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An operational and economic study of a reverse osmosis desalination system for potable water and land irrigation

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HIGHLIGHTS

• A reverse osmosis (RO) desalination plant for the MOD river in Iraq is presented.
• Variation of temperature, pressure and recovery in membrane module are considered.
• Water for drinking and irrigation applications are discussed.
• Salt rejection rates sensitivities to temperature and pressure are highlighted.
• An economic analysis is carried out to determine the total water cost (TWC).

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ABSTRACT

Desalination is a method for producing water for human consumption, irrigation or industrial utilisation. In this study, a reverse osmosis (RO) system for brackish water desalination was theoretically investigated to produce both potable drinking and agricultural water with a lower overall and specific energy consumption. As a case study, the Main Outfall Drain in Iraq is used as the brackish water source. A numerical model based on solution-diffusion theory was developed in Matlab Simulink and used to analyse the design and performance of an RO system. The effect of feed water temperature, pressure, salinity and recovery ratio on the efficiency of the whole RO system was investigated for a wide range of design considerations. The design of an RO system for this application was optimised and economic assessment carried out. Results show that with boosting recovery ratio from 30% to 60%, the specific energy of desalinated water production below 400 ppm was reduced from 2.8 kWh/m³ to a more economically favourable value of 0.8 kWh/m³, when utilizing a pressure exchanger as a recovery device. Salt rejection was reduced from 97% to 88% to obtain large quantities of water for irrigation with an acceptable salinity (<1600 ppm), for agricultural use. The reduction in salt rejection is influenced by the feed water temperature and pressure; also the average pore diameter of the RO membrane and in turn determines the reduction in system energy consumption. It was found that the total cost to produce 24,000 m³/d of water from a feed salinity of 15,000 ppm and a water quality of <400 ppm would be 0.11 £/m³ with a corresponding investment cost of £14.4 million for the drinking water, and for irrigation) obtained product (<1600 ppm) are £0.9/m³ and £11.3 million.

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1. Introduction

In recent years there has been considerable growth in the utilisation of reverse osmosis (RO) processes in major desalination plants [1,2]. An RO purification system uses a semi-permeable membrane to remove ions, proteins, and organic chemicals which are generally not easily removed using other conventional treatments [3]. Among the benefits of RO are its small footprint, a modular design and the possibility of automatic process control and relatively low-cost of water production [4]. RO has been used widely for various water and wastewater treatment processes [5,6], in areas with scarce water supplies (as a means of seawater desalination) and importantly for this study the treatment of brackish water. However RO desalination suffers from a high energy input demand, fouling of the membranes, and low-quality of the water compared to thermal technologies which produce very high quality [2,3]. A high hydraulic pressure is required to overcome the osmotic pressure of the saline feed water solution, which means a high consumption of energy is required when pressurizing the feed flow. Over the past 40 years, as a result of on-going technological advances [6] there has been a significant reduction in the energy required.
### Nomenclature

- \( A \): Area, \( m^2 \)
- \( A_w \): Permeability coefficient, \( m^3/Pa \)
- \( B_s \): Solute transport parameter, \( m/s \)
- \( \xi \): Average salinity through the membrane element, \( mol/m^3 \)
- \( C_{ch} \): Cost of chemical treatment, £/m³
- \( CD_{mem} \): Membrane cost, £
- \( C_r \): Unit power cost, £/kWh
- \( C_f \): Concentration of feed water, \( mol/m^3 \)
- \( C_m \): Solute concentration in the membrane, \( mol/m^3 \)
- \( C_{RO} \): Mass fraction of salt in permeate, %
- \( C_p \): Solute concentration at permeate, \( mol/m^3 \)
- \( C_r \): Concentration in the concentrate, \( mol/m^3 \)
- \( C_w \): Water concentration in the membrane, \( mol/m^3 \)
- \( D_w \): Water diffusivity, \( m^2/s \)
- \( D_s \): Diffusivity of solute, \( m/s \)
- \( E \): Specific energy consumption, kWh/m³
- \( E_{TUR} \): Turbine energy, kWh
- \( E_{mem} \): Membrane activation energy, J/mol
- \( E_{pump} \): Pump energy consumption, kWh
- \( F_1 \): Plant load factor, %
- \( GBP \): Great British Pound, £
- \( i_{eff} \): Effective discount rate relation between the future value and present value
- \( J_s \): Solute transport, \( m/s \)
- \( J_W \): Permeate flux, \( m/s \)
- \( K_s \): Solubility of solute, \( m^3/s \)
- \( n \): Number of membrane elements
- \( P_{C_{mem}} \): Cost per membrane, £
- \( P_{f} \): Feed water pressure, Pa
- \( P_{m} \): Pressure after the intake pump, bar
- \( P_{m} \): Annual membrane replacement factor, %
- \( P_p \): Permeate pressure, Pa
- \( P_r \): Rejected pressure, Pa
- \( \Delta P \): Transmembrane pressure difference, Pa
- \( Q_f \): Feed flow rate, \( m^3/day \)
- \( Q_p \): Permeate flow rate, \( m^3/day \)
- \( Q_{vol} \): Annual volume flow rate of product water, \( m^3 \)
- \( Q_{mem} \): Permeate flow rate per membrane element, \( m^3/s \)
- \( Q_{rej} \): Mass flow rate of permeate in one element, kg/s
- \( Q_{rej} \): Rejected flow rate, \( m^3/day \)
- \( Q_{bypass} \): Amount of water mixes with the permeate to achieve the required salinity, \( m^3 \)
- \( R \): Gas constant, J/mol-k
- \( r_n \): Nominal escalation rate which effects of resource depletion, increased demand and inflation, %
- \( r_r \): Recovery ratio, %
- \( R_s \): Salt rejection, %
- \( T \): Temperature, K
- \( TCF \): Temperature correction factor at T, %
- \( V_w \): Water molar volume, \( m^3 \)
- \( W \): Work, kWh

