v_t = v_p * d / \cos \alpha * 1/ (w_P + d / \tan \alpha) \qquad (1.3a)$$

simplifying,

$$v_t = v_p * d \,(d \sin \alpha + w_P \cos \alpha) \qquad (1.3b)$$

## 4 Cost Factors

Purchase cost of equipment, cost factors used in Equation 2.34 (Coulson J.M. *et al* 1991).

| Equipment | Size unit (S) | Size range | Constant C(£) | Index (n) | Comment |
|---|---|---|---|---|---|
| Plate | Area $m^2$ | 5-50 | 1000 | 0.60 | Carbon steel |
| Tanks<br>Process<br>Vertical<br>Horizontal | Capacity $m^3$ | 1-50<br>10-100 | 500<br>600 | 0.59<br>0.60 | Atmos. press.<br>Carbon steel |

**Table 2: Cost Factors**

# Appendix II Experimental Results

The calibration data obtained for the rotameter was plotted to obtain a calibration curve (see Figure 1).

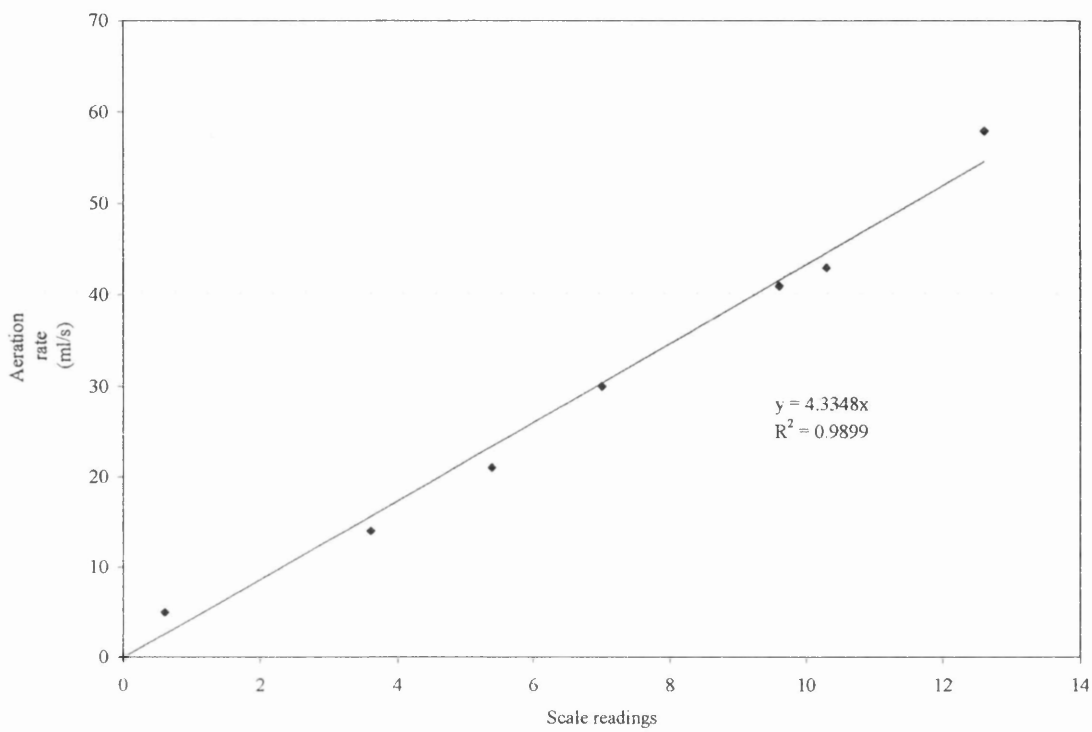



**Figure 1: Calibration curve for the rotameter**

Calculated values of mass transfer coefficient versus aeration rate are presented in Tables 1 to 4 below.

| Aeration rate (ml/s) | Mass transfer coefficient ($s^{-1}$) |
|---|---|
| 48.0 | 0.0039 |
| 38.0 | 0.0035 |
| 24.0 | 0.0024 |
| 9.0 | 0.0020 |

**Table 1: Mass transfer coefficient vs. aeration rate for 16 mm support particle**

| Aeration rate (ml/s) | Mass transfer coefficient ($s^{-1}$) |
|---|---|
| 47.0 | 0.0097 |
| 41.6 | 0.0086 |
| 28.0 | 0.0055 |
| 17.5 | 0.0047 |
| 9.0 | 0.0030 |

**Table 2: Mass transfer coefficient vs. aeration rate for 38 mm support particle**

| Aeration rate (ml/s) | Mass transfer coefficient ($s^{-1}$) |
|---|---|
| 56.0 | 0.0043 |
| 43.6 | 0.0042 |
| 31.1 | 0.0033 |
| 27.0 | 0.0029 |
| 13.0 | 0.0020 |

**Table 3: Mass transfer coefficient vs. aeration rate for 25 mm support particle**

| Aeration rate (ml/s) | Mass transfer coefficient ($s^{-1}$) |
|---|---|
| 50.0 | 0.0058 |
| 40.6 | 0.0035 |
| 30.0 | 0.0021 |
| 15.0 | 0.0020 |
| 10.5 | 0.0015 |

**Table 4: Mass transfer coefficient vs. aeration rate for 50 mm support particle**

# Appendix III Theoretical results: Integration of a horizontal flow separator and an ICB

Input data and results for investigations of integrated processes using a horizontal flow separator and an Immobilized Cell Bioreactor.

Horizontal flow separator performance prediction spreadsheet

Effect of oil concentration
Effect of horizontal flow velocity
Effect of separator volume
Effect of average droplet size
Effect of viscosity
Effect of density difference

Analysing Stokes' Law

# Horizontal flow separator performance prediction spreadsheet

**Influent characteristics**

| | | |
|---|---|---|
| Water flowrate | **10** | $m^3/h$ |
| Water density | **1000** | $kg/m^3$ |
| Water viscosity | **0.002** | Pa.s |
| Oil concentration | **1000** | mg/l |
| Average oil density | **800** | $kg/m^3$ |

**Separator dimensions**

| | | |
|---|---|---|
| Length | **2.3248** | m |
| Width | **1.8288** | m |
| Height | **0.9144** | m |

