SIXTH FRAMEWORK PROGRAMME

Project no.: 018480

EUROMBRA

Membrane bioreactor technology (MBR) with an EU perspective for advanced municipal wastewater treatment strategies for the 21st century.

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D16 – Cost analysis, literature data (incl. pilot plant trials conducted by partners)

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PU	Public	X
PP	Restricted to other programme participants (including the Commission Services)	
RE	Restricted to a group specified by the consortium (including the Commission Services)	
CO	Confidential, only for members of the consortium (including the Commission Services)	

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Rationale

Previous tasks under this work package have established that a simple model can be used to evaluate costs associated with an MBR. This has been through acquisition and collation of heuristic data (D1) and processing and collation of this data in the form of simple unifying normalised parameters (D9). From the outline analysis presented in D9 appropriate ranges of operational parameter values have been established for the two main immersed MBR configurations (flat sheet and hollow fibre) based on both literature and pilot plant data provided by the partners. **The outputs attributed to D16 have therefore already been provided in D9**.

There are several limitations of the model, relating to its basis on heuristic information which thus refers to a wide range of operating conditions, plant size, feedwater quality and membrane process suppliers. This being the case, there is obviously merit in reducing the scope of the model. This has been approached in two ways for D16:

- A full cost model of a simple (package plant) MBR plant, and
- Correlation of flux against aeration.

Objectives

The aims of this Work Package are:

a) to ascertain the overall costs associated with individual modes of operation of existing commercial membrane bioreactor technologies which have studied at pilot scale by several of the partners,
b) to provide information on overall costs, as well as parameter values for appropriate base operating conditions, to be employed throughout the subsequent experimental programme, and
c) to extend the analysis to encompass literature data and data arising from other WPs

Status

Package plant MBR costs

The costs of a small (package plant) MBR have now been evaluated according to the simple cost model outlined in previous a WP report (D9) supplemented with cost data for capital items. The analysis and results have been submitted for publication in a peer reviewed journal. The submitted paper is reproduced in Annex 1 and is summarised below.

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 vii 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.

MBR aeration costs vs. aeration

Introduction

As already stated Membrane aeration is a critical component in the operation of an immersed MBR, contributing significantly to the energy demand. It is generally recognised that flux and/or permeability correlates with membrane aeration. However, the precise quantitative relationship between the two parameters is unclear. Data available from both from the scientific peer-reviewed literature and from full-scale plant are briefly reviewed below.

Literature studies, bench-scale studies

A number of correlations of flux with membrane aeration have been reported in the literature. It has been demonstrated (Ueda et al, 1997; Bouhabila et al, 1998; Liu et al, 2000; Le Clech et al, 2002; Sofia et al, 2003) that there exists a critical aeration rate above which there is no further impact on membrane permeability. The values reported for this parameter are as aeration intensity: air flow/unit floor area in Nm^3 air per m^2 cross section per unit time, rather than specific aeration demand (Nm^3 air per m^2 membrane area per unit time). Example values taken from the literature are shown in Table 1.

Table 1 Laboratory-scale date, critical aeration rate

Reference	*Critical aeration rate ($m^3\ m^{-2}\ s^{-1}$)*	*Flux ($l\ m^{-2}\ h^{-1}$)*	*MLSS ($g\ l^{-1}$)*
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

The data reported in Table 1 provide no clear correlation between flux and approach air velocity. A number of factors may impact upon the relationship, including:

- sludge quality
- aerator design
- membrane material and configuration
- membrane module characteristics

Experiments conducted by Liu et al (2003) showed MLSS to have some effect on the critical aeration of the biomass. Previous research by Bouhabila et al (1998) appears to contradict this. These authors tested a range of sludge concentrations and found the critical air flow to be independent of sludge concentration. In experiments conducted by Sofia et al (2004) coarse and fine bubble aeration over the membrane surface were compared. The results showed fine bubble aeration to prolong the cleaning interval to almost 8 months, whereas coarse bubble aeration only provided a cleaning interval of 4 weeks. This is somewhat contradictory to the perceived wisdom on membrane aeration.

Literature studies, pilot-scale and full-scale data

Pilot-scale and full-scale data is reported in Judd (2006). Data, expressed as specific aeration demand, are highly scattered and have been summarised in previous WP reports. However, specific data for the most widely reported technology can be extracted.

There remains the fundamental question as to how membrane flux and/or permeability correlates with membrane aeration and the factors contributing to the relationship. Central to this is:

- the representation of aeration itself: aeration intensity (i.e. approach air velocity) or specific aeration demand
- the precise definition of "sustainable flux/permeability"

In the following section a generic theoretical representation of aeration demand is represented, with specific reference to an immersed HF MBR (though the same approach can be applied to a FS module). Central to the theoretical development is that flux is linearly related to air flow velocity as measured within the module. It is further assumed that

Theory, HF modules

Packing density given by ratio of fibre surface area to volume

$$\phi = \frac{A_f}{V} \qquad (1)$$

where A_f = surface area of fibres = $N\pi d_f$ (2)

V = module volume

A_v = volume occupied by fibres = $N\pi d_f^2/4$ (3)

So: $$\frac{A_V}{A_f} = \frac{d_f}{4} \qquad (4)$$

Thus $$A_x = \frac{4V - d_f A_f}{4L} \qquad (5)$$

where A_x = free x-sectional area

So: $$Q_A = \frac{3600U}{4L}\left(4V - d_f A_f\right) \quad 6$$

where Q_A = aeration rate in m^3/hr
U = air flow velocity in channels (m/s)

$$SAD_m = \frac{3600U}{4L}\left(\frac{4V}{A_f} - d_f\right) \quad 7$$

where SAD_m = aeration demand with respect to fibre area

Substituting for V/A_f and normalising against flux:

$$SAD_p = \frac{3.6x10^6 U}{4LJ}\left(\frac{4}{\phi} - d_f\right) \quad 9$$

where SAD_p = aeration demand with respect to permeate volume

Evidence suggests that J is a linear function of aeration intensity:

$$J = mU + c \quad 10$$

where m and c are empirical constants

So: $$SAD_p = \frac{3.6x10^6 U}{4L(mU+c)}\left(\frac{4}{\phi} - d_f\right) \quad 11$$

