# ECONOMIC OPTIMISATION OF DUAL PURPOSE, HYBRID POWER-DESALINATION

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**Abstract**

This paper presents an economic optimisation of dual purpose, hybrid power-desalination using a combination of reverse osmosis (RO), membrane technology, and multi stage flash (MSF) distillation technology for desalination and combined cycle power generation technology. The paper uses an integrated system net present value (NPV) approach, in which the dual purpose plant is considered as a single production unit with two revenue streams (power and water), to overcome the problems associated with how to allocate the cost of steam production between the power and desalination plants.

The paper looks at a project required to export 1000 MW of power and 100 MiGD of desalinated water, and investigates the impact of varying both the proportion of RO to MSF, and the number of MSF stages in each MSF unit, on the overall project economics. In each case investigated, the demand for steam and power from the desalination plant is calculated, and then the design of the power plant is adjusted to precisely meet the overall production requirements of 1000 MW export and 100 MiGD.

The paper shows how the optimum number of MSF stages is dictated by the requirements of the power plant, and varies enormously, depending on the ratio of MSF distillate to steam turbine steam flow (equivalent to the power to thermal water ratio). In many cases, the economic optimum number of stages results in a gain output ratio (GOR) which would not meet many consultant specifications, and it is clear that where a consultant specifies a minimum GOR, he removes one of the main optimisation parameters from the process designer, and may force the process designer into a non optimum design.

There are two very significant conclusions from this paper. First, that the ability to manage power to thermal desalination water ratios through the use of hybrid desalination significantly impacts the conventional optimisation guides for thermal desalination processes. And second, that true optimisation of power-desalination processes can only be achieved by consideration of the overall facility, and not by consideration of the power plant and desalination plant in isolation.

CONTENTS

# I. INTRODUCTION

The desalination industry, particularly in the Middle East, is becoming increasingly interested in the use of hybrid desalination, where thermal and membrane technologies are combined. This is evidenced by the recent Fujairah power desalination tender, which was won by hybrid multi stage flash (MSF) and reverse osmosis (RO) technology, against a performance specification which left the technology choice to the tenderer.

The purpose of this paper is to investigate how one of the main design parameters for MSF technology, namely the gain output ratio (GOR), is affected in a hybrid system, by changing the number of MSF stages. In addition to investigating the impact of the number of MSF stages, the paper also looks at differing ratios of MSF to RO, and what effect this has on overall project economics.

Any economic assessment of thermal desalination technology is faced with the key issue of how to cost the steam consumed by the desalination plant. Many methods have been used in the past[1,2,3,4,5], but all of these methods have to contend with the fact that the steam requirement of the desalination plant (flowrate and pressure) can have a very significant effect on the power plant design.

For this reason, this paper considers the combined power and desalination plant as a single, economic facility, which is fed with gas and seawater and outputs two saleable products of power and water. This single facility has capital and operating costs, which are combined with the fuel cost, power & water revenues to calculate the overall project NPV. Since the economic analysis only considers the overall facility, there is no need to calculate the internal cost of transferring steam and electricity from the power plant to the desalination plant.

This study uses process simulation and costing tools for both the desalination plant and the power plant. The MSF process simulator models the process performance (distillate production, steam consumption, power consumption) of an MSF plant for given process conditions. Capital and operating costs are based material quantities.

The RO process simulator[6] models process performance (permeate production, number of elements, feed pressure, recovery etc.) of the RO system. Capital and operating costs are based on material quantities.

The intake & outfall are costed from cost algorithms. It is assumed that the reject cooling water from the MSF plant (or the power plant condenser in the case of 100% RO) is used to feed the RO plant.

The power plant simulator[7] models and costs the power plant. The model accounts for the impact of steam pressure & temperature on the steam turbine and HRSG cost.