**Calculations**

| | | |
|---|---|---|
| Mass of oil in influent | 10 | kg/h |
| Volume of oil in influent | 0.0125 | $m^3/h$ |

| | | |
|---|---|---|
| volume | 3.8876085 | $m^3$ |
| SA | 4.2515404 | $m^2$ |
| XSA | 1.6722547 | $m^2$ cross-sectional area |
| horiz. vel. | 0.004905 | m/s (0.97 ft/min) |
| res. time | 474 | s (7.9 min) |

| | | Influent | | | | Recovered oil | | Water effluent | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Droplet size interval | Mean droplet size (μm) | Number of droplets in interval | Volume of droplets in interval ($m^3$) | Terminal velocity (m/s) | Fraction of droplets removed | Number of droplets removed | Volume of droplets removed ($m^3$) | Number of droplets remaining | Volume of droplets remaining ($m^3$) | Cumulative volume of oil in each interval ($m^3$) | Percentage of oil contributed by each interval | Cumulative percentage of oil in each interval |
| 1 | 5 | 3.970E+07 | 2.598E-09 | 1.4E-06 | 0.0007 | 2.804E+04 | 1.835E-12 | 3.967E+07 | 2.596E-09 | 2.598E-09 | 0.00002 | 0.00002 |
| 2 | 15 | 3.147E+08 | 5.561E-07 | 1.2E-05 | 0.0064 | 2.000E+06 | 3.535E-09 | 3.127E+08 | 5.526E-07 | 5.587E-07 | 0.00445 | 0.00447 |
| 3 | 25 | 1.751E+09 | 1.432E-05 | 3.4E-05 | 0.0177 | 3.091E+07 | 2.529E-07 | 1.720E+09 | 1.407E-05 | 1.488E-05 | 0.11458 | 0.11905 |
| 4 | 35 | 6.334E+09 | 1.422E-04 | 6.7E-05 | 0.0346 | 2.192E+08 | 4.921E-06 | 6.115E+09 | 1.373E-04 | 1.571E-04 | 1.13770 | 1.25675 |
| 5 | 45 | 1.492E+10 | 7.120E-04 | 1.1E-04 | 0.0572 | 8.536E+08 | 4.073E-05 | 1.407E+10 | 6.713E-04 | 8.691E-04 | 5.69642 | 6.95317 |
| 6 | 55 | 2.290E+10 | 1.995E-03 | 1.6E-04 | 0.0855 | 1.957E+09 | 1.705E-04 | 2.094E+10 | 1.824E-03 | 2.864E-03 | 15.96087 | 22.91404 |
| 7 | 65 | 2.290E+10 | 3.293E-03 | 2.3E-04 | 0.1194 | 2.733E+09 | 3.930E-04 | 2.017E+10 | 2.900E-03 | 6.157E-03 | 26.34563 | 49.25968 |
| 8 | 75 | 1.492E+10 | 3.296E-03 | 3.1E-04 | 0.1589 | 2.371E+09 | 5.238E-04 | 1.255E+10 | 2.772E-03 | 9.453E-03 | 26.37231 | 75.63198 |
| 9 | 85 | 6.334E+09 | 2.037E-03 | 3.9E-04 | 0.2041 | 1.293E+09 | 4.157E-04 | 5.042E+09 | 1.621E-03 | 1.149E-02 | 16.29599 | 91.92798 |
| 10 | 95 | 1.751E+09 | 7.859E-04 | 4.9E-04 | 0.2549 | 4.463E+08 | 2.004E-04 | 1.304E+09 | 5.855E-04 | 1.228E-02 | 6.28733 | 98.21531 |
| 11 | 105 | 3.147E+08 | 1.908E-04 | 6.0E-04 | 0.3114 | 9.801E+07 | 5.941E-05 | 2.167E+08 | 1.313E-04 | 1.247E-02 | 1.52614 | 99.74146 |
| 12 | 115 | 3.677E+07 | 2.928E-05 | 7.2E-04 | 0.3736 | 1.374E+07 | 1.094E-05 | 2.303E+07 | 1.834E-05 | 1.250E-02 | 0.23426 | 99.97571 |
| 13 | 125 | 2.789E+06 | 2.852E-06 | 8.5E-04 | 0.4414 | 1.231E+06 | 1.259E-06 | 1.558E+06 | 1.593E-06 | 1.250E-02 | 0.02282 | 99.99853 |
| 14 | 135 | 1.372E+05 | 1.767E-07 | 9.9E-04 | 0.5148 | 7.061E+04 | 9.097E-08 | 6.654E+04 | 8.573E-08 | 1.250E-02 | 0.00141 | 99.99994 |
| 15 | 145 | 4.370E+03 | 6.975E-09 | 1.1E-03 | 0.5939 | 2.595E+03 | 4.143E-09 | 1.774E+03 | 2.832E-09 | 1.250E-02 | 0.00006 | 100.00000 |
| 16 | 155 | 9.007E+01 | 1.756E-10 | 1.3E-03 | 0.6787 | 6.113E+01 | 1.192E-10 | 2.894E+01 | 5.643E-11 | 1.250E-02 | 0.00000 | 100.00000 |
| 17 | 165 | 1.200E+00 | 2.823E-12 | 1.5E-03 | 0.7691 | 9.230E-01 | 2.171E-12 | 2.772E-01 | 6.519E-13 | 1.250E-02 | 0.00000 | 100.00000 |
| 18 | 175 | 1.032E-02 | 2.897E-14 | 1.7E-03 | 0.8651 | 8.931E-03 | 2.506E-14 | 1.392E-03 | 3.907E-15 | 1.250E-02 | 0.00000 | 100.00000 |
| 19 | 185 | 0.000E+00 | 0.000E+00 | 1.9E-03 | 0.9668 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 20 | 195 | 0.000E+00 | 0.000E+00 | 2.1E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 21 | 205 | 0.000E+00 | 0.000E+00 | 2.3E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 22 | 215 | 0.000E+00 | 0.000E+00 | 2.5E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 23 | 225 | 0.000E+00 | 0.000E+00 | 2.8E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 24 | 235 | 0.000E+00 | 0.000E+00 | 3.0E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 25 | 245 | 0.000E+00 | 0.000E+00 | 3.3E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 26 | 255 | 0.000E+00 | 0.000E+00 | 3.5E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| 27 | 265 | 0.000E+00 | 0.000E+00 | 3.8E-03 | 1.0000 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 0.000E+00 | 1.250E-02 | 0.00000 | 100.00000 |
| Totals | | 9.252E+10 | 1.250E-02 | | | 1.002E+10 | 1.821E-03 | 8.251E+10 | 1.068E-02 | | 100.00000 | |

