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Cost of a package plant membrane bioreactor

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MPhil thesis

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2007

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May 2007

This thesis is submitted in partial fulfillment of the requirements for the Degree of Master of Philosophy

Abstract

The capital and operating costs associated with a small package plant MBR for small-scale domestic duty has been appraised based on a medium-strength municipal wastewater. The three main membrane configurations were considered, these being multi-tube, hollow fibre and flat sheet, with the most appropriate plant design chosen for each configuration. The analysis proceeded via a consideration of the estimated amortised capital costs of the plant individual components and their installation, coupled with operating costs based largely on energy demand and residuals management. Energy demand was calculated from aeration and pumping costs, with aeration based on a combination of empirical relationships for membrane aeration and mass balance, and the modified Activated Sludge Model version 2 used for estimating tank size and sludge generation.

Results indicate that it is possible to produce a single household MBR at a capital cost similar to the current market cost for package treatment plants. Desludging and maintenance of these plants is similar but power requirements for an MBR are around 4 times that associated with more conventional package plants. Economies of scale exist from 6-20 p.e. plants but above 20 p.e. there is little cost difference per head, due to the design assumptions made. CAPEX and OPEX are to some extent interchangeable; reductions in CAPEX are associated with an increase in OPEX and vice versa. Whilst costs are high, the market for package MBRs is significantly influenced by the recycling potential of the effluent produced.

A cursory comparison has also been made with conventional package plant treatment processes.

Acknowledgements

This work would not have been possible without the support of my colleagues at Balmoral Group and at Cranfield University. I also extend thanks to Professor Judd for his advice and guidance throughout.

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1. Introduction

1.1. Package plant

Package plants are defined in BS 6297:1983 as "a prefabricated factory built sewage treatment installation." Package treatment plants are used to treat flows from small population ranges - up to 1000 population equivalent (PE). The design of such small plants is quite different from that of full scale municipal plants. Unlike municipal plants package plants are a bulk consumer product and as such must meet a range of customer requirements. The market for sewage treatment plants up to 120 population equivalent (PE) is valued at £25m p.a. A good understanding of the market requirements is vital to producing a successful product.

Package treatment plants are installed for single or multiple domestic dwellings, holiday camps, hotels, restaurants and other semi industrial applications. Such plants will not receive regular maintenance except a service 1–3 times per year. Servicing engineers do not have process expertise so the plant design must be rugged enough to be left for very long periods without manual input. Any solids waste produced must be stored in the plant until removed (desludging). Again, desludging is an infrequent activity, no more than once per year for small plant sizes.

A key factor in full scale, bespoke plant design is a survey of expected plant incoming flows and BOD and ammonia loads. With this information the plant can be tailored to cope with specific pollutants and peaking factors. However, such a survey cannot be afforded in a mass produced plant design so the final product must be robust enough to cope with unexpected loads.

Finally, tank design must be considered. Many package plants are designed for underground use both because of improved hydraulic handling and aesthetics. Mechanical loads on underground plants are high due to soil and water pressing on the tank walls; a smaller tank provides a more stable structure. Also plant design must take into account the transportation of the plant from factory to the installation site. Clearly it is possible to fit a greater number of small tanks on to a single lorry than is the case for larger more cumbersome plants.

1.2. Membrane bioreactors

Membrane bioreactors are suspended growth biological treatment systems coupled with a membrane separation process. The membrane separation process is used to replace the sedimentation step used in more conventional biological treatment systems. A full description of the process including operating parameters can be found in literature (Judd, 2006). Membrane bioreactors (MBRs) are becoming an increasingly popular treatment process for both municipal and industrial wastewater. There are now well over 2000 full-scale installations worldwide (Yang *et al*, 2006). Both capital cost and operational costs of this technology are deceasing significantly, making MBRs a more viable treatment option. The advantages and limitations of MBRs have been comprehensively explored by an ever expanding body of research (Yang *et al*, 2006; Judd, 2006)

Since the sedimentation step is obviated, MBRs require only the aeration tank and no large settlement tank, unlike the conventional activated sludge process. An MBR thus provides a much smaller footprint than a conventional activated sludge process. The sludge age in an MBR is completely decoupled from the hydraulic retention time so the sludge age can be set much higher than for activated sludge. As a consequence of operating with a long sludge retention time MBRs are usually operated at high biomass concentration in the reactor, usually between 8000 and 18000 mg/L, further intensifying treatment. The high MLSS concentration provides efficient treatment at high organic loading rates (Stephenson *et al*, 2000), especially since slow growing bacteria such as autotrophic nitrifiers are favoured. This leads to enhanced nitrification (Chiemchaisri and Yamamoto, 1993; Nah *et al*, 2000). The membrane negates settling difficulties such as bulking sludge and provides a consistent, very high quality effluent (Stephenson *et al*, 2000) – although poor microbiology can negatively impact on performance in other ways.

The operation of membranes is governed by the amount of liquid permeated through a unit area of membrane per time, defined as the permeate flux. Permeate flux takes SI units of $m^3/(m^2s)$, but is more commonly quoted in litres per m^2 per hour ($L/(m^2h)$) or m/day for convenience. The permeate flux is a function of the force driving permeate through the membrane and the resistance to that force. The driving force is usually the trans-membrane pressure (TMP), i.e. the pressure difference across the membrane. The resistance to that pressure comprises the resistance offered by the membrane itself and the resistance of the interfacial region adjacent to the membrane, in accordance with the resistance in series model (Field *et al*, 1995).

Despite improvements in technologies MBRs are still regarded as a high in capital cost and energy demand. It is the precise impact of the capital items of the plant on overall cost which is of interest when considering very small plants, such as package plant.

1.3. Package plant MBR

MBR plants are notionally well suited to treating small flows because of the high sludge handling capacity, small size and robust treatment associated with this technology. However, significant design challenges exist for all market requirements to be met. It is thus of interest to consider the cost implications of producing a package plant MBR in terms of both capital and operating costs to ascertain economic viability.

In order to produce a fair comparison a market survey has been undertaken to ascertain the boundary conditions within which the plant must operate. An overview of treatment processes available is also presented with regard to plant size and cost. This information is used to provide MBR plant specifications and likely range of costs of the individual system components and operating costs pertaining to system design and biokinetics. Available information from existing systems (Judd, 2006) can then be used to correlate membrane permeability with energy demand and maintenance requirements. Designs are considered for the major MBR design layouts based on the three main membrane configurations, these being hollow fibre (HF), flat sheet (FS) and multi-tube (MT). Each design is considered in turn and the cost calculated based on a range of flows (between 6 and 200 population-equivalent, or PE). Calculations are based on assumptions pertaining solely to the UK market, although some supplementary data is provided to set the frame for expansion into the more significant European and US markets.

1.4. Aims and objectives

The aim of this work specifically is to establish whether a generic MBR design can be produced that could be viable within the UK package plant marketplace.

The specific objectives of this work are to:

- provide an assessment of the package plant market and the potential advantages and disadvantages of an MBR within this market;
- generate a set of boundary conditions for package treatment plants from an investigation of standards and existing products;
- investigate the different configurations of an MBR within the boundary conditions specific to the package plant market;
- asses the impact of key assumptions on the results generated from the given boundary conditions.

Related publications by the same author

The MBR Book, Elsevier Publishing 2006, edited by S Judd. Chapter 2 Sections 2.2.3-2.2.5; Chapter 3 Sections 3.1 and 3.3.

The cost of a package plant membrane bioreactor, H. Fletcher, T. Mackley and S Judd. Water Research 41 2637-2635 (2007).

Cost of a package plant MBR: sensitivity analysis, H. Fletcher and S Judd. Presented at 4th IWA Membranes Conference, Membranes for Water and Wastewater Treatment, Harrogate, 15-17 May, 2007.

2. Package plants

To fully understand the challenges to be faced when designing a package plant MBR it is necessary to provide a complete set of boundary conditions. Like many consumer products package sewage treatment plants are subject to legislative standards. In the UK the BS standards apply, and for the wider European market it is the EN standards.

Alongside the legislation it is important to understand the market into which any product is to be introduced. A market survey has thus been carried out, including an opinion survey of customers. Identifying the customers was a significant challenge in this. The customer group chosen were plant installers because of the relationship they hold with manufacturers and the breadth of experience held. However, on occasion water companies use package plants and have their own standards which they expect plant to meet. This information is presented alongside the legislative standards.

Data gained in this section is used to provide boundaries for plant design and assist in any necessary assumptions.

2.1. Package plant standards

Products are required to comply with the British and European standards which then provide complete transparency about the design and test procedures used in the development of the product. Compliance with standards then provides customers with an assurance that the product will be fit for purpose. Some water companies also provide a standard which any works that will be adopted by them must meet. A full list of reviewed standards can be found in Appendix A. The key points pertaining to the design of a package plant MBR have been identified from legislative standards.

- A British Water focus group on package treatment plants has produced a leaflet with expected flows and loads of 200 L/person/day flow. Further flow and load data can be found in British Water Code of Practice – Flows and Loads 2.

- The design of small works, defined as treatment for up to 1000 population by BS 6297:1983, should be based on sewage flows only. No surface water should be taken into the works (BS 6297:1983).
- Access to the plant for routine maintenance, sampling and removal of sludge shall be part of the design.
- The design should prevent unauthorised access to plant and machinery (pr EN 12566-3).
- Temperatures outside of the average UK temperatures and excessive loadings or unusual contaminants, including excessive use of detergents or high loadings from waste disposal units, should be taken into account when designing a small works.
- Site specific requirements should be accounted for. These may include high flows from hospitals or high grease levels from restaurants (BS 6297:1983).
- Flow measurement is important for any works, and even in small plants allowance should be made for measuring effluent flow rate. This may be achieved by installing a v-notch weir at the outlet of the plant to facilitate the use of portable measuring equipment (BS 6297:1983).
- Electrical equipment should comply with IEE wiring regulations (16th edition, 1992). The minimum nominal pipe diameter for inlet/outlet pipes should be 100mm for flows less than or equal to $4m^3$/day and 150mm for flows above that value (pr EN 12566-3).
- All materials used should comply with the relevant EN standards; a full list of references can be found in pr EN 12566-3.

To comply with pr EN 12566-3 new products must be subject to a series of tests to prove suitability for the purpose. The complete plant must be tested for water tightness, treatment efficiency and structural strength. Detailed descriptions of these tests are detailed in Annex's A-C of the standard. To pass the treatment efficiency test the plant must be designed to cope with peak flow discharges from washing machines and baths, as well as accounting for the variants in daily flow. Provision should also be made to cope with a 24 hour power failure and a period of no flow. Tests need to be carried out on both 50% and 200% of nominal hydraulic and BOD loading.

Tanks should be designed with backfill load, hydrostatic load, dynamic load and loads imposed by the fittings and equipment within the plant being considered (pr EN 12566-3). Tanks must be marked with the following information: Manufacturer and product identification, EN number, nominal organic daily load (BOD_5 kg/day), nominal hydraulic daily flow (m^3/day), conditions of use, date of manufacture, name of lab and test report number.

Aside from the general specifications standards have been produced for pre-treatment, primary settlement and suspended growth biological treatment. Although primary settlement is not used in full scale MBR plants a standards review has been conducted, since for certain plant designs considered in this study screening upstream of the reactor can be replaced with settlement.

2.1.1. Pre-treatment

The pre-treatment method to be adopted and is dependent on the process used and the feedwater quality. Excessive grease and oil can create severe operational difficulties and it is recommended (BS 6297:1983) that these are removed at source by means of a grease trap or similar. Rags and other floating debris can cause operational problems and should be removed prior to treatment (BS 6297:1983). This may be achieved by screening or by the application of a static baffle in the primary settlement tank. If screens are used provision should be made to remove and safely dispose of screenings (BS 6297:1983). A means of clearing debris from a baffle in the primary tank should be installed, and human contact with raw influent must be minimised for health and safety reasons.

2.1.2. Primary settlement

Sludge storage should be provided for 6 months minimum in the primary tank (Dwr Cymru and Wessex Water, 2005). The cost of removing stored sludge should be included within the plant operating costs (BS 6297:1983). Sludge storage requirement can be calculated as $0.52m^3$/year per population and should be accommodated in the bottom two-thirds of the settlement tank. The side wall height to be adopted should be not less than 400mm and the gross capacity of the tank should be such as to provide a retention period of not more than 12h (BS 6297:1983). Methods for the calculation of tank volume and area can be found in the British standard.

Upflow settlement is preferable to horizontal settlement since upflow settlement tanks do not have to be completely emptied to be de-sludged (BS 6297:1983). Primary settlement is not necessarily required for plants with less than 100 PE. The angle of slope of the sides of the hopper must not be less than 50° for conical and 60° for pyramid hoppers. Where sludge is removed by suction, which is the case for many package treatment plants, a sloping floor may not be required (BS EN 12255-4).

2.1.3. Suspended growth treatment (activated sludge)

Biological treatment by suspended growth processes is a widely understood and mature technology. For the activated sludge process flow control should be incorporated to minimise risk of sludge washout (BS 6297:1983). Aeration requirements are defined in the standard but should not be less than 2g O_2 / BOD applied (BS 6297:1983). For carbonaceous oxidation the conventional medium rate process is suitable with F/M of 0.25 to 0.5 kg/kg/d, the design MLSS being 2-3g/l and the sludge age (or solids retention time, SRT) being 2-4 day. For nitrification a low rate process is suitable, with an F/M ratio of 0.1-0.155 kg/kg/d, a design MLSS of 3-5g/l and 7-12 day sludge age (BS EN 12255-6).

2.1.4. Screenings estimation

At present no available package plants have screens; primary settlement is used for this purpose. To obtain realistic values for inert screenings that will be contained within the plant a literature study has been undertaken.

During wet weather a higher portion of screenings is faecal matter. This is assumed to be because at high flows the proportion of fine paper washed through the screen is higher (UKWIR, 2000). This study was performed on 6mm screens and the problems associated with solids breakthrough of this nature would be expected to be reduced with finer screens. However maintaining a low velocity through the screen improves the screen effectiveness.

Table 1 shows the composition of screenings at 4 sewage works. These works are all large plants that have influents from a variety of sources and combine storm water drainage as well. In the case of a package plant leaves are unlikely to arise, and with proper education the sanitary products could also be largely eliminated from the waste. Because the screenings are retained in the aeration chamber any faeces break down and are treated. Some portion of the 'other material' is vegetable matter which is also broken down in time. If the plant is designed based on the screenings quantities estimated for a full treatment works the estimated volume for screenings is likely to be conservative.

Table 1. Screenings constituents, 4 different sewage works (UKWIR, 2000)

Plant	Faeces	Sanitary products	Fine paper	Leaves	Other
A	13.6	43.6	34.3	0.8	7.7
A	17.0	14.4	48.4	1.1	19.2
A	27.9	8.1	47.9	0.0	16.1
B	12.7	9.5	59.7	0.3	17.8
B	13.2	29.4	39.8	0.9	16.6
C	14.3	23.6	41.4	0.0	10.7
C	22.0	20.3	38.1	0.0	19.6
C	18.0	20.2	46.1	0.0	15.7
D	12.5	31.6	47.8	2.2	5.9
D	11.2	34.3	39.7	0.0	14.8
D	14.3	29.9	36.0	3.3	16.5
Average	16.1	24.1	43.6	0.8	14.6
Max	27.9	43.6	59.7	3.3	19.6
Min	11.2	8.1	34.3	0.0	5.9

Page (1989) performed some research into variation in screenings volumes between wet and dry weather and found that wet weather screenings collection can be up to 15 times that of dry weather systems (Table 2). Droste (1997) quotes a more conservative value of 5:1 peaking factor, increasing to up to 20 times in combined sewage systems. This is because any matter collecting in the distribution system will be washed into the treatment works in periods of high flow. In a system with no storm water collection the weather has no impact, but during certain times of day the flows into the plant will be far higher. This effect suggests that a multiplier should be applied when designing a screenings store.

Table 2. Comparison of actual and predicted screenings volumes at 5 treatment works (Page, 1989)

Plant	Actual (m^3/day)	Predicted (m^3/day)	Multiplier
1	0.950	0.130	7
2	0.460	0.100	5
3	0.563	0.080	7
4	0.583	0.040	15
5	0.305	0.020	15

To estimate the quantity of screenings that will be found in the sewage a selection of values have been collated from the literature (Table 3). For the purposes of the design calculations a value of five times the average value of these data has been used for screenings production - a somewhat conservative safety factor. Sewage screenings have a high proportion of water. To estimate the actual solids storage needed an average value of the % solids has been taken from the same literature and the storage reduced to this level.

According to the data in Table 3 the average volume required for screenings storage per person per year is:

Estimated Production x Safety Factor x Solids Fraction = Trash Store

$$0.0062x5x0.14 = 0.0045m^3 \quad (1)$$

Table 3. Literature data on screenings production

Reference	m^3 screening /m^3 flow	Estimated Annual Production			% solids
		m3/person	m3/21 PE	m3/100 PE	
Droste, 1997	3.50E-06	0.0003	0.0054	0.0256	10
	3.50E-05	0.0026	0.0537	0.2555	20
Liu & Liptak, 2000	7.40E-05	0.0054	0.1134	0.5400	5
	2.62E-04	0.0191	0.4010	1.9094	20
Metcalf & Eddy, 2003	4.40E-05	0.0032	0.0675	0.3212	10
	1.10E-04	0.0080	0.1686	0.8030	20
Spellman, 1985	3.75E-05	0.0027	0.0575	0.2738	
	9.00E-05	0.0066	0.1380	0.6570	
UKWIR, 2000	5.00E-05	0.0037	0.0767	0.3650	10
	1.50E-04	0.0110	0.2300	1.0950	20
Ave	8.56E-05	0.0062	0.1312	0.6245	14
max	2.62E-04	0.0191	0.4010	1.9094	
min	3.50E-06	0.0003	0.0054	0.0256	

2.2. Package plant market

A survey of the package plants currently commercially available has been conducted. All information in this survey was taken from company websites (Appendix B), and not an independent source, but can none-the-less be assumed to provide representative information concerning package plant products. The companies selected for the survey were chosen so as to provide an overall view of the market at the time of writing. The companies range from specialist small companies (such as Biotank) to large companies providing a diverse range of products (Klargester).

2.2.1. Company overview

Although the companies reviewed vary in size, they fall into two broad categories, consulting engineers and manufacturers. The consulting engineers comprise Biwater, Busse, Copa, KEE and WPL. the manufacturers comprise Balmoral Tanks, Biotank, Clearwater, Conder, Hepworth, Klargester Environmental and Titan Pollution Control. There are links between some of these companies. Balmoral Tanks produce the tanks for Hepworth Ecocell. Copa sells Kubota membrane technology to Conder for their MBR product. Klargester and Titan products are part of the same parent company (Kingspan). These examples of inter-relationships between companies illustrate the commercial complexity of the package plant marketplace.