### Greek symbols

- \( \delta m \): Membrane thickness, m
- \( \eta \): Efficiency, %
- \( \Delta \eta \): Osmotic pressure difference, Pa

### Subscripts

- \( p \): permeate
- \( r \): rejected

Nevertheless, the overall energy utilisation remains considerable for RO desalination and as such the cost-effectiveness of production is highly sensitive to changes in energy prices and policy decisions related to greenhouse gas emissions. Typically, in an RO plant, the production of one cubic meter of freshwater from seawater uses 3–10 kWh of electricity, and between 0.5 and 2.5 kWh from brackish water [7–9]. Another challenge with RO systems is the disposal of brine – an output which has a potentially damaging impact on the local marine environment. The accumulation of solids in the feed solution on the surface of the membranes, i.e. membrane fouling, is a further challenge for an RO system [10]. In fact, this is particularly significant where RO is employed for the wastewater treatment especially where the feed water contains a large amount of solids. Membrane solids is a complex phenomenon involving the deposition of several types of solids on the membrane surface. If it occurs, the permeability of the RO membrane is lowered, which in turn affects the energy requirement [11]. Suitable pre-treatment technologies can minimise membrane fouling to some degree, although it will also have its own energy demands. Improved rejection can be achieved by adding treatment stages or polishing steps, which would lead to substantially higher capital and running costs.

Much work has been carried out to reduce these limitations of RO desalination, including the development of novel membranes with high permeability to water but low permeability to salt [11]. In order to reduce energy consumption and fouling, investigations have been carried out into the hydrodynamics of feed flow inside an RO membrane module. Consideration has been given to various techniques for pre-treatment and post-treatment in combination with the analysis of the characteristics of the feed water. The success of an RO system design requires robust analysis at the feasibility stage to evaluate alternative designs for more efficient design and operation, which can be later applied in its construction. Although membrane manufacturers have developed numerical models to support the design process, the principal area of attention has been on their performance rather than the optimisation of the complete desalination in terms of energy consumption and product water quality. Some research [12–14] has investigated the development of new models for the optimisation of membrane modules and the desalination plant. Nevertheless, the focus of previous research has not been on the impact of various designs and operating conditions on RO desalination performance [15–21]. In addition, several cost models have been developed during recent years; however, they were mostly focused on domestic and municipal [22,23]. Nonetheless, economic data of brackish water on industrials for drinking and irrigation in literature are significantly limited [24–26]. Any system analysis should be underpinned by a feasibility study which supports the selection of appropriate technologies for characterisation of the capital operating and investment costs. The main aims of this paper are to construct a numerical model of an RO system, with and without energy recovery, validate the model against reported measurement data, study the effect of operating parameters such as temperature, salinity, pressure and recovery ratio on RO efficiency, and optimize the RO desalination system in terms of energy requirement, salt rejection and total cost of water production.