## Effect of oil concentration on the integrated process of horizontal flow separator and ICB: Input parameters

| Influent characteristics | | | Separator dimensions | | |
|---|---|---|---|---|---|
| Water flowrate | 10 | $m^3/h$ | Width | 1.8288 | m |
| Water density | 1000 | $kg/m^3$ | Height | 0.9144 | m |
| Water viscosity | 0.001 | Pa.s | Length | 2.3248 | m |
| | | | | | |
| Oil concentration | **VARIABLE** | mg/l | | | |
| Average oil density | 800 | $kg/m^3$ | | | |
| Average droplet size | 60 | μm | | | |
| Standard deviation | 15 | μm | | | |

| Basis of costing | | |
|---|---|---|
| Cost of oil | 0.62 | £/$m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of oil concentration on the integrated process of horizontal flow separator and ICB: Results

| Influent oil concentration (mg/l) | Oil flow to separator ($m^3/h$) | Separator volume ($m^3$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| **1000** | 0.01250 | 3.888 | 7.9 | 0.0036 | 709 | 29.1 | 0.00886 | 2.46 | 113.1 |
| **1200** | 0.01500 | 3.888 | 7.9 | 0.0044 | 850 | 29.1 | 0.01063 | 2.95 | 135.7 |
| **1300** | 0.01625 | 3.888 | 7.9 | 0.0047 | 921 | 29.1 | 0.01151 | 3.20 | 146.9 |
| **1400** | 0.01750 | 3.888 | 7.9 | 0.0051 | 992 | 29.1 | 0.01239 | 3.44 | 158.2 |
| **1500** | 0.01875 | 3.888 | 7.9 | 0.0055 | 1063 | 29.1 | 0.01328 | 3.69 | 169.5 |

| Influent oil concentration (mg/l) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| **1000** | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| **1200** | 44.56 | 10.84 | 33.72 | 632.49 | 5509.79 | 6142.28 | 6176.00 |
| **1300** | 44.56 | 11.74 | 32.82 | 666.20 | 5967.95 | 6634.15 | 6666.97 |
| **1400** | 44.56 | 12.65 | 31.91 | 699.07 | 6426.97 | 7126.04 | 7157.96 |
| **1500** | 44.56 | 13.55 | 31.01 | 731.16 | 6886.39 | 7617.55 | 7648.56 |

## Effect of horizontal flow velocity on the integrated process of horizontal flow separator and ICB: Input parameters

### Influent characteristics

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | 0.001 | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | 60 | μm |
| Standard deviation | 15 | μm |

### Separator dimensions

| | |
|---|---|
| Width | 1.8288 m |
| Height | 0.9144 m |
| Length | 2.3248 m |

### Basis of costing

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of horizontal flow velocity on the integrated process of horizontal flow separator and ICB: Results

| Horizontal velocity (m/s) | Oil flow to separator ($m^3/h$) | Separator volume ($m^3$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| **0.0049050** | 0.0125 | 3.888 | 7.9 | 0.0036 | 709 | 29.1 | 0.00886 | 2.46 | 113.1 |
| **0.0024525** | 0.0125 | 3.888 | 15.8 | 0.0072 | 424 | 57.6 | 0.00529 | 1.47 | 67.6 |
| **0.0019620** | 0.0125 | 3.888 | 19.75 | 0.0087 | 303 | 69.7 | 0.00378 | 1.05 | 48.3 |
| **0.0014715** | 0.0125 | 3.888 | 26.33 | 0.0104 | 168 | 83.3 | 0.00209 | 0.58 | 26.7 |
| **0.0009810** | 0.0125 | 3.888 | 39.5 | 0.0118 | 56 | 94.4 | 0.00070 | 0.19 | 8.9 |
| **0.0004905** | 0.0125 | 3.888 | 78.99 | 0.0124 | 4 | 99.6 | 0.00005 | 0.01 | 0.7 |

| Horizontal velocity (m/s) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| **0.0049050** | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| **0.0024525** | 44.56 | 17.87 | 26.69 | 300.62 | 2743.91 | 3044.52 | 3071.21 |
| **0.0019620** | 44.56 | 21.61 | 22.95 | 197.09 | 1962.52 | 2159.61 | 2182.56 |
| **0.0014715** | 44.56 | 25.81 | 18.75 | 138.93 | 1085.04 | 1223.98 | 1242.73 |
| **0.0009810** | 44.56 | 29.27 | 15.29 | 72.79 | 362.73 | 435.51 | 450.81 |
| **0.0004905** | 44.56 | 30.87 | 13.69 | 15.97 | 27.73 | 43.70 | 57.39 |

Effect of separator volume on the integrated process of horizontal flow separator and ICB: Input parameters

| Influent characteristics | | Separator dimensions | |
|---|---|---|---|
| Water flowrate | 10 $m^3/h$ | Width | 1.8288 m |
| Water density | 1000 $kg/m^3$ | Height | 0.9144 m |
| Water viscosity | 0.001 Pa.s | Length | **VARIABLE** m |
| | | | |
| Oil concentration | 1000 mg/l | | |
| Average oil density | 800 $kg/m^3$ | | |
| Average droplet size | 60 µm | | |
| Standard deviation | 15 µm | | |

Basis of costing

| | |
|---|---|
| Cost of oil | 0.62 £/$m^3$ |
| Equipment cost model | $CS^n$ |
| Daily hours of operation | 16 h/day |
| Annual days of operation | 250 days/yr |
| Period of operation | 25 yr |

Effect of separator volume on the integrated process of horizontal flow separator and ICB: Results

| Separator volume ($m^3$) | Separator length (m) | Oil flow to separator ($m^3/h$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| **3.888** | 2.3248 | 0.0125 | 7.9 | 0.0036 | 709 | 29.1 | 0.00886 | 2.46 | 113.1 |
| **7.851** | 4.6950 | 0.0125 | 16.0 | 0.0073 | 419 | 58.1 | 0.00523 | 1.45 | 66.8 |
| **11.663** | 6.9743 | 0.0125 | 23.7 | 0.0098 | 212 | 78.8 | 0.00265 | 0.74 | 33.9 |
| **15.550** | 9.2991 | 0.0125 | 31.6 | 0.0112 | 103 | 89.8 | 0.00128 | 0.36 | 16.3 |
| **19.438** | 11.6239 | 0.0125 | 39.5 | 0.0118 | 56 | 94.4 | 0.00070 | 0.19 | 8.9 |
| **23.326** | 13.9486 | 0.0125 | 47.4 | 0.0122 | 25 | 97.5 | 0.00032 | 0.09 | 4.1 |
| **27.213** | 16.2734 | 0.0125 | 55.3 | 0.0123 | 18 | 98.2 | 0.00023 | 0.06 | 2.9 |
| **31.101** | 18.5982 | 0.0125 | 63.2 | 0.0124 | 11 | 98.9 | 0.00013 | 0.04 | 1.7 |
| **34.988** | 20.9229 | 0.0125 | 71.1 | 0.0124 | 5 | 99.5 | 0.00006 | 0.02 | 0.8 |
| **38.876** | 23.2477 | 0.0125 | 79.0 | 0.0124 | 4 | 99.6 | 0.00005 | 0.01 | 0.7 |
| **46.652** | 27.8974 | 0.0125 | 94.8 | 0.0125 | 3 | 99.7 | 0.00003 | 0.01 | 0.4 |

