

Synergistic extraction and separation of Dysprosium and Europium by supported liquid membrane

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Abstract—Synergistic extraction and separation of Dysprosium (Dy) and Europium (Eu) via supported liquid membrane (SLM) were investigated using mixture of di-2-ethylhexyl phosphoric acid (D2EHPA) and Bis (2,4,4-trimethylpentyl) phosphinic acid (Cyanex272). The results showed that separation of Dy and Eu was highly dependent on the pH of feed solution. Dy can be extracted more than Eu, because the equilibrium constant for Dy using the mixture of D2EHPA and Cyanex272 was greater than that for Eu. Various parameters were optimized to achieve maximum separation factor, namely feed phase pH, carrier concentration and stripping phase concentration. The membrane was stable at six cycles of operation.

Keywords: Dysprosium, Europium, Supported Liquid Membrane, Synergistic Separation

INTRODUCTION

Rare earth elements (REE) have become crucial and strategic materials from an industrial perspective. The widest application of rare earth elements is found in some lucrative domains such as catalysts, glassmaking, lighting, and metallurgy. These materials are also used in high-growth markets such as battery alloys, ceramics, and permanent magnets [1]. Dysprosium (Dy), for example, is mainly used in neodymium-iron-boron magnets to increase strength and corrosion resistance. It is also added to the terfenol-D compound, which is a highly practical material for future technology applications. Dy is found in the minerals such as monazite and basnasite, often with Eu or other rare earth elements. Therefore, separation of these elements is very important [2].

Liquid membrane (LM) technique, as an advanced solvent extraction procedure, provides a simple and effective method for the selective separation and concentration of rare earth elements [3]. LMs combine extraction and stripping processes in a single stage, and they have the advantage of nonequilibrium mass transfer, where the solute can move from a low to high concentration solution [4]. Supported liquid membrane (SLM) and emulsion liquid membrane (ELM) are the two configurations of the membranes that are generally used for facilitated transport of the metal ions [5]. Recently, the SLM process has been regarded as an attractive alternative for conventional solvent extraction for metal separation because limitations such as emulsification, flooding, loss of solvent, phase disengagement and large solvent inventory can be avoided [6]. The

advantages possessed by SLMs, such as less capital and low operating cost, considerable potential for energy-saving, minimal use of expensive carriers, simple operation, and easy scale-up, favor their use in experimental studies [7].

Synergistic extraction has drawn much attention in the last decades. In the synergistic extraction, mixtures of two different extractants enhance the extraction of a metal as compared with the normal additive effect of these extractants separately [8]. Many authors reported the synergistic solvent extraction systems for extracting and separating rare earths using acidic, neutral and solvating extractants [9-16], but in the case of liquid membrane, there are limited number of investigations devoted to the synergistic extraction of rare earth elements. Synergistic effect between Cyanex272 and TBP was observed in the separation of yttrium ions from the mixture of rare earths by a microporous hydrophobic hollow fiber supported liquid membrane [17]. Synergistic extraction of Eu from nitrate medium using TOA and TBP was studied in a contained supported liquid membrane (CSLM) system. The reported results showed better permeability in comparison with the individual extractant [18]. Joshi et al. studied the transport of Uranium(VI) and Europium(III) across a supported liquid membrane using D2EHPA and different organo phosphorous oxodonors (TBP, DBBP, TOPO and Cyanex923). They showed that the SLM formed by TOPO and D2EHPA has a better U(VI) transport from the phosphoric acid feed with a good selectivity over Eu(III) [5]. The synergistic effect of TOPO and D2EHPA was also observed during the separation of neodymium from lanthanide series via hollow fiber supported liquid membrane [19]. Synergistic extraction and separation of lanthanum and neodymium were found by using a mixture of HTTA and TOA by Ramakul et al. [20].

Acidic Cyanex series extractants offer high extraction power, and

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their solubility in aqueous solutions is very low, but they are rather expensive to be used commercially, while D2EHPA is considerably cheaper [10,20]. In this study, the synergistic separation of Dy and Eu using mixture of D2EHPA and Cyanex272 extractants has been investigated for the first time. Various parameters such as feed phase pH, carrier concentration and stripping phase concentration, which affect the transport of Dy and Eu, were studied in detail. Stability of the membrane against the leaching of the carrier was also investigated. A simple kinetic model was proposed to evaluate the membrane diffusion coefficient and mass transfer coefficient of Dy under the current experimental conditions.