Aeration energy demand in kWh per m^3 permeate is then given by

$$E_A = 0.0303 SAD_p \gamma \frac{\left(\frac{10x+101}{101}\right)^{\left(1-\frac{1}{\gamma}\right)} - 1}{(\gamma-1)\zeta} \quad 12$$

where γ = aerator constant = 1.4
ζ = blower efficiency = 0.5
x = aerator depth = 3 m

So $$E_A = \frac{Uk}{L(mU+c)}\left(\frac{4}{\phi} - d_f\right) \quad 13$$

where $k = \frac{0.0303x3.6x10^6}{4}\gamma \frac{\left(\frac{10y+101}{101}\right)^{\left(1-\frac{1}{\gamma}\right)} - 1}{(\gamma-1)\zeta}$ and is thus constant for a given system

Now, commercial technical data for available membrane modules (Judd, 2006) suggests that, for packing density and fibre diameter respectively in m^{-1} and m:

$$\frac{1}{\phi} \approx g d_f + 0.001 f \quad 14$$

where f = 0.7-1.7 and g = 0.9-1.1

Thus $$E_A = \frac{Uk}{L(mU+c)}\left((4/g-1)d_f + 0.004 f\right) \quad 15$$

For the four main MBR HF membrane suppliers, $g = 0.89$ ($R^2 = 0.97$) and $f = 1.7$, and thus:

$$E_A = \frac{Uk}{L(mU + c)}\left(3.5d_f + 0.0068\right) \qquad 16$$

Energy demand can thus be presented as a function of aeration velocity and fibre diameter for existing commercial systems, provided that Equation 10 holds. According to the data from Table 1, ignoring outliers:

$$J = 342U + 9 \qquad R^2 = 0.78\text{, 7 data points} \qquad 17$$

For full-scale and pilot-scale data from Zenon plant provided in previous reports:

$$J = 553U - 18 \qquad R^2 = 0.97\text{, 6 data points} \qquad 18$$

Results and discussion

Substituting the values for m and c into Equation 16 produces markedly differing correlations. The pilot and full scale data (Fig. 2) suggest that no flux is possible without an air flow velocity of ~0.032 m/s. Beyond this value energy demand peaks sharply and then decreases with aeration rate. This is clearly counter intuitive and suggests that:

- Equation 18 does not adequately describe the relationship between flux and air flow velocity, and
- these plant are probably operating sub-optimally, particularly for plants operating at lower fluxes; operation at low flux would allow reduction in the aeration rate, but this is clarly not practical for full-scale plants.

Only data beyond an air flow velocity of ~0.041 m/s (when the decrease in energy demand with air flow is at its shallowest) would appear to be valid.

For the bench-scale data (Fig. 1), which conversely indicate that a flux of 9 LMH is sustainable at zero aeration, the energy demand increases with aeration. Clearly, the trend depicted is more in keeping with expectation, and probably reflects the nature of the studies reported: the data refer to optimised conditions of flux and aeration.

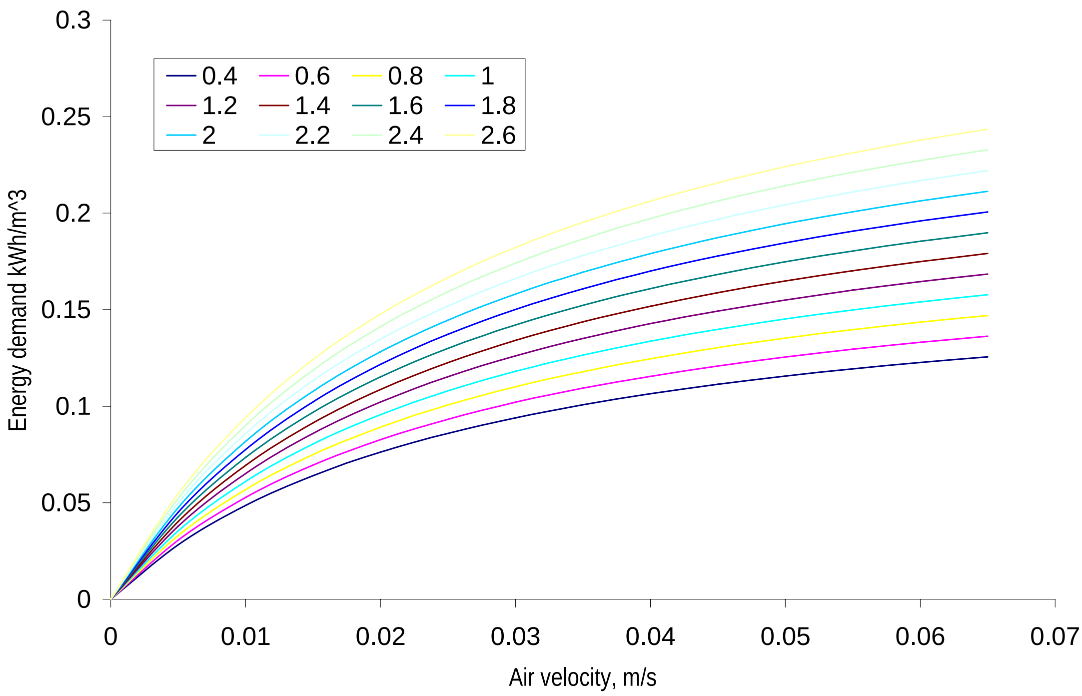

Fig. 1 Energy demand vs. air velocity based on bench-scale data (Equation 18); d_f = 0.4-2.6 µm and Equation 14 applies.