## II. STUDY PARAMETERS

### General Parameters 2.1

Table 1 shows the overall parameters used in this study.

| | |
|---|---|
| Seawater Temperature | 33 °C |
| Seawater Salinity | 40 g/kg |
| Gas Cost | US$1.50/GJ |
| Water Tariff | 70 US¢/$m^3$ |
| Power Tariff | 2.5 US¢/kW.hr |
| Water Production | 100 MiGD |
| Power Export | 1000 MW |

**Table 1 General Study Parameters**

### Multi Stage Flash Design Parameters 2.2

The basis for the MSF plant is the 100 MiGD design described by Nada [8], which is used to derive the design information listed in Table 2.

| | Brine Heater | Heat Recovery Stages | Heat Reject Stages |
|---|---|---|---|
| Number of Stages | 1 | Variable | 3 |
| Tube Surface Area | Variable to give a condensing temperature of 117 °C | 6348 $m^2$ | Variable |
| Fouling Factor | 0.3 x $10^{-3}$ $m^2K/W$ | 0.2 x $10^{-3}$ $m^2K/W$ | 0.2 x $10^{-3}$ $m^2K/W$ |
| Tube Length | Variable | 19 m | 19 m |
| Tube Outside Diameter | 29 mm | 29 mm | 29 mm |
| Tube Thickness | 0.9 mm | 0.9 mm | 0.9 mm |
| Tube Material | 66/30/2/2 Cu/Ni | 66/30/2/2 Cu/Ni tube side temperature > 90 °C or 90/10 CuNi | 66/30/2/2 Cu/Ni |
| Number of Tubes | 3,800 | 3,668 | Variable |
| Tube Side Velocity | 2.3 m/s | 2.3 m/s | 2.3 m/s |

**Table 2 Assumed and Derived Tube Properties**

For each case in the study, the following procedure was used to design the MSF plant:

1. Adjust the number of tubes in each heat reject stage until the average velocity leaving the first (hottest) heat reject stage is as required.
2. Adjust the flowrate through the heat reject section (repeating step 2 above for each new value) until the temperature leaving the heat reject section is 40 °C.
3. Adjust the brine heater tube length until the steam condensing temperature is as required.
4. Adjust total number of MSF units to the integer value which gives close to the required total distillate production.
5. Adjust brine heater, recovery & reject section tube number, brine recycle and reject flowrates by target capacity / total capacity so that the total MSF production capacity is as required.
6. Record cost and operating data for financial analysis.

## Reverse Osmosis Design Parameters 2.3

The basis for the design of the reverse osmosis system is shown in Table 3. Pre-treatment for the reverse osmosis system comprises screening, chlorination-dechlorination, acidification, coagulant dosing and single stage, pressure filtration. The system includes a partial second pass, where necessary, to achieve a blended water chloride concentration not exceeding 250 mg/l (this being the World Health Organisation guideline value).

| | First Pass | Second Pass |
|---|---|---|
| System Flux | 14 l/(m$^2$.hr) | 30 l/(m$^2$.hr) |
| Feed Pressure | 67.4 Barg | 8.7 Barg |
| Maximum Lead Element Flux | 26 l/(m$^2$.hr) | 51 l/(m$^2$.hr) |
| Element Type | Hydranautics SWC3 | Hydranautics ESPA4 |
| Recovery | 42.5% | 85% |
| Elements per Pressure Vessel | 6 | 6 |
| Average Element Age | 3 years | 3 years |
| Salt Passage Increase Factor | 10% per year | 0% per year |
| Flux Decline Factor | 9% per year | 0% per year |
| Maximum Permeate Chloride Concentration | 500 mg/l | 50 mg/l |

**Table 3 Reverse Osmosis System Design Parameters**

## Power Plant Design Parameters 2.4

The basis of the power plant design is combined cycle blocks, with each block comprising 2 General Electric GE9FA gas turbines, 2 heat recovery steam generators (HRSGs), and a single, backpressure steam turbine, with a single pressure, non re-heat steam cycle. The gas turbine air inlet includes a fogger, to increase the gas turbine output. However, for systems with a higher low pressure steam requirement, it was necessary to use the smaller, GE6FA gas turbine, which allows for more HRSGs, and hence more duct firing. For very high steam demands, the GE6FA-HRSG arrangement was used to raise low pressure steam for the desalination process directly, without going through a steam turbine.

| | |
|---|---|
| Ambient Air Temperature | 40 °C |
| Maximum Steam Temperature | 550 °C |
| Maximum Duct Firing per HRSG | 150 MWth LHV |
| Relative Humidity | 40% |
| Ambient Air Pressure | 1.013 Bara |
| Steam Pressure Leaving Power Plant | 0.2 Bara above condensing pressure in brine heater |
| Condensate Return Temperature | 5 °C below condensing temperature in brine heater |
| Proportion of Condensate Returned | 99% |
| Make-up Water Temperature | 40 °C |

**Table 4 Power Plant Design Parameters**

The following procedure is used to design the power plant:

1. Input steam pressure and flowrate requirement from the desal plant.
2. Calculate the net power requirement for the power plant (power export + desal plant power consumption).

