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RENEWABLE ENERGY DESALINATION PILOT PLANT: ENERGY, MASS AND VOLUME BALANCES

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Advisor: Prof. Dr. Marcos Rogério Szeliga

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ABSTRACT

Facing the current scenario looking for water scarcity solutions, water desalination is an alternative for this problem, because it can treat water with a high concentration of total dissolved solids (TDS) and making it potable. The present work had as objectives to detail the energy consumption and elaborate the balances of energy, mass and volume of a pilot plant of brackish water desalination with production capacity of 1 $m^3.h^{-1}$. The pilot plant was composed by an ultrafiltration system (UF) and softening as a pre-treatment of the reverse osmosis (RO), and fed with brackish water due to the mixture of sea and river waters. The energy supply by means of integration between the conventional grid, and solar converted in electricity using photovoltaic panels. The system was installed in the dependencies of the water treatment plant of Sanepar, in Praia de Leste, located in the county of Pontal do Parana, Parana State. There were realized 20 operations: the first 10 treated brackish water with initial concentration of TDS of 3,500 ± 100 $g.m^{-3}$ and the last 10 with TDS initial concentration of the brackish water of 7,000 ± 100 $g.m^{-3}$. The total consumption of energy during the 20 operations was 388.61 kWh, the production of solar energy during the study period was 111 kWh and the quantity of energy provided by the conventional grid was 277.61 kWh. The effective consumption of energy for the permeate production was approximately 4.30 $kWh.m^{-3}$. The total consumption of energy composed by the consumption of preparing the brackish water and the UF and RO systems during the first 10 operations was 5.84 $kWh.m^{-3}$ to produce 28.06 m^3 of permeate with TDS concentration of 28 $g.m^{-3}$. In the last 10 operations the total consumption of energy was 5.34 $kWh.m^{-3}$ and produced 26,85 m^3 of permeate with TDS concentration of 67 $g.m^{-3}$. At the monitoring it was observed that the removal of color and turbidity in water occurred in the UF system and the removal of TDS happened in the RO system.

Keywords: Brackish Water, Desalination, Energy Balance, Mass Balance, Volume Balance.

RESUMO

Diante do cenário atual de busca por soluções para a escassez de água, a dessalinização da água é uma alternativa para esse problema, pois pode tratar a água com alta concentração de sólidos dissolvidos totais (SDT) e torná-la potável. O presente trabalho teve como objetivos detalhar o consumo de energia e elaborar os balanços de energia, massa e volume de uma estação piloto de dessalinização de água salobra com capacidade de produção de 1 $m^3.h^{-1}$. A estação piloto foi composta por um sistema de ultrafiltração (UF) e abrandamento como pré-tratamento da osmose reversa (OR), e alimentada com água salobra a partir da mistura de águas do mar e rio. O fornecimento de energia ocorreu por meio da integração entre a rede convencional e a energia solar convertida em eletricidade usando painéis fotovoltaicos. O sistema foi instalado nas dependências da estação de tratamento de água de Sanepar, Praia de Leste, localizada no município de Pontal do Paraná, estado do Paraná. Foram realizadas 20 operações: as 10 primeiras com água salobra com concentração inicial de TDS de 3.500 ± 100 $g.m^{-3}$ e as 10 últimas com concentração inicial de SDT da água salobra de 7.000 ± 100 $g.m^{-3}$. O consumo total de energia durante as 20 operações foi de 388,61 kWh, a produção de energia solar durante o período de estudo foi de 111 kWh e a quantidade de energia fornecida pela rede convencional foi de 277,61 kWh. O consumo efetivo de energia para a produção de permeado foi de aproximadamente 4,30 $kWh.m^{-3}$. O consumo total de energia composto pelo consumo dos sistemas de água salobra e UF e RO durante as 10 primeiras operações foi de 5,84 $kWh.m^{-3}$ para produzir 28,06 m^3 de permeado com concentração de TDS de 28 $g.m^{-3}$. Nas últimas 10 operações o consumo total de energia foi de 5,34 $kWh.m^{-3}$ e produziu 26,85 m^3 de permeado com concentração de TDS de 67 $g.m^{-3}$. No monitoramento observou-se que a remoção de cor e turbidez na água ocorreu no sistema de UF e a remoção do SDT ocorreu no sistema de RO.

Palavras-chave: Água Salobra, Dessalinização, Balanço Energético, Balanço de Massa, Balanço de Volume.

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ACRONYMS LIST

BWD	Brackish water desalination
CD	Capacitive Deionization
CONAMA	Conselho Nacional do Meio Ambiente (National Environment Council)
COPEL	Companhia Paranaense de Energia (Company of Energy Parana)
ED	Electrodialysis
EDR	Electrodialysis reversion
IBGE	Instituto Brasileiro de Geografia e Estatística (Brazilian Institute of Geography and Statistics)
MD	Membrane Distillation
MED	Multiple Effects Destilation
MF	Microfiltration
MH	Mixture and homogenization
MSC	Mechanical Steam Compression
MSD	Multiple Stages Deionization
NF	Nanofiltration
PAC	Polyaluminium chloride
PV	Photovoltaic Panel
RO	Reverse Osmosis
SANEPAR	Companhia de Saneamento do Parana (Sanitation Company from Parana)
TDS	Total Dissolved Solids
TSC	Thermal Steam Compression
UEPG	Universidade Estadual de Ponta Grossa (State University of Ponta Grossa)
UF	Ultrafiltration
UV	Ultraviolet
WTP	Water Treatment Plant

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INTRODUCTION

Water is a natural resource that is essential to life, being destined to human supply, industry, agriculture, energy supply, among others. Besides its extreme importance, the availability of water has been decreasing due to the climate changes and supply sources pollution, that affects considerably the water quality to be treated and later consumed or by the increase of population that as consequence generates a raise of consumption of supply water, in the industrial and agricultural use (EL-BIALY et al., 2016; QIBLAWEY; BANAT, 2008).

Facing this scenario, desalination presents as a solution for the water unavailability, because it can treat and turning the saltwater potable. Desalination plants are installed in more than 150 countries, and 44% of these plants are found in the Middle East and Northern Africa. In Brazil exists the Programa Água Doce (Fresh Water Program), that is a program of the Federal Government in partnership with Federal, State, County and civil society institutions and has the objective the installation of desalination systems, mainly in the Semi-arid region of the country (QTAISHAT; BANAT, 2013; IDA, 2018; IWA, 2018; BRASIL, 2018).

Desalination processes are characterized by the high energy consumption. Thermal processes of distillation demand great availability of heat and membrane separation processes require high pressures (QTAISHAT; BANAT, 2013).

The demand for energy availability to treat water encourages the use for alternative sources, which has objectives: cost reduction of the treatment, reduction of fossil fuels consumption, reduction of air pollution, or substitute the use of conventional energy which is not present in some regions that require the use of desalination (QIBLAWEY; BANAT, 2008).

In this context of search for solutions for the unavailability of drinking water by mean of desalination and reduction of energy costs of it, the *Universidade Estadual de Ponta Grossa* (UEPG) in partnership with University of North Texas, University College London and the *Companhia de Saneamento do Parana* (Sanepar), have been developed studies in a brackish water desalination pilot plant using ultrafiltration followed by softening and reverse osmosis supplied by renewable energy, in the county of Pontal do Parana, balneary of Praia de Leste.

This study is important to improve the quality of life of the local community, as periods of low river flow drought and high tide periods can cause saline intrusions in the springs, hampering the efficiency of conventional treatment. Saltwater intrusions may also occur in groundwater used for well supplies, rendering saltwater unfit for consumption.

Facing the presented facts, the general objective of this work is the monitoring and evaluation of a desalination pilot plant composed by ultrafiltration followed by softening and reverse osmosis, using renewable energy with the elaboration of energy, mass and volume balances of the system.

The present work was divided into three chapters: the first chapter shows a generic Bibliographic Review.

The second chapter had as general objective the detailing of the energy cost of the desalination pilot system with the use of solar energy combined with energy provided by the conventional grid and their respective energy balance. The specific objectives were:

- Monitoring the production and energy consumption;
- Establishing energy consumption for each subsystem and for the group, identifying effective and indirect incidences in the final generated product;
- Obtaining the final energy consumption of production and the relation between conventional and alternative energy.

The third chapter had as general objective the detailing of the mass and volume balances of the desalination pilot system with different initial concentrations of total dissolved solids (TDS). The specific objectives were:

- Monitoring the volume and mass in the subsystems of the plant by measuring the volume and analyzing TDS concentration;
- Application of continuity model to identify the mass and volume distribution between the subsystems based in measured values from the monitoring, establishing the mass and volume balances.
- Obtaining ratios among processed, rejected and recovered volumes and the definition of systems efficiency.

1 BIBLIOGRAPHIC REVIEW

According to Resolution 357 of the National Environment Council (CONAMA), the water is classified as freshwater when it has salinity equal or inferior to 0,5 ‰, brackish when it has salinity between 0,5 ‰ and 30 ‰ and salt when the salinity is above 30 ‰. According to Drinking Water Regulation 2914 of the Ministry of Health, potable water for human consumption must have the maximum TDS concentration of 1000 $mg.L^{-1}$. These solids are composed by inorganic salts as calcium, magnesium, potassium, sodium, bicarbonates, chlorides, sulfates and small amounts of organic matter dissolved in water, that may cause hardness, salty and bitter flavors and increase the corrosivity level (BRASIL. Ministério do Meio Ambiente, 2018; BRASIL. Ministério da Saúde, 2018; WATER RESEARCH CENTER, 2018).

While the conventional treatment technologies are proper to treat water with dissolved solids concentration below 3000 $mg.L^{-1}$, the desalination is used in the TDS removal of the brackish or salt water, resulting in the outputs of the permeate, is the treated water, and the brine, is the reject with high saline concentration. Reverse osmosis (RO) is an example that may treat brackish water with TDS concentration up to 45000 $mg.L^{-1}$ and provide treated water with TDS concentration inferior to 500 $mg.L^{-1}$ (WADE, 2001; QIBLAWEY; BANAT, 2008; FRITZMANN et al., 2007).

Desalination is used mainly to treat sea water (67%), brackish water (19%), river water (8%) and residual water (6%) (AL-KARAGHOULI; KAZMERSKI, 2011).

According to the International Desalination Organization (2017), the accumulated hired capacity of worldly desalination is of 99.8 million of $m^3.day^{-1}$, while the operating installation capacity is of 92.5 million of $m^3.day^{-1}$ (ZARZO; PRATS, 2018).

1.1 WATER DESALINATION PROCESSES

The desalination processes currently available fit in the categories in which there is or there is not phase change – which are thermal, membrane and hybrid processes. In the thermal process, occur changes in the phase of water. In the membrane process, a phase change does not happen, but a separation of water from the dissolved salts. The hybrid process involves both processes previously cited, membrane process and phase change processes, such is the case of distillation by membrane. (QTAISHAT; BANAT, 2013).

1.1.1 Thermal processes

In the thermal processes, the separation of salt in water is realized by evaporation and condensation, such is the case of the multiple effect distillation (MED), multiple stage distillation (MSD), evaporation by compression and evaporation lagoons, mechanic steam compression (MSC) thermal steam compression (TSC). These processes are characterized for demanding high quantity of energy to evaporate the water (FRITZMANN et al., 2007; QTAISHAT; BANAT, 2013; AL-KARAGHOULI; KAZMERSKI, 2013).

If compared to the membrane processes of distillation, that require high quantity of energy for example, the MSD that consumes 27,25 kWh.m^{-3} and the MED from 14.45 to 21.35 kWh.m^{-3} of water, while the RO consumes about 4 to 6 kWh.m^{-3} to treat salt water and between 1.5 to 2.5 kWh.m^{-3} to treat brackish water (ABDELKAREEM et al., 2017).

1.1.2 Membrane processes

The processes which use membranes under pressure in water treatment are known as microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), electrodialysis (ED), electrodialysis reverse (EDR), capacitive deionization (CD), membrane distillation (MD) and RO. MF and UF can remove final colloidal particles, bacteria, virus and bigger particles, such as proteins. NF and RO are used to remove smaller particles, like dissolved salts (FRITZMANN et al., 2007; ABDELKAREEM et al., 2017).

Among the technologies that use membrane processes – except emerging technologies, like evaporation by membrane and pervaporation – there are some that use electricity to pressurize them, like NF, UF and RO, and those that use electricity by means of an electro-separation process, in which the difference of electric potential is used as driving force to move the ions through the membranes, that is the ED case (MROILLO et al., 2014; ZARZO; PRATS, 2018).

According to the United States Environmental Protection Agency, the RO is the best technology capable to remove compounds that are harmful to health such as arsenic, barium and nitrate (FRITZMANN et al., 2007).

In RO, the membrane is fed with high pressure by the saltwater so it can be superior to the osmotic pressure of the water supply, and then, the water with low concentration of salt is collected on the permeate side and the brine is rejected. The pressure for seawater application is around 60 to 80 bar (AL-KARAGHOULI; RENNE; KAZMERSKI, 2010).

The feed water of low quality can cause damage to the membranes applied in the desalination process, due to scaling and fouling. The first one occurs when there is deposition of calcium carbonate and magnesium hydroxide in the membranes that are found in seawater, the second one happens when there are biological, suspended or colloidal rejects deposition in the membrane (EL-DAHSHAN, 2001; POTTS; AHLERT; WANG, 1981; RICHARDS; SCHÄFER, 2002).

UF membrane can be used as pre-treatment for the RO, because it acts as pre-filter, being able to remove virus and bacteria – being enough for waters that have low salinity. In the desalination system, UF and RO set offers a double barrier against microorganisms and allows the cleaning solicitation reduction of the RO membranes, due to the water quality that leaves from the UF (SCHA, 2007).

According to the International Desalination Association, in 2015 with a total of 86,5 millions of m³ of treated water produced per day by desalination in the world, the most used method was the RO with 65%, followed by MSD with 21%, MED with 7%, ED with 3% and other methods with 4% (ABDELKAREEM et al., 2017).

The main method of desalination adopted in Brazil is by the RO membranes. In Fernando de Noronha archipelago, Pernambuco, the installed system can treat saline water and produce 54 $m^3.h^{-1}$. By mean of the *Programa Água Doce* communities of low income from the Brazilian Semi-arid are supplied with potable water obtained from saline or brackish water from wells. Another desalination example used in Brazil is the installation of these systems in ships, submarines and oil extraction platforms that obtain potable water from seawater (ROSAS, 2013).

1.2 DESALINATION AND ENERGY

1.2.1 Energy consumption in desalination process

The energy used in desalination can be electric or thermal. The first one can originate from fossil fuels as coal, oil and gas, from nuclear sources and renewable energy. The second one can be provided by fossil fuels that feed the boilers, electric heating power plants, industrial residue sources and renewable energy (AL-KARAGHOULI; KAZMERSKI, 2013).

The energy consumption in the process is intense. The demand of energy depends on the adopted technology, the salinity and temperature of the feed water and the treated water quality. According to Acumed, a Spanish public energy company responsible for about 60% of expenses with treated water production (GHENAI et al., 2018;ZARZO; PRATS, 2018).

Among thermal processes, MSD consumes between 19.58 and 27.25 $kWh.m^{-3}$, with the MED between 14.45 and 21.35 $kWh.m^{-3}$, the MSC from 7 to 12 $kWh.m^{-3}$ and TSC consumes approximately 16 $kWh.m^{-3}$. These processes can provide treated water with about 10 $mg.L^{-1}$ of salinity concentration (AL-KARAGHOULI; KAZMERSKI, 2013).

Among the processes that use membranes, the RO consumes between 4 to 6 $kWh.m^{-3}$ to treat saline water and provide water with salinity concentration between 400 and 500 $mg.L^{-1}$. To treat brackish water, RO uses between 1.5 to 2.5 $kWh.m^{-3}$ and can produce water with salinity between 200 and 500 $mg.L^{-1}$. The ED requires from 0.7 to 2.5 $kWh.m^{-3}$ to treat water that possesses salinity smaller than 2500 $mg.L^{-1}$ and between 2,64 to 5,5 $kWh.m^{-3}$ to treat water with salinity concentration between 2500 to 5000 $mg.L^{-1}$. This process generates water with salinity concentration between 150 to 500 $mg.L^{-1}$ (AL-KARAGHOULI; KAZMERSKI, 2013).

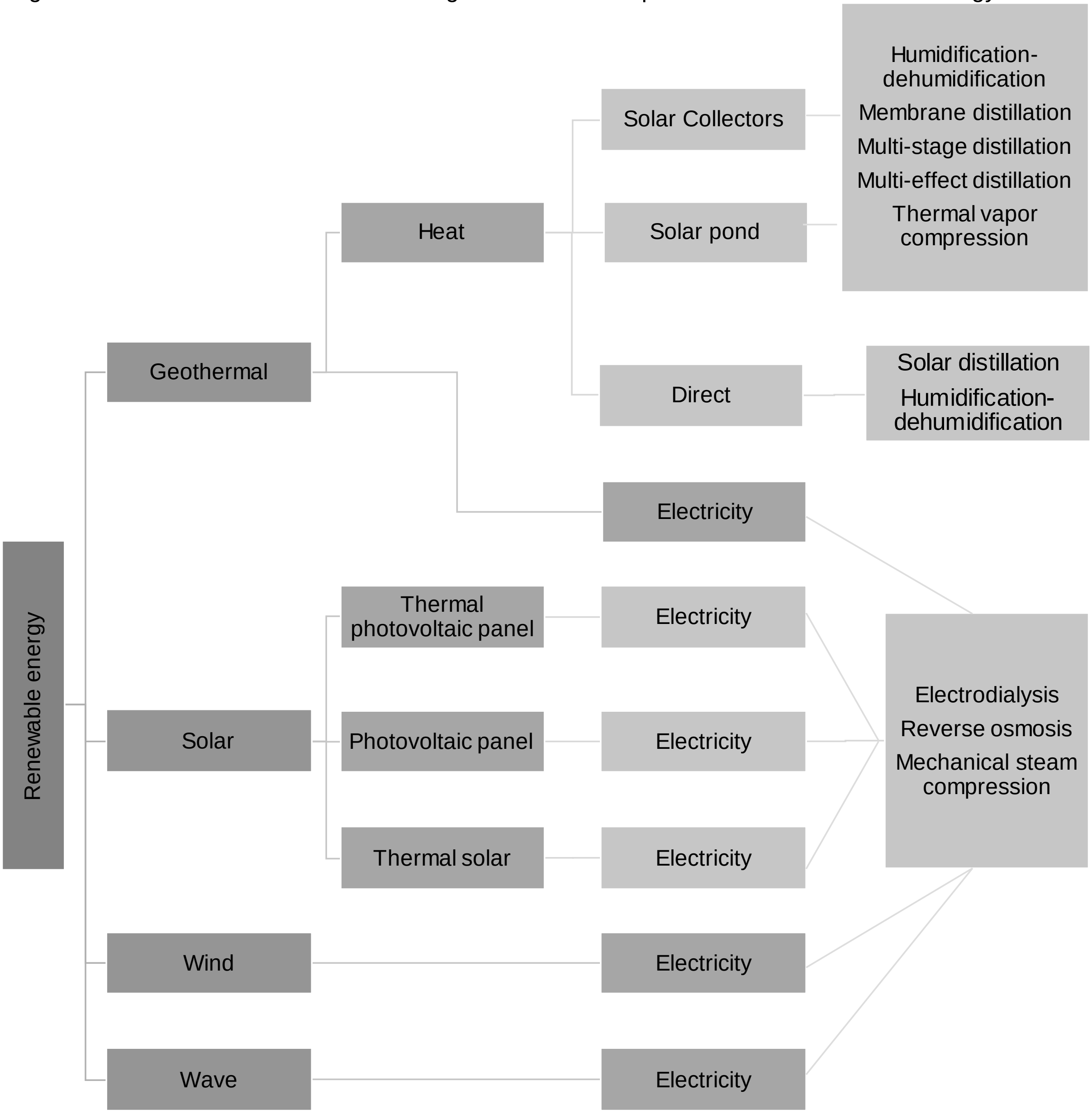
1.2.2 Energy production for the desalination process

The search for renewable energy to conduct the desalination plants has as objective the cost reduction, the use reduction of fossil fuels and allow that areas away from the conventional electric network can also have access to desalination (QIBLAWEY; BANAT, 2008).

The research for renewable sources can be justified, because the main expenses with desalination are due to feed water salinity, energy consumed in the process and the size of the treatment plant (WINTER, 2002).

The Figure 1.1 shows possible combinations that can be done between the renewable energy sources and the desalination processes (MATHIOULAKIS; BELESSIOTIS; DELYANNIS, 2007).

Figure 1.1 - Possible combinations among the desalination processes and renewable energy sources



Source: Adapted from MATHIOULAKIS; BELESSIOTIS; DELYANNIS, 2007

Among the available renewable energy sources, the solar is the most abundant, possessing the higher potential for use in desalination (BLANCO et al., 2009).

The solar desalination plants are described as of free energy and with smaller operation cost, they are indicated for arid areas, that have great availability of insolation and where conventional electric energy is scarce (NAIMA; ABD; KAWIB, 2002).