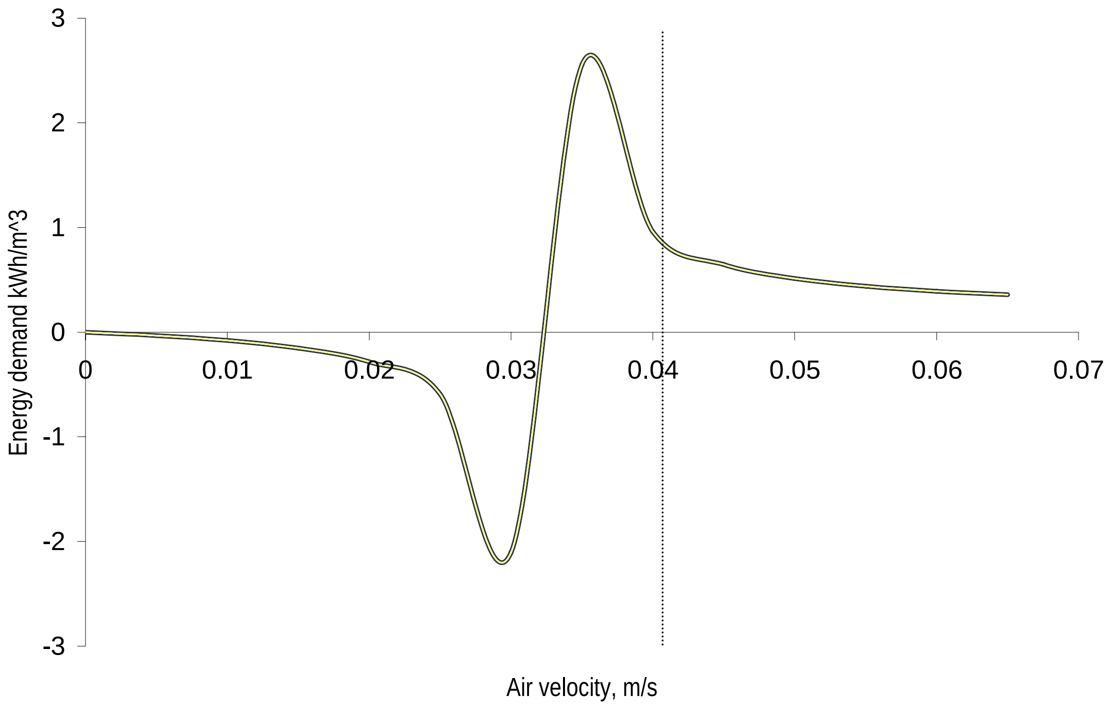

Fig. 2 Energy demand vs. air velocity based on pilot and full-scale data from *Zenon* plant (Equation 18); $d_f = 1.9$ µm

Future model development and ratification

The full cost model provides a useful starting point for assessing whole life costs and so comparing the costs of MBRs based on the three different configurations. The model has now been formatted to comprise a simple colour-coded Excel spreadsheet which can be placed on the web site and made available to all partners.

According to this simple theoretical development, it is evident, from Equation 16, that for a given commercial membrane the energy demand is critically dependent on the function $m/(mU+c)$. It is important, therefore, to acquire as much data from pilot plant studies as possible which relates specifically to *optimum* membrane flux and aeration values. Of key interest is the nature of c, and in particular whether this value is positive or negative. A positive value (implying a threshold flux) indicates that specific aeration demand always increases with air flow rate, whereas for a negative value the reverse is true and a threshold aeration rate must exist. Clearly, the former is intuitively correct, and suggests that some plants providing the data from which Equation 18 was generated operate at a higher air rate than necessary.

Annex 1 The cost of a package plant membrane bioreactor

A.1 NOMENCLATURE

C – Cost (€)
C – dissolved oxygen concentration (kg/m^3)
C^* – saturated oxygen concentration (kg/m^3)
c_c – chemical concentration (g/m^3)
C_e – effluent COD concentration (g/m^3)
C_i – inlet COD concentration (g/m^3)
C_{sup} – biomass supernatant COD concentration (g/m^3)
d – tank diameter (m)
g – acceleration due to gravity (m/s^2)
H – dig height (m)
h – water height (m)
h_m – membrane module height (m)
I_d – desludge interval (years)
k_e – endogenous decay coefficient (1/d)
k_La – oxygen transfer coefficient (1/d)
m_0 – biological oxygen requirement (g/d)
N – number of service/desludge visits per year
n – physical membrane cleans per chemical clean
N_e – effluent TKN concentration (g/m^3)
N_i – inlet TKN concentration (g/m^3)
NO_x – oxidised ammonia (g/m^3)
OTE – oxygen transfer efficiency (transfer/m tank depth)
$OTR_{cleanwater}$ – oxygen transfer rate for clean water
$P_{A,1}$ – inlet air pressure (Pa)
$P_{A,2}$ – outlet air pressure (Pa)
PE – population equivalent
P_p – pump head (m)
P_x – sludge production (g/d)
Q – system flowrate (m^3/d)
$q_{,m}$ – membrane aeration intensity (m/h)
Q_A – air flow (m^3/s)
$Q_{A,m}$ – membrane air flow (m^3/h)
Q_b – backflush flow (m^3/d)
Q_e –effluent flowrate (m^3/d)
Q_i – influent flowrate (m^3/d)
Q_{pump} – Pumped liquid flow (m^3/d)
R_{COD} – ratio of COD to BOD
S – influent BOD concentration (g/m^3)
SAD_m – air flow per unit membrane area (m/h)
S_e – influent BOD concentration (g/m^3)
T – wastewater temperature (°C)
t_c – time between chemical cleans (h)
$T_{K,1}$ – air temperature (K)
V – tank volume (m^3)

v_c – volume of reagent per chemical clean (m3)
V_{dig} – dig volume (m^3)
V_m – membrane module volume (m^3)
V_p – primary tank volume (m3)
X – MLSS (g/m^3)
Y – yield factor (kg VSS / kg COD)
α – process water correction for oxygen transfer
β – salts, surfactants and particulates correction for oxygen transfer
θ_x – sludge age (d)
ρ – density (kg/m^3)
τ_c – time for chemical clean (h)
φ – temperature correction for oxygen transfer

