This report provides a User's Manual and Help Menu for the Solar Desalination Analysis Tool (SDAT) developed at Columbia University under the auspices of the Solar Energy Technologies Office (SETO) of the United States Department of Energy (US-DOE)

To see the Help system in SDAT, click Help in the main window, or press the F1 key (command-? in MacOS) from any page in SDAT.

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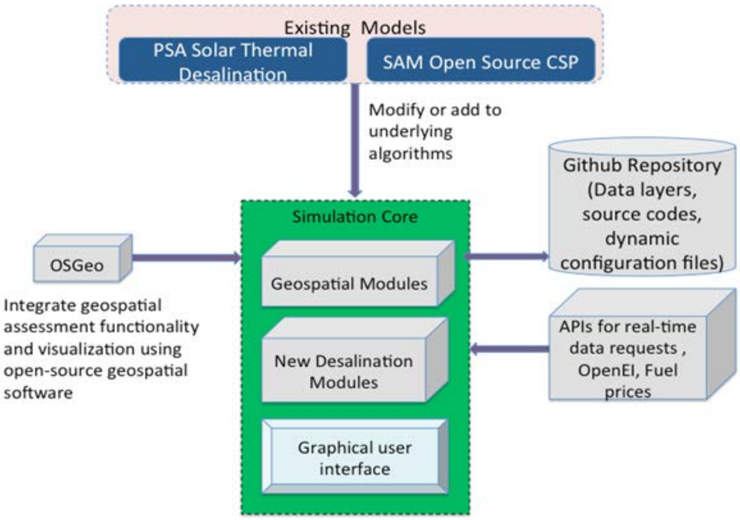
1. Introduction

1.1 Objective

The Objective is

1.2 Tool Structure

The structure of this tool of analysis is shown in fig. 1.



1.3 Databases

Table 1. Simplified summary of primary GIS data source types. Sub-parameters highlighted in green pertain to data or estimates that have been the focus of this quarter.

Table 1. Simplified summary of primary GIS data source types. Sub-parameters highlighted in green pertain to data or estimates that have been the focus of this quarter.

Parameter	Sub-parameter	Spatial Level of Detail
Alternative water sources	Municipal and industrial wastewaters	County, local (points)
	Brackish water	Local
	Agricultural drainage water	County
	Seawater	Local
Water demand	Areas of projected population growth	Local
	Sector water withdrawals (e.g., domestic, agricultural, industrial)	County and state
	Water pricing	Local, county
	Solar Direct Normal Irradiance (DNI)	Local

Energy sources	Waste heat (power plants)	Local, county
	Natural heat (geothermal)	Local
	Commercial & industrial prices for electricity and fuel (gas and diesel)	State
Site selection factors	Land cover, physical restrictions (slope, protected areas)	Local (gridded and detailed mapping units)
	Regulations and Permitting requirements	Administrative (local, state, federal)

1.1 Regulatory/Permitting Data (ST1.4)

A cursory review of federal regulations revealed a handful of agencies with jurisdiction over desalination developments including the U.S. Environmental Protection Agency, the U.S. Army Corps of Engineers (USACE), the U.S. Bureau of Reclamation and the U.S. Coastal Guard, among others. These federal agencies oversee the permitting associated with discharges as outlined by the National Pollution Discharge Elimination System (NPDES) permit program, alongside permitting for plant construction in the vicinity of U.S navigable waters and environmental impact on aquatic flora and fauna. The USEPA empowers most states to handle the permitting, administrative and enforcement of the NPDES program while retaining oversight power.

A review of the regulatory and permitting practices abroad by the Water Environment & Reuse Foundation reveals that other countries where desalination has become a mainstay of the water supply, such as Australia and Spain, have a more streamlined process, with a key difference between the U.S. and these countries being the amount of time needed to issue the permits, with the United States being the only country where the permitting process, 3-10 years on average, can take longer than the construction time, 2 to 3 years. This discrepancy in time is not limited to international waters, however. States within the U.S have such an independent and fragmented permitting system that the permitting process for the Carlsbad Desalination Project (California) took 10 years, from project inception in 2000 to permit approval in 2010. Conversely, that of East Cherry Creek Valley Reverse Osmosis Plant (Colorado) took 2 years, from the pilot testing of brine disposal methods in 2008 to the discharge permit approval in 2010.
The data show that state by state differentiation is important. For example, California and Texas have fashioned slightly different requirements for a wide range of desalination facilities within the states. In 2014, the State Water Quality Control Board of California instituted the California Ocean Plan which contains specific requirements pertaining to discharge from desalination plants. The plan covers only seawater facilities and prescribes key constraints on their development such as mandatory use of subsurface intakes for source water, a requirement for impingement and entrainment mitigation where applicable, discouragement of co-location with power plants so as to reduce open channel withdrawal and discharge, as well as a requirement that the discharge salinity at the edge of the mixing zone be lower or equal to 2 parts per thousand (ppt) above the ambient water salinity, necessitating large dilution levels. On the other hand, Texas has a step-wise, case-by-case approach, with a permitting decision tree model for determining which existing permits apply to the user's plant specifics.

Thus, for the regulatory and permitting portion of the solar thermal desalination tool under development at the Columbia Center for Life Cycle Analysis (CLCA), states in the southeast and southwest of the United States were selected for the project, namely Texas, Arizona, Nevada, Florida, California and Colorado. Available in the open literature information on the permitting requirements applicable to existing desalination plants was compiled, and the associated state and county permitting requirements where synthesized into tabular forms that will be shown to the user of the software, together with links to associated agencies and permitting forms per state and county as available and applicable. It is noted that the CLCA database, while useful to stakeholders, is not all-encompassing and should only be used as a preliminary guideline for permitting requirements. A snapshot of our Regulatory Database is shown below (Fig. 1).

Permit Type	Permit Purpose	Additional Details	Permit Name	Issuing Agency	Link to Site
Construction	Coastal Development Permit		Coastal Development Permit (CDP)	California Coastal Commission	Application
Discharge	NPDES Industrial Stormwater General Permit		Industrial General Permit (IGP)	California State Water Resources Board	Application
Discharge	Brackish groundwater and Reverse Osmosis concentrate		Extracted Brackish Groundwater and Reverse Osmosis Concentrate	San Francisco Regional Water Quality Control Board	Application
Discharge	Permit for discharges of stormwater associated with construction activity		Construction General Permit	California State Water Resources Control Board	Application
Discharge	Construction NPDES permit for Lake Tahoe Hydrological Unit		Constructional NPDES General Permit for the Lake Tahoe Basin	Lahontan Regional Water Board	Application
Discharge	NPDES permit for drinking water system discharges	Applies to both surface and ground water systems	General NPDES Permit for Drinking Water System Discharges	California State Water Resources Control Board	Application
Dicharge	Industrial wastewater NPDES permit		Water Discharge Requirements & Water Recycling Requirements	North Coast Water Resources Quality Control Board	Application
Discharge	NPDES Report of process wastewater discharge		State Application Form 200	California State Water Resources Control Board	Application

Figure 1: Snapshot of Regulatory Database

1.4 Techno-economic Models

Table 2. List of SAM models integrated in the software

CSP models	Financial models
Parabolic Trough Physical (PT)	Commercial (distributed)
Linear Fresnel Molten Salt (MSLF)	PPA single owner (utility)
Linear Fresnel Direct Steam (DSLF)	PPA partnership flip with debt (utility)
Power Tower Molten Salt (MSPT)	PPA partnership flip without debt (utility)
Power Tower Direct Steam (DSPT)	PPA sale leaseback (utility)
Integrated Solar Combined Cycle	LCOE calculator
Industrial Process Heat (IPH) Parabolic Trough	
Industrial Process Heat Linear Direct Steam	

1.5 Quick Access Web-based Simplified Tool

1.6 Downloading and Installing SDAT

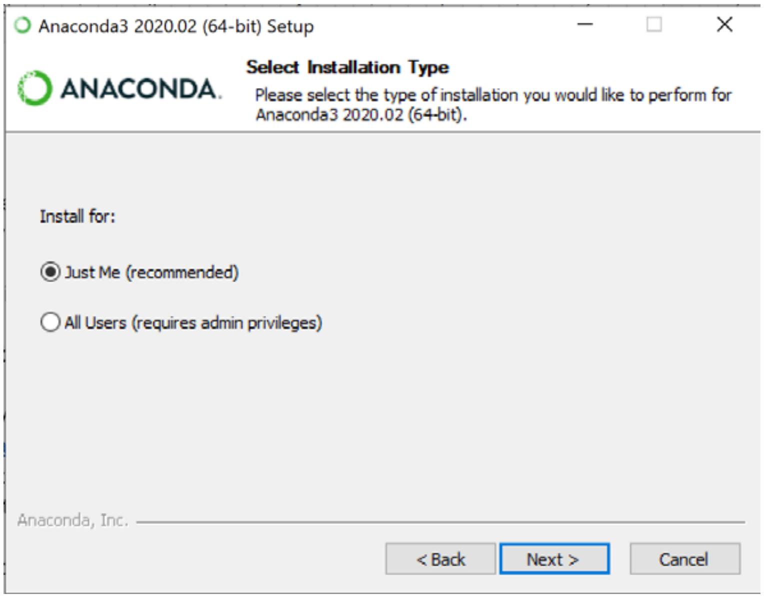
Windows Installation

1.1 Download and install Anaconda; if you already have a recent version of Anaconda, you can **skip** this step)

https://www.anaconda.com/products/individual

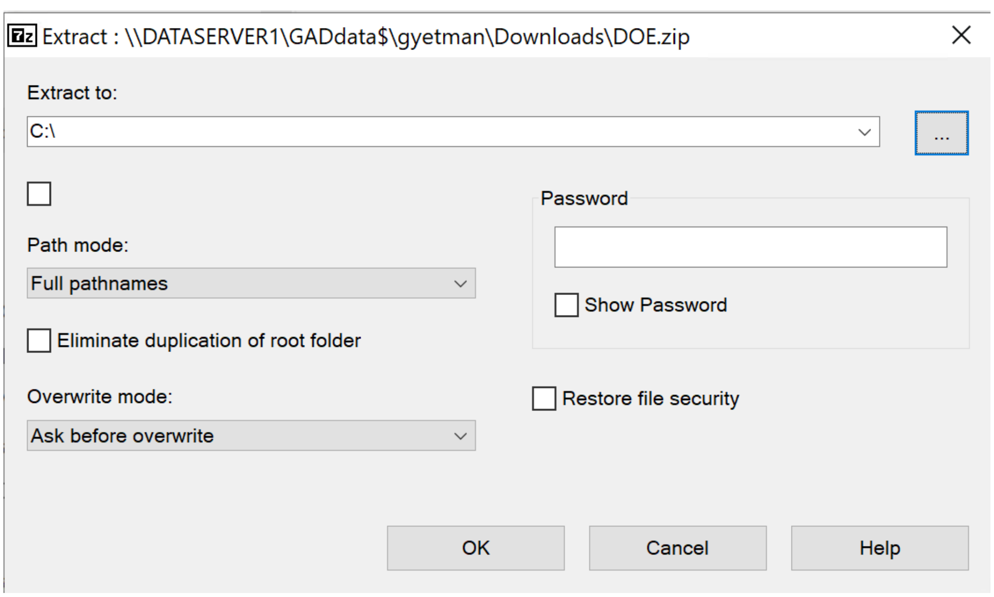
Choose the 64 bit installer.

While installing, selection the option to install it for **Just Me**



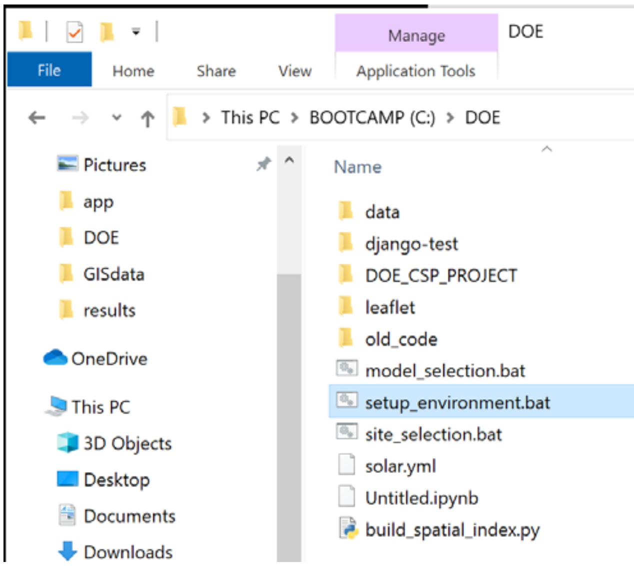
1.2 Download the bundled application and extract it to the **root** of your **C:** drive. Note that if you installed the alpha version of the software, you should first **delete** the C:\DOE folder from your computer before downloading and extracting the new version from:
https://drive.google.com/file/d/1pttB0l05JH6ebqT2YAKlQYFXtjaBDAKY/view?usp=sharing

This should create a folder named C:\DOE that contains the application and subfolders.

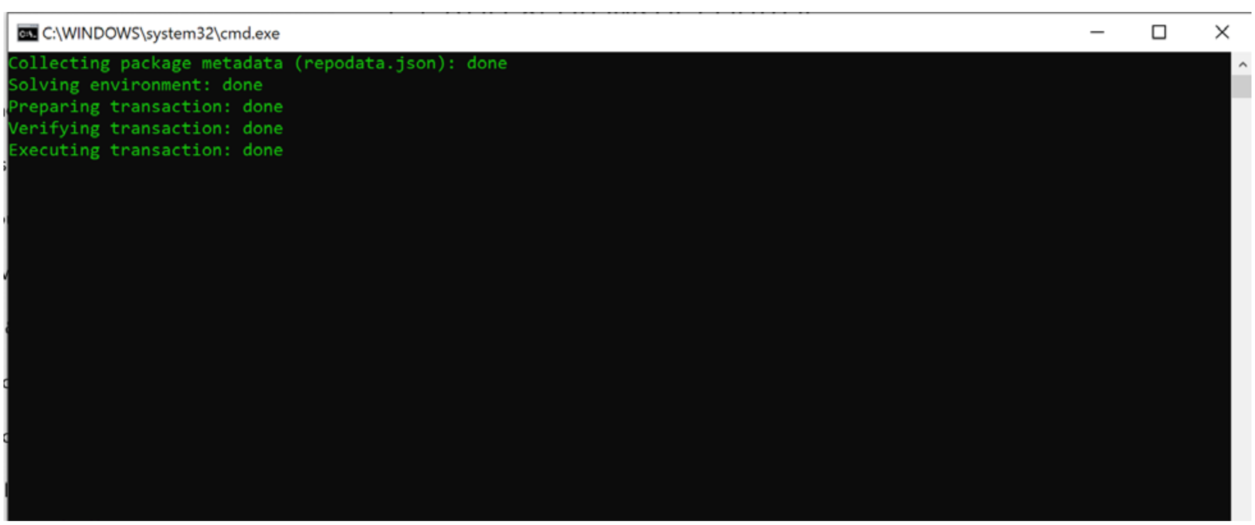


Make certain to extract to the root folder (C:\), so that you don't end up with a nested folder structure (C:\DOE\DOE), which will not work.

