

**MEMBRANE GAS ABSORBERS FOR H₂S REMOVAL – DESIGN,
OPERATION AND TECHNOLOGY INTEGRATION INTO EXISTING ODOUR
TREATMENT STRATEGIES**

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ABSTRACT

A hollow fibre (HF) polypropylene membrane gas absorber was investigated for the removal of hydrogen sulphide (H₂S) from gas streams. Gas concentrations between 25-2010 ppmV were fed into the shell side of a membrane module whilst water-NaOH solutions flowed counter-currently in the fibre lumens. The process was effective at removing the H₂S (96% at G:L ratios up to 50 and pH 13) from the gas phase in a single pass through the membrane at all the concentrations of H₂S investigated. Analysis of the mass transfer process revealed the rate of transfer to be controlled by the gas phase transfer coefficient with a value between 1 and $25 \times 10^{-4} \text{ m.s}^{-1}$. The possible integration of a membrane absorber system into existing odour treatment strategies was assessed by comparing the membrane system, based on the experimentally determined mass transfer coefficient, with existing full scale biofiltration plants. The membrane system became economically favourable at gas flow rates lower than $1630 \text{ m}^3.\text{h}^{-1}$.

Keywords: Gas absorption, hollow fibre membrane, hydrogen sulphide, economic
assessment

INTRODUCTION

Odours in wastewater treatment arise mainly from the anaerobic degradation of sewage during transportation and treatment. Other odours arise from sources such as industrial wastewaters (solvents, volatile organic compounds) or sulphurous effluents [1]. Many compounds have been identified in sewage works odours and include reduced sulphur and nitrogen compounds, organic acids, aldehydes and ketones. The most significant of these is hydrogen sulphide (H_2S) which is the most abundant odour source in municipal sewage works. The formation of hydrogen sulphide arises from either the reduction of sulphate or the desulphurisation of organic compounds containing sulphur.

Gas/liquid contacting units, which have been employed for the removal of odourous gases through absorption, have been conventionally performed in systems such as packed towers, spray towers and bioscrubbers. Such set-ups aim to provide as much interfacial area as possible through the selection of the correct packing media or dispersion system. Although high removal efficiencies in such systems are obtained, they are limited in performance by factors such as foaming, unloading and flooding [2,3].

Absorption using microporous membranes has been successfully applied for odorous gas removal without many of the problems associated with conventional systems. In such systems, removal occurs as the gas and liquid absorbent flow across either side of the membrane, with gas/liquid mass transfer occurring at the pore mouth of the membrane. Membrane absorbers offer a number of advantages over conventional technologies, with many of the associated disadvantages being consistently diminished or eliminated through continual research and development (Table 1). Principal attributes of

the membrane system include the provision of a constant surface area for absorption, decoupling of flow rates and reduced operating and maintenance costs due to the absence of moving parts [2]. The main disadvantage of membrane absorbers is the additional resistance to mass transfer imposed by the membrane which can have a detrimental effect on the overall rate of transfer.

Membrane contactors have been used in a variety of applications, from blood oxygenation [4], to the absorption of a number of acid and alkaline gases such as, carbon dioxide, sulphur, ammonia etc into various solvents or aqueous solutions (water, soda etc), as demonstrated by Qi and Cussler [5] and Karoor and Sirkar [3]. Commercially gas absorbers are used for the carbonation/nitrogenation of beverages in the drinks industry [6].

In the present study, a HF membrane based gas/liquid absorption process, with the intention of being employed for use for contaminated gas cleanup, was investigated. The test contaminant was hydrogen sulphide and the absorbent, a dilute caustic solution. The aim of the current research was to derive design numbers and assess how this technology could be integrated into an existing odour treatment strategy thorough comparative analysis with published case studies. Analysis of H₂S removal was studied under various conditions such as water chemistry, inlet gas concentration and operating conditions of the absorber.

Mass Transfer Theory of a Microporous Membrane in Non-Wetted Mode

In this configuration, the hydrophobic membrane pores are gas filled, with liquid prevented from entering the membrane. To maintain this mode, the gas pressure, P_g , has to be higher than that of the liquid to prevent the latter from dispersing as drops into the gas phase (Figure 1). The gas liquid interface is at the pore mouth of the hydrophobic membrane on the liquid side, with the diffusion of gas occurring firstly, through the gas film in the pores, and finally dissolving into the contact liquid absorbent.

In a membrane contactor, operated under non-wetted mode, three mass transfer resistances present themselves in series: namely the gas, membrane and liquid phases. If the system is operated under steady state conditions, with gas and liquid equilibrium at the membrane wall, the concentration profile is discontinuous at the membrane interface, according to Henry's Equilibrium Law. The mass transfer coefficient in series [7] can be described as

$$\frac{1}{K_G d_i} = \frac{1}{k_g d_o} + \frac{1}{k_m d_{lm}} + \frac{1}{H k_l d_i} \quad (i)$$

where K_G (m.s^{-1}) is the overall mass transfer coefficient; k_g (m.s^{-1}), k_m (m.s^{-1}), and k_l (m.s^{-1}) are the local gas phase, membrane and liquid phase mass transfer coefficients; and d_o (m), d_{lm} (m) and d_i (m) representing the outer, logarithmic mean, and inner fibre diameter. H is the Henry's law constant (dimensionless).