3. With full GT fogging, adjust HRSG duct firing until the required steam mass flowrate is achieved, using the maximum high pressure steam temperature and the greatest turbine inlet pressure which results in dry steam leaving the steam turbine.
4. If the net power requirement is not met, increase the number of power blocks.
5. If the net power requirement is exceeded, reduce the degree of fogging until the net power requirement is matched.
6. If the net power requirement is exceeded with no fogging, reduce the steam turbine inlet pressure until the net power requirement is matched, adjusting the steam temperature to maintain dry, saturated steam at the steam turbine outlet.
7. If the net power requirement is exceeded with the minimum steam pressure, turndown the GT.
8. If the maximum duct firing per HRSG is exceeded, increase the number of power blocks.

**Financial Parameters 2.5**

In this study, all options are assessed using an overall project Net Present Value (NPV), with all costs discounted to the project commencement date. A typical power-desalination project takes 2 or 3 years to build, and for this study, it is assumed that the project takes 3 years from project commencement date to commercial operation, and that the capital cost is spent at a constant rate during these 3 years. Construction interest is paid in each year on half of the capital spent in that year as well as all capital and interest spent in previous years.

It is assumed that annual operating costs, fuel cost, power and water tariffs remain constant (no inflation and no demand growth/decline).

All annual costs (during construction and operation) are discounted to the mid-point of that year.

Based on the above, it is possible to calculate NPV multipliers for both capital, operating costs and revenues. In this manner, the NPV is simply calculated by multiplying the total capital cost (power plant and desal plant) by the capital cost multiplier and taking this away from the total annual net operating revenue (water revenue + power revenue – fuel cost – fixed power & water operating costs – variable power & water operating costs).

| | |
|---|---|
| Construction Interest Rate | 8% |
| Construction Period | 3 Years |
| Proportion of Equity | 30% |
| Interest Rate Charge on Equity | 15% |
| Proportion of Debt | 70% |
| Interest Rate Charged on Debt | 6% |
| Overall NPV Discount Rate | 8.7% |
| Operating Period | 20 Years |
| Capital Cost Multiplier | 0.99 |
| Operating Cost/Revenue Multiplier | 7.57 |

**Table 5 Financial Parameters**

## III. RESULTS

Figure 1 shows the overall project NPV plotted against the total number of MSF stages (recovery and reject) for varying capacities of MSF output. In each case, the total distillate production capacity is 100 MiGD, and the difference between the MSF capacity and the total capacity is made up of RO. In the case of 100% RO, there is a single point, which is plotted as a horizontal line against all values of MSF stages.