EXPERIMENTS

1. Materials

The commercial extractant di-2-ethylhexyl phosphoric acid (D2EHPA) and Bis(2,4,4-trimethylpentyl) phosphinic acid (Cyanex272) were purchased from Sigma Aldrich (Germany) and used without further purification. The aqueous solutions of Dysprosium (III) and Europium(III) were prepared by dissolving their nitrate salts ($\text{Dy}(\text{NO}_3)_3 \cdot 6\text{H}_2\text{O}$, $\text{Eu}(\text{NO}_3)_3 \cdot 6\text{H}_2\text{O}$, Middle East Ferro Alloy Company, 99.9% purity) in deionized water. Kerosene was obtained from Alfa Aesar (Germany) and HNO_3 (65%) was purchased from Merck (Germany). All other chemicals used in this work were of A.R. grade. During the study we used flat-sheet type PTFE hydrophobic microporous membranes, procured from Millipore (USA), as membrane support. The characteristics of these membrane filters are given in Table 1. The effective area of the membrane support was determined to be 8.18 cm^2 that was computed according to the geometrical area (which was determined to be 9.62 cm^2) and the membrane porosity.

2. Apparatus

A digital pH meter (Sartorius, USA), calibrated daily with standard buffer solutions, was used for measuring the pH values of the aqueous phases.

Equilibrium experiments were carried out in a temperature controlled shaking water bath (INFORS AG).

Quantitative determination of Dy and Eu was made using inductively coupled plasma-Atomic emission spectrometry (ICP-AES, Optima 7300 DV, America). The wavelengths of Dy and Eu ions were 353.17 and 381.96 nm, respectively. The method was proposed by Wiel [23].

3. Methods

3-1. Solvent Extraction Studies

In the solvent extraction experiments, appropriate volumes (10 ml) of aqueous and organic phases were contacted in reagent bottles using thermostatic shaking water adjusted to 25°C for 30 min, enough time to reach equilibrium. Then the two phases were separated and Dy or Eu concentration in the aqueous phase before and after

extraction was determined by ICP-AES. The concentration of metal ion in the organic phase was determined by mass balance. The distribution coefficient (D) was calculated as the ratio of metal concentration in the organic phase to that in the aqueous phase in equilibrium. The experiments were repeated two times. Average error for distribution coefficient was 5%.

3-2. Membrane Studies

The membrane transport experiments were carried out in a permeation cell consisting of two 200 cm^3 Perspex compartments separated by a microporous membrane, one for feed solution and the other for stripping solution. Initial concentrations of Dy and Eu in the feed phase were $0.33 \times 10^{-3} \text{ mol L}^{-1}$ in all the SLM studies. The stirring rate of the feed and stripping solution was 500 rpm.

The liquid membrane phase was prepared by dissolving Cyanex272 and D2EHPA in kerosene. The PTFE support was impregnated with the carrier solution for 24 h, then removed from the solution and wiped carefully with a tissue paper to remove the excess carrier. After preparation, the membrane was put in the transport cell and the extraction rate was determined by monitoring Dy and Eu concentrations by ICP-AES in the feed and stripping phases as a function of time. The percentage of extraction was determined as follows:

$$\% \text{Extraction} = \frac{[\text{M}^{3+}]_0 - [\text{M}^{3+}]_t}{[\text{M}^{3+}]_0} \times 100 \quad (1)$$

where $[\text{M}^{3+}]_0$ and $[\text{M}^{3+}]_t$ are the concentration of metal ions in the feed phase at the start of the experiment ($t=0$) and at time t , respectively.

The separation factor $S_F(AB)$ is the ratio of concentrations A and B in the strip and feed phase. This was calculated as in Eq. (2):

$$S_F(AB) = \frac{([\text{M}^{3+}]_A / [\text{M}^{3+}]_B)_{\text{strip}}}{([\text{M}^{3+}]_A / [\text{M}^{3+}]_B)_{\text{feed}}} \quad (2)$$

where $[\text{M}^{3+}]_A$ and $[\text{M}^{3+}]_B$ are the concentrations of Dy and Eu, respectively.

All experiments were at ambient temperature (25°C).

