



# Effect of electrode area on demineralization performance and the distribution of current density in an industrial-scale electro dialysis stack



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## HIGHLIGHTS

- Reduction of electrode area negatively affects performance.
- Reduction of area generates current density peaks at the edges of the electrode.
- The macrohomogeneous 2D model can be used to study the effect of electrode area.
- The macrohomogeneous 2D model cannot be used for prediction of performance.

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## ABSTRACT

Effect of electrode area on demineralization performance and the distribution of current density in an industrial-scale electro dialysis stack was studied. A segmented electrode was used for this purpose. Several combinations of active/inactive segments of the electrode were tested and results obtained were mutually compared. It was found that any reduction of the electrode area caused the demineralization performance to decline and generated extremely high local current densities at the edges of the electrode. The experimental results were qualitatively in correlation with a macrohomogeneous 2D model for computation of the distribution of electric potential, electrolyte concentration in the dilute and current density in the electro dialysis stack.

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## 1. Introduction

Electrodialysis (ED) is an electromembrane separation process that is mainly used for partial demineralization of whey, low TDS waters and aqueous solutions such as river or brackish waters, municipal waste waters or some industrial waste waters.

Current trends in development of conventional ED are toward a higher level of modularity and reliability, simplicity of operation and lower maintenance of the process modules. At the same time, competition of other (membrane) separation processes in the field of low TDS water treatment such as reverse osmosis (RO) or capacitive deionization (CapDI) requires significant reduction of costs for production of industrial-scale ED stacks. The research activities are mainly focused on the development of novel, low cost ion-exchange membranes in this field as the ion-exchange membranes are the major part of manufacturing costs. Ti/Pt plate electrodes are another important part of manufacturing costs of industrial-scale ED stacks, especially in cases

when electrode polarity reversal is used. For instance, Ti/Pt electrodes constitute approximately 20% of manufacturing costs of EDR-III stacks with a higher number of cell pairs (above 400) available from MEGA a.s. (Czech Republic). These electrodes can be replaced with cheaper alternatives such as mixed metal oxide (MMO) electrodes in some applications. Recently, use of CapDI electrodes in ED is of particular interest in some applications, especially owing to the possibility of complete elimination of the electrode loop and associated equipment [1]. Those who are skilled in the art know that electrodes for electrochemical systems are available also in other than plate forms. For example, mesh or rod electrodes are available. These alternatives have been adopted on a limited scale or have never been adopted in ED, however.

In some ED stacks available in the market, electrode area is reduced compared to the active area of an ion-exchange membrane. This might be an alternative way of reduction of manufacturing costs. To date, relatively little has been published on the effect of electrode area on demineralization performance of ED stacks. It is therefore the aim of this work to evaluate the feasibility of building industrial-scale ED stacks with conventional plate electrodes and reduced electrode area in order to reduce manufacturing costs. Successful completion of this

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**List of symbols**

$C$	Salt removal (%)
$F$	Faraday constant ( $= 96485 \text{ C mol}^{-1}$ )
$J_S$	Local flux density ( $\text{mol m}^{-2} \text{ s}^{-1}$ )
$L$	Active length of an ion-exchange membrane (flow path length) (m)
$Q_D$	Dilute flow rate ( $\text{m}^3 \text{ s}^{-1}$ )
$R$	Molar gas constant ( $= 8.31446 \text{ J K}^{-1} \text{ mol}^{-1}$ )
$R_A$	Sheet resistance of an ion-exchange membrane ( $\Omega \text{ m}^2$ )
$T$	Thermodynamic temperature (K)
$U$	Voltage (V)
$a_i$	$i$ -th coefficient of the kinetics of electrolyte separation from the dilute ( $\text{mol}^{1-i} \text{ m}^{3i-2} \text{ s}^{-1}$ )
$c_S$	Molar concentration of an electrolyte ( $\text{mol m}^{-3}$ )
$d_C$	Thickness of a flow chamber (m)
$d_{CP}$	Thickness of a cell pair (m)
$d_M$	Thickness of an ion-exchange membrane (m)
$d_{stack}$	Thickness of a membrane (electrodialysis) stack (m)
$\mathbf{j}$	Current density vector ( $\text{A m}^{-2}$ )
$n$	Number of cell pairs
$x$	Horizontal coordinate (m)
$y$	Vertical coordinate (m)
$x_{rel}$	Relative horizontal coordinate $x_{rel} = \frac{x}{d_{stack}}$ (-)
$y_{rel}$	Relative vertical coordinate $y_{rel} = \frac{y}{L}$ (-)
$w$	Active width of an ion-exchange membrane (m)
$m$	Degree of a polynomial (-)
$\Lambda_S$	Molar conductance of an electrolyte ( $\text{S m}^2 \text{ mol}^{-1}$ )
$\alpha$	Coefficient related to the effect of a spacer on ohmic resistance of a flow chamber in the ED stack (-)
$\phi$	Electric potential (V)
$\rho$	Electric resistance of an ion-exchange membrane ( $\Omega \text{ m}$ )
<b>Subscripts</b>	
$AM$	Related to the anion-exchange membrane
$C$	Related to the concentrate
$CIN$	Related to the concentrate inlet into the ED stack
$CM$	Related to the cation-exchange membrane
$D$	Related to the dilute
$DIN$	Related to the dilute inlet into the ED stack

task requires that the potential performance loss associated with the reduction of electrode area is lower than the reduction of manufacturing costs.