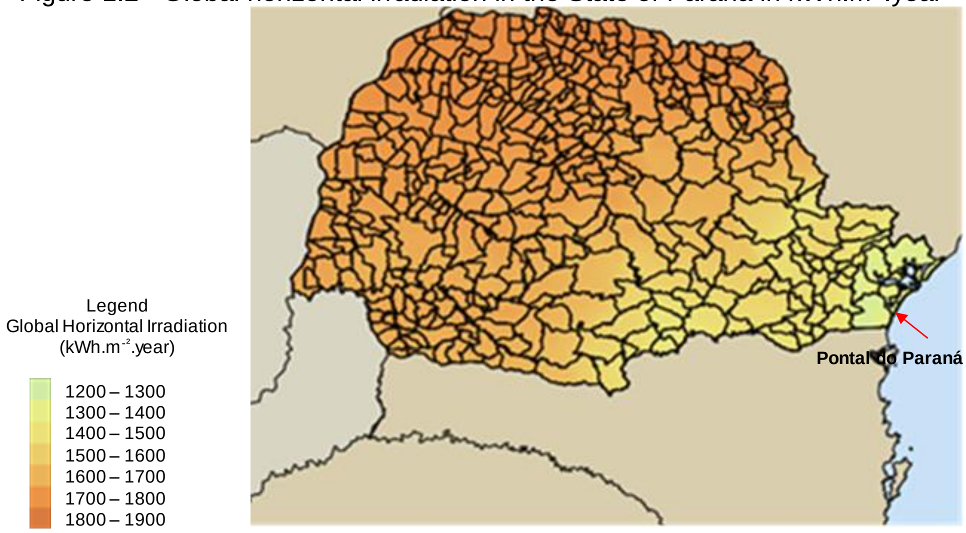
The solar energy used in desalination can be used as direct or indirect. Systems that use direct solar energy are those that produce the distilled water in the solar collector itself. And the systems that use indirect solar energy are those that generate electricity or heat, which is the case of the membrane processes (GARCIA-RODRIGUEZ; GMEZ-CAMACHO, 2001).

1.2.3 Energy and solar radiation

The solar radiation that affects Earth's surface in W.m^{-2}, is formed by direct and diffuse components. The direct irradiation is a parcel of the solar radiation that passes through Earth's atmosphere and do not suffer any changes in its original direction and is composed by normal direct irradiation and horizontal direct. The first one is the energy rate by area unity that comes straight from the Sun and that affects in a perpendicular to the Earth's surface form, and the second one is the energy rate by area unity of the direct solar beam in a horizontal surface. Horizontal solar irradiation is a portion of solar radiation that reaches a horizontal surface per area and is responsible for changing the occurrence and mirroring processes that occur in the atmosphere through the presence of particulate gases in the atmosphere. The global horizontal irradiance is the total rate of energy, per area unity, that affects in a horizontal surface, and the irradiance on the inclined plane is the energy rate per area unity that affects over an inclined plane. According to Solar Atlas from Parana, from 2017, the Northeast region of Brazil presents an inter annual average variability of solar irradiation between 5.39 and 5.59 kWh.m^{-2}, in the Southern region this average variability is between 4.53 and 4.61 kWh.m^{-2}, in the Northern region between 4.61 and 4.69 kWh.m^{-2}, the Southeast region has annual average variety of radiation between 4.97 and 5.11 kWh.m^{-2}, and in the Midwest this value is between 5 and 5.2 kWh.m^{-2} (PEREIRA et al., 2017).

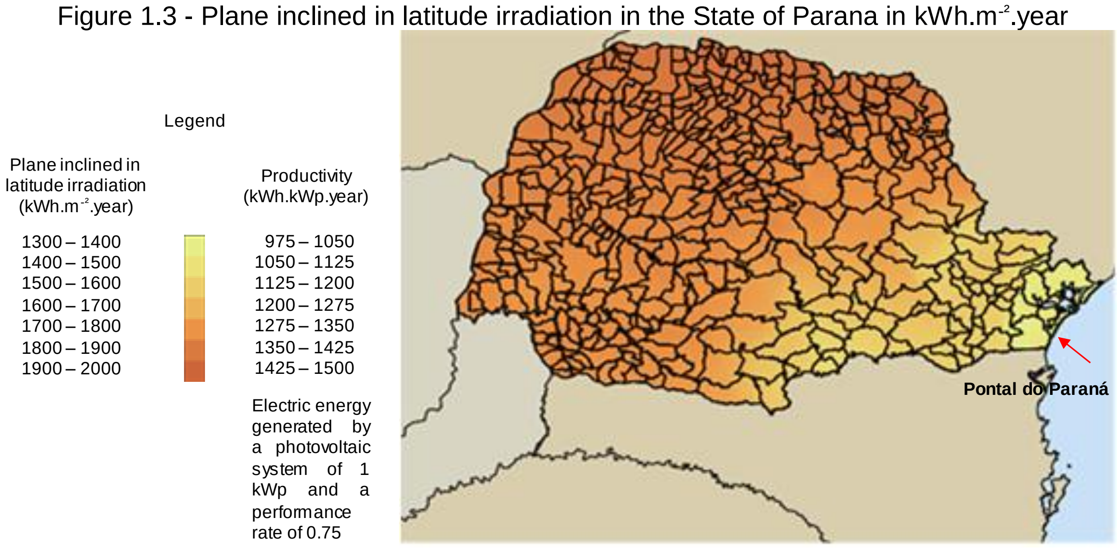
The Figure 1.2 shows the global horizontal irradiation, in Figure 1.3 the inclined plane of latitude, the normal direct in Figure 1.4 and the diffuse in Figure 1.5, all in kWh.m^{-2}.year, and all of them also show that the irradiation is higher depending on how western in the State is located the county (ATLAS DE ENERGIA SOLAR DO ESTADO DO PARANA, 2018).

Figure 1.2 - Global horizontal irradiation in the State of Parana in $kWh.m^{-2}.year$



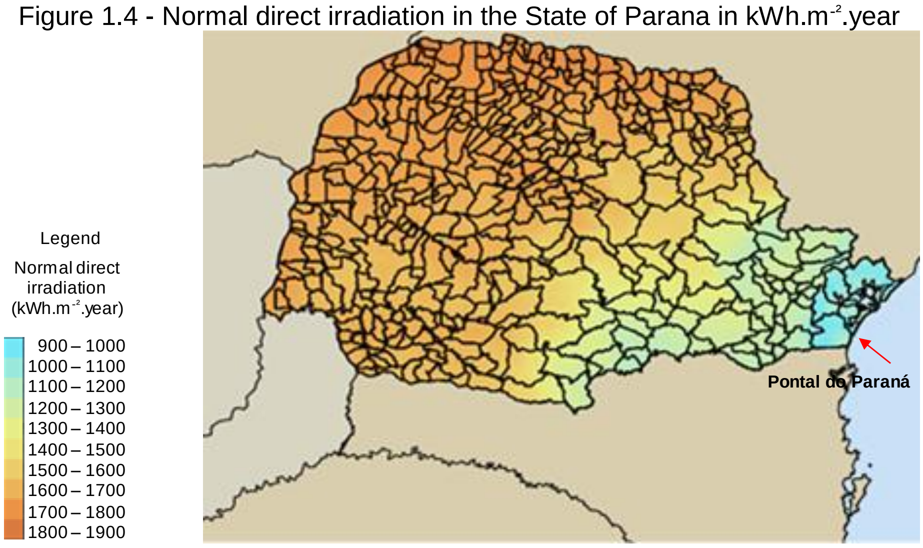
Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

Figure 1.3 - Plane inclined in latitude irradiation in the State of Parana in $kWh.m^{-2}.year$



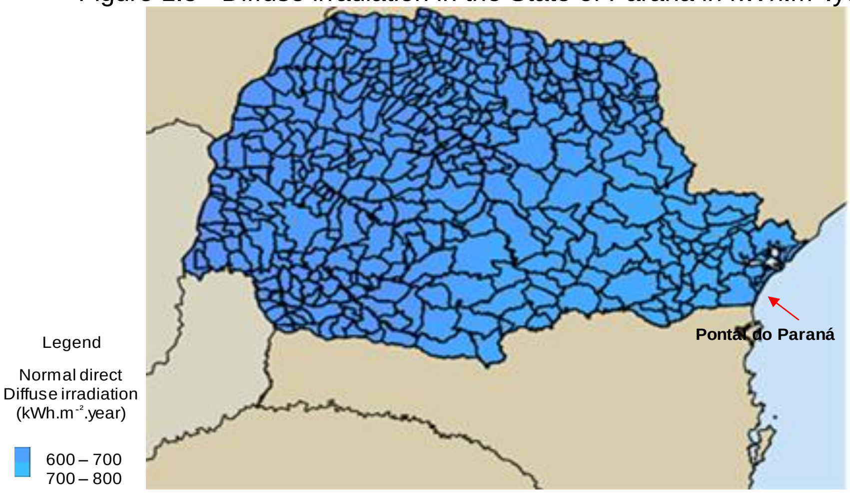
Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

Figure 1.4 - Normal direct irradiation in the State of Parana in $kWh.m^{-2}.year$



Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

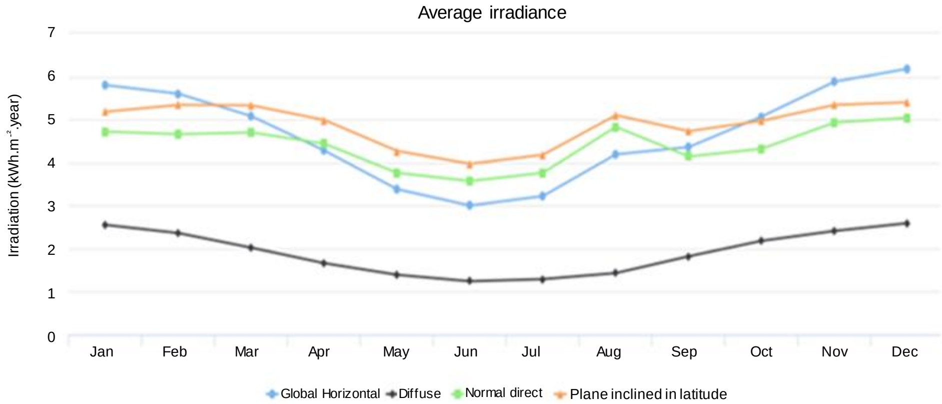
Figure 1.5 - Diffuse irradiation in the State of Parana in kWh.m^{-2}.year



Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

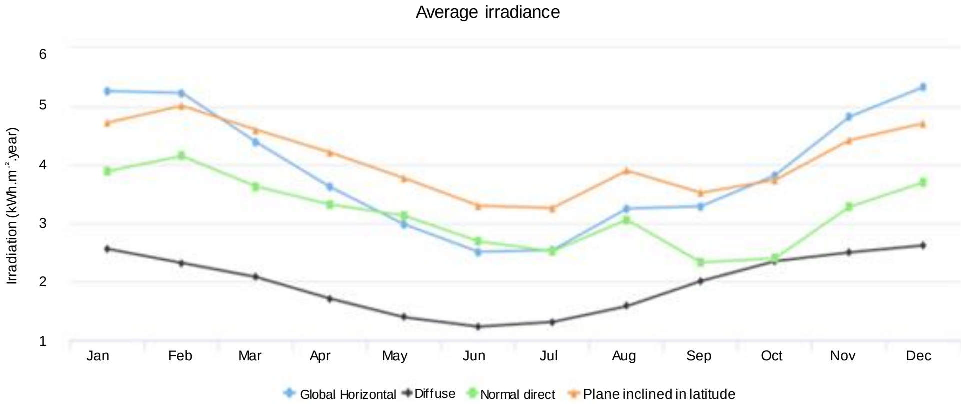
Figure 1.6 shows the daily average irradiations for the State. For the county of Pontal do Parana the daily average is illustrated in Figure 1.7.

Figure 1.6 - Average global horizontal diffused irradiation, normal direct and inclined plane in latitude in the State of Parana in kWh.m^{-2}.day



Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

Figure 1.7 - Average global horizontal irradiation, diffused, normal direct and inclined plane in latitude in the county of Pontal do Parana in kWh.m^{-2}.day



Source: Adapted from Atlas de Energia Solar do Estado do Paraná, 2018

The daily global horizontal irradiation in Parana is 6.2 KWh.m^{-2}, and occurs in December, and the minimum of 3.0 kWh.m^{-2}, in June. In the county of Pontal do Parana, this irradiation is 5.3 KWh.m^{-2}, in December, and the minimum of 2.5 kWh.m^{-2}, in June and July. The daily minimum diffused from the State is 1.2 kWh.m^{-2}, in June, and the maximum in December and January. In Pontal do Parana is equal to 1.2 kWh.m^{-2}, in June and the maximum of 2.6 kWh.m^{-2}, in December and January. The daily minimum normal direct of the State is 3.6 kWh.m^{-2}, in June, and the maximum is 5.0 in December kWh.m^{-2}. In Pontal do Parana is 2.3 kWh.m^{-2}, in the month of June, the maximum is 4.2 kWh.m^{-2} in February. Lastly the plane inclined irradiation the minimum daily latitude in the State is 4.0 kWh.m^{-2}, in June and the maximum is 5.4 kWh.m^{-2}, in December, and in Pontal do Parana the daily minimum is 3.3 kWh.m^{-2}, in June and July the maximum is 5.0 kWh.m^{-2}, in February (ATLAS DE ENERGIA SOLAR DO ESTADO DO PARANA, 2018).

1.2.4 Photovoltaic systems

One of the possible combinations between desalination systems and renewable energy is the RO driven by electricity created by solar energy, captured from photovoltaic panels (MATHIOULAKIS; BELESSIOTIS; DELYANNIS, 2007).

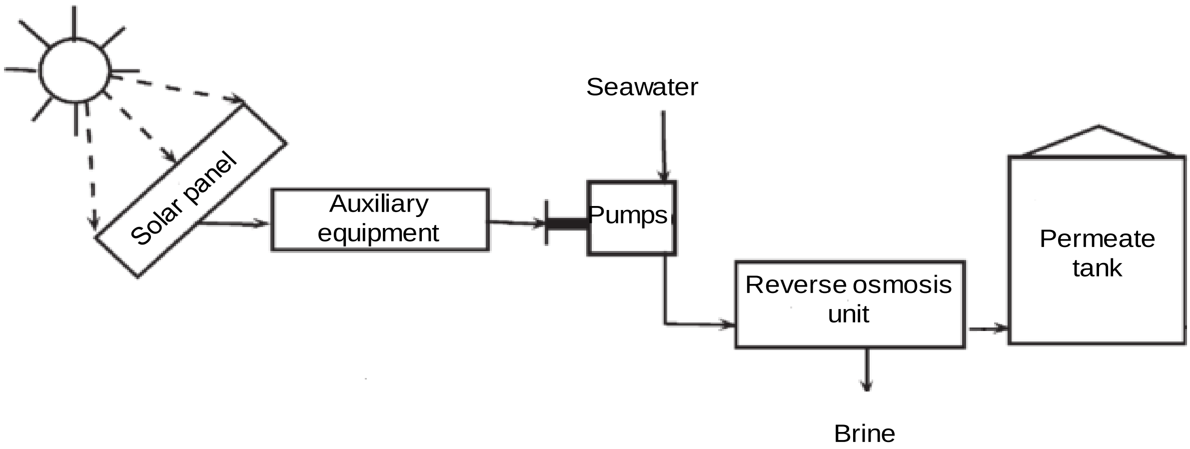
In a RO system driven by photovoltaic cells, both the systems work in an independent form, in an economic and trustworthy way. The main characteristics are to consume few energy, does not require great areas to be built, the system can be expanded if necessary, easy to operate, doesn’t generate air or sound pollution, requires low maintenance and keeps the efficiency during his lifespan, besides that the accelerated development of these technologies in a combined way transformed solar energy into the most recommended to feed the desalination systems (LI et al., 2012; AL SULEIMANI; NAIR, 2000;AHMAD et al., 2015).

A factor that must be observed is about the installation of photovoltaic panels in desalination plants in remote areas is the need of a reserve system that is used when solar light is unavailable, for example, at night. This question is solved using batteries, a hybrid diesel-solar energy system, or constructing a plant with treatment capacity superior to the daily requirement, so, treated water can be stored and used later (HELAL; AL-MALEK; AL-KATHEERI, 2008).

The photovoltaic cell is semi-conductive device that converts sunlight into electric energy and usually is made of silicon (monocrystalline, polycrystalline or amorphous). Many photovoltaic cells can be arranged in a way to form a photovoltaic module, this can be arranged as many modules and form a photovoltaic matrix, according to the energy that the desalination system uses. Depending on the desalination plant, the installation of supply equipments might be necessary, such as charge controller and inverters, and the energy storage, as batteries. The charge controller is used to protect the batteries from overcharge, the inverters convert the direct current that is produced by the photovoltaic cells in alternating current. The lifespan of photovoltaic systems is from 20 to 30 years (AL-KARAGHOULI; KAZMERSKI, 2013).

Figure 1.8 presents a schematic desalination plant that uses reverse osmosis driven by solar energy without batteries.

Figure 1.8 - Schematic of desalination plant using reverse osmosis powered by solar energy



Source: Adapted from KUMARASAMY; NARASIMHAN; NARASIMHAN, 2015

The main energy expenses in desalination plants that use RO are the high pressure systems, the feed water collection pumps, the pre-treatment stage, the membrane stage, the high pressure pumps, the energy recovery devices, the post treatment and the auxiliary equipments, such as compressors, from the CIP – clean in place systems, the instruments, the illumination, among others. In the collecting state, the energy cost can be optimized by choosing the pump and reducing pressure drops. The pre-treatment can work by gravity or pressurized in different stages. The membrane stage is the main stage in desalination systems that use RO and is also the main stage where the energy decrease can occur in the process, decreasing the membrane pressure, but without decreasing the quality of the water. High-pressure pumps are the ones that generate the most energy consumption in desalination, and their consumption can be reduced adopting more efficient pumps or fitting pressure for the original saline concentration and treatment level (ZARZO; PRATS, 2018).

1.3 WATER QUALITY INDICATORS

Water quality indicators are used to describe their physical, chemical and biological characteristics (SPERLING, 2017).

The monitoring of indicators is made to follow the performance of the water treatment, that must fit established patterns and avoid desalination membrane damage, being it by their deposition by calcium carbonate and magnesium hydroxide or incrustation of biological suspended or colloidal rejects (PRIHASTO; LIU; KIM, 2009).

A form to identify the presence of suspended solids, from mineral origin, such as clay, silt, sand, or organic as vegetable and animal microorganisms, is by means of water turbidity analysis. It is obtained by means of inference of suspended particle concentration in the water obtained from the passage of a light beam in the sample (SPERLING, 2017; LIBÂNIO, 2005).

The presence of dissolved solids, normally from organic origin and iron and manganese is detected by mean of analysis of the water coloration. The organic matter is also responsible for adding taste to the water (SPERLING, 2017).

TDS concentration monitoring is one of the oldest methods to squire water quality. The simplicity and low analysis cost helped to spread its use, as an example the wide employment following desalination (KABSCH-KROBUTOWICZ, M. et al, 2011; GUSTAFSON; BEHRMAN, 1939; LINDSEY; RUPERT, 2012; GILMROE; LUONG, 2016).

Another index to be evaluated for water quality control and desalination process is hardness. It indicates the presence of calcium and magnesium dissolved compounds, and in smaller quantity aluminum, iron, manganese and strontium. The presence of these compounds can implicate in a reduction in the foam formation of cleaning products and incrustations in hot water pipes, due to precipitation caused by high temperatures. It may also cause incrustations in desalination membranes (LIBÂNIO, 2005; RAHARDIANTO et al., 2007).

The incrustations must be avoided because besides spoiling the water treatment process, they must raise the operational costs, because they increase the energy demand and the membrane washing processes, besides reducing their lifespan (PONTIÉ et al., 2005).

According to the Drinking Water Regulation 2914 from the Ministry of Health, potable water must have turbidity inferior to 5 NTU, apparent color inferior to 15 uH, total hardness inferior to 500 $mg.L^{-1}$ and TDS inferior to 1000 $mg.L^{-1}$ (BRASIL, 2018).

Some parameters of the desalination system must be followed to be available to observe its efficiency with the energy costs. For example, about desalination system, the total dissolved solids rates, the permeate and brine production, pressure, temperature, among others. About the energy system some parameters to be followed are insolation, the daily sun-time duration and active days for example. (EL-SAYED, 2007).

1.4 ULTRAFILTRATION AND SOFTENING AS REVERSE OSMOSIS PRE-TREATMENT

High quality feeding water is essential for the good RO performance, in contrary, the membranes are going to present short operation periods, increase in maintenance and reduction in their lifespan. UF is used as RO pre-treatment, because it has the capacity to remove final particles of sizes between 0.01 to 0.02 micrometer. Between these particles that must be removed are suspended solids, dissolved solids, organic contaminants, immiscible liquids and moderately soluble solids. UF use before RO allows the incrustation reduction in membranes and the use of chemical products and optimizes the process operation time (PRIHASTO; LIU; KIM, 2009; PEARCE, 2007).

The softening is indicated to remove hardness from the water. One of the used methods consists in ion exchange while the water passes through the filter – a cationic resin. Removal of hardness occurs when the hardness ions, mainly calcium and magnesium, are exchanged for sodium ions. When saturation of the cationic resin occurs, there should be a regeneration of the softener, using a saturated solution of sodium chloride. In this process, calcium and magnesium ions are reversed in sodium chloride (PERMUTION, 2016; Water Quality Research Council, 1990 quoted by BRASTAD; HE, 2013).