A.1 Introduction

A package plant is a complete unit fabricated in a factory and shipped to location for direct installation as opposed to more traditional plant that is installed on site. The main aerobic process technologies used for these factory-produced plants are the submerged aerated filter (SAF), conventional activated sludge (CAS), rotating biological contactor (RBCs), sequencing batch reactor (SBR), trickling filter (TF) and biological activated filter (BAF). Each of these processes have specific advantages but none produce disinfected or highly-clarified effluent.

MBRs have achieved considerable market penetration in the municipal water treatment sector over the past 15 years (Hanft, 2006). Their 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 p.e.), based on flat sheet membrane configuration. Other commercial package MBR plant technologies tend to be targeted at higher flows, upwards of 125 p.e. (25 m^3/day).

Package treatment plants are subject to very specific constraints that are not the same as bespoke municipal plants. They are often left unattended for 3-12 months and construction and process design must therefore be sufficiently robust to cope with this maintenance regime. It is highly desirable to produce a plant that is simple to install since installation is often carried out by parties specialising in groundworks rather than wastewater treatment and drainage. Most importantly, the capital expenditure must be low; operational costs are rarely considered in this market since the total energy demand per unit time is usually low, even if the specific energy demand (per volume effluent treated) is relatively high. The design must also be flexible enough to be applicable to a wide range of feed water qualities since, unlike bespoke on-site installations, the process technology is generally limited to a single plant design so as to reduce manufacturing costs through mass production.

Notwithstanding the conventional view based on capital expenditure, it is none-the-less 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. The calculation proceeds through a consideration of the specification 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. Energy demand arises primarily from a combination of aeration and liquid pumping, with a small fraction devoted to maintaining of electrical control equipment. The extent of liquid pumping and aeration is dependent on the system design. Each design is considered in turn and the cost implications over a range of flows (between 6 and 200 p.e.).

A.2 Methods

A.2.1 Boundary conditions

A European standard (prEN125566-3) is currently available to ensure that all package treatment plants are designed to the same specification. The scope of the standard is to specify "the functional requirements, process performance, testing, marking and quality control requirements" for plants up to 50 p.e.. Plants above 50 p.e. can be covered by the standards issued by Dwr Cymru and Wessex Water for adoptable package plants. This paper deals with package plants for which this benchmark can be used for convenient comparison. From this information, alongside that relating to MBR plants at various scales (Judd, 2006), some key assumptions concerning a package plant MBR can be made:

1. Flow capacity of 200 L/(p.e.) (litres per day per person).
2. Maximum of 10% of daily flow discharged over a period of one hour, i.e. 20 L/(h.person).
3. Influent quality of 300 mg/L BOD, 600 mg/L COD, 375 mg/L suspended solids and 45 mg/L NH_4-N.
4. No nutrient removal required: only an aerobic bio-zone used.
5. Effluent quality of 20:0:5 COD:SS:NH_4-N (Côte et al, 1998; Tao et al, 2005)
6. Commercially-available tanks comprising vertical cylinders of polyethylene construction.
7. Installation costs based on excavation of soil with no concrete lining required; installation volume based on a square hole with sides of the same width as the tank diameter; each side excavated at 45°angle (the angle of repose) to prevent hole collapsing; excavation costs estimated at €80 per m^3 of soil removed.
8. 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.
9. Plants capable of sustainable operation for 6 months without maintenance visits.
10. Plant capacity range of 6-49 p.e. with no redundancy provided; 50% redundancy at 50-200 p.e..
11. Aeration demand determined by generic membrane configuration (i.e. FS, HF or MT for flat sheet, hollow fibre or multi-tube respectively), independent of supplier.

System components used in this comparison are listed in Table 1, and operating costs assumed outlined in Table 2. 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 (Figs. 1-3):

a) Membrane-aerated HF iMBR or HF
b) Membrane-aerated FS iMBR or FS
c) Pumped MT sMBR or MT