From the above equation (i), it is apparent that Henry's Law constant is a predominant factor in the control of the mass transfer process. For the absorption of gases with low solubilities such as O_2 and CO_2 through a hydrophobic membrane (i), H is reduced and mass transfer becomes controlled by the liquid film layer with both the gas and membrane resistances becoming negligible.

$$\frac{1}{K_G d_i} \cong \frac{1}{H k_l d_i} \quad (ii)$$

For highly soluble gases, on the other hand, the liquid resistance is removed from the overall equation due to H being several magnitudes higher. In this instance the overall transfer equation simplifies to:

$$\frac{1}{K_G d_i} \cong \frac{1}{k_g d_o} + \frac{1}{k_m d_{lm}} \quad (\text{iii})$$

Hydrogen sulphide absorption in NaOH solutions has also been reported as gas film controlled, and so mass transfer coefficients can be calculated from the following, as described by Costello et al. [8] and Li et al. [9].

$$K_G = \frac{Q_G}{A_T} \ln \left(\frac{C_{ig}}{C_{og}} \right) \quad (\text{iv})$$

where Q_g is the gas flow rate ($\text{m}^3 \cdot \text{s}^{-1}$) and A_T represents the transfer area (m^2). C_{ig} (ppmV) and C_{og} (ppmV), represent the inlet and outlet concentrations, respectively.

MATERIAL AND METHODS

Experimental Procedure for H_2S absorption

The microporous membrane gas absorber, illustrated in Figure 2, was similar in design to that described by Kong and Li [10]. The HF cartridge used in the study was supplied by Intersep Ltd (UK) with the estimated membrane specifications listed in Table 2. The feed gas, which was a mixture of H_2S in balance with synthetic air ($\text{N}_2 + \text{O}_2$), was taken from 25, 500, 1000 and 2000ppmV bulk cylinders (BOC Gases, UK) and various concentrations obtained by diluting with air through an additional rotameter. The gas mixture was then passed through the fibre shell at the desired flow rate and a dilute caustic

solution was pumped counter-currently from the bottom of the shell through the fibre lumens.

When the contact solvent was used at pH 7, samples were taken directly into 25 ml bottles for sulphate analysis at a later date through spectrophotometry using a wet chemical method (HACH Method 8131, HACH Ltd, 1998). When using the contact solvent at pH 13 (10% NaOH), samples were taken and stored with an anti-oxidant solution (Standards Methods 4500 S²⁻.G) with the purpose of inhibiting the oxidation of sulphide and to provide a constant ionic strength solution. Sulphide measurements were obtained using an ion probe and sulphurous compounds measurements checked through atomic absorption. Obtained experimental data was compared to theoretical values which were found by determining the mass balance through the system.

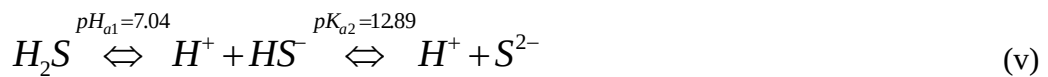
Determining Outlet Gas Concentraions

Outlet gas concentrations were measured using a gas data logger (Oialog, Australia), which had a range of 1 to 200 ppmV H₂S. The instrument was calibrated prior to each run against a concentration of 25 ppmV or 170 ppmV, depending on the gas flow concentration used for that particular experiment.

RESULTS AND DISCUSSION

H₂S Absorption

After establishing equilibrium, which was found to occur within five minutes, gas phase data showed that the removal efficiency of H₂S, upon a single pass through the membrane unit, to be strongly dependent on the liquid contactor pH. At a pH of 7.5, removal efficiency was found to decrease in a linear fashion as a function of gas to liquid (G:L) ratio, with a gradient of -2.06 and a regression coefficient of 0.66, illustrated in Figure 3. An example of this relationship, can be seen when the G:L ratio was increased from 1 to 21 there was a corresponding decrease in removal efficiency, from 95 to 56%, respectively. In contrast to this finding, at a contact solvent pH of 13, removal efficiency remained at a constant level of over 97% for G:L ratios between 1 and 50. It is these contrasting findings that demonstrate the impact of altering the alkalinity of the contact liquid on H₂S removal. The function of pH on removal performance (v) is due to the fact that H₂S is a weak dibasic acid which dissociates in water [11].



Estimates of pK_{a2} are in the range of approximately 12.2-17.1 at 25°C [12] with the latest estimates tending to be higher in value [12].

At a neutral pH, approximately 50% of the total H₂S is present in its molecular form and it is in this state that odour problems occur [14]. Above pH 7, the majority of sulphide is present in ionic form, as either HS⁻ or S²⁻. If pH is increased further, to pH 13, a second passover occurs and approximately 50% of the sulphide present is in each ionic form. Sulphide may also be oxidised to form elemental sulphur, sulphite, thiosulphate or sulphate. Data presented by Birkner and O'Brien [15] suggested that pH and sulphide/oxygen ratios were significant factors in determining oxidation reaction

products. In the case where sulphide/oxygen ratios are high, elemental sulphur was reported to be formed. At lower pHs and lower sulphide/oxygen ratios, however, sulphate tended to be formed, whilst under alkaline conditions sulphite, thiosulphate and sulphate were produced. The researchers suggested that this was related to the dissociation products, i.e. dissolved H_2S or HS^- . Buisman et al. [16] found sulphide concentration to have similar effects on the formation of certain dissociation products. At high sulphide concentrations elemental sulphur was dominant, whereas at lower sulphide concentrations sulphite, thiosulphate and sulphate were observed as reaction products. At lower pHs the reaction only tends to elemental sulphur, with the process occurring at a very slow rate in the absence of a catalyst. Liquid phase data obtained in the present study revealed that all the absorbed H_2S existed in either S^{2-} form or the sulphate form indicating that the reaction was occurring at the gas/liquid interface at pH 13 and pH 7, respectively. These findings were in agreement with results published by Qi and Cussler [5], who found that the reaction of H_2S in a caustic solution to be almost instantaneous, thus negating the liquid side resistance to mass transfer.