### 2. Case study

Iraq has been experiencing an extreme water shortage in recent years, over the last four decades the amount of available water has diminished because of the use by upstream countries such as Turkey and Syria [27]. Recently, agricultural lands have been adversely affected by these shortages, and there is a need for a very large quantity of water...
to revive the dried marshlands in Southern Iraq. The Mesopotamian marshlands are the largest wetlands ecosystem in the Middle East and western Eurasia. They are crucial in terms of ecological, economical and hermetic importance, but they are in drought [28–30]. The special significance of these marshlands is habitat provided for migratory birds support for endangered species and the support provided for

### Table 1

| Zones          | M1     | M2     | M3     | M4     | M5     | M6     | M7     | M8     | M9     | M10   |
|----------------|--------|--------|--------|--------|--------|--------|--------|--------|--------|-------|
| PH             | 7.68   | 7.28   | 6.99   | 7.52   | 7.05   | 7.54   | 6.92   | 7.12   | 7.24   | 7.90  |
| Electrical conductivity (ds/m) | 2.92   | 3.85   | 5.10   | 7.31   | 7.67   | 9.35   | 9.52   | 10.53  | 14.02  | 14.96 |
| TDS (mg/l)     | 1868   | 2432   | 3264   | 4691   | 4966   | 6112   | 6092   | 6739   | 8972   | 9574  |
| Dissolved positive ions (mg/l) Ca | 1050   | 1000   | 1500   | 1215   | 1466   | 300    | 2975   | 720    | 670    | 8800  |
| Mg             | 350    | 385    | 550    | 330    | 520    | 170    | 987    | 390    | 180    | 2960  |
| K              | 14     | 14.5   | 14.5   | 14     | 15     | 8.0    | 18.5   | 18.5   | 10.5   | 8.0   |
| NH₄            | 0.65   | 8.89   | 9.38   | 4.93   | 0.26   | 0.29   | 0.86   | 6.16   | 0.76   | 7.67  |
| Dissolved negative ions (mg/l) NO₃ | 0.012  | 0.028  | 0.044  | 0.007  | 0.010  | 0.055  | 0.017  | 0.009  | 0.017  | 0.006 |
| SO₄            | 1585   | 1823   | 2272   | 1842   | 2089   | 913    | 3230   | 1141   | 1031   | 4955  |
| Cl (mg/l)      | 1738   | 1624   | 2669   | 1897   | 2692   | 568    | 5680   | 1363   | 874    | 1848  |
| Alkalinity (Mg/l) | 175   | 220    | 290    | 235    | 250    | 160    | 135    | 195    | 165    | 150   |
| Total organic carbon sediment (TOCS) (%) | 0.12  | 3.12   | 2.23   | 3.5    | 0.38   | 0.28   | 0.23   | 1.44   | 1.53   | 1.34  |
freshwater fisheries. Aboriginal communities have been living in these marshlands for millennia and include culturally significant historical sites such as the Garden of Eden [28–30].

Around 150,000 km² of Iraqi agricultural lands drain into by the Main Outfall Drain [1], formerly called The Third River, shown in Fig. 1. The MOD is situated between the Tigris and Euphrates Rivers, and passes through the main Mesopotamian marshlands south of these rivers. This work considers that the MOD river could provide Southern Iraq with a lifeline for potable and irrigation water and could supply water for 1.8 million people in All-Nasiriya City and irrigate 150,000 km² of farmland and replenish 20,000 km² of the southern marshlands [31].

According to the Iraqi Ministry of Agriculture, the river has suffered from very high salt concentrations since it was constructed in 2008. It has a length of 565 km from north of Baghdad to the Arabian Gulf with the total discharge of 210 m³/s. Drainage water is released into Shat AL Basra canal and then flows into the Arabian Gulf. The chemical properties of the water for different zones Fig. 1 close to Al-Nasiriya City are shown in Table 1 [32]. The salinity level increases gradually as it flows from the south of Iraq close to city of All-Nasiriya with total dissolved solid (TDS) levels between 6000 and 8500 ppm [33]. Recently, the MOD water has been used to revive the dried marshlands but this has resulted in a negative effect on diversity of life and agricultural due to its high salinity. Therefore, to satisfy the demands of agricultural irrigation, marshland revival and domestic water user the desalination of some of the MOD water would be hugely beneficial.