2.2.2. Tank material

The package plant market is dominated by steel and GRP (glass-reinforced polymer) tanks, with only Balmoral tanks and Hepworth drainage manufacturing products in polyethylene tanks. Of the companies surveyed 6 (50%) produce products in GRP tanks and 7 (60%) produce products in steel tanks. There is a tendency to fabricate the larger, above ground products in steel, whereas many of the single household units are produced in GRP tanks.

2.2.3. Population range

Package plants vary in size from single household units, usually designed around 4-6 population equivalent (PE), to units capable of treating effluent from 1000's of PE in a single tank. Companies marketing single household units in the UK include Biwater, Clearwater, Klargester, WPL, Titan, Balmoral, Biotank and Conder.

2.2.4. Processes

Processes based on biodegradation can be classified according to the process configuration, feeding regime, and oxidation state (Table 4). Process configuration defines the way in which the water is contacted with the biomass, which can either form a layer on some supporting media to form a fixed biofilm or else be suspended in the reactor. Suspended growth systems provide higher mass transfer but the biomass subsequently needs to be separated from the water and excess biomass, or sludge, needs to be disposed of. Feeding regime defines the way in which the feedwater is introduced, which can be either continuous or batch-wise. Feeding in batches allows the same vessel to be used both for biodegradation and separation, thus saving on space. This is the case for the sequencing batch reactor, or SBR. Finally, the reduction-oxidation (redox) conditions are defined by the presence of either dissolved oxygen (DO) (aerobic conditions) or some other species capable of providing oxygen for bioactivity (anoxic conditions) or the complete absence of any oxygen (anaerobic conditions). The different redox conditions favour different microbial communities and are used to provide different types of treatment.

Aerobic treatment is used to remove organic compounds (BOD or COD) and to oxidise ammonia to nitrate. Aerobic tanks are often combined with anoxic and anaerobic tanks to provide biological nutrient removal in full scale treatment plants, but this is rare in package treatment plant since nutrient removal is rarely specified in plants below 3000 PE. The various facets of biological processes in general are described in more detail in various reference books (Metcalf and Eddy, 2003). Almost all package treatment processes are configured according to the sub-categories listed in Table 4, and their function and performance depends on which specific sub-categories apply.

In all biotreatment processes, the treated water must be separated from the biomass. Fixed film processes require a smaller settlement capacity for this, since the biomass is fixed onto some supporting matrix which substantially retains it in the bioreactor and only dead biomass which is sloughed off the media must be settled out. The clumps of dead cells sloughed off a fixed film process are larger than the flocculated solids from a suspended growth process. The MBR technology employs membrane filtration to quantitatively retain all the biomass in the reactor, eliminating the requirement for substantial floc growth associated with AS processes employing downstream sedimentation. However measures must be taken to suppress membrane fouling (Section 4.1). Those processes employed in currently available package plant treatment products are illustrated in Figure

1. This figure indicates that the SAF process is the most common type of product followed by AS. These are both simple processes requiring little technology to operate.

Table 4. Examples of biological processes and their characteristics

	Process configuration		Feeding regime	
	Fixed film	Suspended growth	Continuous	Fed-batch
AS		X	X	
BAF	X		X	
RBC	X		X	
SAF	X		X	
SBR		X		X
TF	X			
MBR		X	X	

AS Activated sludge process
BAF Biological aerated filters
RBC Rotating biological contactor
SAF Submerged aerated filter
SBR Sequencing batch reactor
TF Tricking filter

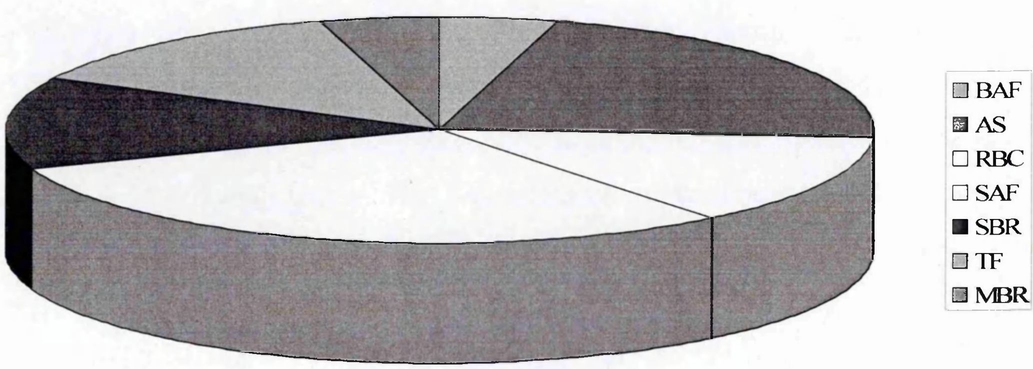

Figure 1 % of Single PE Plants

2.3. Customer survey

Most sales of package plants are through installers. Because of this, market share can be increased by making plants more attractive to the installers. To gain a clear understanding of what is required many installers were contacted to establish what is perceived as important when specifying plant. The findings of this survey are discussed in this section of the report. A full list of contacts can be found in Appendix B.

2.3.1. Results from customer feedback

Cost is the driving force when specifying products, installers will specify the cheapest plant which meets the site requirements in most cases. A plant providing a higher quality effluent than is required by the discharge consent will none-the-less lose out to the less expensive competing product. The majority of customers are not willing to invest in a more expensive product to process beyond the minimum requirements.

2.3.2. Tank design and installation

Tank geometry significantly impacts on the ease of installation of a plant. A major concern to installers is the depth of the tank; feedback from most suggests that shallower tanks are preferable. In areas with a high water table a shallow tank design not only eases installation but can also remove the need for costly de-watering pumps during excavation. Shallow tanks are also favoured by the fact that excavation is easier for a hole that is shallow and wide than for one that is deep and narrow. Deep holes can lead to the requirement to shore up the sides, and health and safety then become a greater concern. Furthermore, shallow tanks permit greater flexibility in sites with a rock layer under the surface. The installation issues were raised by the majority of those surveyed. This is an issue that is of greater concern to installers than end users. Installers who do select deep tanks often make that choice based on historical loyalties with particular manufacturers.

Most of the plants on the market are constructed of GRP (Section 2.2.2). The majority of installers questioned favour this material, since it is stiff and allows for less expensive tanks. GRP has become the industry standard as its advantages outweigh its disadvantages, and any alternative material (such as polyethylene, PE) has to offer clear advantages over GRP for package plant duties. A clear benefit of GRP tanks is that they can be buried in high water table areas without mechanical failure, something the installers perceive may not be possible for tanks constructed of PE.

The possibility of installing above-ground plants was not popular with the surveyed installers. The main issue was aesthetics, with installer expressing concern over pumping into the plant itself from the lower invert sewer. A pumping station would be required in this instance, with either a large bore passage or a macerator included. Whilst the installers accepted that a saving would be made through obviating excavation work there was little enthusiasm for highly visible treatment plants. Almost all existing plants are thus installed below ground, particularly smaller units.

2.3.3. Tank equipment

The internal equipment of the tanks was not found to significantly impact on customer preference. Having simple internals is a benefit and is usually highlighted in sales literature, but there are many successful plants available that use complex components (e.g. the Klargester Biodisc). However complex the process is the factors governing the success of a product are firstly reliability and secondly ease of maintenance, with the reliability issue obviously being of high priority. Installers operate largely on historical trends, and a plant is likely to be commercially successful if it can be shown to perform well even if its cost slightly exceeds that of a less reliable competitive technology. Many of the installers are small companies and cannot afford to take large risks.

The power consumption of the plant was not highlighted as a major selling point by any of the installers. Usually clients do not pay any attention to the electrical energy demand of the system, and rarely compare two different products on this basis. Many of plants use diaphragm blowers whose power consumption is comparable to a low wattage incandescent light bulb. Coupled with this, the blowers are usually low-noise and generate little heat, and therefore are largely unobtrusive.

2.3.4. Operation and maintenance

The power requirement varies from 3 watts to 123 watts per person. The power consumption follows no pattern with the processes. As previously mentioned, power consumption is not regarded as a factor impacting on plant sales.

All plant manufacturers offer maintenance contacts which includes desludging the facility. The desludging frequency varies between 1 month and 1 year. Smaller plants generally require less frequent sludge removal, with most having sludge storage for a whole year. The cost of desludging a plant is usually the most significant component of the operational cost of a package treatment plant. Minimal sludge production is regarded as highly advantageous in a package plant.

2.3.5. Odour

Odour minimization is considered a highly import factor in the design of small treatment plants because of their proximity to domestic dwellings or commercial properties such as hotels. Odour represents the most invasive of the sensory perception elements (the others being noise and visibility); it is a problem that customers can easily detect and has to be dealt with quickly. Any new design must consider odour minimisation as a high priority. Smaller plants almost always have venting through a soil stack. Most of the installers did not report any problems with odour. Odour problems have usually been ameliorated by de-sludging the tanks.

2.3.6. Summary

The key outcomes from the installer survey can be summarised as follows:

- Plants producing high quality output must none-the-less be able to compete with existing systems on price.
- Plants must be based on relatively shallow tanks.
- Plants must be shown to be easily maintainable and of sound build quality, with proof of reliability being available.

- Plant desludging represents the highest portion of maintenance costs for a conventional package plant; operational savings would be thus most noticeable for a plant providing a lower sludge yield.
- Plants must be designed to combat odour release.
- Plants constructed of GRP would be subject to less resistance by the installers since it is a well established material; any other material (such as PE) must be shown to have advantages over GRP.
- Any above-ground plant must be unobtrusive; only very small tanks could be employed for this configuration, limiting the design to high intensity processes such as membrane bioreactors.
- Plants should not demand excessive power, but this is less of a priority than capital expenditure and desludging frequency.

3. Process options

The first stage in assessing the feasibility of a package plant MBR is to undertake a high level plant design based on empirical evidence and plant size. The customer and installer survey (Section 2.3) has highlighted primary concerns of plant size as pertaining to ease of installation. The majority of systems are installed below ground since this minimises visual impact and obviates supplementary liquid pumping. However, below ground installations require excavation and this adds significantly to the overall cost. Plant size is thus an important factor in considering package treatment plant marketability.

3.1. Process volume

Figure 2 illustrates the average plant size for each process taken from available data about the existing products on the market. One common feature that is illustrated in Figure 2 is the tendency for manufacturers to undersize package plants somewhat. All sizes are given as m^3/PE, the volume given includes settlement zones. To compare this data with the European standard settlement tank sizes for a plant size have been obtained using the formula given in BS 6297:1983 and added to reactor tanks which were sized based on the BOD loading rate (kg BOD_5/m^3) given in the pr EN 12255-6 standards. Data for MBR loading rates was obtained from published literature (Stephenson *et al*, 2000), since none were available from the market survey. It can be seen from the figure that the MBR is significantly smaller than the other plants, making it ideally suited to package treatment were the choice based on size alone. It is worth noting that the data depicted refer to plants having no balance tank of any sort; the inclusion of any flow equalisation will obviously impact upon plant size.

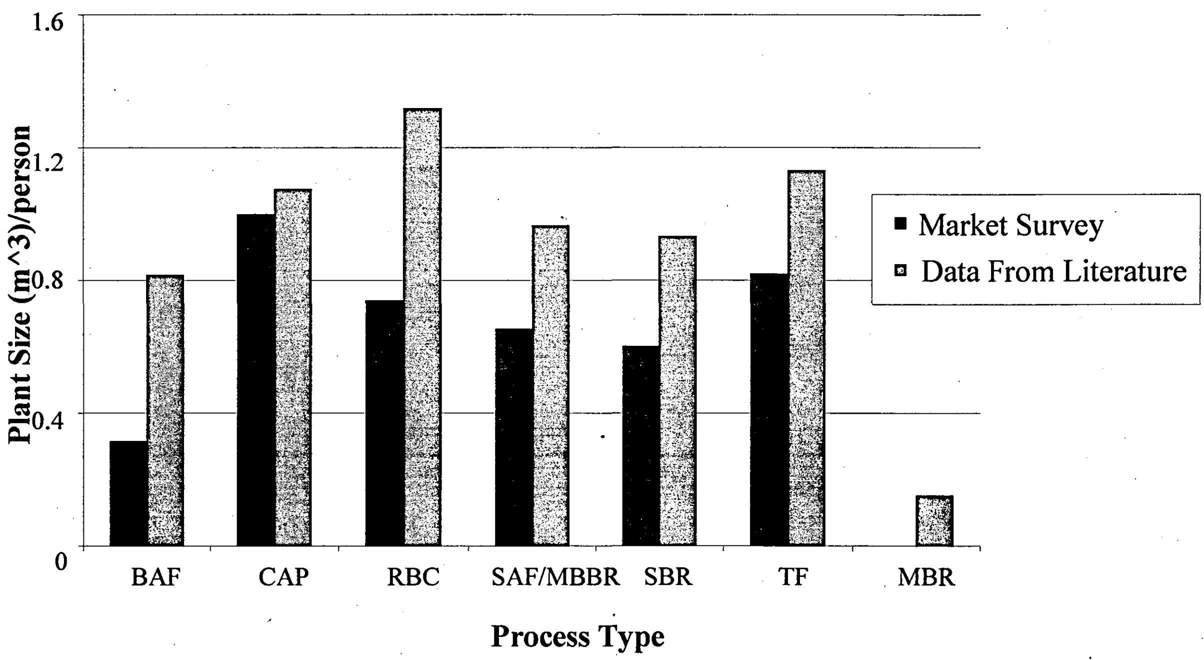

Figure 2 *Plant sizes by process*

3.2. Process cost

The following plants were selected for the cost comparison:

MBR (Membrane Bioreactor)

SBR (Sequencing Batch Reactor)

AS (Activated Sludge)

SAF (Submerged Aerated Filter)

Cost comparison of different process types

3.2.1. Analysis criteria

To determine overall costs for the production of these plants a set of specific costs were selected to represent the predominant influences on the overall cost of a new plant. Costs of the following key items were estimated:

- Tank size and consequent installation cost
- Membranes, media, membrane support rig etc.
- Aeration requirement
- Control system requirement

- Pumps
- Blowers

The aim of the exercise was to provide costs relating to actual production, using supplier pricing as a guide. It should be noted at this stage that these costs are an estimate only, and more comparative than absolute. Plants were designed based on the same influent and effluent qualities, and two plant sizes considered: 6 and 50 PE. Flow and water quality data were taken from the British Water table of loadings for the influent:

Flow - 200 L/day/person

BOD5 - 60 g/day/person

NH_3–N - 8 g/day/person

The effluent discharge consent the limit was set as 20:30:20 (BOD:SS:NH_3-N)

Calculations for the aeration vessel sizes, sludge production and air requirements for each plant were performed individually using the design spreadsheets (Appendix C). Component prices were obtained from existing suppliers in the marketplace where possible. Tank costs were estimated as rotamoulded, high density polyethylene (HDPE) tanks using company data provided by Balmoral Group (Appendix B). Installation costs were estimated as £50/m^3 soil removed based on excavating a simple rectangular hole with sloped sides to prevent collapse.

3.2.2. Plant specifications

3.2.2.1. Activated sludge process

6 PE

Primary Tank Volume = 2.4m^3 (assuming 6 month interval)

Aeration tank volume = 0.34m^3 (assuming 2500 mg/l MLSS)

Clarifying tank = 0.6m^3

Continuous cylindrical tank containing all the compartments assumed

Dimensions: Total volume 3.34m^3, length 2.0m, diameter 1.5m

Air flow requirement = 11.47 m^3/day

Cost of aeration system for diffusers = £100

50 PE

Primary Tank Volume = $18m^3$ (assuming 6 month interval)

Aeration tank volume = $2.73m^3$ (assuming 2500 mg/l MLSS)

Clarifying tank = $3.76m^3$

Continuous cylindrical tank containing all the compartments assumed

Dimensions: Total volume $24.5m^3$, length 5m, diameter 2.5m

Air flow requirement = 100 m^3/day

Cost of aeration system for diffusers = £300

3.2.2.2. Membrane bioreactor

The control system was specified as a low level float switch and a timer which both control the operation of the membrane. This system assumes no back pulsing only relax mode. The MBR tanks have been increased from 160L to 640L and 1360L to 5360L for the 6 PE and 50 PE respectively for the aeration vessel. This is to allow for a buffer capacity within the vessel and is calculated from 40% x Influent Flow + Aeration requirement (PrEN 12566-3). Design flux was based on 75% of standard membrane manufacturer's recommendations for full scale plants. The reduction in flux was to reduce maintenance requirements associated with high flux fouling.

6 PE

Aeration tank volume = 640 L (based on 5000 mg/l MLSS)

Air flow requirement = 34 m^3/day

Assume continuous vertical cylindrical tank

Dimensions: Total volume 640L, diameter 0.8m, height 1.4m

Cost of aeration system for diffusers = £140

Cost of permeate pump = £140

Estimated operating flux = 10 $l/m^2/h$

Membrane area = 5 m^2

Estimated membrane cost = £1000

Cost of float switch and timer = £200

50 PE

Aeration tank volume	=	5360 L (assume 5000 mg/l MLSS)
Air flow requirement	=	290 m^3/day

Assume continuous vertical cylindrical tank

Dimensions: Total volume 5.4m^3, diameter 1.7m, height 2.5m

Cost of aeration system for diffusers	=	£450
Cost of permeate pump	=	£350
Estimated operating flux	=	10 L/m^2/h
Membrane area	=	42 m^2
Estimated membrane cost	=	£3000
Cost of float switch and timer	=	£200

3.2.2.3. Sequencing batch reactor

The value of the SBR controller was determined from existing supplier costs of the kiosk (the current kiosks are supplied complete with a blower and full control equipment). The estimated cost of the internal control components are £750 and £1000 for the 6 PE and 50 PE plants respectively. The cost of airlift pumps for the reactors was estimated as approximately £30 each for both PE sizes.