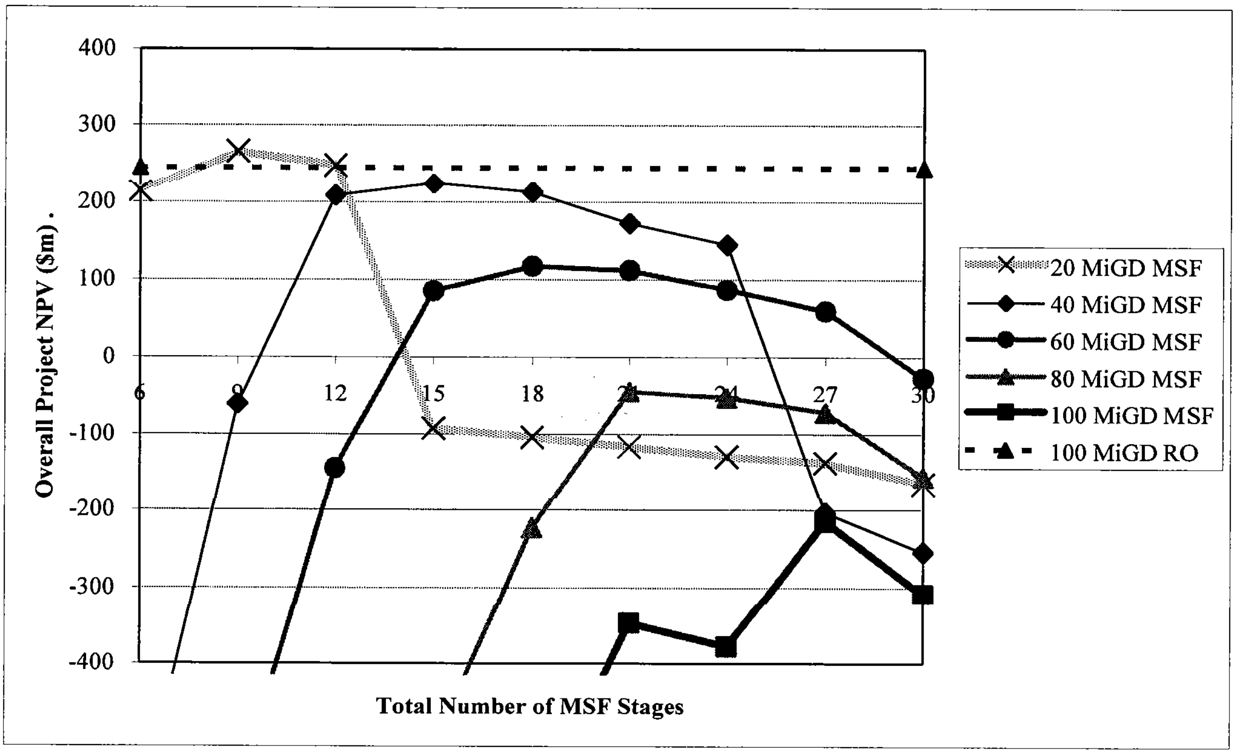


**Figure 1 NPV as a Function of Proportion MSF and Number of MSF Stages**

Figure 1 shows that there is an optimum number of MSF stages, and that this optimum number of stages is greater for higher proportions of MSF.

Figure 2 provides a breakdown of the MSF performance parameters as a function of the number of MSF stages. The data is for the 40 MiGD MSF case. This figure shows that increasing the number of stages reduces the LP steam consumption. This is because the greater heat exchange area provided by the extra stages allows the system to operate with a smaller temperature driving force, hence allowing a closer approach between the temperature of the tube side brine leaving the hottest heat recovery stage and the brine entering the first flash stage. This, in turn, reduces the heat load required of the brine heater.

The figure also shows a virtually proportional relationship between the LP steam consumption and the reject flow per stage. This is because the reject flow per stage is the heat sink for the LP steam supply, and the temperature rise in this heat sink is constant (temperature leaving final heat reject stage is fixed at 40 °C). It should be noted that one of the design requirements for the MSF system was to have a constant tube velocity in the heat reject section. Therefore, increasing the reject flow per stage requires the number of tubes to be increased proportionately. As a result, the number of reject tubes, and the cost of the reject section, decreases with increasing total number of stages.

