



## On the cost of electrodialysis for the desalination of high salinity feeds



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### HIGHLIGHTS

- Feed water was desalinated from 192 down to 0.24 g NaCl/kg solution.
- Energy requirements are similar but treatment costs are lower than evaporation.
- Voltage optimisation can further reduce ED treatment costs by 30–60%.

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### ABSTRACT

We propose the use of electrodialysis to desalinate produced waters from shale formations in order to facilitate water reuse in subsequent hydraulic fracturing processes. We focus on establishing the energy and equipment size required for the desalination of feed waters containing total dissolved solids of up to 192,000 ppm, and we do this by experimentally replicating the performance of a 10-stage electrodialysis system. We find that energy requirements are similar to current vapour compression desalination processes for feedwaters ranging between roughly 40,000–90,000 ppm TDS, but we project water costs to potentially be lower. We also find that the cost per unit salt removed is significantly lower when removed from a high salinity stream as opposed to a low salinity stream, pointing towards the potential of ED to operate as a partial desalination process for high salinity waters. We then develop a numerical model for the system, validate it against experimental results and use this model to minimise salt removal costs by optimising the stack voltage. We find that the higher the salinity of the water from which salt is removed the smaller should be the ratio of the electrical current to its limiting value. We conclude, on the basis of energy and equipment costs, that electrodialysis processes are potentially feasible for the desalination of high salinity waters but require further investigation of robustness to fouling under field conditions.

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### 1. Introduction

We have experimentally investigated factors affecting the cost of electrodialysis (ED) for the desalination of high salinity feeds, focusing on the dependence of the cost of salt removal upon diluate salinity. We have also developed a numerical model for the system, validated it against the experimental results and identified a strategy to optimise the stack voltage such that the sum of equipment and energy costs are minimised. Our motivation for this investigation was the desalination of produced waters in unconventional oil and gas extraction where, amongst other factors, the presence of high levels of total dissolved solids can disincentivise water reuse.

Water reuse in hydraulic fracturing is of great interest both from an environmental perspective, as it reduces water use and minimises disposal through deep-well injection, but also from an economic perspective as water management costs can account for between 5% and 15% of drilling costs [1].

For the purpose of this investigation, we were most interested in flows of water during the life-cycle of a well, which are depicted in Fig. D.1. For reuse to be economical, the savings in the sourcing, disposal and transport of water must outweigh any increased costs of treatment or of chemicals in the formulation of the fracturing fluid. This means that regional differences in recycling rates are strongly influenced by regional differences in sourcing, disposal and transport costs. For example, reuse rates are currently greatest in the Marcellus shales [3] (reused water makes up 10–15% of the water needed to fracture a well) where transport and disposal costs can reach \$15–18/bbl (\$94–113/m<sup>3</sup>) [4]. The initial rate at

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## Nomenclature

### Roman symbols

$A_m$	membrane area, $m^2$
$C$	concentration, $\text{mol}/m^3$
$D$	diffusivity, $m^2/s$
$E^s$	specific energy of salt removal, $\text{kW h}/\text{lb}$ or $\text{kW h}/\text{kg}$
$E^w$	specific energy of water produced, $\text{kW h}/\text{bbl}$ or $\text{kW h}/m^3$
$h$	channel height, $m$
$i$	current density, $A/m^2$
$I$	current, $A$
$k$	conductivity, $S/m$
$K_E$	energy price, $\$/\text{kW h}$
$K_Q$	area normalised equipment price, $\$/m^2$ membrane
$L_s$	membrane salt permeability, $m^2/s$
$L_w$	membrane water permeability, $\text{mol}/m^2 \text{ s bar}$
$m$	slope
$M$	molar mass, $\text{kg}/\text{mol}$
$m_s$	molal concentration, $\text{mol}/\text{kg w}$
$N$	number of moles, $\text{mol}$
$n_{cp}$	number of cell pairs, –
$\bar{r}$	membrane surface resistance, $\Omega m^2$
$R$	universal gas constant, $J/\text{mol K}$
Re	Reynolds number
Sc	Schmidt number
Sh	Sherwood number
$t$	process time, $s$
$T$	system life, years
$\bar{T}_{cu}$	integral membrane counterion transport number, –
$t_{cu}$	solution counter-ion transport number, –
$T_s^{cp}$	cell pair salt transport number, –
$T_w^{cp}$	cell pair water transport number, –
$V_{corr}$	stack voltage corrected for concentration polarisation, $V$
$V_{stack}$	stack voltage, $V$
$V$	volume, $m^3$
$\dot{V}$	volume flow rate, $m^3/s$