After a single pass of H_2S through the membrane module, the residual gas concentration, under an operating pH of 7.5, C_{go} (ppmV), showed a logarithmic trend with respect to the liquid flow rate ($\text{ml} \cdot \text{min}^{-1}$), in the form:

$$C_{go} = a \ln(Q_L) + b \quad (\text{vi})$$

Where a ($\text{min} \cdot \text{m}^{-3}$) and b (m^{-3}) are constants. It was discovered that the value of exponent a decreased in a linear relationship ($R^2 = 0.96$), from -4.9 to -2.1, as the rate of the inlet gas flow was varied between 1200 and 850 $\text{mL} \cdot \text{min}^{-1}$, respectively. From this it was obvious that the process of absorption in the membrane unit was more sensitive to liquid flow rate at higher gas flows (Table 3). An example of this relationship between liquid

and gas flow rate can be seen when operating at a gas flow rate of 1200 mL.min⁻¹. In this instance the outlet gas concentration decreased from 17 to 4 ppmV as the liquid flow rate increased from 50 to 650 mL.min⁻¹. Alternatively, when the gas flow rate was set at 850 mL.min⁻¹, the effluent gas concentration decreased from 7 to 1 ppmV over the same range of liquid flow rates (Figure 4).

Mass Transport Analysis

Overall mass transfer coefficients were calculated as both a function of gas and liquid flow rates at pHs between 7.5 and 13. The mass transfer coefficient ranged between 0.17×10^{-4} and 4.07×10^{-4} , at a pH of 7.5, across a liquid velocity range of 0.003 to 0.04 m.s⁻¹ (Figure 5). The relationship between K_G and liquid velocity showed either a slight increase or no overall change as a function of liquid velocity (Table 4). Experimental data produced mean K_G ranges from 0.72 to 3.45×10^{-4} m.s⁻¹ as the gas flow rate was increased from 850 to 5000 mL.min⁻¹, respectively, showing a greater dependency on the gas flow rate in the system. Similar observations were made when the pH was altered to 13, with no change in K_G occurring by varying liquid phase velocities between 0.006 and 0.034 m.s⁻¹. The magnitude of K_G varied as a function of both the gas flow rate and the inlet gas concentration, within the limits of 1.46×10^{-4} and 25.4×10^{-4} m.s⁻¹, at corresponding gas phase concentrations of 600, 524 and 5000 and 2010 mL.min⁻¹, respectively (Figure 6). From these findings, no direct correlation could be found by combining the two parameters in the form of inlet mass loading rate.

In contrast to the findings mentioned previously, K_G exhibited a linear relationship with gas velocity under all experimental conditions. Figure 7 shows that at

pH 13, an increase in the gas phase velocity from 0.008 to 0.067 m.s⁻¹ resulted in the subsequent increase in K_G values from 1.5×10^{-4} to 25.4×10^{-4} m.s⁻¹. By reducing the pH to 7.5, at the same gas velocity range, the effect on K_G was much smaller, with values ranging from 1×10^{-4} to 4×10^{-4} m.s⁻¹. In order to examine the relative effects of the different operating variables the linear gradient and the overall mass transfer coefficients were compared (Table 4). It can be seen that the greatest influence on mass transfer was the gas velocity of the H₂S and the pH of the contact liquid absorbent. The variation in linear gradient of K_G , as a function of gas velocity, varied between 0.005 and 0.002 at pH 7.5, compared to an average experimental value of 0.037 at pH 13 (Figure 8). An order of magnitude difference in the overall mass transfer coefficient was observed, as a function of pH, at higher gas velocities when compared to those values obtained at lower pHs.

Experimental K_G values reported in this study, ranging from 1.46×10^{-4} and 25.4×10^{-4} m.s⁻¹, are similar to values cited in the literature for H₂S absorption systems. Kreulen et al. [17], found the overall mass transfer coefficient to vary between 25×10^{-4} for H₂S mixed with N₂O, to 0.02×10^{-4} for pure H₂S, with the latter value showing absorption in that system to be controlled by the liquid phase, producing a lower overall transfer coefficient. Similar mass transfer coefficients were also reported by Li et al. [9], with ranges between 1×10^{-4} and 90×10^{-4} , for H₂S absorption into a 10% NaOH contact solution with the gas flowing in the lumen side of the membrane.

As mentioned previously, low solubility gases such as H₂S, CO₂ and SO₂ would be expected to be liquid phase controlled due to the species greater affinity for the gas phase (i.e. small H). This theory was confirmed experimentally by Karoor and Sirkar [3], with the absorption of CO₂ and SO₂ into water. However, work carried out by Yang and

Cussler [18] and Qi and Cussler [5] on CO₂ and H₂S, respectively showed that a chemical reaction takes place at the gas/liquid interface, depending on the nature of solvent used, with the effect of negating the liquid side resistance to mass transfer.

Membrane Resistance

By determining the individual mass transfer coefficient for the membrane, k_m [9,17] and comparing it to the overall coefficient (iii), the relative impact of the membrane and gas boundary layer resistances were evaluated. The membrane transfer coefficient of 0.14 m.s⁻¹ was calculated from the membrane data supplied by the manufacturer: porosity of 40% and a thickness of 15 µm. The tortuosity of the pores was estimated from values cited in the literature [18] and the thickness of the membrane was assumed to impart a resistance. Experimental values of k_m ranged from 2.27x10⁻³ to 2.63x10⁻³ m.s⁻¹, as determined from a classical Wilson plot with a velocity exponent of one (Figure 9). A similar difference in calculated and experimentally obtained membrane coefficient values were also noted by Li et al. [9], proposing that this was due to the partial wetting of the membrane pores and the possibility that the membrane substrates contributed towards a significant resistance. Comparable experimental data for the membrane transfer coefficient were also in accordance to results published by Qi and Cussler [5], using a polypropylene membrane (7.3x10⁻³ m.s⁻¹). They too found higher experimental values for polysulfone membranes (0.0123 – 0.025 m.s⁻¹) and lower values for polyethersulfone membranes (5x10⁻⁴ m.s⁻¹), demonstrating the importance of membrane material selection as regards membrane resistance on the absorption process [9]. In the current study, the porosity of the polysulfone membrane was considerably

lower than that for the polypropylene membrane, but still produced a transfer coefficient which was an order of magnitude higher than the latter.