Effect of separator volume on the integrated process of horizontal flow separator and ICB: Results

| Separator volume ($m^3$) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| **3.888** | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| **7.851** | 67.46 | 18.02 | 49.44 | 298.55 | 2712.53 | 3011.08 | 3060.52 |
| **11.663** | 85.20 | 24.42 | 60.78 | 159.78 | 1375.18 | 1534.96 | 1595.75 |
| **15.550** | 100.96 | 27.82 | 73.14 | 103.99 | 664.02 | 768.01 | 841.15 |
| **19.438** | 115.17 | 29.27 | 85.90 | 72.79 | 362.73 | 435.51 | 521.42 |
| **23.326** | 128.25 | 30.21 | 98.04 | 45.70 | 164.83 | 210.53 | 308.57 |
| **27.213** | 140.46 | 30.44 | 110.02 | 37.39 | 117.27 | 154.65 | 264.67 |
| **31.101** | 151.97 | 30.67 | 121.30 | 27.51 | 69.71 | 97.22 | 218.52 |
| **34.988** | 162.91 | 30.84 | 132.07 | 17.72 | 33.09 | 50.82 | 182.88 |
| **38.876** | 173.36 | 30.87 | 142.49 | 15.97 | 27.73 | 43.70 | 186.19 |
| **46.652** | 193.05 | 30.92 | 162.13 | 11.97 | 17.00 | 28.97 | 191.09 |

Effect of average droplet size on the integrated process of horizontal flow separator and ICB: Input parameters

**Influent characteristics**

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | 0.001 | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | **VARIABLE** | µm |
| Standard deviation | 15 | µm |

**Separator dimensions**

| | | |
|---|---|---|
| Width | 1.8288 | m |
| Height | 0.9144 | m |
| Length | 2.3248 | m |

**Basis of costing**

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of average droplet size on the integrated process of horizontal flow separator and ICB: Results

| Average droplet size (μm) | Oil flow to separator ($m^3/h$) | Separator volume ($m^3$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| **60** | 0.0125 | 3.888 | 7.9 | 0.0036 | 709 | 29.1 | 0.00886 | 2.46 | 113.1 |
| **70** | 0.0125 | 3.888 | 7.9 | 0.0046 | 634 | 36.6 | 0.00792 | 2.20 | 101.2 |
| **80** | 0.0125 | 3.888 | 7.9 | 0.0056 | 549 | 45.1 | 0.00686 | 1.90 | 87.5 |
| **90** | 0.0125 | 3.888 | 7.9 | 0.0068 | 453 | 54.7 | 0.00566 | 1.57 | 72.2 |
| **100** | 0.0125 | 3.888 | 7.9 | 0.0081 | 348 | 65.2 | 0.00435 | 1.21 | 55.5 |
| **150** | 0.0125 | 3.888 | 7.9 | 0.0124 | 7 | 99.3 | 0.00009 | 0.02 | 1.1 |

| Average droplet size (μm) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| **60** | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| **70** | 44.56 | 11.34 | 33.22 | 522.65 | 4108.49 | 4631.14 | 4664.36 |
| **80** | 44.56 | 13.99 | 30.57 | 351.16 | 3555.08 | 3906.23 | 3936.81 |
| **90** | 44.56 | 16.96 | 27.60 | 312.88 | 2933.02 | 3245.90 | 3273.50 |
| **100** | 44.56 | 20.21 | 24.35 | 267.20 | 2254.62 | 2521.82 | 2546.17 |
| **150** | 44.56 | 30.78 | 13.78 | 21.46 | 45.78 | 67.25 | 81.03 |

## Effect of viscosity on the integrated process of horizontal flow separator and ICB: Input parameters

**Influent characteristics**

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | **VARIABLE** | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | 60 | μm |
| Standard deviation | 15 | μm |

**Separator dimensions**

| | | |
|---|---|---|
| Width | 1.8288 | m |
| Height | 0.9144 | m |
| Length | 2.3248 | m |

**Basis of costing**

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of viscosity on the integrated process of horizontal flow separator and ICB: Results

| Water viscosity (Pa.s) | Oil flow to separator ($m^3/h$) | Separator volume ($m^3$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| 0.00100 | 0.0125 | 3.888 | 7.9 | 0.0036 | 709 | 29.1 | 0.00886 | 2.46 | 113.1 |
| 0.00065 | 0.0125 | 3.888 | 7.9 | 0.0056 | 554 | 44.6 | 0.00692 | 1.92 | 88.4 |
| 0.00047 | 0.0125 | 3.888 | 7.9 | 0.0076 | 389 | 61.1 | 0.00485 | 1.35 | 62.0 |
| 0.00036 | 0.0125 | 3.888 | 7.9 | 0.0094 | 246 | 75.5 | 0.00307 | 0.85 | 39.2 |
| 0.00028 | 0.0125 | 3.888 | 7.9 | 0.0107 | 147 | 85.3 | 0.00183 | 0.51 | 23.4 |

| Water viscosity (Pa.s) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| 0.00100 | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| 0.00065 | 44.56 | 13.82 | 30.74 | 353.26 | 3590.63 | 3943.89 | 3974.63 |
| 0.00047 | 44.56 | 18.96 | 25.60 | 285.44 | 2516.91 | 2802.34 | 2827.95 |
| 0.00036 | 44.56 | 23.39 | 21.17 | 174.08 | 1590.25 | 1764.33 | 1785.50 |
| 0.00028 | 44.56 | 26.45 | 18.11 | 128.51 | 950.67 | 1079.17 | 1097.28 |

## Effect of density difference on the integrated process of horizontal flow separator and ICB: Input parameters

**Influent characteristics**

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | 0.001 | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | 60 | μm |
| Standard deviation | 15 | μm |

**Separator dimensions**

| | | |
|---|---|---|
| Width | 1.8288 | m |
| Height | 0.9144 | m |
| Length | 2.3248 | m |