3. Model development, model based analysis and optimisation

The Matlab/Simulink and Thermlib blocks software tools were used to design an RO system numerical model. The schematic diagram of the RO desalination system with Turbine is shown in Fig. 2 and with Pressure exchanger is shown in Fig. 3, and the modelling equations as shown in Table 2. The main components of the RO system are a pump unit which supplies high pressure feed water, \( P_f \) and flow rate, \( Q_f \) to a membrane, a group of membrane modules, and an energy recovery device (hydraulic turbine and pressure exchanger) which generates energy from the rejected brine stream and directly powers a pump. The model was designed to predict the system performance and support

![Fig. 3. Schematic representation of the RO desalination model with PX energy recovery system.](image-url)

### Table 2
Summary of RO model equations.

| Meaning                  | Equation                                      | No. Reference | Reference |
|--------------------------|-----------------------------------------------|---------------|-----------|
| Permeate flux            | \( J_{PW} = A_w (\Delta P - \Delta n) \)     | 1             | [36,38,40]|
| Solvent permeability coefficient | \( A_w = \frac{Dw Cw Vw}{\delta m RT} \) | 2             | [36,40,41]|
| Solute transport         | \( J_s = B_s (C_m - C_p) \)                  | 3             | [38–40,42]|
| Solute permeability coefficient | \( B_s = \frac{D_s K_s}{\delta m} \)       | 4             | [40,41]   |
| Salt rejection           | \( R_s = \left[ 1 + \left( \frac{b_{water}}{b_{salt}} \right) \right]^{-1} \) | 5             | [40]      |
| Osmotic Pressure         | \( \Delta \pi = RT \sum (n/v) \)            | 6             | [38,41]   |
| Temperature correction factor | \( TCF = exp \left[ \frac{1}{2} \left( \frac{1}{C_p} - \frac{1}{C_f} \right) \right] \) | 7     | [43,44]   |
| Specific energy          | \( E = \frac{P_f Q_p}{\eta_{s,pump}} \)     | 8             | –         |
| Recovery ratio           | \( R = \frac{Q_f}{Q_p} \)                   | 9             | –         |
| Total mass balance       | \( Q_C = Q_f C_f + Q_R C_r \)               | 10            | –         |
| Delta pressure           | \( \Delta P = P_f - P_p \)                  | 11            | –         |

### Table 3
Model validation for RO model against reported measurement data.

| Parameters                  | Unites | Reported measurement data | Model | Percentage error |
|-----------------------------|--------|---------------------------|-------|------------------|
| Normal flow                 | Feed T/H | 327.6                     | 327.58| 0.01%            |
| Permeate T/H                | 147.4  | 147.41                    |       | 0.01%            |
| Rejected T/H                | 180.2  | 180.17                    |       | 0.6%             |
| Temperature                 | Feed °C | 25–34                     | 25    | –                |
| Permeate °C                 | 25–34  | 25                         | –     | –                |
| Rejected °C                 | 25–34  | 25                         | –     | –                |
| Pressure                    | Feed bar| 64                        | 65    | 1.5%             |
| Permeate bar                | 1.5    | 1.5                       | –     | –                |
| Rejected bar                | 62     | 63                        | 1.5%  | –                |
| Total Dissolved Solids (TDS)| Feed mg/l | 36,000                    | 36,036| 0.1%             |
| Permeate mg/l               | 500    | 477                       | 4.6%  | –                |
| Rejected mg/l               | 65,100 | 65,130                    | 0.05% | –                |
the optimisation of the permeate quality and flow rate. In this paper, the following assumptions were imposed:

- the solution-diffusion model is valid for the transport of water and solute through the RO membrane;
- the efficiency of the pump and turbine are fixed at 84% and 70% respectively;
- The pressure drop in feed stream is taken as the dead state 101.3 kPa;
- the salt feed water stream is considered to be a dilute solution and is treated as an ideal solution;
- the concentration polarization effect is negligible; [21,34,35].