It was determined that the individual gas transfer coefficient (k_g) contributed up to 93% of the total resistance. The percentage influence of k_g was found to decrease to the extent of 62% when gas velocity was increased, indicating a reduction in the thickness of the boundary layer as velocity increased. Beyond a velocity of 0.026 m.s^{-1} no further change in the contribution of k_g was noted which was in accordance with findings published by Li et al. [9], who reported no change in the overall transfer coefficient as function of gas velocity up to values of 5 m.s^{-1} . The membrane used by Li et al. [9] was reported to have the controlling resistance with regard to the mass transfer in the system due to the low effective porosity of the membrane material (0.017). The membrane in the current study operated at a much lower range of gas velocities ($0.005 - 0.07 \text{ m.s}^{-1}$), which was up to two orders of magnitude lower than the work presented by Li et al.[9], and had a much higher porosity (0.4). This insensitivity to gas velocity in Li's study is likely to be due to the gas phase boundary layer being reduced to a minimum, even at the lowest investigated velocities.

Industrial Case Study

MAS design data, as determined through the current investigation, was compared to existing site information from a series of UK odour treatment sites all of which were designed by the same company, at confidential locations throughout the UK. Site inclusion was limited to locations with sufficient data for comparison. In total 12 sites comprising of: 10 biofilters, 1 catalytic iron filter and 1 adsorption filter were included

(Table 5). The available information was restricted to flow rate, inlet and outlet concentration, and capital cost. In particular, no operating cost data was available restricting the comparison to capital costs rather than whole life costs. Whilst the latter is the preferred option the lack of available data reflects that in practice capital cost is often the over riding decision factor in selection of ostensibly bolt on technology solutions. Removal efficiencies of H₂S varied from 93 to 98% across the sites, with plants treating gas flows up to 12112 m³.h⁻¹, at gas concentrations of between 6.5 and 59 ppmV, producing a residual concentration as low as 0.4 ppmV (Table 5).

A simplified economic assessment was performed to take into account the paucity of data and was restricted to a comparison with the capital cost data from the biofilters sites. The approach followed traditional methods whereby the cost of the membrane plant was determined in relation to the cost of the membrane material, per unit area, and an appropriate Lang factor (LF) to adjust to total plant capital cost [19]. The product of the two components generates a cost factor (CF) which represents the total price per unit membrane area for the whole plant and was set to obtain economic parity with the existing biofilters cost.

$$CF = \text{£/m}^2 \times LF = \frac{\text{cost of biofilter}}{A} \quad (7)$$

Actual hollow fibre membrane costs can vary but normally remain within the range £30-50.m², based on UK location in 2001 . Lang factors for membrane plants of this type are reported in the literature to vary between 1.66 and 2 and combining the two ranges generates a likely range for the cost factor for a membrane absorber plant which lies between 50 and 100 [20]. The membrane option was assessed to be economically favorable when the available cost factor fell below this level.

The gas flow rate range needed to achieve economic parity, due the variability of the actual cost factors, is between 580 and 3752 m³.h⁻¹, and this reflects the importance of membrane cost on the overall economics of the MAS. For a typical membrane costs of £40.m⁻² parity occurs at a gas flow rate of 1630 m³.h⁻¹ and indicates that technologies such as the membrane absorber are more likely to be cost competitive at low to medium flow rates. The differences with flow rate reflect the respective economy of scales between the two technologies. Membrane systems are modular and hence exhibit an economy of scale exponent of nearly one, generating a linear cost with flow rate function. However, biofilters exhibit a more classical 0.66 exponent reflecting the ratio of area to volume increase that occurs in such systems.

CONCLUSION

Experiments conducted with the HF membrane absorber have demonstrated that this technology can effectively remove H₂S, with efficiencies of 96% at a pH of 13, at G:L ratios up to 50 and inlet concentrations of up to 2000 ppmV. From this study it was also discovered that the transfer of H₂S into a dilute caustic solution was gas phase controlled. The overall mass transfer coefficient of the absorption process was found to vary between 1 m.s⁻¹ and 25x10⁻⁴ m.s⁻¹. The overall mass transfer coefficient was found to vary as a linear function of gas velocity, but was insensitive to liquid velocity or inlet gas concentration. With respect to comparing the economics of this membrane system for odour removal, economic parity with a biofiltration plant was determined at gas flow rates between 580 and 3752 m³.h⁻¹, with these values being ultimately dependent on the procurement cost of the membrane.

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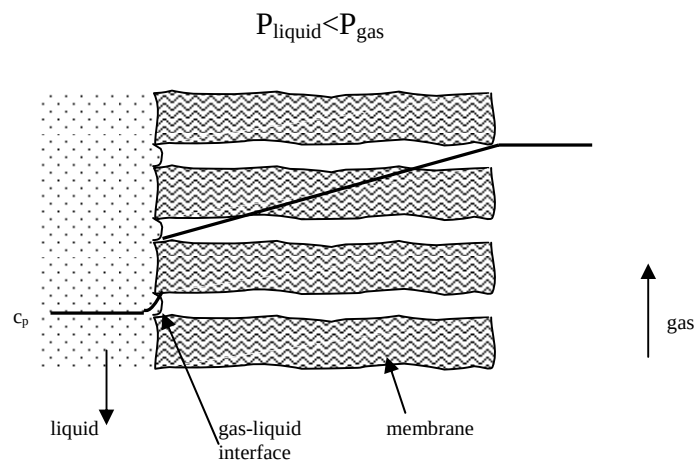
Advantages	Disadvantages
Decoupling of liquid and gas flow	Additional resistance due to membrane
Constant surface area	Gas bypassing on the shell side
Simple scale up design	Membrane fouling
Low height of mass transfer unit (HTU)	Membranes have a finite life
No moving parts	The achievable number of equilibrium stages is limited by pressure drop constraints