1.3 Open the C:\DOE folder, with Windows Explorer you should see a file named **setup_environment.bat**. Double-click the file to execute it. Some versions of Windows will give a security warning and not allow the .bat file to run. If this occurs, in the warning dialog, click "**More Info"** and then click "**Run Anyway**". This will configure the Anaconda Python environment you installed in Step 1. This may take some time (up to 10 minutes, depending on your Internet speed) to complete as it downloads files from the Internet.



Note: if a terminal window does not open and proceed to download the necessary data, you may need to override your Windows Security settings and restart your computer before trying again. To override your security settings, go to "App & Browser Control" and turn off "Check apps and files." From here, restart your computer and try opening **setup_environment.bat** again. A terminal window like the one shown below should open. This process takes a few minutes and depends on your internet connection.



Once this is done, the installation is complete!

Mac Installation

1. If you already have Python 3 and Anaconda installed, skip to the next step. You can test this by opening a terminal window and entering the command:
   ```
   conda list anaconda
   ```
   If the result includes a python build of 3.6 or later, you can proceed to step 2. Otherwise, download and install anaconda from https://www.anaconda.com/products/individual
   Accept the default options for the installation.
2. Download the .zip archive from this link: https://drive.google.com/file/d/1MtlWCC4bU0W6WoZPMfu3CXg3iW8at0Pu/view?usp=sharing
3. Extract the contents to your home directory (/users/<your_username>/). It should create a folder named DOE.
4. Open a terminal (shell) and change to the DOE folder (~/DOE). Run the script solar-setup.sh by entering the code:
   ```
   ./solar-setup.sh
   ```

1.7 User Support

Email: xxx

2 Running SDAT

2.1 Visualize Maps

3 Steps to Run the software

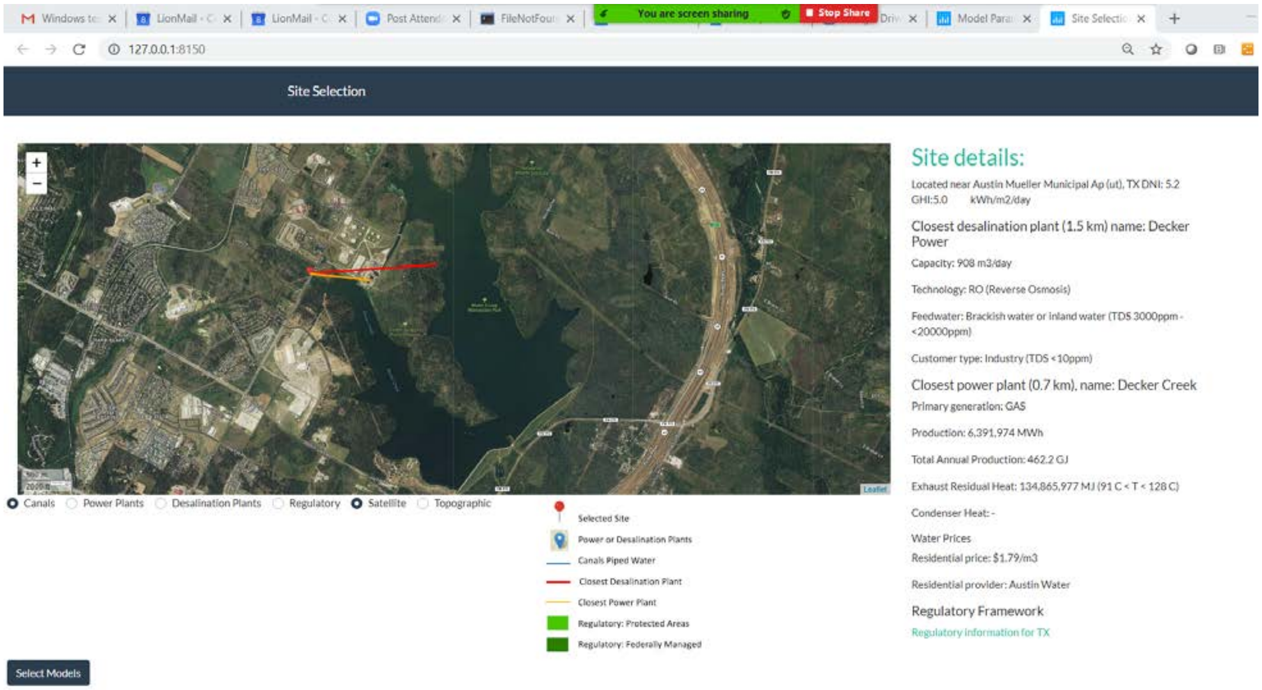
3.1 Location Specific

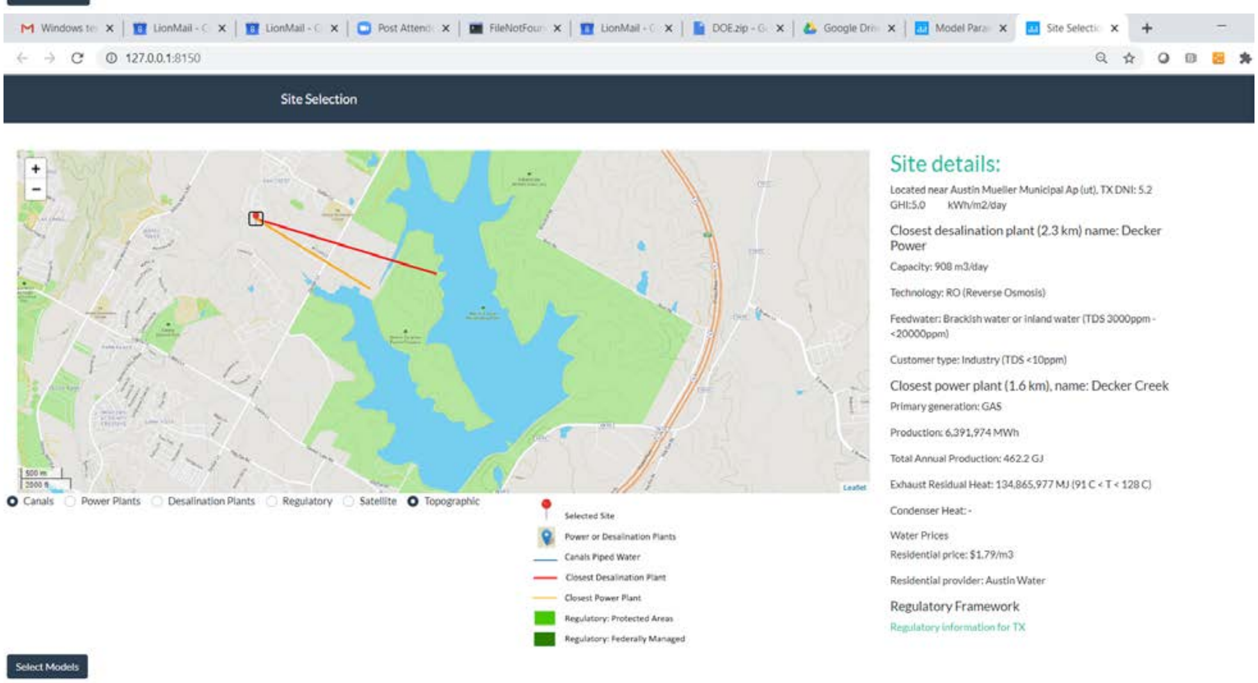
Windows systems

Run the site_selection.bat file in the C:\DOE folder by double-clicking it. It should automatically open your web browser to the site selection page (http://127.0.0.1:8150/)

You can see maps with locations of Canals, Desalination plants, and Power plants and you can move on the map and zoom in and out to smaller or larger areas. You can select a location on either a satellite or topographic map and see, by clicking there, the nearest desalination and power plants, associated distances and capacities and also water prices for closest demand locations.

A couple of example images are shown below.





Then you can proceed to techno-economic desalination modeling by clicking at "Select Models" (left bottom of the screen).

Mac systems

Open a terminal and change directory to the ~/DOE directory. Run the command below in the terminal:

```
source site-selection.sh
```

After about 10 seconds, your browser window should open to the map. You can go directly from site selection to model runs via the application. When you are finished, simply close the terminal windows that open*.

***Troubleshooting:** Depending on your OS software version, closing terminal windows may not suffice to restart the software without issue. If that is the case, when opening a new terminal to run the software again, you may see an error message in the terminal window that reads

`[Errno 48] Address already in use`

This is because any python processes that should have ended after closing the terminal are still running in the background. Here are two options to resolve this issue:

(Option 1) Open Activity Monitor >> locate any Python processes that are running and force each process to quit >> rerun the command to open either the Model Selection or Site Selection interface

(Option 2) Open a Terminal window>> enter the command

```
sudo pkill python
```

>>enter your computer's password when prompted >> then type the following command to verify that all Python processes have ended. This command should not produce any output.

```
sudo pgrep python
```

>> rerun the command to open either the Model Selection or Site Selection interface

2.3 To run Model Selection for a default Tucson, Arizona location (or the last location from a previous Site Selection run)

Open a terminal (shell) and change into the directory ~/DOE

1. Run the command below in the terminal
   `source model-selection.sh`

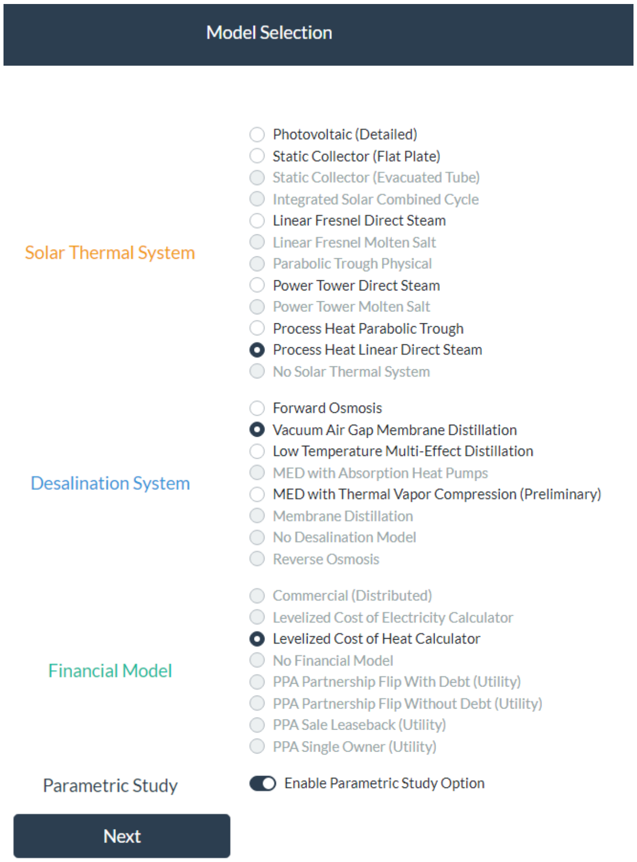
2. Your browser should automatically open the menu (http://127.0.0.1:8077/model-selection)

From this stage on, you can follow the instructions shown in the Windows section above.

3.2 Design Specific

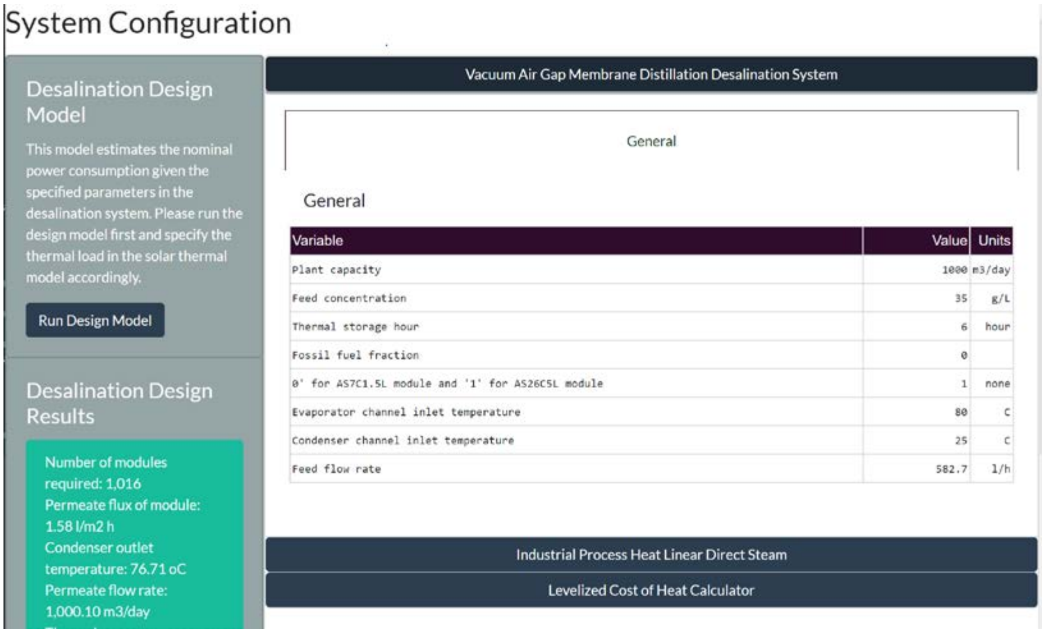
To run Model Selection for a default location in Tucson, Arizona (or the last location from a Site_Selection run)

- Run the **model_selection.bat** file in the C:\DOE folder by double clicking it. It should automatically open your web browser to the model selection page (http://127.0.0.1:8077/model-selection)
- Choose the model combination that you want to run using the radio buttons. Models not available in the GUI application cannot be selected.



You have the option to Enable a parametric study if you wish. However, if you enable it then you would have to introduce additional variables as described in a previous section .

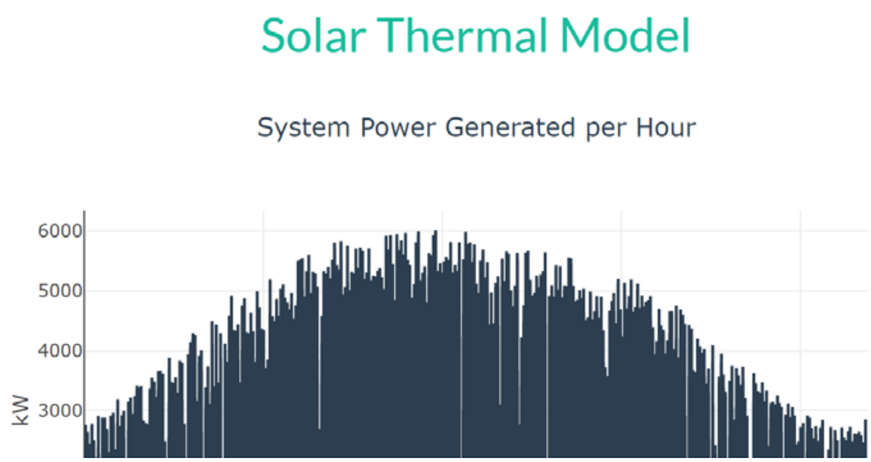
Click the next button to review parameters. The menu that allows you to review and edit parameters as needed.



- Review the parameters, you can expand model sections by clicking the model names. The design model recommended parameters are shown on the left, these can be updated by choosing "Run Design Model" on the left. When you are satisfied with the inputs, scroll down to choose "**Run Simulation Model**".