In relation to the RO, the presence of calcium and magnesium ions may cause incrustation on the membranes, reducing the recovery rate of the operation, what increases the generated brine volume and consequently less permeate will be produced (RAHARDIANTO et al., 2007).

1.5 POST TREATMENT OF REVERSE OSMOSIS

Desalination processes generate a slightly acid permeate, demineralized, with low tampon capacity and low concentration of calcium and magnesium. Water that presents these characteristics cannot be distributed for consumption or irrigation, in contrary, it may cause damage in the water distribution system, in the installations of the final consumers, for the treatment of downstream wastewater and for consumers reusing downstream waste in agriculture (BIRNHACK; VOUTCHKOV; LAHAV, 2011; DELION; MAUGUIN; CROSIN, 2004).

The permeate needs post-treatment, that has as objective the water potabilization, remineralization and disinfection (KHAWAJI; WIE, 1994).

In the disinfection stage, chloramines, ozone, chlorine dioxide and ultraviolet light (UV) can be used to inactive any residual pathogenic microorganism. The disinfection with UV is viable because it does not use heat neither chemical products (Montgomery Watson Harza, 2005 quoted by KIM; AMY; KARANFIL, 2015; GANDHIBDAN; AL-MOJEL, 2009).

1.6 ENERGY, MASS AND VOLUME BALANCES IN DESALINATION

In what relates to energy, exists the law of self-conservation of energy, that affirms that energy of bodies or particles in a closed system remain constant (BRITANNICA ACAMEDIC, 2018).

According to this law, the energy provided to the desalination system must be the same consumed during the process, both in direct form, related to the treatment process, and in indirect form, related to indirect consumption.

The principle of mass conservation estimates that it remains constant in a system. The continuity equation represents mathematically a balance of mass for a volume control, and it says that the liquid mass stream that crosses a control surface summed to the mass variation rate inside the volume control must be equal to zero (LIVI, 2004).

Following this principle, in the volumetric balance of desalination the feed water that supplies the system must coincide with the volume treated in the diverse stages of the operation. About the mass balance, it demonstrates the operational performance by monitoring the compounds removal present in the water in the diverse stages of the treatment.

1.7 DESCRIPTION OF THE BRACKISH WATER DESALINATION PILOT PLANT

The county of Pontal do Parana, Parana State, is limited by the Atlantic Ocean to the North and East, West with Paranaguá county and south Matinhos county, as shown in Figure 1.9, being part of the Paranagua Metropolitan region, at approximately 100km of the Parana State capital, Curitiba.

Figure 1.9 - Location of Pontal do Parana

Source: Google Earth (2018)

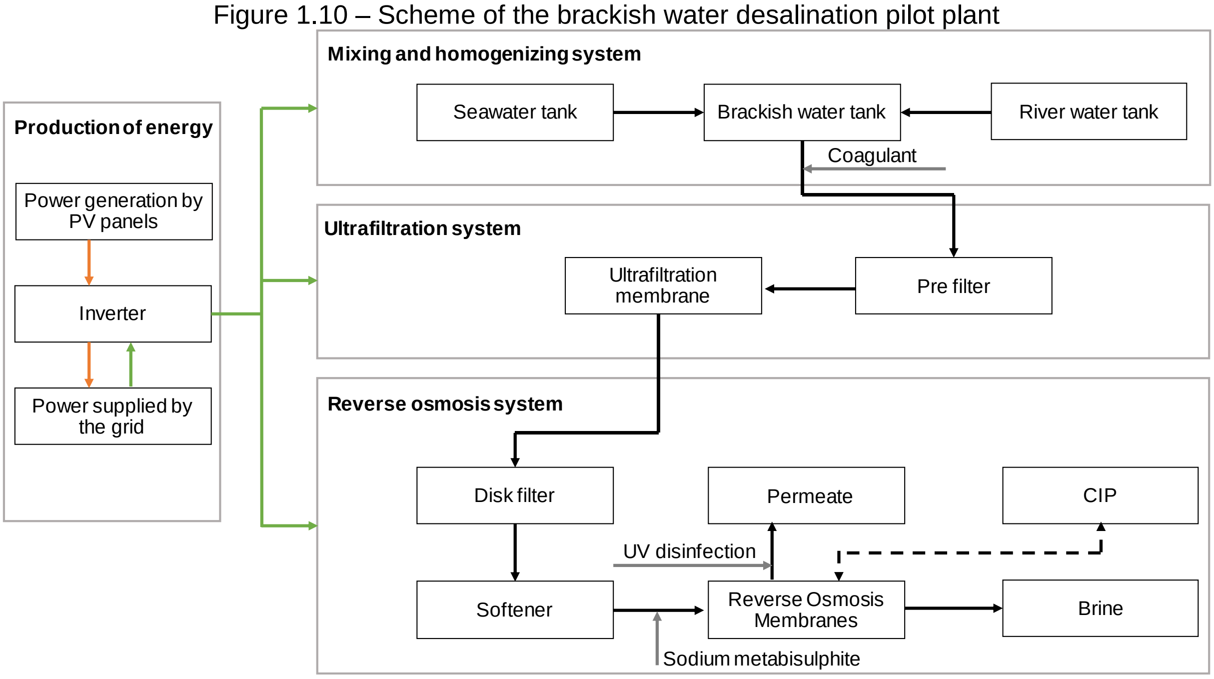
The county fits in the coastal plain of Praia de Leste, it possesses a soft beard and low altitude, generically called *restinga*. The summer is the rainy plant, with about 350 mm of rain, the winter is the driest season, with 100 mm of precipitation, and the annual average of precipitation is about 2000 mm (PREFEITURA MUNICIPAL DE PONTAL DO PARANA, 2015).

The water that supplies the county is collected in the Pombas River, it is treated by conventional way by Sanepar, with coagulation, flocculation, decantation, filtration, disinfection and pH adjustment. The river is about 10 km away from the water treatment plant (WTP) (ALMEIDA, 2017).

The installation of the desalination pilot plant was annex to the conventional WTP of Sanepar, in Praia de Leste, one of the 48 beaches of Pontal do Parana (PREFEITURA MUNICIPAL DE PONTAL DO PARANA, 2015).

Figure 1.10 shows the scheme of the brackish water desalination pilot plant (BWD) used in the study and the stages of the energy production process.

Figure 1.10 – Scheme of the brackish water desalination pilot plant



The energy supply was composed by 8 photovoltaic panels modules, the inverter and the connection to the Energy Company from Parana (*Companhia Paranaense de Energia*), Copel.

The water treatment was formed by the mixture and homogenization of brackish water tank, filter, UF membrane, pre-filter, softener, RO membranes and permeate and brine tanks.

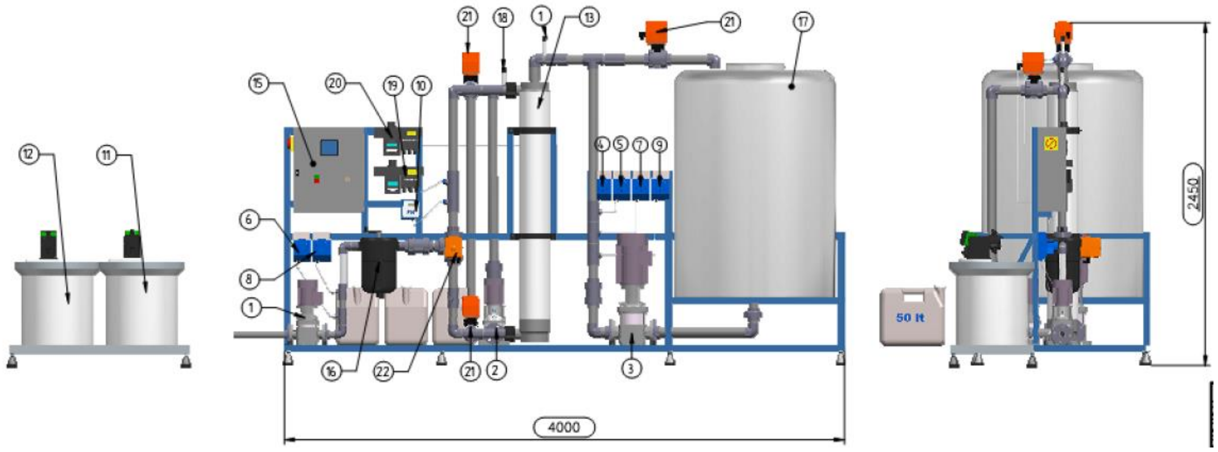
Figure 1.11 illustrates the brackish water mixture tanks of the system. The left tank stored the water from the Pombas River, collected in the WTP. The right tank stored sea water, that was collected in Marine Aquaculture and Restocking Center, from Federal University of Parana and transported by truck to the pilot WTP. In the central tank the mixture and homogenization (MH) was made from the water of the other two tanks.

Figure 1.11 - Mixture water tanks

Figure 1.12 shows the installation of the UF system and Figure 1.13 shows the details of the same system.

Figure 1.12 - UF system

Figure 1.13 - UF system details



At where :

1 Capture pump	2 Circulation pump	3 Backwash pump
4 Chlorine residual metering pump	5 Chlorine dosing pump - backwash	6 Hydrochloric acid metering pump - inlet
7 Hydrochloric acid metering pump - backwashing	8 Sodium hydroxide metering pump - inlet	9 Sodium hydroxide dosing pump - backwashing
10 PH Controller	11 Binder dosing machine	12 Oxidizer dosing machine
13 Aquaflex UF membrane 64 m²	14 Static mixer	15 Control panel
16 Pre filter 300 mc	17 1500 L permeate tank	18 Pressure Transmitter
19 Turbidimeter 0-1000 NTU	20 Turbidimeter 0-100 NTU	21 2 way valve – 2”
22 3/2-way valve– 2”		

Source: Manual de instruções do sistema piloto de ultrafiltração 1 $m^{-3}.h^{-1}$ - H_2LIFE BRASIL, 2015.

The coagulant polyaluminium chloride (PAC) was added after leaving the brackish water mixture tank,. Later the water went to the pore filter of 300 µm that was previously installed in the UF membrane with the function of preserving its performance. After the filter, the water went to the UF membrane. This system produced by the company H_2Life Brazil, presents capacity permeate production of 2.7 $m^3.h^{-1}$ and recovery rate between 90 to 98%. The membrane model X-Flow Aquaflex 55 produced by Pentair has pores of 20 nm. This system worked as pre-treatment to the RO (PENTAIR, 2018).

Feed water for the UF system must had pH between 2 to 12. The UF system can remove 99.9999% of bacteria, 99.99% of viruses and producing permeate with turbidity inferior to 0.1 NTU (PENTAIR, 2018).

UF permeate was stored in a 1500 L tank, which feed the RO system and the reject generated during the ultrafiltration operation was sent to the Sanepar sludge treatment tank.

The nominal flow rate of the RO system was 1 $m^3.h^{-1}$ and the and the pH of the feed water should be between 4 and 10. The system was able of treating water with TDS concentration up to 10000 $g.m^{-3}$ and removing until 99.5% of the salts. The recovery rate of the system was 70% when treated water with a concentration 3,500 $g.m^{-3}$ and 50% when treated water with concentration of 7,000 $g.m^{-3}$ (PERMUTION, 2016).

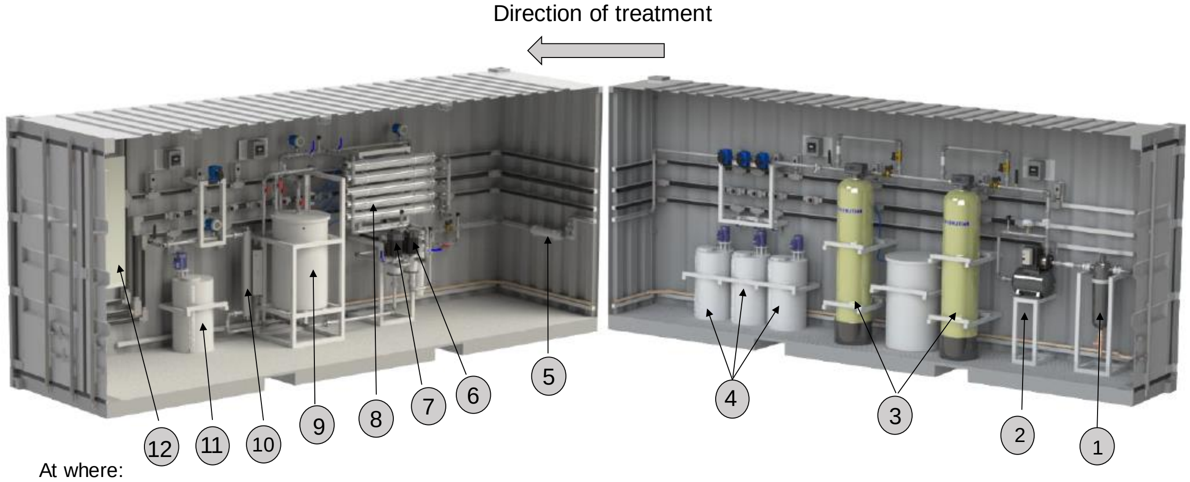
UF permeate passed through the polypropylene disk filter, to retain particles larger than 130 μm, and to the softener, whose objective was removing hardness by means of ions exchange of calcium and magnesium present in the water by sodium ions.

Sodium metabisulfite was added, which had the function of oxidizing free chlorine present in the feed water of the system, to avoid harm to the RO membranes. Then, the water was pumped to the 5 RO membranes, produced by the company Vontron Technology Co. Ltd., and have pores of 1 ηm.

The RO permeate was routed to the permeate water tank. Water disinfection could be made by addition chlorine or by UV, however, the second option was chosen. Water treated and disinfected went to a tank with 5000 L capacity and the brine to another 5000L tank. RO permeate was also used in clean in place, that consisted in the backwash of the membranes from this system. Figures 1.14 and 1.15 illustrate the RO system.

Figure 1.14 - RO unity

Figure 1.15 - Schematic cut from the RO system



At where:

1 Disc filter 130 µm	2 Pressurizing pump	3 Softeners
4 Dosing pumps and chemical tanks	5 Static mixer	6 Reverse osmosis pump
7 CIP pump	8 Reverse Osmosis Membranes	9 CIP Wash Tank
10 UV disinfection	11 Dosing Pump and Chlorine Reservoir	12 Control panel

Source: Adapted from Manual técnico de operação e manutenção da unidade de tratamento avançado por osmose reversa - Modelo ROH 010054 – $1m^3.h^{-1}$, Permution, 2016

In relation to the management and reuse of the RO brine, parallel studies were realized with the objective of obtaining an adequate disposal destination for this reject, by means of wetlands systems.

Rodrigues (2018) developed his research in the same pilot plant and obtained several recovery rates from the RO system for different TDS concentrations of brackish water. For TDS concentration of 1515 $g.m^{-3}$ the recovery rate was 27.99%. For TDS concentration of 1482 $g.m^{-3}$ the recovery rate was 32.40%. When TDS concentration was 1491 $g.m^{-3}$, was obtained recovery rate of 45.91%, and for brackish water with TDS concentration of 1045 $g.m^{-3}$ the obtained recovery rate was 69.13%.

As previously mentioned, this work has been divided into two chapters. The first one, entitled Energetic cost of pilot desalination system of brackish water – the use of smart grid as a solar energy had as general objective the detailing of the energy cost of the pilot desalination system with the use of solar energy combined with the energy supplied by the conventional network and their respective energy balance. The second chapter titled Mass and volume balances in a pilot desalination station of brackish water had the general objective of detailing mass and volume balances of the pilot desalination system with different initial concentrations of TDS.

2 ENERGY CONSUMPTION OF A BRACKISH WATER DESALINATION PILOT PLANT – THE USE OF SMART GRID WITH SOLAR ENERGY

ABSTRACT

Allied to the desalination question is the search for clean energy sources to supply this system, being it to reduce environmental pollution or to reduce the costs of this process. This work had as objective detailing the energetic balance and the energy consumption of a brackish water desalination pilot plant with the use of solar energy and energy provided by the grid, installed in the water treatment facilities of Sanepar in Praia de Leste in Pontal do Parana, Parana State. The system operated between May and June. The system had permeate production capacity of 1 $m^3.h^{-1}$ and utilizes ultrafiltration and softening as pre-treatment of reverse osmosis. Ten operations were performed in brackish water with initial total solids dissolved (TDS) concentration of 3,500 ± 100 $g.m^{-3}$ and ten operations with initial TDS concentration of 7,000 ± 100 $g.m^{-3}$ in brackish water. During the 20 operations the system consumed 388.61 kWh of energy, being 199.16 kWh spent in the 10 first operations to generate 27,94 m^3 of permeate at a consumption per m^3 of 5.83 $kWh.m^{-3}$. The remaining 189.45 kWh were consumed in the last 10 operations in the production of 26.85 m^3 of permeate at a consumption per m^3 of 5.34 $kWh.m^{-3}$. The effective consumption per m^3 of the treatment for both initial TDS concentrations was close to 4.30 $kWh.m^{-3}$. During the system operation days, the solar energy system, using 8 solar photovoltaic panels produced 111 kWh, 84 kWh of those between the 10 firsts operations and 27 kWh during the last 10 operations. It was observed that the solar energy system production was insufficient to supply the system demanding electric energy from the grid, working in a Smart Grid system.

Keywords: Ultrafiltration/ Reverse Osmosis, Solar Energy, Photovoltaic Panels, Produced Energy, Consumed Energy.

2.1 INTRODUCTION

The ocean is an extraordinary source of water, but the high salinity is a great issue. In this context, desalination is an important alternative for the lack of water, especially where it is scarce. There is as characteristic the high demand of energy, principally in the membrane filtration stages that require high pressures (ZARZO; PRATS, 2018; CHAFIDZ et al., 2014).

One aspect that marks many regions that suffer with water scarcity is the high index of solar radiation. It encourages the study of desalination systems propelled by solar energy, because it is an abundant, safe, clean and non-aggressive energy source to the environment. According to the International Energy Agency, until 2050, this kind of energy can become something bigger than electricity provided by the conventional grid (FRANKL; TANAKA, 2014;ZHANG et al., 2018; EL-BIALY et al., 2016).

There is an interest in cost reduction in desalination processes and an alternative is the use of the Smart Grid*,* an intelligent network that is a trustworthy system, integrated, reconfigurable and electronically controlled. In this system in which are used Information Technology (IT), the distribution grids from the dealerships integrate the local producers/consumers and interact receiving or providing energy according to the demands and providing a higher performance to the system. One of the integration possibilities to the smart grid is the use of photovoltaic (PV) panels for collecting solar energy to be later converted in electric energy (BRITANNICA ACAMEDIC, 2018; JOSHI; RAHMAN, 2015).

In Jordan, a brackish water desalination plant using RO propelled by photovoltaic solar energy was proposed together with a computer code generated to simulate and predict treated water production during one year. It was demonstrated that the system would be able to produce in the region of Tafila as an example, 1679 $m^3.year^{-1}$ of treated water from brackish water with TDS concentration of 7,000 $mg.L^{-1}$. In the region of Queira, for the same TDS concentration the permeate production would be equal to 1473 $m^3.year^{-1}$. The country has the annual total global solar irradiation incidence of between 1600 to 2300 $kWh.m^{-2}$ (HRAYSHAT, 2008).

Also, in Jordan a compact plant of BWD propelled by photovoltaic energy, treating water with TDS concentration of 1700 $mg.L^{-1}$ and production of 0.43 $m^3.day^{-1}$, presented a specific energy consumption of 16 $kWh.m^{-3}$, at 0.6 MPa of pressure for RO operation. The price of cubic meter of the water was reduced from US$ 15.6.

Besides the non-competitive price, this system can be used to treat water in exceptional situations (QIBLAWEY; BANAT; AL-NASSER, 2011).

Between 2006 and 2013, a BWD plant propelled by solar energy with PV panels was used to treat groundwater in a village with 300 inhabitants in Tunisia. It can produce 15 million of liters of treat water with TDS inferior to 300 $mg.L^{-1}$, the monthly average production rate between 3.26 and 12.8 $m^3.day^{-1}$ and specific energy consumption between 1.64 and 3.13 $kWh.m^{-3}$ (PEÑATE; GARCÍA-RODRÍGUEZ, 2012).

A study was developed from a desalination plant propelled by solar energy for the Masdar Institute of Science and Technology, located in Abu Dhabi, with the objective of treating 200 $m^3.day^{-1}$ and TDS concentration of 16,290 $mg.L^{-1}$. The permeate production capacity was 1,344 $m^3.day^{-1}$, at a cost of US$ 0.825. The energy production was 1,758 $MWh.year^{-1}$, being the estimate consumption of energy for water production of 6.99 $kWh.m^{-3}$. The justification of this low value is that the photovoltaic system capacity was largely superior to the need by the RO, what allowed the selling of the energy excess to the electric energy company (ALSHEGHRI et al., 2015).