The solution-diffusion model formed the basis for the design of the model. The difference between permeability coefficient, \( A_w \) and solute transport parameter, \( B_s \) determines the separation performance of the RO system. Any excess of the hydraulic pressure applied, \( P_f \) was assumed to be proportional to water permeation over the osmotic pressure, \( \pi_m \). Where \( J_w \) is the permeate flux, \( A_w \) is the apparent water permeability of the membrane which is used to characterise the membrane itself, \( \Delta P \) is the pressure applied across the membrane, and \( \Delta \pi \) is the osmotic pressure difference between the feed and the permeate [36,37]. Furthermore, the flux of the dissolved salts is proportional to the trans-membrane concentration difference. However, \( A_w \) and \( B_s \), are dependent upon temperature and they can be defined by a viscosity temperature function which is considered to have sufficient accuracy for engineering analysis carried out here. Changes in \( A_w \), result in proportional changes in \( J_w \) and \( J_s \), which means the temperature dependency of \( A_w \) and \( B_s \), must at least in principle be considered. For practical purposes, however, the temperature dependency of \( B_s \), is often neglected, while only the temperature dependency \( A_w \), which is much more important is included. The selectivity of a membrane considered, using the rejection coefficient, which explains the more frequent use of a rejection coefficient \( R \) and permeate flux \( J \) than the use of \( A_w \) and \( B_s \), membrane constants. However, such use of \( R \) and \( J \) only has significance when linked with precise information regarding the conditions, i.e. transmembrane pressure difference, salt concentration of the feed solution and membrane flux conditions. Thus, the formation of a concentration layer occurs at the membrane surface: The strong effect of concentration is hidden in the osmotic pressure difference \( \Delta \pi \). At least for highly dilute solutions, the relationship between osmotic pressure and concentration is linear: The transmembrane osmotic pressure is determined by Eq. (1). Membrane salt rejection is a measure of performance for overall membrane system for example manufacturers of membrane technologies usually define a specific salt rejection for each

| Temperature, °C | TDS, kg/kg | Pressure, bar | \( Q_f \), m³/h | \( C_p \), ppm |
|----------------|-----------|--------------|----------------|-------------|
| 25             | 0.015     | 30           | 2224           | <400        |

Fig. 4. Utilizing operating parameters as function of temperature.
commercial membrane available. Salt rejection through an RO membrane is nominally given by:

\[ R_s = \left( 1 - \frac{C_p}{C_r} \right) \times 100\% \]  

(12)

The feed water becomes gradually concentrated from the beginning to the end of the tube in a spiral wound element, and the salt rejection is described by:

\[ R_s = \left( 1 - \frac{C_p}{\left( \frac{C_f + C_r}{2} \right)} \right) \times 100\% \]  

(13)

Where \( C_r \) is the ion concentration in the concentrate. RO membranes achieve NaCl rejections of 98–99.8\% [45].

The stream numbers on the schematic representation are indicators of thermodynamic properties, as shown in Figs. 2 and 3. Stream no.1 is used as feed water which takes on the properties of brackish water for the purposes of validation. The RO model was built and validated against previously reported measurement data [46]. Table 3 establishes the difference between the RO model and reported measurement data as <4.8%.

4. Results and discussion

4.1. Effect of feed water temperature and recovery on RO efficiency

Based on the model conditions outlined in Table 4 and as a result of temperature increase mechanisms, the water passes more easily through the membrane due to a reduction in the water viscosity and change in the structure of membrane. As presented in Fig. 4(a), both the permeate flow rate and concentration increase with temperature from 20 °C to above 50 °C and this leads to increase of recovery, results in an increase in the mechanical power consumption (Fig. 4(b)). These outcomes are in good agreement with other reported observations [47,48,49] of similar systems. Furthermore, temperature plays a significant role in the performance of the RO filtration. For these calculations, the temperature range applied was between 20 °C and 50 °C. The specific energy increases with reducing temperature because of the corresponding reduction of the solvent transport constant, \( A_w \) (Eq. (2)), and the reduction of permeate flow rate (Eq. (7)). Fig. 4(c) shows the specific energy consumption for different scenarios, by using the pressure exchanger, the specific energy is reduced by about over 50% and by using turbine is about 30%. However, the concentration of TDS in the permeate decreased with the reduction in temperature, leading to a more considerable rejection of TDS. It should

Fig. 5. Effect of Recovery on permeate flow rate, area, power consumption and specific energy.
be noted that both the solvent transport constant, $A_w$ in Eq. (2) and the solute transport constant, $B_s$ in Eq. (4) increase with increases in temperature. By rearranging Eqs. (1) and (3), the solute concentration in permeate can be expressed as:

$$C_p = \frac{C_m}{\frac{A_w}{B_s} (\Delta p - \Delta \pi)} + 1$$  \hfill (14)

Hence, there is trade-off between temperature dependence of $A_w$ and that of $B_s$ which ultimately determines the overall temperature dependence of solute rejection. In addition as shown in Fig. 4(b), the percentage of recovery is effected by temperature which increased from $<45\%$ to $>50\%$ at $20\, ^\circ C$ and $60\, ^\circ C$ respectively.