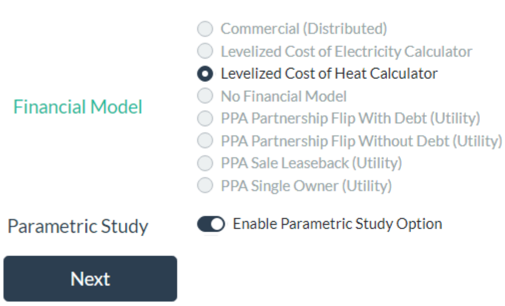
- When the model run is complete, you can click "View Results to see the results charts and report.

- Use the Charts and Report buttons on the top-right to switch views.



- To run Model Selection for a default location in Tucson, Arizona while enabling parametric study

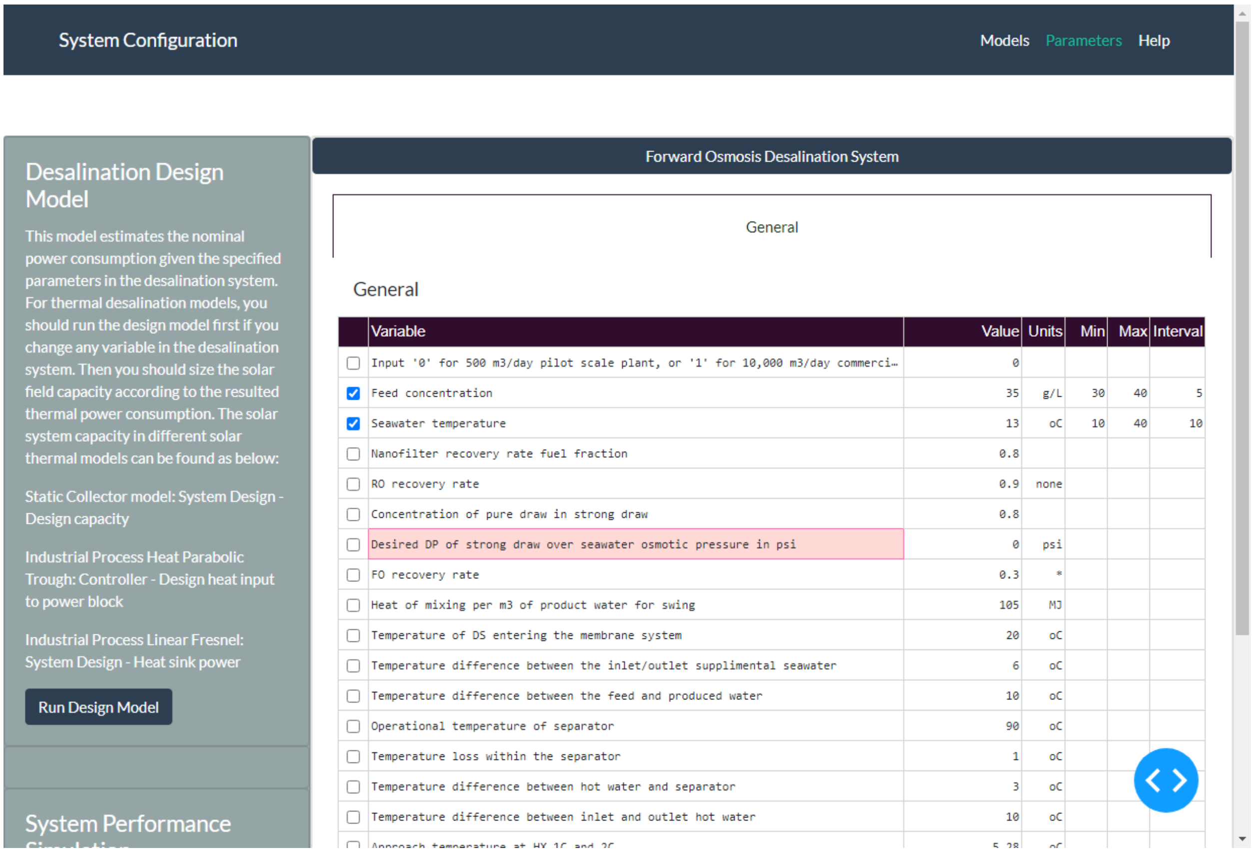
Choose the model combination that you want to run using the radio buttons. Then toggle the button in the end to enable the parametric study.



Select the variables you're interested in for the parametric study. Please remember to:

a. Check the box on the left of the target variables
b. Input the min, max and interval values for the parametric study (No need to change the column "Value")
c. Modify the values of other variables as you see fit

Please note that currently you cannot select more than 2 variables.



Please allow a few minutes for the models to run. Then you can check the bar chart results: (The "Report" button doesn't work for parametric study for now)

3.3 Select Models

3.4 Inputs Browser/Review and Select Inputs

3.5 Work Flow for Running Simulations

The data flow diagram shown in Figure XX walks through a high-level workflow for site and model selection using the desktop application for a site-selection driven workflow. The steps shown in the data flow diagram are as follows.

1. User starts desktop application.
2. The application loads the GIS data from local files and a base map from the web, and renders selected layers over the base map. A base map provides context (highways, topography, place names, etc.). At this point, the user can zoom and pan on the map and select a site location.
3. The site details are loaded and shown to the user, including the annual DNI data, water sources (wells), waste heat sources, and existing desalination facilities within a cutoff distance.
    a. The user can interactively choose a new site and review the site details repeatedly until satisfied with the location. This is indicated by the loop "user updates location".
4. Once the location has been finalized, a model selection can be made by the user from the following menus:
    a. Solar/thermal model
    b. Desalination model
    c. Financial model
5. The application loads the relevant parameters from the GIS data and configuration files that store parameters and default values for the selected model.
6. A subset of the parameters most commonly selected by the user are displayed in the application interface; these may be edited and revised.
7. Optionally, a user may examine and modify extended parameters (those not usually modified but perhaps of interest to some).

8. Once the user submits the parameters, the techno-economic model or models are run (status updates will be provided).
9. Once completed, the model outputs are displayed to the user (graphs/charts and tables). The user may also access the files generated.
10. At this point the workflow ends. The process could be started again, or the application closed.

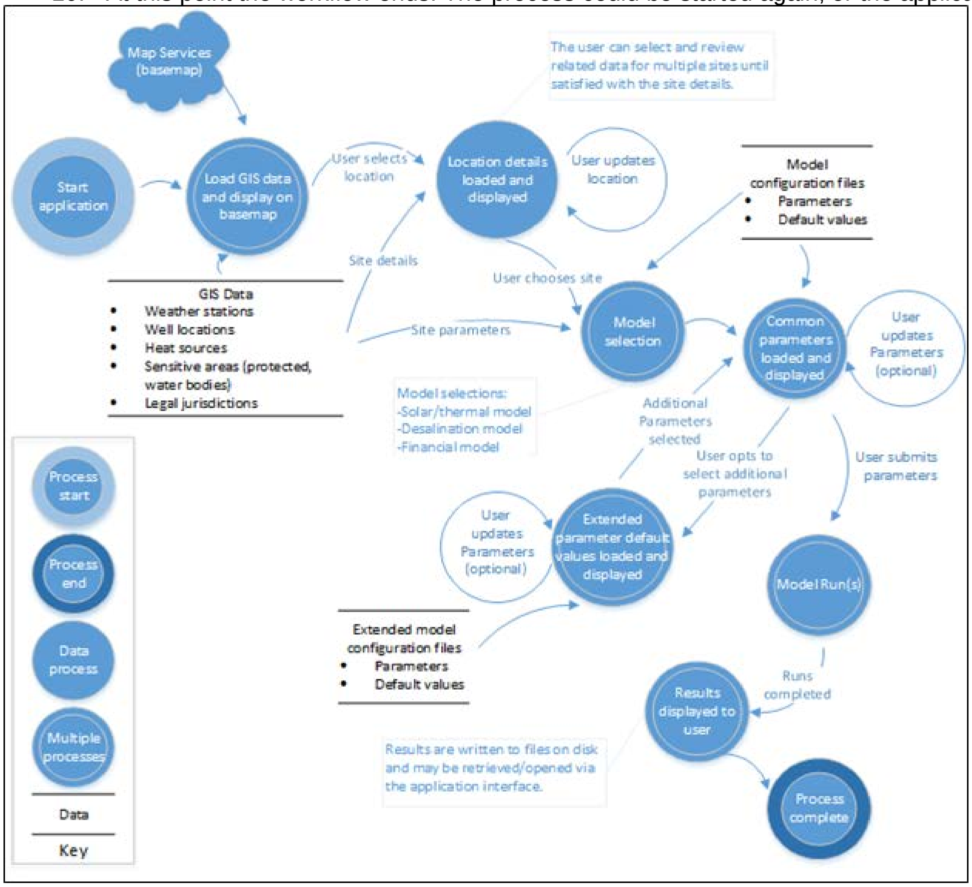


Figure xx. Data flow diagram for a typical workflow of selecting a site, model, reviewing parameters, and model output.

3.6 Parametric Simulations

3.7 Review Results

4 File Management

4.1 File Menu

4.2 Case Reports

4.3 Case Graphs

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5 Databases

5.1 Solar Resource

5.2 Alternative Water

5.3 Power Plant

5.4 Desalination Plant

5.5 Sensitive Areas

5.6 Water Networks

5.7 Water Utility Prices

5.8 Regulatory/Permitting

6 Solar Thermal Energy Generation Models (summaries/details in SAM manual)

6.1 Static Solar Collector

The SAM solar collector model is limited to residential water heating; thus, we elected to code and integrate a model developed by our PSA collaborators which has been verified with measurements at a pilot plant in Almeria. The new static collector class was created mostly by converting PSA's MATLAB functions to Python and modifying the code to be more dynamic. The class allows the user to select either flat-plate or evacuated tube collectors. The user will be able to design the collector field and simulate hourly performance.

Figure xx . Part of the Solar Flat Plate Collector at PSA

ADD schematic and DESCRIPTION

6.2 CSP Parabolic Trough

ADD SUMMARY FROM SAM

Figure xx. Part of the Parabolic Trough CSP facility at PSA

6.3 CSP Linear Fresnel

6.4 Industrial Process Heat

6.5 Financial models

ADD FROM SAM

Levelized Cost of Heat (LCOH) calculation
In order to evaluate the economic performance of the thermal desalination system, we should determine the levelized cost of heat (LCOH) used in desalination. However, the financial models in the SAM CSP models, do not report LCOH. Thus, we made the following additions to the SAM code.

2.1.3.1 LCOH calculator (replicating the LCOE calculator in SAM)

This is a simpler calculation, where cash flow is neglected:
Starting from:

$$LCOE = \frac{FCR \times TCC + FOC}{AEP} + VOC(\$/kWh)$$

where FCR is the fixed charge rate, TCC is the capital cost ($), FOC is the fixed annual operating cost ($), VOC is the variable operating cost ($/kWh), FCR is the fixed charge rate, AEP is the annual electricity production (kWh)

and LCOH is derived by replacing the electricity production with the thermal energy production:

$$LCOH = \frac{FCR \times TCC + FOC}{AHP} + \frac{AHP}{AEP} \times VOC(\$/kWh)$$

where AHP is the annual heat (thermal energy) production (kWh-t)

2.1.3.2 LCOH considering cash flow (used in most LCOE financial models)

To evaluate the levelized cost at present, the cash flow is considered:

$$LCOH = \frac{\sum_{i=0}^{n} \frac{C_n}{(1+d_n)^n}}{\sum_{i=1}^{n} \frac{Q_n}{(1+d_r)^n}}$$

where C_n is the total annual (capital+operation) cost in year n, Q_n is the thermal energy generation in year n (kWh-t), d_n is the nominal discount rate and d_r is the real discount rate.

Levelized Cost of Water (LCOW) calculation

Capital cost and O&M cost are included:

$$LCOW = CAPEX + OPEX\ (\$/m^3)$$

where $CAPEX$ is the unit capital cost ($\$/m^3$), to evaluate the levelized cost at present, the cash flow is considered:

$$CAPEX(\$/m^3) = \frac{C_0}{Q_p} \times \frac{r(1+r)^t}{(1+r)^t - 1}$$

where C_0 is the initial capital cost for the desalination plant (\$), Q_p is the annual water production (m^3), r is the interest rate, and t is the plant life time.

$OPEX$ is the O&M cost per unit production, including energy cost and other costs:

$$OPEX(\$/m^3) = SEC \times LCOE + STEC \times LCOH + Opex_{other}$$

SEC and $STEC$ are the specific electricity and heat consumption (kWh/m^3). $LCOE$ and $LCOH$ are the levelized cost of electricity and heat, respectively. $Opex_{other}$ is other O&M cost such as maintenance, labor, insurance and so on.

7 Thermal Desalination Models

7.1 Solar LT-MED

Multi-effect distillation (MED) is currently the most thermodynamically efficient, commercialized thermal desalination technology. Its coupling with solar thermal energy has been intensely investigated by our partners at the Solar Desalination Unit of the Plataforma Solar de Almeria (PSA). The PSA pilots include flat plate solar collectors (Fig. 8a), parabolic trough concentrators (Fig. 8b), and thermal storage units (Fig. 9), connected with a 14-effect MED pilot plant (Fig. 10).

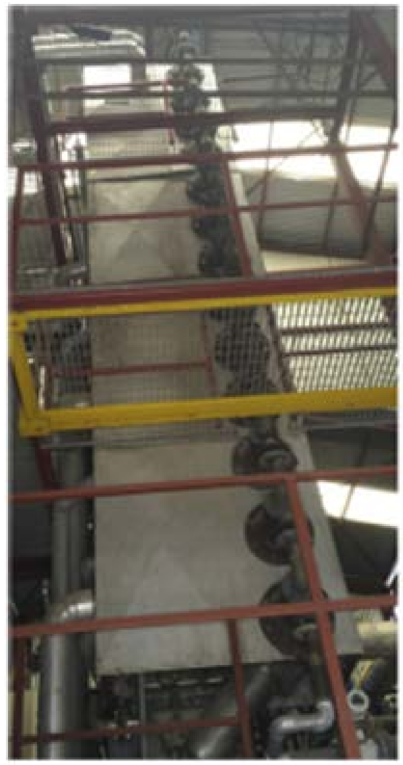


Figure 9. Solar Thermal Units at PSA *Figure 10. Front of the MED Unit at PSA* *(source: Chorak et al., 2017*[2]*)*

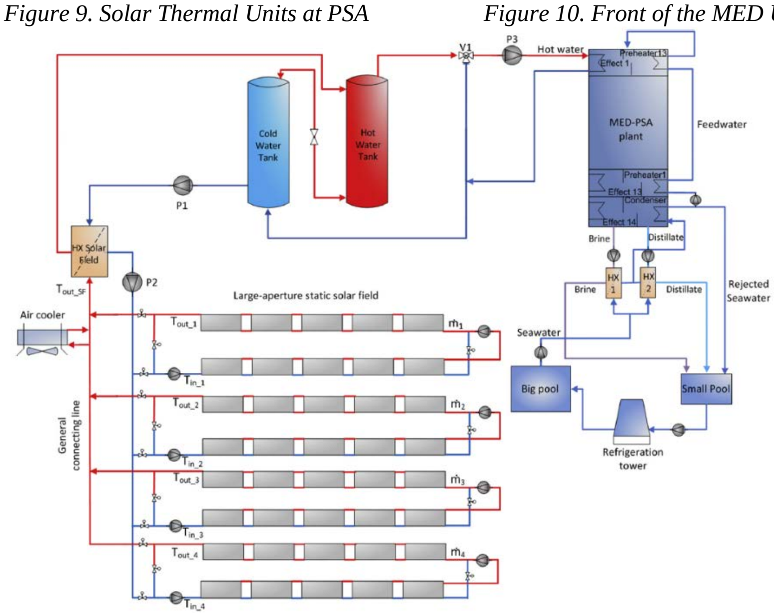


Figure 14. Schematic diagram of the flat plate collector integrated with LT- MED system[1]

2.1 Integration of the LT-MED Model

[1] Chorak, A., Palenzuela, P., Alarcón-Padilla, D. C., & Abdellah, A. B. (2017). Experimental characterization of a multi-effect distillation system coupled to a flat plate solar collector field: Empirical correlations. *Applied Thermal Engineering*, *120*, 298-313.