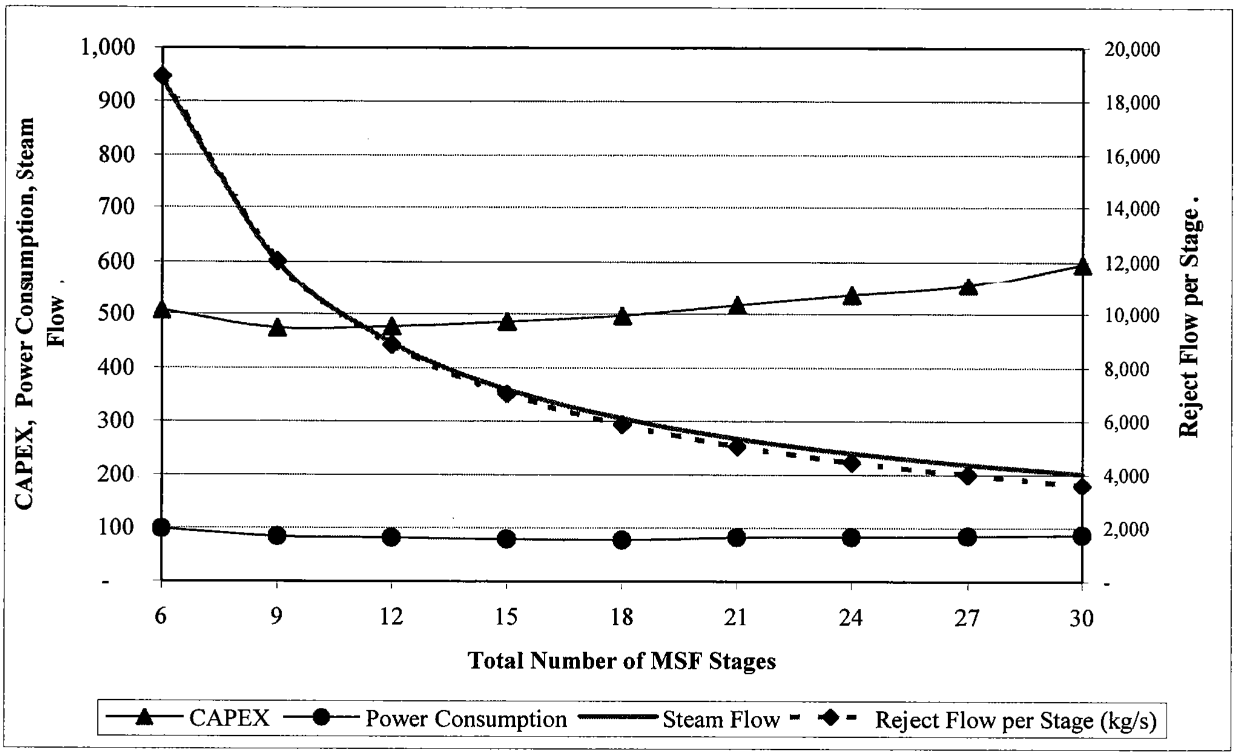


**Figure 2 MSF Main Performance Parameters as a Function of Number of Stages**

The capital cost of the heat recovery section is proportional to the number of stages. However, for very low numbers of stages, the capital cost of the reject section and the seawater intake/outfall system are more important than the cost of the heat recovery section, and increasing the number of stages reduces overall capital cost. This is true upto a total number of stages of about 9. For greater number of stages than 9, the capital cost increases of providing more recovery stages are greater than the savings in reject tubes and intake/outfall systems.

A similar optimum point can be seen in the power consumption. By far the biggest power consumers on an MSF plant are the seawater feed pump and the brine recycle pump. The power consumption of the seawater intake pump is proportional to the reject flowrate, hence increasing the total number of stages reduces the intake pump power consumption. On the other hand, the brine recycle flowrate is constant and the brine recycle pump power consumption is proportional to the brine recycle pump delivered pressure. Thus the brine recycle pump power consumption is proportional to the number of heat recovery stages. Hence increasing the number of recovery stages increases the brine recycle pump power consumption. Figure 2 shows that for increases in the total number of stages upto about 15, the power savings from the seawater intake pump are greater than the increases in the brine recycle pump power consumption, and the overall power consumption reduces. However, for greater than 15 stages, the brine recycle pump power consumption increases are more significant, and the overall power consumption increases.

Because the power export quantity is defined (as 1000 MW in all cases), and the desalination plant power consumption is less than 10% of the exported power, as indicated in Figure 2, the generating capacity for the power plant is approximately constant in all cases. As a result, the NPV of the power plant, and consequently the overall NPV, is greatly affected by the LP steam flow, which explains why the optimum number of MSF stages increases with increasing MSF distillate production.

Figure 3 shows how the specific capital cost and energy efficiency of the power plant are related to the total low pressure steam flowrate. The figure also indicates the design of the power plant in each region of the plot, by reference to Table 6.

