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REPLACEMENT OF SECONDARY CLARIFICATION BY MEMBRANE SEPARATION – RESULTS WITH TUBULAR, PLATE AND HOLLOW FIBRE MODULES

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ABSTRACT

Membrane separation systems can replace the final clarification step to separate mixed liquor suspended solids (MLSS) in the activated sludge processes. Mixed liquor suspended solids concentrations as high as 20 g/l can be obtained compared with the typical 3-4 g/l for conventional activated sludge/secondary clarifier systems. This leads to much smaller reactor volumes. In addition, excellent, solids free effluent qualities can be achieved with this process technology. This paper reports about the parallel investigation of three membrane systems installed within or outside bioreactors of 7 to 9 m^3 volume and flow rates from 1 to 3 m^3/h. The different membrane modules were investigated: plate module (80 m^2 membrane surface), hollow fibre module (80 m^2) and tubular module (45 m^2). At MLSS concentrations up to 25 g/l and water temperatures from 10 to 25°C a stable operation of the membrane systems was achieved for a period of more than one year. The energy consumption was approximately 1.5 kWh/m^3 for the plate and hollow fibre and 3.0 kWh/m^3 for the tubular module system. © 1999 IAWQ Published by Elsevier Science Ltd. All rights reserved

KEYWORDS

Activated sludge process; energy consumption; final clarification; mixed liquor suspended solids (MLSS); hollow fibre modules; membrane filtration; plate modules; tubular modules.

INTRODUCTION

Activated sludge process and membrane separation

The activated sludge process is a common technology for the treatment of wastewater. The purification takes place by suspended, aerobic microorganisms in the activated sludge tank. In the secondary clarification the suspended solids are separated from the treated water by sedimentation and recycled to the reaction tank. This separation step depends mainly on the sedimentation properties of the sludge and has several disadvantages:

- low concentration of mixed liquor suspended solids (3 - 4 g/l max.) in the activated sludge tank
- fine suspended solids in the effluent (up to 20 mg/l)
- no elimination of germs

To meet stringent nutrient removal requirements as well as bathing water standards, the conventional activated sludge process has to be enlarged and combined with further tertiary treatment steps such as filtration and/or disinfection. By replacing the secondary clarification with membrane systems the highest requirements could be met by only one system. With MLSS concentrations 3 to 5 times higher than for conventional systems, reactor volumes can be drastically reduced or the treatment capacity of existing plants can be significantly increased without enlarging bioreactor volumes. Figure 1 shows the maximum performance of membrane separation in the activated sludge process.

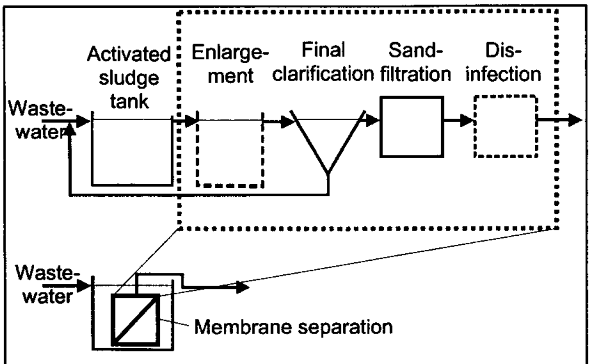

Figure 1. Maximum performance of membrane separation in the activated sludge process.

The separation of activated sludge by membranes has developed to a widespread technology in the last few years. Usually tubular modules - arranged outside the bioreactor - with an internal diameter of 6 to 25 mm were used for separation (Krauth and Staab, 1988). Recently new membrane modules have become available which can be submerged directly into the activated sludge tank (Chiemchaisri *et al.*, 1993).

For continuous separation of activated sludge with high MLSS concentrations only the cross-flow technique is suitable. The cake layer formation is controlled by shear strength and a stable capacity of the membrane system can be achieved. The shear strength can be induced either by high liquid velocities in the tubular modules or by the turbulence of uprising air and liquid in the submerged plate and hollow fibre modules.