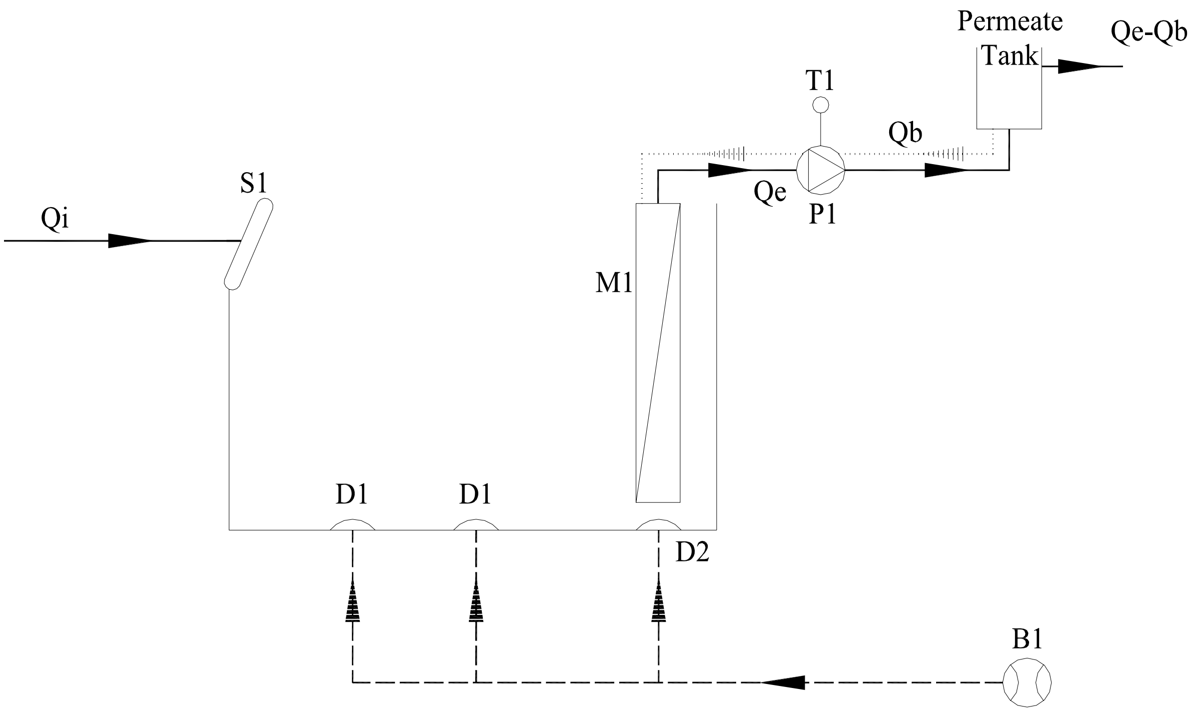

Figure 1: HF iMBR layout

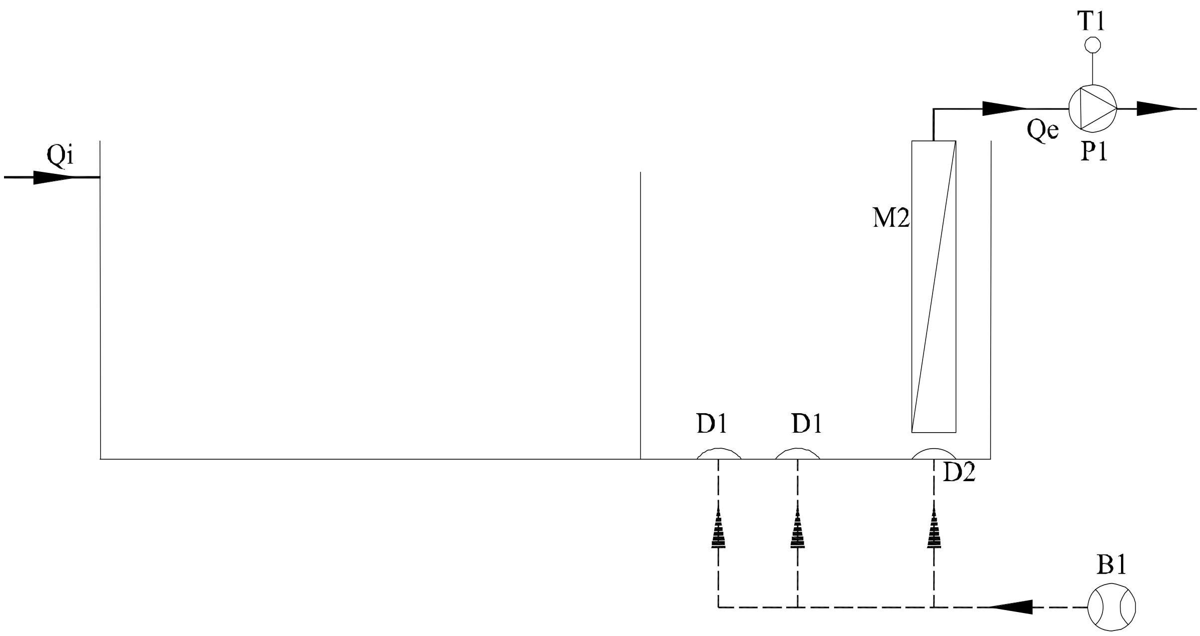

Figure 2: FS iMBR layout, including primary sedimentation tank

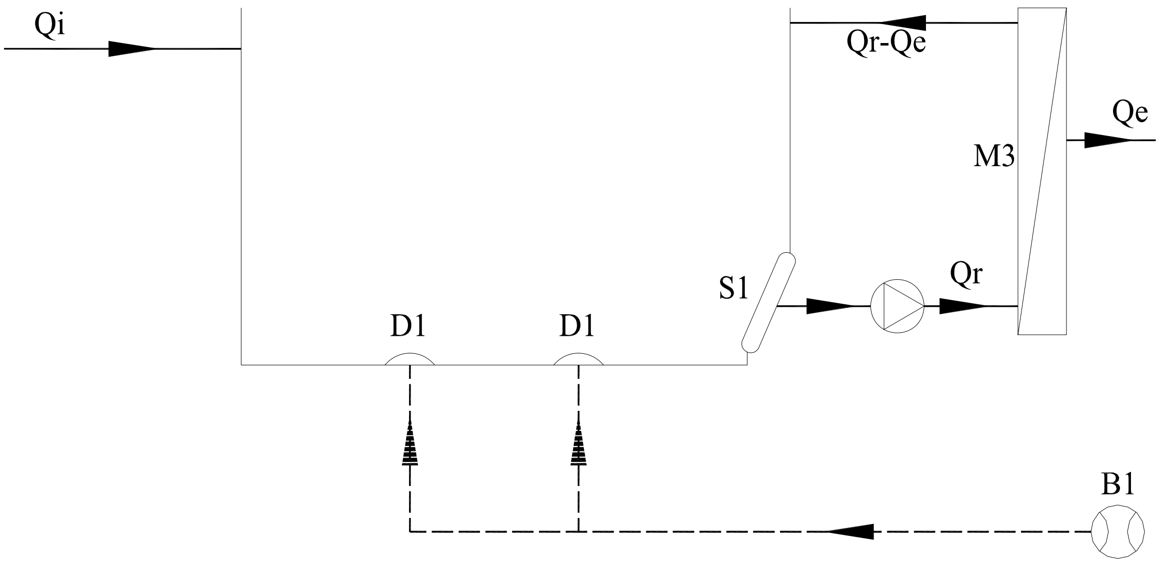

Figure 3: MT sMBR layout

Table 1: Range of capital items

Component	Life, years	Comments	ID
Tank	20	Vertical PE cylinder rotamoulded at a cost of: C = 1000 + 520 x *V*. Cylinder diameter given by: $d = \sqrt{4V/\pi h}$	V1 V2 V3
Installation	20	Total installed tank depth estimated by: H = h + 0.8. Total dig volume calculated as: $V_{dig} = d^2 H + 2dH^2$.	
Membrane	10	€150/m^2 membrane area assumed for all technologies.	M1 M2 M3
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.	P1 P2 P3
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).	B1
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.	D1 D2
Screen	10	HF, AL & MT fitted with 0.5mm screen, €1000. FS operated without screen but with primary settlement designed to BS6397:1983.	S1
Timer switch	10	Solid state timer (€85) for: • reversing permeate flow through HF module (iMBR) • relax permeate flow for FS module (iMBR)	T1 T2

Table 2: Range of operational costs assumed

Parameter	Cost	Normalised Cost per year per p.e.
Blower power cost, €/(day.person)	€0.18/kWh (70% 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 p.e 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

A.2.2 Design: biotreatment

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):

$$V_p = \frac{10x52xPE}{1000xI_d} . \frac{3}{2} = \frac{0.78PE}{I_d} \qquad 1$$

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_4-N.

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, 2004; 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 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:

$$V = \frac{QY\theta_x(C_i - C_e)}{X(1 + k_e\theta_x) - Y(C_i - C_{\sup})} \qquad 2$$

Assuming 85% of COD is removed in the bioreactor and 12% by the membrane separation (Xing et al, 2000), then C_{sup} is $0.15C_i$ and C_e is $0.03C_i$. Equation 2 thus simplifies to:

$$V = \frac{QY\theta_x 0.97C_i}{X(1 + k_e\theta_x) - 0.15YC_i} \qquad 3$$

Sludge production can be estimated from:

$$P_x = \frac{VX}{\theta_x} \qquad 4$$

Table 3: 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

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):

$$m_o = {Q(C_i - C_e)} \big/ {R_{COD}} - 1.42P_x + 4.33Q(NO_x) \qquad 5$$

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.

$$NO_x = N_i - N_e + 0.12P_x \qquad 6$$

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:

$$OTR_{cleanwater} = k_L a(C^* - C) \qquad 8$$

where C and C^* are the dissolved and saturated oxygen concentration values in kg/m^3. 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:

$$OTE_{process} = \frac{OTE_{cleanwater}}{\alpha\beta\varphi} \qquad 9$$

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:

$$\varphi = 1.024^{(T-20)} \qquad 10$$

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:

$$\alpha = e^{-0.084.X} \qquad 11$$

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.

A.2.3 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.