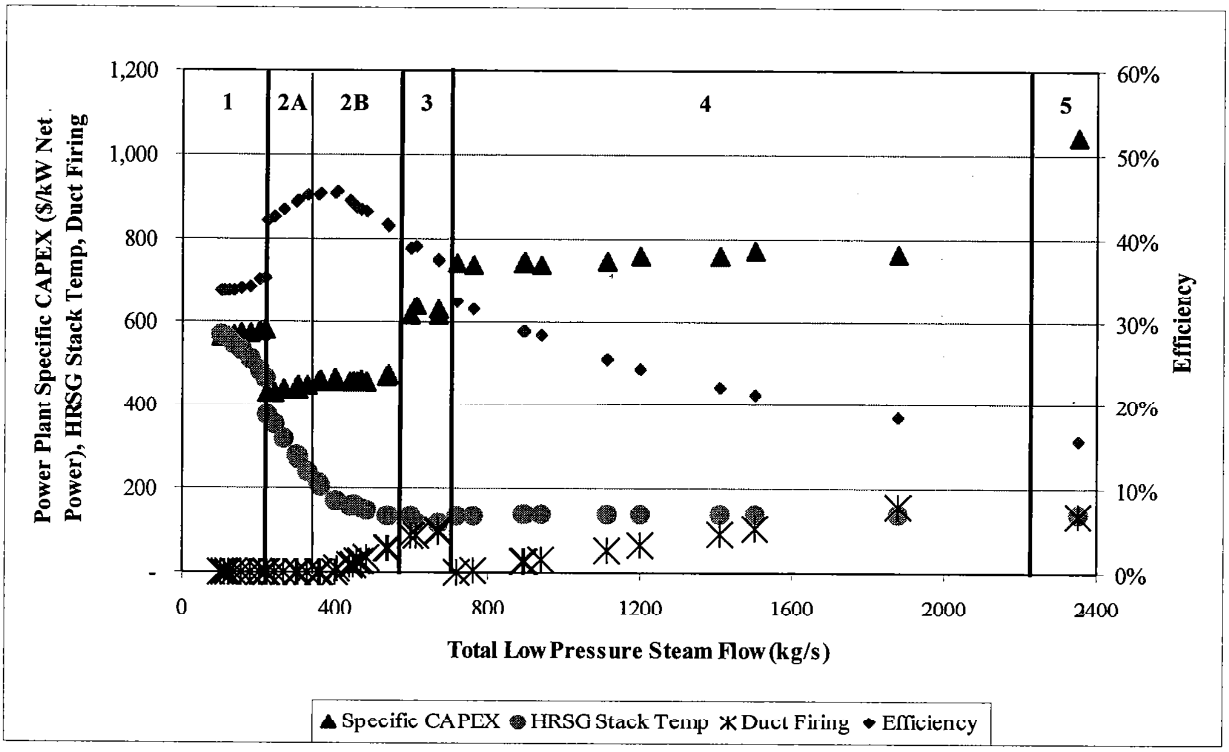


**Figure 3 Power Plant Design Parameters as a Function of LP Steam Flow**

| Region | Power Plant Configuration | Power to Steam Ratio Control |
|---|---|---|
| 1 | GE9FA; 3 Blocks; 2-2-1 | GT Turndown |
| 2A | GE9FA; 2 Blocks; 2-2-1 | Fogging |
| 2B | GE9FA; 2 Blocks; 2-2-1 | ST Inlet Temperature/Pressure |
| 3 | GE6FA; 5 Blocks; 2-2-1 | Fogging |
| 4 | GE6FA; 17 Blocks; 1-1-0 | Fogging |
| 5 | GE6FA; 24 Blocks; 1-1-0 | GT Turndown |

**Table 6 Power Plant Configurations in Figure 3**

The optimum power plant configuration investigated in this study was 2 blocks of 2:2:1, based on the GE9FA gas turbine. This is evidenced from Figure 3, which shows both the lowest specific CAPEX and the highest efficiency from this configuration. This configuration is valid for steam flows ranging from 223 to 548 kg/s. For the lower steam flows (223 to 329 kg/s), control of the balance between power and steam is achieved by adjusting the level of fogging the gas turbines. By increasing fogging, the gas turbine power output is increased, reducing the power, and thereby the LP steam flow, required from the steam turbine. However, as the steam flowrate requirement is further reduced, there comes a point at which the power demand cannot be met, even with the maximum gas turbine fogging, because not enough power can be generated in the steam turbine. At this point, it is necessary to add extra gas turbine generating capacity, or in other words, provide an additional power block. This is required for steam flows of 220 kg/s and less.

The two blocks of 2:2:1 configuration, using GE9FA gas turbines can supply upto 548 kg/s of LP steam. Steam flows of 360 to 548 kg/s can be achieved by reducing the steam turbine inlet temperature/pressure, without any fogging of the gas turbine. Reducing steam turbine inlet temperature/pressure reduces the power that can be extracted for a given steam flow as well as reducing the enthalpy at which the steam is generated in the HRSGs. Consequently, the available HRSG heat is capable of generating more steam, and less power is extracted from that steam.

It can be seen from Figure 3 that for steam flows upto 401 kg/s, there is no HRSG duct firing, and there is a steady increase in plant efficiency, and specific CAPEX with increasing steam flow. Below 401 kg/s, there is more heat available in the gas turbine exhaust than is required for generating steam, and this results in excess heat being vented up the power plant stack, as evidenced from the HRSG stack temperature in Figure 3. This is an obvious waste of energy, which results in reduced system efficiency. Once the steam demand is such that HRSG duct firing is required, the HRSG stack temperature is fairly constant, being dictated by the minimum pinch between the boiler and the HRSG gas side. However, as steam flow increases, as already mentioned above, the steam turbine inlet temperature/pressure is reduced to balance the power and steam requirement. This reduction in steam turbine inlet temperature/pressure reduces the overall system efficiency. In addition, increasing duct firing to reduce the proportion of power generated from the gas turbines also reduces efficiency.