**Basis of costing**

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of density difference on the integrated process of horizontal flow separator and ICB: Results

| Density difference ($kg/m^3$) | Oil flow to separator ($m^3/h$) | Separator volume ($m^3$) | Residence time (min) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Separator efficiency (%) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) |
|---|---|---|---|---|---|---|---|---|---|
| 200 | 0.0125 | 3.888 | 7.9 | 0.0036 | 709 | 29 | 0.00886 | 2.46 | 113.1 |
| 225 | 0.0125 | 3.888 | 7.9 | 0.0041 | 672 | 33 | 0.00840 | 2.33 | 107.2 |
| 235 | 0.0125 | 3.888 | 7.9 | 0.0043 | 658 | 34 | 0.00822 | 2.28 | 104.9 |

| Density difference ($kg/m^3$) | Cost of separator (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|
| 200 | 44.56 | 9.03 | 35.53 | 561.84 | 4591.89 | 5153.72 | 5189.25 |
| 225 | 44.56 | 10.16 | 34.40 | 542.88 | 4355.69 | 4898.57 | 4932.96 |
| 235 | 44.56 | 10.61 | 33.95 | 535.20 | 4261.27 | 4796.47 | 4830.42 |

## Analysing Stokes' law

$v = d^2 (\rho_w - \rho_o)\, g / 18\mu$

| | |
|---|---|
| $v$ = | droplet rise velocity (m/s) |
| $d$ = | particle diameter (μm) |
| $g$ = | acceleration due to gravity (m/s²) |
| $\rho_w$ = | density of water (kg/m³) |
| $\rho_o$ = | density of oil (kg/m³) |
| $\mu$ = | viscosity of water (Pa.s) |

## Effect of increasing droplet size on rise velocity

| Droplet size (μm) | Increase in droplet size (%) | Droplet rise velocity (m/s) | Increase in rise velocity (%) |
|---|---|---|---|
| 20 | 0 | 4.360E-06 | 0 |
| 22 | 10 | 5.276E-06 | 21 |
| 24 | 20 | 6.278E-06 | 44 |
| 26 | 30 | 7.368E-06 | 69 |
| 28 | 40 | 8.546E-06 | 96 |
| 30 | 50 | 9.810E-06 | 125 |
| 32 | 60 | 1.116E-05 | 156 |
| 34 | 70 | 1.260E-05 | 189 |
| 36 | 80 | 1.413E-05 | 224 |
| 38 | 90 | 1.574E-05 | 261 |
| 40 | 100 | 1.744E-05 | 300 |
| 42 | 110 | 1.923E-05 | 341 |
| 44 | 120 | 2.110E-05 | 384 |

Effect of increasing density difference on droplet rise velocity

| Density difference (kg/m³) | Increase in density difference (%) | Droplet rise velocity (m/s) | Increase in rise velocity (%) |
|---|---|---|---|
| 200 | 0 | 3.924E-05 | 0 |
| 220 | 10 | 4.316E-05 | 10 |
| 230 | 15 | 4.513E-05 | 15 |
| 247 | 23 | 4.840E-05 | 23 |
| 262 | 31 | 5.134E-05 | 31 |
| 277 | 38 | 5.428E-05 | 38 |
| 292 | 46 | 5.723E-05 | 46 |
| 307 | 53 | 6.017E-05 | 53 |
| 322 | 61 | 6.311E-05 | 61 |
| 337 | 68 | 6.605E-05 | 68 |
| 352 | 76 | 6.900E-05 | 76 |
| 367 | 83 | 7.194E-05 | 83 |
| 382 | 91 | 7.488E-05 | 91 |

Effect of decreasing viscosity on droplet rise velocity

| Water viscosity (cP) | Decrease in water viscosity (%) | Droplet rise velocity (m/s) | Increase in rise velocity (%) |
|---|---|---|---|
| 0.13 | 0 | 3.018E-06 | 0 |
| 0.12 | 8 | 3.270E-06 | 8 |
| 0.11 | 15 | 3.567E-06 | 18 |
| 0.1 | 23 | 3.924E-06 | 30 |
| 0.09 | 31 | 4.360E-06 | 44 |
| 0.08 | 38 | 4.905E-06 | 63 |
| 0.07 | 46 | 5.606E-06 | 86 |
| 0.06 | 54 | 6.540E-06 | 117 |
| 0.05 | 62 | 7.848E-06 | 160 |
| 0.04 | 69 | 9.810E-06 | 225 |
| 0.03 | 77 | 1.308E-05 | 333 |
| 0.02 | 85 | 1.962E-05 | 550 |
| 0.01 | 92 | 3.924E-05 | 1200 |

# Appendix IV Theoretical results: Integration of a CPI and an ICB

Input data and results for investigations of integrated processes using a Corrugated Plate Interceptor and an Immobilized Cell Bioreactor.

Influent oil characteristics for the CPI

Effect of plate width
Effect of plate angle
Effect of plate spacing
Effect of plate length
Effect of number of plates