Furthermore, the specific energy and membrane performance are influenced by the percentage of recovery. As shown in Fig. 5(a), with an increase in recovery percentage, the driving force required for an increase in flux increases due to greater salt concentration in feed stream, so the corresponding permeate flow rate increases thus requiring a large area of membrane. Production of high quantity of permeate water is positively affected on the specific energy which is reduced from 2.8 kWh/m³ to a more economical 0.8 kWh/m³ at 30% and 60% respectively, when using pressure exchanger, as shown in Fig. 5(b). Due to relatively high energy demands, most RO systems are fitted with a device to recover energy from the pressurized RO concentrate leaving the system. The primary objective here is to recover as much of the energy held in the pressurized RO concentrate stream as possible, and it is very clear from Fig. 5(b), the PX device is the best option to recover the rejected energy. The concentrate is sent through an energy recovery device, and this energy is used to supplement the power to the pumps. Thus the efficiency of the energy recovery device has important role in the overall RO system energy consumption. Fig. 5(c) demonstrates a high level of pump power consumed without any energy recovery compare with ERT consumption. In circumstances where the brine flow rate,
4.2. Effect of applied pressure on permeate concentration, area of membrane, ERT power saving and flux

The performance of RO membrane was analysed at different feed water pressures. Fig. 6(a) shows the changes in permeate flux at different applied pressure with constant temperature (25 °C). In general, the results illustrated that an increase in applied pressure yielded in an increase in the feed water flux. These data are in agreement with the observations of Ahmed et al. [50], Mohammadi et al. [51] and Hyun et al. [48]. Based on Darcy’s law, the permeate flux increases with increasing pressure gradient (see Eq. (1)) whereas, the membrane area decreases, as shown in Fig. 6(b). The solute concentration decreased gradually as the feed pressure increased, indicating that less permeate TDS is produced. Fig. 6(c) illustrates the concentration of permeate is reduced with increasing pressure. As shown in Fig. 6(d), another interesting observation was that the energy recovery turbine saved energy by approximately 30% by increasing the applied pressure and PX device saved >50%.

4.3. Salt rejection as function of temperature and pressure

Salt rejection coefficient is an appropriate measure for the selectivity of a membrane and is more often used than the membrane constants \( A_w \) and \( B_s \) (shown in Eqs. (2) and (4)) [52]. Whilst, the rejection coefficient is not a membrane constant, it is a function of the operating conditions. Fig. 7 shows that high pressures increase the salt rejection from 96.8% to >98.8% when pressure is increased from 25 bar to 63 bar (at 25 °C). Also, salt rejection increases with reducing temperature as shown in Fig. 8, which means that with higher temperatures, much more TDS is in permeate. This is due to a reduction of solvent viscosity and the pore size

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**Figure 8.** Salt rejection as a function of temperature.

**Figure 9.** Diagram of total water cost components.
effect. Both temperature and pressure regulate the amount of TDS in permeate. In order to produce higher quantities of water with specific quality (i.e. for irrigation applications), particularly with brackish feed water, control of the salt rejection ratio has been analysed. When salt rejection coefficient increases, less salt concentration appears in permeate and simultaneously higher energy is required. Thus, to mitigate the excess salt rejection, the pore size diameter can be increased. To match the feed brackish water in the proposed case study, with its specific requirements of freshwater properties and energy consumption, a change in temperature affected the permeate concentration and flow rate, it proved insufficient to meet the requirements of freshwater production volume. Therefore, physical changes of pore size diameter for RO membrane are required to tackle the agricultural demand for water.

5. Economic analysis

Among the major determining factors for estimating the cost of water is the cost of available energy. The principal cost factors considered include capital investment, maintenance cost and the cost of supplying saline water to the desalination system. The labour cost can vary greatly, and is subject to the local economy. Cost balance equations for the required components in the system are presented in Table 6. The expenditure connected with setting up and operation of a desalination plant, include the initial concept, design, obtaining of permits, finance, construction and the commissioning and acceptance testing for normal operation [53] is defined here as the capital cost [24].