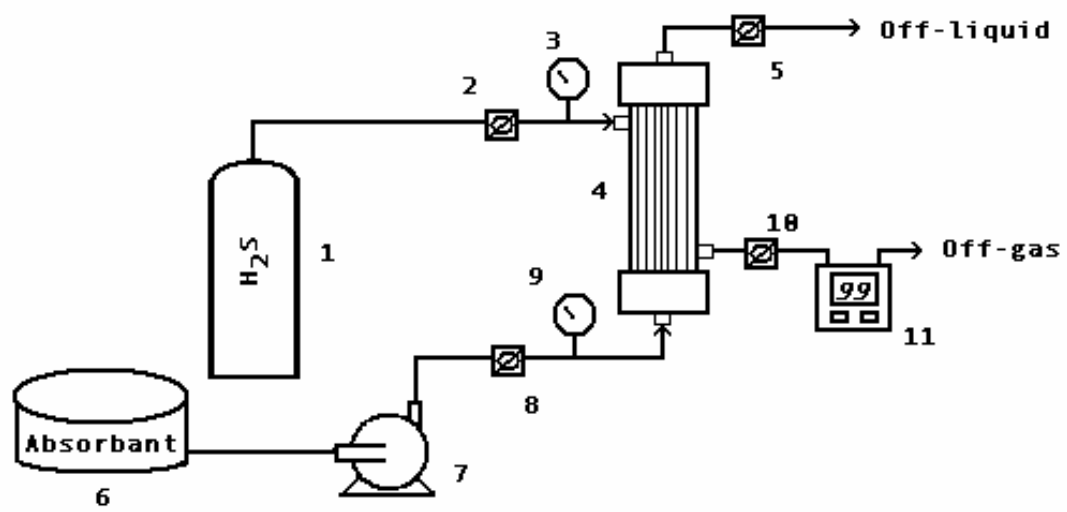
Parameter	Module
Material	Polypropylene
Inner diameter (μm)	330
Outer diameter (μm)	360
Pore size (μm)	0.6
Number of fibers	1930
Active module length (cm)	20
Shell inner diameter (mm)	35
Inner contact area (m^2)	0.4
Outer contact area (m^2)	0.4363
Packing fraction (total fibre diameter/shell volume, $\text{m}^3.\text{m}^{-3}$)	0.204
Surface to volume fraction ($\text{m}^2.\text{m}^{-3}$)	2400

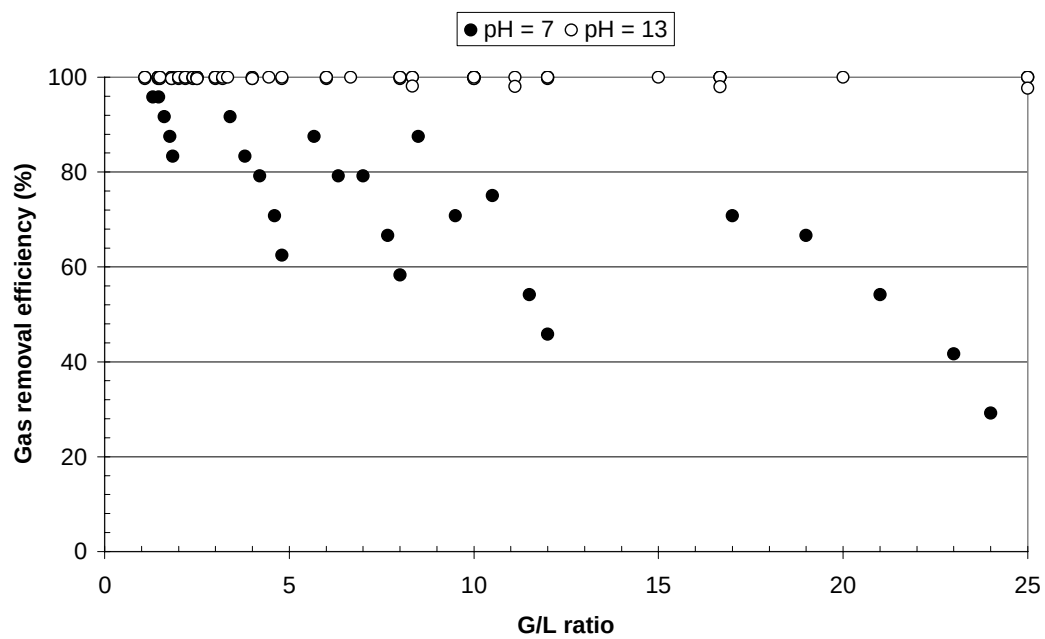
Q_G (mL.min ⁻¹)	A	b	R^2
1200	-4.9	35.8	0.98
1150	-4.2	30.3	0.98
1050	-3.1	21	0.87
950	-2.8	19.3	0.92
850	-2.1	14	0.8