6 PE

Primary Tank Volume	=	2.4m^3	(assuming 6 month interval)
Aeration tank volume	=	1.17m^3	(assuming 2500 mg/l MLSS)

Assume continuous cylindrical tank containing all the compartments

Dimensions: Total volume 3.6m^3, length 3.24m, diameter 1.62m

Air flow requirement	=	45 L/min (based on 2m water depth)
Cost of aeration system for diffusers	=	£135

50 PE

Primary Tank Volume	=	18m^3	(assuming 6 month interval)

Aeration tank volume	=	$9.72m^3$	(assuming 2500 mg/l MLSS)

Assume continuous cylindrical tank containing all the compartments

Dimensions: Total volume $27.7m^3$, length 10.9m, diameter 2.5m

Air flow requirement	=	375 l/min (based on 2m water depth)
Cost of aeration system for diffusers	=	£736

3.2.2.4. Submerged aerated filter

6 PE

Primary Tank Volume	=	$2.4m^3$	(assuming 6 month interval)
Aeration tank volume	=	$0.24m^3$	(assuming 2500 mg/l MLSS)
Clarifying tank	=	$0.62m^3$	

Assume continuous cylindrical tank containing all the compartments

Dimensions: Total volume $3.3m^3$, length 2.6m, diameter 1.3m

Air flow requirement	=	28 l/min (assume 2m water depth)
Cost of aeration system for diffusers	=	£95
Media	=	£60

50 PE

Primary Tank Volume	=	$18m^3$	(assuming 6 month interval)
Aeration tank volume	=	$2m^3$	(assuming 2500 mg/l MLSS)
Clarifying tank	=	$3.76m^3$	

Assume continuous cylindrical tank containing all the compartments

Dimensions: Total volume $23.8m^3$, length 5m, diameter 2.5m

Air flow requirement	=	235 l/min (assume 2m water depth)
Aeration system for diffusers	=	£452
Media	=	£250

Note: it was assumed that the SAF clarifier tanks are the same as for as conventional AS plant.

A summary of this cursory cost comparison is given in Table 5. From these data, it can be seen that for both the 6 PE and 50 PE the MBR plant is the most expensive product to manufacture. When installation costs are accounted for the plant becomes a more economically viable, and potentially significantly less expensive than the SBR option which is considered to provide the highest effluent quality of the existing commercially-available technologies.

Table 5. Cost comparison summary

6 PE

	MBR	SBR	SAF	AS
Installation	£715	£2,434	£1,443	£1,700
Controller	£200	£750	NA	NA
Blower	£140	£135	£95	£100
Liquid Pump In	NA	£30	NA	NA
Liquid Pump Out	£140	£30	NA	NA
Media	NA	NA	£60	NA
Membrane	£1,000	NA	NA	NA
Totals (no installation)	£1,480	£945	£155	£100
Totals	£2,195	£3,379	£1,598	£1,800

50 PE

	MBR	SBR	SAF	AS
Installation	£4,382	£12,074	£7,564	£7,564
Controller	£200	£1,339	NA	NA
Blower	£450	£736	£452	£300
Liquid Pump In	NA	£60	NA	NA
Liquid Pump Out	£350	£60	NA	NA
Media	NA	NA	£250	NA
Membrane	£3,000	NA	NA	NA
Totals (no installation)	£4,000	£2,195	£702	£300
Totals	£8,382	£14,269	£8,266	£7,864

3.2.3. Decision matrix

Information presented in Section 2.3 indicates plant cost is the most important factor when selecting a package plant, but that other factors also contribute to technology selection. As a further means of comparison a decision matrix has been constructed to illustrate the criteria identified in Section 2.3. The decision matrix was constructed and graded quantitavely by two design engineers and reviewed by the board of a package plant manufacturing company.

The operational cost of the plant impacts on sales but is not regarded as crucial; this is shown in the matrix as life time cost (depreciation and replacement parts) and air requirement (and hence power cost). A small tank size is desirable to reduce potential installation problems, although the market survey (Section 2.3) suggests that large tanks are marketable. Once the criteria had been selected they were weighted from 1-5 where 5 impacts heavily on potential sales and 1 has minimal impact.

3.2.3.1. Grading methodology

Each process was graded from 1-10 for every criteria, where 1 is desirable and 10 is undesirable. The basis for each decision is summarised in the following list.
Purchasing and Installation costs decisions were based on the results from Section 3.2.

- Life time costs were obtained by establishing which components are likely to need replacing (e.g. blowers and membranes) and their life from installation to replacement. From this a 10 year cost to maintain each plant was estimated. The cost of replacing membranes is high, impacting significantly on total cost. Controlling an SBR requires probes and pumps which wear out and will eventually need replacing. SAFs may need media replacement occasionally. AS plants have the least number of moving parts and consumables and so require minimal maintenance.
- Tank sizes were graded based on the information in Section 3.2.3.

- Process simplicity was assessed by establishing how difficult each process is to operate and maintain. An SBR has a complex control system which requires advanced understanding to maintain. An MBR has membrane units which must be cleaned and replaced periodically. Both an SAF and an AS plant have no control or complex components but do require some care as the biological growth must be maintained.
- The air requirement is based on the results from Section 3.2.2.
- Process reliability is a measure of how consistent the effluent quality is under stress conditions of high or low flow. An MBR has a physical barrier which guarantees high effluent quality, and also the high biomass concentrations allow for higher stress loading. An SBR operates under high biomass concentrations so has potential to cope with stress loading, and because it is a fed-batch process with buffer storage hydraulic washout of biomass is not possible. Because the biomass is grown on media in an SAF it should not wash out of the system under high hydraulic loading but the process has limited potential to deal with stress loading. An AS is highly susceptible to any changes in conditions, since biomass can wash out of the system easily and consequently the process must be carefully controlled.

Table 6. Process Decision Matrix

Criteria	SBR	AS	MBR	SAF	*Weighting*
Purchasing Cost	7	2	9	2	*5*
Installation Cost	9	6	2	5	*5*
Life Time Cost	5	2	9	4	*3*
Tank Size	8	6	2	5	*3*
Process Simplicity	9	2	7	3	*2*
O_2 required	10	6	9	6	*1*
Reliability	3	9	1	5	*4*
Weighted total	159	110	115	94	

When all factors are considered the prospect of marketing an MBR is shown to be realistic.

4. Analysis scope

MBRs are now a well established technology and a number of textbooks have been written on the subject covering membrane configurations and process design features (Judd, 2006; Stephenson *et al*, 2000) MBRs have achieved considerable market penetration in the municipal water treatment sector over the past 15 years (Hanft, 2006). The advantages over conventional processes are well documented (Stephenson *et al*, 2000), as are the constraints imposed by membrane fouling (Le Clech *et al.*, 2006). Applications of the increasingly diverse range of commercial technologies available have tended to be restricted to the range between 10 and 50,000 m^3/day of installed capacity, although larger MBRs are being built year-on-year. On the other hand, increasing water scarcity coupled with stringent regulations have meant a single-household MBR (<5 m^3/day), with the effluent being recycled for non-human contact applications such as irrigation, washing and toilet flushing, is potentially economically viable. However, a single-household MBR is believed costly compared with established freshwater supply and effluent discharge. Indeed, only one established product exists in mainland Europe for flows of 0.8-1.6 m^3/day (4-8 population equivalent, or PE), based on flat sheet membrane configuration. Other commercial package MBR plant technologies tend to be targeted at higher flows, upwards of 125 PE (25 m^3/day).

4.1. Membrane fouling

Membrane fouling has been understood as the most challenging issue in operating an MBR (Chang *et al*, 2002; Chang *et al*, 2001a). As a consequence much of the research performed on MBRs has aimed to gain understanding and attempt to model fouling within the system in order to control it (Chang and Lee, 1998; Tardieu *et al*, 1999; Defrance and Jaffrin, 1999; Ognier *et al* 2001). Reviews on the subject have been published which describe in detail the mechanisms (Judd, 2004; LeClech *et al*, 2006). In depth analysis of fouling is not pertinent here and key operational aspects will be highlighted instead.

Membrane fouling results in a reduction of permeate flux per unit pressure applied; this translates as a loss of system efficiency. When a membrane is in contact with biological suspension, matter will deposit either on the surface of the membrane or within the membrane structure. The amassed foulants may be colloidal, particulate or solute material of organic or inorganic composition (Judd, 2004). Much research has been carried out to fractionate the foulants this is summarised in LeClech at al (2006). Material deposited on the surface of the membrane is known as the cake layer and may be removed using a physical procedure such as air scour or backpulsing, known as reversible fouling (Chang *et al*, 2002). Membrane aeration is a significant portion of the operational cost associated with an MBR, this is discussed in more detail in section 4.9.3. The formation of a cake layer can be controlled by operating at sub-critical flux for a given system. The critical flux is defined as the lowest flux that creates an irreversible deposit on the membrane surface (Espinasse *et al*, 2002). Critical flux is system and feedwater specific so it is impossible to define for each package plant sold. Therefore operating flux should be set at a conservative value to ensure the plant is within the sub-critical zone.

Foulants that have accumulated within the membrane pores cannot be removed by physical means and are known as irreversible fouling. The removal of dissolved material adsorbed into the membrane pores requires chemical cleaning.

The biomass characteristics will govern the biological fouling within an MBR. There is much confusing and contradictory evidence about a link between MLSS and fouling propensity. Some authors have reported a direct link between MLSS and cake resistance (Chang *et al*, 2001b; Shimizu *et al*, 1996). Early research shows that above an MLSS concentration of 30000 mg/L the flux decreases suddenly (Yamamoto *et al*, 1989) but latterly evidence has been presented that MLSS below 6000 g/L also has a negative influence on cake formation. Lubbecke *et al* (1995) suggested that MLSS concentrations up to 30000 mg/L are not responsible for irreversible fouling. A recent study by Rosenburger *et al* (2005) illustrated that between MLSS 8000-12000 mg/L there is no significant effect on cake formation. It is pertinent to design systems to operate within these boundaries.

During crossflow operation of a side stream MBR differing fluid velocities within the membrane module generate high shear forces. Shear forces break flocs into smaller particles within the reactor (Wisniewski and Grasmick, 1998). Floc breakage has two detrimental effects, smaller particles create a denser cake layer on the membrane surface allowing less fluid to pass through and the break-up of flocs releases extracellular polymers (EPS) which have been linked to catastrophic fouling (Chang *et al*, 2001a). Shear forces are lower when the membrane module is submerged within the biomass hence larger particles are found (Zhang *et al*, 1997).

Whatever the cause of fouling it must be successfully controlled for proper operation of an MBR. Full scale MBRs operate a combination of relax, backpulse, chemical cleaning and air scouring to achieve this. The exact combination depends on the membrane technology, large amounts of data on this subject has been correlated by Judd (2006). Package plants are not frequently maintained so the strategy for fouling amelioration must account for this. However, complex plants with a lot of mechanical equipment are also likely to fail so the final design should provide a balance of cleaning without a lot of complex machinery. Package plant MBRs should be operated at a very low flux to minimise the risk of catastrophic fouling.

4.2. Membrane aeration

It is necessary to aerate a submerged membrane unit in an MBR to promote cross flow filtration. The supply of aeration to the membrane unit reduces the change in trans-membrane pressure (TMP) with time (i.e. dTMP/dt) during constant flux operation. This change in suction pressure is commonly regarded as a measure of membrane fouling since an increase in the resistance to filtration would require greater suction pressure to achieve the same filtrate flow. Using a modified form of Darcy's law membrane flux can be defined as:

$$J = \frac{TMP}{\eta\left(R_m + R_c + R_f\right)} \qquad (2)$$

where TMP is the pressure difference across the membrane (kg/m/s^2), η is permeate viscosity (kg/m/s), R_m is the membrane resistance, R_c is the cake layer resistance and R_f is the internal fouling resistance (m^{-1}) (Chang and Judd, 2002). R_m is solely dependant on the membrane selected. R_f relates to pore clogging and is not affected by the aeration of the membrane module, therefore the difference in pressure seen is solely related to the reduction of the cake layer that forms on the membrane surface. The exact portion of the fouling that the cake layer represents is subject to the exact system conditions. Chang and Judd (2002) found this to be a small portion of the total resistance, just 8% (operating at 3.2 g/l MLSS). However other authors found that the cake layer represents a much higher proportion of the fouling on the membrane (Bouhabila *et al*, 1998). Because of this discrepancy there is clearly a maximum effect that aeration will have on a system and the desired effect is to minimise the cake layer.

The hydrodynamics of a system can have a profound effect on the rate of initial particle deposition (Choo and Lee, 1998). Coarse bubble aeration along the membrane surface is often used to reduce the build-up of material. Deposits can be either scoured off by the bubbles or, in the case of hollow fibre membranes, shaken off the surface by movement of the membrane. In either case it is generally recognised that an increase in membrane aeration increases membrane permeability, with generally a linear relationship being reported up to some threshold value.

The idea of a critical aeration value above which there is no further influence on pressure across the membrane was introduced by Ueda *et al* (1997). This concept has been confirmed by several authors (Liu *et al*, 2000; Bouhabila *et al*, 1998; Sofia *et al*, 2003). Ueda *et al* (1997) also hypothesised that the turbulence of the flow had such an effect on filtration pressure because the membrane fibres (in a hollow fibre unit) are shaken. However, a comparison between a submerged membrane and a rigid tubular membrane indicated that the effects of membrane movement are small with respect to reduction of the cake layer (Shimizu *et al*, 1996)

Experiments conducted by Liu *et al* (2003) show that the MLSS has some effect on the critical aeration of the biomass. This would appear to contradict other reported findings. Bouhabila *et al* (1998) tested a range of sludge concentrations and found the critical air flow was the same for all concentrations. The most significant factor affecting the efficient use of aeration is the physical dimensions of the setup. Ueda *et al* (1997) showed that by reducing the floor area by a third the resistance to filtration decreased 5 times. It is the difference in physical dimensions that changes the aeration intensity (air flow/unit floor area). Values pertaining to critical aeration intensity from published pilot and bench-scale studies are presented in Table 7. These are compared with values found from full scale installations (Table 8), where the latter data is likely to refer to sub critical operation

Table 7. Critical aeration intensity

Reference	Critical Aeration Intensity ($m^3/m^2/s$)	Flux ($l/m^2/h$)	MLSS (g/l)
Liu *et al* (2000)	0.014	5.2	4
Liu *et al* (2003)	0.0048	10	2
Liu *et al* (2003)	0.0048	20	2
Liu *et al* (2003)	0.0085	10	10
Liu *et al* (2003)	0.0085	20	10
McAdam *et al* (2005)*	0.017	13.51	3.5
Sofia *et al* (2003)	0.017	16.9	9+
Sofia *et al* (2003)	0.026	16.9	9+
Ueda *et al* (1997)	0.0068	12.08	10+
Ueda *et al* (1997)	0.01	12.08	10+

* Experiments performed on an SBR with a membrane

\+ average values

Data for membrane aeration rate per unit membrane area (SAD_m), in $Nm^3/hr/m^2$, are provided by Judd (2006). These data (Table 8) can been manipulated, using the available information on packing density, to provide the aeration intensity q_m:

$$q_m = SAD_m h_{module} \rho_m \quad (3)$$

where SAD_m is the aeration requirement per membrane area ($Nm^3/hr/m^2$), h_{module} the module height (m) and ρ_m the membrane packing density (m^2/m^3)

Table 8. ***Full Scale MBR Installation Aeration (adapted from Judd, 2006)***

Membrane Type	Aeration Intensity ($m^3/m^2/s$)	Flux ($L/m^2/h$)	MLSS (g/L)
FS			
Membrane 1	0.024	20	12-18
Membrane 1	0.034	33	8-12
Membrane 1	0.018	25	-
Membrane 2	0.044	27	12-15
Membrane 3	0.043	25	6-18
Membrane 3	0.032	21.6	22
Membrane 4	0.022	25	-
HF			
Membrane 5	0.158	18	15
Membrane 5	0.046	18	8-10
Membrane 5	0.103	12	10-15
Membrane 5	0.063	25	12
Membrane 6	0.126	10	12
Membrane 7	0.032	16	12
Membrane 8	0.036	25	8
Membrane 9	0.128	16	-

There is no clear pattern found within this data, and further research is needed to expand this work to a full relationship. However, the information in Table 7 and 8 indicates that the two key MBR membrane configurations operate within the following ranges of aeration intensity and flux:

0.018-0.044 $m^3/m/s$ and 20-33 $L/m^2/h$ for flat sheet and

0.032-0.158 $m^3/m^2/s$ and 10-25 $L/m^2/h$ for hollow fibre

4.3. Biomass aeration

Biomass aeration is the dissolution of oxygen into the liquor to provide an available source of oxygen for the metabolising microorganisms to digest successfully the organic pollutants. The aeration is also used to mix the liquor and keep the biomass in suspension. The most suitable aeration system will me*et al*l of the system requirements with the minimum power demand. There are several methods of achieving this and the most suitable should take the specific requirements of a package plant into account. These are

- Low cost
- Reliable operation of mechanical parts
- Submerged parts that operate without fouling by biological growth
- No unnecessary odour
- Quiet operation
- Reasonable power requirement (ie maximum electrical efficiency)

Two main types of air blower are available, centrifugal and diaphragm blowers. Previous experience from Balmoral Group company data suggests that the diaphragm blower types are both quieter and more reliable than centrifugal type blowers. Because of this evidence the air supply should be provided by a diaphragm blower where an external air source is needed. There is a maximum air flow which a diaphragm blower can efficiently provide, for plant sizes above that airflow it is necessary to move to a centrifugal type blower.

There are three main types of aeration used in wastewater plants; diffused air, jet aeration and mechanical agitation. Each system is considered on its suitability for use in the proposed MBR package treatment plant.

4.3.1. Jet aeration

In a jet aeration system liquor is pumped around the system. The flow passes through a venturi nozzle and draws air in because of the pressure drop associated with the increased fluid velocity at the neck (Figure 3).

Jet aerators provide very small bubbles, increasing the air liquid surface interface available for oxygen to dissolve into the liquid. Jet aerators also provide effective mixing because of the horizontal velocity given to the fluid by pumping. However, a jet aeration system will require both liquid pump, which will increase the cost of the system overall as liquid pumps are more expensive than air blowers. In certain system designs the air supply is pressurised to improve performance further increasing mechanical complexity. As Table 9 illustrates, the improvements in efficiency do not justify the added complexity with this type of aeration.

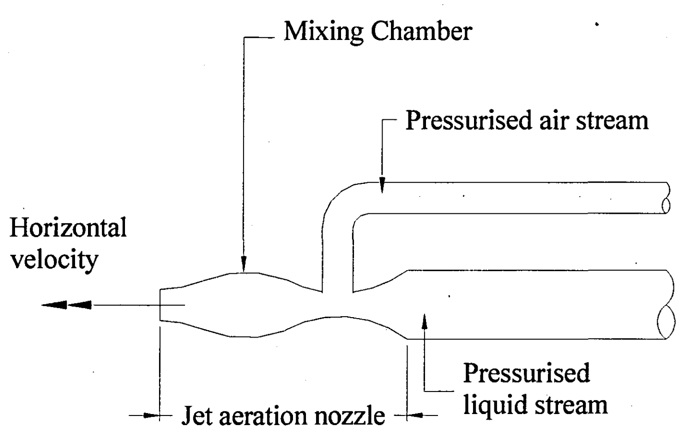

Figure 3 ***Jet Aeration***

4.3.2. Mechanical agitation

The main part of mechanical aerators is a rotor to which a number of blades are fitted the mixing action produces turbulence which entrains atmospheric air from the liquid surface (Figure 4). Tank geometry plays a significant part in the effects of a mechanical aerator and should be considered in detail when designing this type of aeration system. Circular tanks require less power to provide the same level of oxygen transfer (Rao *et al*, 2004).