Influent oil characteristics for the CPI

## Oily Wastewater oil droplet size distribution

Influent characteristics

Water flowrate **10** $m^3/h$

Oil concentration **1000** mg/l
Average oil density **800** $kg/m^3$

Average droplet diameter **60** µm
Standard deviation **15** µm

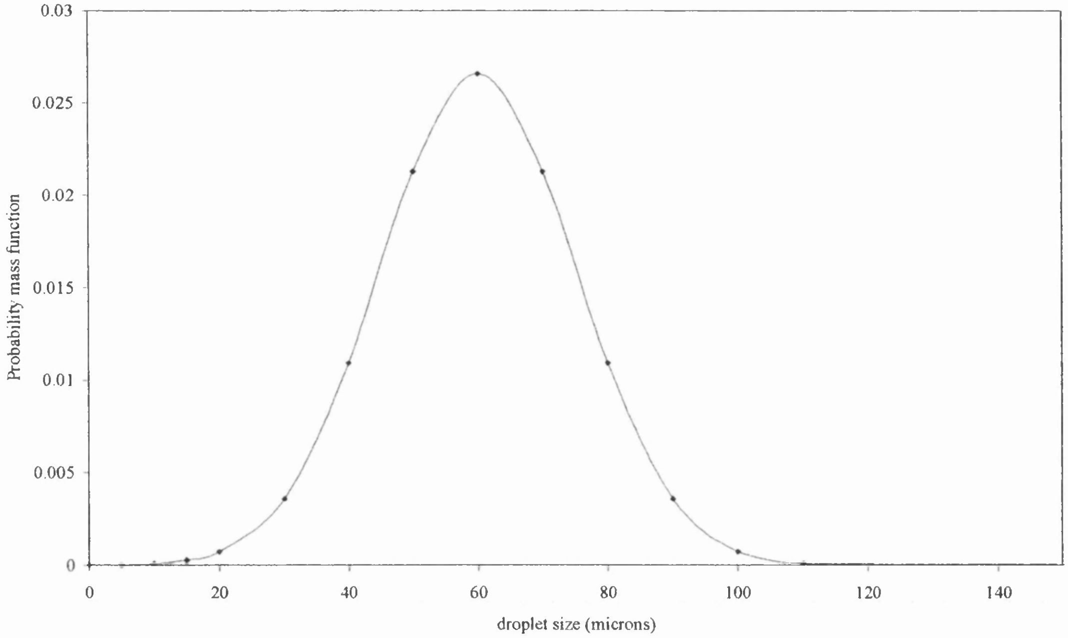
Calculations

Mass of oil in influent 10 kg/h
Volume of oil in influent 0.0125 $m^3/h$

No. of droplets in influent 9.25E+10 - Solver variable

| Droplet size interval | Minimum droplet size µm | Maximum droplet size µm | Mean droplet size µm | Cumulative distribution function | Fraction of droplets in interval | Number of droplets in interval | Volume of droplets in interval $m^3$ |
|---|---|---|---|---|---|---|---|
| 1 | 0 | 10 | 5 | 0.00043 | 0.00043 | 3.970E+07 | 2.6E-09 |
| 2 | 10 | 20 | 15 | 0.00383 | 0.00340 | 3.147E+08 | 5.56E-07 |
| 3 | 20 | 30 | 25 | 0.02275 | 0.01892 | 1.751E+09 | 1.43E-05 |
| 4 | 30 | 40 | 35 | 0.09121 | 0.06846 | 6.334E+09 | 0.000142 |
| 5 | 40 | 50 | 45 | 0.25249 | 0.16128 | 1.492E+10 | 0.000712 |
| 6 | 50 | 60 | 55 | 0.50000 | 0.24751 | 2.290E+10 | 0.001995 |
| 7 | 60 | 70 | 65 | 0.74751 | 0.24751 | 2.290E+10 | 0.003293 |
| 8 | 70 | 80 | 75 | 0.90879 | 0.16128 | 1.492E+10 | 0.003296 |
| 9 | 80 | 90 | 85 | 0.97725 | 0.06846 | 6.334E+09 | 0.002037 |
| 10 | 90 | 100 | 95 | 0.99617 | 0.01892 | 1.751E+09 | 0.000786 |
| 11 | 100 | 110 | 105 | 0.99957 | 0.00340 | 3.147E+08 | 0.000191 |
| 12 | 110 | 120 | 115 | 0.99997 | 0.00040 | 3.677E+07 | 2.93E-05 |
| 13 | 120 | 130 | 125 | 1.00000 | 0.00003 | 2.789E+06 | 2.85E-06 |
| 14 | 130 | 140 | 135 | 1.00000 | 0.00000 | 1.372E+05 | 1.77E-07 |
| 15 | 140 | 150 | 145 | 1.00000 | 0.00000 | 4.370E+03 | 6.98E-09 |
| 16 | 150 | 160 | 155 | 1.00000 | 0.00000 | 9.007E+01 | 1.76E-10 |
| 17 | 160 | 170 | 165 | 1.00000 | 0.00000 | 1.200E+00 | 2.82E-12 |
| 18 | 170 | 180 | 175 | 1.00000 | 0.00000 | 1.032E-02 | 2.9E-14 |
| 19 | 180 | 190 | 185 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 20 | 190 | 200 | 195 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 21 | 200 | 210 | 205 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 22 | 210 | 220 | 215 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 23 | 220 | 230 | 225 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 24 | 230 | 240 | 235 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 25 | 240 | 250 | 245 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 26 | 250 | 260 | 255 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| 27 | 260 | 270 | 265 | 1.00000 | 0.00000 | 0.000E+00 | 0 |
| Totals | | | | | 1.00000 | 9.252E+10 | 0.012499 |

Difference between calculated oil volumes 1E-06 - use Solver to make it zero.




Oil droplet distribution

## Effect of plate width on the integrated process of CPI and ICB: Input parameters

### Influent characteristics

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | 0.001 | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | 60 | μm |
| Standard deviation | 15 | μm |

### Interceptor dimensions

| | | |
|---|---|---|
| Width | VARIABLE | m |
| Height | 1.32 | m |
| Length | 1.08 | m |
| Number of plates | 5 | |
| Plate width | **VARIABLE** | m |
| Plate length | 1 | m |
| Plate spacing | 0.02 | m |
| Plate angle to horizontal | 60 | ° |

### Basis of costing

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of plate width on the integrated process of CPI and ICB: Results

| Plate width (m) | Oil flow to interceptor ($m^3/h$) | Interceptor width (m) | Total plate area ($m^2$) | Effective plate area ($m^2$) | Interceptor volume ($m^3$) | Velocity through plates (m/s) | Reynolds Number | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Interceptor efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| **1** | 0.01250 | 1.5 | 5.0 | 3.0 | 2.144 | 0.0278 | 556 | 0.0065 | 479 | 52.1 |
| **2** | 0.01250 | 2.5 | 10.0 | 5.0 | 3.573 | 0.0139 | 278 | 0.0107 | 142 | 85.8 |
| **3** | 0.01250 | 3.5 | 15.0 | 8.0 | 5.002 | 0.0093 | 185 | 0.0120 | 43 | 95.7 |
| **4** | 0.01250 | 4.5 | 20.0 | 10.0 | 6.431 | 0.0069 | 139 | 0.0123 | 17 | 98.3 |
| **5** | 0.01250 | 5.5 | 25.0 | 13.0 | 7.860 | 0.0056 | 111 | 0.0124 | 5 | 99.5 |

| Plate width (m) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) | Cost of interceptor (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|---|---|---|
| **1** | 0.00599 | 1.66 | 76.4 | 136.42 | 16.14 | 120.28 | 323.73 | 3104.54 | 3428.27 | 3548.55 |
| **2** | 0.00178 | 0.49 | 22.7 | 201.64 | 26.59 | 175.05 | 126.12 | 920.90 | 1047.02 | 1222.07 |
| **3** | 0.00053 | 0.15 | 6.8 | 254.81 | 29.68 | 225.13 | 61.86 | 275.36 | 337.22 | 562.35 |
| **4** | 0.00021 | 0.06 | 2.7 | 301.33 | 30.48 | 270.86 | 35.69 | 108.40 | 144.10 | 414.95 |
| **5** | 0.00006 | 0.02 | 0.8 | 343.45 | 30.84 | 312.61 | 17.75 | 33.18 | 50.93 | 363.55 |