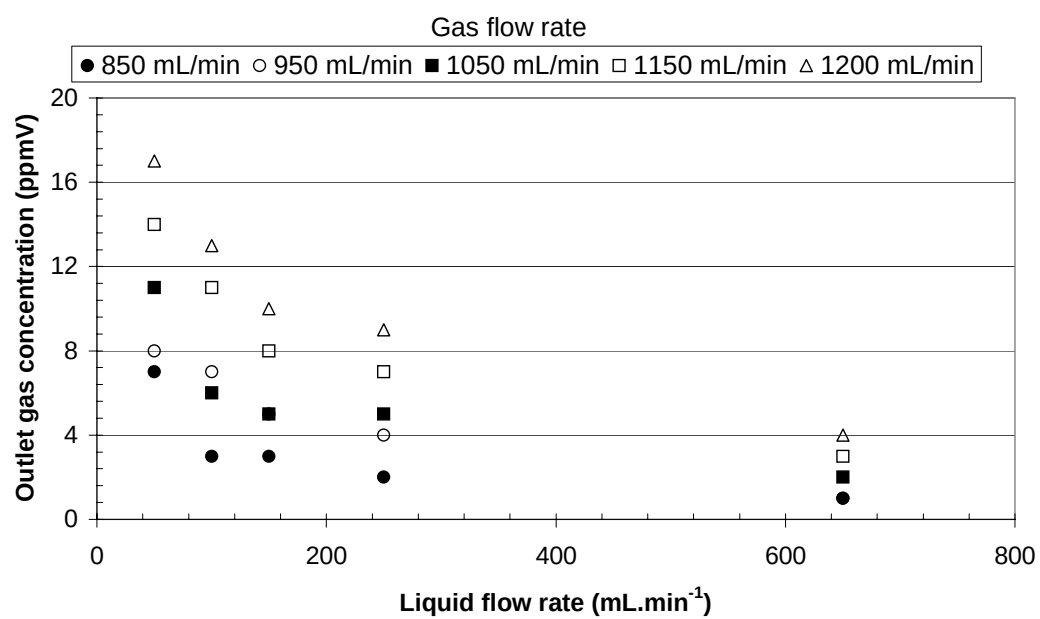
pH	C_{gi} (ppmv)	K_G (m.s ⁻¹) x10 ⁻⁴			Linear gradient as a	
		Min	Max	average	function of:	
					v_L	v_g
7.5	24	0.17	4.07	1.28	0-0.0038	0.005
7.5	514	1.66	3.58	2.35	-	0.0219
7.5	2010	0.59	1.40	0.88	0.0024	-
13	514	1.46	22.6	6.37	No	0.037
13	1024	2.89	24.1	10.1	No	0.0356
13	2010	3.01	25.4	9.54	No	0.0347

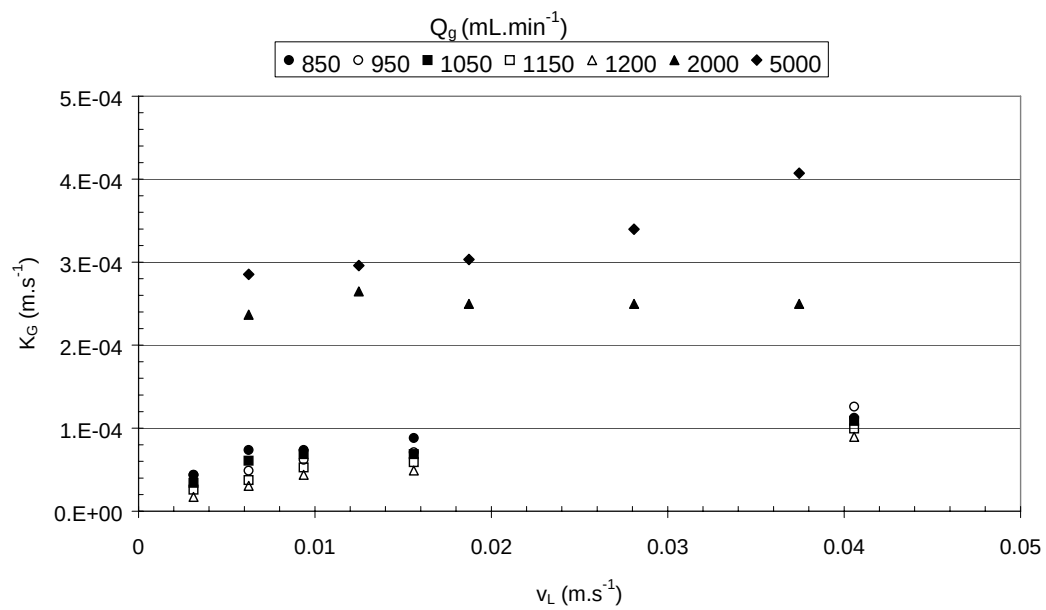
Site	Technology	Current Unit						Membrane Contactor			
		C_{gi}	C_{go}	E	Q_g	Volume	Cost	Membrane area (m^2)		Cost factor	
		(mg/l)	(mg/l)	(%)	(m^3/h)	(m^3)	(£)	MIN	MAX	MAX	MIN
A.1	Biofilter	0.09	0.0044	95	802	35	28500	269	672	106	42
A.2	Biofilter	0.04	0.0018	95.5	2309	60	52500	796	1989	66	26
A.3	Biofilter	0.09	0.0044	95	7905	90	110000	2651	6627	41	17
A.4	Biofilter	0.01	0.0007	93	10170	350	-	3005	7512	-	-
A.5	Biofilter	0.01	0.0003	97	1500	14	34000	584	1461	58	23
A.6	Biofilter	0.08	0.0042	94.7	1893	29	66000	620	1550	106	43
A.7	Biofilter	0.07	0.0036	94.8	2773	32	55000	914	2286	60	24
A.8	Biofilter	0.09	0.0046	94.8	723	25	55000	239	597	230	92
B.1	Adsorption	0.02	0.0006	97	33	0.6	33000	13	32	2567	1027
B.2	CIF + Adsorption	0.03	0.0010	96.6	2227	16	111000	86	214	1294	518
C.1	Biofilter	0.01	0.0002	98	2352	-	-	1022	2556	-	-
C.2	Biofilter	0.02	0.0005	97.5	12112	-	-	4964	12411	-	-