Surface aerators are simple systems with little operational complexity although there is no data available for surface aerators used with an MBR. It is a significant risk that the turbines used in surface aeration will damage the membrane. By their nature surface aerators can be very noisy. Another problem with mechanical aerators is they produce odorous aerosol gasses by spraying the substrate into the atmosphere; however this may be overcome by appropriate venting systems. Finally the tank structure must be structurally strong enough to support the motor and rotor assembly, unlike a blower system the motor cannot be sited remotely. Despite the fact that surface aerators have the lowest operational costs the process risk, noise and odour problems suggest that this is not a suitable system for residential plants.

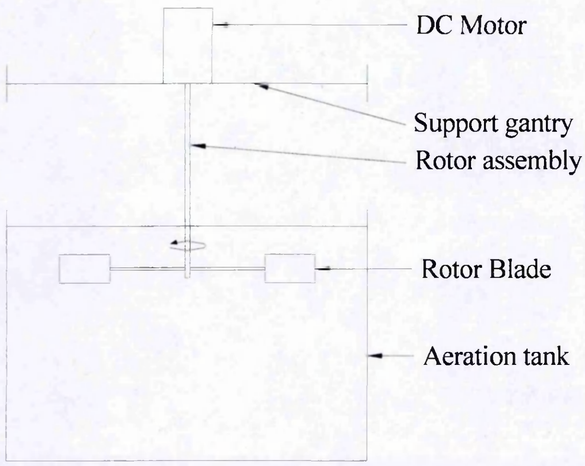

Figure 4 *Mechanical Aerator*

4.3.3. Diffused aeration

In a diffused aeration system air is introduced into the liquor as bubbles blown through a diffuser (Figure 5). Diffused aeration is broken into two main categories, fine bubble and coarse bubble. Fine bubble aeration is provided by porous diffusers, usually ceramic or membrane diffusers where as coarse bubble aeration is usually provided by a steel or plastic tube with small holes. Coarse bubble diffusion produces bubbles of 6-10mm diameter and fine bubble diffusers produce bubbles typically 2-5mm diameter. Fine bubble diffusers increase oxygen transfer efficiency because the smaller bubbles provide a greater interfacial area per unit volume of air (EPA, 1989) but the pressure drop across the diffuser is higher for fine bubble systems. Fine bubble diffusers are more susceptible to clogging which can severely detract from performance and requires more maintenance than coarse bubble systems (EPA, 1989). Recent developments in porous membrane material used in fine bubble diffusers reduces the maintenance problems associated with fine bubble systems (Stenstrom and Redmon, 1996) and it is worthy of note that the currently available systems use fine bubble diffusers and that no fouling problems have been reported during the routine maintenance. Because of the system simplicity diffusers for coarse bubble aeration are less expensive than those used in fine bubble diffusion (Stenstrom and Redmon, 1996), price quotations obtained from equipment suppliers are given in Table 9. Because of the higher air flows related to coarse bubble diffusion more turbulence is created promoting better mixing properties within the tank and membrane scour.

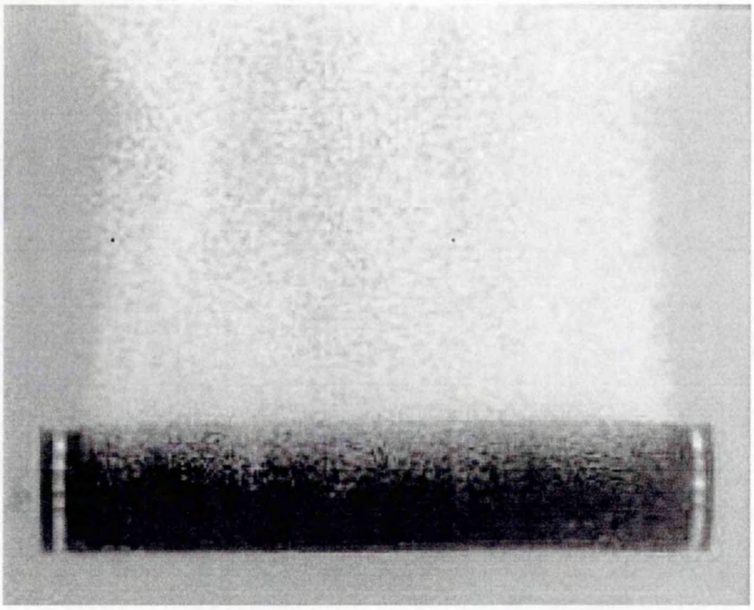

Figure 5 Diffuser Tube

4.3.4. Cost comparison

Cost comparison data are summarised in Table 9. These data were taken from a combination of sales literature from manufacture and from research data. Unsurprisingly the values given in the research data are more conservative than those from sales literature. Typically MBRs use coarse bubble aeration to scour the membrane which is supplemented by fine bubble aeration to provide sufficient air flow to the biomass. This is the aerator configuration that the design is to be based on.

4.4. Blower power consumption

An air blower is a mechanical process that increases the pressure of air by applying shaft work. The first law of thermodynamics states that "for any mass system the net heat supplied to the system equals the increase in energy of the system plus all the energy that leaves the system as work is done" (Massey, 1989). This can be expressed algebraically as

$$\Delta Q = \Delta E + \Delta W \quad (4)$$

Where

Q	= heat transferred
E	= system energy
W	= work done by the system

The system energy comprises:

- Kinetic energy ($\frac{1}{2}mu^2$)
- Potential energy, usually taken as gravitational energy (mgh)
- Internal energy (e)

The work done by the system is made up of:

W	= external work
p/ρ	= work done against forces due to pressure

From this information it is possible to derive equation (5) for a fluid moving at steady flow between points 1 and 2 as illustrated in Figure 6.

Table 9. Oxygen transfer efficiency and annual operation cost of aeration equipment

Aerator type	Transfer Efficiency	Operation cost per PE per year (Assume 6p/kWh)	Mixing	Comments
Fine Bubble	3-10% O_2 supply dissolved per m depth	£8-£29 £14 average	Low	The large variation in efficiency is because of the diffuser placement. Higher efficiencies are for tanks with an even distribution of air across the whole surface area of the tank.
Coarse Bubble	1-3% O_2 supply dissolved per m depth	£11-£28 £18 average	High	Low pressure across diffuser.
Jet Aeration	5% O_2 supply dissolved per m depth	£7-£8 + pumping costs	Medium	Liquid pump and air blower required for jet aeration. Very susceptible to negative influence of surfactants.
Mechanical Aeration	0.5-1.4 kg O_2 delivered per kWh	£1.50-£3.28	High	Produces potentially odourous aerosols. Noisy mechanical installations. Motor must be directly mounted above aerator and the tank must be strong enough to support installation.

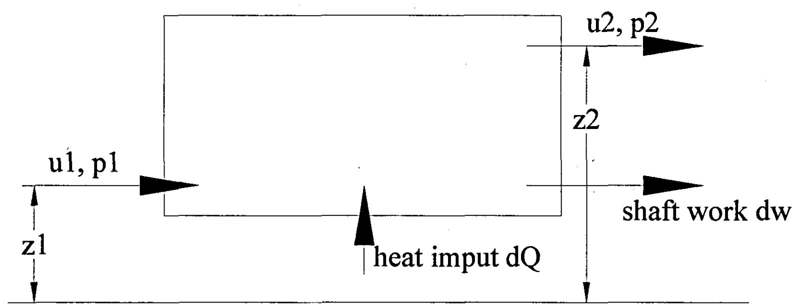

Figure 6 Steady flow system

$$\frac{\delta Q}{\delta m} = (e_2 - e_1) + \frac{(u_2^2 - u_1^2)}{2} + g(z_2 - z_1) + \frac{\delta W}{\delta m} + \frac{(p_2 - p_1)}{\rho} \qquad (5)$$

This equation is known as the steady-flow energy equation. The full derivation of which can be found in any good fluid dynamics text such as Massey (1989).

In this case the change in internal, kinetic and potential energy of the fluid are negligible. And the associated terms will be ignored from equation (5). In a blower there is no heat transferred assuming the effects of friction are ignored (this will be rectified later), therefore $\Delta Q = 0$. Because a blower exerts work on the fluid the work term will be negative in this case and to simplify the equation this will be ignored and regarded as positive. The fluid density in this system is not constant and thus it is necessary to use a differential form of equation (5)

$$dw = \frac{dp}{\rho} \qquad (6)$$

w is the work done per unit mass of fluid

for an ideal gas under adiabatic conditions

$$\frac{p}{\rho^{\gamma}} = const$$

or

$$\frac{p_1}{\rho_1^{\gamma}} = \frac{p}{\rho^{\gamma}}$$

or

$$\rho = \frac{\rho_1}{p_1^{\left(\frac{1}{\gamma}\right)}} p^{\left(\frac{1}{\gamma}\right)} \qquad (7)$$

where

γ is the ratio of specific heat capacity at constant pressure to specific heat capacity at constant volume (c_p/c_v) and is assumed to be a constant value. Substitute equation (7) into equation (6) and integrate between points 1 and 2.

$$w = \frac{p_1^{\frac{1}{\gamma}}}{\rho_1} \int_{p_1}^{p_2} \frac{dp}{p^{\frac{1}{\gamma}}} = \frac{p_1^{\frac{1}{\gamma}}}{\left(1 - \frac{1}{\gamma}\right)\rho_1} \left(p_2^{\left(1-\frac{1}{\gamma}\right)} - p_1^{\left(1-\frac{1}{\gamma}\right)} \right)$$

This equation can be manipulated to give

$$w = \frac{p_1 \gamma}{(\gamma - 1)\rho_1} \left[\left(\frac{p_2}{p_1} \right)^{1-\frac{1}{\gamma}} - 1 \right] \qquad (8)$$

This equation provides the work done per unit mass by a blower that is 100% efficient. No mechanical process is 100% efficient so a term for the blower efficiency (η) is added, this accounts for any mechanical losses through the system including friction. w is work per unit mass and power is total work per second. Equation (8) can be converted into power consumption in kW by dividing by time in seconds. It is also important to remember that the density of air varies directly with temperature and to compensate for that a temperature correction should be used in the form

$$\rho_t = \rho_0 \frac{T_0}{T_t} \qquad (9)$$

T_0 is the temperature in kelvin at 0°C (standard temperature) so the power consumed becomes:

$$\frac{P}{m} = \frac{p_1 \gamma}{1000\eta(\gamma - 1)\rho_0 t} \frac{T_1}{273} \left[\left(\frac{p_2}{p_1} \right)^{1-\frac{1}{\gamma}} - 1 \right] \qquad (10)$$

but

$$V = qt \qquad (11)$$

$$\rho = \frac{m}{V} = \frac{m}{qt} \qquad (12)$$

substituting Equation 12 into 10 leaves us with a theoretical blower power of

$$P = \frac{p_1 T_1 \gamma q}{2.73x10^5 \eta(\gamma - 1)}\left[\left(\frac{p_2}{p_1}\right)^{1-\frac{1}{\gamma}} - 1\right] \tag{13}$$

4.4.1.1. Blower efficiency

Equation (13) has an efficiency term used to account for the losses through the blower system. This efficiency term is a percentage expressed as a decimal. To get a clear idea of the efficiency of air blowers, data from manufacturers and distributors has been compared to the theoretical power at 100% efficiency. Figure 7 shows the efficiency plotted against flow rate. This figure illustrates that there is no clear correlation between flow rate or back pressure and the blower efficiency. However since most of the values lie between 20-60% efficiency, for the purpose of this work an efficiency of 40% has been assumed.

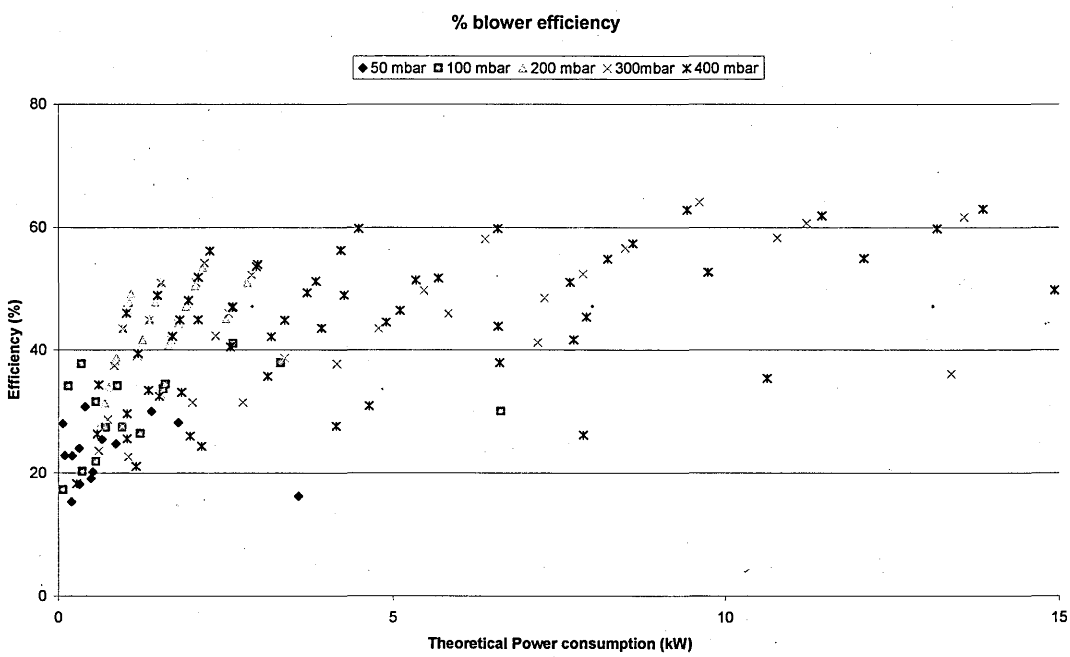

Figure 7 Blower Efficiency

4.5. Temperature effects

The effects of temperature have an impact on the system design and operating costs. In a package plant it is assumed that since the pipe runs are short the feed is still warm from the indoor use so design is carried out on wastewater at 12°C. To test the effect of this assumption temperature corrections have been built into the design and assessed in relation to capital and operational costs. Temperature affects all three main areas of process design in an MBR.

- Membrane critical flux
- Biological growth and reaction rates
- Oxygen transfer

A generic temperature correction relationship has is accepted as

$$k_T = k_{20}\varphi^{(T-20)} \qquad (14),$$

known as the Van't Hoff-Arrhenius relationship

4.5.1. Membrane critical flux

Temperature has an effect on the critical flux. In a series of experiments undertaken by Fan *et al* (2006) the effect of increasing temperature was plotted against flux. From this work an empirical relationship was derived (equation 15)

$$J_{c,T} = J_{c,20} x 1.025^{(T-20)} \qquad (15)$$

This follows the Van't Hoff-Arrhenius temperature relationship.

4.5.2. Biological effects

Using the van't Hoff-Arrhenius temperature relationship θ coefficients have been shown for biological parameters used in activated sludge modelling (Metcalf and Eddy, 2003). The θ correction for k_e, the endogenous decay coefficient, is 1.04.

The solids retention time is defined as:

$$\theta_x = \frac{1}{\mu} \qquad (16)$$

where μ is the specific growth rate. It is understood that μ is changed by temperature with a φ coefficient of 1.07 (Metcalf and Eddy, 2003). The SRT can be adjusted for temperature by the relationship:

$$\theta_{x,T} = \frac{\theta_{x,20}}{1.07^{(T-20)}} \qquad (17)$$

4.5.3. Oxygen transfer

Similarly a temperature correction factor for aeration temperature is commonly accepted as (EPA, 1989)

$$\varphi = 1.024^{(T-20)} \qquad (18)$$

4.6. Boundary conditions

Using information presented in Section 2.3 alongside knowledge gathered of MBR plants at various scales (Judd, 2006), some key assumptions concerning a package plant MBR can be made:

- Flow capacity of 200 L/(PE) (litres per day per person).
- Maximum of 10% of daily flow discharged over a period of one hour, i.e. 20 L/h/person.
- Influent quality of 300 mg/L BOD, 600 mg/L COD, 375 mg/L suspended solids, 45 mg/L NH_3-N (Britsh Water flows and loads at a BOD:COD ratio of 2:1; Metcalf and Eddy, 2003).
- No nutrient removal required: only an aerobic bio-zone used.
- Effluent quality of 20:0:5 BOD:SS:NH_3-N (Côte *et al*, 1998; Tao *et al*, 2005)
- Tanks assumed to be commercially-available vertical cylinders of polyethylene construction.
- Installation costs based on excavation of soil with no concrete lining required. The installation volume is based on a square hole with sides of the same width as the tank diameter. Each side must be excavated at an angle to prevent the hole collapsing (the angle of repose), this being taken as 45°. Excavation costs are estimated at €80 per m^3 of soil removed.
- Additional 600 mm height required for access and 200 mm air gap giving a total additional dig depth of 800 mm on top of the design water depth.
- Plants capable of sustainable operation for 6 months without maintenance visits.
- Plant capacity range of 6-49 PE. with no redundancy provided; 50% redundancy at 50-200 PE.
- Aeration demand of a technology is determined by generic membrane configuration (i.e. FS, HF or MT for flat sheet, hollow fibre or multi-tube respectively), independent of supplier.