In Malaysia a BWD system propelled with PV panels was developed, capable of treating feed water with TDS concentration of 5,000 $mg.L^{-1}$, generating 5.1 $m^3.day^{-1}$ of permeate with less than 50 $mg.L^{-1}$ of TDS, consuming 1.1$kWh.m^{-3}$. The peninsula region with lesser minimum solar irradiation is the South and presents average annual solar radiation of 4,794 $kW.year^{-2}$(ALGHOUL et al., 2016).

Facing this context of searching performance optimization of the desalination process and use of renewable energy to drive it, this chapter has as objective detailing the energetic balance and the energy consumption of the brackish water desalination pilot plant with the use of solar energy and energy provided by the grid. The system, installed in Praia de Leste, Pontal do Parana, Parana State, Brazil, has permeate production of 1 $m^3.h^{-1}$ and uses the ultrafiltration (UF) and softening as pre-treatment for reverse osmosis (RO).

2.2 MATERIALS AND METHODS

2.2.1 Experiments in the BWD pilot plant

Twenty water treatment experiments were realized, the first 10 with TDS concentration in brackish water of 3,500 ± 100 $g.m^{-3}$ and the last 10 with TDS concentration of 7,000 ± 100 $g.m^{-3}$. The duration of each experiment was 3h, totalizing 60 worked hours.

The mixture of water from Pombas River and seawater for the brackish water preparation had concentration correction for each experiment. For operations with water TDS of 3,500 ± 100 $g.m^{-3}$, the seawater tank pump transferred 0.61 m^3 of water to the brackish water tank. For operations with TDS concentration of water of 7,000 ± 100 $g.m^{-3}$, the seawater tank pump transferred 1.22 m^3 of water to the brackish water tank. In the brackish water homogenization process by pump recirculation of the motor and flow mirroring, it was kept turned on for the entire operation.

By means of jar test experiments, it was determined the coagulant PAC application was 492 $mL.h^{-1}$ for the operations with TDS of 3,500 ± 100 $g.m^{-3}$ and of 542 $mL.h^{-1}$ for operations with TDS of 7,000 ± 100 $g.m^{-3}$. The acid addition was not necessary, because PAC was able to control the average pH value of the water, of 6.48 in the first 10 operations and 6.87 in the last 10 operations.

An automatic backwash of the membrane happened every 30 minutes of the UF with 30 seconds of duration, resulting in 6 backwashes per operation. The backwash used the UF permeate in a reverse cycle to clean the membrane and the outflow was destined to the sludge reception tank from WTP.

Acid and basic chemical backwashes were realized in the UF membrane, with the objective of increasing its performance. The first one was previously executed previously to the 10 operations, the second was done in the beginning of the operations with 7,000 ± 100 $g.m^{-3}$ and the third between the fourteenth and the fifteenth operations. In the basic chemical cleaning, 2 L at 10% of sodium hydroxide solution was used in the acid washing, chloridric acid was used in 2 L at 10% solution.

About the RO system, in the 10 first treatment operations the softener realized the regeneration every 5h of operation. With the objective of increasing hardness removal, for the 10 last operations the regeneration gap was reduced to every 3h of operation.

The average pressure by the high-pressure pump of the RO for the operations with TDS concentration of 3,500 $g.m^{-3}$ was 10 bars and of 13 bars for TDS concentration of 7,000 $g.m^{-3}$.

The hydraulic membrane washes of the RO were done once after 3h of each experiment. The CIP used the permeate from the own RO in the cleaning operation and had a 14 minutes duration. This process used an average permeate volume of 0.12 m^3 in each operation.

The quantity of sodium metabisulfite added to the water after passing through the softener was 2 $mL.min^{-1}$ during the operation. When the system was not operated for more than two days, 20 $mL.min^{-1}$ of sodium metabisulfite were added for 10 minutes to preserve membrane integrity.

In relation to the solar energy production system, the PV panels had their positions corrected according to the month of operation. The 10 first experiments were realized between May 28th to June 21th. The 10 last experiments were realized between June 22nd to 29th. In May and June, the inclination angles of the PV panels were 30° and 35°, respectively.

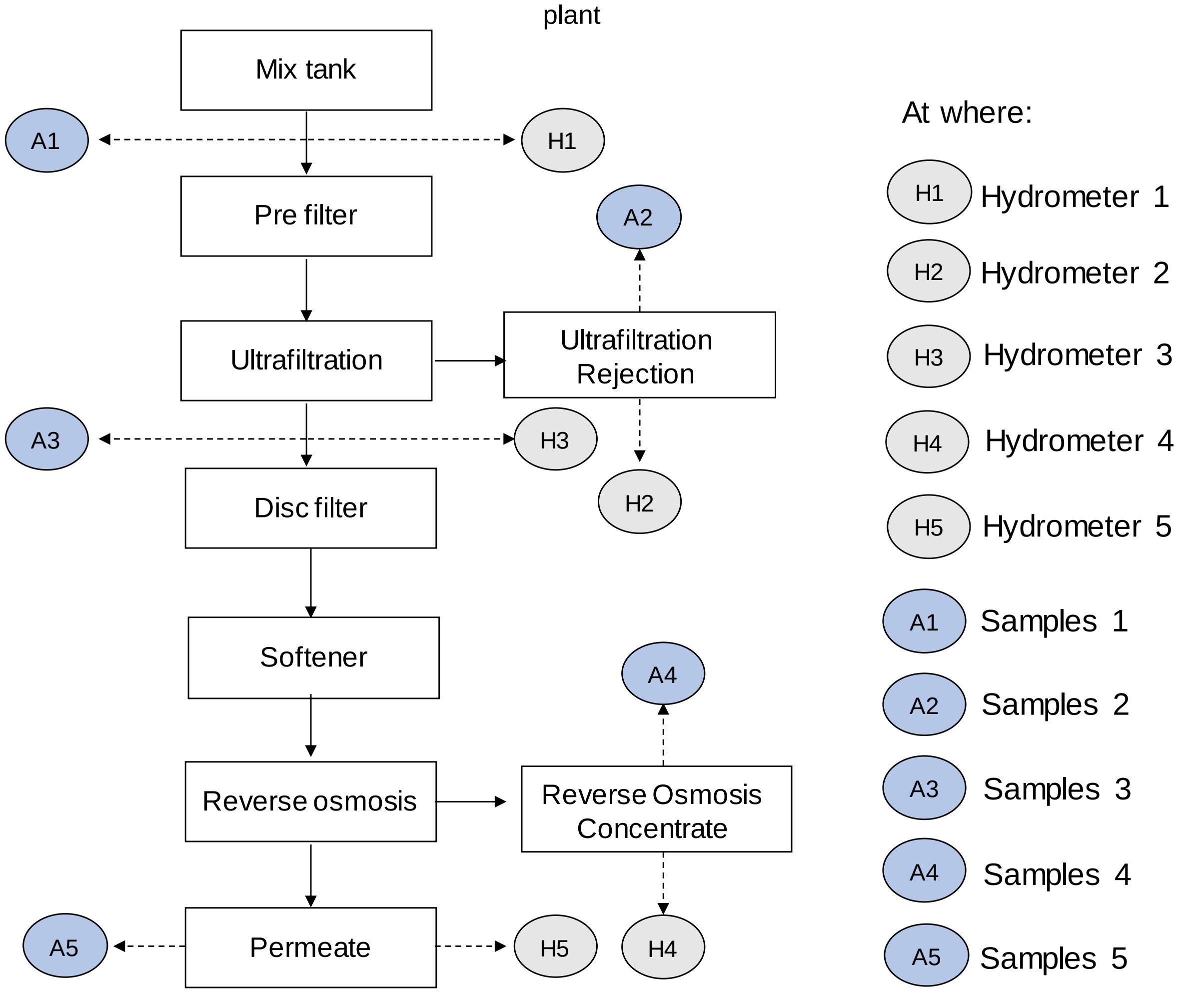
For the TDS concentration analysis in the diverse stages of the treatment was used the multi-parameter device Ultrameter II Model 6PV PPC^E from the company Myron. TDS, turbidity and pH analysis were done in the pilot desalination plant. The hardness and color evaluation were done in an analysis lab. For the attendance of the processed volumes were used hydrometers described in Table 2.1. Figure 2.1 shows the flow table of sample collection points and the hydrometers.

Table 2.1 - Hydrometers used in the BWD pilot plant

Hydrometer	Brand	Model	Class	Nominal diameter (mm)	Maximum flow ($m^3.h^{-1}$)	Rated flow ($m^3.h^{-1}$)	Minimum flow ($L.h^{-1}$)	Transition flow ($L.h^{-1}$)	Maximum pressure (bar)	Cargo loss (bar)
H1	Itron	Multimag TM II	B	40	20	10	200	800	10	<1
H2	Itron	Multimag TM II	B	40	20	10	200	800	10	<1
H3	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75
H4	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75
H5	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75

Source: Itron, 2018; Fae, 2018.

Figure 2.1 - Flow table of the hydrometers and sample collection points installed at BWD pilot plant



Point A1 were collected samples from brackish water that supplied the UF. Point A2 were collected samples from the reject generated by the UF backwash operation. Point A3 samples were collected after passing through the UF membrane that supplied the RO. Point A4 concentrate samples from the RO, generated during the CIP treatment were collected and at A5 RO permeate were collected.

In relation to the hydrometers, H1 showed the brackish water volume that supply the UF, H2 showed the reject volume generated during UF backwash, H3 showed the o permeate volume from the UF that supply the RO, H4 showed the concentrate volume that was generated during the RO and during the CIP and H5 showed the permeate volume produced by the RO.

The operations have 3h duration, and the hydrometer readings were done in the beginning and in the end of each hour of operation, totalizing 6 readings in each operation.

About the samples, they were also collected at each hour of operation: 5 minutes after the first UF backwash, 15 minutes after the first backwash and 25 minutes after the first backwash, totalizing 9 samples collected in each operation.

2.2.2 Description of the energy system

The electric energy that supplied the desalination pilot plant was composed by electric energy incoming from the grid, provided by Copel (*Companhia Paranaense de Energia*), and by the solar energy system, captured by PV panels.

The photovoltaic panels generated electricity in direct current by means of the photoelectric effect and depended on light incidence. The inverter that was installed after the photovoltaic modules allowed a synchrony between the modules and the grid, converting the direct current into alternate and injecting it in the grid. In the case of the energy distribution be turned off the inverter was mandatorily turned disconnected, needing the automatic shutdown. So, accidents can be avoided because the energy that is produced by the solar panels would not return to the dealership network. The photovoltaic system was formed by 8 photovoltaic modules model HR 250 P, Elco, with a power 250 Wp (Watts-peak) each, totalizing 2,000 Wp and 1 inverter Grid-Tie model SF 1600 TL B&B, with nominal power of 1,600 W and exit tension of 220 V. The area used by each panel was 1.49 m² and by the 8 panels was 11.88 m². The estimated daily energy production from the solar system was 6.76 $kWh.day^{-1}$, the monthly estimative was 209.44 $KWh.month^{-1}$ and the annual estimative was 251.23 $kWh.year^{-1}$ (EVEHX-SOLARH$_2$, 2015).

Table 2.2 presents the monthly average irradiation values in Praia de Leste and the ideal inclination angle for the solar panels. Inclination correction was done in a way to increase solar beam captation and consequently increase electric energy productivity.

Table 2.2 - Monthly average irradiation and ideal angles of the sun panels

Months	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Irradiation ($Wh.m^{-2}.day^{-1}$)	5390	5110	4400	3790	3110	2860	3080	3670	3870	4650	5310	5430
Ideal angle (°)	6	15	21	25	30	35	38	32	23	20	9	8

Source: EVEHX-SOLARH$_2$, 2015.

Figures 2.2 and 2.3 show the photovoltaic modules used in electric energy production for the desalination pilot plant and the current inverter, respectively.

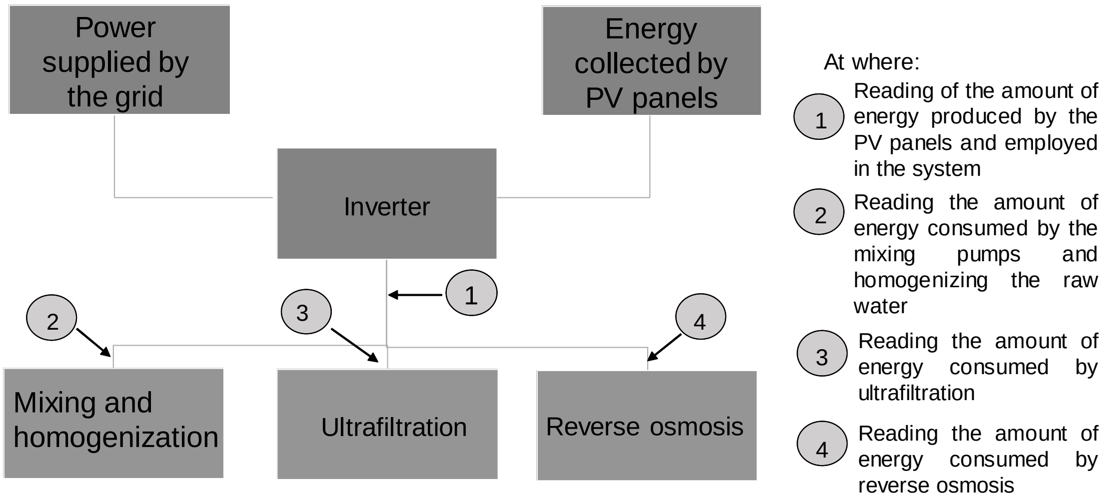
Figure 2.2 - Photovoltaic Modules from the BWD pilot plant

Figure 2.3 - Power inverter from the BWD pilot plant

2.2.3 Measurement points of electric energy production and consumption in the system

Figure 2.4 shows a schematic representation of where the energy meters were installed, the inverter, that measured the amount of energy captured by the PV panels, and the consume meters in the mixture and homogenization (MH) pumps of brackish water, the UF and RO. The meters were manufactured by the company Kron, and the model Konect.

Figure 2.4 -Schematic representation of energy measuring points



Point 1 were read the quantity of electric energy provided by the solar system. The inverter, besides converting direct current/alternate current, also promoted the connection with the electric network (Grid-Tie) injecting energy on the network that supplies the desalination system. The readings of energy production were realized during every system operation day at 11a.m.

Point 2 the readings of electric energy consumed during the brackish water MH processes were done.

Point 3 the energy consumption during the UF was measured.

Point 4 measurements of energy consumed during RO and CIP were noted. Readings were done and stored by the meters automatically.

The energy consumption data were collected at the beginning and at the end of each operation aiming the obtention of the consumed energy value per hour.

The energy data provided by the grid were obtained by means of difference between energy consumption and production by the photovoltaic system.

2.3.4 Electrics equipments installed in the desalination pilot plant

Equipments used in the system were present in the three stages of treatment and are described in Table 2.3.

Table 2.3 - Electrics equipments installed in the BWD pilot plant

System	Equipment	Brand	Model	Power (kW)
MH	Sea water pump	Schneider	BCR-2010	0,56
	Fresh water pump	Schneider	BCR-2010	0,56
	MH pump	Schneider	BCR-2011	0,75
UF	Capture pump	Grundfos	CR 3-3	0,55
	PAC agitator pump	Siemens	IEC 60034	0,09
	Circulation pump	Grundfos	CRN 3-3	0,25
	Backwash pump	Grundfos	CRN 3-3	3
RO	Feeding RO pump	Schneider	BC-98	0,37
	Pressurization pump	Grundfos	CM 3-3	0,75
	RO high pressure pump	Grundfos	CRI 3-23	3
	CIP pump	Grundfos	CM 1-8	1,12
Others	Lights Panels Instruments			0,5
Total power installed in the system				11,5

In the MH stage, sea water, river water and brackish water pumps were installed for TDS concentration adjustments and water homogenization.

In the UF stage were installed the PAC agitator pump, the captation pump, that had the function of permeating the water through the membrane, the circulation pump, that was activated only during the chemical washing of the membrane and the backwash pump, that was activated during the hydraulic membrane backwash operation.

Lastly, the pumps that were installed in the RO stage were the supplying pump, that transported the permeate from the UF to the RO, the pressurization pump, the RO high pressure pump, the pump that supplied the membranes and separated water in permeate and concentrate and the CIP pump, in charged the hydraulic membrane backwash.

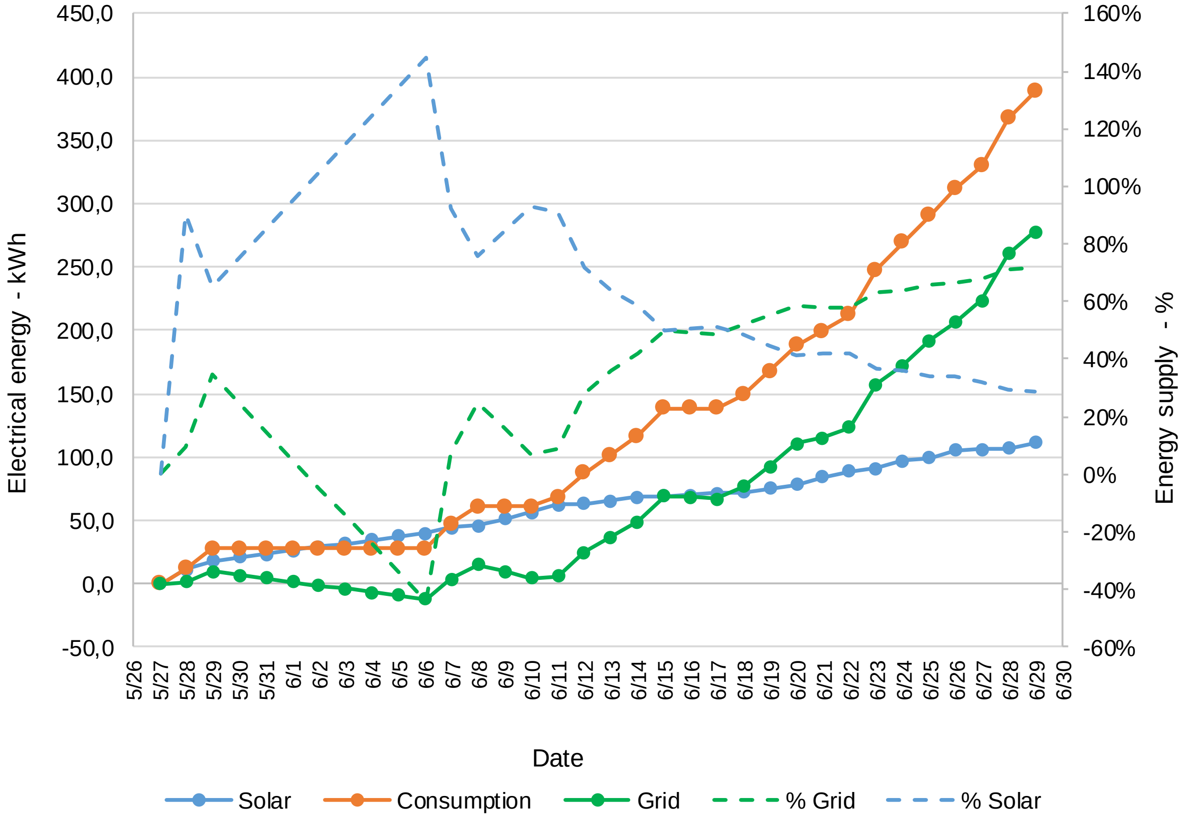
2.4 RESULTS AND DISCUSSION

The BWD pilot plant was operated during 60h: ten 3h experiments for the initial TDS concentration of 3,500 ± 100 $g.m^{-3}$ realized between May 28th to June 21st and ten more 3 hours experiments for the initial TDS concentration of 7,000 ± 100 $g.m^{-3}$ realized between June 22nd and 29th.

The BWD pilot plant was not operated between May 30th and June 6th, between June 9th and 11th, June 16th to 20th, and that on June 23rd and 28th two brackish water treatment operations were realized.

Figure 2.5 shows the accumulated curves of solar energy production, energy consumption by the system and energy provided by the grid, it also shows the solar energy and conventional grid energy percentage curves.

Figure 2.5 – Electric energy production, supplying and consumption in kWh between May 27th and June 29th of 2018



In this graphic is possible to observe that the energy produced by the PV panels was superior to the energy provided by the grid until June 18th. This is due the fact that in the same days in which the system was not operated, the PV panels gathered and produced energy that generated credit to be lately used during the operations. Starting from June 18th the electric energy provided by the PV panels was not enough to supply the system, being necessary the use of electric energy provided by the grid. As the operations with initial TDS concentration of 7,000 ± 100 $g.m^{-3}$ were realized without any days off, there was no time to generate energy balance produced by the PV panels, resulting in electric energy compensation by the network.

When the percentage curve provided by the conventional energy grid is decreasing, it shows that the solar energy was enough to sustain the system and by means of injections in the grid it creates a positive balance of solar electric energy. The days in which there is positive balance of solar energy are the same that the system was not operated, occurring an ascending conformation in the solar energy percentage curve. As mentioned before, starting from June 18th there was a higher demand for the electric energy from the grid by the system. On the general computation of consumed energy, about 26% was supplied by solar production and 74% by the grid.