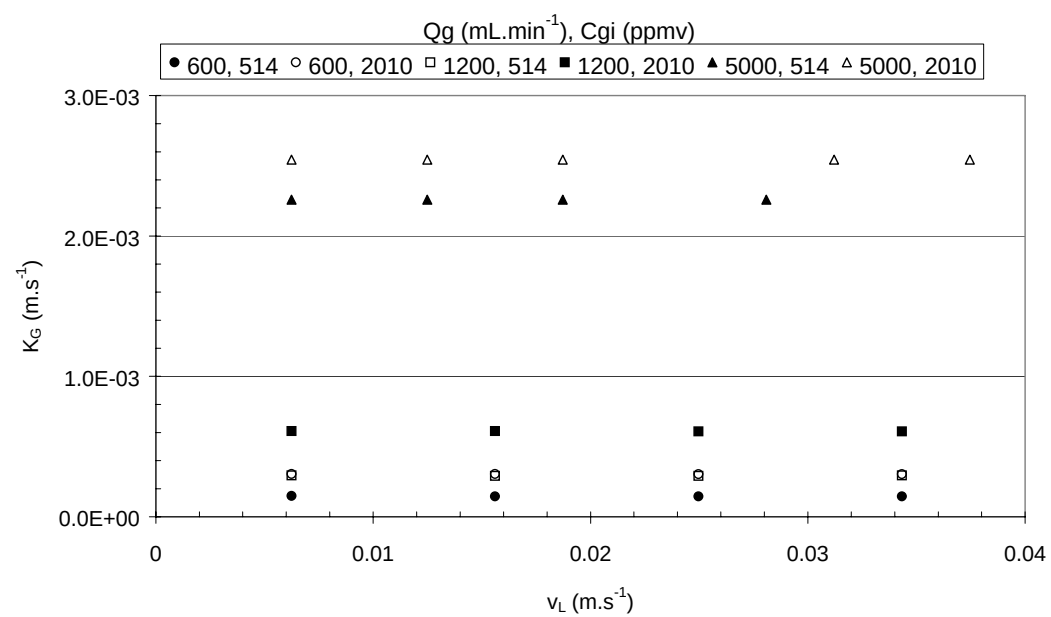


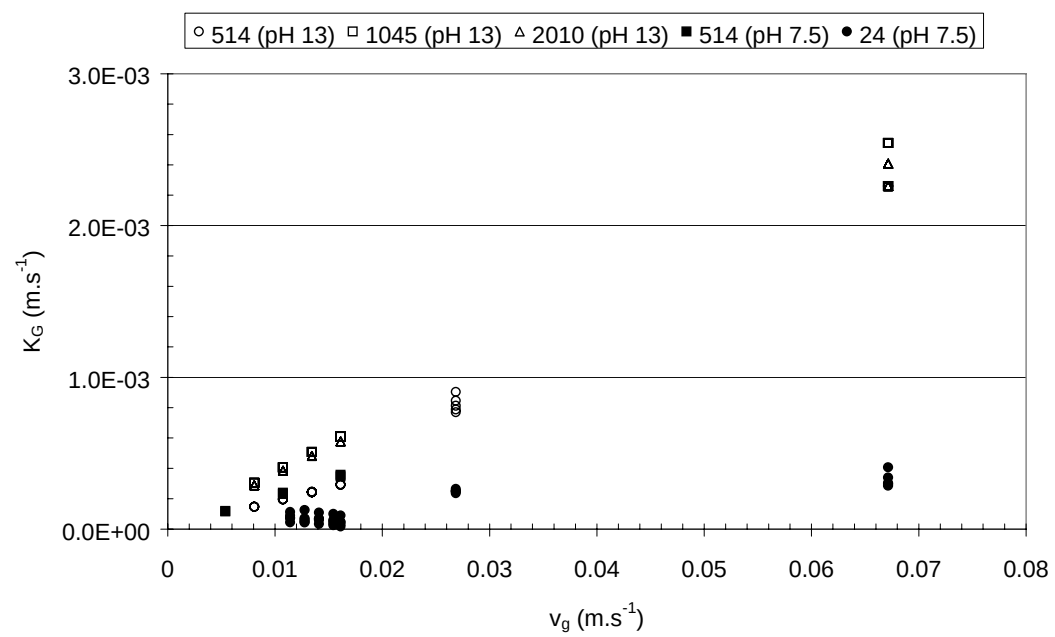


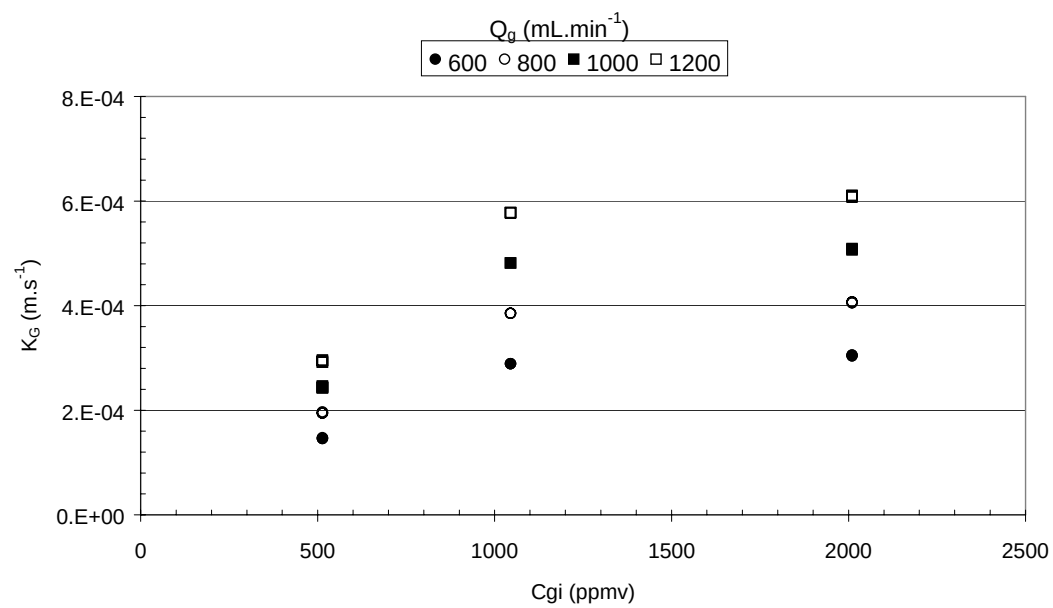


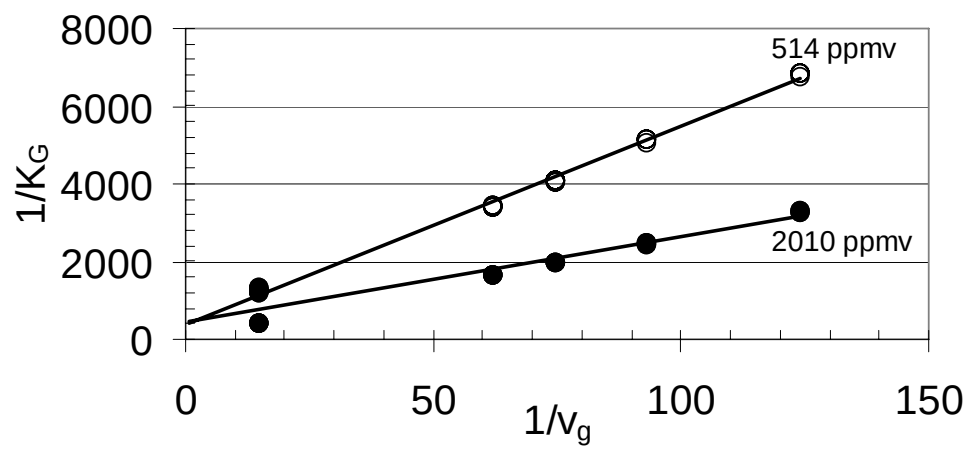


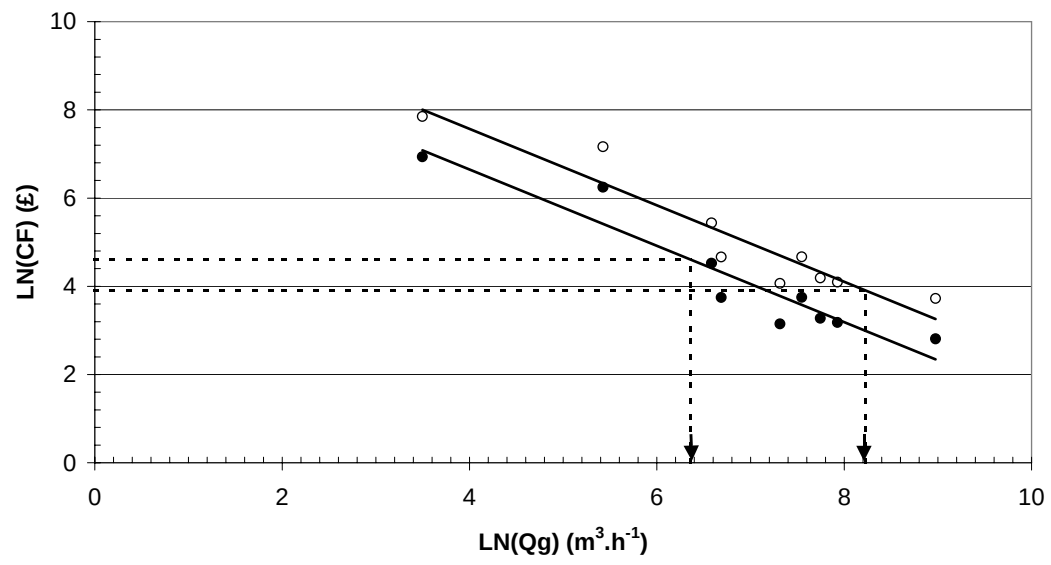












CAPTIONS TO TABLE AND FIGURE TITLES

- Table 1. Advantages and disadvantages of membrane gas absorbers
- Table 2. Membrane module specification
- Table 3. Summary data of logarithmic trend from Figure 6
- Table 4. Summary of mass transfer results
- Table 5. Summary of case study data
-
- Figure 1. Membrane contactor in a non-wetted mode with gas filled pores
- Figure 2. Schematic of the H₂S absorber unit: 1. H₂S gas cylinder (BOC gases, UK); 2. Flow regulator (Platon, range 0.1-1.2 ml/min); 3. Pressure gauge (Platon, range 0.2-5 bar); 4. Membrane module (Intersep Ltd); 5. Pressure regulator, liquid side (Platon, needle valve); 6. PVC Tank with absorbant (50 l, PVC); 7. Centrifugal motor-pump (Lafert, 0.25 KW, 1360 rpm); 8. Flow regulator (Platon, needle valve); 9. Pressure gauge (Platon, range 0.2-5 bar); 10. Pressure regulator, gas side (Platon, needle valve); 11. Gas meter (Odalog, Australia).
- Figure 3. Removal efficiency of H₂S from the gas phase as function of gas to liquid ratio.
- Figure 4. Residual gas concentration as a function of liquid flow rate when operated at pH 7 (liquid phase in the fibre lumens).
- Figure 5. Effect of liquid velocity on the overall mass transfer coefficient at different gas flow rates (pH 7.5, $C_{gi} = 24$ ppmV).

- Figure 6. Effect of liquid velocity on the overall mass transfer coefficient at different gas flow rates and influent gas concentrations (pH 13).
- Figure 7. Effect of gas velocity on the overall mass transfer coefficient at different influent gas concentrations and solvent pH.
- Figure 8. Effect of influent gas concentration on the overall mass transfer coefficient for different gas flow rates (pH 13).
- Figure 9. Wilson plot based on gas velocity and aqueous phase pH 13.
- Figure 10. Economic assessment of membrane absorber in comparison to a biofilter system.



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Membrane Gas Absorbers for H₂S Removal - Design, Operation and Technology Integration into Existing Odour Treatment Strategies.

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