RESULTS AND DISCUSSION

1. Solvent Extraction Studies

The extraction behavior of Dy was studied with D2EHPA, Cyanex272 and mixtures of them by the solvent extraction method. The aqueous solutions of Dy ($4.31 \times 10^{-3} \text{ mol L}^{-1}$) were used for the extraction studies. The synergistic enhancement factor (R) could be obtained based on the experimental data [24],

$$R = \frac{D_{AB}}{D_A + D_B} \quad (3)$$

where D_{AB} , D_A and D_B are the distribution coefficients when Dy is extracted by D2EHPA+Cyanex272, D2EHPA and Cyanex272, respectively. The synergistic extraction occurs when $R > 1$, whereas $R < 1$ means antagonistic extraction. Table 2 shows the R value for Dy as a function of D2EHPA mole fraction (X_{D2EHPA}). As can be seen, there are synergistic effects in the presence of both extractants. The maximum value of R was observed when D2EHPA mole fraction was 0.6 in the organic phase. With this background of sol-

Table 1. Characteristics of the membrane filters used

Membrane	Diameter (mm)	Nominal pore size (μm)	Porosity (ϵ)	Thickness (μm)	$\tau = 1/\epsilon$ [22]
PTFE	47	0.22	85%	150	1.17

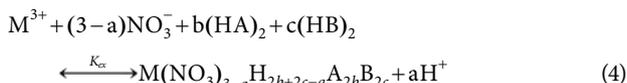
Table 2. Synergistic enhancement factor (R) for the extraction of Dy using different concentration of D2EHPA and Cyanex272

X_{D2EHPA}	D2EHPA concentration (M)	Cyanex272 concentration (M)	Synergistic enhancement factor (R)
0	0	0.060	1
0.1	0.006	0.054	1.41
0.4	0.024	0.036	1.93
0.5	0.030	0.030	2.25
0.6	0.036	0.024	2.97
0.8	0.048	0.012	1.90
0.9	0.054	0.006	1.13
1	0.060	0	1

Experimental conditions: Phase ratio=1; T=298.15 K; Sum of two extractants concentration equivalent to 0.060 M in their mixture; aqueous phase pH: 4

vent extraction data, to study the synergistic separation of Dy and Eu in the SLM system, the mole fraction of D2EHPA in the mixture of the two extractants in the liquid membrane is adjusted to be 0.6.

In the mixture of D2EHPA and Cyanex272 for Dy or Eu extraction, the synergistic extraction reaction could be written as:



where $(HA)_2$ and $(HB)_2$ represent dimeric form of D2EHPA and Cyanex272 in kerosene [2]. The equilibrium constant, K_{ex} is defined as:

$$K_{ex} = \frac{D_{AB}[H^+]^a}{[(HA)_2]^b[(HB)_2]^c[NO_3^-]^{(3-a)}} \quad (5)$$

Taking the logarithm of Eq. (5), and rearranging the formula:

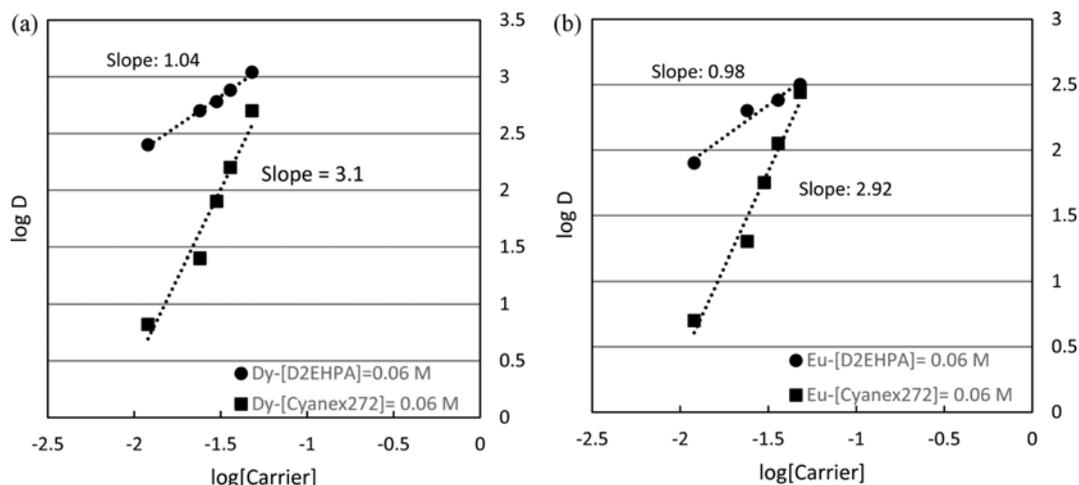


Fig. 2. Relationship between distribution coefficient (D) and equilibrium concentration of (D2EHPA+Cyanex272) for (a) Dy (b) Eu (aqueous phase pH: 4, phase ratio=1, T=298.15 K).