4.7. System design

System components used in this comparison are listed in Table 10, and operating costs assumed outlined in Table 11. Assumptions made are dependent upon the overall system design. Immersed and sidestream (iMBR and sMBR respectively) options are considered, and these are further categorised according to membrane type (FS or HF) in an iMBR or MT for an sMBR. iMBRs membranes are assumed to be aerated whereas sMBR MT membranes are pumped. The process configurations considered are thus (Figures 8-10):

- Membrane-aerated HF iMBR – denoted HF
- Membrane-aerated FS iMBR – denoted FS
- Pumped MT sMBR – denoted MT

To simplify the schematics of process configuration notation has been used to define various items this is as follows:

B1 - Low pressure air blower
D1 - Fine bubble diffuser
D2 - Coarse bubble diffuser
M1 - Hollow Fibre membrane module
M2 - Flat sheet membrane module
M3 - Multi-tube membrane module
P1 - Permeate pump
P2 - Recirculation pump
Qb - Backpulse flow
Qe - Effluent flow
Qi - Influent flow
Qr - Recirculation flow
S1 - Fine screen
T1 - Timer switch

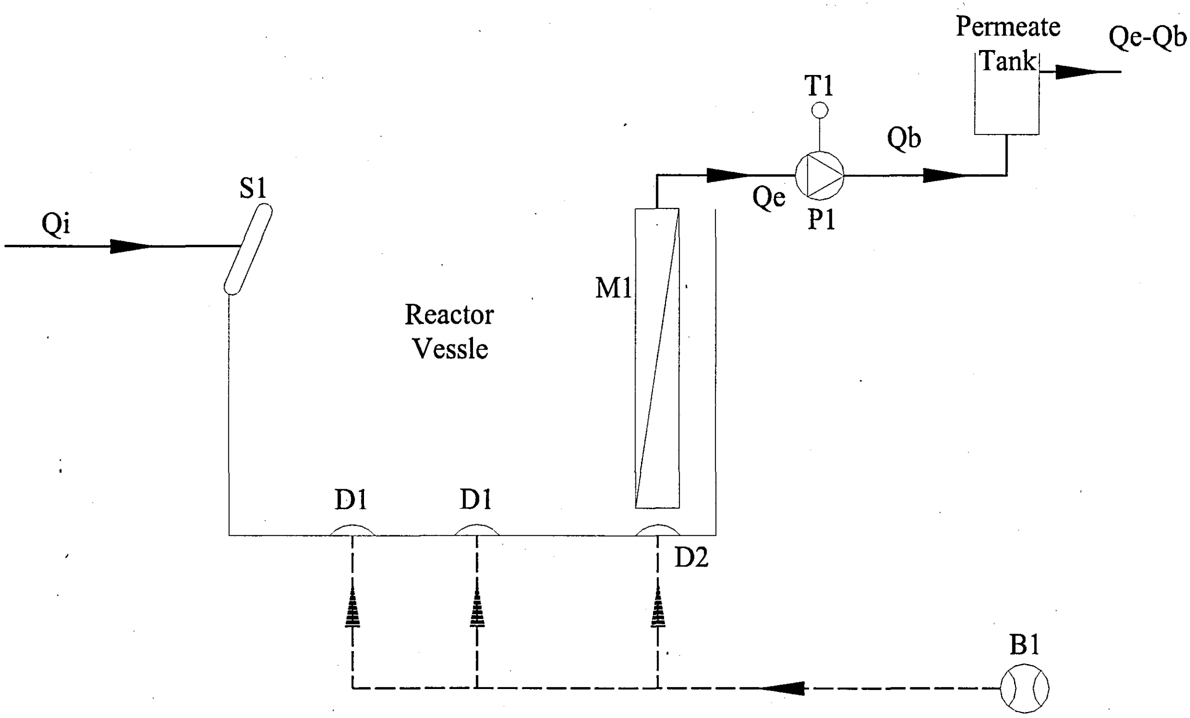

Figure 8 *HF iMBR layout*

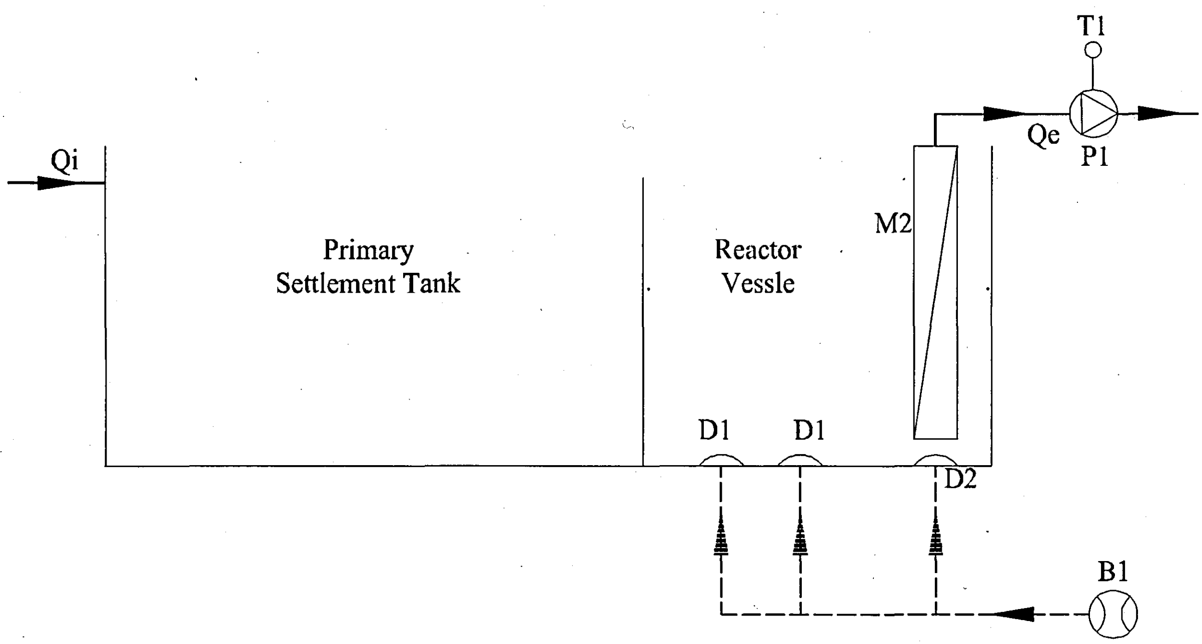

Figure 9 *FS iMBR layout*

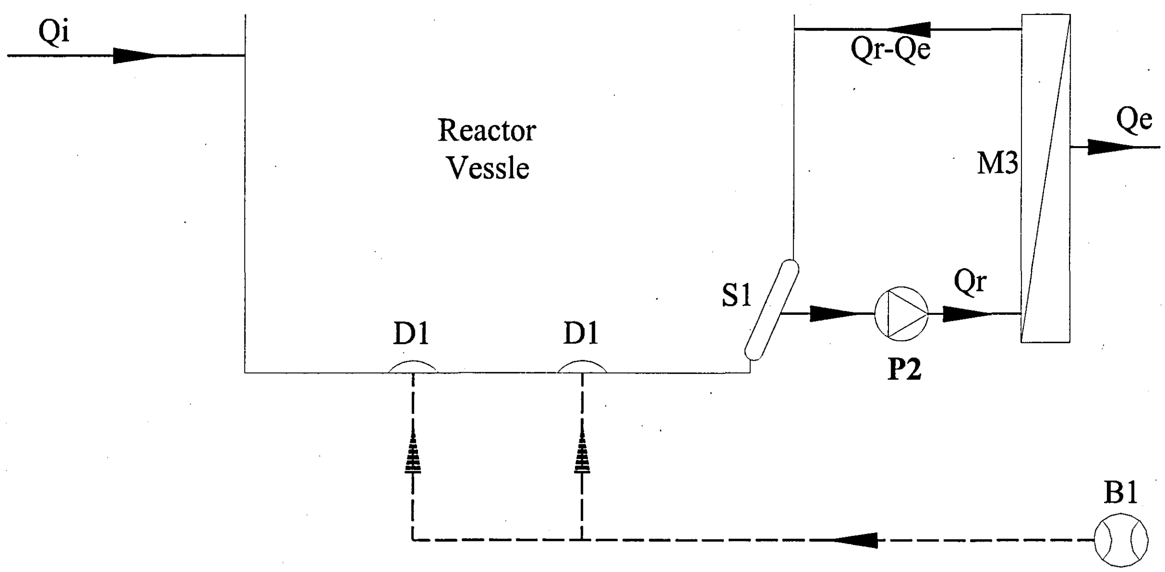

Figure 10 MT sMBR layout

Table 10. ***Range of capital items***

Component	Life, years	Comments
Tank	20	Vertical PE cylinder rotamoulded at a cost of: C = 1000 + 520 x V. Cylinder diameter given by: $d = \sqrt{4V/\pi h}$
Installation	20	Total installed tank depth estimated by: H = h + 0.8. Total dig volume calculated as: $V_{dig} = d^2H + 2dH^2$.
Membrane	10	€150/m^2 membrane area assumed for all technologies.
Liquid pumps (up to 3 bar)	5	Reversible pump for permeate suction and backpulse of HF iMBR, €400 per 20 m^3/h capacity. Permeate suction pump only for FS iMBR, €400 per 20 m^3/h. Retentate cross flow pump for MT sMBR, €400 per 20 m^3/h.
Air blower	5	Provides sufficient airflow for both biological aeration and membrane aeration (iMBR only). €126 for 85 l/min (up to 1m head) or €368 for 205 l/min (up to 2.5m head).
Air diffusers	10	Fine bubble for biological aeration, €24 per 7m^3/h flow. Coarse bubble for membrane aeration (not used in MT system), €8 per 15m^3/h flow.
Screen	10	HF, AL & MT fitted with 0.5mm screen, €1000. FS operated without screen but with primary settlement designed to BS6397:1983.
Timer switch	10	Solid state timer (€85) for: reversing permeate flow through HF module (iMBR) relax permeate flow for FS module (iMBR)

Table 11. ***Range of operational costs assumed***

Parameter	Cost	Normalised Cost per year per PE
Blower power cost, €/(day.person)	€0.18/kWh (40% efficiency)	$= \frac{P_{A,1}T_{K,1}}{1.263x10^4}\left[\left(\frac{P_{A,2}}{P_{A,1}}\right)^{0.286} - 1\right]\left\{\frac{Q_A}{PE}\right\}$
Liquid pumping power, €/(day.person)	€0.18/kWh (70% efficiency)	$= \frac{0.00617\rho g P_p Q_{pump}}{PE}$
Sludge disposal	€480 per desludge	$= \frac{480N}{PE}$
Maintenance visits	€11 per PE per visit	$= 11N$
Cleaning chemical costs	€0.48/kg sodium hypochlorite	$= n\frac{(t_c + \tau_c)c_c v_c}{8760PE}$

n = Number of visits

4.8. Design: biotreatment

4.8.1. Primary tank sizing method

For the FS iMBR a primary settling tank is used. The tank stores 10 L/PE/week sludge in the bottom two-thirds of the tank (BS 6297:1983):

$$V_p = \frac{10x52xPE}{1000xI_d}.\frac{3}{2} = \frac{0.78PE}{I_d} \quad (19)$$

Settled sewage has less BOD and SS load than raw sewage which impacts upon the downstream aeration tank volume, sludge production and process air requirement. The assumed influent strength into the reactor chamber is 90:75:45 COD:TSS:NH_3-N.

4.8.2. Reactor design

Much work has been performed on modelling MBR biokinetics (Huang *et al*, 2001; Fan *et al*, 1996; Lee *et al*, 2002; Yildiz *et al*, 2005; Liu *et al*, 2005; Wen *et al*, 1999; Xing *et al*, 2003), providing a range of values for key parameters for the MBR system (Table 12). Specific results are presented in Appendix D.

Table 12. Kinetic Parameters

Constant	Range	Value used	Unit
k_e	0.023-0.075	0.05	per day
Y	0.25-0.61	0.3	kg VSS/kg COD

4.8.3. Tank size and excess sludge production

Rearranging the equation for sludge concentration in the bioreactor or MLSS of Wen *et al* (1999), the tank volume can be calculated from:

$$V = \frac{QY\theta_x (C_i - C_e)}{X(1 + k_e\theta_x) - Y(C_i - C_{\sup})} \qquad (20)$$

Assuming 85% of COD is removed in the bioreactor and 12% by the membrane separation (Xing *et al*, 2001), then C_{sup} is 0.15Ci and C_e is $0.03C_i$. Equation 20 thus simplifies to:

$$V = \frac{QY\theta_x 0.97C_i}{X(1 + k_e\theta_x) - 0.15YC_i} \qquad (21)$$

Sludge production can be estimated from:

$$P_x = \frac{VX}{\theta_x} \qquad (22)$$

4.8.4. Aeration

The oxygen requirement to maintain a community of micro-organisms and degrade COD and ammonia and nitrite to nitrate can be found from a mass balance on the system (Metcalf and Eddy, 2003):

$$m_o = Q(C_i - C_e)\big/R_{COD} - 1.42P_x + 4.33Q(NO_x) \qquad (23)$$

Only the biodegradable fraction of the COD will consume oxygen and thus this equation gives a conservative estimate. NO_x is the amount of ammonia that is oxidised by the system, which can be calculated from a nitrogen balance on the system.

$$NO_x = N_i - N_e + 0.12P_x \qquad (24)$$

Much of the oxygen bubbled through the biomass remains undissolved; mass transfer effects must be taken into account, as defined by the volumetric mass transfer coefficient k_La per unit time. The rate of oxygen transfer into a liquid can be determined by:

$$OTR_{cleanwater} = k_L a(C^* - C) \qquad (25)$$

where C and C^* are the dissolved and saturated oxygen concentration values in kg/m3. For pure water and equilibrium conditions C is found using Henry's Law. Oxygen transfer is directly proportional to the depth of the water, since bubbles naturally rise so deeper tanks have a longer bubble residence time. Diffuser manufacturers provide an estimate of the oxygen transfer efficiency for their product. A market survey reveals the OTE per m depth to be around 2.5% for coarse bubble and 4.5% for fine bubble aerators. This can be converted to process conditions by the application of three correction factors (α, β and φ) which account for those sludge properties which impact on oxygen transfer:

$$OTE_{process} = \frac{OTE_{cleanwater}}{\alpha\beta\varphi} \qquad (26)$$

β accounts for the effects of salts and particulates, usually around 0.95 for wastewater (EPA, 1989) and φ relates to the effect of temperature given by equation 18 where T has been assumed to be 12°C on average.

The α factor is the difference in mass transfer (k_La) between clean and process water, and has the most significant impact on aeration efficiency of all three conversion factors. Studies of the impact of solids concentration on oxygen transfer in biological wastewater treatment systems have all indicated a decrease in OTE with increasing solids concentration regardless of the system studied, though the relationship is system and feedwater dependent (Chatellier and Audic, 2001; Muller *et al*, 1995; Krampe and Krauth, 2003; Gunder, 2001; Chang *et al*, 1999; Lindert *et al*, 1992; Fujie *et al*, 1992). In a number of studies of sewage treatment, an exponential relationship between α-factor and MLSS concentration has been observed, an exponent value of -0.084 (Krampe and Krauth, 2003; Gunder, 2001) is taken:

$$\alpha = e^{-0.084.X} \qquad (27)$$

In an immersed MBR some of the oxygen used for membrane aeration will transfer into the biomass and can reduce the oxygen demand. This transfer is taken into account within the model by reversing the calculation.

4.9. Design: membrane

Membrane permeability is governed primarily by crossflow velocity in a sidestream system and aeration in a submerged system. A combination of supplementary backflushing (HF systems), relaxation (HF and FS systems) and chemical cleaning is also employed to maintain permeability at an acceptable level.

4.9.1. Flux

Design flux dictates both aeration demand (for iMBR) or crossflow velocity (for sMBR) and membrane area requirement. Reasonably conservative estimates of average sustainable net flux, taken from real plant data (Judd, 2006), are taken as:

HF iMBR	10 $L/m^2/h$
FS iMBR	15 $L/m^2/h$
MT sMBR	50 $L/m^2/h$

4.9.2. Physical and chemical cleaning

As with aeration and backflush regimes, real plant data suggest the following protocol for maintaining permeability through cleaning (Judd, 2006):

Physical cleaning interval – 10 min

Physical cleaning duration – 1 min

Chemical cleaning interval – 6 months

Chemical cleaning duration – 2 hours

Cleaning reagent strength – 500 g/m^3

Cleaning reagent volume = reactor tank volume

4.9.3. Membrane aeration

It is necessary to aerate a submerged membrane unit in an MBR to promote cross flow filtration. Increasing the membrane module height thus increases the bubble path. However, package plants are required to be relatively shallow units to reduce installation problems associated with high water table and shallow bedrock. Clearly this must be reconciled with the requirement to produce narrow, deep units to maximise membrane aeration efficiency. A survey of the available data pertaining to membrane aeration intensity is presented in section 4.2. Aeration intensity can be converted to the air flow required for each particular membrane unit by:

$$Q_{A,m} = \frac{q_m V_m}{h_m} \tag{28}$$

the aeration intensity for FS has been taken as 110 $m^3/m^2/h$ and for HF as 310 $m^3/m^2/h$. In this case the air flow becomes a function of tank depth which for the main part has been assumed to be 1m.

4.9.4. Operating parameters

Table 13. Operating parameters

	HF iMBR	FS iMBR	MT sMBR
SRT, d (θx)	25	30	30
MLSS, mg/L (X)	8000	12000	13000
OTE – Fine bubble, % per m depth	4.5	4.5	4.5
OTE – Coarse bubble, % per m depth	2.5	2.5	na
Membrane permeability (L/m^2/h/bar)	135	300	200
Operating flux (L/m^2/h)	15	15	50
Membrane aeration intensity ($m^3/m^2/h$)	310	110	-
Cross flow velocity (m/s)	-	-	3
Fibre diameter/Plate spacing (mm)	3	10	8
Physical clean interval (min)	10	10	-
Physical clean duration (min)	1	1	-
Backflush flux (L/m^2/h)	17	0	-
Chemical clean interval (months)	6	6	6
Chemical clean duration (hours)	2	2	2
Chemical clean: equivalent volume	Reactor tank	Reactor tank	Reactor tank

4.10. *Cost development*

It is likely that a household would either have to borrow money to purchase a package plant or would save the money otherwise spent. In order to account for interest losses the annual cost has been corrected as a simple compound interest calculation at the Bank Of England base Rate, which is 5.25% at the time of writing.

5. Results and discussion

5.1. Plant costs

Single-household package plants are currently available for €2880-€9600 per unit. Installation costs depend on unit size and shape. Applying the same method as used in this paper yields an average installation cost of €3200. Assuming production of around 60% of material costs of the plant, to allow for company overheads and profit margin, the total plant cost ranges from €1728 to €5760, and is €4400 on average. This range of costs makes the MBR technology one of the more expensive for single-household use but is within the price range of existing commercial products. Much of the annual operating cost of a package plant is the plant desludge and maintenance at around €1080 per year. These costs are the same for an MBR as for a traditional plant. Power costs traditionally largely relate to aeration for aerobic treatment of COD and ammonia, approximately €4-6 per PE per year. For an MBR additional aeration is demanded for membrane scouring, and sidestream processes also demand power for pumping – a higher-energy process than aeration in iMBRs.

Unlike full scale plants operational costs are rarely accounted for when selecting a package treatment plant, since the cost per unit time is low. However, for an sMBR the power requirement is 20 times that of a conventional package plant.

5.2. Plant size

Figure 11 illustrates the total annual plant cost per person per year for 6-50 PE plants. The total cost incorporates capital equipment costs and installation costs amortised over the plant lifetime and operational costs. All technologies showed a sharp reduction in plant cost per person at very small plant sizes, with the trend approaching a constant value at around 20 PE.. The difference in total annual cost per person between 4 and 20 PE plant is around €250 for all plant types, whereas between 20 and 49 PE the difference is around €40. At 50 PE there is a sharp increase in plant cost (€40-70) because of the inclusion of 50% redundancy in the plant. However above 50 PE there is little difference in annual cost (~€30) up to 200 PE and this trend is not greatly affected by plant type. The annual operational and capital costs for 6, 20, 49, 50 and 200 PE are provided in Table 14.