$w$	mass, lbs or kg
$x$	concentration, $\text{mol salt}/\text{mol water}$

### Greek symbols

$\Delta$	change
$\epsilon$	error
$\Lambda$	molar conductivity, $S m^2/\text{mol}$
$\mu$	chemical potential, $J/\text{mol}$
$\nu$	viscosity, $m^2/s$
$\Xi^s$	specific cost of salt, $\$/\text{lb}$ or $\$/\text{kg}$
$\Xi^w$	specific cost of water, $\$/\text{bbl}$ or $\$/m^3$
$\pi$	osmotic pressure, $\text{bar}$
$\rho$	density, $\text{kg}/m^3$
$\sigma$	spacer shadow factor, –
$\tau^s$	specific process time, $\text{days}/\text{lb}$ or $\text{days}/\text{kg}$
$\tau^w$	specific process time, $\text{days}/\text{bbl}$ or $\text{days}/m^3$

### Subscripts

<i>am</i>	anion exchange membrane
<i>c</i>	concentrate
<i>circ</i>	circuit
<i>cm</i>	cation exchange membrane
<i>d</i>	diluate
<i>el</i>	electrode
<i>i</i>	stage number
<i>j</i>	time period
<i>m</i>	membrane surface
<i>p</i>	pump
<i>r</i>	rinse
<i>s</i>	salt
<i>s</i>	water

### Superscripts

<i>f</i>	final
<i>in</i>	initial

which produced water flows to the surface (e.g. within the first 10 days) also influences the viability of reuse as low initial produced water volume flow rates making the logistics of reuse more difficult [3,5].

Moving to the costs of reuse, and setting aside the expense associated with logistics, the costs come primarily in the form of: increased water treatment costs; increased chemical costs in the formulation of the hydraulic fracturing fluid to mitigate undesirable feed water properties; and/or reduced oil or gas production from the well. By and large, the increase in treatment costs is highest, and the increase in chemical costs lowest, when produced water is treated with mechanical vapour compression. Vapour compression provides high purity water for the formulation of the hydraulic fracturing fluid but is expensive – ranges of roughly 5–8  $\text{kW h}/\text{bbl}$  (32–50  $\text{kW h}/m^3$ ) of distillate<sup>1</sup> [7] and \$3.50–6.25/bbl (\$22–39/ $m^3$ ) of distillate [1] have been reported for the treatment of produced waters. Direct reuse, whereby produced water is directly blended with freshwater before formulation of the fracturing fluid, results, by and large, in the lowest treatment costs but greater chemical costs for fluid formulation and perhaps a decline in the well's production. Increased costs associated with reuse, depending on the degree of treatment employed, can come in the form of: increased friction reducer and scale inhibitor demand with high

chloride contents; increased scaling within the shale formation with the presence of divalent ions; increased corrosion of pipes; increased levels of sulphate reducing bacteria resulting in the production of  $H_2S$  gas [8]; and a reduction in the performance of coagulation/flocculation, flotation, gravity settling and plate and frame dewatering equipment due to residual unbroken polymer gel [9].

Many of the challenges faced in reuse can be dealt with through primary treatment that removes suspended solids, oil, iron, unbroken polymers and bacteria [9], generally at a cost much below complete desalination (circa \$1/bbl (\$6.3/ $m^3$ ) compared to \$3.50–6.25/bbl (\$22–39/ $m^3$ ) for complete desalination [1]). The need for the removal of all solids, suspended and dissolved, is less clear. Opinions vary as to the level of total dissolved solids (TDS) that can be tolerated [10] and a complete understanding of issues of chemical compatibility remains elusive [2]. There is evidence that, with improved chemical formulations, high salinity produced waters may be reused without desalination, particularly in the formulation of fluids for slickwater processes [11–16] (processes with high volume flow rates to avoid premature settling of sand, which serves to maintain fractures propped open) and to some extent for cross-linked gel fracturing processes [17] (lower volume flow rate processes employing low molecular weight guar gum based gels to ensure proppant remains suspended). However, the increase in chemical costs associated with such formulations not evident. Depending on the fracturing fluid desired, chemical use can be significant. Fedotov et al. [9] indicated that the use of drag reducing

<sup>1</sup> 6.4  $\text{kW h}/\text{bbl}$  (40  $\text{kW h}/m^3$ ) has been reported for 72.5% recovery of feedwater with total dissolved solids of 50,000  $\text{mg}/\text{L}$  [6].

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