As we were integrating the PSA LT-MED model into the software, we noticed that the code needed changes so that the user could specify the desalination capacity. The model was written as a design tool using steam and feed flow rates as inputs to calculate the corresponding heat transfer area and the water production. The model was changed so that steam and feed flow rates could be determined from a user-specified desalination plant capacity. This was accomplished via parametric analysis using the design model developed by PSA for a Forward-Feed Low-Temperature MED plant (FF-LT-MED). The inputs of **distillate flow rate (qD)** and **heating steam temperature (Ts)** were changed within their ranges of applicability and polynomial equations were obtained by fitting the results from the simulation with Curve Fitting Tool in MATLAB; these were subsequently converted to Python scripts. The outputs considered are: **Gain Output Ratio (GOR),** the **steam mass flow rate (qs),** and the **feedwater mass flow rate (qF)**. Due to the difficulty of considering the number of effects as an input in the parametric analysis, the polynomial equations have been obtained keeping the number of effects fixed in seawater desalination systems. Different capacity and steam temperature have been considered; currently, polynomial equations have been obtained for the following number of effects ($N_{effects}$) per scenario: 12, 14 and 16. The resulting polynomial equations are shown below.

Gained output ratio: GOR= p0 + p1*Ts + p2*qD+p3*Ts^2+p4*Ts*qD

$N_{effects}$	p0	p1	p2	p3	p4	R^2
12	8.744	-0.02817	$-4.426\cdot10^{-6}$	0.0004547	$3.932\cdot10^{-8}$	0.996
14	9.683	-0.03048	$-7.639\cdot10^{-6}$	$4.896\cdot10^{-4}$	$7.227\cdot10^{-8}$	0.999
16	10.55	-0.03405	$-8.205\cdot10^{-6}$	0.0005344	$7.227\cdot10^{-8}$	0.999

Steam flow rate (kg/s)**:** qs= p0 + p1*Ts + p2*qD + p3*Ts^2 + p4*Ts*qD

$N_{effects}$	p0	p1	p2	p3	p4	R^2
12	-1.211	0.0321	0.001666	$-2.12\cdot10^{-4}$	$-5.55\cdot10^{-6}$	1
14	-1.176	0.03098	0.001503	$-2.039\cdot10^{-4}$	$-4.94\cdot10^{-6}$	1
16	-1.127	0.02981	0.001381	$-1.97\cdot10^{-4}$	$-4.48\cdot10^{-6}$	1

Feed water flow rate (kg/s)**:** qF= p0 + p1*Ts + p2*qD + p3*Ts^2 + p4*Ts*qD

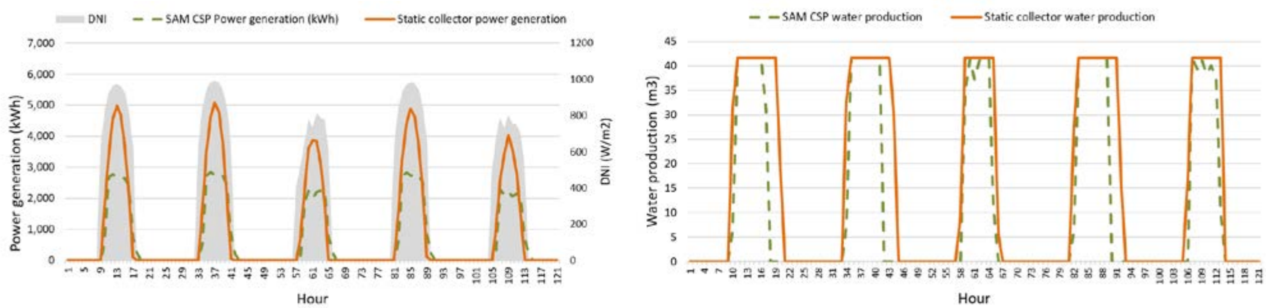
$N_{effects}$	p0	p1	p2	p3	p4	R^2
12	0.3312	-0.00881	0.03835	$5.832\cdot10^{-5}$	$-1.13\cdot10^{-6}$	1
14	0.3513	-0.0094	0.03834	$6.24\cdot10^{-5}$	$-9.89\cdot10^{-7}$	1
16	0.3312	-0.00881	0.03835	$5.894\cdot10^{-5}$	$-1.128\cdot10^{-6}$	1

Valid for the following specifications:

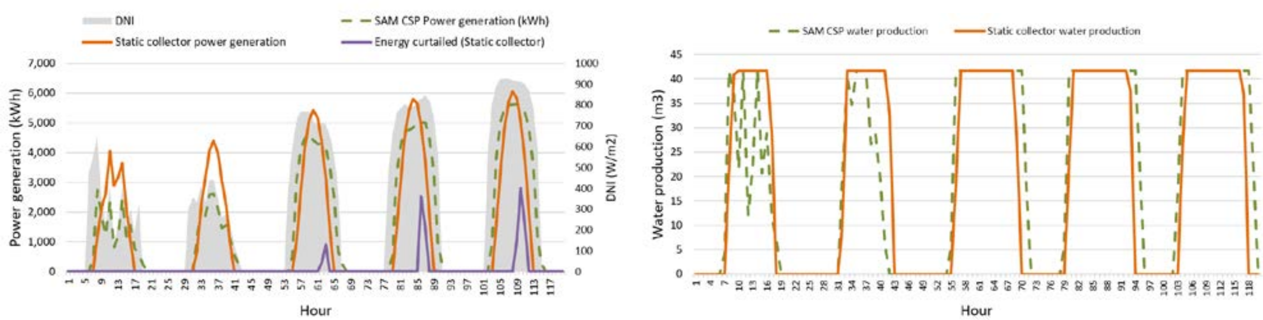
- Ts: 65-85 ºC
- qD: 2,000-10,000 m^3/day
- Intake seawater temperature: 22 ºC
- Salinity of the intake seawater: 35,000 ppm
- Salinity of the brine leaving the last effect: 50,000 ppm
- Last effect vapor temperature: 37 ºC
- Temperature difference in the condenser (i.e. temperature difference between the inlet and outlet seawater temperatures): 7.3 ºC
- DTT in the condenser (i.e. temperature difference between the temperature of the vapor that enters the condenser and the temperature of the cooling water at the outlet of the condenser: 4 ºC

- Feedwater temperature at the outlet of the last: 7ºC lower than the heating steam temperature

This MED model has been integrated with both the static collector and SAM industrial process heat (Linear Fresnel) models. A quick comparison is made between these two combinations. For a 1000 m3/day LT-MED plant located in Phoenix, AZ, the thermal energy consumption is 2.24 MW. With a solar multiple of 1.8, the solar design models result in a flat-plate static collector capacity of 4.03 MW, whereas SAM gives a Linear Fresnel (LF) solar field capacity of 4.27 MW. Both configurations include a 6-hr thermal storage system. The annual water production is estimated to be 157,571 m^3 for a flat-plate collectors, and 158,248 m^3 for SAM's LF collectors; the annual thermal power production is 9.8 GWh and 10.5 GWh respectively. Fig.1 shows the results of hourly simulations of these systems for 5 days in January and June of a typical meteorological year.



(a) January



(b) June

Fig. 1 Thermal power generation and water production for 4.27 MW SAM CSP-LF and 4.03 MW PSA flat-plate static collector systems, combined with a 1000 m3/day LT-MED plant in Phoenix, AZ

The results suggest the annual performance of the two systems are similar (7% difference in annual thermal power generation and 0.5% difference in annual water production), considering the actual capacity for the two solar fields are different (6% difference). It is also worthy to note that the linear Fresnel system tends to perform better than the flat-plate collector system during the summer since it operates longer during the morning and evening. On the other hand, the flat-plate collector system outperforms the linear Fresnel system during the winter as it utilizes global irradiation on tilted plane, whereas the LF system uses only DNI. The thermal storage system stores the excess energy from the oversized solar field and provide it to the desalination system. The overall energy loss (curtailment) during a year is 5.4% for the static collector and 0% for the LF system, which is a very small component considering a solar multiple of 1.8. The capital cost of the storage tank is compensated by the increased energy utilization and water production, so that the LCOW is reduced comparing to a non-storage system. It is also suggested in the last quarterly report.

MED Cost Model

The capital cost for MED plants, designed for seawater desalination, is estimated by an empirical equation reported by Kosmadakis et al.[2] (Eqn. 1). This equation takes the plant capacity (D) and the heat exchanger area (HEX area) as inputs; it was verified by 6 MED seawater desalination plants in the Middle East and 1 plant in Italy. The capacity of these plants varies from 9,000 to 270,000 m3/day. This cost relationship is applicable to both LT-MED and MED-TVC systems, with the HEX area being the major difference between the two types of plants.

$$C_{MED} = 6291D^{-0.135}\left[1 - f_{HEX} + f_{HEX}\left(\frac{HEXarea}{HEXarea,ref}\right)^{0.8}\right] \tag{1}$$

The O&M cost was broken down into major components including chemicals, labor, maintenance, brine discharge and miscellaneous cost based on the study from Papapetrou et al.[3]

An example of input parameters for the LT-MED cost model is shown in Table 2.

Table 2. Input parameters for a reference LT-MED costing model

Capital cost			**Output from simulation model**
Design Capacity	1000	m3/day	
Annual water production	328500	m3	Yes
HEX area	4157	m2	Yes
Expected plant lifetime	20	yr	
Average interest rate	0.04		
Cost fraction of the evaporator	0.4		
Thermal storage tank	30	$/kWh	
Energy expenses			
Specific thermal energy consumption	58.3	kWh/m3	Yes
Specific electricity consumption	1.5	kWh/m3	
Cost of heat	0.01	$/kWh	Yes
Cost of electricity	0.04	$/kWh	
Other O&M costs			
Chemicals	0.04	$/m3	
Labor	0.033	$/m3	
Maintenance	2%	% to CAPEX	
Brine disposal/discharge	0.02	$/m3	

[2] Correlations between MED capital costs and design parameters. "Kosmadakis, G., Papapetrou, M., Ortega-Delgado, B., Cipollina, A., & Alarcón-Padilla, D. C. (2018). Correlations for estimating the specific capital cost of multi-effect distillation plants considering the main design trends and operating conditions. Desalination, 447, 74-83."

[3] Summary of "Papapetrou, M., Cipollina, A., La Compare, U., Micale, G., Zaragoza, G., & Kosmadakis, G. (2017). Assessment of methodologies and data used to calculate desalination costs. Desalination, 419, 8-19."

Assuming a LCOH of $0.01/kWh, the LCOW for a relatively small (1000 m³/d) plant is 1.28 $/m³, comprising an annualized specific CAPEX of 0.53 $/m³ and 0.85 $/m³ OPEX. For large plants the LCOW goes down to $1.10/m³; the water production cost is a strong function of the LCOH; when the LCOH is essentially zero (waste heat available on site), then the LCOW can be down to $0.50/m³ (Fig. 3).

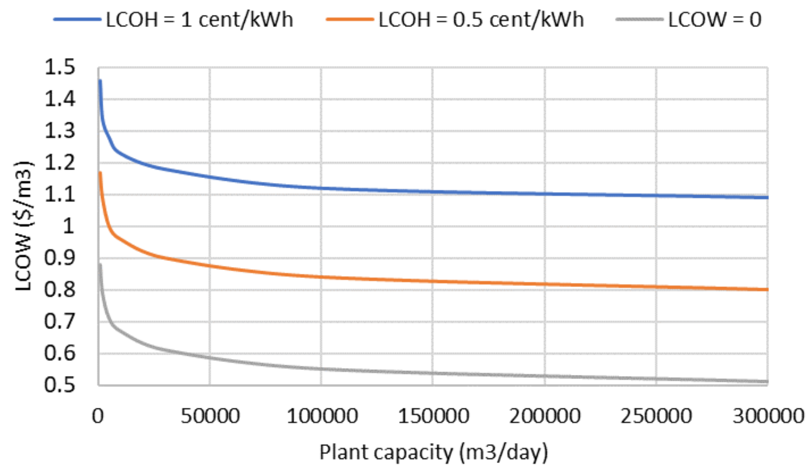


Fig. 3. The dependence of levelized cost of water on MED plant size

For the RO cost model, the reference case is described in Table 3. The LCOW is 0.95 $/m³, comprising an annualized specific CAPEX of 0.56 $/m³ and an OPEX of 0.39 $/m³.

Table 3. Cost parameters for a reference RO model

		User Inputs	**Model Estimates**
Capital Costs			
Design Capacity	m³/day	1000	
Annual water production	m³		328500
Unit Capital Cost	$/ m³/day	2500	
Expected plant lifetime	yrs	20	
Average interest rate		0.04	
Membrane area	m²		2284.8
Membrane cost	$/m²	50	
Energy Use			
Specific electricity consumption	kWh/ m³		2.5
Cost of electricity	$/kWh		0.05
O&M			

Chemicals	$/ m^3	0.05	
Labor	$/ m^3	0.1	
Membrane replacement	$/ m^3	0.05	
Brine disposal/discharge	$/ m^3	0.03	
Other maintenance	$/ m^3	0.03	

Figure 4 shows the unit capital cost of MED as a function of capacity, based on the Global Water Intelligence (GWI) dataset. Additionally, the figure shows cost-capacity curves compiled from data in the literature[4,5] as well as our own fit to the GWI data (last updated in 2019) after bin-

averaging (sample size= 701, bin size= 10000). While the data agree with the prediction curve by Kosmadakis et al. for large plants, the Kosmadakis curve seems to underestimate costs for smaller plants under 10,000 m3/d.

7.2 Solar TVC-MED

A model of a TVC-MED, developed by Drs. Palenzuela, and Alarcón-Padilla, is based on the Trapani, Italy 12-effects MED desalination plant and is validated by actual performance data from this plant. The model is built in Engineering Equation Solver (EES), and it can calculate the gain output ratio (GOR) from different motive pressure and different thermo-compressor locations by operating balance equations of energy, mass, and salt of each effect. We converted the EES model into Python so that it will be available for public use in this open-access software and used approximations and empirical relationships to simplify the equations. Running of a reference case with the TVC at the last effect, produced the results with less than 1% deviation with most of the measurements from the Trapani plant, with the exception of cooling mass flow rate and number of tubes for which the results differ by 12% and 6% correspondingly.