Table 14. Annual cost per person at three different plant sizes

Configuration	Size	Cost/PE/year	CAPEX/PE	OPEX/PE/year
FS	6	307	905	214
	20	166	632	102
	49	129	552	74
	50	176	636	110
	200	149	581	89
HF	6	319	861	218
	20	166	506	106
	49	126	410	78
	50	183	553	116
	200	156	496	95
MT	6	324	662	247
	20	182	362	135
	49	142	264	106
	50	208	344	160
	200	181	293	138

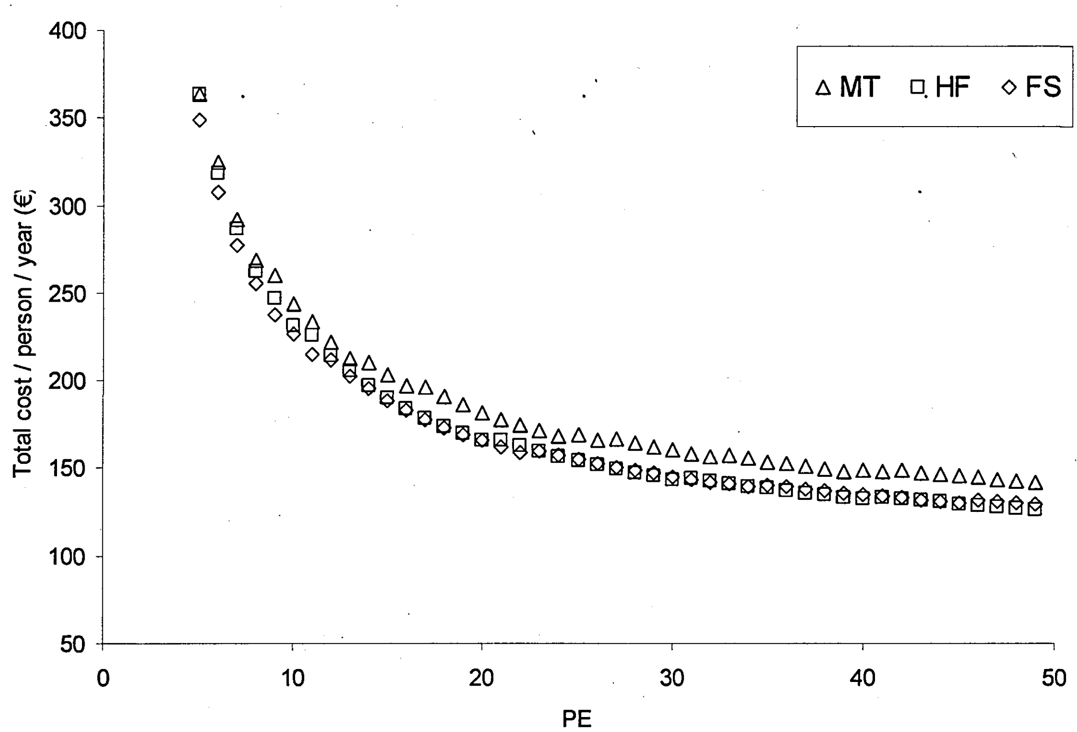

Figure 11 Annual cost per person as PE increases

5.3. Plant type

All plant types show a similar trend in terms of economies of scale but the absolute costs differ. The HF and FS systems have approximately the same overall cost the and MT system is consistently more expensive as an annual cost including CAPEX and OPEX (Figure 11). Operational costs are illustrated in Figure 12 and CAPEX including equipment purchase and installation are illustrated in Figure 13. The FS system provides the lowest operating but the highest purchase and installation costs and the reverse is true for the MT system. If the results for total cost are taken as absolute then the features selected for the FS system are clearly preferable to other types of plant but with package plants it is often the purchase cost that is the critical factor. Table 7 gives the estimated total production cost for each plant type. At 6 PE the difference in cost between the lowest cost (MT) and highest cost plant (FS) is €1459, equating to almost 30% higher price. The difference is even more marked at higher population ranges increasing to 50% increase for 200 PE plant.

A further factor particularly pertinent to package plant systems is operational complexity. The inclusion of a screen in the HF and MT systems may create reliability issues. Because the HF and FS systems require backflushing or relaxation a timer switch must be included in the which, coupled with the additional wear on the permeate pump from starting and stopping, makes maintenance of these systems more onerous. Diffuser cleaning is part of regular servicing of package plants, and the additional diffusers in the submerged systems will add an extra component to maintain. Another important factor is the time used for plant assembly both at the factory and on site. Additional components will add to this time, increasing the purchase cost.

The cost of installing primary settlement in a FS system has been accounted for within the installation and tank costs. Other factors mitigating against selection of a large plant have not been considered. If space is at a premium, which in gardens is likely to be the case, then the smaller systems are likely to be more attractive. Replacing the primary settlement of the FS system with a screen would reduce plant size and CAPEX but increase process complexity and OPEX.

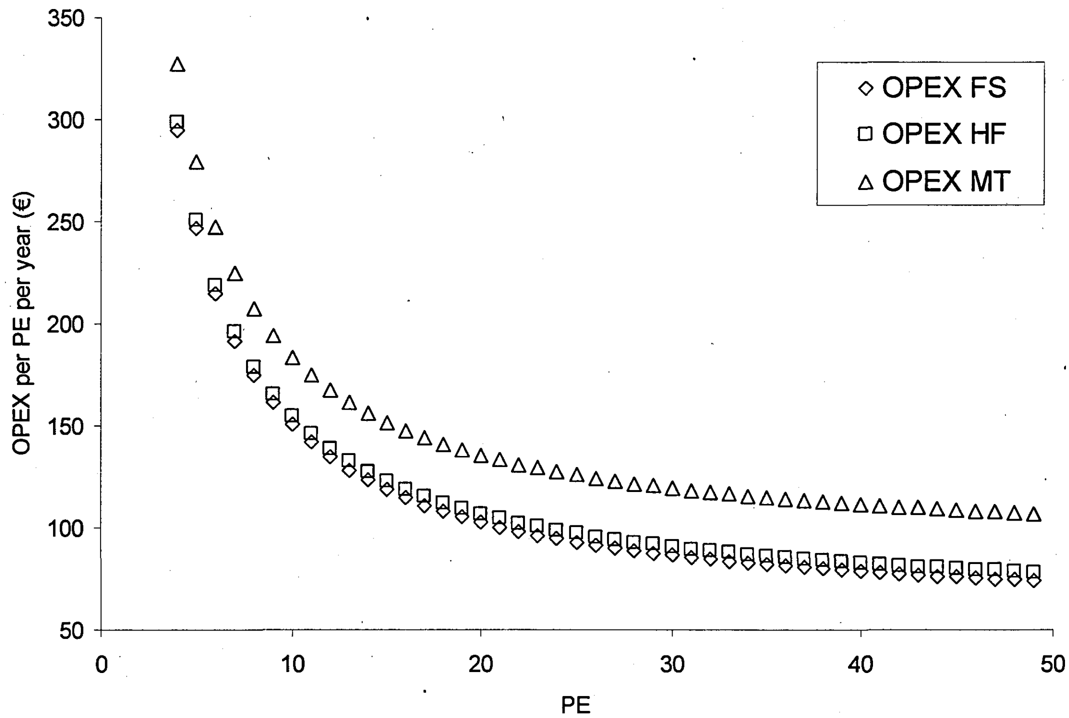

Figure 12 *OPEX vs. PE*

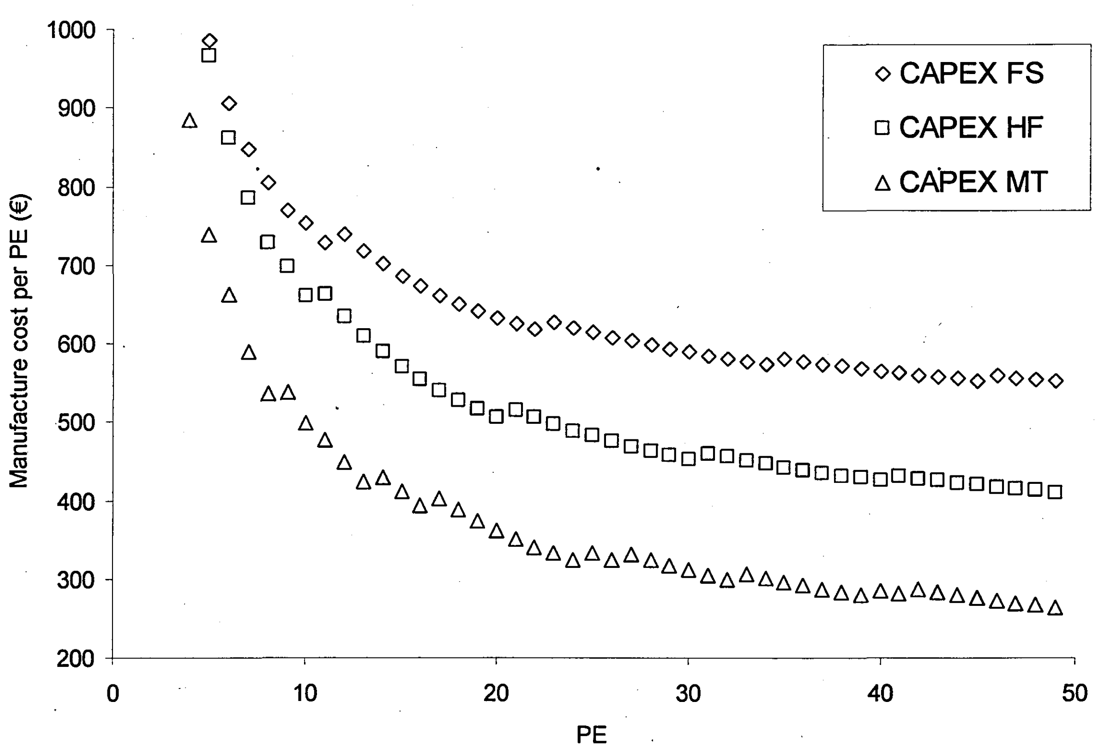

Figure 13 *CAPEX vs. PE*

Table 15. Absolute costs and power requirements

Plant Size PE	Plant Type	Power Cost €	Plant Capital cost €
6	FS iMBR	195	5433
	HF iMBR	219	5167
	MT sMBR	392	3974
20	FS iMBR	649	12645
	HF iMBR	731	10123
	MT sMBR	1307	7241
50	FS iMBR	2435	31822
	HF iMBR	1792	27659
	MT sMBR	4899	17201
100	FS iMBR	4869	60307
	HF iMBR	5485	51255
	MT sMBR	9799	30913
200	FS iMBR	9739	116208
	HF iMBR	10972	99378
	MT sMBR	19598	58591

5.4. Tank depth

The depth of package systems is an interesting feature since shallow systems are preferred for ease of installation whilst operational costs for these systems are higher due to decreased oxygen transfer efficiency. The change in costs for a 100 PE plant are illustrated in Figure 14. Taller units lead to reduced floor area under the membrane module which in turn reduces operational costs in submerged systems. A longer membrane path length in the MT system provides energy savings for an MT system, these savings are even greater than for submerged systems because liquid pumping costs are higher than blower power costs. The change in OPEX for the MT system is higher than for the FS or HF systems which have similar, shallow slopes (Figure 14).

Deeper excavations require greater and increasingly complex and costly side wall support. In this model this has been estimating by allowing for 45° slope on the side wall. The extra cost of this increasing additional excavation provides an estimate of the effect of depth on installation cost. In reality a number of shoring options are available to support side walls but assessing the most cost effective method of excavation is beyond the scope of this work. Figure 14 illustrates the trade off between installation and operational costs.

A further consideration of deep tank design is the possibility of encountering a high water table or bedrock. Either of these obstacles significantly increase the cost, complexity and time for installation so where possible customers wish to avoid this scenario.

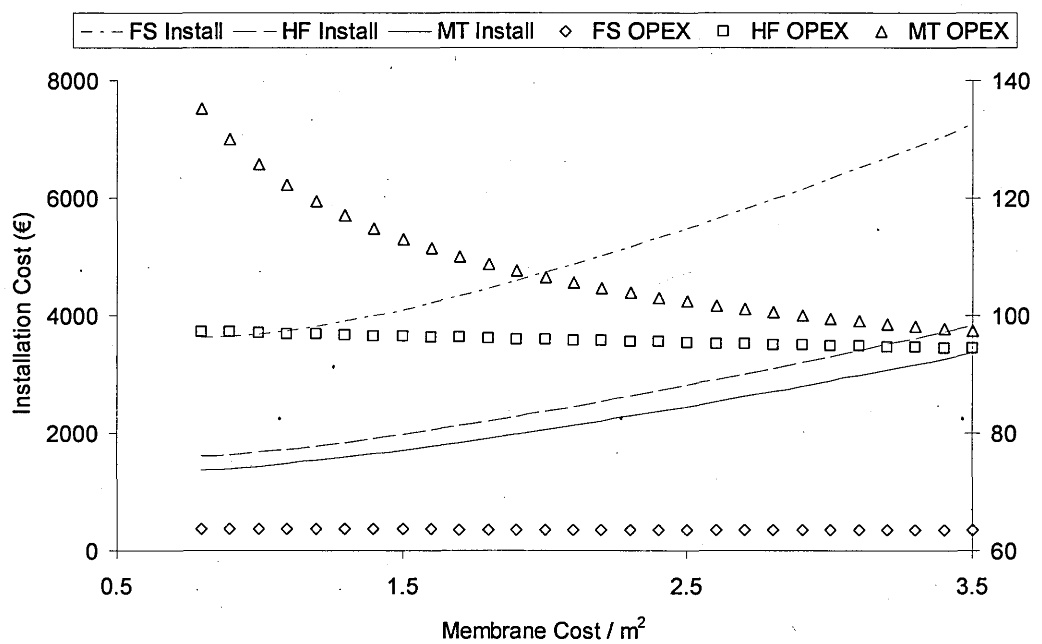

Figure 14 ***Change in operational and installation cost with tank depth (100 PE plant)***

5.5. Operational cost breakdown

It is of great interest to asses what portion of operational costs is attributed to which activity within the plant. With this knowledge it is possible to direct design economies to provide the greatest savings within the plant.

In a package plant there is a set charge to have the desludge lorry tanker come to the plant irrespective of plant size. As Figure 15 as plant size increases there is a greater population to bear the constant cost of desludging the plant. For all three designs this is the predominant operational cost at 6 PE but by the 50 PE plant size this cost is overtaken by power costs in all cases.

Unsurprisingly at large plant sizes power cost makes up the majority of operational cost. As the plant size approaches the size of a full scale plant costs associated with maintenance and sludge removal become less prevalent. Power requirements for both submerged MBR plants (FS and HF) are similar but there is a noticeably higher power requirement for the MT system. The reason for this is because of the increase in liquid pumping requirements. For the iMBR systems the pumping power makes up less than 1% of the total power demand for the plant whereas for the sMBR systems the pump makes up 56% of the plant power. The aeration requirements for the sMBR are only 45% of those for iMBR plants as there is no call for membrane aeration. This is in line with conventional MBR wisdom.

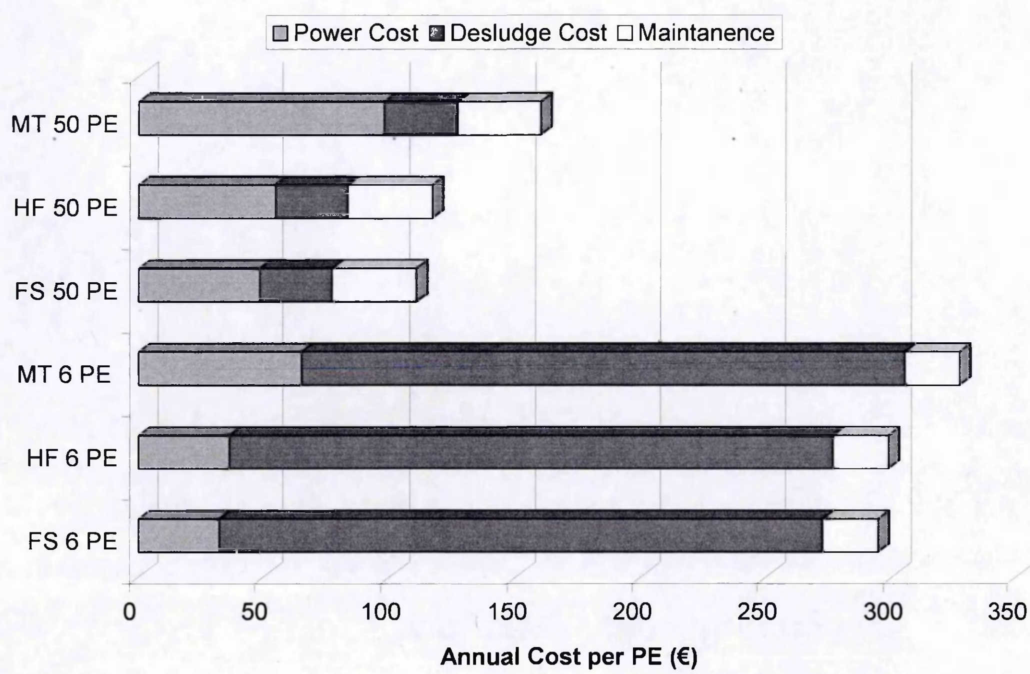

Figure 15 ***Specific contributions to OPEX***

5.6. Temperature effects

At lower temperatures plants must be made larger aeration demand increases. Figure 16 illustrates the change in manufacture cost due to temperature. The cost change can be attributed to the following:

- Reduced biological activity at lower temperature means that a longer HRT is required and hence a larger tank. This impacts not only on the tank purchase cost but also on the installation cost.
- Lower effective flux rate requires a greater membrane area to permeate the same flow
- Reduced oxygen transfer increases the aeration requirement to the plant. The increased air flow may require a larger blower, hence the steps in the curves illustrated in Figure 16

The HF MBR is most adversely affected by dropping temperature whilst the MT MBR is least affected. This difference is related to the operating flux and hence the increasing membrane cost. This issue highlights a potential problem with an HF design from the point of view of robustness.

The increased operational cost is related to a higher power requirement associated with the air blower. The effects on pumping power and chemical costs are negligible by comparison. Increased air requirement is a result of decreasing biological activity with colder temperature so more oxygen must be provided to produce the same removal efficiency. Decreasing flux at lower temperature (equation 15) also leads to higher membrane area requirement, and hence membrane aeration needs.