## 2. Experimental

EDR-III/400–0.68 Type 13 industrial-scale ED stack available from MEGA a.s. (Czech Republic) was used in this work. The stack comprised 400 cell pairs of RALEX® CM-PES and AM-PES heterogeneous ion-exchange membranes available from the same manufacturer. Standard electrodes are divided into two parts, each part having a size of approximately half of the membrane active area in this stack. The purpose for this solution is a quite large membrane active area ( $350 \times 1300 \text{ mm}$ ) and limited commercial availability of anodes of that size, but also mechanical issues such as bending of large electrodes manufactured from a relatively thin (2 mm) titanium plate and practical issues when installing such electrodes.

In order to evaluate the effect of an electrode with reduced area in the stack, the standard two part electrodes were replaced with segmented electrodes. This is a common technique frequently used in electrochemistry including characterization of ED stacks. For example, Tanaka used an electrode divided into 8 segments to study the

distribution of current density and the limiting current density in an ED stack [2]. Kodým et al. used an electrode divided into 6 segments in order to verify their 2D mathematical model of mass and charge transfer in an ED stack [3].

In this work, each electrode was formed of 10 segments. The active surface area of each electrode segment was  $350 \times 125 \text{ mm}$ . Adjacent segments were separated with a 5 mm wide gap. A power cable to each electrode segment was equipped with an on/off switch and an ammeter, see Fig. 1, which allowed for different combinations of active/inactive segments of the electrode to be tested and mutually compared.

An overview of the studied combinations of active/inactive electrode segments is shown in Table 1. All tested combinations of active/inactive electrode segments are symmetrical about the two planes that bisect the ED stack vertically and horizontally. Case S1 is equal to the standard full-area electrode. Case S2 simulates reduction of electrode area by 20% and cases S3, S5 and S6 simulate reduction of electrode area by 40%. Case S6 was not tested, but its potential benefits are only discussed based on prediction of its performance using a mathematical model.

Described segmentation and tested combinations of active/inactive segments of the electrode were chosen with regard to the potential replacement of the standard electrodes with alternative electrode arrangements, which requires a small number (2 to 3) of segments and easy installation in the end-plate of the ED stack.

Both diluting and concentrating loops of the ED stack were fed with the same aqueous solution of NaCl at a temperature of  $25^\circ \text{C}$  from a tank. After one pass of the liquid through the stack, both dilute and concentrate were returned back to the tank, thus maintaining constant composition of the feed solution. To avoid chlorine formation during the tests, the electrode loop was separated from the diluting and concentrating one and an aqueous solution of  $10 \text{ kg m}^{-3} \text{ Na}_2\text{SO}_4$  was circulated in the electrode loop. Electric potential was applied on the electrode terminals from the external DC power supply.

The ED stack was gradually tested at all combinations of the following operating conditions:

- an aqueous feed solutions of 17.11, 51.33 and  $85.56 \text{ mol m}^{-3} \text{ NaCl}$ ,
- a dilute flow rate of 24, 32 and  $40 \text{ m}^3 \text{ h}^{-1}$ ,
- a voltage of 0.25 to 1.25 V/cell pair with a step of 0.25 V/cell pair,
- an electrode configuration of S1, S2, S3 and S5, see Table 1.

The total electric current, electric current in the respective electrode segments, dilute and concentrate pH and conductivity at the outlet from the stack were monitored during the tests.

The ED stack was operated under each combination of operating conditions for at least 10 min, although much less time was required to reach the steady state operation. Experimental data readings obtained at the steady state were used for further evaluation.

## 3. Mathematical modeling

To evaluate a dilute flow rate needed to achieve a certain salt removal per a single pass of the liquid through the ED stack at the given feed concentration and voltage, means of approximating the dependence of salt removal on the dilute flow rate is required. Since only three experimental data points are available for three flow rate conditions, linear regression with an univariate quadratic function can be used. It can be shown that this approach provides sufficient accuracy for the purposes of this work. This is only due to a relatively narrow range of tested flow rates, however. Therefore, an alternative and generally better approach was used in this work and is described below.

The local mass balance of an electrolyte in a volume element of the diluting chamber is

$$\frac{\partial c_{S,D}}{\partial y} = -\frac{nwJ_S}{Q_D} \quad (1)$$

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