[4] Kosmadakis G, Papapetrou M, Ortega-Delgado B, Cipollina A, Alarcón-Padilla D-C. Correlations for estimating the specific capital cost of multi-effect distillation plants considering the main design trends and operating conditions. Desalination. 2018 Dec 1;447:74–83.

[5] Rahimi B, May J, Christ A, Regenauer-Lieb K, Chua HT. Thermo-economic analysis of two novel low grade sensible heat driven desalination processes. Desalination. 2015 Jun 1;365:316–28.

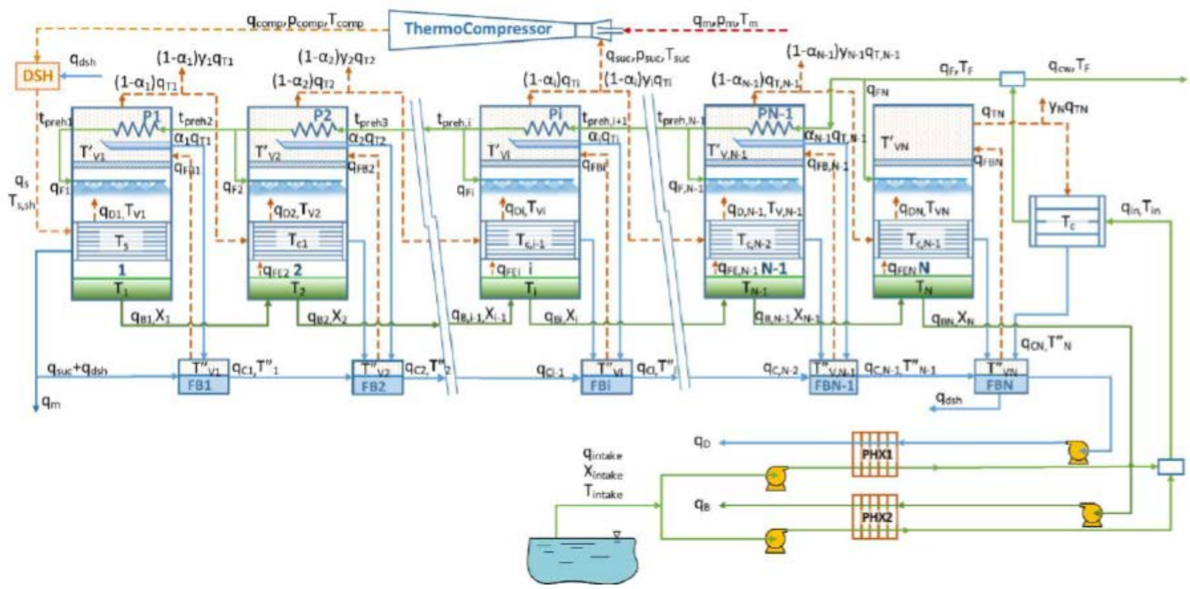


Fig. 3. Schematic of the PC-MED-TVC plant[i]

7.3 Solar MD

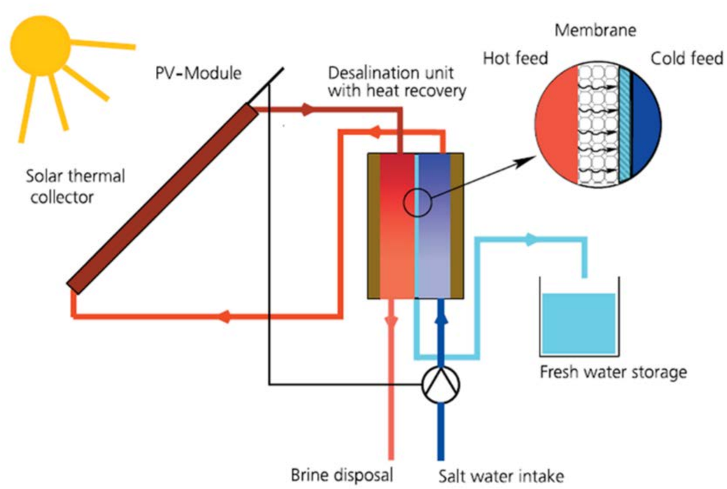


Figure YY. Schematic diagram of the flat plate collector and MD system

Table 4 below specifies the system design parameters of the solar field given the same desalination capacity. PSA described the performance of the new MD system this year, and as a comparison to the old one, it exhibits a lower specific energy consumption although it is connected with the same type of flat plat collector-AGM system. In Table 4 we list the specifications of both systems to highlight the recent drastic improvement in technology development. The results and discussion in this section relate to modeling the new system.

The distillate water production is estimated from the thermal energy provided by the solar field. For LT-MED system, the nominal required thermal energy (P_{req}, kW) is calculated from the gained output ratio (GOR), which is assumed to be 10 in this case.

$$P_{req} = \frac{1}{GOR} \times Capacity \times (h_{vsat} - h_{lsat}) \times \rho_{dist} \times \frac{1}{24 \times 60 \times 60}$$
$$M_{d\_rel} = -0.23064P_{rel}^2 + 1.4384P_{rel} - 0.184$$

M_{d_rel}: Distillate water production relative to nominal production rate (%)
P_{rel}: Thermal load relative to nominal thermal energy consumption (%)

For AGMD system, AS24C5L modules from Aquastill were selected for our modeling. The following equations were derived from PSA's experimental data to estimate the permeate flux ($P_{flux},\ l/h \cdot m^2$) and the specific thermal energy consumption ($STEC,\ kWh/m^3$):

$$P_{flux} = -1.088 + 0.024 \cdot T_{evap} - 0.018 \cdot T_{cond} - 0.001 \cdot F + 0.00006 \cdot T_{evap} \cdot F$$
$$STEC = 479.373 - 39913 \cdot T_{evap} + 0.16429 \cdot F$$

T_{evap}: Evaporator inlet temperature (℃), the nominal value is 80 ℃
T_{cond}: Condenser outlet temperature (℃), the nominal value is 25 ℃
F: Feed flow rate (l/h), the value is optimized to 582.7 l/h to maximize P_{flux} while minimizing STEC

Zaragoza et al. developed new empirical equations for simulating the performance of the Aquastill V-AGMD (vacuum air-gap membrane distillation) pilot plant, their best performing MD system so far. A schematic of this pilot is shown in Fig. 4. Its performance was compared with AGMD system with the same modules, which include AS7C1.5L and AS26C2.7L, with an effective membrane area of 7.2 m^2 and 25.9 m^2 respectively. Both modules have spiral-wound geometry, and the membrane material is the same for both, made of low-density polyethylene. AS7 has 6 spiral envelops, while AS26 has 12 envelopes. The differences in the number of channels and their lengths lead to different residence times, which in AS26 module is 3.6 times longer than that in AS7 module.

The Zaragoza et al. model is based on their actual testing at PSA [6]. The model is based on empirical equations applicable for a feed salinity range of 35 g/L to 105 g/L, while the pilot-plant investigation of the AGMD and V-AGMD systems took place at an extended feed salinity range i.e., 35 g/L to 292 g/L. Fig. 5 shows the advantages of V-AGMD and AGMD in terms of water production and thermal energy consumption over a large range of feed salinity. Zaragoza believes that MD using multi-envelope spiral wound modules in V-AGMD operation is currently the best available technology for desalination of highly concentrated solutions and he expects drastic reductions of the STEC to be accomplished during the current year of testing at the PSA facility.

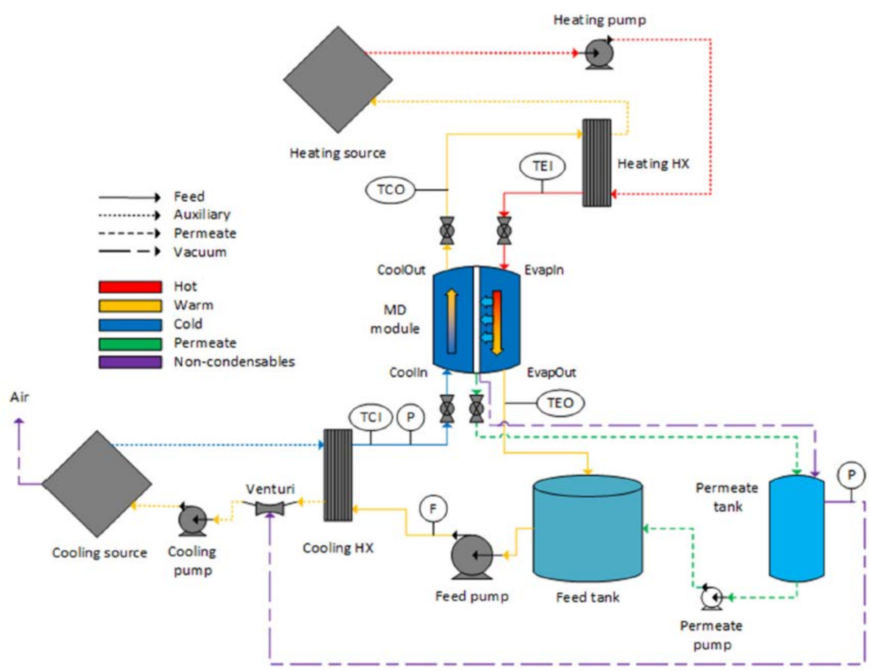


Figure 4. Layout of the Aquastill V-AGMD system[1]

[6] Andrés-Mañas, J. A., Ruiz-Aguirre, A., Acién, F. G., & Zaragoza, G. (2020). Performance increase of membrane distillation pilot scale modules operating in vacuum-enhanced air-gap configuration. *Desalination*, *475*, 114202.

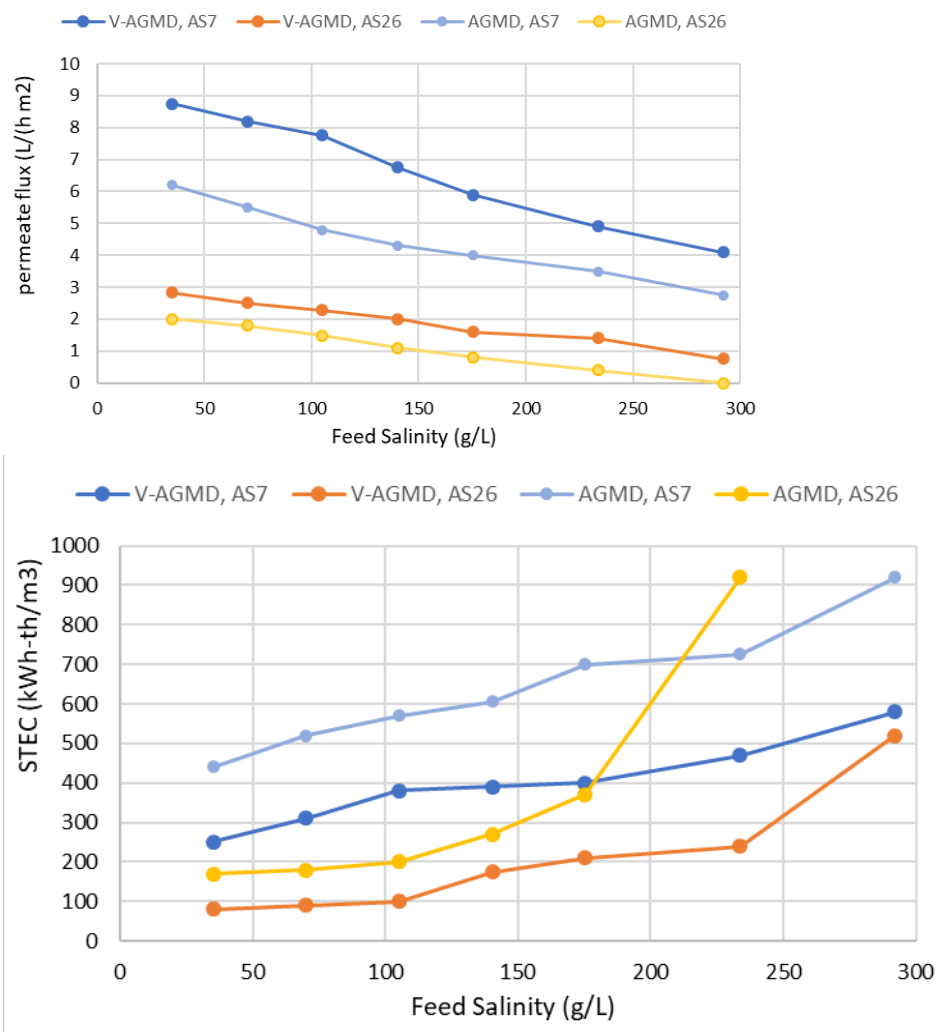


Figure 5. Comparative results of permeate flux and STEC in experiments of AGMD and V-AGMD. (Feed flow rate =1100 l/h, inlet evaporator temperature=80 °C, inlet condenser temperature = 25 °C)

As shown in Fig. 5, the shorter residence time VAGMD, AS7 module exhibits the highest permeable flux (PF) whereas the VAGMD, AS26 module shows the best (lowest) STEC. It appears that the reason for this is that for the same driving force (vapor pressure), it's harder for a vapor flux to take place through the membrane with larger residence time.

For the thermal energy consumption, theoretically the STEC = Qth/V water, is proportional to the evaporator inlet temp (TEI) minus the condenser outlet temp (TCO), where TEI is fixed at 80 deg C (parameters shown in Fig. 4). As shown, TCO for AS26 would be larger than that for AS7, when TCI (condenser inlet temp) is the same, considering a longer heat exchange delta T. So the total thermal energy consumption is actually 3 times larger in AS7 than AS26. Even with a higher Pflux, the STEC for AS7 is still larger. The Zaragoza team is currently working to optimize the PF and STEC combination.

The Zaragoza et al. model was scripted in Matlab. We successfully converted this to Python and compared the results of the new model with those of the original model and the field data. The modeling results from the Python script were identical to those from the Matlab model (0% error) and the modeling results agreed within 5% with the field measurements of permeate flux, and within 15% with the specific thermal energy consumption (STEC) data.

Then, we simulated a case study for a system operating in Phoenix, AZ. The solar field is sized as 224 kW, the nominal power required for a 100 m^3/day LT-MED desalination system. The feed salinity is a constant 35 g/L. Fig. 6 shows comparisons of the performance of the two VAGMD systems, given the same thermal energy input. Fig. 7 provides a close look at the hourly water generation for these three systems.

It can be seen that MED has in overall a higher thermal efficiency than VAGMD systems. As discussed before, AS26 module has a lower STEC and thus can produce more water than AS7 does with the same solar field area. On the other side, it would need more modules as the permeate flux is lower. In this case, 38 modules are needed for AS26 while 13 modules are enough for AS7.