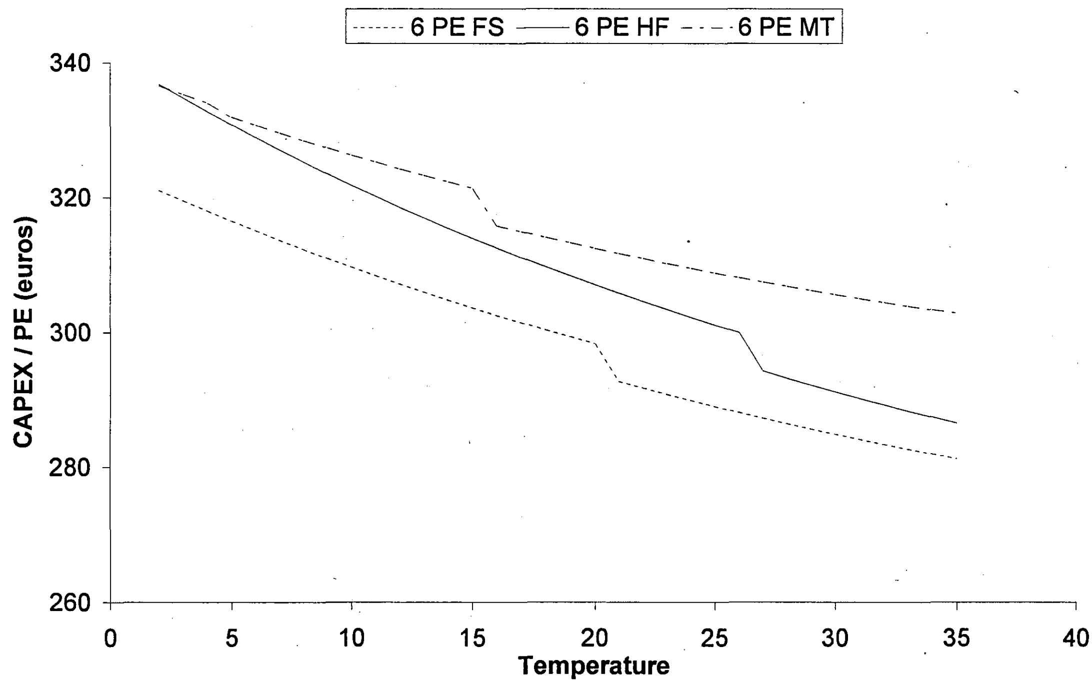

Figure 16 Manufacturing cost variation with temperature.

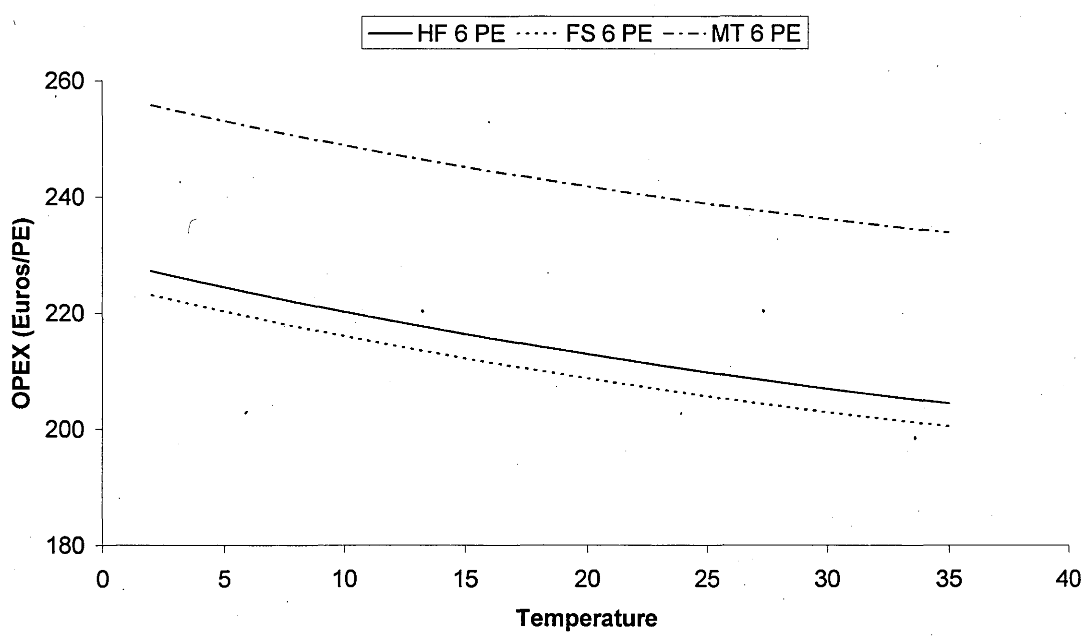

Figure 17 Operational cost variation with temperature.

5.7. Sensitivity analysis

This preceding was based on a number of key assumptions which can impact on the outputs of the analysis. To assess the sensitivity of the outputs to the assumptions made a number of plots have been produced.

5.7.1. Membrane cost

The membrane capital cost was assumed to be €150 per m^2, Figure 18 illustrates how changing this cost affects each plant. The sensitivity is greatest at the lowest flux, such that for an HF module operating at the lowest flux, and thus the highest membrane area, the cost of the membrane is the most significant proportion of the overall plant cost. In the case of the MT unit the membrane is a relatively small fraction of the overall plant cost and thus the impact of membrane cost is much lower. A breakdown of plant costs has been provided in Appendix E. The current trend is for decreasing membrane costs. This may ultimately lead capital cost trends of an HF unit more resembling that of an MT unit.

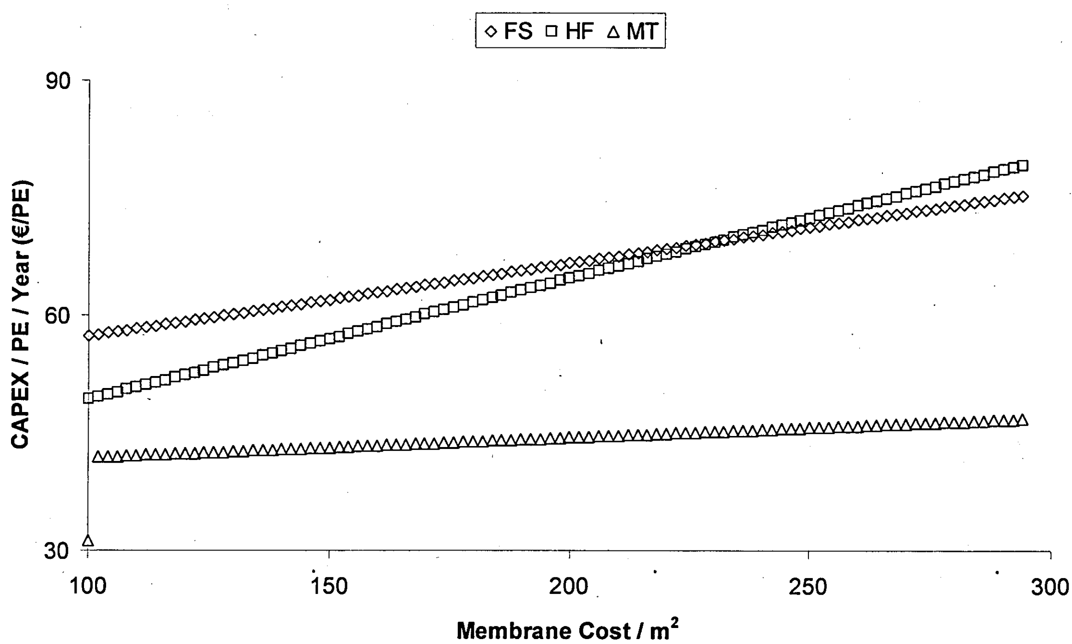

Figure 18 Impact of membrane cost on CAPEX

5.7.2. Blower and pump efficiency

Figure 7 illustrates that there is a significant variation in mechanical efficiency of rotating machines. Efficiency figures have been assumed for both the air blower and the pump (Table 11). If products having reduced or improved efficiency are used then this will affect the annual OPEX of the plant to an extent. Figure 19 illustrates how changing blower efficiency impacts on plant OPEX. The immersed plants are more susceptible to the change because the overall aerating requirement is higher for these plants. Conversely because the sMBR power requirements are dominated by the pumping necessary to scour the membrane the pump efficiency has the biggest impact on the plant operational cost in this case (Figure 20). The sMBR is still affected by changes in blower efficiency since biological aeration demand is responsible for around 40% of the overall power cost, hence the equipment selection for an sMBR should be undertaken with great care to avoid excessive OPEX from inefficiencies.

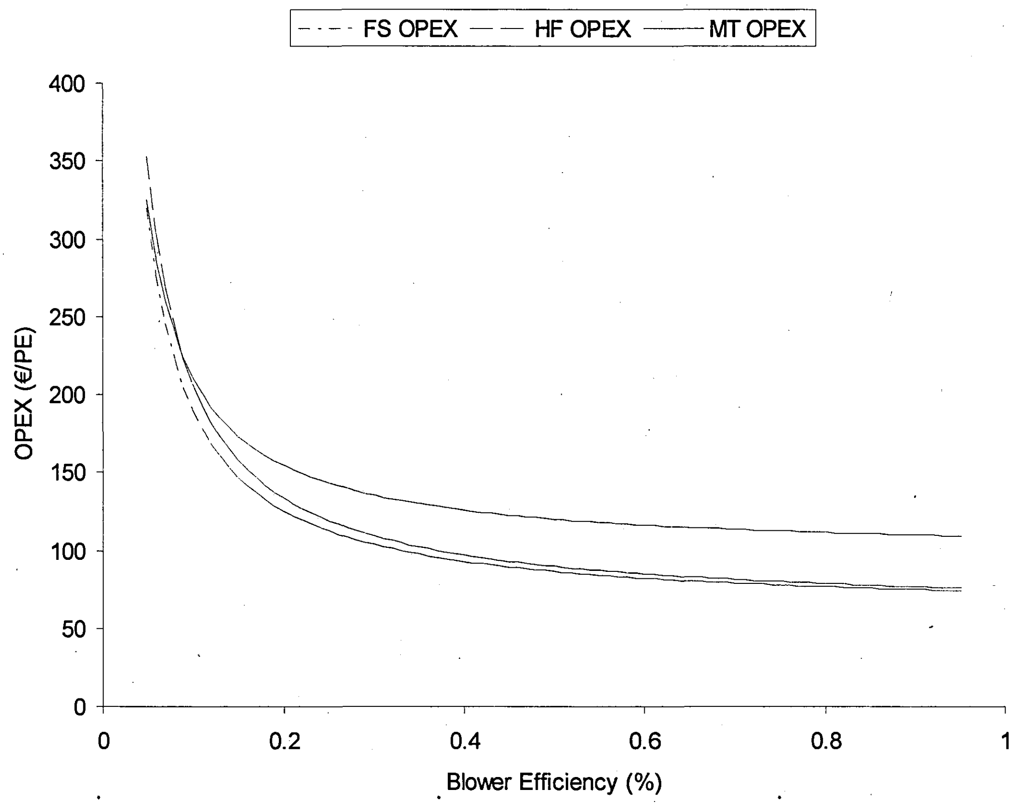

Figure 19 OPEX against blower efficiency

5.7.3. Oxygen transfer efficiency

At full scale significant investments have been made to improve the oxygen transfer into solution. The cost of aerating a plant makes up the largest portion of OPEX, particularly for an iMBR. Figure21 illustrates OPEX savings with increasing oxygen transfer efficiency. Both the iMBR systems reach a plateau where more oxygen is used for membrane aeration than is required for biological maintenance. The sMBR system continues to improve with improving OTE. Package treatment systems do not generally have having an aeration system designed specifically for technology employed, and diffuser layout is primarily governed by what is available. As a result the OTE is usually lower than expected in a full scale system.

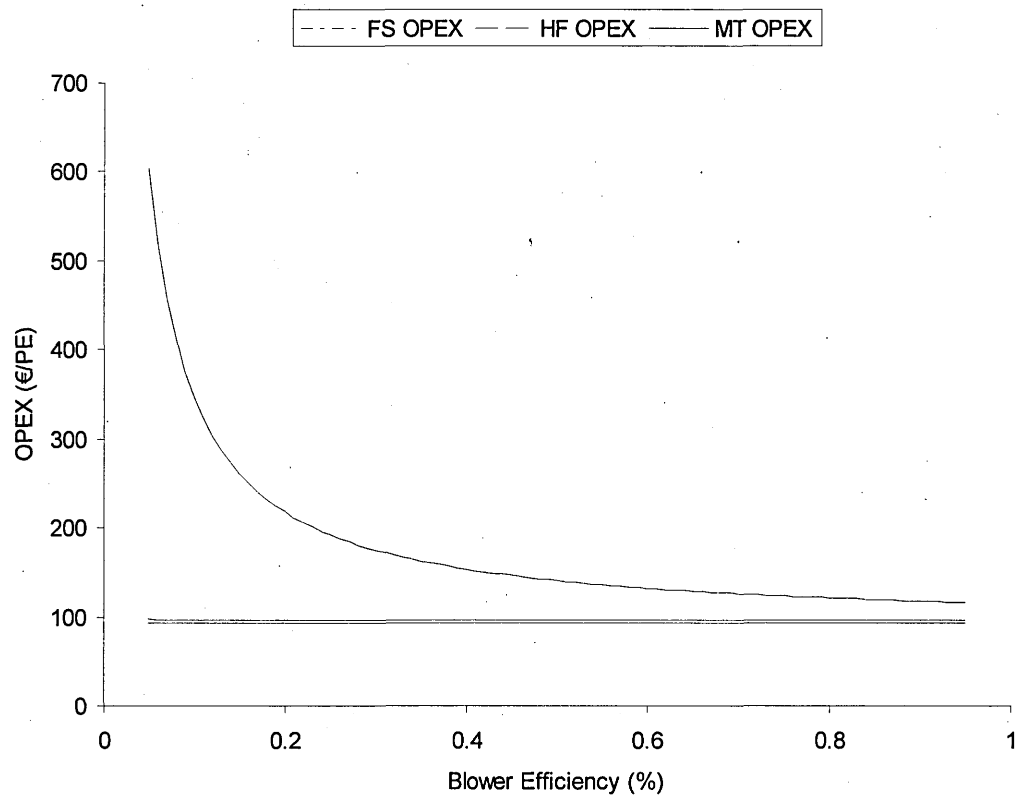

Figure 20 ***OPEX against pump efficiency***

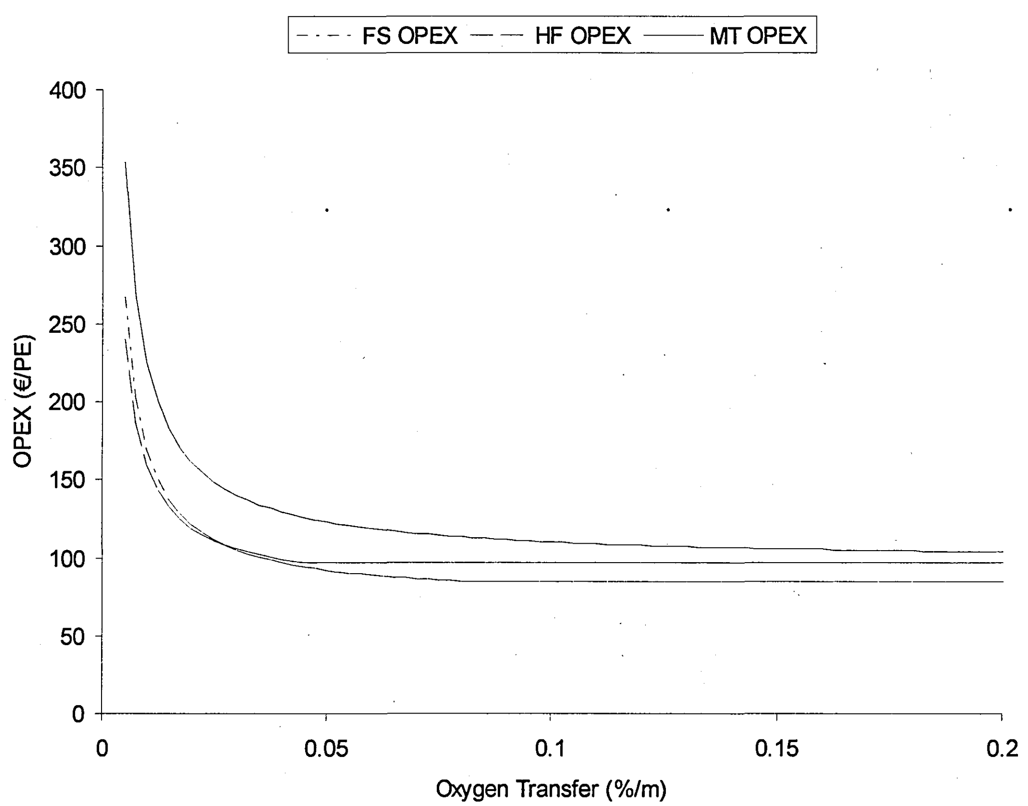

Figure 21 ***Impact of oxygen transfer efficiency***

6. Conclusions

Based on the assumptions made in this study:

- A single-household package plant MBR can be produced at a capital cost which is within the boundaries of commercially-available package plants, albeit at the high end of the range (i.e. €1728 to €5760; €4400 on average).
- Economies of scale exist from 6-20 PE plants; above this size the change in specific cost with size is low. This is mainly due to the impact of desludging on opex is high for small plants, but are overtaken by other opex components (such as power demand) above ~20 PE.
- The operational costs of an MBR significantly exceed those of more conventional package plant designs, primarily due to energy demand. MBRs can be as much as 20 times more energy intensive than conventional package plants.
- Of those designs considered, the ones that are most expensive plants to produce (i.e. above €12k for a 20 PE plant) also provide the lowest operational costs since they incorporate design elements (submerged membrane operation) which make the system more efficient.
- Although the lifetime cost of the sidestream system is high compared to that of the submerged system the nature of package plant market, being driven by CAPEX, may make the low plant capital cost and simple operation the most attractive option.
- The market for package MBRs is significantly influenced by the recycling potential of the effluent produced. Further research is needed to asses the financial and environmental benefits offered by such a technology for recycling duties specifically.