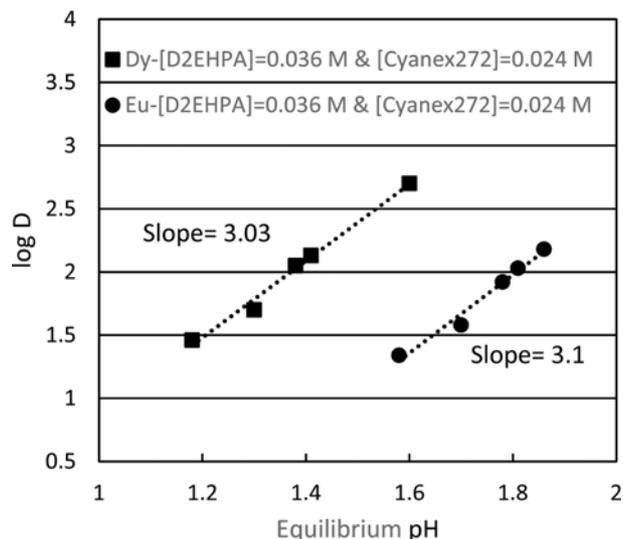


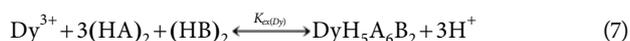
Fig. 1. Relationship between distribution coefficient (D) and equilibrium pH for Dy and Eu (organic phase: 0.036 M D2EHPA + 0.024 M Cyanex272, phase ratio=1, T=298.15 K).

$$\log(D_{AB}) = \log(K_{ex}) + (3-a)\log[NO_3^-] + b\log[(HA)_2] + c\log[(HB)_2] + a\text{pH} \quad (6)$$

A series of experiments was carried out to determine the numbers of stoichiometry, using slope analysis method. Typical experimental distribution curve $\log(D_{AB})$ vs. equilibrium pH at a fixed concentration of extractants and initial Dy and Eu concentrations of 4.31×10^{-3} mol L⁻¹ is shown in Fig. 1. The results show straight lines with a slope of about 3 for both of metal ions. To determine the values of "b" and "c", the relationship between the concentration of one extractant and distribution ratio was studied. The experiments were carried out at feed pH of 4 and fixed concentration of other extractant. Figs. 2(a) and 2(b) show that the plots of $\log(D_{AB})$ for Dy and Eu versus $\log[(HA)_2]$ and $\log[(HB)_2]$ are linear with slopes of about 3 and 1, respectively. So Eq. (4) for Dy and Eu can be presented as:

Table 3. Equilibrium constants of Dy and Eu extraction by mixture of D2EHPA and Cyanex272

Metal	System	log (K_{ex})
Dy	0.036 M D2EHPA+0.024 M Cyanex272	2.78
Eu	0.036 M D2EHPA+0.024 M Cyanex272	1.84



These mechanisms of the reaction indicate participation of three D2EHPA molecules and one Cyanex272 molecule in the extraction of Dy and Eu ions.

The equilibrium constants of Dy and Eu extracted with mixtures of D2EHPA and Cyanex272 were obtained with multiple linear regression, as shown in Table 3.

2. Membrane Transport Studies

In this investigation the liquid membrane phase contains the mixture of D2EHPA and Cyanex272 carriers, and kerosene as diluent, which are hydrophobic in nature. The liquid membrane phase is embedded in a polymeric microporous supported film that is also a hydrophobic PTFE film. Therefore, the direct transport of metal ions through the membrane is impossible, and the only way is the transport of metal ions via formation of metal-carrier complexes [25]. Transport mechanism of Dy and Eu ions through the liquid membrane is shown in Fig. 3. As observed, Dy and Eu ions react with the carrier at the feed-membrane interface to form complex species. After that, the complex species diffuse through the liquid membrane to react with the stripping solution at the membrane-strip interface. Finally, Dy and Eu ions are released in the strip solution and diffuse from the membrane-strip interface into the bulk strip solution due to concentration gradient [7].

2-1. Influence of Feed Solution pH

When Dy and Eu ions are carried from the feed phase to the stripping phase, protons are transported in the other direction (Fig. 3) and the gradient in proton concentration is the driving force for

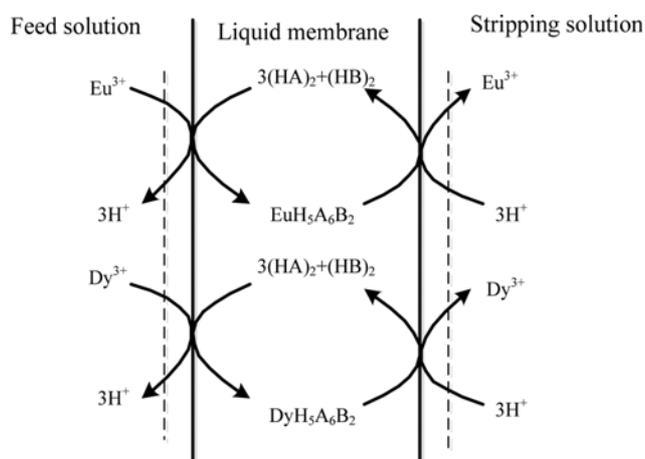


Fig. 3. Pertraction mechanism of Dy and Eu ions through liquid membrane using mixture of D2EHPA and Cyanex272 as carrier.