Demands on membrane technology for application in (municipal) wastewater treatment

For any application of membrane technology in municipal wastewater treatment the hydraulic performance of the membranes is of fundamental importance. Most plants have peak flows at least double the normal average flow. Diurnal flow fluctuations are even more accentuated. While daily flow variations could be balanced, building large tanks to equalize days and sometimes weeks of stormwater flows is not practical and too expensive.

Hence, a membrane separation system for municipal wastewater treatment must be able to cope with flow variations between:

• Minimum flow	Q_{min}	about 10 % of Design Q_{dwf}
• Design dry weather flow	Design Q_{dwf}	100 %
• Maximum plant inflow*	Q_{max}	about 200 % of Design Q_{dwf}

* Note: In Germany typically only 2 x the Design Q_{dwf} is treated, the rest flows in stormwater retention basins.

For an economical optimization of membrane systems it is advisable to design the separation step for the maximum dry weather flow. Turning off some modules could easily provide the necessary elasticity towards smaller flows. Higher diurnal flows, which normally last only for a few hours, could be managed by raising the flux. Under these dynamic flow conditions membranes should react fast to variable flows and should not produce damaging and lasting effects under peak flows.

Objective of the investigations

An existing sewage treatment plant (40,000 p.e., Q_{dwf} = 700 m^3/h, Q_{max} = 1,260 m^3/h) discharging directly into Lake Konstanz (*Bodensee*) had to be upgraded to improve its effluent quality. Lake Konstanz is the drinking water reservoir for approximately 4 million people as well as bathing water. To maintain the eco-system of the lake, stringent nutrient removal requirements (P_{tot} < 0.3 mg/l in effluent) have to be met by the treatment plant. For a nearby bathing area a disinfection of the effluent is desired. To achieve the requirements it was planned to enlarge the aeration basin volume, expand the secondary clarification and install a post flocculation filtration.

As an alternative, the use of available membrane technology was evaluated in order to avoid the extension of the plant, hopefully leading to a more economical overall solution. To test the membrane technology experimental research with industrial-scale modules was performed from April 1997 to June 1998.

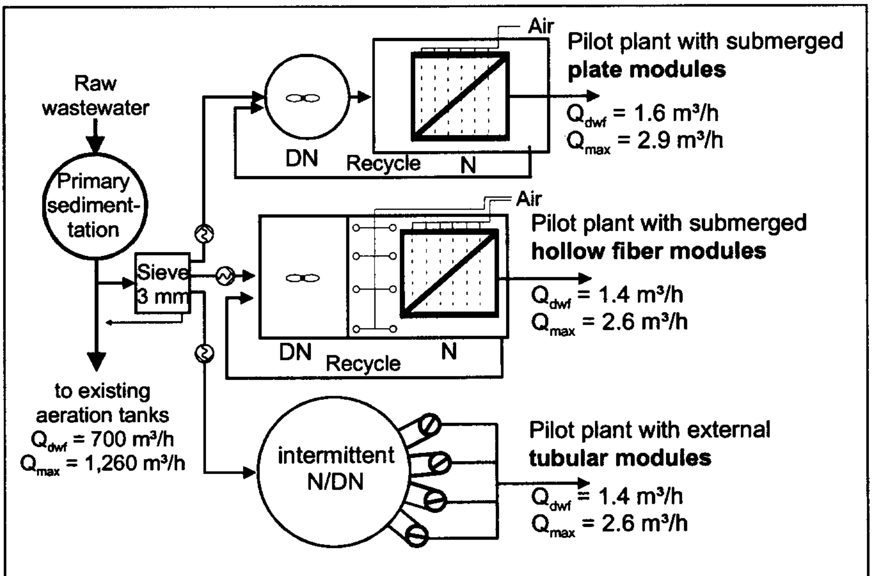

Figure 2. Process flow sheet of pilot plants.