There is, therefore, a very significant optimum point for the power plant, at which there is no duct firing, and no venting of waste heat up the HRSG stack. In Figure 3 it can be seen that this optimum point occurs when the steam turbine inlet temperature/pressure is being reduced to control power to steam demand. The power plant would actually be better optimised if this point occurred in the fogging control region, which would require a slightly higher power export demand, and since fogging is a cheap way of providing additional power capacity, without an efficiency penalty, the overall optimum point occurs with the maximum level of fogging.

Where more LP steam is required than can be provided by two blocks of 2:2:1 with GE9FA gas turbines, it is necessary to increase the proportion of power generated by duct firing and steam turbines. This can be achieved by providing smaller gas turbines and more HRSGs. Thus for steam flows in the range 600 to 670 kg/s, the gas turbine model is changed to the GE6FA, whilst maintaining the 2:2:1 configuration. In these cases, 5 blocks are required to provide the necessary power export. There is a major CAPEX penalty with changing to this configuration, and since the level of duct firing is being increased with increasing steam flow, efficiency falls with increasing steam flow. The steam flow which can be achieved using this configuration is limited by the maximum level of HRSG duct firing.

Once steam demand increases beyond 720, it is not possible to satisfy steam demand with the 5 blocks of GE6FA 2:2:1 configuration because of HRSG duct firing limits. In fact, the steam demand is so great that it is impractical to generate power from this steam, but rather, it makes more sense to generate low pressure process steam directly in the HRSGs. This is done in a simple 1:1 configuration, using the GE6FA gas turbine, and duct firing to achieve the necessary steam flow. The total power export requirement can be achieved using 17 gas turbines in this configuration, which can cater for steam flows from 720 to 1880 by adjusting duct firing. It should be noted that increasing duct firing dramatically reduces system efficiency (since no power at all is generated from duct fired gas), and also increases specific CAPEX (more HRSG surface area is required to raise the additional steam).

Once the maximum steam with this configuration is reached (due to maximum limit on HRSG duct firing), it is necessary to provide more GT-HRSGs, but to turndown the GTs to meet the power and steam demands. This configuration has a very high specific CAPEX, and also a very low efficiency.

To demonstrate the significance of the optimum power plant design point on the overall system optimisation, Figure 4 plots the low pressure steam demand as a function of number of MSF stages for the different MSF/RO combinations. The optimum power plant steam demand of 402 kg/s is shown as a horizontal line in this plot.