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Appendix A – Related Legislation and Standards

Urban Wastewater Treatment Directive (UWWTD)

Scottish Water Design Standards & Specification

Wastewater Treatment Works For Adoption, version 1, April 2003 – Dwr Cymru (Welsh Water)

Guide And Procedure For The Adoption Of Package Sewage Treatment Plants, Draft – Wessex Water

Code Of Practice: Flows and Loads – Sizing Criteria, Treatment Capacity for Small Wastewater Treatment Systems (Package Plants) – British Water

BS 6297:1983 – Design and installation of small sewage treatment works and cesspools

prEN 12566-3 – Small wastewater treatment systems for up to 50 PT – part 3:Packaged and/or site assembled domestic wastewater treatment plants

BS EN 12255-1:2002 Wastewater treatment plants – Part 1: General construction principals

BS EN 12255-4:2002 Wastewater treatment plants – Part 4: Primary settlement

BS EN 12255-6:2002 Wastewater treatment plants – Part 6: Activated sludge processes

BS EN 12255-7:2002 Wastewater treatment plants – Part 7: Biological fixed-film reactors

BS EN 12255-10:2002 Wastewater treatment plants – Part 10: Safety principals

BS EN 12255-11:2002 Wastewater treatment plants – Part 11: General data required

Appendix B – Package Plant Company Websites

Balmoral Tanks is a medium sized company based in Scotland. The company business is primarily the manufacture of tanks.
http://www.balmoral-group.com/tanks/environmental.asp

Biotank is a small company based in England. The company manufacture and distribute sewage treatment products.
http://www.biotank.co.uk/

Biwater is a worldwide company specialising in water and wastewater treatment. The company also offers consultancy services.
http://www.biwater.co.uk/

Busse are independent consulting engineers based in Germany. This company has been included because they manufacture a single house MBR unit.
http://www.busse-gmbh.de/en/development/biomir/structure-of-plant.htm

Clearwater are a UK based company providing package treatment plants, oil separators, pump stations, tanks and filters
http://www.clearwaterpolcon.co.uk/

Conder is manufacture GRP pollution control products including septic tanks, oil separators and sewage treatment plant.
http://www.conderproducts.com/main.htm

Copa is a UK based company which trades internationally and has two main areas to its business - CSO technology and process technology for municipal wastewater treatment.
http://www.copa.co.uk/

Hepworth is UK based company producing clay and plastic drainage products.

http://www.hepworthdrainage.co.uk/

Kee is an international company specialising in the design manufacture and installation of wastewater treatment processes.
http://www.keeprocess.com/html/wastewater_.html

Klargester Environmental is an international supplier of specialist pollution control equipment for off-mains sewage solutions, fuel and oil separation and rainwater recycling
http://www.klargester.com/

Titan Pollution Control manufactures GRP pollution control products including septic tanks, oil separators and sewage treatment plant.
http://www.titanpc.co.uk/

WPL Ltd specialise in the manufacture of ecological wastewater treatment products, for both commercial and domestic markets.
http://www.wpl.co.uk/

Appendix C – Example Calculations for Plant Comparison

Equations taken from BS 6297:1983 *(section 11.3.2)*

The design equation used for primary clarifier is:

C	=	**$180\ P^{0.85}$**
Where C	=	Capacity (l)
P	=	Design population

The results for the populations are:

6 POP	=	825 litres
50 POP	=	5000 litres

The sludge storage capacity was also calculated to determine the additional volume required above the settling volume (calculated above). This was performed using the basis of 10L of sludge per head per week. Assuming a de-sludging interval of 6 months, the sludge storage requires a capacity of:

6 POP	=	1560 litres
50 POP	=	13000 litres

When these two capacities above are combined for the respective PE size they equal the total tank volumes of:

6 POP	=	$2.4m^3$
50 POP	=	$18m^3$

Secondary Clarifier Sizing

Equations taken from BS 6297:1983 *(section 11.5.2)*

The design equation used for secondary clarifier is:

C	=	**$135\ P^{0.85}$**
Where C	=	Capacity (l)
P	=	Design population

Aeration Chamber

Requirements			Kinetic Constants			
BOD	300	mg/l	K_1	20	Kg BOD utilised/KgVSS.day	
Input Flowrate	1.2	m^3/day	K_s	400	mg BOD_u/l	
TKN	40	mg/l	K_d	0.1	day^{-1}	
Phosphorus	10	mg/l	Y	0.5	Kg VSS / Kg BOD utilised	
Aeration Tank X	2500	mg/l	$(u_m)_N$	0.4	d^{-1} at 20deg and pH 8.2	
Recycle X_R	10000	mg/l	K_N	2	mg NH_4^+-N/l	
			Y_N	0.05	Kg VSS/Kg NH_4^+-N Nitrified	
Output BOD <	10	mg/l				
N_e <	20	mg/l	Tank Temp	5	degrees	
			pH	8		
Output Values						
Nitrification Requirement			**NH^+ - N**			
New $(u_m)_N$	0.125	d^{-1}	NH^+ - N Removed	0.011	Kg/day	(from assimulation)
Sludge Age_{min}	8.0	Days	Residual NH^+ - N	30.5	mg/l	
Sludge Age	9	Days	Biomass Conc	17.0	mg VSS/l	
			Nitrifying Fraction	0.007		OK
Effluent Soluble BOD			Effluent NO_3^- - N	10.5	mg/l	
S_e (Output BOD)	8.7	mg BOD/l				
C Oxidation Check			**Recycle Ratio, R**			
Sludge Age $_{min}$	0.24	days	R	0.33		
Aeration Tank Volume			**Effluent Phoshorus**			
V_a	0.33	m^3	Removed by	0.0019	Kg/day	

			assimilation			
			Effluent P	8.45	mg/l	
Food/Microorganism Ratio						
F/M check =	0.44		**Oxygen Utilisation Rate**			
			Oxygen Required	0.28	Kg O_2/day	
Heterotrophic Sludge Production H.S.P.			Volume of Air Required	1.15	m^3/day	
H.S.P.	0.093	Kg VSS/day	Air (accounting for Diff)	11.47	m^3/day	based on 10% diffuser efficency

Appendix D – Biokinetic Parameters

Parameter	Value	Units	Reference	Process	Wastewater tested and additional comments
$k_{d,n}$	0.21	per d	Dinçer and Kargi, 2000	ASP with denitrification.	Synthetic wastewater (100:100 COD:NH4-N)
k_{dn}	0.05-0.15	per d	Metcalf & Eddy, 2003		0.08 typical
k_e	0.06-0.2	per d	Metcalf & Eddy, 2003		0.12 typical
k_e	0.067	per d	Yenkie (1992)	High rate CAS	22°C Synthetic wastewater (1090:872 COD:BOD), SRT 0.3d
k_e	0.050	per d	Fan *et al*, 1996	MBR	municipal wastewater, 30°C, 411-72 COD: 26-53 NH_4-N, 20d SRT
k_e	$0.85\theta_x^{-0.62}$	per d	Huang *et al*, 2001	MBR	Domestic wastewater (~250:20:170 COD:NH3-N:SS) SRT 5-40d,
k_e	0.023	per d	Liu *et al*, 2005	MBR	Synthetic wastewater (220-512mg/L COD, 36-72mg/L NH_4-N) infinite SRT
k_e	0.08	per d	Wen *et al*, 1999	MBR	30°C Urban wastewater (~500 COD) SRT 5-30d
k_e	0.025-0.075	per d	Xing *et al*,	MBR	Variable (30-2234 mg/l COD) Municipal wastewater, SRT 5-

			2003		30d
k_e	0.048	per d	Yilditz *et al*, 2005	MBR	Synthetic wastewater, 26°C, 1090 COD, SRT 0.3d
$k_{d,n}$	0.12	per d	Harremoes and Sinkjaer, 1994	ASP with denitrification.	municipal wastewater, 20°C, 397-256 COD: 40-35 TKN, 18-21d SRT
K_n	0.1-0.4	g/m^3	Harremoes and Sinkjaer, 1994	ASP with denitrification.	municipal wastewater, 20°C, 397-256 COD: 40-35 TKN, 18-21d SRT
K_n	0.5-1	g/m^3	Metcalf & Eddy, 2003		0.74 typical
K_n	0.1-0.15	g/m^3	Manser *et al*, 2005	MBR & CAS in parallel	Domestic Wastewater (quality not given) SRT 20d
K_n	0.85	g/m^3	Wyffels *et al*, 2004	MBR with low DO	30°C, ~ Sludge digester supernatant 605:931 COD:TAN (Total ammonium nitrogen), SRT >650d
K_n	0.01-0.34	g/m^3	Groweneweg *et al*, 1994	Nitrifying chemostat	Synthetic wastewater, 30°C, 392 mg NH_x-N/L. Values were dependant on pH, Temp and bacterial species
K_s	5-40	g/m^3	Metcalf & Eddy, 2003		20 Typical
K_s	80	g/m^3	Yenkie (1992)	High rate CAS	22°C Synthetic wastewater (1090:872 COD:BOD), SRT 0.3d
K_s	192	g/m^3	Yilditz *et al*,	MBR	Synthetic wastewater, 26°C, 1090 COD, SRT 0.3d

			2005		
Y	0.3-0.5	g VSS/ g bCOD	Metcalf & Eddy, 2003		0.4 typical
Y	0.44	per d	Yenkie (1992)	High rate CAS	22°C Synthetic wastewater (1090:872 COD:BOD), SRT 0.3d
Y	0.61	g VSS/ g COD	Fan *et al*, 1996	MBR	municipal wastewater, 30°C, 411-72 COD: 26-53 NH_4-N, 20d SRT
Y	0.28-0.37	g VSS/ g COD	Huang *et al*, 2001	MBR	Domestic wastewater (~250:20:170 COD:NH3-N:SS) SRT 5-40d
Y	0.288	g VSS/ g COD	Liu *et al*, 2005	MBR	Synthetic wastewater (220-512mg/L COD, 36-72mg/L NH_4-N) infinite SRT
Y	0.40-0.45	g VSS/ g BOD	Lübbecke *et al*, 1995	MBR	Synthetic wastewater (8500-17600mg/L COD, 36-72mg/L NH_4-N), SRT 1.5-8d
Y	0.56	g VSS/ g COD	Wen *et al*, 1999	MBR	30°C Urban wastewater (~500 COD) SRT 5-30d
Y	0.25-0.40	g VSS/ g COD	Xing *et al*, 2003	MBR	Variable (30-2234 mg/l COD) Municipal wastewater, SRT 5-30d
Y	0.58	g VSS/ g COD	Yilditz *et al*, 2005	MBR	Synthetic wastewater, 26°C, 1090 COD, SRT 0.3d
Y_n	0.34	g TSS/ g N	Dinçer and	ASP with	Synthetic wastewater (100:100 COD:NH4-N)

			Kargi, 2000	denitrification.	
Y_n	0.16	g VSS/ g N	Harremoes and Sinkjaer, 1994	ASP with denitrification.	municipal wastewater, 7.5°C, 397-256 COD: 40-35 TKN, 18-21d SRT
Y_{obs}	0.31-0.36	g VSS/ g COD	Tao *et al*, 2005	3 MBR plants in parallel	SRT 14-28d, Settled sewage, 265 COD
Y_{obs}	0.11	g VSS/ g COD	Liu *et al*, 2005	MBR	Synthetic wastewater (220-512mg/L COD, 36-72mg/L NH_4-N) infinite SRT
Y_{obs}	0.16-0.38	g VSS/ g COD	Wen *et al*, 1999	MBR	30°C Urban wastewater (~500 COD) SRT 5-30d
μ_{max}	3-13.2	per d	Metcalfe and Eddy, 2003		6 Typical
μ_{max}	0.125	per d	Yenkie (1992)	High rate CAS	22°C Synthetic wastewater (1090:872 COD:BOD), SRT 0.3d
μ_{max}	3.24	per d	Yilditz *et al*, 2005	MBR	Synthetic wastewater, 26°C, 1090 COD, SRT 0.3d
μ_{max}/K_s	0.001-0.01	per d	Wen *et al*, 1999	MBR	Lower than ASP. 30°C Urban wastewater (~500 COD) SRT 5-30d
$\mu_{n,max}$	0.1-0.2	per d	Fan *et al*, 1996	MBR	municipal wastewater, 30°C, 411-72 COD: 26-53 NH_4-N, 20d SRT
$\mu_{n,max}$	0.2-0.9	per d	Metcalf &		0.75 typical

$\mu_{n,max}$	0.2-0.9	per d	Metcalf & Eddy, 2003		0.75 typical
$\mu_{n,max}$	2.02	per d	Wyffels *et al*, 2004	MBR with low DO	30°C, ~ Sludge digester supernatant 605:931 COD:TAN (Total ammonium nitrogen), SRT >650d

Appendix E – CAPEX Breakdown

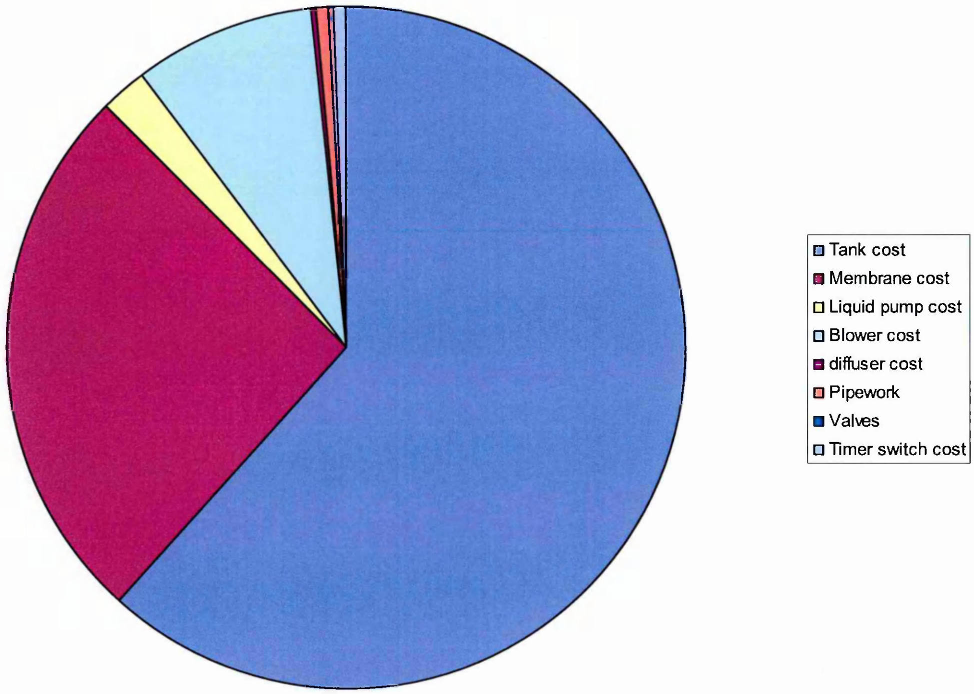

Figure 22 Component contribution to CAPEX FS

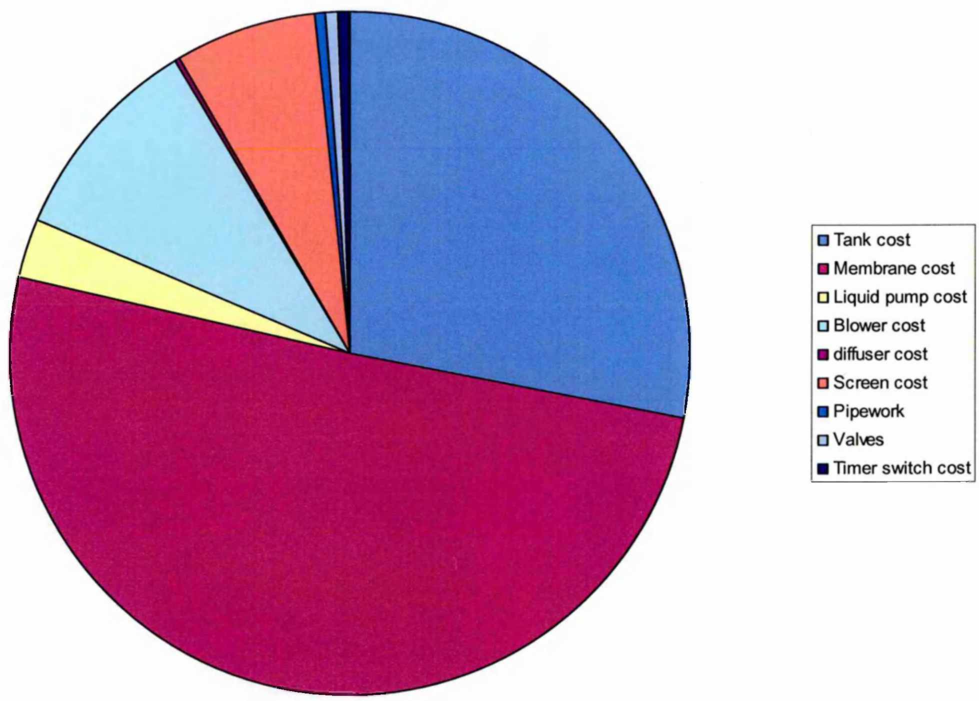

Figure 23   Component contribution to CAPEX HF

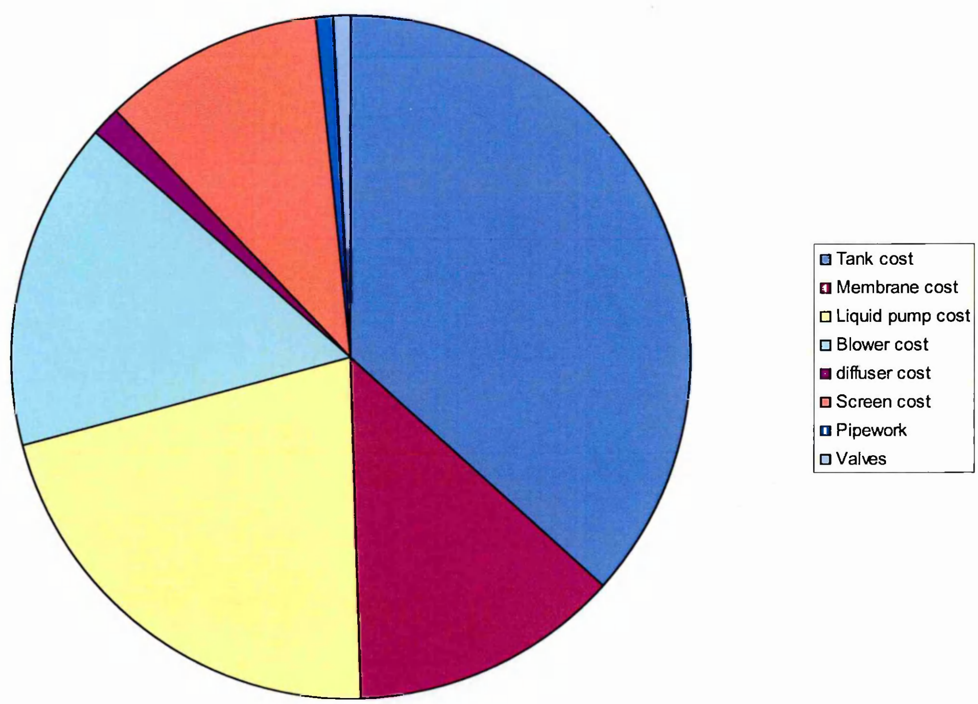

Figure 24   Component contribution to CAPEX MT