MATERIALS AND METHODS

Description of the pilot plants

The three pilot plants received an identical wastewater feed (primary clarified municipal wastewater, see Table 3) and performed the activated sludge treatment process. The systems with plate and hollow fiber modules used pre-denitrification (DN) and nitrification (N). The recycle in these systems was about 300 % of the influent flow rate. The system with tubular modules performed the intermittent nitrification and denitrification (N/DN) by periodical aeration. The plate modules and the hollow fiber modules were submerged directly into the nitrification basin and the turbulence for membrane cleaning (cross-flow) was performed by supplying air below the modules. The specific cross-flow aeration was about 0.75 to 1.0 for the plate system and 1.0 $m^3_{air}/(m^2_{membrane} \cdot h)$ for the hollow fiber system. In the plate module system no additional air was supplied, whereas to the hollow fiber system additional air for the biological reaction was supplied by fine bubble diffuser. The tubular modules were arranged outside the bioreactor and the cross-flow was generated by a pump. The cross flow velocity was about 2 m/s. The concentrate was redirected to the surface of the reactor and performed the aeration of the system. In the denitrification period the concentrate was redirected under the water surface. Figure 2 shows the schematic process flow sheets and Table 1 summarizes the technical parameters of the installed pilot plants.

Table 1. Technical Parameter of the pilot plants

	Tubular	Plate	Hollow fibre
Membranes			
Surface	**45 m^2**	**80 m^2**	**83.4 m^2**
Pore size	**0.1 µm**	**0.4 µm**	**0.1 µm**
Reactors			
Nitrification		6.3 m^3	1.9 m^3
Filtration			2.2 m^3
Denitrification		2.75 m^3	2.8 m^3
Total volume	**9.0 m^3**	**9.0 m^3**	**6.9 m^3**
V_{DN}/V_{Total}	-	0.3	0.4
HRT	3.1-5.6 h	3.1-5.6 h	2.7-4.9
Aeration			
	surface aeration; 4 pumps with 100 m^3/h	fine bubble; 50-80 Nm^3/h	fine bubble; max. 60 Nm^3/h; coarse bubble; max. 90 Nm^3/h

Description of the membrane modules installed in the pilot plants

The material for all membranes was organic polymer. One tubular module (4 in total) consists of 110 membrane tubes with an inner diameter of 11.5 mm. The plate module consists of 100 panels arranged vertically at a distance of about 8 mm. Each panel itself consists of a support plate with welded flat membranes over a drainage system on both sides. The hollow fibre module consists of pressure resistant and self carrying hollow fibre membranes with an outer diameter of 3 mm, put together in a bundle and welded into a filtrate collector pipe (Figure 3). The hollow fibres are flexible and move in the uprising flow of air and liquid. Different flow rates were achieved by different transmembrane pressures at the plate and hollow fibre system and with different numbers of modules in use at the tubular module system.

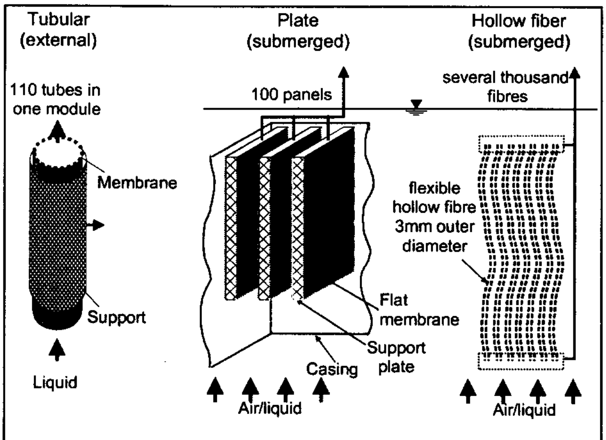

Figure 3. Tubular, plate and hollow fibre modules used in the pilot plants.

Operation, control, analysis and monitoring

Analytical parameters such as pH, O_2, temperature, transmembrane pressure and flow rates were registered continuously by automatic data logging equipment. NO_3^--N, NH_4^+-N and PO_4^{3-}-P were determined online in the effluent of the test installations. In addition, individual and flow proportional influent and effluent samples were taken. These samples were analyzed for the most important wastewater characteristics according to German Standard Methods (DEV).

Operating conditions

After a 30 days start-up period the plate and the hollow fibre plants were in continuous operation for 423 days. The tubular system was built later and started operation at the end of run 3. The major runs are described in Table 2.