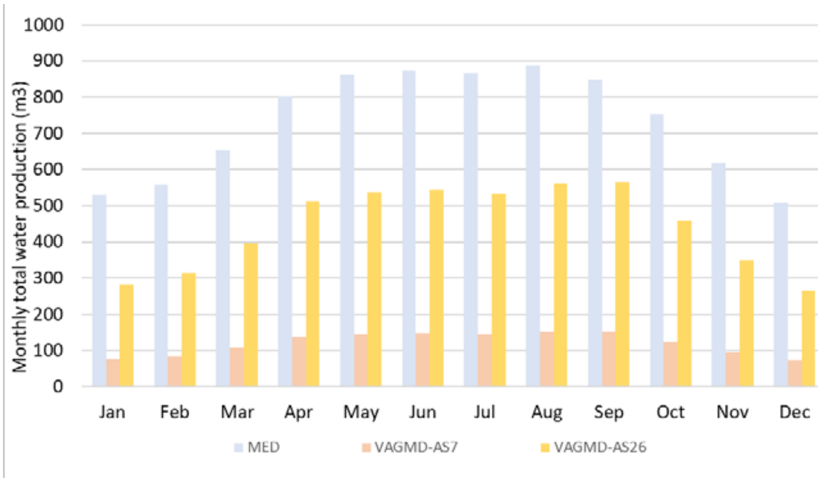


Figure 6. Monthly water production for MED and VAGMD powered by 224 kW flat collector in Phoenix, AZ (38 modules for AS26 and 13 modules for AS7)

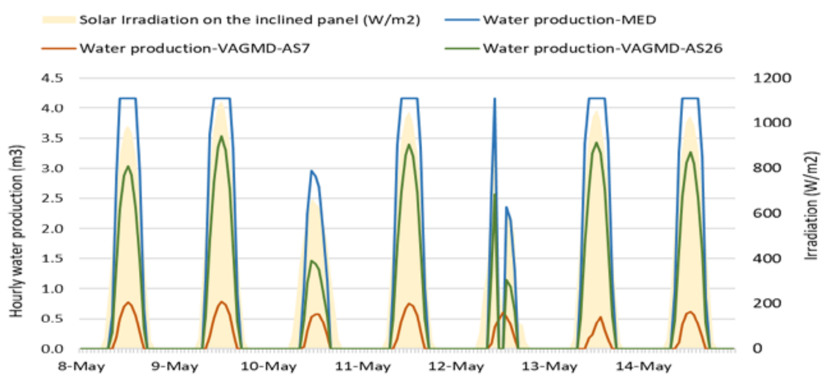


Figure 7. Hourly water production for MED and VAGMD during a week

VAGMD Cost Model

Based on the performance of the pilots at PSA, we developed a preliminary model of costing MD systems the results of which are summarized below.

Assumptions:
Evaporator Channel Inlet Temperature = 80 ºC
Condenser Channel Inlet Temperature = 25 ºC
Feed Flow Rate = 582.7 L/(h*module)
Feed Concentration = 35 g/L
It is noted that these operation conditions were used at PSA V-AGMD pilot plant testing to maximize the energy efficiency of the larger V-AGMD module (25.9 m^2), under which the lowest specific thermal energy consumption of 49 kWh_{th}/m^3, equivalent to the highest GOR of 13.5, is achieved.
Our cost modeling yields the cost curves shown in Fig. 8 for two V-AGMD modules that were studied at PSA; AS7C1.5L stands for the smaller module with channel length of 1.5 m and effective module area of 7.2 m^2 and AS26C2.7L stands for the larger module with channel length of 2.7 m and effective module area of 25.9 m^2. Under the conditions listed above, the larger module gives the lowest LCOW.

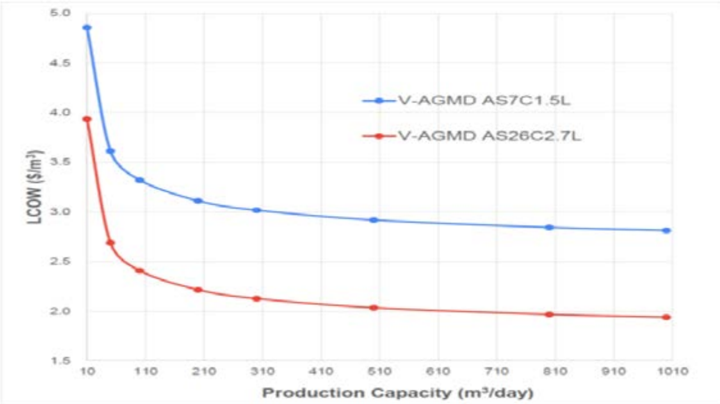


Figure 8. Cost comparison between two V-AGMD systems

However, the operating conditions can greatly affect the cost-based selection of one or the other module. For example, when assuming the following conditions, the smaller module gives the lowest LCOW (Fig. 9):
a) Evaporator Channel Inlet Temperature = 80 ºC (normally maximized at 80 ºC to ensure the highest possible energy efficiency for each operation).
b) Condenser Channel Inlet Temperature = 20 ºC (which is the lowest allowed temperature in PSA V-AGMD performance modeling)

c) Feed Flow Rate = 400 L/(h*module) (which is the lowest allowed feed flow rate in PSA V-AGMD performance modeling)
d) Feed Concentration = 105 g/L (which is the highest allowed seawater salinity in PSA V-AGMD performance modeling)

This result shows that not only designated daily production capacity, but also the operational parameters play a major role in affecting the final fresh water production costs. In this case, the high-salinity feed source seawater reduces the vapor pressure across the V-AGMD module, resulting in an even lower permeate flux at a rather low feed flow rate for the larger module. Even if it has a larger module area, the resultant low permeate flow rate requires a higher number of V-AGMD modules in parallel operation to reach the nominal fresh water production capacity. A sharp increase on capital expenditures is caused correspondingly, for which the advantage on its operational expenditures due to low thermal energy consumption is not able to compensate. All in all, there is no guarantee regarding which module works the best in terms of costs. In principle, the operational parameters will change the extent to relative advantage and disadvantage of CAPEX and OPEX between two modules at any time, thereby altering the LCOW advantage of one versus the other.

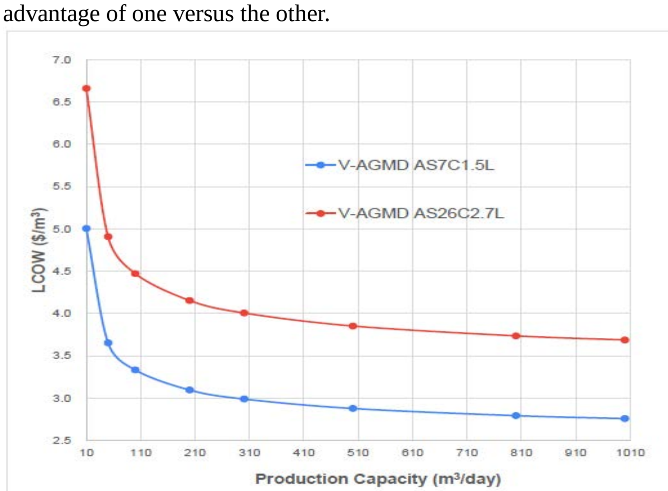


Figure 9. Cost comparison between two V-AGMD systems under different operating conditions

In our cost model, operational parameters include evaporator channel inlet temperature, condenser channel inlet temperature, feed flow rate per module and feed concentration. Additionally, there are V-AGMD plant design parameters which are part of user's inputs and will be listed in detail in the user's manual. Since those design parameters are specific to each project location and condition, there is no way to generalize their ranges accurately. As was the case with the LT-MED example described in section 2.1, the biggest parameter among all the operational and design parameters that is affecting the costs of fresh water production turns out to be the unit cost of heat ($/kWh$_{th}$). As of now, it is left as 0.01 $/kWh$_{th}$ in our model, and the modeling results show that thermal energy cost accounts for over 80% of OPEX. If the user is able to implement free waste heat or low-price low-temperature geothermal energy as thermal energy source for MD system, where the unit cost of heat could be even lower than 0.002 $/kWh_{th}$, then the percentage of thermal energy cost can be lower than 50% and OPEX can be largely reduced by over 70%. Eventually, LCOW is reduced by more than 60%.

7.4 Solar MD Batch

Batch MD pilot experiments at Plataforma Solar de Almeria (PSA) by the Zaragoza team, employed the Aquastill modules AS7C1.5L and AS26C2.7L. Models of those systems were developed applying a Box-Behnken design of experiments and were tested and validated under multiple stationary operating conditions.

The system shown in Figure 6 works in recirculation. There is a permeate tank that releases intermittently (every 3,22 litres); this was used as discretization interval for the batch model, since it would be more physically adequate than employing equal time increments.

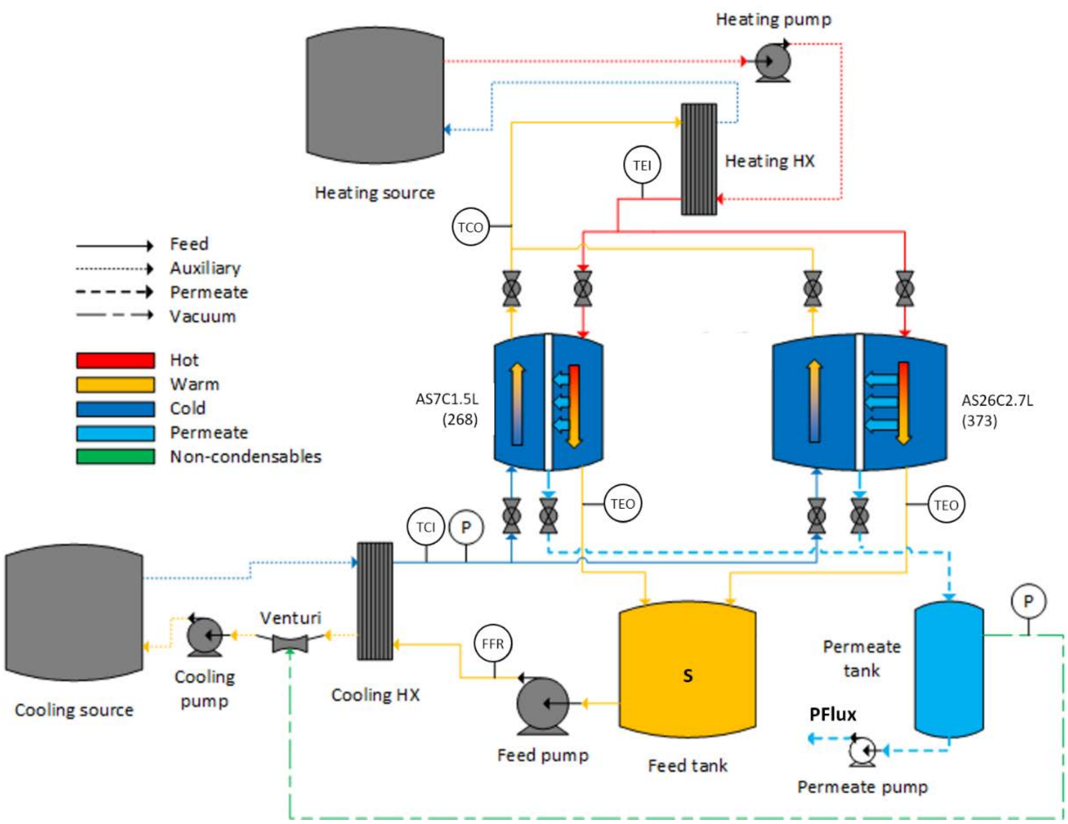


Figure 6. Schematic of the MD-Batch-MD pilot plant at the PSA field-test facility.

The experiments allow to simulate the permeate productivity (PFlux), the outlet temperature of the feed stream that circulates through the evaporation channels (TEO), and the outlet temperature of the feed stream that circulates through the cooling channels of the module (TCO), which is used to calculate the thermal energy consumption. Batch MD operation is performed in a single recirculation step from an initial feed volume, taking as inputs the inlet temperature of the feed into the evaporation channels (TEI), the feed flow rate (FFR), the inlet temperature of the feed into the cooling channels (TCI), and the initial feed salinity (S). Based on the TCO, the model estimates the thermal power (ThPower) and the thermal efficiency, that is given commonly in thermal desalination by the specific thermal energy consumption (STEC) and, alternatively, by the gained output ratio (GOR). Valid ranges of operating conditions are 60 ^{o}C $\leqslant$ TEI $\leqslant$ 80 ^{o}C; 20 ^{o}C $\leqslant$ TCI $\leqslant$ 30 ^{o}C; 400 l/h $\leqslant$ FFR $\leqslant$ 1100 l/h. The allowed range of feed salinity (S) is 35 g/l $\leqslant$ S $\leqslant$ 175.3 g/l for module AS7C1.5L(268) and 35 g/l $\leqslant$ S $\leqslant$ 105 g/l for module AS26C2.7L(373).

To develop the models, TEI, FFR and TCI are considered constant throughout the duration of the treatment. In addition, full mixing in the feed tank is considered, without temperature or concentration gradients, and feed losses through the permeate channels are supposed negligible. When the code is executed, the user is asked for the required inputs to carry out the simulation. The module to be simulated, the initial feed salinity, and the recovery ratio (RR) to be achieved will be specified. The software asks subsequently for the initial batch volume and the operating conditions, i.e., TEI, FFR and TCI. With all the required inputs, the simulation begins automatically, taking a constant permeate volume of 3.22 litres as the discretization interval of the process. This corresponds to the volume that the permeate tank discharges from the Aquastill unit during the operation. Permeate discharge is carried out discontinuously by means of a system that allows to maintain the vacuum level within the modules, even when the permeate tank is opened to the outside. By choosing this volume, the discretization is done in more physically similar intervals regarding the operation than if a constant time was chosen.

First of all, PFlux is calculated, and then the treatment time (t), the permeate flow rate (PFR) and the current RR, using the membrane area (A) of the simulated module. Subsequently, TEO and TCO are calculated for the current iteration with the model equations, and from them the mean logarithmic temperature difference (ATml), the thermal power (ThPower), the temperature in the feed tank (Ttank), the thermal energy per iteration (ThEnergy) and accumulated (AccThEnergy), the cooling power (CPower), the cooling energy per iteration (CEnergy) and accumulated (AccCEnergy), and the typical thermal performance indicators in thermal desalination: Gained Output Ratio (GOR) and Specific Thermal Energy Consumption (STEC). The equations of the model are solved iteratively until the target Recovery Rate given by the user at the beginning of the execution is reached, and a table with the results of each iteration is presented to the user.

So far, the model has a limitation in maximum salinity (especially in the large module) because our collaborators at the PSA, Spain (Zaragoza and coworkers) have not been able to complete the experiments yet due to COVID-19. Mostly they have been dealing with snags on instrumentation. Their aim is to extend the salinity range of the experiments and reach with the 2.7L module at least the same high salinity as with the 1.5L module. They have also worked with 250 g/L brine feed, but the life of the module is jeopardized. Thus, they do not think that they can operate sustainably with current commercial modules at those higher concentrations because of limitations mostly associated with commercial membranes. Regarding the advancement of the MD technology, the next step at PSA is to build larger units with several modules operating together in parallel, in order to increase the capacity and optimize the costs. Large pumps are cheaper and more efficient that small pumps, so working with larger flow rates in units comprised of ten or more modules would decrease the resulting LCOW (reducing specific CAPEX and also OPEX associated to the electrical consumption of the pumps, without altering the specific thermal energy consumption given the modularity of the system).

