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ECONOMICS OF DOWNHOLE SEPARATION OF GASES

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ABSTRACT

This paper quantifies the economic advantages and drawbacks of downhole separation of non-condensable geothermal gases. Using details of the performance of the downhole separator (DHS) and estimates of the costs of placement and workovers a realistic assessment can be made.

Such an assessment was carried out for a (hypothetical) 50 MWe power plant at Ngawha geothermal field. The use of the DHS gives rise to a reduction in the cost of gas extraction equipment. However high costs for workovers on DHS wells far outweighs the savings which result from using the DHS. The likely cost could be as high as $2,452,000 p.a. Even given (impossible) optimistic circumstances there would still be a nett cost of about $1,420,000 p.a. over the life of the project. The DHS is not economically viable for New Zealand geothermal fields.

INTRODUCTION

The concept and means for downhole separation of non-condensable geothermal gases was conceived by DSIR Chemistry Division who developed the Downhole Separator (DHS) to the point of testing a laboratory scale air/water model. The next stage in DHS development called for testing using geothermal fluids in an above ground test rig. Construction and operation of the test rig was funded by the Electricity Division, Ministry of Energy and progress on the DHS development has been achieved by a joint effort of Chemistry Division, Electricity Division, Ministry of Works and Development (Wairakei Project) and the Geothermal Institute. MWD constructed the geothermal test rig and provided the engineering input on DHS placement and well workovers. The Geothermal Institute assisted with the experimental work and contributed to co-ordination of the DHS project.

The preceding papers have discussed aspects of the DHS project. Kanyua (1) treated the fluid mechanics theory of the DHS. Grant-Taylor (2) looked at the chemical aspects of DHS operation and the experimental confirmation of DHS performance. Hattersley (3) discussed methods and costs for placing the DHS downhole. The final aspect, to be treated in this paper, is the evaluation of overall economics. This will be done taking the real performance of the DHS and by using the DHS costs quantify the economic advantages and drawbacks of using the DHS.

This paper considers the economics of using the DHS for a (hypothetical) 50 MWe power plant at Ngawha. There are (presently) only two high gas Liquid dominated geothermal fields in New Zealand. At Ngawha, gas in the total fluid is about 1 to 2 percent by weight. The gas content of the steam construction and tenders have been let for gas exhausters; Thus the DHS might only be applicable to Ngawha (in New Zealand).

ESSENTIAL DETAILS OF PERFORMANCE, PLACEMENT AND WORKOVERS

The essential parameters describing DHS performance are the vapour and liquid separation efficiencies. The vapour separation efficiency describes the proportion of vapour (existing in the DHS) which is separated to the gas pipe. Similarly the liquid separation efficiency is the proportion of liquid (existing in the DHS) which is separated to the annulus. In their experimental work Grant-Taylor and Kanyua (4) observed vapour separation efficiencies of about 90% but noted that this parameter was sensitive to changes in volumetric flow rate. For the purpose of this economic assessment a time averaged vapour separation efficiency of 80% will be adopted - this allows for the MIS to operate off optimum for part of the time. Liquid separation efficiency was found to be very good with about 99.8% of the liquid phase going to the annulus in some cases. To represent time averaged performance it is assumed that 0.5% of the liquid phase is entrained in the gas pipe flow.

Hattersley (3) has described the method which was proposed for DHS downhole placement. The essential details are that it appears practicable to place the DHS downhole and that the additional cost (over a normal well) is $182,000. Subsequent workovers would cost $220,000 (in addition to the normal cost of a workover). These are the estimated costs as at April 1984 which are based on MWD experience with the dual completion at well NG 9, Ngawha.

ADVANTAGES OF USING THE DHS

By reducing the amount of non-condensable gas reaching the power plant there will be a reduction in the capacity of gas extraction equipment. This will give both capital and operating and maintenance savings. Additionally reduced parasitic power requirements would result. For this economic evaluation it is assumed that gas extraction will be by centrifugal exhausters - this is the method adopted for Ohaaki Power Station.

METHODOLOGY FOR EVALUATION OF ECONOMICS

The approach adopted in this paper is to evaluate the economics for an "average" case and then look at the effect of off-average parameters on the overall economics. This is intuitively satisfactory because of the natural variation in well output and gas content and the trade-offs which might result. From a production pint of view a large well output is attractive since more power can be yeneratad for the same DHS costs. However, for a given gas content a same DHS costs and this is less attractive. Lower gas content means lower gas pipe loadings but is tending towards low gas situations where conventional gas extraction systems will be more suitable.