Table 2. Operation conditions of the pilot plants

run	Description	days
1	constant loading with dry weather flow Q_{dwf}	54
2	variable loading with "simulated" mixed water flow	45
3	variable loading proportional to the inflow of the sewage treatment plant	117
4	ditto run 3 with phosphorous removal by precipitation	63
5	constant loading with high sludge retention time	144

Characterization of wastewater

The average wastewater concentrations for each run are summarized in Table 3.

Table 3. Average wastewater concentrations

run	COD mg/l	BOD_5 mg/l	SS mg/l	TKN mg/l	P_{tot} mg/l
1	202	128	106	49	5.6
2	100	79	81	27	4.1
3	226	123	126	51	5.4
4	270	147	190	40	6.0
5	258	147	160	40	5.7

RESULTS AND DISCUSSION

Hydraulics

As the major parameter to describe the hydraulic capacity of the membrane systems, the permeability at comparable water temperature of 10°C was chosen. This is the quotient of flux to transmembrane pressure corrected with the viscosity of water at 10°C. In the first run the pilot plants (only plate and hollow fibre) were loaded constantly. At comparable flux rates of 16 to 18 $l/(m^2 \cdot h)$ the transmembrane pressure differences reached 0.08 bar (8 kPa) for the plate and 0.25 bar (25 kPa) for the hollow fibre system. The permeability of both membrane systems was constant during run 1 (see Figure 4; note: the continuous data registration started from day 20).

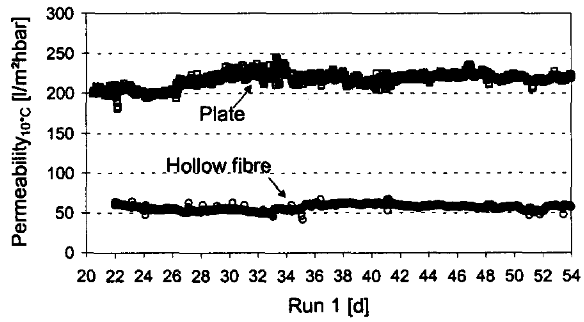

Figure 4. Permeability in run 1 (plate and hollow fibre).

In the second experimental period the pilot plants were switched over to a variable load. Flows were varied matching the variation between minimum night flow and maximum storm water flow Q_{max}. A corresponding flux rate of 7.5 to 38 l/(m²·h) was reached with the plate module. The hollow fibre modules were loaded with flux rates from 6 to 30 l/(m²·h). Figure 5 shows the flux and pressure of the plate system for a period of 5 days.

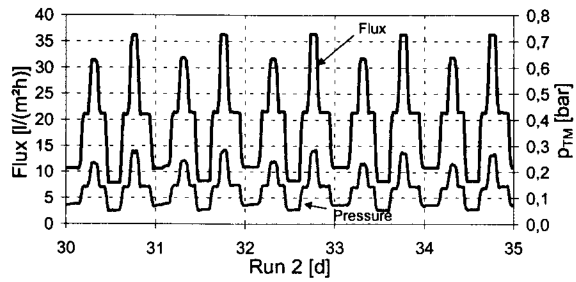

Figure 5. Flux and pressure for the plate system in run 2 (detail).

During run 2 the permeability decreased for the plate system from 210 to 110 l/(m²·h·bar) in 20 days and increased during the following period again to 140 l/(m²·h·bar). On the hollow fibre system the permeability decreased from 60 to 48 l/(m²·h·bar) and remained constant at this level (see Figure 6). It has to be emphasized that no chemical cleaning of the two membrane systems occurred.

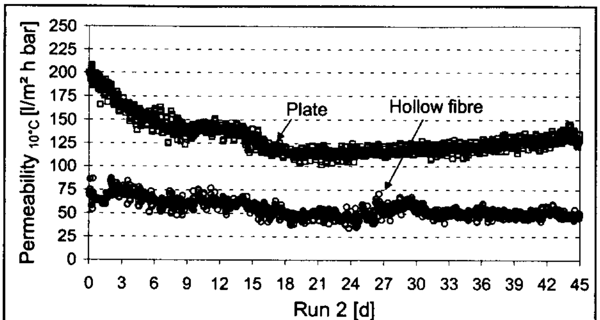

Figure 6. Permeability in run 2 (plate and hollow fibre).