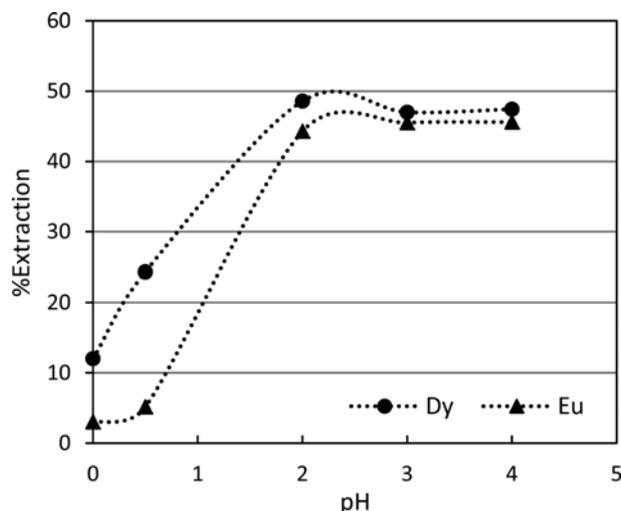


Fig. 4. Influence of feed phase pH on the extraction rate, stripping phase: 1 M HNO₃; carrier: (0.27 M D2EHPA+0.18 M Cyanex272); time of transport: 120 min.

the Dy and Eu transport. The dependence of the percentage of Dy and Eu extraction on the feed phase pH is shown in Fig. 4. The liquid membrane composition was kept at 0.27 M D2EHPA and 0.18 M Cyanex272 in kerosene and 1 M HNO₃ was maintained as stripping solution during the experiments. As observed, the percentage of extraction increases with increasing pH value of the feed phase. Dysprosium can be extracted more than Europium, because the equilibrium constant for Dysprosium using the mixture of D2EHPA and Cyanex272 was greater than that for Europium (see Table 3). The extraction of Dy and Eu ions by synergistic effect can be expressed as illustrated in Eqs. (7) and (8). According to both equations, when H⁺ concentration decreases, the more metal ions are extracted. However, maximum selectivity was observed for pH value of 0.5. At low pH values, the hard nature of D2EHPA increases;

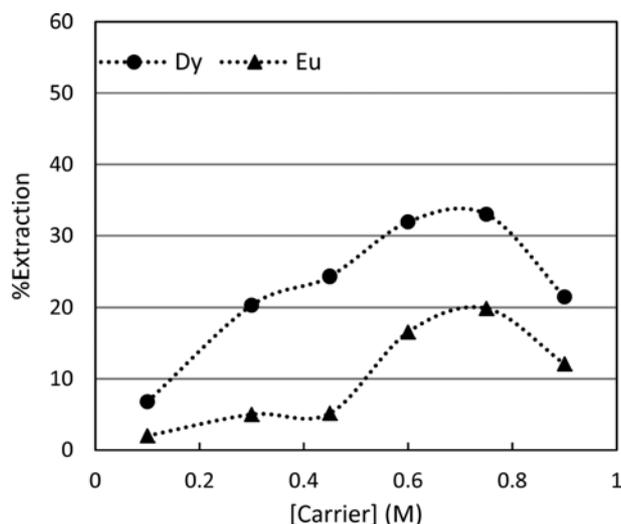


Fig. 5. Influence of carrier concentration on the extraction rate, stripping phase: 1 M HNO₃; feed phase pH: 0.5; time of transport: 120 min.

thereby according to hard and soft acid base theory (HSAB) [26], the harder ion (Dy) tends to interact more with D2EHPA molecules.