7.5 Solar FO

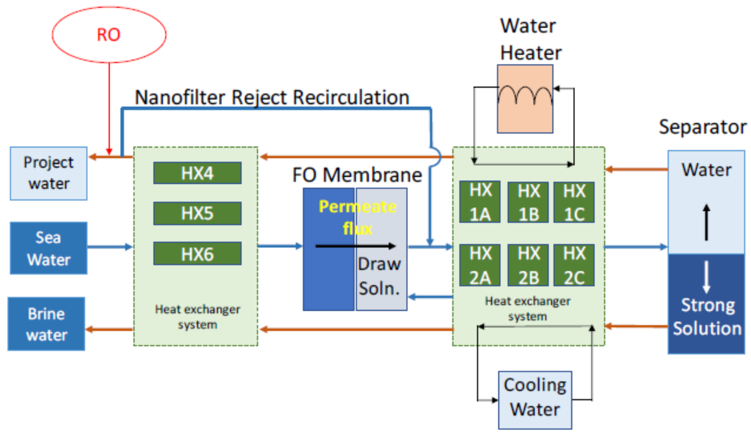


Fig 9. Simplified schematic of mass flows in the Trevi Systems FO prototype

Generalization of Trevi's FO model

Converted from Excel to Python and generalized a Trevi FO desalination model designed for seawater desalination at constant temperature and specific draw solution; a schematic of the model is shown in Fig. 3. The model's functionality was enhanced to simulating different feed salinities, feed water temperatures, system capacities and draw solution properties.

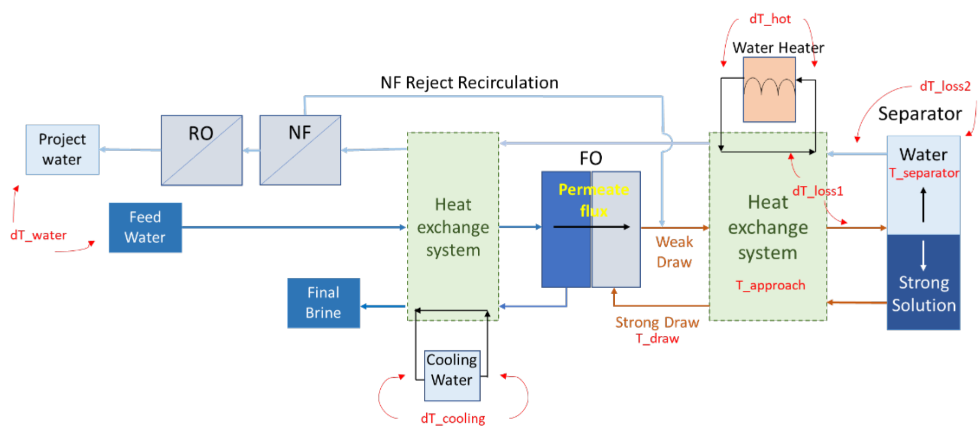


Fig. 3. Schematic of the Trevi FO desalination prototype

Our enhancements of the model included the following:

2.1.1 Generalizing design inputs

Introduced design parameters to significantly reduce the number of operational parameters and make the model friendlier to the user and more dynamic with fewer input variables. These parameters are shown in Table 1 and labeled in Fig. 3.

Table 1. New Operational Parameters in the FO Model

New operational parameter	Defaulted value

	(Based on Excel)
Temp. of the draw solution entering the membrane system (T_draw)	20 ºC
Temp. difference between the inlet/outlet supplemental seawater (dT_cooling)	6 ºC
Temp. difference between the feed and produced water (dT_water)	10 ºC
Operational temperature of separator (T_separator)	90 ºC
Temp. loss in the separator (dT_loss2)	1 ºC
Temp. difference between hot water and solution entering the separator (dT_loss1)	3 ºC
Temp. difference between inlet and outlet hot water (dT_hot)	10 ºC
Approach temperature at HX 1C and 2C (T_approach)	5.28 ºC

These variables should be determined by the user demand and the properties of draw solution, and they help to determine the solution temperature within the heat exchangers.

2.1.2 Coding Variable Dependencies

We identified dependencies among variables and created solvers to further reduce the number of model variables. These solvers include:

- Membrane temperature solver (function T_memb_solver in the Python code):
  This solver optimizes the strong draw solution flowrate and the brine temperature, so that the heat balance in the membrane system and HX 4&5 converge. It helps determine the solution properties entering HXs downstream.

- Membrane delta_T solver (function find_deltaT in the Python code):
  It's a solver embedded in the first solver that estimates the heat transferred to weak draw and outgoing brine at the membrane, and thus determine the delta_T.

- Approach temperature solver (function T_app_solver in the Python code):
  This solver is designed to determine the solution temperature within a certain HX, given the designed approach temperature.

Fig. 4 shows the simulated STEC (specific thermal energy consumption) of the system with different seawater temperature, feed salinity and FO recovery rate. We noticed that STEC increases as the seawater temperature increases. The reason is that the brine osmotic pressure increases as its temperature increases (a function of inlet seawater temperature). It results in a higher minimum osmotic pressure for the draw solution, and the draw solution cannot be diluted as much. As a consequence, more draw solution needs to be circulated for a certain amount of water production, thus the STEC increases.

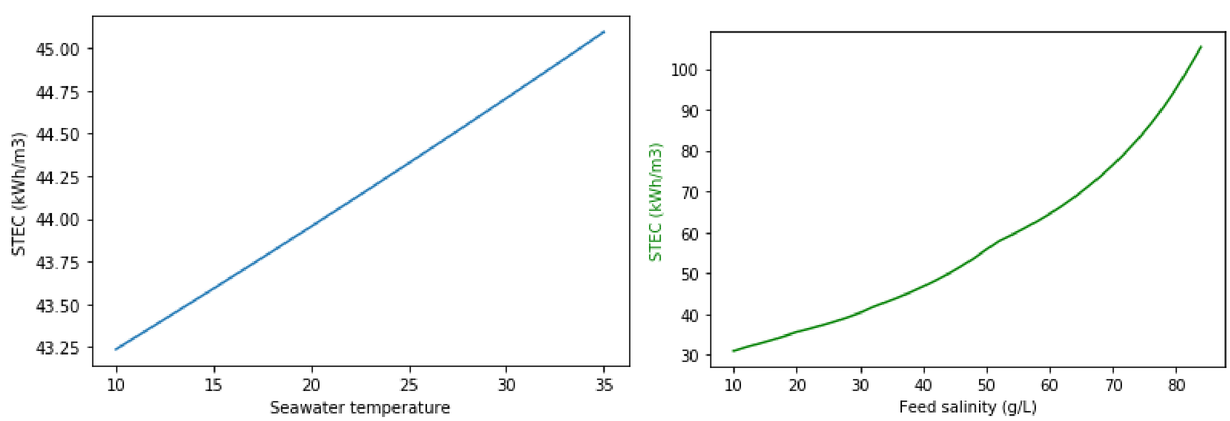


Fig. 4. System STEC varies with temperature (left: salinity fixed as 35g/L) and

feed salinity (right: feedwater temperature fixed as 15 oC and FO recovery rate as 30%)

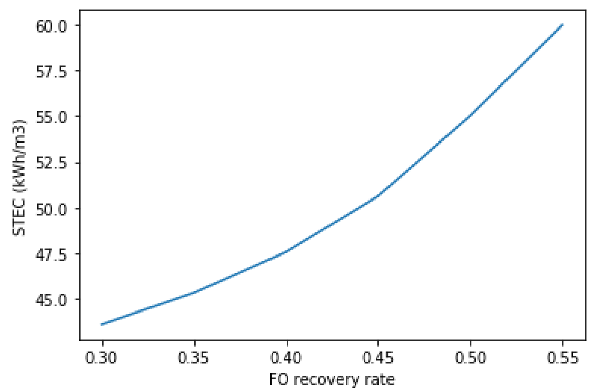


Fig. 5. System STEC varies with FO recovery rate

2.1.3 Updating Cost Estimates

We also gathered and implemented the latest costing numbers on the FO system from Trevi Systems Inc. For large scale plant (~10,000 m^3/day) and pilot scale plant (~500 m^3/day), we applied \$4,500,000 and \$1,333,000 as the capital cost respectively. The CAPEX per unit capacity is \$450 and \$2,667 per m^3/day for these two scales, and they are implemented in the GUI separately.

The breakdown of the CAPEX is shown in Table 2.

Table 2. FO Cost breakdown (source Trevi Systems)

CAPEX breakdown	CAPEX Percentage
FO membranes	11.6%
Heat exchangers	13.9%
Construction	22.5%
Draw solution	9.0%
Coalescers	4.9%
Structural	4.5%
Polishing	8.3%
Pipes and plumbing	5.3%
Pre-filtration	4.9%
Controls/Electrical	3.4%
Pumps	4.7%
Instrumentation	2.6%
Valves	2.4%
CIP	1.1%
Tanks	1.1%

Fig. 8. New menu enabling Parametric Studies

At this stage, one or two numerical variables can be selected from either technical models or financial models for parametric study; a sample input to parametric studies is shown in Fig. 9 and sample outputs in Figures 10 and 11.

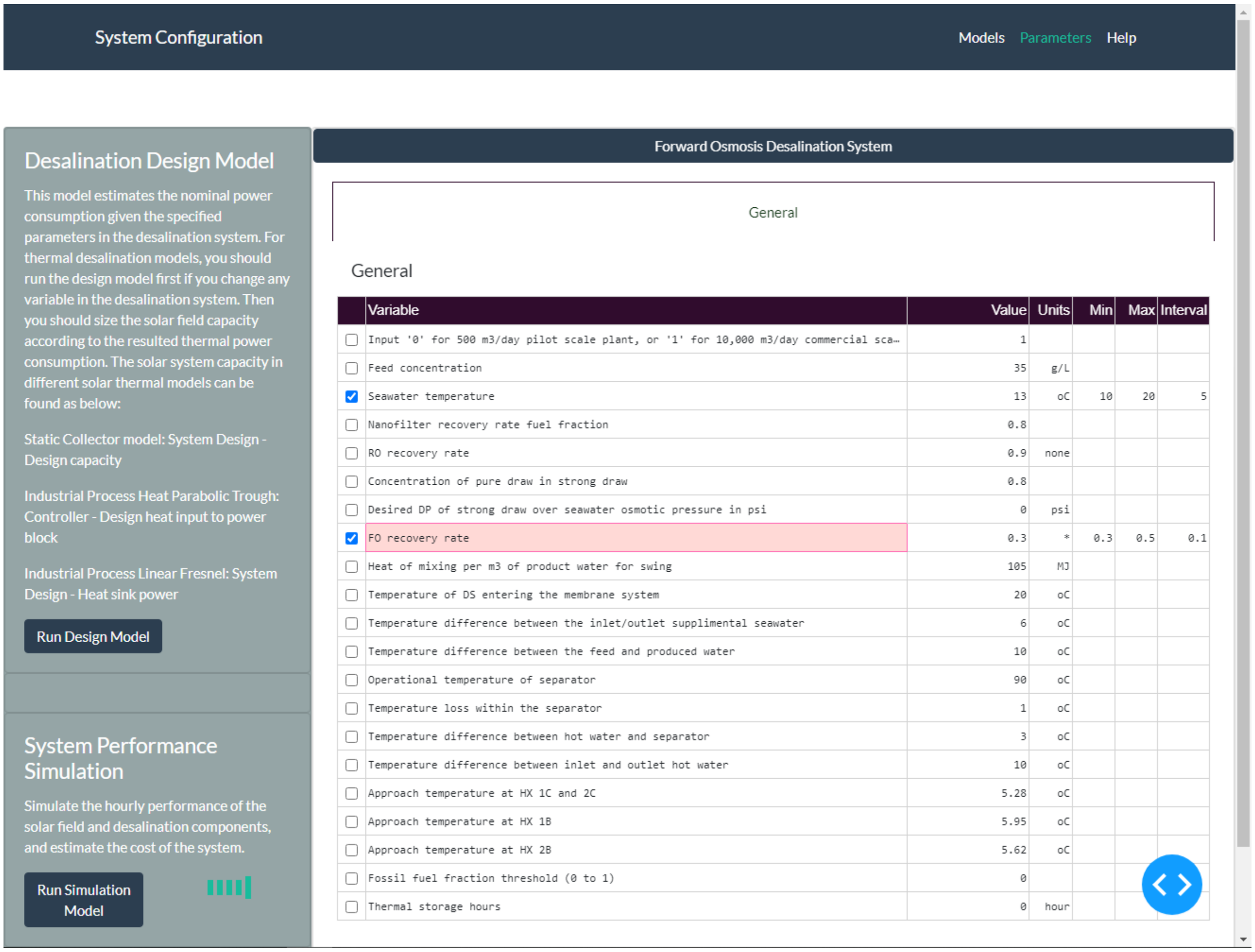


Fig. 9. Seawater temperature (10, 15 and 20 ºC) and FO recovery rate (30%, 40% and 50%) are selected for parametric study

The user will be able to specify a range with increment in the parametric mode. The model will execute multiple times based on combinations defined by the users. In the example shown in Fig. 9, nine cases will be simulated and the results will be stored separately. We collected the results and generated bar graphs in the next figures for a quick comparison of those cases.

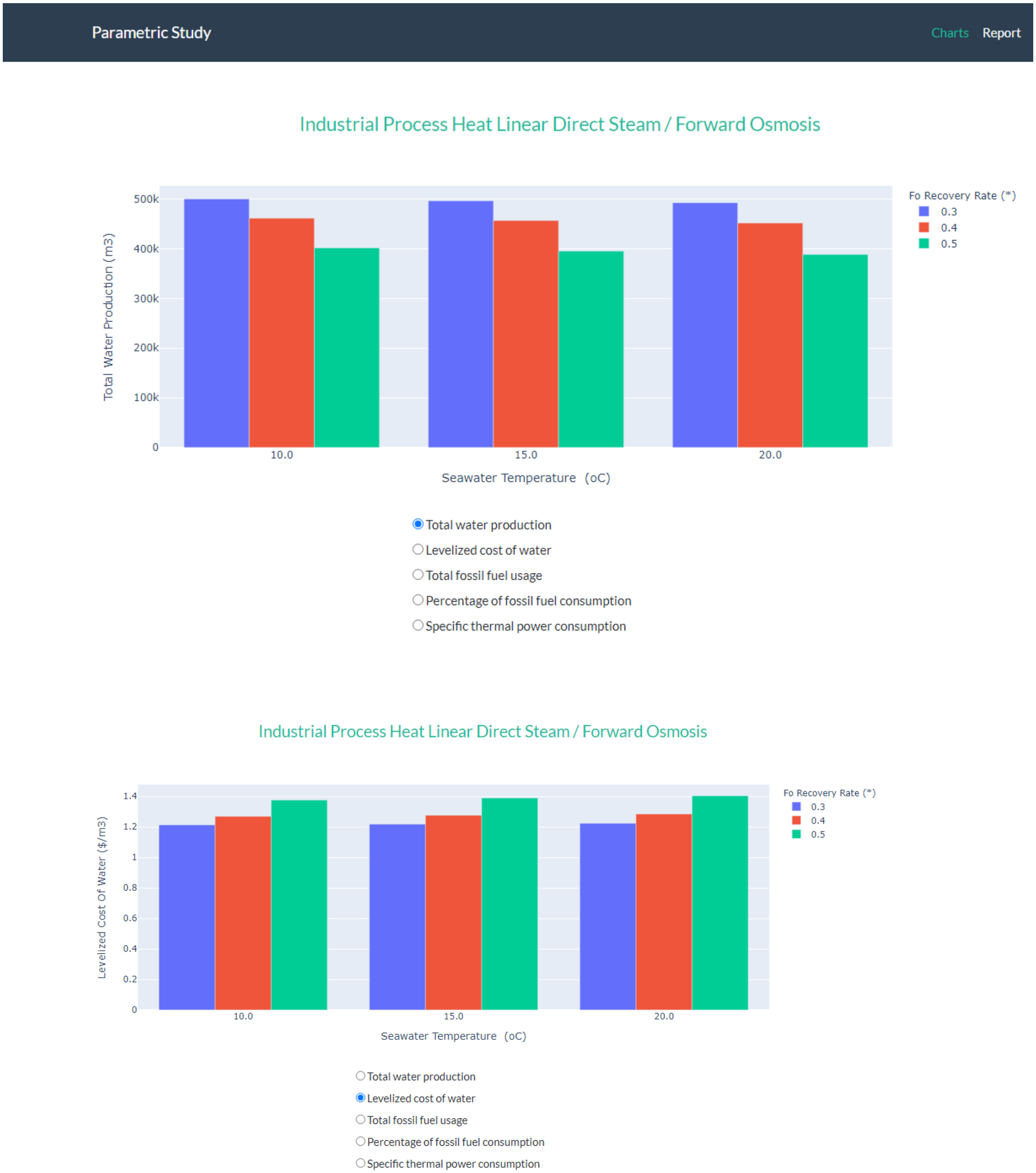


Fig. 10. Sample outputs from parametric study. Bar graphs of total water production(top) and LCOW(bottom) of a 10,000 m3/day FO plant coupled with a 20 MW industrial process heat linear Fresnel solar field