## Effect of plate angle on the integrated process of CPI and ICB: Input parameters

**Influent characteristics**

| | | |
|---|---|---|
| Water flowrate | 10 | $m^3/h$ |
| Water density | 1000 | $kg/m^3$ |
| Water viscosity | 0.001 | Pa.s |
| Oil concentration | 1000 | mg/l |
| Average oil density | 800 | $kg/m^3$ |
| Average droplet size | 60 | µm |
| Standard deviation | 15 | µm |

**Interceptor dimensions**

| | | |
|---|---|---|
| Width | 1.50 | m |
| Height | VARIABLE | m |
| Length | VARIABLE | m |
| Number of plates | 5 | |
| Plate width | 1 | m |
| Plate length | 1 | m |
| Plate spacing | 0.02 | m |
| Plate angle to horizontal | **VARIABLE** | ° |

**Basis of costing**

| | | |
|---|---|---|
| Cost of oil | 0.62 | $£/m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of plate angle on the integrated process of CPI and ICB: Results

| Plate angle (°) | Oil flow to interceptor ($m^3/h$) | Interceptor length (m) | Interceptor height ($m^2$) | Total plate area ($m^2$) | Effective plate area ($m^2$) | Interceptor volume ($m^3$) | Velocity through plates (m/s) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Interceptor efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| **60** | 0.01250 | 1.08 | 1.32 | 5 | 3 | 2.144 | 0.0278 | 0.0065 | 479 | 52.1 |
| **40** | 0.01250 | 1.61 | 1.10 | 5 | 4 | 2.660 | 0.0278 | 0.0092 | 266 | 73.4 |
| **30** | 0.01250 | 1.81 | 0.96 | 5 | 4 | 2.602 | 0.0278 | 0.0099 | 205 | 79.5 |
| **20** | 0.01250 | 1.96 | 0.80 | 5 | 5 | 2.349 | 0.0278 | 0.0103 | 172 | 82.8 |
| **10** | 0.01250 | 2.05 | 0.63 | 5 | 5 | 1.939 | 0.0278 | 0.0105 | 158 | 84.2 |

| Plate angle (°) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) | Cost of interceptor (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|---|---|---|
| **60** | 0.00599 | 1.66 | 76.4 | 136.42 | 16.14 | 120.28 | 323.73 | 3104.54 | 3428.27 | 3548.55 |
| **40** | 0.00332 | 0.92 | 42.4 | 140.68 | 22.76 | 117.93 | 182.43 | 1721.55 | 1903.98 | 2021.90 |
| **30** | 0.00256 | 0.71 | 32.6 | 140.22 | 24.65 | 115.57 | 156.35 | 1325.43 | 1481.78 | 1597.35 |
| **20** | 0.00215 | 0.60 | 27.4 | 138.16 | 25.66 | 112.50 | 141.12 | 1114.11 | 1255.23 | 1367.73 |
| **10** | 0.00197 | 0.55 | 25.2 | 134.63 | 26.10 | 108.52 | 134.16 | 1022.56 | 1156.72 | 1265.24 |

Effect of plate spacing on the integrated process of CPI and ICB: Input parameters

| Influent characteristics | | | Interceptor dimensions | | |
|---|---|---|---|---|---|
| Water flowrate | 10 | $m^3/h$ | Width | 1.50 | m |
| Water density | 1000 | $kg/m^3$ | Height | 1.32 | m |
| Water viscosity | 0.001 | Pa.s | Length | VARIABLE | m |
| | | | | | |
| Oil concentration | 1000 | mg/l | Number of plates | 5 | |
| Average oil density | 800 | $kg/m^3$ | Plate width | 1 | m |
| Average droplet size | 60 | μm | Plate length | 1 | m |
| Standard deviation | 15 | μm | Plate spacing | **VARIABLE** | m |
| | | | Plate angle to horizontal | 60 | ° |

Basis of costing

| | | |
|---|---|---|
| Cost of oil | 0.62 | £/$m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of plate spacing on the integrated process of CPI and ICB: Results

| Plate spacing (m) | Oil flow to interceptor ($m^3/h$) | Interceptor length (m) | Total plate area ($m^2$) | Effective plate area ($m^2$) | Interceptor volume ($m^3$) | Velocity through plates (m/s) | Reynolds Number | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Interceptor efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| **0.020** | 0.01250 | 1.08 | 5.0 | 3.0 | 2.144 | 0.0278 | 556 | 0.0065 | 479 | 52.1 |
| **0.025** | 0.01250 | 1.10 | 5.0 | 3.0 | 2.183 | 0.0222 | 556 | 0.0066 | 475 | 52.5 |
| **0.030** | 0.01250 | 1.12 | 5.0 | 3.0 | 2.223 | 0.0185 | 556 | 0.0066 | 471 | 52.9 |
| **0.040** | 0.01250 | 1.16 | 5.0 | 3.0 | 2.302 | 0.0139 | 556 | 0.0067 | 462 | 53.7 |
| **0.050** | 0.01250 | 1.20 | 5.0 | 3.0 | 2.382 | 0.0111 | 556 | 0.0068 | 454 | 54.6 |

| Plate spacing (m) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) | Cost of interceptor (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|---|---|---|
| **0.020** | 0.00599 | 1.66 | 76.4 | 136.42 | 16.14 | 120.28 | 323.73 | 3104.54 | 3428.27 | 3548.55 |
| **0.025** | 0.00593 | 1.65 | 75.8 | 136.76 | 16.27 | 120.49 | 322.02 | 3077.26 | 3399.28 | 3519.78 |
| **0.030** | 0.00588 | 1.63 | 75.1 | 137.10 | 16.40 | 120.70 | 320.31 | 3049.98 | 3370.29 | 3490.99 |
| **0.040** | 0.00578 | 1.60 | 73.7 | 137.77 | 16.66 | 121.11 | 316.86 | 2995.43 | 3312.29 | 3433.40 |
| **0.050** | 0.00567 | 1.58 | 72.4 | 138.43 | 16.92 | 121.51 | 313.38 | 2940.88 | 3254.26 | 3375.77 |