A.2.4 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 15 LMH
- FS iMBR 15 LMH
- MT sMBR 50 LMH

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

Membrane aeration

It is necessary to aerate a submerged membrane unit in an MBR to promote cross flow filtration. Ueda et al (1997) identified aeration intensity (air flow/unit floor area) to have the primary impact on membrane fouling. 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.

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

$$\dot{q}_m = \frac{SAD_m h_m}{\rho_m} \qquad 12$$

aeration intensity can be converted to the air flow required for each particular membrane unit by:

$$Q_{A,m} = \frac{q_m V_m}{h_m} \qquad 13$$

From this summary data average values for permeability and aeration intensity are provided taken for the two immersed technologies.

- FS: q_m = 100 m^3/m^2/hr
- HF: q_m = 220 m^3/m^2/hr

Table 4: Operating parameters

	HF iMBR	FS iMBR	MT sMBR
SRT, d (θ_x)	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 (LMH/bar)	135	300	200
Operating flux (LMH)	15	15	50
Membrane aeration intensity ($m^3/(m^2.h)$)	220	100	-
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 (LMH)	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

Pilot scale investigations into the effects of cross flow velocity on membrane permeability for MT sMBRs have been conducted at cross flow velocities between 1.5 and 4.7, producing permeabilities between 4 and 227 (Tardieu et al, 1998; Krauth and Stab, 1993; Defrance and Jaffrin, 1999; Huisman and Trägårdh, 1999). Defrance and Jaffrin (1999) observed a linear relationship between cross flow velocity and critical flux in their study of a ceramic multi-tube membrane. An average value of 3 m/s has been taken for this work.

A.3 Results and discussion

A.3.1 Plant costs

Single-household package plants are currently available for £1800-£6000 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 £2000. 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 £3080 to £5600, 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 maintanence 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 €20-30 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.

A.3.2 Plant size

Figure 4 illustrates the total annual plant cost per person per year for 6-49 p.e. plants, with the corresponding capital and operating costs given in Figures 5 and 6 respectively. The total cost incorporates capital equipment and installation costs, amortised over the plant lifetime, and operational costs. All technologies showed the expected sharp reduction in plant cost per person at very small plant sizes, with the trend approaching a constant value at around 20 p.e.. The difference in total annual cost per person between 4 and 20 p.e. plant ranges from €130-140 depending on plant type, whereas between 20 and 49 p.e. the difference is €36-38. At 50 p.e. there is a sharp increase in plant cost (of €40-70, Table 6) because of the inclusion of 50% redundancy in the plant. However above 50 p.e. there is little difference in annual cost (~€26) up to 200 p.e.; this trend is not greatly affected by plant type.