8 Pressure-Driven Desalination Models

8.1 RO

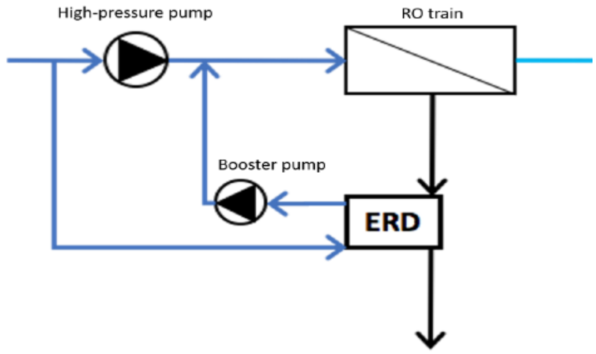


Figure 10. Simplified schematic of one-stage RO

8.2 Enhanced RO

9 Hybrid Desalination Models

9.1 RO-FO

2.5 Preliminary modeling of a hybrid RO-FO system

A proposed[7] design of an RO-FO system, for treating low salinity water (6,000 ppm) with a claimed recovery rate of 75% and 90% for RO and FO correspondingly, is shown in Figure 10.

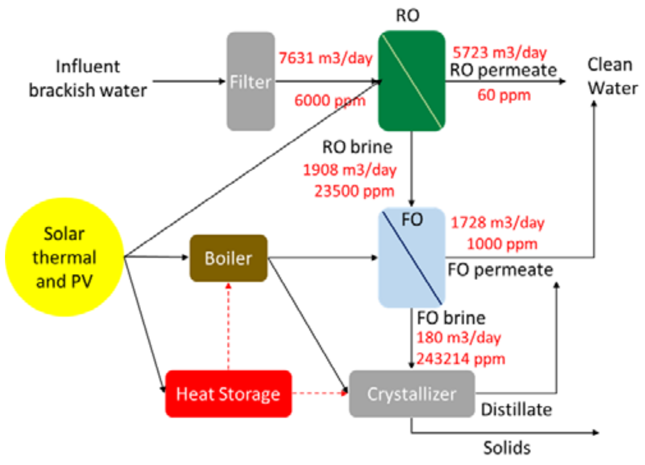


Figure 10. Conceptual design of hybrid RO-FO system with operational parameters from WaterFX

With the current RO model from Columbia group and FO model from Trevi Company, we simulated this conceptual hybrid system powered by PV and static flat collector. Table 4 shows the energy consumption for RO, FO and crystallization process and Table 5 gives the size of PV and flat collector fields.

Table 4. Energy consumption for each process

	RO	FO	Crystallizer	Total

[7] WaterFX conceptual design report. The FO system was developed by Oasys and is currently owned by Woteer Company, China.

Specific electricity energy (kWh/m^3) *	3.16	2.60	0.77	-
Specific thermal energy (kWh/m^3) *		29.5	178	-
Nominal electricity consumption (kW)	754	187	6	947
Nominal thermal energy consumption (kW)		2121	1335	3456

* The specific energy consumption regards to 1 m^3 of permeate flow for RO and FO process or 1 m^3 of brine fed into the crystallizer.

Table 5. Size of the solar field

	Module	Nameplate capacity	Total module area (m^2)	Capacity factor (%)
PV (tilt-fixed)	SunPower SPR-E19-310-COM	1132 (kWdc)	5948	20.9
Flat collector	Wagner Solar LBM HTF10	3457 (kWth)	6444	21.9

Figure 11 shows an overall performance of the system located in Phoenix, AZ, where the annual DNI is 2,520 kWh/m^2. With solar multiple of 1 and 0-hour storage system, the capacity factor of desalination is around 20%. Figures 12 and 13 give a detailed hourly simulation of the system at the same location, over a course of a week in May. The load rate indicates the percentage of energy provided from the solar field. It is shown that at different hours either electricity or thermal energy may limit the performance of the system. In most days, thermal energy generation is limiting the water production in the morning and evening (with lower load rates), while electricity generation is the limiting factor in the middle of the day. We plan to develop an algorithm for optimizing the sizing of the PV-static collector combination.

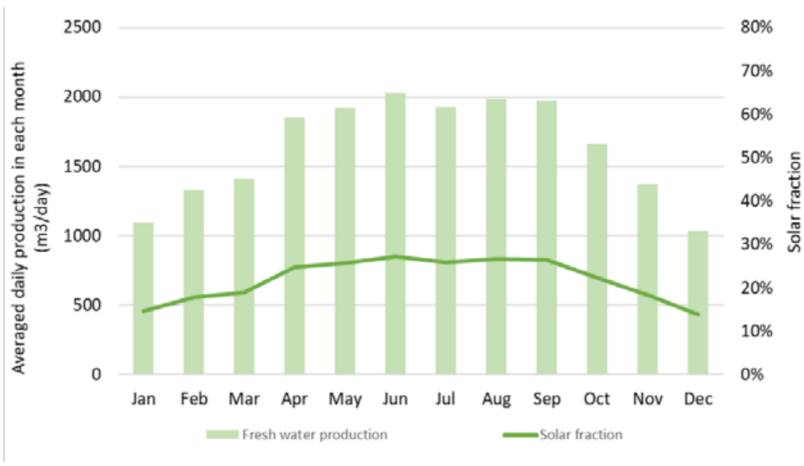


Figure 11. Monthly water production (m^3/day) and averaged solar fraction (energy contribution from solar field) for a RO-FO-crystallizer system (Production capacity: 310 m^3/h) in Phoenix, AZ

Future implementations include the logic of partial load operation in cases of hybrid desalination system. A proper sizing strategy between PV, solar thermal and storage can be developed to maximize the load rate while minimizing the solar field area.

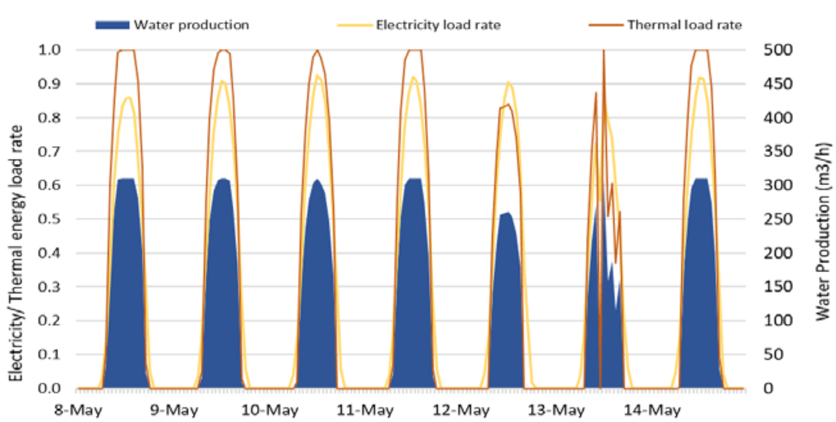


Figure 12. System performance during a week in May in terms of power load rate (powered by with a 1,123-kW PV field and a 3,456-kWthermal flat-plate collector field)

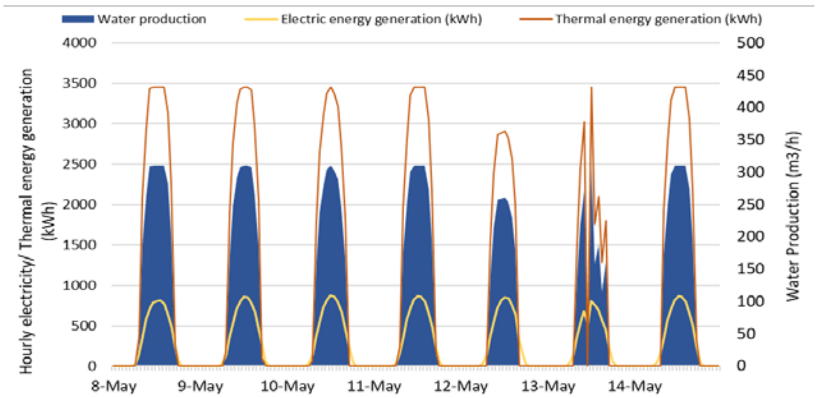


Figure 13. System performance during a week showing hourly electricity and thermal energy demand of a RO-FO system powered by 1,123-kW PV field and a 3,456-kW flat-plate collector field.

9.2 RO-MD

9.3 MED-MD?

9.4

10 Results of Sample Cases

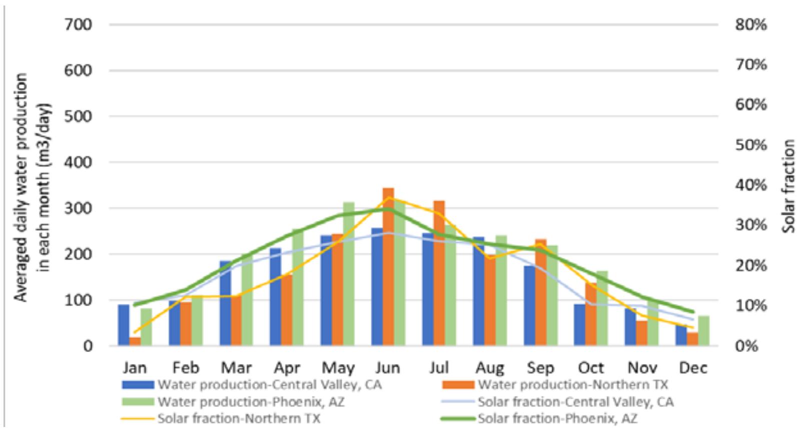
10.1 …

Integration of Industrial Process Heat and LT-MED: Application in TX, AZ and CA

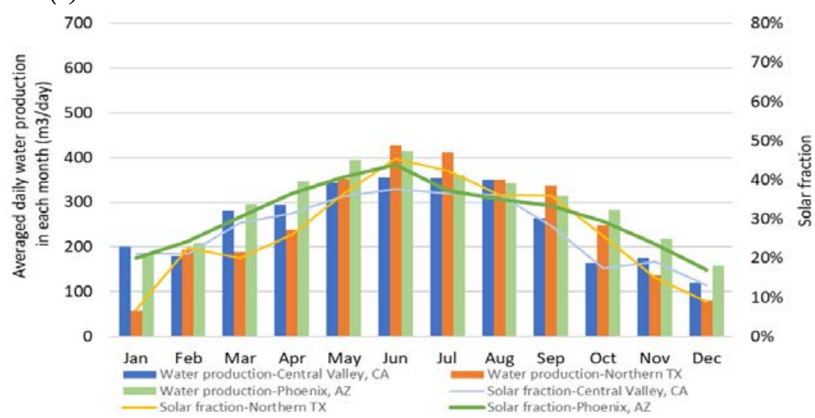
The SAM industrial process heat (IPH) model simulates the thermal energy collected from the solar field that can be used for thermal applications. We simulated a case study of coupling an IPH (Linear Fresnel process heat) system with a LT- MED plant in three locations, namely Central Valley, CA, Phoenix, AZ and near Dallas, TX.

The LT-MED system was specified as 1000 m^3/day capacity with a GOR of 12 for 35 g/L feed water. The nominal thermal power required for this system is 2.24 MWth. The solar field was sized at 2 times of the nameplate power demand of the MED (solar multiple =2). The actual field output is 4.98 MWth with a field area of 1.52 acres. Scenarios with and without thermal storage were simulated.

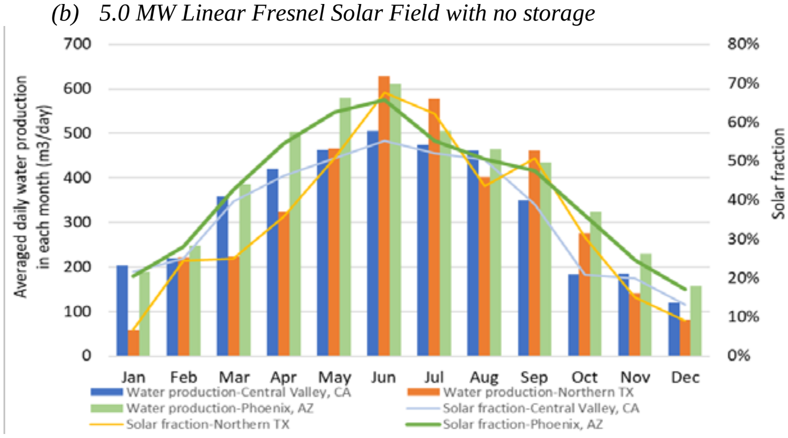
Figure 3 shows the monthly performance of the system in each of the three locations. It is noted that a 6-hr storage system could almost double the annual water production. Its benefit is dependent on the seasons; in the winter (November, December and January) the monthly product volume is increased, with storage, by about 5%, while the increment can reach 48% during each of the summer months. However, even in the best case (June with storage) the solar system alone cannot enable more than 70% of the desalination plant. Our software will allow the user to compliment the solar system with auxiliary power and/or waste or geothermal heat, if available near the selected site.



(a) 2.5 MW Linear Fresnel Solar Field



(b) 5.0 MW Linear Fresnel Solar Field with no storage



(c) 5.0 MW Linear Fresnel Solar Field Coupled with a 6-hr storage system

Figure 3. Averaged daily water production (m^3/day) of a 1,000 m^3/day LT-MED system powered by Linear Fresnel solar field and the corresponding solar fraction (energy contribution from solar field) for locations in Central Valley, CA, Northern TX and Phoenix, AZ

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10.2 Sensitivity/Uncertainty Analysis

10.3 General Desalination Model

10.4 User Water Price Input

10.5 Future Projections

11 Bibliography