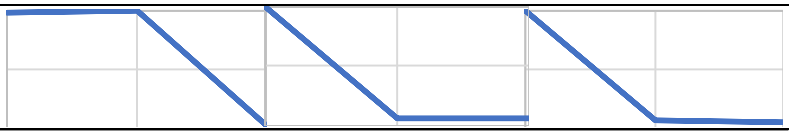
The average electric energy consumption during the 20 operations was 12.72 kWh. The minimum consumption was 10.97 kWh and occurred during the operations with initial TDS concentration of 3,500 ± 100 g.m^{-3}. The higher consumption was 21.63 kWh and occurred during the operations with initial TDS concentration of 7,000 ± 100 g.m^{-3}. The standard deviation of energy consume was 7 kWh.

The average daily production of solar energy considered from the first to the last days of operation was 3.70 kWh, with the minimum daily production of 1 kWh, that occurred in June 12th and 15th and maximum of 11 kWh that occurred in May 28th. The standard deviation is 2.22 kWh.

The average of electric energy provided by the conventional grid was 9.25 kWh, being the minimum value equal to 1.18 kWh, that occurred in May 27th and maximum equal to 36.53 kWh that occurred in June 28th, in that day two operations with initial TDS concentrations of 7,000±100 g.m^{-3} were done. The standard deviation was 8.29 kWh.

The Table 2.4 shows the relation of energy consumption with the color and turbidity removal graphics that happened in the UF system and the TDS removal that happened in the RO system for brackish water with initial TDS concentration of 3,500 ± 100 g.m^{-3}. To obtain the graphics the average data of the analyzed standard of water quality were used. For the electric energy consume data the accumulated percentage of each stage were used.

Table 2.4 – Energy consumption in the removal of color, turbidity and TDS for the initial concentration of 3,500 ± 100 g.m^{-3}

	%	TDS (g.m^{-3})	Color (uH)	Turbidity (NTU)
Removal	0,0 50,0 100,0			
System		MH UF RO	MH UF RO	MH UF RO
Consumption (% kWh)		57.62	4.34	

On the board it is possible to confirm that there was not TDS removal by the UF system but from the RO membrane. In this stage of the experiments the average TDS concentration of brackish water was 3,528 g.m^{-3}, equivalent to 99% of TDS, the average after passing through the UF membrane was 3,556 g.m^{-3}, equivalent to 100% of the TDS and after passing through the RO membrane the average was 28 g.m^{-3} that is equivalent to 1% of the TDS.

In relation to color removal it is possible to observe that the same happened in the UF stage. The average brackish water color was 44.27uH and after passing through the membrane these average values were 0.00uH, that is, the UF membrane removed 100% of the color.

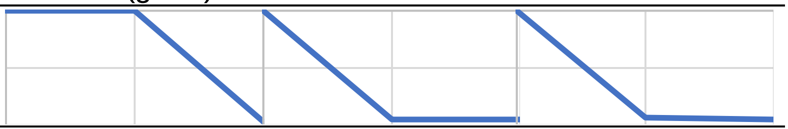
Turbidity was removed in the UF stage. The average turbidity of brackish water was 5.54NTU, that equals 100% of the turbidity, after passing through the UF membrane the average was 0.34NTU that is equivalent to 6% of turbidity and after the RO membrane the average value was 0.27NTU, equivalent to 5% of turbidity.

Table 2.4 also shows that the electric energy consumed to remove TDS was 57.62%, that is equivalent to 114.77 kWh. To remove color and turbidity this consume was 4.34% of the total consume, equivalent to 8.64 kWh. The remaining 38.04%, equivalent to 75.76 kWh, were consumed during the mixture and homogenization stages.

The standard values of water obtained after the UF and the RO treatments were proper to the established by Drinking Water Regulation 2914 from the Ministry of Health.

On Table 2.5 are presented the energy consumption graphics for the removal of color and turbidity by the UF system and the TDS removal by the RO system for brackish water with initial TDS concentration of 7,000 ± 100 g.m^{-3}. In this case the average standards for quality of the analyzed water were also used. For the electric energy consumption curve were used accumulated values.

Table 2.5 – Energy consumption in the removal of color, turbidity and TDS for the initial concentration of 7,000 ± 100 $g.m^{-3}$

	%	TDS ($g.m^{-3}$)	Color (uH)	Turbidity (NTU)
Removal	0,0 50,0 100,0			
System		MH UF OR	MH UF OR	MH UF OR
Consumption (% kWh)		56.82	7.82	

The performance of the UF and RO systems was similar both for experiments with brackish water with initial TDS concentration of 3,500 ± 100 $g.m^{-3}$ and for the experiments with initial TDS concentration of 7,000 ± 100 $g.m^{-3}$.

On Board 2.5 it is possible to observe that TDS removal occurred in the RO membranes. The average TDS concentration of brackish water was 7,022 $g.m^{-3}$, the average after the UF membrane was 7,038 $g.m^{-3}$ after passing through the RO membranes the average was 67 $g.m^{-3}$, in percentage the TDS removal through the RO membranes was 99%.

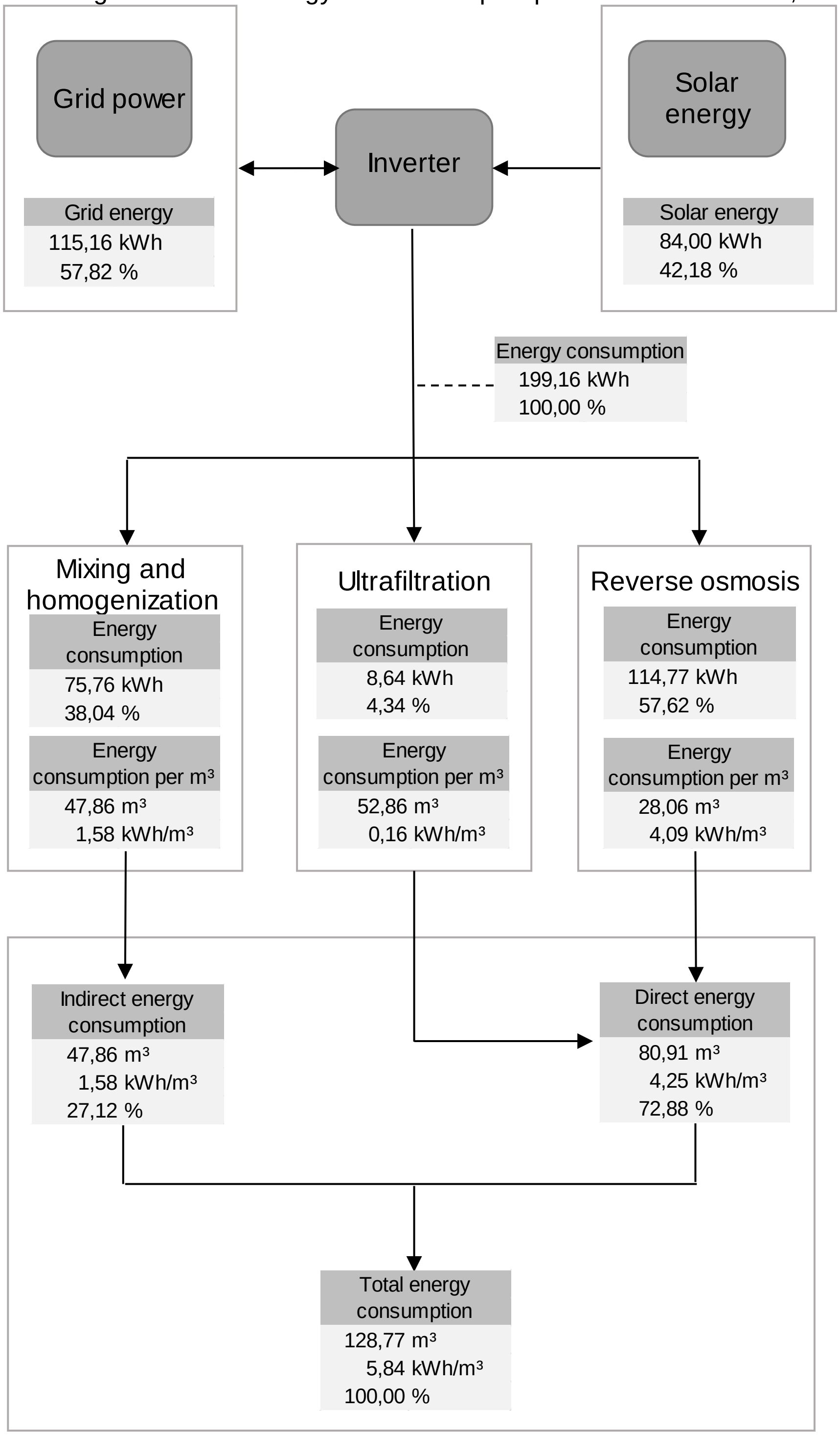
The board also shows that the color and turbidity removal happened during the UF stage. The average color of brackish water was 34.97uH and the average turbidity of brackish water was 5.20NTU. After passing through the UF membrane, 100% of color was removed. In relation to turbidity, the brackish water average was 5.20NTU, after passing through the UF membrane it was 0.33NTU and after passing through the RO membrane it was 0.25NTU. These values show that the UF membrane removed 100% of color and turbidity.

The electric energy consumed to remove TDS was 56.82%, equivalent to 107.65 kWh. To remove color and turbidity the electric energy consumption was 7.82%, equivalent to 14.81 kWh. The remaining 35.37%, equivalent to 67.00 kWh, were consumed in the mixture and homogenization of brackish water stage.

In this stage of the treatment the standard values for water obtained after the UF and RO treatments were proper to the patterns established by Drinking Water Regulation 2914 from the Ministry of Health.

In Figure 2.6 a diagram of the produced, consume and provided by the grid to the BWD pilot plant for the electric energy flow for operations with TDS of 3,500 ± 100 g.m^{-3} is presented.

Figure 2.6 - Flow diagram of used energy in the BWD pilot plant – initial TDS of 3,500 ± 100 g.m^{-3}



The diagram shows that the solar energy produced in the first 10 experiments was equivalent to 42.18% of the total consumed energy during the 3 operations. The remaining 57.82% of the energy consumed in these experiments were provided by the grid. The total energy consumed in this stage of the study was 199.16 kWh.

TDS adjustment and MH stages consumed 75.76 kWh to prepare 47.86 m^3 of brackish water. This consumption was equivalent to 38.04% of the total energy consumed during the three stages.

The UF consumed 4.34% of the total energy, that is, it spent 8.64 kWh to generate 52.86 m^3 of permeate. The permeate volume in this stage was bigger than the volume that was prepared during the previous stage, because the UF production was not able to follow the RO production for the initial TDS concentration of 3,500 ± 100 $g.m^{-3}$. To overcome this deficit, it was necessary to operate the UF stage previously to the 3 hours operations, what resulted in permeate volume of 4.28 m^3 superior to the brackish water prepared in the TDS adjustment and MH stages.

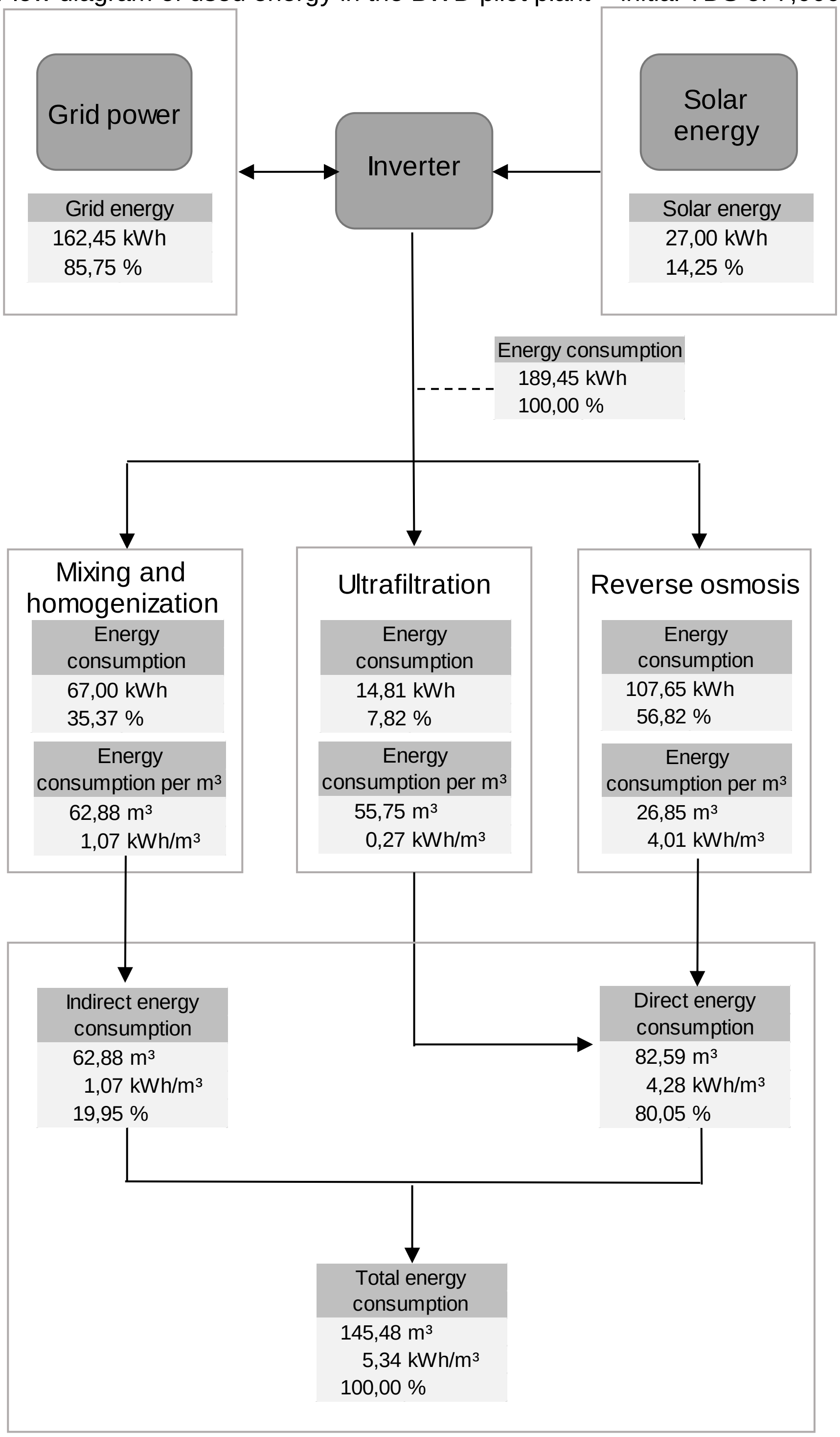
RO stage was responsible for the higher energy expense of the process. It consumed 114.77 kWh to generate 28.06 m^3 of permeate, that represents 57.62% of the total energy used.

As can be observed in Table 2.3 the total power installed in the system was 11.5 kW that multiplied by the total of worked hours during the first stage corresponds to the nominal electric energy of 345 kWh.

Figure 2.6 shows that the total consumed energy was inferior to the nominal energy and equal to 199.16 kWh, an equivalent value to 57.73% of the nominal power. The average useful power developed, equivalent to the energy cost per hour and obtained by mean of the ratio between the total electric energy consumed by the quantity of worked hours in this stage of the experiments was 6.65 kW.

Energy production and consumption data for the operations with TDS of 7,000 ± 100 $g.m^{-3}$ are detailed in the diagram in Figure 2.7.

Figure 2.7 - Flow diagram of used energy in the BWD pilot plant – initial TDS of 7,000 ± 100 g.m^{-3}



In the last 10 operations realized in the BWD pilot plant, the solar energy system was responsible for providing 27 kWh, that is, 14.25% of the energy that was consumed in the water treatment process. The grid complemented the remaining 85.75% of consumed energy that is equivalent to 162.45 kWh. The total energy consumed in this stage was equal to 189.45 kWh.

The TDS adjustment and MH stages were responsible for 35.37% of the consumed energy. It spent 67 kWh to process 62.88 m^3 of brackish water. The UF spent 7.82% of the consumed energy, that is equivalent to 14.81 kWh to generate 55.75 m^3 of permeate for the RO. The RO generated 26.85 m^3 of permeate consuming 107.65 kWh, that is, it consumed 56.82% of the total energy spent between the three stages of treatment.

As well as in the first stage of the experiments the electric energy that was consumed in the second stage of the experiments, equals to 189.45 kWh, it was inferior to the nominal electric energy of the system that was equal to 345 kWh. The energy consumption value is equivalent to 54.91% of the nominal electric energy. The cost per working hour of this experiment phase was 6.32 kW.

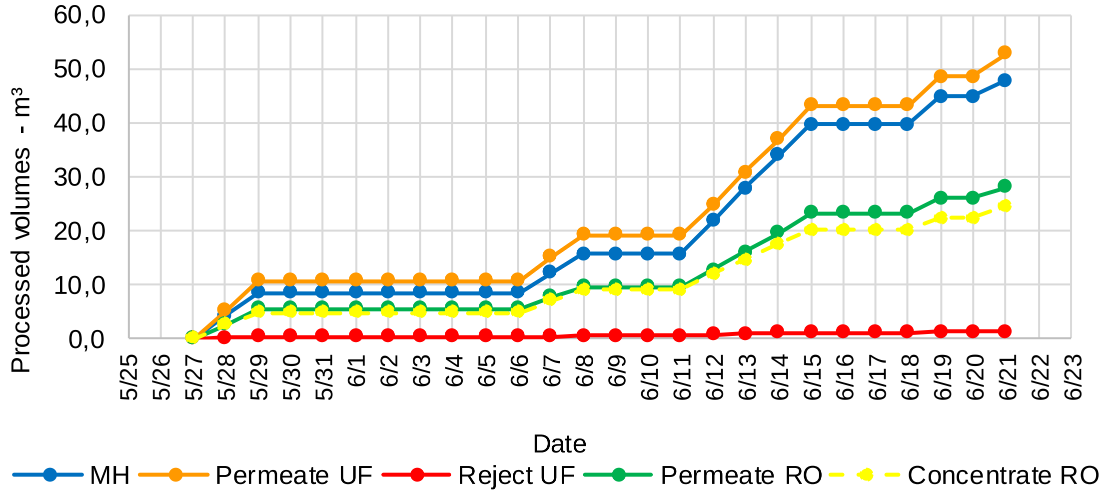
Considering the two stages of the treatment, together they consumed 388.61 kWh. This value is equivalent to 56.32% of the nominal power installed in both stages that equals 690 kWh.

The electric solar energy production during the 20 operations was 111 kWh, but as it can be seen in the flow diagram of Figure 2.8, the energy produced by the PV panels was superior in the first 10 operations in relation to the 10 last, due to the days in which no consecutive operation was done in the TDS of 3,500 ± 100 $g.m^{-3}$ phase.

Other fact that contributed for the low solar energy production was June was the month with the smaller average of solar irradiation in Praia de Leste, consequently the solar energy production in this month tends to be inferior in relation to other months of the year.

Figure 2.8 are presented the accumulated processed volumes in the BWD pilot plant for brackish water with initial TDS concentration of 3,500 ± 100 $g.m^{-3}$.

Figure 2.8 – Processed volumes in the BWD pilot plant – initial TDS of 3,500 ± 100 g.m^{-3}



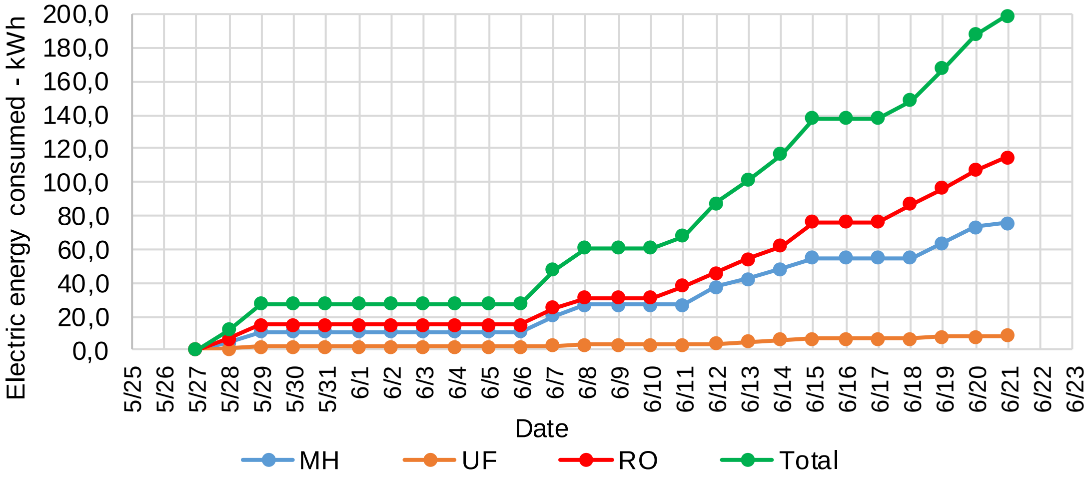
The accumulated volume of water that was processed in the MH stage during the 10 first operations with duration of 3 hours was 47.86 m³, corresponding to 84.69% of the volume that supplied the system. The remaining 15.31% are equivalent to 8.65 m³, that supplied the system being a result of the balance from the UF permeate tank volume between UF operations and also previous UF operations, so it was a sufficient volume balance to supply the RO, because in the first experiment phase the UF permeate production capacity was insufficient to supply the RO when both were working simultaneously.