The total water cost (TWC) is estimated by adding the capital cost to the operating cost for the length of the contract (Fig. 9) and dividing the total of the amortized (annualized) capital costs and the annual Operating and Maintenance (O&M) costs by the average annual potable water production volume. These parameters are set out Tables 5 and 8. As is typical, the TWC excludes distribution costs, especially where alternative delivery contracts are concerned [53]. The O&M costs are specific to the site but consist of both fixed (insurance and amortization) and variable costs (cost of labour, energy, consumables, maintenance, and spare parts etc.). Features of capital cost are direct (process equipment, auxiliary equipment and the associated piping and instrumentation, site civil works, intake and brine discharge infrastructures, buildings, roads and laboratories) and indirect costs. The contract agreement establishes the land cost, which may vary from zero to an agreed lump sum according to the site characteristics [54]. Typically, 50–85% of the total capital cost are construction costs. Indirect capital costs, usually calculated as a percentage of the direct capital costs, averaging 40% [54], 15–50% [53] or 30–45% [55], but very project specific, are composed of interest accruing during construction, working capital, freight and insurance, contingencies, import duties, project management, and Architectural and Engineering (A&E) fees.

The TWC and the investment costs for the MOD brackish water plant are GBP £0.11/m³ and GBP £14.4 million respectively. This is to produce drinking water with total capacity of 24,000 m³/day (obtained product <400 ppm) and feed salinity of 15,000 ppm with PX device. Also the TWC and the investment costs production for irrigation) 24,000 m³/day, feed salinity of 15,000 ppm and obtained product <1600 ppm) are GBP £0.9/m³ and GBP £11.3 million respectively. As shown in Fig. 10, the cost of production is affected by recovery devices, pressure exchanger reduced the cost by 11% of the cubic meter of the production and the turbine is only 3%, and these costs are influenced by the different salinities. These costs are in good agreement with the findings outlined in Tables 9 and 10. Comparing the model results of this work with findings of previous publications (Tables 9 and 10) confirms the rapid decline in TWC with increasing plant capacity. It can be concluded that the MOD investment cost and TWC arrived at in this work is in agreement with other findings. The annual investment cost

### Table 5
Economic data of the BWRO plant.

| Economic parameters (assumed) | |
|-----------------------------|------------------|
| Interest rate               | 8%               |
| Nominal escalation rate, \( r_s \) | 5%               |
| Economic life time, \( n \) | 16 year          |
| Effective discount rate, \( i_{eff} \) | 8%               |
| Membrane life               | 5 year           |
| Annual operating hours      | 7884 h           |

### Table 6
Equations of calculation of the capital and operating cost.

| Description                              | Equation                                      | No. | Ref. |
|------------------------------------------|-----------------------------------------------|-----|------|
| Cost of the intake and pretreatment      | \( C_{BHPP} = 996 (Q_f/10^6) \)               | 15  | [26,56]|
| Annual cost of the energy of the intake pump | \( C_{BHPP} = C_{HPP} \cdot \eta \cdot Q_f \) | 16  | [57,58]|
| Cost of chemical treatment in the pretreatment | \( C_{C,CHPP} = Q_f/10 \cdot \eta \cdot C_{CH} \) | 17  | [58,59]|
| Power of high pressure pump              | \( \log_{10}(PC_{HCPP}) = 3.3892 + 0.0536 \log_{10}(W_{HCPP}) + 0.1538(\log_{10}(W_{HCPP}))^2 \) | 18  | [60]|
| Annual cost of the power provided to the HPP | \( C_{PP,HPP} = P_{PP,HPP}/Q_f \cdot \eta \cdot C_{PP} \) | 19  | [58]|
| Capital cost of the RO membrane          | \( C_{RO} = N \cdot PC_{m} \)                  | 20  | [26,57]|
| No. of elements                          | \( N = K_{PO} \cdot Q_f/\bar{Q}_{P} \)         | 21  | [26,57]|
| Cost per membrane                        | \( PC_{m} = 10 \cdot A \)                      | 22  | [26,58]|
| Area                                     | \( A = Q_f + C_{m}/(Q_f/10^6) \)              | 23  | [26,61]|
| Average salinity through the membrane element | \( A \)                                      | 24  | [26,61]|
| Amount of bypass water                   | \( Q_{bypass} = Q_f \cdot C_{m}/(Q_f/10^6) \) | 25  | [26,61]|
| Cost of membrane elements replacement    | \( C_{RO} = N \cdot PC_{m} \)                  | 26  | [26,62]|
| Power of turbine                         | \( \log_{10}(PC_{f}) = 2.2476 + 1.4965 \log_{10}(W_f) - 0.1618(\log_{10}(W_f))^2 \) | 27  | [60]|
| Total annual O&M cost                    | \( C_{O,M} = 0.002X_1 \cdot Q_f \cdot C_{m} \) | 28  | [63]|
| Constant escalation levitzation factor   | \( CELF = CRF - \frac{1}{10^{(1+\eta_{max})}} \) | 29  | [64]|
| Constant factor                          | \( K = \frac{1}{\eta_{max}} \)                 | 30  | [64]|
| Capital recovery factor                  | \( CRF = \eta(1+\log_{10}(\frac{Q_{bypass}}{Q_f}))^2 \) | 31  | [64]|