Table 5: Annual cost per person at three different plant sizes

Configuration	Size	Cost/PE/year	CAPEX/PE	OPEX/PE/year
FS	6	265	819	205
	20	135	571	93
	49	100	485	65
	50	138	541	97
	200	112	488	75
HF	6	271	738	206
	20	131	404	94
	49	93	302	65
	50	132	379	98
	200	107	328	76
MT	6	304	645	246
	20	170	347	134
	49	131	241	105
	50	193	304	158
	200	167	257	136

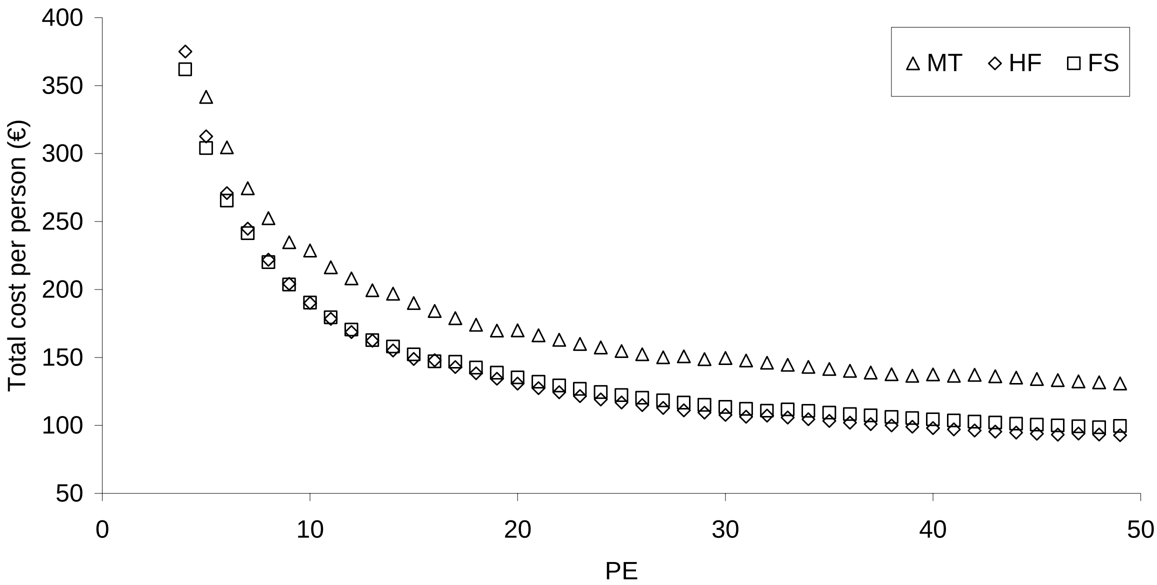

Figure 4: Annual cost per person as PE increases

All plant types show a similar trend in terms of economies of scale but the absolute costs differ. The HF system is the least expensive overall and the MT the most expensive (Figure 4). The FS system provides the lowest operating but the highest purchase and installation costs and the reverse is true for the MT system (Figures 5 and 6). If the results for total cost are taken as absolute then the features selected for the HF system are clearly preferable to those of other types of plant. However, for package plants it is often the purchase cost that is the critical factor. According to Table 7, which gives the estimated total production cost for each plant type, the difference in cost between the lowest cost (MT) and highest cost plant (FS) for the 6 p.e. plant is €1050 - a 30% difference. This trend applies to all plant sizes studied.

A further critical factor in package plant systems is operational complexity. The inclusion of a screen in the HF and MT systems may create reliability issues. Also, because the HF and FS systems require backflushing or relaxation a timer switch must be included. The inclusion of this, along 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, 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.

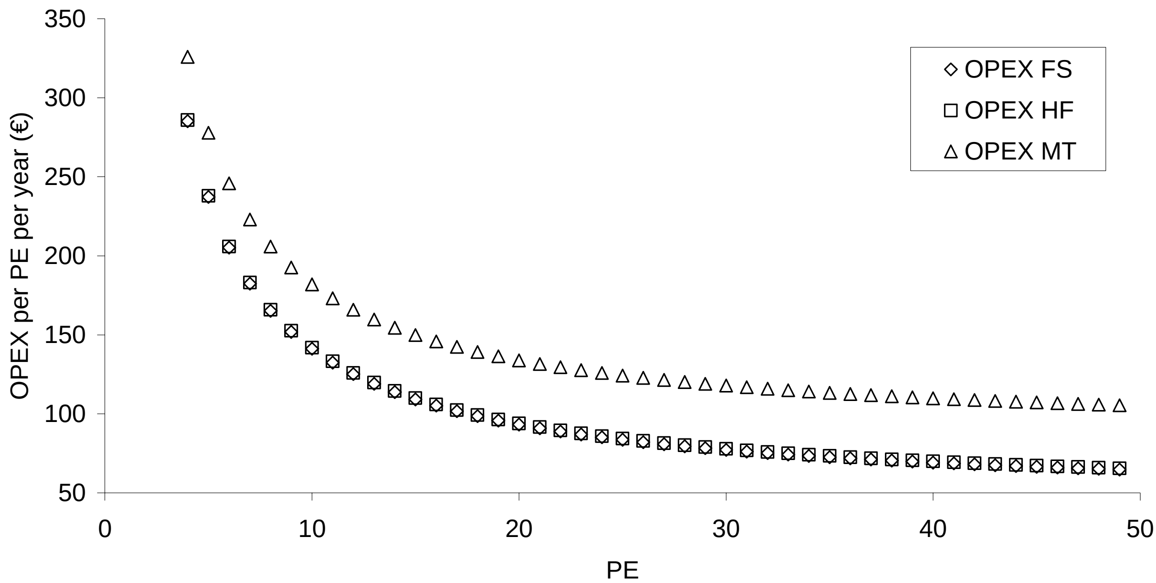

Figure 5: OPEX vs. PE

A.3.3 Tank depth

The depth of package systems is particularly pertinent 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 p.e. plant are illustrated in Figure 7. The reduced floor area under the membrane unit reduces operational costs in all systems but because pumping costs are higher than blower power costs for the MT system the reduction in OPEX is steeper in this case.