2-2. Influence of Carrier Concentration in Liquid Membrane

To investigate the role of the carrier concentration in the selectivity of Dy and Eu, various concentrations of D2EHPA and Cyanex272 with mole fraction ratio of 0.6/0.4 were used in combination of the liquid membrane. The feed phase pH was maintained at 0.5 and 1 M HNO₃ was maintained as stripping solution during the experiments. Fig. 5 shows that with an increase in carrier concentration, the extraction rate increases due to an increase in the amount of carrier-metal complexes, which enhances the Dy and Eu pertraction efficiency [27]. On the other hand, the availability of carrier at the feed-membrane interface increases with an increase in the carrier concentration. Any further increase in carrier concentration, results in a decrease in Dy and Eu extraction rate because, with a higher carrier concentration, the viscosity of the liquid membrane increases, thereby resulting in a reduction in the diffusion coefficients of the complexes [28]:

$$D = \frac{7.4 \times 10^{-8} (M_B)^{0.5} T}{\mu V_A^{0.6}} \quad (9)$$

where D is the diffusion-coefficient ($\text{cm}^2 \text{s}^{-1}$), M_B is diluent molar mass (g mol^{-1}), T is the absolute temperature (K), μ is the dynamic viscosity of the liquid membrane (mPa s) and V_A is the molar volume of solute A at its boiling temperature ($\text{cm}^3 \text{mol}^{-1}$). As is evident from this equation, D is inversely proportional to μ . However, the difference between Dy and Eu is barely dependent on the carrier concentration. The maximum selectivity was found for 0.45 M (Cyanex272+D2EHPA) in the liquid membrane.

2-3. Influence of Stripping Agent Concentration

The driving force for the Dy and Eu ions pertraction is the pH gradient between the feed and stripping phases. Accordingly, the influence of this gradient was studied by determining extraction rate as a function of proton concentration in the stripping phase,

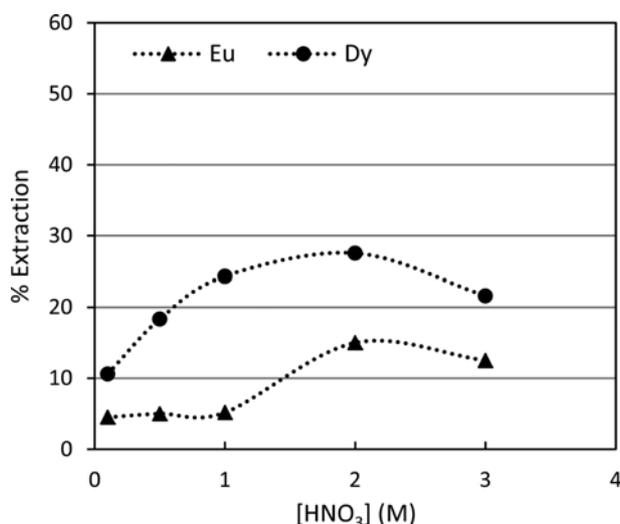


Fig. 6. Influence of stripping phase concentration on the extraction rate, carrier: (0.27 M D2EHPA+0.18 M Cyanex272); feed phase pH: 0.5; time of transport: 120 min.

as shown in Fig. 6. It indicates that, with an increase in HNO₃ concentration (up to 2 M), Dy and Eu pertraction increases due to the increase of H⁺ concentration, which accelerates the dissociation of metal-carrier complexes at the membrane-strip interface. Beyond 2 M HNO₃, the decrease in the metal transport may be due to the combination of carrier molecules with the hydrogen ions at the membrane-strip interface, so there will be a smaller amount of carrier for synergistic complex formation [29]. Obviously, the effective HNO₃ concentration for the transport and separation of Dy and Eu was 1 M. Under the optimal conditions assessed, the maximum separation factor, $S_F(\text{Dy}/\text{Eu})$, was calculated to be 4.7.

2-4. Kinetic Modeling of Dy Pertraction

As shown in Fig. 3, the transport process of Dy ions across SLM involves the following steps: (i) the Dy ions diffuse through the aqueous feed boundary layer and react with the carrier at feed-membrane interface resulting in the formation of Dy complex, (ii) the Dy complex diffuses through the membrane to the stripping phase due to concentration gradient of the Dy complex, (iii) the Dy is released to the stripping phase at the membrane-strip interface. The extraction of Dy from nitrate medium using synergistic mixture of D2EHPA and Cyanex272 is described by Eq. (7).

By applying Fick's first law, the flux across the aqueous diffusion layer (J_f) and the flux across the membrane layer (J_m) could be described as:

$$J_f = \frac{D_f}{d_f} ([\text{Dy}^{3+}]_{f,i} - [\text{Dy}^{3+}]_{f,e}) \quad (10)$$

$$J_m = \frac{D_m}{d_m \tau} ([\text{DyH}_5\text{A}_6\text{B}_2]_{m,f} - [\text{DyH}_5\text{A}_6\text{B}_2]_{m,s}) \quad (11)$$

where $[\text{Dy}^{3+}]_f$ corresponds to the concentration of Dy on the feed side, $[\text{Dy}^{3+}]_{f,i}$ corresponds to the concentration of Dy at the feed-membrane interface, and D_f and D_m are the diffusion coefficients of Dy in the feed phase and membrane, respectively. The thickness of the diffusion layer in the feed phase is represented by d_f and d_m as the membrane thickness. Tortuosity factor is required to convert the membrane thickness to an effective diffusion path length. $[\text{DyH}_5\text{A}_6\text{B}_2]_{m,f}$ corresponds to the concentration of the Dy complex in the feed-membrane interface and $[\text{DyH}_5\text{A}_6\text{B}_2]_{m,s}$ corresponds to the concentration of the Dy complex at the membrane-strip interface.