From the total volume that supplied the UF, 52.86 m³, equivalent to 93.54%, it was permeate by the RO and 1.27 m³, equivalent to 2.25% were discarded during the backwashes. The remaining 2.38 m³ that supplied the UF correspond to 4.21%, they were used in sample collections and eventual reading and operating mistakes.

From the 52.86 m³ that supplied the RO, 28.06 m³, that is equivalent to 53.08% of the volume that supplied the RO and 49.65% of all volume that supplied the UF/RO are referent to the permeate and 24.43m³ are referent to the RO concentrate, equivalent to 46.23% of the volume that supplied the RO and 43.24% of the volume that supplied the entire UF/RO system. The remaining 0.37 m³ that are equivalent to 0.70% of the volume that supplied the RO and 0.65% of the entire volume that supplied the UF/RO system were used in sample collection and eventual reading or operating mistakes.

Figure 2.9 presents a graphic with the accumulated electric energy consumption in the BWD pilot plant for brackish water with initial TDS concentration of 3,500 ± 100 g.m^{-3}.

Figure 2.9 – Electric energy consumed during the processes in the BWD pilot plant – initial of 3,500 ± 100 g.m^{-3}



The total electric energy consumed in the MH stage was 75.76 kWh, corresponding to 38.04% of all the energy that was consumed during the first stage of experiments. In the UF stage 8.64 kWh were spent, this corresponds to 4.34% of the consumed electric energy. RO was the stage that consumed more energy, it spent 114.77 kWh, that is equivalent to 57.62%. The total electric energy consumed in the first stage of the experiments was 199.16 kWh.

Figure 2.10 shows the graphic of the energy consumption per m³ to produce permeate in the BWD system for brackish water with initial TDS concentration of 3,500 ± 100 g.m^{-3}, and it was obtained by the ratio between the electric energy consumption demanded by the operation, by the permeate volume that was produced during the operation.

Figure 2.10 – Energy consumption per m³ to produce permeate in the BWD pilot plant – TDS initial of 3,500 ± 100 g.m^{-3}

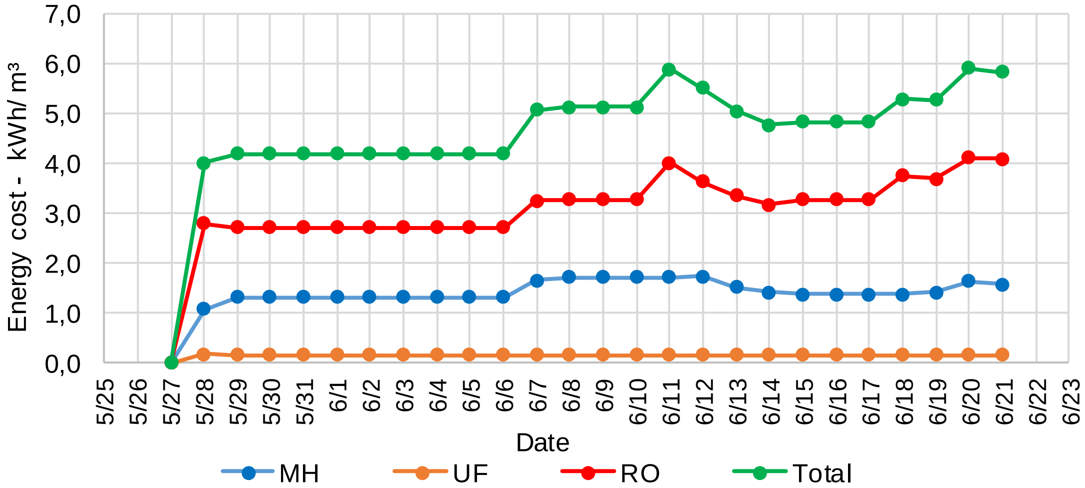
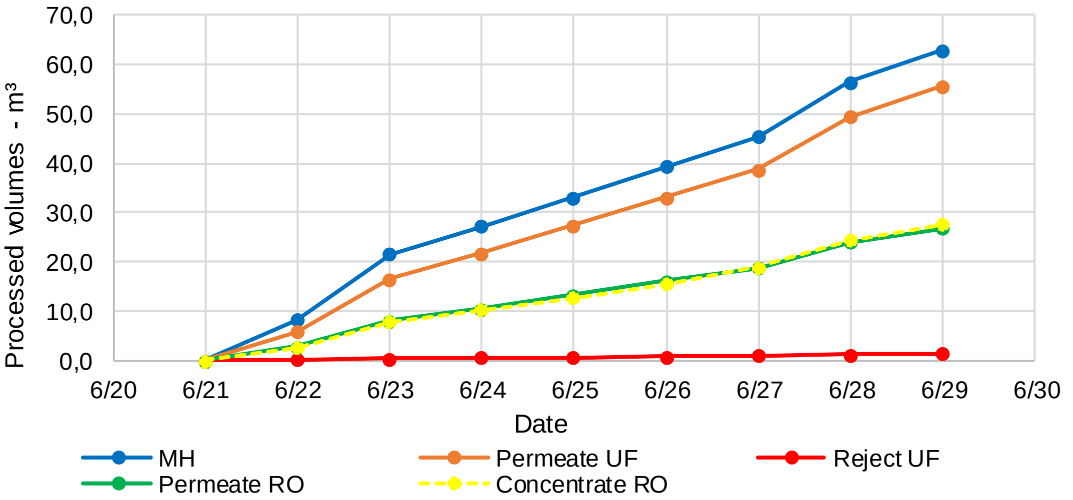


Figure 2.10 shows that the energy consumption per m³ in the MH stage of brackish water, considered indirect was 1.58 kWh.m^{-3} and it is equivalent to 27.12%. For the permeate production from the UF, the energy consumption was 0.16kWh.m^{-3} and it is equivalent to 2.80% of the energy consumed. For the RO permeate production, the energy consumption was equal to 4.09kWh.m^{-3} and it is equivalent to 70.08%. These last two summed were considered as direct cost since these stages were, in fact, responsible for treating the water and they are equivalent to 4.25kWh.m^{-3}, that is, 72.88% of the electric energy consumption of the system.

The energy consumption per m³ to prepare brackish water was considered as indirect, because in usual situations the water to be treated can avoid TDS adjustment. The consumption corresponded to 27.12% of the energy consumption to treat water in the 10 operations. The energy consumption per m³ was 5.84kWh.m^{-3}.

Figure 2.11 presents the graphic that is referent to the accumulated volumes processed by the BWD pilot plant during the experiments with brackish water with initial TDS concentration of 7,000 ± 100 g.m^{-3}.

Figure 2.11 – Processed volumes in the BWD pilot plant – initial TDS of 7,000 ± 100 g.m^{-3}



The accumulated volume of brackish water processed in the MH stage was 63.61 m³, equivalent to 100% of the volume that supplied the UF. From this volume 62.88 m³ were results from the operations with the 3 hour duration and this is equivalent to 98.86%. The remaining 0.72 m³ was the residue volume left in the UF permeate tank between the operations and is equivalent to 1.14% of the brackish water that supplied the system.

From the total volume that supplied the UF 55.75 m³ and its equivalent were composed of the permeate and supplied the RO. The accumulated volume of generated rejects in this stage was 1.30 m³ and it is equivalent to 2.07%. The remaining 5.83 m³, equivalent to 9.28% were used in sample collection and in eventual reading and operating mistakes.

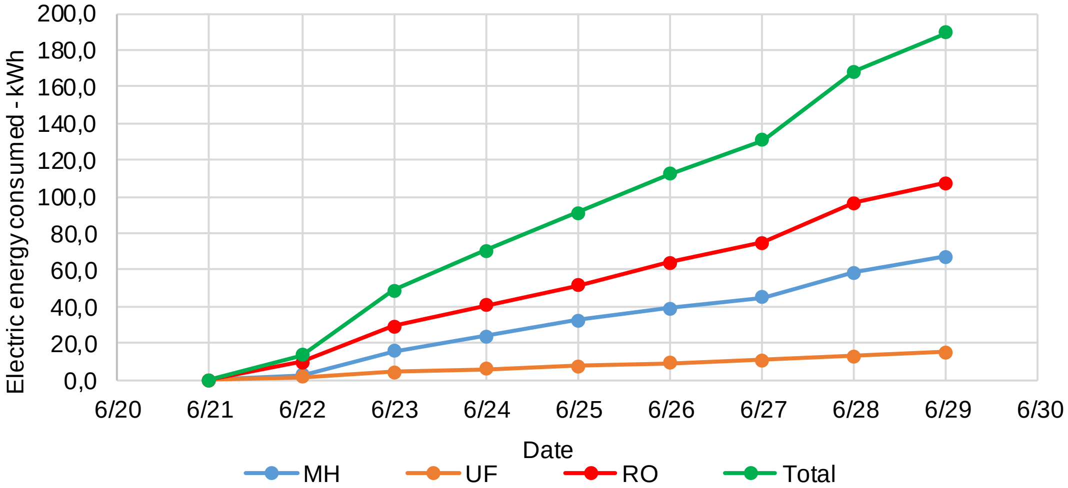
From the 55.75 m³ of UF permeate, 26.85 m³ were treated, that is equivalent to 48.16% of the volume that reached the RO leaving the UF and it is equivalent to 42.21% of the total processed volume in this stage of the experiments.

The concentrate volume generated by the RO was 27.61 m³, corresponding to 49.53% of the volume that supplied the RO and 43.41% of the entire volume that was processed in this stage of the experiments.

The remaining 1.29 m³ were used in sample collection and in eventual reading and system operation errors. This value is equivalent to 2.31% of the volume that supplied the UF and to 2.03% of all processed volume.

Figure 2.12 shows the graphic that refers to the accumulated consumption of electric energy during the processes in the BWD pilot plant for the brackish water with initial TDS concentration of 7,000 ± 100 g.m^{-3}.

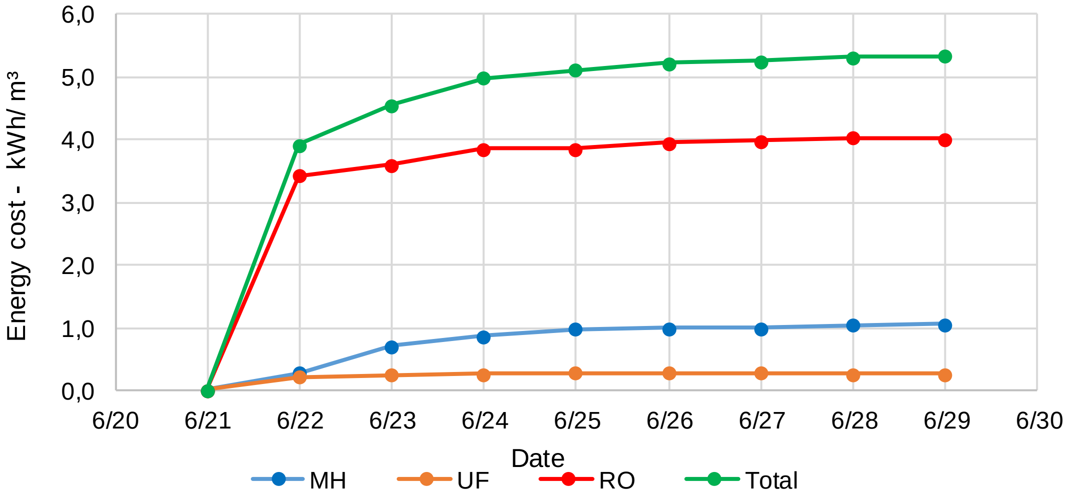
Figure 2.12 – Electric energy consumed during the processes in the BWD pilot plant – initial TDS of 7,000 ± 100 g.m^{-3}



This stage of the experiments consumed 189.45 kWh of electric energy. In the MH stage of brackish water were consumed 67.00 kWh, equivalent to 35.37% of the total. UF stage consumed 14.81 kWh, equivalent to 7.82% of the total cost. RO was the stage that consumed the most, spending 107.65 kWh, a value that corresponds to 56.82% of the total energy consumed.

Figure 2.13 shows the graphic that refers to energy consumption per m³ to produce permeate in the BWD pilot plant for brackish water with initial TDS concentration of 7,000 ± 100 g.m^{-3}. This value was obtained by mean of the ration between the electric energy consumption in the operation by the permeate volume generated in the operation.

Figure 2.13 – Energy consumption per m³ to produce permeate in the BWD pilot plant – initial TDS of 7,000 ± 100 g.m^{-3}



The energy consumption per m³ to prepare the brackish water, that was considered as indirect, was 1.07 kWh.m^{-3}, equivalent to 19.95% of the energy consumption per m³. The energy consumption per m³ to produce permeate from UF and RO were respectively 0.27 and 4.01 kWh.m^{-3}, respectively equivalent to 4.97% and 75.08% of the total energy consumption per m³ that together generated a direct consumption of 4,28 kWh.m^{-3} and equivalent to 80.05% the total energy cost was 5.34 kWh.m$^{-3.}$

In relation to the total consumed energy, the 10 operations that treated water with initial TDS concentration of 7,000 ± 100 g.m^{-3} was inferior to the energy consumed in the operations with initial TDS of 3,500 ± 100 g.m^{-3}. The lowest energy consumption happened in the last 10 operations in the stages that prepared brackish water and in the RO. In total, the system consumed 388.61 kWh during the 20 operations, and from this value, the conventional grid provided 277.61 kWh.

The only stage in which the energy consumption in the last 10 operations was superior to the 10 first was in the UF, because two chemical backwashes more were done in this membrane due to the higher TDS concentration. The first wash happened in the beginning of the last 10 operations and the second wash happened after 15 operations. Consequently, the total cost to treat the water was also inferior in the last 10 operations.

About the energy consumption in each stage of the operation, the most economic was the UF, followed by TDS adjustment and HM of brackish water. As expected what spent more electric energy was RO, because such as shown in Table 2.3, the more powerful pumps were installed in this stage, that were the high membrane pressure and the CIP.

The average utile power developed (energetic consumption per hour) in the first monitoring stage was 6.65 kW. In the second stage the utile power was 6.32 kW. In relation to the total power installed, the system consumed 56.3% of the nominal energy to the full power, indicating an operational mode in moderate regimen.

In the Southern Seawater Desalination Plant, of Binningup, Western Australia, with production capacity of 306000 $m^3.day^{-1}$, the energy consumption of the UF was 0.05 $kWh.m^{-3}$, equivalent to 1.3% of the energy consumption of the treatment, and the RO consumed 2,65 $kWh.m^{-3}$, that is 71.4 % of the energy expenses (ZARZO; PRATS, 2018).

Richards and Schäfer developed in 2002, in White Cliffs, in the interior of Australia, a prototype of a desalination system supplied with electric energy produced by PV panels and that used UF as a pre-treatment to RO. They tested diverse configurations of waters that supply the system, among them the water from White Cliffs barrage with salinity of 150 $g.m^{-3}$, but too turbidity and the water from the well in Glenhope Plant, that had salinity of 3,500 $g.m^{-3}$. To treat the water from the barrage the system consumed 2.5 $kWh.m^{-3}$ of energy and generated permeate with saline concentration of 10.5 $g.m^{-3}$. The UF removed all turbidity, viruses and bacteria. In its turn, the energy needed to treat the water from the well was 6.5 $kWh.m^{-3}$ and generated permeate with salinity of 175 $g.m^{-3}$, this consumption was similar to the obtained in the pilot desalination plant compared in Praia de Leste (RICHARDS; SCHÄFER, 2003).

2.5 CONCLUSION

The total energy consumption during the 20 experiments was 388.61 kWh, of which 199.16 kWh were consumed in the first 10 and 189.45 kWh in the last 10 experiments. The effective energy consumption for the 20 trials was close to 4.30 $kWh.m^{-3}$, and, due to the preparation of raw water, the total energy consumption per m^3 in the first 10 trials was raised to 5.83 $kWh.m^{-3}$ to produce 27.94 m 3 of permeate. In the last 10 tests the energy consumption per m^3 was 5.34 $kWh.m^{-3}$ to produce 26.85 m^3 of permeate.

By means of the results obtained in this study it was possible to observe that the preparation of brackish water elevated the energy cost from the first and second stages of the experiments in approximately 27% and 20%, respectively.

If only the effective energy cost to treat the water in the UF and RO systems are taken for consideration, it was very close for both initial TDS concentrations of brackish water: the consumption percentage of UF would be 7% and the RO 93% in the 10 first experiment. In the last 10 experiments by their turn the percentages would be of 12% in the UF system UF and 88% in the RO system. The total energy cost of the 10 first experiments was 0.49 $kWh.m^{-3}$ bigger in relation to the last 10 experiments.

During the days the system was operated and using 8 PV panels, the solar system produced 111 kWh of which 84 kWh were produced between the first 10 operations and 27 kWh produced during the last 10 operations.

About the solar electric energy production, it was not enough to supply the BWD pilot plant because it was operated in the time of the year with the lesser index of solar radiation in the place of study, that is, for the solar energy system be enough to supply the system, 28 PV panels of the same model used in this monitoring would be necessary and the PV panels should have a minimum power of 11.5 kW.

3 MASS AND VOLUME BALANCES IN A BRACKISH WATER DESALINATION PILOT PLANT

ABSTRACT

Encouraged by the growing demand of potable water, studies that refer to desalination have been developed with the aim of improve the performances of these systems. In this context, this work had as objective the detailing of mass and volume balances of a brackish water desalination pilot plant of with permeate production capacity of 1 $m^3.h^{-1}$, formed by ultrafiltration (UF) and softening as pre-treatment of reverse osmosis (RO). The system was installed in the facilities of the water treatment plant belonging to Sanepar, in Praia de Leste, Pontal do Parana, Parana State. Twenty operations were realized in it, being the 10 first done with brackish water with initial total dissolved solids (TDS) concentration of 3,500 ± 100 $g.m^{-3}$ and the last 10 done with initial concentration of brackish water of 7,000 ± 100 $g.m^{-3}$. By mean of the mass balance it was possible to observe that the RO permeate represents about 0,40% of the TDS quantity that enters in the system, that is, the average TDS concentration of the permeate of the 10 first operations was 28 $g.m^{-3}$ and the last 10 was 67 $g.m^{-3}$. By mean of the volumetric balance it was observed that the permeate production was superior when the initial TDS concentration was smaller: 50% of the water that supplied the UF-RO set transformed into permeate in the first 10 operations and 42% of the brackish water that supplied the system in the last 10 operations that transformed the permeate.

Keywords: Ultrafiltration / Reverse Osmosis, Desalination, Brackish Water, Mass Balance, Volume Balance.

3.1 INTRODUCTION

The current scenario of searching for water supply source due to the increase in demand and water scarcity increase the interest by desalination technologies. This process in its turn requires a pre-treatment to keep its good progress. In the processes that use membranes the pre-treatment must be capable of removing the suspended solids, being them of mineral or biological origin. The advantages of use the ultrafiltration (UF) system as a pre-treatment are producing supply water in a constant volume and high quality for the Reverse Osmosis (RO), incrustation reduction, that results in less cleaning and increasing the RO membrane lifespan (NATIONAL RESEARCH COUNCIL, 2008).

Halpern, McArdle and Antrim realized a case study in a desalination pilot plant in Ewe Beach, Hawaii. The pilot plant composed by UF followed by RO in two stages treated raw water from a well 488 m deep and average initial TDS concentration of 34,842 $g.m^{-3}$. After passing through the first RO stage the permeate had a TDS concentration between 150 and 325 $g.m^{-3}$, and after passing through the second stage the TDS concentration was between 1 and 37 $g.m^{-3}$ (HALPERN; MCARDLE; ANTRIM, 2005).

Xu et al. developed a pilot study in a desalination plant of sea water installed in Qingdao Jiaozhou Bay, at the Yellow Sea in China. The plant used UF as a pre-treatment of RO and had permeate production capacity of 1.17 $m^3.h^{-1}$. The system can produce permeate with TDS concentration between 224 and 320 $g.m^{-3}$ starting with raw water with TDS concentration between 26,000 and 30,120 $g.m^{-3}$ (XU et al., 2007).

Tomaszewska and Bodzek realized a study in a pilot desalination plant with production capacity of 1 $m^3.h^{-1}$. It was installed in the Mineral Research and Energy Economy Institute from the Academy of Sciences of Poland, and it used UF as pre-treatment for 2 stages in series RO, due to the high concentration of boron. The treated water coming from the geothermal region had an initial TDS concentration equal to 2,473.37 $g.m^{-3}$. The TDS removal rate after passing through the first RO stage was 93% and after passing through the second RO stage it was 95% (TOMASZEWSKA; BODZEK, 2013).

Considering the search for alternative sources of consumption water, this work had as objective detailing the mass and volume balances of the pilot desalination system of brackish water, installed in Praia de Leste, Pontal do Parana, Parana, Brazil, composed by UF and softening as pre-treatments of RO with different initial concentrations of total dissolved solids in brackish water.

3.2 MATERIALS AND METHODS

3.2.1 Experiments in the BWD pilot plant

The water treatment experiments were divided in two stages: 10 operations with initial TDS concentration in brackish water of 3,500 ± 100 $g.m^{-3}$ and 10 operations with initial TDS concentration of 7,000 ± 100 $g.m^{-3}$. Each experiment had a 3 hour operation, totalizing 60 worked hours.