Table 5: Absolute costs and power requirements

Plant Size PE	**Plant Type**	**Power Cost** €	**Plant Capital cost** €
6	FS iMBR	139	4918
	HF iMBR	143	4431
	MT sMBR	493	3870
20	FS iMBR	463	11429
	HF iMBR	477	8088
50	FS iMBR	1737	27070
	HF iMBR	1788	18947
	MT sMBR	6168	16811
100	FS iMBR	3475	50904
	HF iMBR	3578	34307
	MT sMBR	12337	29854
200	FS iMBR	6951	97559
	HF iMBR	7161	65625
	MT sMBR	24675	57332

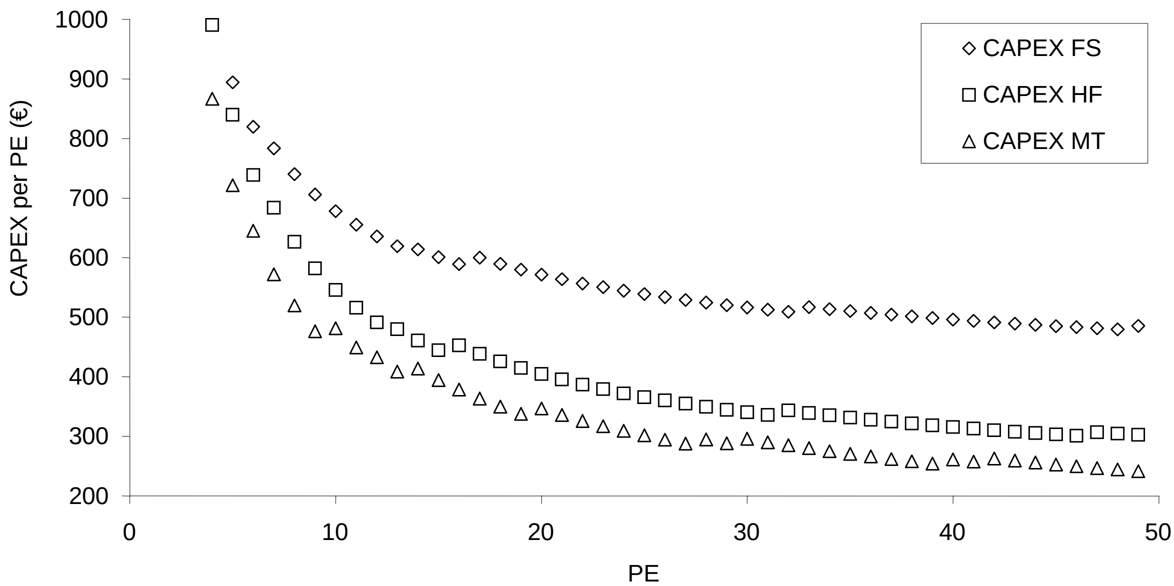

Figure 6: CAPEX vs. PE

A.4 Conclusions

Based on the assumptions made in this study:

- A single-household package plant MBR can be produced at a capital cost that is within the boundaries of commercially-available package plants, albeit at the high end of the range.
- Economies of scale exist from 6-20 p.e. plants; above this size the change in specific cost with size is low due to the assumption of the requirement for 50% redundancy (based on water company specifications).
- The operational costs of an MBR significantly exceed those of more conventional package plant designs.
- The most expensive plants to produce provide the lowest operational costs since they incorporate design elements 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.

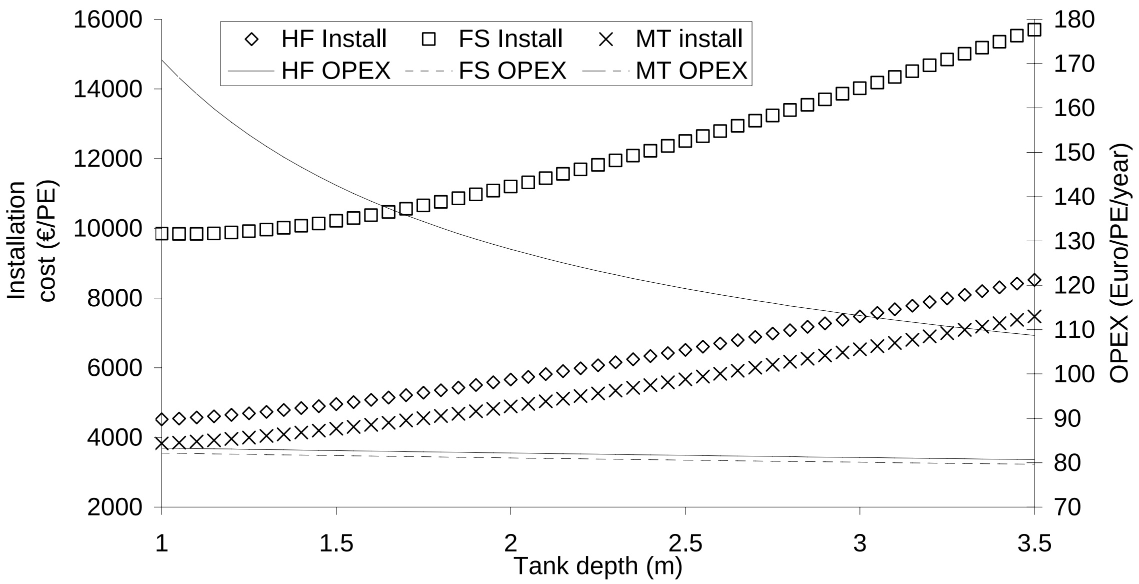

Figure 7: Change in operational and installation cost with tank depth (100 p.e. plant)

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