Some consideration needs to be given to the economic consequences of the decline in well output between workovers. The presence of the gas pipe in the well casing will affect the well output due to increased resistance to fluid flow. The gas pipe flow will be controlled by the desired well head pressure and the gas pipe site.

THE AVERAGE CASE FOR NGAWHA
PHASE FLOWS

Of the wells drilled at Ngawha, the maximum flow, enthalpy and well head pressure for NG 4, NG 8, NG 9, NG 11, NG 12, and NG 13 suggest that these wells would be suitable to supply a power plant. From the data on these exploratory wells (5), it is a straight forward matter to calculate mass weighted averages of maximum flow, enthalpy, and gas content:

Maximum flow = 360 t/h
Enthalpy = 970 kJ/kg
Gas Content = 1.3 weight % of total flow

Assuming that the non-condensable gas is wholly carbon dioxide, the component flows are:

Total gas flow = 0.013 x 360 = 4.68 t/h of CO_2
Total water flow = 360 − 4.68 = 355.3 t/h

At the DHS it is necessary to have a vapour fraction that is sufficient to ensure a reasonable proportion of the gas resides in the vapour phase. Too low a vapour fraction would mean that the non-condensables are retained in the liquid phase defeating the purpose of the DHS. For the average case, take the DHS operating pressure to be 21 bars. (All pressures in this paper-arcabsolute.) For the water component of the well flow, the vapour fraction (i.e. steam fraction) is:

$$x_{steam} = (970 - 920)/1880 = 0.0266$$

The phase flows of water (assuming phase equilibrium prevails) are:

Water vapour flow = 355.3 x 0.0266 = 9.45 t/h
Liquid water flow = 355.3 − 9.45 = 345.85 t/h

The distribution of carbon dioxide between the phases depends on equilibrium as well as kinetics but it is assumed that CO_2 is in equilibrium for the DHS operating temperature of 215°C. At this temperature the distribution coefficient is 3.72×10^{-3} (4) i.e.

$$3.72 \times 10^{-3} = \frac{\text{kmoles of } CO_2 \text{ in liquid phase/kg liquid}}{\text{kmoles of } CO_2 \text{ in vapour phase/kg vapour}}$$

A CO_2 balance can be made:

let C_1 = holes of CO_2 in vapour (one hour basis)
and C_2 = kmoles of CO_2 in liquid

then

$$\frac{C_1}{\text{kg vapour}} = 269 \frac{C_2}{\text{kg liquid}}$$

and $C_1 + C_2 = 4.68 \times 10^3/44 = 106.36$ kmoles

thus $C_1 + (C_1 \times 345.85)/(9.45 \times 269) = 106.36$

$\therefore$ $C_1 = 93.63$ kmoles (i.e. 4.120 t/h) and
$C_2 = 12.74$ holes (i.e. *0.560* t/h)

As a result the phases consist of:

Liquid phase: liquid water 345.85 t/h
gas 0.56 t/h
346.41 t/h

The flows of fluid to the annulus and the gas pipe can now be calculated using the vapour and liquid separation efficiencies. The annulus flow will be:

209 of vapour phase + 99.5% of liquid phase

gas=0.2 x 4.12	+ 0.995	x	0.56	=	1.38 t/h
liquid water=0.2 x 0	+ 0.995	x	345.85	=	344.14 t/h
water vapour=0.2 x 9.45	+ 0.995	x	0	=	1.89 t/h

Total annulus flow = 347.41 t/h

Similarly the gas pipe flow will be:

80% of vapour phase + 0.5% of liquid phase

gas = 3.30 t/h) Total gas pipe
liquid water = 1.73 t/h) flow = 12.59 t/h
water vapour = 7.56 t/h)

(As a check the total flow is

347.41 + 12.59 = 360.00 t/h)

At later stage it will be necessary to use estimates of the enthalpy of each stream. Since water predominates, the water enthalpy of each stream is used. For the DHS operating pressure of 21 bars, h_f = 920 kJ/kg and h_g = 2800 kJ/kg

For the annulus flow:

$$h_{ann} = (344.14 \times 920 + 1.89 \times 2800)/(344.14 + 1.89) = 930 \text{ kJ/kg}$$

For the gas pipe flow:

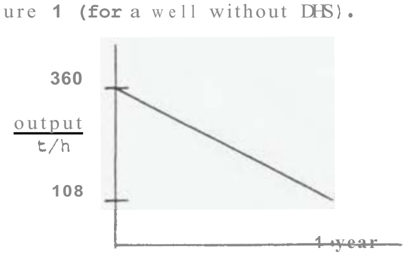
$$h_{gp} = (1.73 \times 920 + 7.56 \times 2800)/(1.73 + 7.56) = 2450 \text{ kJ/kg}$$

FACTORS AFFECTING WELL OUTPUT

The foregoing calculations are based on the maximum output of the "average" well. However, two factors will significantly affect the well output.