For the brackish water preparation, it was realized a mixture of water from Pombas River with sea water to correct the TDS concentration before each operation. For the operations with initial TDS concentration of brackish water equal to 3,500 ± 100 $g.m^{-3}$, the seawater tank pump transferred 0.61 m^3 of water to the brackish water tank, and for the 10 operations with TDS concentration equal to 7,000 ± 100 $g.m^{-3}$ the seawater tank pump transferred 1.22 m^3 of water to the brackish water tank. The homogenization of brackish water, by mean of recirculation of the motor pump and flow spreading was kept during all the operation time.

PAC coagulant application was determined after jar test experiments. For the water with TDS of 3,500 ± 100 $g.m^{-3}$ the application was 492 $mL.h^{-1}$ and allowed that the pH was 6.48. For water with TDS of 7,000 ± 100 $g.m^{-3}$ the application was 542 $mL.h^{-1}$ and the pH value of the water was 6.87.

In the UF system at each 30 minutes of operation the mechanical membrane backwash happened, that used permeate from the UF system. Each backwash had 30 seconds duration. There were realized 6 backwashes per operation.

Besides the mechanical membrane backwashes of the UF, some basic and acid chemical washes were also realized in the UF membrane. The first was before the beginning of the operations, the second was done in the beginning of the operations with TDS of 7,000 ± 100 $g.m^{-3}$ and the third was done in the middle of the experiments with this TDS concentration. On the basic chemical backwash, was used the sodium hydroxide with solution of 2L at 10% and in the acid backwash the hydrochloric acid with solution of 2L at 10%.

In relation to the RO in the 10 first treatment operations the softener realized the regeneration at each 5 hours of work. For the 10 final operations the regeneration gap was reduced for each 3 hours of operation.

In relation with the RO osmotic pressure, the average during the operations with TDS of 3,500 $g.m^{-3}$ was 10 bar and for TDS of 7,000 $g.m^{-3}$ it was 13 bar.

The hydraulic backwashes of the RO membranes had 14 minutes duration. They were executed with the RO permeate and were done once every 3 hours of experiment. The CIP used an average volume of permeate of 0.12 m^3 per operation.

With the objective of preserving the RO membranes, was added the sodium metabisulfite in the water after passing through the softener and application flow was 2 $mL.min^{-1}$ during the operation. For operation pauses superior to two days 20 $mL.min^{-1}$ of sodium metabisulfite were added to the water during 10 minutes.

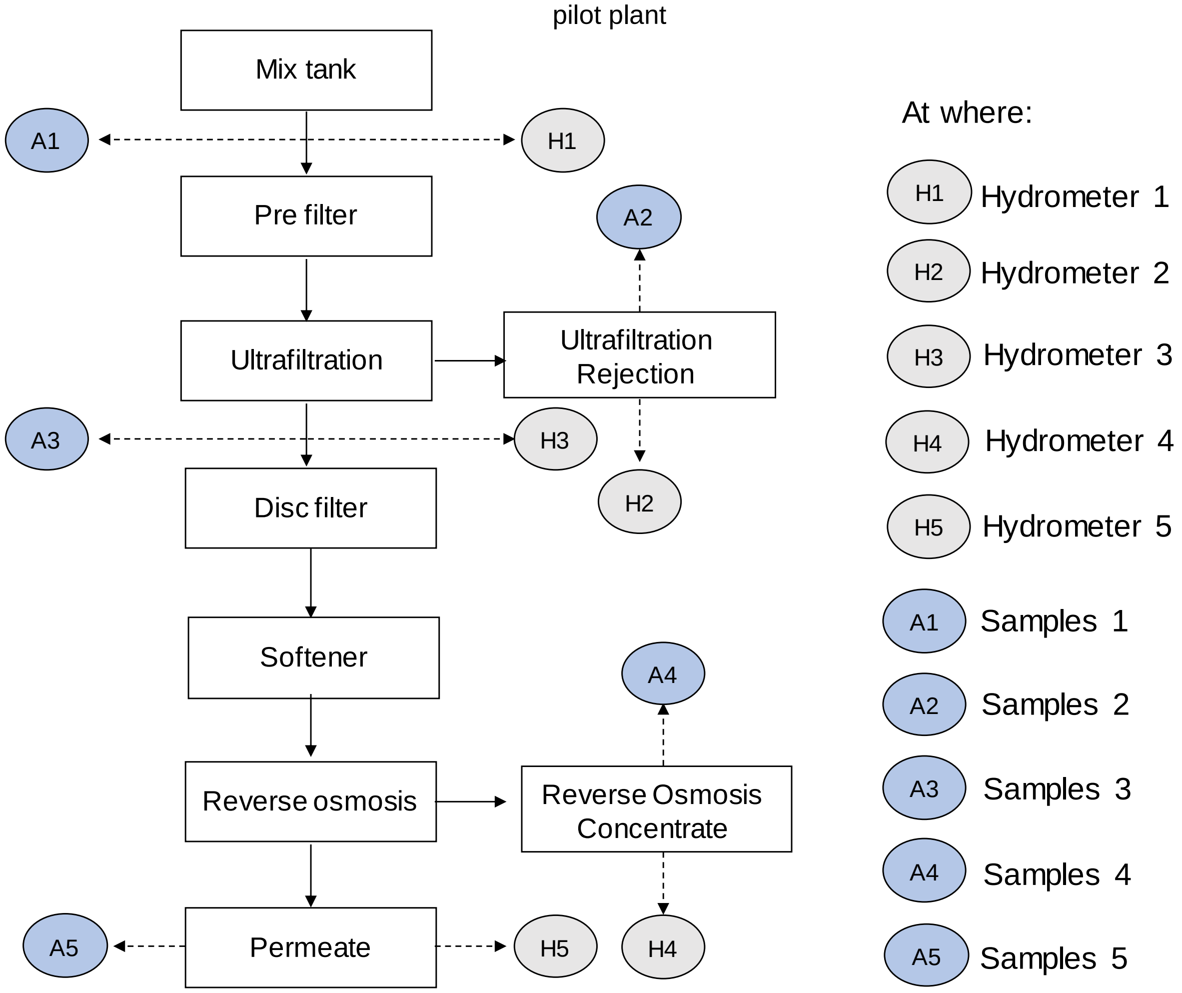
For the formulation of the mass balance samples were collected and an analysis of the TDS concentration was done using the multi-parameter device Ultrameter II Model 6PV PPC^E from the manufacturer Myron. To elaborate the volume balance, readings of the hydrometers described in Table 3.1 were done. Figure 3.1 shows a flow table of where the sample collection points and the hydrometers were installed.

Table 3.1 - Hydrometers used in the BWD pilot plant

Hydrometer	Brand	Model	Class	Nominal diameter (mm)	Maximum flow ($m^3.h^{-1}$)	Rated flow ($m^3.h^{-1}$)	Minimum flow ($L.h^{-1}$)	Transition flow ($L.h^{-1}$)	Maximum pressure (bar)	Cargo loss (bar)
H1	Itron	Multimag TM II	B	40	20	10	200	800	10	<1
H2	Itron	Multimag TM II	B	40	20	10	200	800	10	<1
H3	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75
H4	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75
H5	Fae	Delta MTF Multijato	B	20	3	1,5	30	120	10	0,75

Source: Itron, 2018; Fae, 2018.

Figure 3.1 - Flow table of the hydrometers and sample collection points installed in the BWD pilot plant



About the hydrometers, H1 indicated the brackish water volume that supplied the system, H2 indicated the volume of reject produced during the UF backwash, H3 indicated the permeate volume produced by the UF and that fed the RO, H4 indicated the concentrate volume generated during the RO and during CIP and H5 indicated the permeate volume produced by the RO.

In relation to samples, in A1 were collected brackish water from the system, in A2 were collected reject samples produced by the UF backwash, in A3 were collected UF permeate that supplied the RO samples, in A4 were collected RO concentrate samples generated during the treatment operation and CIP in A5 were collected the RO permeate samples.

Each operation had duration of 3 hours, and the hydrometers readings were done at the beginning and at the end of each operation, resulting in 6 readings per operation in each hydrometer.

The sample collections were done in the five points per hour of operation: 5 minutes after the first UF backwash, 15 minutes after the first backwash and 25 minutes after the first backwash, totalizing 9 sample collections per operation in each point.

3.3 RESULTS AND DISCUSSION

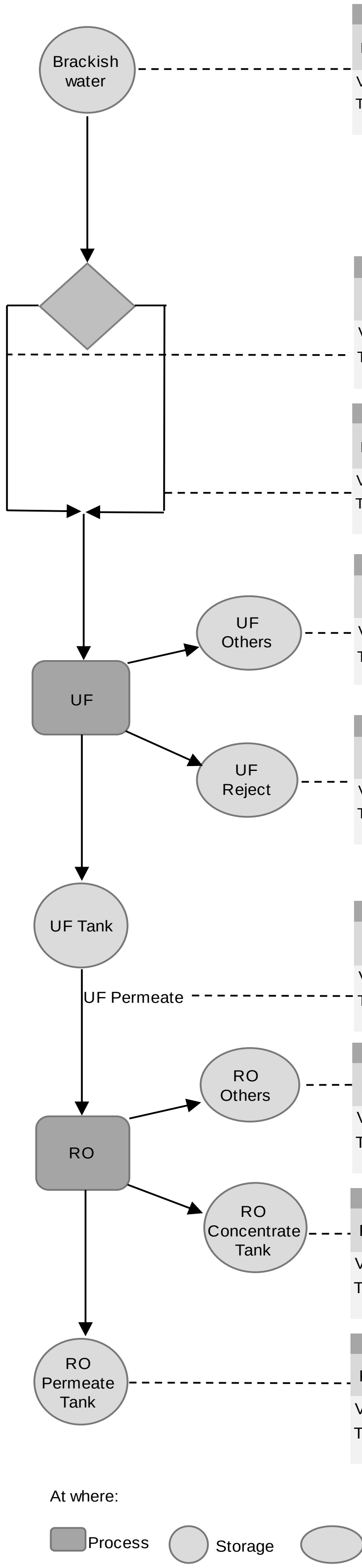
The brackish water desalination pilot plant was operated in two stages, the first consisted in 10 experiments of 3 hours for the initial TDS concentration of 3,500 ± 100 $g.m^{-3}$ and the second consisted in 10 experiments with 3 hours duration for the initial TDS concentration of 7,000 ± 100 $g.m^{-3}$, in a total of 60 hours of operation.

In the first stage of the experiments the average volume that was processed per 3 hour operation was 4.79 m^3 of brackish water, the UF permeate was 5.29 m^3 (considering the pre-existent 0.50 m^3 in the UF permeate storage tank), of the balance in the UF tank was 0.87 m^3, of the reject that was generated by the UF was 0.13 m^3, of the RO permeate was 2.81 m^3 and the concentrate that was generated by the RO was 2.44 m^3.

The average of processed flow rate per hour in the first stage of the experiments was 1.60 $m^3.h^{-1}$ in brackish water input, from the UF permeate it was 1.76 $m^3.h^{-1}$, of the UF permeate to supply the RO was 0.29 $m^3.h^{-1}$, of the reject generated by the UF was 0.04 $m^3.h^{-1}$, of the RO permeate was 0.94 $m^3.h^{-1}$ and of the concentrate generated by the RO was 0.81 $m^3.h^{-1}$. UF permeate production was superior to the specified by the manufacturer, while the RO had its production inferior to what was specified by the manufacturer.

Figure 3.2 shows a flowchart with the accumulated results obtained from the mass and volume balances of the system for the initial TDS operations of brackish water being 3,500 ± 100 $g.m^{-3}$ and details the values found during the same experiments.

Figure 3.2 – Flowchart of the desalination pilot plant – accumulated results obtained in the mass and volume balances with initial TDS concentration of 3,500 ± 100 g.m^{-3}



Treatment Stage: Brackish Water - during the experiments

Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	56,51	100,00	-0,01	6,14	5,65	2,20
TDS (g.m^{-3})	3556		3486	3595	3543	30
Mass (g)	200939,5	100,00	-44,4	21825,4	20022,5	7728,6

Treatment Stage: Brackish Water - pre experiments

Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	47,86	84,69	3,01	6,14	4,79	1,22
TDS (g.m^{-3})	3556		3486	3595	3543	28
Mass (g)	170181,5	84,69	10649,3	21825,4	17018,1	4269,5

Treatment Stage: UF Tank

Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	8,65	15,31	-0,01	1,28	0,87	0,39
TDS (g.m^{-3})	3556		3486	3595	3543	28
Mass (g)	30756,1	15,31	-44,4	4531,9	3075,6	1382,0

Treatment Stage: UF Others

Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	2,38	4,21	-0,30	1,21	0,24	0,49
TDS (g.m^{-3})	3546		3518	3579	3546	24
Mass (g)	8434,6	4,20	-1066,7	4324,7	843,5	1741,8

Treatment Stage: UF Reject

Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	1,27	2,25	0,05	0,18	0,13	0,04
TDS (g.m^{-3})	3550		3527	3584	3550	20
Mass (g)	4513,6	2,25	190,0	653,1	451,4	137,5

Treatment Stage: UF Permeate

Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	52,86	100,00	93,54	4,08	6,38	5,29	0,79
TDS (g.m^{-3})	3556			3517	3595	3556	28
Mass (g)	187956,7	100,00	93,54	14419,5	222530,0	18795,7	2738,7

Treatment Stage: RO Others

Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	0,37	0,70	0,65	-0,51	0,24	0,04	0,29
TDS (g.m^{-3})	4216			3673	4614	4216	330
Mass (g)	1549,3	0,82	0,77	-2300,0	1033,6	154,9	1280,7

Treatment Stage: RO Concentrate

Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	24,43	46,23	43,24	2,07	3,05	2,44	0,33
TDS (g.m^{-3})	8404			7315	9205	8404	664
Mass (g)	205338,4	109,25	102,19	15128,1	27520,1	20533,8	4124,1

Treatment Stage: RO Permeate

Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	28,06	53,08	49,65	1,87	3,67	2,81	0,62
TDS (g.m^{-3})	28			23	39	28	5
Mass (g)	776,5	0,41	0,39	56,3	87,4	77,7	9,9

The accumulated volumes during the 30 hours of operations presented in the flowchart were obtained by mean of the sum of the registered volumes in the hydrometers corresponding to each operation. The TDS concentration values were determined by mean of the average of the TDS concentrations that were read during the operations. The mass values were obtained by mean of multiplying the processed volumes in each stage by the average TDS concentration in the same stage.

The flowchart of Figure 3.2 shows that the total brackish water volume that supplied the system was 56.51 m^3 that corresponded to 100% of the processed volume. From this value, 47.86 m^3 were processed during the 3 hours treatment operations, that is, 84.69% of the processed volume. The remaining 8.65 m^3, that corresponded to 15.31% of the volume, were processed in before the experiment operations, because the permeate volume produced by the UF system was not fully supplying the RO system demand. That is, in this stage of the experiments it was necessary to operate only the UF system previously to the experiments so there would not be a lack of water during the operations.

The average volume of brackish water processed in each operation was 5.65 m^3, considering that from this value, 4.79 m^3 were processed during the 3 hours operations, with values variation between 3.01 m^3 and 6.14 m^3, and has a standard deviation of 1.22 m^3. In operations that were before the experiments an average value of 0.87 m^3 were processed, with value variation between -0.01 m^3 and 1.28 m^3, with standard deviation of 0.39 m^3. The negative value is due the volume variations between the UF tank operations.

The volume values of this stage were accounted with the PAC coagulant addition and the total of PAC added in this stage of the experiments was 15 L.

The average TDS concentration of brackish water was 3,556 $g.m^{-3}$ that corresponded to the average TDS concentration in brackish water processed during the 3 hours operations, also considering the average TDS concentration in the preliminary operations of the UF system. The TDS value in brackish water varied between 3,486 $g.m^{-3}$ and 3595 $g.m^{-3}$. The standard deviation of these values is of 30$g.m^{-3}$.

The total TDS mass coming from brackish water was 200,939.5 g, corresponding to 100% of all the processed mass. From this value 170,181.5 g were provided by the mass that was detected during the 3 hours operation experiments, that is equivalent to 84.69%. The remaining 30756.1 g were provided by the preliminary operations of the UF system that is equivalent to 15.31% of the processed mass.

The average TDS mass of the brackish water processed by the operations during the standard experiments was 17,018.1 g, with value variation between 10649 g and 21,825.4 g with the standard deviation of 4,269.5 g. The average TDS mass of brackish water that was processed during the previous experiment operations was 3,075.6 g, with values between -44.4 g e 4,531.9 g and standard deviation of 1,382 g. The negative value is due to the volume variations between the operations in the UF tank.

The accumulated volume of rejects generated during the UF membrane backwash was 1.27 m^3 that is equivalent to 2.25% of the total volume of processed water. The average volume of reject generated in each operation was 0.13 m^3, with value variation between 0.05 m^3 and 0.18 m^3 with standard deviation of 0.04 m^3.

The average TDS concentration in the reject was 3,550 $g.m^{-3}$, with values between 3,527 $g.m^{-3}$ and 3,584 $g.m^{-3}$ with standard deviation of 20g.m^{-3}.

The discarded mass accumulated during the UF backwash was 4,513.6 g. This value is equivalent to 2.25% of all the processed mass. The average mass discarded in each operation was 451.3 g with value variation from 190.0 to 653.1 g and standard deviation of 137.5 g.

The values that refer to the TDS of “Other UF” were obtained from the average TDS values of brackish water, the permeate and the UF rejects and represented the values that were used in sample collection, leaks and eventual operating errors and system readings.

The accumulated volume that refers to this data was 2.38 m^3 equivalent to 4.21% of the processed volume. The average of the volume of this stage that was processed in each operation was 0.24 m^{-3}, with value variation between -0.30 m^3 and 1,21 m^3 the standard deviation of 0,49 m^3. The negative value is due to the volume variations between the UF tank operations.

The average of TDS that refers to the same data was 3,546 $g.m^{-3}$, with values variation between 3,518 $g.m^{-3}$ and 3,579 $g.m^{-3}$ and standard deviation of 24 $g.m^{-3}$.

The accumulated mass destined to this stage was 8,434.6 g, that corresponded to 4.21% of the mass. The average of mass in each operation of this stage was 843.5 g, with value variation between -1,066.7 g and 4,324.7 g and standard deviation of 1,741.8 g. The negative value is due to the volume variations between the operations in the UF tank.

The permeate volume produced by the UF system was 52.86 m^3. This value corresponded to 100% of the water that supplied the RO system and 93.54% of the total volume that was processed in this stage of the experiments.

The average value of permeate that was processed in each operation was 5.29 m^3 with the value variation between 4.08 m^3 and 6.38 m^3 and standard deviation of 0,79 m^3.

The average TDS concentration in the UF permeate was 3,556 $g.m^{-3}$, with the value variation between 3,517 $g.m^{-3}$ and 3,595 $g.m^{-3}$ and standard deviation of 28 $g.m^{-3}$.

The accumulated volume of mass that was processed by the UF system was 187,956.7 g, equivalent to 93.87% of all the mass that passed through the UF system and to 100% of all the mass that reached the RO system.

The average of the mass that was processed by the UF system in each operation was 18,795.7 g, with the value variation between 14,419.5 g and 22,530.0 g and the standard deviation of 2,738.7 g.

The accumulated volume of concentrate that was generated during the operations and in the CIP by the RO was 24.43 m^3, that is equivalent to 46.23% of the volume that reached it and to 43.24% of the volume that was processed by the system at all.

The average volume of the RO concentrate generated in each operation was 2.44 m^3, with value variation between 2.07 m^3 and 3.05 m^3 and standard deviation of 0.33 m^3.

The average TDS concentration in the RO concentrate was 8,404 $g.m^{-3}$ with value variation between 7,315 $g.m^{-3}$ and 9,205 $g.m^{-3}$ with standard deviation of 664$g.m^{-3}$.

The TDS mass in the RO concentrate generated during the process was 205,338.4 g, equivalent to 109.25% of the mass that entered the RO system and to 102.55% of the entire mass processed by the system. The values above 100% can be explained by the addition of sodium metabisulfite and by the ion exchange of calcium and magnesium by sodium ions during the regeneration processes of the RO system.

The average mass of RO concentrate generated in each operation was 20,533.8 g, with value variation between 15128.1 g and 27520.1 g with standard deviation of 4124.1 g.

The data that refers to “Other RO” in this case are also equivalent values to sample collection and eventual reading and operating errors. The values that refer to the TDS in this stage were obtained from the average values of RO permeate and rejects.

The processed volume in this stage was 0.37 m^3, equivalent to 0.70% of the volume that reached the RO system and to 0.65% of the volume processed in the entire system.

The average of volume that is processed in this stage in each operation was 0.04 m^3, with value variation between -0.51 m^3 and 0.24 m^3 and standard deviation of 0.29 m^3. The negative value is due to the volume that supplied the RO being inferior to the permeate and concentrate volumes produced during this operation.

The average TDS concentration in this stage was 4,216 $g.m^{-3}$, with value variation between 3,673 $g.m^{-3}$ and 4,614 $g.m^{-3}$ with standard deviation of 330 $g.m^{-3}$.

The mass that is attributed to this stage of the treatment was 1,549.3 g, corresponding to 0.82% of the mass that entered the RO system and 0.77% of all the processed mass.