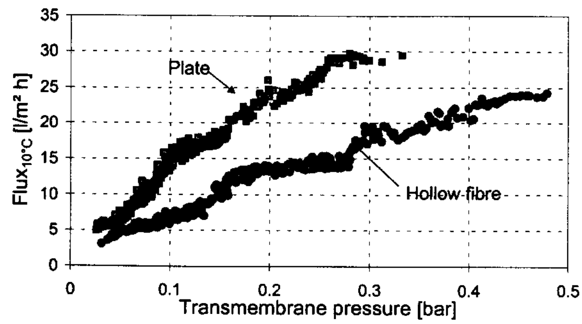

Figure 7. Transmembrane pressure and flux in run 2.

Figure 7 shows the relationship between fluxes (at 10°C) and transmembrane pressures in run 2 for both systems. The flux for both systems is almost proportional to the pressure. With the plate system maximum fluxes of 30 l/(m^2·h) at transmembrane pressures of 0.3 bar, and with the hollow fibre system 24 l/(m^2·h) at pressures of 0.45 bar were achieved.

In run 3 the flow rates to the pilot plants were proportional to the flow rate of the full sewage treatment plant. The fluxes varied from 5 to 40 for the plate system and from 5 to 35 l/(m^2·h) for the hollow fibre system with transmembrane pressures from 0.04 to 0.35 bar for the plate and 0.1 to 0.54 bar for the hollow fibre system. For the tubular system the variation of flow was achieved by different numbers of modules in use at a constant transmembrane pressure of 0.4 bar. On day 20 of run 3 the hollow fibre modules were replaced by new ones with an improved air sparging system.

As shown in Figure 8, the permeability of the plate membranes increased from about 140 to 170 l/(m^2·h·bar) during run 3. An explanation was not found. The permeability of the new hollow fibre modules decreased significantly from above 300 to about 100 l/(m^2·h·bar) in 20 days. The permeability of the tubular modules was between 40 and 60 l/(m^2·h·bar).

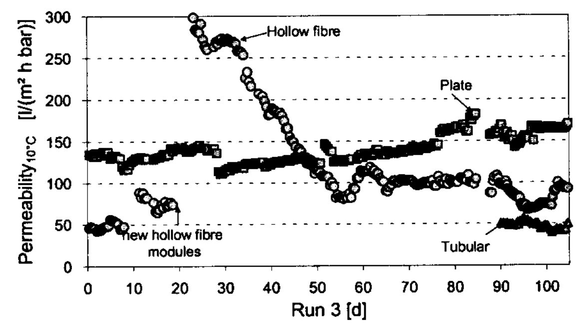

Figure 8. Permeability in run 3.

Chemical and physical parameters

The MLSS varied in all three pilot plants between 10 and 25 g/l. A significant correlation between MLSS and flux was not seen. The temperature varied from 10 to 25°C and the pH was constant at about 7 to 7.5 during the complete test period.

COD and SS removal

In the effluents of the pilot plants, suspended solids were completely removed by the membranes during the total period. Corresponding to the COD in the influent the COD in the effluent varied between 10 and 30 mg/l. The DOC in the effluent varied between 5 and 10 mg/l. The COD-elimination was approximately 90 to 95%. No differences in the effluent qualities were found due to the different pore size of 0.4 µm (plates), 0.1 µm (hollow fibre) or 0.1 µm (tubular).

Nitrogen elimination

Due to insufficient air supply, resp. oxygen supply, full nitrification was not achieved all the time with the plate and tubular system. With the hollow fibre module reactor the NH_4^+-N effluent concentrations were below detection limits nearly all the time. The total nitrogen elimination was about 60% in all test installations and was insufficient due to the slow denitrification rate. The slow denitrification rate was caused by unfavourable influent wastewater characteristics (TKN/COD ratio often higher than 0.2). It is unrelated to the membrane separation process.