Assuming the fast chemical reaction compared to the diffusion rate, local equilibrium is reached at the interface [30]. Thus, at steady state, $J_f = J_m = J$ and combining Eqs. (5), (10) and (11), the following formula is obtained:

$$J = \frac{1}{\frac{d_f}{D_f} + \frac{d_m \tau [\text{H}^+]^3}{D_m K_{\text{ex}} [(\text{HA})_2]^3 [(\text{HB})_2]}} [\text{Dy}^{3+}]_f \quad (12)$$

$$= \frac{1}{\Delta_f + \frac{\Delta_m [\text{H}^+]^3}{K_{\text{ex}} [(\text{HA})_2]^3 [(\text{HB})_2]}} [\text{Dy}^{3+}]_f$$

where Δ_f and Δ_m are the resistance terms in the membrane and diffusion layer in the feed phase, respectively.

The permeability coefficient can be defined as:

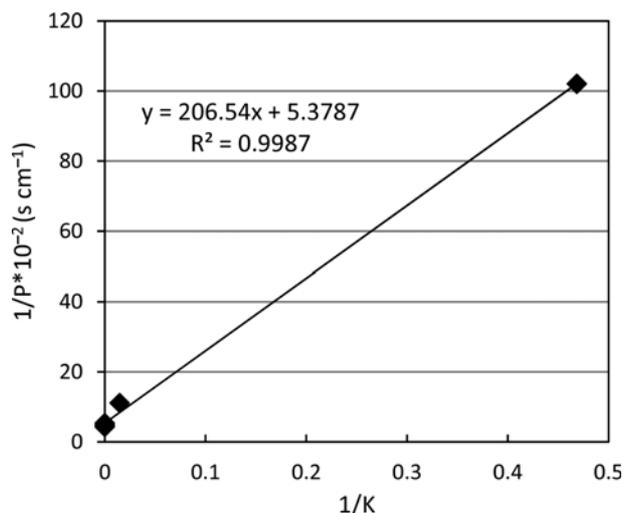


Fig. 7. Plot of $1/P$ vs $1/K$ at different feed phase pH; stripping phase: 1 M HNO_3 ; carrier: (0.27 M D2EHPA+0.18 M Cyanex272).

$$P = \frac{J}{[\text{Dy}^{3+}]_f} = \frac{1}{\Delta_f + \frac{\Delta_m [\text{H}^+]^3}{K_{\text{ex}}[(\text{HA})_2]^3[(\text{HB})_2]}} \quad (13)$$

$$\frac{1}{P} = \frac{\Delta_m [\text{H}^+]^3}{K_{\text{ex}}[(\text{HA})_2]^3[(\text{HB})_2]} + \Delta_f \quad (14)$$

The plot of $1/P$ as a function of $1/K$, where $K = (K_{\text{ex}}[(\text{HA})_2]^3[(\text{HB})_2][\text{H}^+]^3)$ for different feed acidity and at constant extractant concentrations (0.27 M D2EHPA and 0.18 M Cyanex272), gives a straight line with a slope Δ_m and intercept Δ_f (Fig. 7). The values of Δ_m and Δ_f were determined from the proposed kinetic model as $2.06 \times 10^4 \text{ s cm}^{-1}$ and 537.87 s cm^{-1} , respectively.

The transport resistance in the membrane can be expressed as follows: $\Delta_m = \tau d_m / D_m$. Considering the value of τ as 1.17 (see Table 1) and d_m as $150 \times 10^{-4} \text{ cm}$, the D_m value was calculated as $8.77 \times 10^{-7} \text{ cm}^2 \text{ s}^{-1}$. The mass transfer coefficient in the feed phase was evaluated as $\Delta_f^{-1} = 1.86 \times 10^{-3} \text{ cm s}^{-1}$.