## Effect of plate length on the integrated process of CPI and ICB: Input parameters

| Influent characteristics | | | Interceptor dimensions | | |
|---|---|---|---|---|---|
| Water flowrate | 10 | $m^3/h$ | Width | 1.50 | m |
| Water density | 1000 | $kg/m^3$ | Height | VARIABLE | m |
| Water viscosity | 0.001 | Pa.s | Length | VARIABLE | m |
| | | | | | |
| Oil concentration | 1000 | mg/l | Number of plates | 5 | |
| Average oil density | 800 | $kg/m^3$ | Plate width | 1 | m |
| Average droplet size | 60 | µm | Plate length | **VARIABLE** | m |
| Standard deviation | 15 | µm | Plate spacing | 0.02 | m |
| | | | Plate angle to horizontal | 60 | ° |

**Basis of costing**

| | | |
|---|---|---|
| Cost of oil | 0.62 | £/$m^3$ |
| Equipment cost model | $CS^n$ | |
| Daily hours of operation | 16 | h/day |
| Annual days of operation | 250 | days/yr |
| Period of operation | 25 | yr |

Effect of plate length on the integrated process of CPI and ICB: Results

| Plate length (m) | Oil flow to interceptor ($m^3/h$) | Interceptor length (m) | Interceptor height (m) | Total plate area ($m^2$) | Effective plate area ($m^2$) | Interceptor volume ($m^3$) | Velocity through plates (m/s) | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Interceptor efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| **1** | 0.01250 | 1.08 | 1.32 | 5.0 | 3.0 | 2.144 | 0.0278 | 0.0065 | 479 | 52.1 |
| **2** | 0.01250 | 2.08 | 2.19 | 10.0 | 5.0 | 6.830 | 0.0278 | 0.0106 | 148 | 85.2 |
| **3** | 0.01250 | 3.08 | 3.06 | 15.0 | 8.0 | 14.115 | 0.0278 | 0.0119 | 47 | 95.3 |
| **4** | 0.01250 | 4.08 | 3.92 | 20.0 | 10.0 | 23.998 | 0.0278 | 0.0123 | 18 | 98.2 |
| **5** | 0.01250 | 5.08 | 4.79 | 25.0 | 13.0 | 36.479 | 0.0278 | 0.0124 | 5 | 99.5 |

| Plate length (m) | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) | Cost of interceptor (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|---|---|---|
| **1** | 0.00599 | 1.66 | 76.4 | 136.42 | 16.14 | 120.28 | 323.73 | 3104.54 | 3428.27 | 3548.55 |
| **2** | 0.00185 | 0.51 | 23.6 | 221.38 | 26.41 | 194.97 | 129.16 | 958.88 | 1088.04 | 1283.02 |
| **3** | 0.00058 | 0.16 | 7.4 | 298.46 | 29.55 | 268.91 | 65.37 | 302.33 | 367.70 | 636.60 |
| **4** | 0.00023 | 0.06 | 2.9 | 371.78 | 30.44 | 341.35 | 37.33 | 116.99 | 154.32 | 495.67 |
| **5** | 0.00007 | 0.02 | 0.8 | 442.92 | 30.83 | 412.08 | 18.16 | 34.47 | 52.63 | 464.71 |

Effect of number of plates on the integrated process of CPI and ICB: Input parameters

| Influent characteristics | | Interceptor dimensions | |
|---|---|---|---|
| Water flowrate | 10 $m^3/h$ | Width | 1.50 m |
| Water density | 1000 $kg/m^3$ | Height | 1.32 m |
| Water viscosity | 0.001 Pa.s | Length | VARIABLE m |
| | | | |
| Oil concentration | 1000 mg/l | Number of plates | **VARIABLE** |
| Average oil density | 800 $kg/m^3$ | Plate width | 1 m |
| Average droplet size | 60 µm | Plate length | 1 m |
| Standard deviation | 15 µm | Plate spacing | 0.02 m |
| | | Plate angle to horizontal | 60 ° |

Basis of costing

| | |
|---|---|
| Cost of oil | 0.62 £/$m^3$ |
| Equipment cost model | $CS^n$ |
| Daily hours of operation | 16 h/day |
| Annual days of operation | 250 days/yr |
| Period of operation | 25 yr |

Effect of number of plates on the integrated process of CPI and ICB: Input parameters

| Number of plates | Oil flow to interceptor ($m^3/h$) | Interceptor length (m) | Total plate area ($m^2$) | Effective plate area ($m^2$) | Interceptor volume ($m^3$) | Velocity through plates (m/s) | Reynolds Number | Oil recovery ($m^3/h$) | Effluent oil concentration (mg/l) | Interceptor efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| **5** | 0.01250 | 1.08 | 5.0 | 3.0 | 2.144 | 0.0278 | 556 | 0.0065 | 479 | 52.1 |
| **10** | 0.01250 | 1.18 | 10.0 | 5.0 | 2.342 | 0.0139 | 278 | 0.0107 | 142 | 85.8 |
| **15** | 0.01250 | 1.28 | 15.0 | 8.0 | 2.541 | 0.0093 | 185 | 0.0120 | 43 | 95.7 |
| **20** | 0.01250 | 1.38 | 20.0 | 10.0 | 2.739 | 0.0069 | 139 | 0.0123 | 17 | 98.3 |
| **25** | 0.01250 | 1.48 | 25.0 | 13.0 | 2.938 | 0.0056 | 111 | 0.0124 | 5 | 99.5 |

| Number of plates | Oil flow to bioreactor ($m^3/h$) | Oxygen demand (g/l.h) | Bioreactor volume ($m^3$) | Cost of interceptor (£/yr) | Value of recovered oil (£/yr) | Cost of separation (£/yr) | Cost of bioreactor (£/yr) | Cost of electricity (£/yr) | Cost of bioprocess (£/yr) | Cost of int. process (£/yr) |
|---|---|---|---|---|---|---|---|---|---|---|
| **5** | 0.00599 | 1.66 | 76.4 | 136.42 | 16.14 | 120.28 | 323.73 | 3104.54 | 3428.27 | 3548.55 |
| **10** | 0.00178 | 0.49 | 22.7 | 192.29 | 26.59 | 165.70 | 126.12 | 920.90 | 1047.02 | 1212.72 |
| **15** | 0.00053 | 0.15 | 6.8 | 237.77 | 29.68 | 208.09 | 61.86 | 275.36 | 337.22 | 545.32 |
| **20** | 0.00021 | 0.06 | 2.7 | 277.61 | 30.48 | 247.13 | 35.69 | 108.40 | 144.10 | 391.23 |
| **25** | 0.00006 | 0.02 | 0.8 | 313.72 | 30.84 | 282.88 | 17.75 | 33.18 | 50.93 | 333.81 |