Phosphorus removal

In run 4 phosphate was removed by simultaneous precipitation. $FeCl_3$ was dosed directly into the nitrification tank. The PO_4^{3-}-P in the effluent decreased rapidly after dosing, from a concentration of 5 to below 0.3 mg/l at all three systems. The precipitant feed was controlled by the P-concentration in the effluent and was adjusted to a rated value of 0.3 mg/l. During a failure in the precipitant feed the effluent concentrations increased rapidly. The molar ratio (mole metal/mole total influent phosphorus) was 1.0 to 1.5. Regarding only the precipitated phosphorus, the molar ratio was about 2.0 (mole metal/mole precipitated phosphorus). Any influences on the membrane characteristics were not observed. Figure 9 shows the phosphate concentration in the effluent for the tubular system.

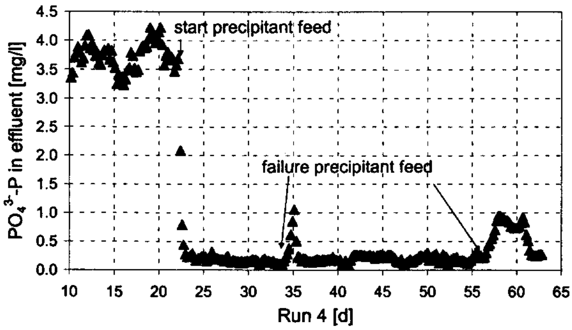

Figure 9. PO_4^{3-}-P in the effluent of the tubular pilot plant.

Sludge production

At COD-loads from 0.03 to 0.1 g COD/(g MLSS·d) the specific sludge production varied from 0.35 to 0.53 g MLSS/(g COD_{in}). In run 4 with precipitation the sludge production increased to 0.65 g MLSS/(g COD_{in}) at a sludge load of 0.1 g/(g·d).

Microbiological parameters

The effluent was examined for microbiological parameters. In the effluents of all plants salmonella were not detectable. All indicator bacteria (total coliforms, faecal coliforms, faecal streptococci) were completely removed (or below detection limit) in the system with plate modules. Due to conversions of pipe connections at the hollow fibre and the tubular system, sometimes higher values were measured in the effluent of these systems (Figure 10).

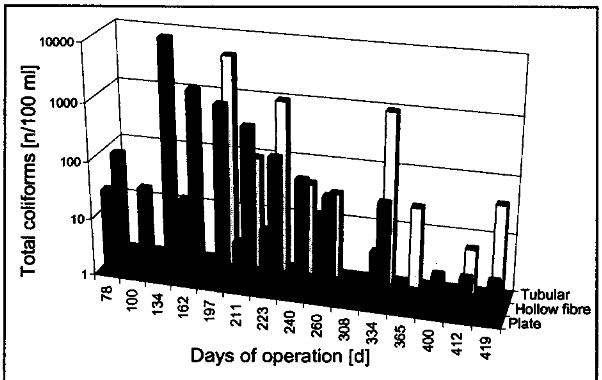

Figure 10. Total coliforms in effluent of the pilot plants.

Oxygen transfer (α-factor)

With increasing MLSS concentration the oxygen transfer decreased significantly at constant aeration conditions. The measured α-factor, which expresses the relationship between oxygen transfer in activated sludge and oxygen transfer in water decreased from 0.5 at MLSS of 8 g/l to 0.15 at MLSS of 25 g/l. Significant differences between the activated sludge of the different systems were not found (see Figure 11).

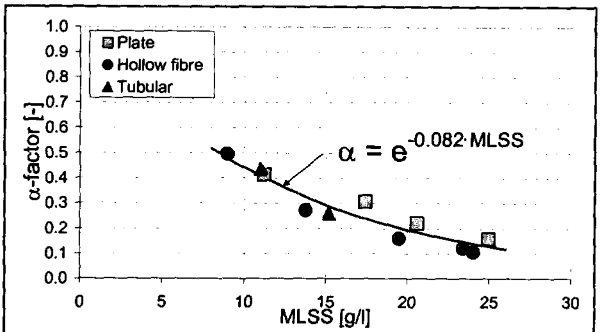

Figure 11. Relationship between α-factor and MLSS.