The average mass processed in each operation in this stage was 154.9 g, with value variation between -2,300 g and 1,033.6 g and standard deviation of 1,280.7 g. The negative value is due to the volume that supplied the RO being inferior to the permeate and concentrate values produced in this operation.

The volume of permeate that was produced by the RO was 28.06 m^3, equivalent to 53.08% of the entering value in the RO system and to 49.65% and of all the water that was processed during the experiments. The average amount of permeate produced in each operation by the RO was 2.81 m^3, with value variation between 1.87 m^3 and 3.67 m^3 of standard deviation 0.62 m^3.

The average TDS concentration in the RO permeate was 28 $g.m^{-3}$, with value variation between 23 $g.m^{-3}$ and 39 $g.m^{-3}$, with standard deviation of 5 $g.m^{-3}$.

The mass that was permeate by the RO system was 776.5 g, that corresponded to 0.41% of the mass that entered the RO system and 0.39% of all the mass that was processed during the experiments.

The average RO permeate mass processed in each operation was 77.7 g, with value variation between 56.3 g and 87.4 g the standard deviation was 9.9 g.

Figure 3.3 shows a flowchart with accumulated results obtained in the mass and volume balances of the system for operations with initial brackish water TDS of 7,000 ± 100 $g.m^{-3}$ and details the values obtained during these same experiments.

During the second stage of the experiments the average volume processed in the brackish water stage was 6.36 m^3, from the UF permeate was 6.29 m^3, from the balance of UF permeate used to supply the RO was 0.07 m^3, from the reject generated by the UF was 0.13 m^3, from the RO permeate was 2.68 m^3 and of the concentrate generated by the RO was 2.76 m^3.

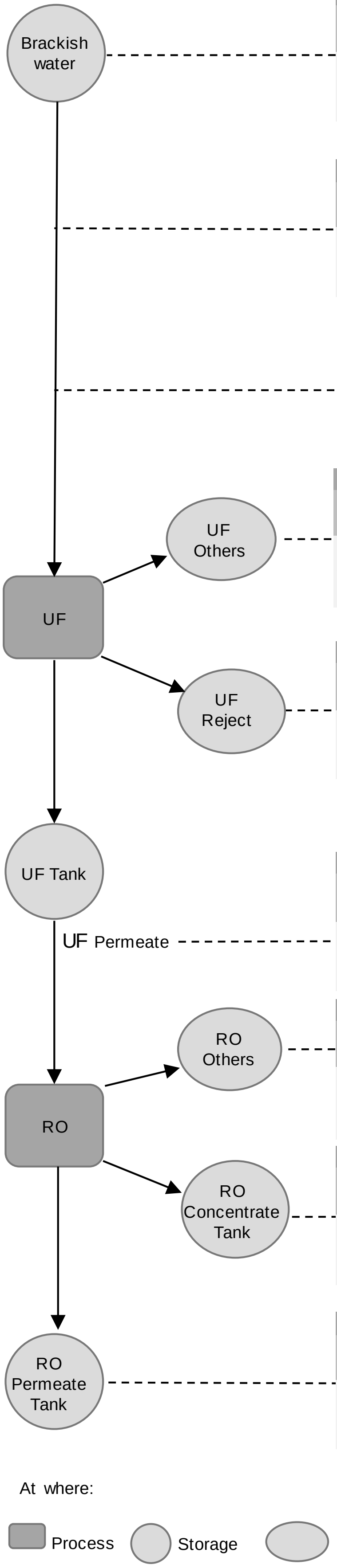
The average flow rate processed per hour in the second stage of the experiments was 2.10 $m^3.h^{-1}$ in the brackish water stage, from the UF permeate it was 1.86 $m^3.h^{-1}$, from the balance of UF permeate to supply the RO it was 0.02 $m^3.h^{-1}$, from the reject generated by the UF it was 0.04 $m^3.h^{-1}$, from the RO permeate it was 0.90 $m^3.h^{-1}$ and from the concentrate generated by the RO it was 0.92 $m^3.h^{-1}$.

Just like it occurred in the first 10 operations, the production of UF permeate was superior to the specified by the manufacturer and the RO permeate production was inferior to the specified by the manufacturer.

The flowchart in Figure 3.3 shows that the accumulated volume of brackish water that supplied the system was 63.60 m^3 that is equivalent to 100% of the volume. Most of this volume 62.88 m^3 were obtained during the operations with the 3 hours duration that is 98.86%. The remaining 0.72 m^3 were the residual volume left in the UF permeate tank between one operation and the next, this residue is equivalent to 1.14% of the brackish water that supplied the system.

The average volume of processed brackish water was 6.36 m^3. Considering 6.29 m^3, with values variation between 5.56 m^3 and 8.36 m^3 the standard deviation was 1.87 m^3, as the processed volume during the 3 hours operations and the remaining 0.07 m^3, with values variation between -0.52 m^3 e 1.12 m^3 with standard deviation of 0.59 m^3 are residual volumes that stayed in the UF tank. The negative value is due to the volume variations between the UF tank operations.

Figure 3.3 – Flowchart of the desalination pilot plant – accumulated results obtained from mass and volume balances with initial TDS concentration of 7,000 ± 100 g.m^{-3}



Treatment Stage: Brackish Water - during the experiments						
Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	63,60	100,00	-0,52	8,36	6,36	3,20
TDS (g.m^{-3})	7041		7018	7091	7040	27
Mass (g)	447842,3	100,00	-3709,2	92747,3	44775,5	31489,7

Treatment Stage: Brackish Water - residual between experiments						
Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	62,88	98,86	5,56	8,36	6,29	1,87
TDS (g.m^{-3})	7041		7018	7091	7040	27
Mass (g)	442755,1	98,86	40397,9	92747,3	44274,7	19879,4

Treatment Stage: UF Tank						
Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	0,72	1,14	-0,52	1,12	0,07	0,59
TDS (g.m^{-3})	7041		7018	7091	7040	27
Mass (g)	5087,2	1,14	-3709,2	7827,3	508,5	4380,5

Treatment Stage: UF Others						
Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	5,83	9,17	-0,06	2,46	0,58	0,83
TDS (g.m^{-3})	7031		7010	7076	7031	26
Mass (g)	41019,8	9,16	-427,6	17309,0	4102,0	6821,7

Treatment Stage: UF Reject						
Parameters	Accumulated Value	% Local	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	1,30	2,05	0,09	0,16	0,13	0,03
TDS (g.m^{-3})	7013		6986	7072	7013	1860
Mass (g)	9138,1	2,04	647,1	1912,8	913,8	481,3

Treatment Stage: UF Permeate							
Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	55,75	100,00	87,64	5,17	6,29	5,57	0,34
TDS (g.m^{-3})	7038			7019	7089	7038	27
Mass (g)	392356,4	100,00	87,61	36593,9	49043,6	39235,6	16583,4

Treatment Stage: RO Others							
Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	1,31	2,35	2,06	-0,36	0,33	0,13	0,24
TDS (g.m^{-3})	7816			7439	8086	7816	224
Mass (g)	10232,1	2,61	2,28	-2821,6	3075,0	1023,2	2027,4

Treatment Stage: RO Concentrate							
Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	27,61	49,53	43,41	2,42	3,27	2,76	0,30
TDS (g.m^{-3})	15564			14811	16109	15564	448
Mass (g)	429713,9	109,52	95,95	35803,4	88246,8	42971,4	19188,3

Treatment Stage: RO Permeate							
Parameters	Accumulated Value	% Local	% Global	Minimum Value	Maximum Value	Average	Standard Deviation
Volume (m³)	26,85	48,16	42,21	2,43	2,87	2,68	0,14
TDS (g.m^{-3})	67			63	69	67	2
Mass (g)	1787,8	0,46	0,40	164,5	381,6	178,8	82,5

The addition of the PAC coagulant was done before the brackish water volume reading. The coagulant addition corresponded to 16 L.

The average TDS concentration in brackish water was 7,041 $g.m^{-3}$ and it was obtained by the average value of TDS concentration of brackish water used during the operations and the residue in the UF permeate tank. The values of TDS concentration in brackish water varied between 7,018 $g.m^{-3}$ and 7,091 $g.m^{-3}$. The standard deviation was 27 $g.m^{-3}$.

In the same form as in the first stage of experiments, the mass calculation of the brackish water that supplied the system was obtained without considering coagulant addition in the water, because the sample collection point was placed before the PAC addition point.

The total mass supplied to the system was 447,842.3 g that is equivalent to 100% of the mass entry. From this value, 442,755.1 g entered the system during the 3 hours experiments, what corresponded to 98.86% of the brackish water mass. The remaining mass that entered the system was the residual mass that stayed in the UF permeate tank between the operations. This value was 5,087.2 g, what is equivalent to 1.14% of the mass that entered the system.

The average TDS mass of brackish water that supplied the system was 44,775.5 g. Considering that 44,274.7 g, with values between 40,397.9 g and 92,747.3 g and standard deviation of 19,879.4 g were processed during the 3 hours operations. The remaining 508.5 g, with value variation between -3,709.2 g and 7,827.3 g with standard deviation of 4,380.5 g are referent to the residual mass in the UF tank UF. The negative value is due the volume variations between the operations in the UF tank.

The accumulated reject volume that was produced by the backwash in the UF system was 1.30 m^3 and corresponded to 2.05% of the volume that entered the system. The average volume of reject generated in each operation was 0.13 m^3, with value variation between 0.09 and 0.16 m^3 and standard deviation of 0.03 m^3.

The average TDS concentration of the reject was 7,013 $g.m^{-3}$, with value variation between 6,986 $g.m^{-3}$ e 7,072 $g.m^{-3}$ and standard deviation of 1,860 $g.m^{-3}$.

The accumulated mass that was discharged during the backwash was 9,138.1 g, value that corresponded to 2.04% of all the mass that was processed by the system.

The average TDS mass of the processed rejects was 913.8 g, with value variation between 647.1 g and 1,912.8 g with standard deviation of 481.3 g.

The “Other UF” values refer to data of sample collections that were realized, leaks and eventual reading and operational errors and were obtained from the brackish water, permeate and UF concentrate data.

The accumulated volume in these items was 5.83 m³, that is equal to 9.17% of the quantity used in the system. The average volume processed between the operations was 0.58 m³ with value variation between -0.06 m³ and 2.46 m³ and standard deviation of 0.83 m³. The negative value is due to the volume variations between the operations in the UF tank.

The average TDS concentration of “Other UF” was 7,031 $g.m^{-3}$, with value variation between 7,010 $g.m^{-3}$ and 7,076 $g.m^{-3}$ and standard deviation of 26 $g.m^{-3}$.

The accumulated mass discarded in this stage was 41,019.8 g that corresponded to 9.16% of the processed mass.

The average mass discarded by backwash in each operation was 4,102.0 g, with value variation between -427.6 g and 17,309.0 g with standard deviation of 6,821.7g. The negative value is due to the volume variation between the operations in the UF tank.

The accumulated volume of the permeate that was produced by the UF system and that supplied the RO system was 55.75 m³, value that corresponded to 100% of the RO backwash RO and 87.64% of the volume processed by the system as whole.

The average UF permeate volume in each operation was 5.58 m³, with value variation between 5.17 m³ and 6.29 m³ with standard deviation of 0.34 m³.

The average TDS concentration of the UF permeate was 7,038 $g.m^{-3}$, with value variation between 7,019 $g.m^{-3}$ and 7,089 $g.m^{-3}$ with standard deviation of 27 $g.m^{-3}$.

The UF permeate mass was 392,356.4 g, value that corresponded to 100% of the mass that entered the RO system and 87.63% of all the mass processed by the system.

The average mass of the UF permeate processed in each operation was 39,235.6 g, with value variation between 36,593.9 g and 49,043.6 g with standard deviation of 16,583.4 g.

The volume of concentrate that was generated during the CIP operations by the RO system was 27.61 m³. This value is equivalent to 49.53% of the volume that supplied this system and 43.41% of the volume that was processed during the entire system.

The average volume of RO concentrate that was processed during the operations was 2.77 m^3, with value variation between 2.42 m^3 and 3.27 m^3 and standard deviation of 0.30 m^3.

The average TDS concentration of the concentrate was 15,564 $g.m^{-3}$, with volume variation between 14,811$g.m^{-3}$ and 16,109 $g.m^{-3}$, with standard deviation of 448$g.m^{-3}$.

The accumulated mass in the RO concentrate was 429,713.9 g, value that is equivalent to 109.52% of all the mass that entered the RO system and to 95.97% of the mass that was processed by the system. This value above 100% is explained by the addition of sodium metabisulfite and by the calcium and magnesium ions exchange for sodium ions during the RO system regeneration process.

The average TDS mass of the RO concentrate that was generated in each operation was 42,971.4 g with value variation between 35,804.4 g and 88,246.8 g and standard deviation of 19,188,3 g.

The “Other RO” data also refer to the sample collection values and eventual reading and operational errors and were obtained from the UF permeate, RO permeate and concentrate data.

The accumulated volume that refer to these data was 1.31 m^3, that corresponded to 2.35% of the volume that supplied the RO system and to 2.06% of the volume that was processed by the whole system.

The average volume of “Other RO” processed during the operations was 0.13 m^3 with value variation between -0.36 m^3 and 0.33 m^3 with standard deviation of 0.24 m^3. The negative value is due to the volume that supplied the RO being inferior to the permeate and concentrate values produced during this operation.

The average TDS concentration in these data was 7,816 $g.m^{-3}$, with value variation between 7,439 $g.m^{-3}$ and 8086 $g.m^{-3}$ with standard deviation of 224$g.m^{-3}$.

The accumulated mass that refers to these data was 10,232.1 g, that is equivalent to 2.61% of the mass that entered the RO system and 2.29% of the mass that was processed in the whole system.

The average mass of these data that was processed during the operations was 1,023.2 g, with the value variation between –2,821.6 g and 3,065.0 g with standard deviation of 2,027.4 g. The negative value is due to the volume that supplied the RO being inferior to the concentrate and permeate volumes produced during this operation.

The accumulated volume of permeate produced by the RO was 26.85 m³, a value that is equivalent to 48.13% of the volume that supplied the system and 42.21% of the volume processed by the whole system.

The average volume of RO permeate in each operation was 2.69 m³, with value variation between 2.43 m³ and 2.87 m³ and with standard deviation of 0.14 m³.

The average TDS concentration of the RO permeate was 67 $g.m^{-3}$, with value variation between 63$g.m^{-3}$ and 69$g.m^{-3}$, with standard deviation of 2 $g.m^{-3}$.

The accumulated mass in the RO permeate was 1787.8 g, that corresponded to 0.46% of the mass that entered the RO system and to 0.40% of all the mass that was processed by the system.

The average mass processed by the RO in the operations was 178.8 g, with value variation between 164.5 g and 361.6 g with standard deviation of 82.5 g.

By mean of the flow tables it is possible to observe that in relation to the volume balance the quantity of brackish water that is demanded in the second stage of the experiments was superior to the first stage, because the necessary volume for the permeate production from brackish water with higher TDS concentration is superior to a brackish water with less concentration of it.

It is also possible to observe that in the first stage of the experiments it would not be possible to operate the UF and RO systems simultaneously, because the production of the first would not be able to provide permeate for the second system. For this reason, it was decided to utilize a chemical wash in the middle of the second stage of the experiments, because with the operations there was an accumulation of solids and consequently clogging of the UF membrane, what resulted in the loss of permeate production by it.

The reject volumes discarded by the UF membrane in the two stages of the experiments were similar. In the first phase the discard was smaller due to the bigger clogging in the UF membrane. The chemical wash caused an increase of permeate and reject production.

The production of permeate by the UF system was superior in the second stage of the experiments, but the RO permeate production in the RO system was inferior in the same stage, because the water that was treated in the second stage of the experiments presented the double of the TDS concentration that the water that entered in the first stage had.

In relation to the concentrate that is generated by the RO production, the second stage produced a higher volume than the first. This volume was superior due to the higher TDS concentration in this stage.

About the TDS concentration in both stages of experiments its remotion happened in the RO system.

The TDS concentration in the concentrate was elevated in both stages of the monitoring. It is observed the efficiency of the RO system with a high salt rejection rate from the Reverse Osmosis membranes causing the saline reject. Other factor that has influence in high the concentration of the concentrate is that in the softening stage happens the calcium and magnesium exchange with sodium ions that lately were removed from the RO membrane.

The average useful power developed is equal to the energetic cost per hour in the first stage of monitoring was 6.65 kW. In the second stage the useful power was 6.32 kW. In relation to the total power installed in the system it consumed 56.3% of the nominal energy to the full power.

The recovery rate of the UF system in the first stage of the experiments was 93.54% and fitted the gap in production capacity informed by the manufacturer. In the second stage of the experiments it was 87.64%, the permeate production capacity was inferior to what the manufacturer informed, that was between 90 and 98%.

The recovery rate from the RO system for the experiments with TDS concentration water of 3,500 ± 100 $g.m^{-3}$ was 53.08%, that is, 16.92% inferior to the recovery rate specified by the manufacturer. For the experiments with TDS concentration in water of 7,000 ± 100 $g.m^{-3}$ the recovery rate was 48.16%, that is, 1.84% inferior to the value specified by the manufacturer. In relation to the total volume that supplied the system, in the first stage, 50% of the supply water was transformed in permeate and in the second stage this value was 42%.

Elasaad et al. developed a research in the village of La Mancalona, in Mexico, where the brackish water from the well with high concentration and hardness and rain water stored in cisterns and surface ponds were treated in a desalination system powered by solar energy with a permeate production capacity of 0.04 $m^3.h^{-1}$. The brackish water from the well had the TDS concentration equal to 2145 $g.m^{-3}$ and the brackish water from the cistern had TDS concentration equal to 69 $g.m^{-3}$. After the desalination process both started to own a TDS concentration of 10 $g.m^{-3}$. The recovery rate to treat brackish water was 33% and to treat water from the cistern was 50% (ELASAAD et al., 2015).

Khanzada, Khan and Davies realized a study in a RO pilot desalination plant powered by solar energy. The plant was installed in the National University of Sciences and Technology, in Islamabad, Pakistan, and had as objective evaluating the performance of pre-treatment technology of two RO membrane models. Between the studied pre-treatment types was the UF and the treated waters had various initial TDS concentrations. The RO membranes used in the study were the Hydranautics model CPA5-LD-4040 and the Filmtec model LC-LE-4040 (KHANZADA; KHAN; DAVIES, 2017).

Utilizing the Filmtec membrane in the TDS concentration of 3,500 $g.m^{-3}$, its rejection rate was 98% and the permeate that was produced presented a TDS concentration equal to 68 $g.m^{-3}$ and using the Hydranautics membrane the TDS rejection rate was 97% and the permeate with TDS concentration of 98 $g.m^{-3}$. For the initial TDS concentration of 4000 $g.m^{-3}$ and using the RO membrane from the brand Filmtec, the rejection rate was 97% and the permeate presented TDS concentration of 110 $g.m^{-3}$ (KHANZADA; KHAN; DAVIES, 2017).

3.4 CONCLUSION

From the mass balance it was possible to observe that the RO permeate represents about 0.40% of the amount of TDS entering the system, that is, the average TDS concentration of the first 10 operations was 28 $g.m^{-3}$ and the last 10 were 67 $g.m^{-3}$.

Based on the volumetric balance it was observed that the permeate production was higher when the initial TDS concentration was lower: 50% of the water supplying the UF-RO set became permeate in the first 10 operations and 42% in the brackish water, which provided the system in the last 10 operations that transformed the permeate.

By means of the volumetric balance for the first 10 experiments, it is possible to observe that from the volume that entered the pilot system, 50% were considered useful, while the volume that provided the system in the last 10 experiments, 42% were considered useful.

The mass balance of this study showed the efficiency of the use of UF as a pretreatment of the RO system. The UF was able to completely remove the suspended solids that provide the color of the water. The turbidity for the most part was also removed by the UF. The RO removed the remaining suspended solids that were not retained by the UF and removed from the TDS.

FINAL CONSIDERATIONS

The first chapter of this work was dedicated to the energetic question of a pilot desalination plant of brackish water inserted in a smart grid system, that used the solar electric energy together with the energy provided by the conventional grid. It was observed that the pilot desalination system demanded electric energy from the distribution grid in superior proportion to the quantity produced by the solar energy system. In relation to energy consumption, the expenses to treat water with initial TDS concentration of 3,500 ± 100 $g.m^{-3}$ and of 7,000 ± 100 $g.m^{-3}$ were similar, being that in the water treatment with lesser TDS concentration the consumption was higher, but some factors happened, such as the prolonged time for maintenance, specific operations before the standardized experiments and others.

The second chapter by its turn was dedicated to the elaboration of mass and volume balances of the same pilot desalination system of brackish water. The system composed by the UF stages, softening and RO can treat water, generating permeate with about 0.40% of the TDS mass from the brackish water input. As expected, when operated with water with a smaller concentration of TDS, the system generated a larger volume of permeate even consuming a smaller volume of supply water.

The benefits of using desalination in the search for potable water are innumerable and allying this method with renewable energy the gains are even higher, both economically and ecologically. Besides the obtained results in this work in relation to the solar energy production not being enough to supply the system, it is necessary to remind that the study was developed in the month with less incidence of solar radiation of the entire year, that is, on the other months the solar electric energy would be bigger. In relation to the mass and volume balances, these showed that the system was capable of treating water producing permeate with quantity and quality predicted by the manufacturer.

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