2-5. Stability of SLM

To date, instability of SLM is the major problem that restricts a large-scale process application of the system. Stability means that no direct connection between the feed and the stripping phase occurs, which leads to non-facilitated transport [31]. Neplenbroek et al. [32] showed that the instability in SLM was created by a loss of the membrane phase through the formation of emulsion droplets induced by lateral shear forces. The membrane is more stable when a less stable emulsion is formed between the LM phase and the feed solution. Several instability mechanisms have been proposed, including organic solubility, pressure difference over the membrane, osmotic pressure, pore-blocking and shear-induced emulsion formation [33]. Leepipatpiboon et al. [34] showed that the polarity of the diluent is the main factor influencing the stability of the SLM. In the current study, experiments were performed to observe the stability of the SLM in terms of the carrier leaching from the support. The membrane stability was checked by operating the same membrane continuously over a period of six days. As is evident

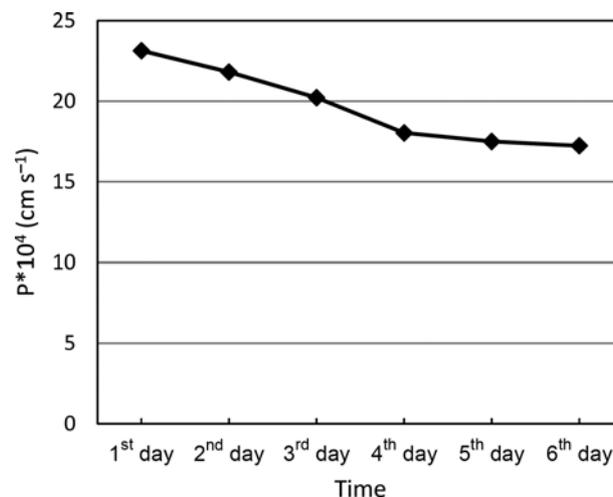


Fig. 8. Membrane stability over six consecutive days of operation, stripping phase: 1 M HNO_3 ; feed phase pH: 0.5; carrier: (0.27 M D2EHPA+0.18 M Cyanex272).

from Fig. 8, there is no significant decrease in permeability coefficient within a period of six days; hence the stability of the membrane was satisfactory.

CONCLUSION

The synergistic extraction and separation of Dy and Eu through SLM impregnated with a mixture of D2EHPA and Cyanex272 dissolved in kerosene were investigated.

Solvent extraction studies showed that the most synergistic effect for the extraction of Dy was produced by the mixture of D2EHPA and Cyanex272 with mole fraction ratio of 0.6/0.4. Utilizing this composition of carriers in the SLM system, the maximum separation factor (Dy/Eu) of 4.7 was observed for 0.45 M (D2EHPA+ Cyanex272) as a carrier, keeping the feed pH at 0.5 and 1 M HNO_3 as a stripping phase. The results showed that separation of Dy and Eu was highly dependent on the pH of the feed solution.

Using the kinetic modeling, the diffusion coefficient and mass transfer coefficient of Dy were evaluated as $8.77 \times 10^{-7} \text{ cm}^2 \text{ s}^{-1}$ and $1.86 \times 10^{-3} \text{ cm s}^{-1}$, respectively.

Stability of the membrane was found to be satisfactory over six cycles of operation indicating that efficient liquid membrane system could be operated using mixture of D2EHPA and Cyanex272 as a carrier for Dy pertraction from nitrate media.

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NOMENCLATURE

D : distribution coefficient
 $[\text{M}^{3+}]$: metal concentration $[\text{mol L}^{-1}]$
 S_f : selectivity factor

R	: synergistic enhancement factor
K_{ex}	: equilibrium constant
D_m	: diffusion coefficient in the membrane phase [$\text{cm}^2 \text{s}^{-1}$]
D_f	: diffusion coefficient in the feed phase [$\text{cm}^2 \text{s}^{-1}$]
d_m	: thickness of membrane [cm]
d_f	: thickness of the diffusion layer in the feed phase [cm]
Δ_m	: resistance in the membrane [s cm^{-1}]
Δ_f	: resistance in the diffusion layer in the feed phase [s cm^{-1}]
P	: permeability coefficient [cm s^{-1}]
M_B	: diluent molar mass [g mol^{-1}]
T	: temperature [K]
V_A	: molar volume of solute A at its boiling temperature [$\text{cm}^3 \text{mol}^{-1}$]
J_m	: flux across the membrane [$\text{mol cm}^{-2} \text{s}^{-1}$]
J_f	: flux across the diffusion layer in the feed phase [$\text{mol cm}^{-2} \text{s}^{-1}$]

Greek Letters

ε	: porosity
τ	: tortuosity factor
μ	: dynamic viscosity of the liquid membrane [mPa s]

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