Energy consumption

The energy consumption per m^3 treated water is a major parameter of operational costs. In the plate and hollow fibre system only the energy for suction pumps and the blowers for aeration were considered. In case of the tubular system the energy for circulation pumps and for the feed pump was considered. The pilot plants were operated at COD space loadings of approximately 800 g COD/(m^3·d). At MLSS less than 15 g/l the plate and hollow fibre system consumed 1.0 to 2.0 kWh per m^3 treated water and the tubular system 2.5 to 3.5 kWh/m^3 (Figure 12).

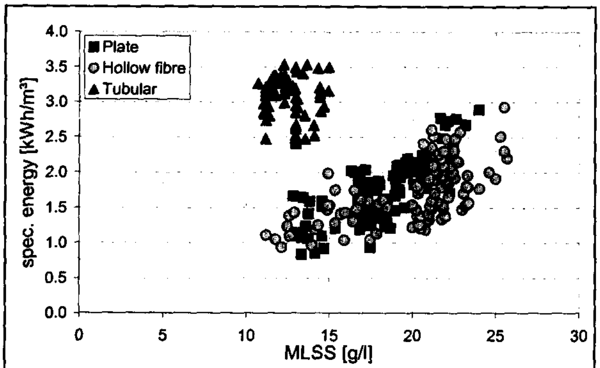

Figure 12. Relationship between specific energy consumption and MLSS.

A large dependence was evaluated between the energy consumption and the MLSS in the reactor due to the poor oxygen transfer at high MLSS concentrations. The specific energy consumption increased from 1 kWh/m^3 at MLSS concentrations of about 12 g/l to 3 kWh/m^3 at MLSS of 25 g/l for the plate and hollow fibre systems.

CONCLUSIONS

In experimental studies for a period of more than one year it could be shown that replacing the secondary clarification by membrane filtration combines the simplicity of the activated sludge process with the advantages of the membrane separation technology at stable operation conditions. The studies were carried out with submerged plate and hollow fibre modules as well as with external tubular modules. At hydraulic retention times of 3 to 6 hours, solids retention times of 20 to 30 days and MLSS concentrations up to 25 g/l, effluent COD levels well below 30 mg/l can be achieved. Suspended solids in the effluent could not be detected and – in case of no work at external pipe connections - the bacteriological indicators were around or below detection limits. The nitrogen elimination and the sludge production depend on the influent waste water characteristics and are similar to the conventional activated sludge process. With simultaneous precipitation, phosphorus concentrations in the effluent well below 0.3 mg/l can be easily achieved at a molar ratio of 2 regarding the precipitated phosphorus.

At flux rates of 16 $l/(m^2 \cdot h)$ for the hollow fibre module and 18 $l/(m^2 \cdot h)$ for the plate module stable operation is possible without deterioration of the membranes. Temporarily higher fluxes up to 24 for the hollow fibre and 30 $l/(m^2 \cdot h)$ for the plate module system can be easily achieved with higher transmembrane pressures but lead to a reduction of the permeability.

The energy consumption - only for filtration and aeration - is between 1.0 and 2.0 for the plate and hollow fibre system and between 2.5 and 3.5 kWh/m^3 for the tubular system at MLSS less than 15 g/l. Due to poor oxygen transfer in high concentrated activated sludge the energy consumption increases up to 3 kWh/m^3 for the plate and hollow fibre system at MLSS of 25 g/l.

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REFERENCES

Chiemchaisri, C., Yamamoto, K. and Vigneswaran, S. (1993). Household membrane bioreactor in domestic wastewater treatment. *Wat. Sci. Tech.*, **27**(1), 171-178.

Krauth, Kh. and Staab, K. F. (1988). Ersatz der Nachklärung durch Membranfiltration bei Anlagen mit erhöhtem Betriebsdruck. Schlußbericht zum BMFT-Forschungsvorhaben, Universität Stuttgart.