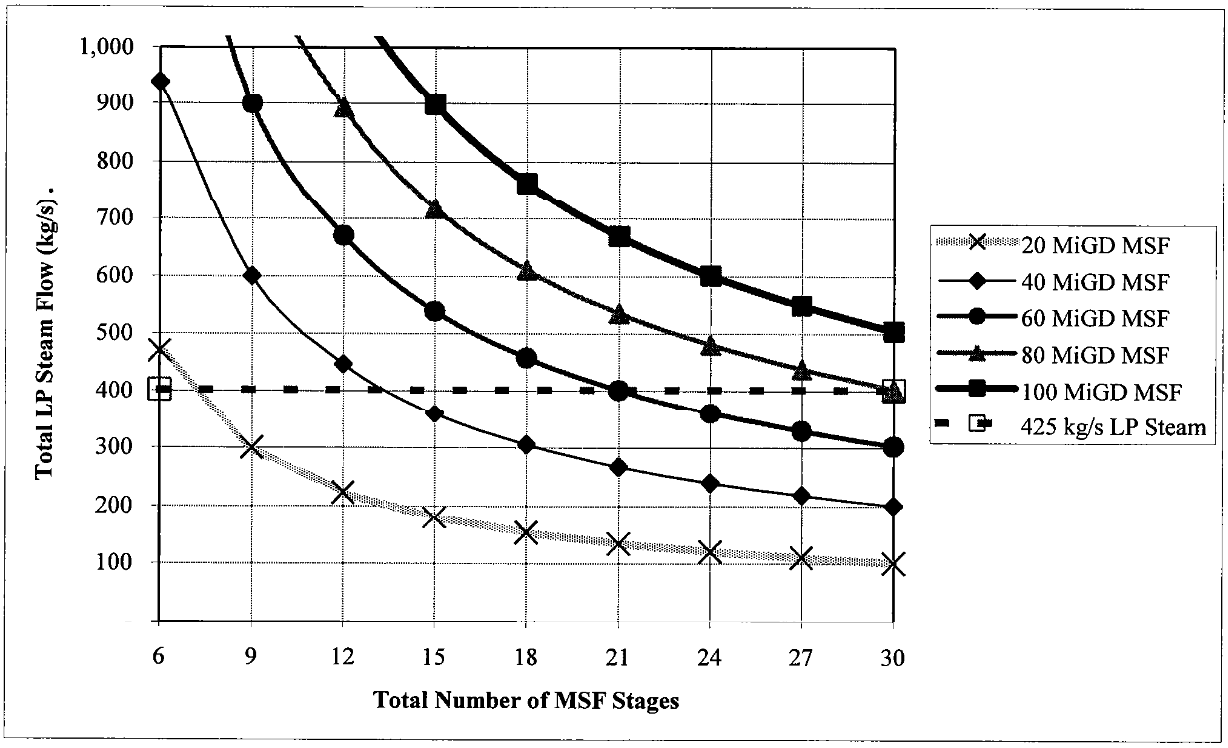


**Figure 4 Total LP Steam Flow as a Function of MSF Stages & Proportion of MSF**

From Figure 4, the optimum number of MSF stages required by the power plant can be read as the point at which the 402 kg/s steam line crosses the steam flow curve. These values are given, along with the optimum number of stages from Figure 1, in Table 7.

| MSF Capacity | Optimum Number of Stages from Figure 1 | Optimum Number of Stages from Figure 4 |
|---|---|---|
| 20 MiGD | $9 \pm 2$ | 7 |
| 40 MiGD | $15 \pm 2$ | 13 |
| 60 MiGD | $18 \pm 2$ | 20 |
| 80 MiGD | $21 \pm 2$ | 28 |
| 100 MiGD | $27 \pm 2$ | >30 |

**Table 7 Comparison of Optimum MSF Stages from Power Plant Optimum Steam Flow and Overall NPV**

Table 7 clearly demonstrates good agreement between the number of MSF stages required to give the optimum power plant low pressure steam flow, and the number of MSF stages which gives the optimum overall project NPV, although the power plant does tend to over-estimate the number of stages at high numbers of stages. This is because, as the number of stages increases, the capital cost of unit reduction in steam demand increases.

## IV. DISCUSSION

Because power to water ratios are generally set for the overall desalination and power plant, the use of hybrid technology allows process designers to adjust the power to MSF water ratio (that between total power production, and water production from the MSF process). Thus a 50:50 hybrid plant doubles the effective power to water ratio for the MSF plant.

Non hybrid power-MSF plants will typically have minimum GORs specified, and a typical minimum value is 8 kg distillate per kg steam. This is a particularly useful requirement if the MSF plant is being tendered in isolation, since it enables the power plant designer to know what the low pressure steam demand is going to be. However, in the case of a hybrid system, specifying the MSF GOR greatly restricts the number of viable hybrid configurations, and may well prevent some of the greatest cost saving potentials of the hybrid system being realised.

Where a client wishes tenderers to use their process expertise to provide the lowest whole life project cost for a combined power-desalination project, he should limit his performance specification to the total requirements for power and water, and be clear about the evaluation criteria (capital cost, operating cost, NPV etc.). Specifying a minimum GOR for a hybrid system effectively specifies a minimum proportion of MSF. However, where a client requires a minimum proportion of MSF, this should be specified explicitly, and the process designer should be free to select the optimum GOR.

## V. CONCLUSIONS

- There is an optimum combination of low pressure steam flowrate and power export requirement for any power plant configuration (gas turbine type and number).
- The optimum point for a combined cycle power plant is the point at which there is no venting of waste heat up the HRSG stack and no duct firing.
- The optimum power to steam ratio for a given configuration is given by maximum gas turbine fogging combined with no duct firing and no waste heat up the HRSG stack.
- The optimisation of the power plant dominates the optimisation of the desalination system.
- The ability to manage power to thermal desalination water ratios through the use of hybrid desalination significantly impacts the conventional optimisation guides for thermal desalination processes.
- True optimisation of power-desalination processes can only be achieved by consideration of the overall facility, and not by consideration of the power plant and desalination plant in isolation.

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