...
Calculation is based on the assumption that Table 10 Capacity of desalination unit and cost of water produced.

| No.  | Type of feed water salinity, ppm | Plant capacity, m³/d | Cost, £/m³ | Source of information |
|------|---------------------------------|----------------------|------------|----------------------|
| 1    | Brackish, 5700                  | 50                   | 4.7        | [65]                 |
| 2    | Brackish                        | <20                  | 3.3—7.5    | [9]                  |
| 3    | Brackish                        | 20—1200              | 0.45—0.77  | [9]                  |
| 4    | Brackish, 8116                  | 6000                 | 0.22       | [66]                 |
| 5    | Brackish, 4221                  | 10,000               | 0.15       | [66]                 |
| 6    | Brackish, 5844–11,688           | 30,000               | 0.18       | [66]                 |
| 7    | Brackish, 10,000                | 38,000               | 0.35       | [67]                 |
| 8    | Sea water, 26,000               | 95,000               | 0.34       | [67]                 |
| 9    | Brackish, 3000                  | 38,000               | 0.21       | [67]                 |
| 10   | Brackish, 2300                  | 92,000               | 0.19       | [68]                 |
| 11   | Brackish, 5000                  | 46,000               | 0.17       | [69]                 |
| 12   | Brackish                         | 5000–60,000          | 0.15—0.31  | [70]                 |
| 13   | Brackish                         | 40,000–46,000        | 0.15—0.31  | [9]                  |
| 14   | Brackish                         | 19,000               | 0.15       | [9]                  |
| 15   | Brackish, large scale            |                      | 0.13—0.26  | [7]                  |
| 16   | Brackish                         | 38,000               | 0.12       | [9]                  |

Calculation is based on the assumption that

1 GBP = 1.54 $  
1 GBP = 1.37 €

Table 10 Water and capital cost for different projects [7,67].

| Feed water TDS, ppm | Capacity, m³/d | Capital cost, £/m³ | Cost, £/m³ |
|---------------------|----------------|-------------------|------------|
| Sea water 20,000    | 13             | 0.42              | 0.42       |
| Brackish water 1380 | 28,400         | 16                | 0.2        |
| Sea water 34,000    | 40–60          | 0.25–0.41         | 0.25–0.41  |
| Sea water 45,000    | 45.5           | 0.36              | 0.36       |
| Brackish water 2550 | 55,670         | 56.5              | 0.27       |

Calculation is based on the assumption that

1 GBP = 1.54 $  
1 GBP = 1.37 €

Fig. 10. Cost of production for different scenarios.

of any component is achieved by adding the annual total capital investment rate and the annual O&M cost rate, estimated by dividing the capital investment cost for each component by the plant annual operating hours, as shown in Tables 5, 7 and 8.

6. Conclusion

Numerical analysis has been conducted to study the performance of reverse osmosis membrane for a brackish water desalination process. The RO model has been designed for analysis of a case study based in Iraq and the model developed using Matlab/Simulink and Thermolib software. The process can produce water for several purposes namely for domestic utilisation, agricultural irrigation and survive Marshlands. A detailed analysis has been carried out to reduce losses and to specify efficiencies of individual components. As a result of the analysis using the model, it can be seen that, salt rejection can be reduced from 97% to 88% to obtain high quantities of fresh water with an agriculturally acceptable (lower quality). Increasing water feed temperature or reducing the feed water pressure, and physically by increasing the average pore size diameter, led to significantly reduced power energy consumption. Moreover, the specific energy was reduced from 2.8 kWh/m³ to a more economical 0.8 kWh/m³ by producing high quantities of drinking water. In addition, it was demonstrated that utilizing energy recovery device turbine and PX in brackish feed water led to a further power saving of around 30% and over 50% respectively. Another interesting finding was that total water cost of a MOD brackish water plant with a total capacity of 24,000 m³/d and feed salinity of 15,000 ppm (obtained product <400 ppm) is GBP £0.11/m³ and the investment costs is GBP £14.4 million, and the cost of production for irrigation) 24,000 m³/d, feed salinity of 15,000 ppm and obtained product <1600 ppm) is GBP £0.9/m³ and with an investment cost of GBP £11.3 million.

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