By introducing a gas pipe into the well casing there is a reduction in area available for flow. Experience with NG 9 (3) suggests that for a 3½" gas pipe well output is likely to be reduced by 309. For the "average" well then the maximum flow would be 360 x 0.70 = 252 t/h.

Mineral deposition is expected to be a major problem at Ngawha (5). Indications from the short term discharge of NG 9 are that the well production rate will drop by 700 after one year of continuous operation. The time dependence of such decline in output will determine the cumulative mass discharge and hence the annual kWh which the well could produce. This economic assessment assumes that the decline curve is straight line to 300 over one year as in figure 1 (for a well without DHS).



The measure of the power producing capability of a well is the equivalent continuous flow rate. For a non-DHS well this is 360(1 + 0.3)/2 = 234 t/h. (This represents the field and time averaged output of a well).

If mineral deposition were to occur above the DHS then the decline would also be straight line. However, it is envisaged that most deposition would occur below the DHS (6). In this case well output would be reduced as the mineral deposition becomes as dominant as the added well friction due to the gas pipe. For a well with a DHS the output might tend flat then drop off to become the usual straight line decline. How long would the well produce at 252 t/h until calciting reduces the output? It is assumed that mineral deposition is proportional to the cumulative flow. A non-DHS well will decline to 252 t/h in ((360 - 252)/(360 - 108) =) 0.4286 yr. The cummulative flow is 1.149 Mt. For the DHS well the same amount of calciting occurs in $(1.149 \times 10^6/(252 \times 8760) =)$ 0.52 yr. At this point the decline is taken as straight line - figure 2.

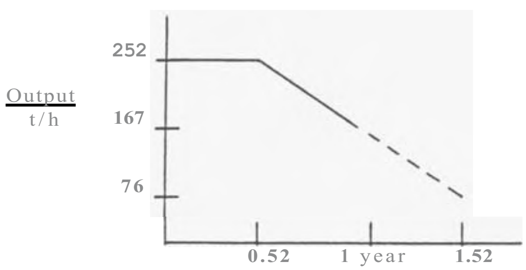


Fig 2

The equivalent continuous flow for a DHS well is calculated as 231.7 t/h.

GAS PIPE PRESSURE DROP

The gas pipe well head pressure will be determined by the flow and the gas pipe size. It will also depend on the nature of the flow in the gas pipe. For the maximum flow in a DHS well (252 t/h) the gas pipe flows would be:

gas	3.30 x 252/360 = 2.31 t/h
liquid water	1.73 x 252/360 = 1.21 t/h
water vapour	7.56 x 252/360 = 5.29 t/h

When the gas pipe pressure drops to say 18 bars the phase flows would (assuming no heat transfer) be:

$$h_{gp} = 2450 \text{ kJ/kg} = 885 + 1912 \times X$$
$$X = 0.819$$

$$\therefore \text{ vapour flow} = 0.819 \times 6.50 + 2.31 \times 18/44 = 6.27 \text{ t/h}$$

(the CO_2 is represented by a water vapour equivalent)

$$\text{and liquid flow} = (1 - X) \times 6.50 = 1.18 \text{ t/h}$$

For a gas pipe size of 3½" Sched 40 the inside diameter is 90.1 mm and the cross-sectional flow area is 0.006376 m^2. The superficial velocities are:

liquid: $V_{sl} = 1.18/3.6 \text{ kg/s} \times 1 \text{ m}^3/856 \text{ kg}/0.006376 \text{ m}^2 = 0.06 \text{ m/s}$

vapour: $V_{sg} = 6.27/3.6 \text{ kg/s} \times 0.1104 \text{ m}^3/\text{kg}/0.006376 \text{ m}^2 = 30.14 \text{ m/s}$

and the flow is annular. As the pressure drops the flow will tend to become more like single phase (vapour) flow. In this paper it is assumed that the

Using the equation for steam flow in horizontal pipes it is possible to estimate the gas pipe pressure drop by:

$$P_{WHP}^2 = P_{DHS}^2 - KW^2LP_{DHS}/\rho_{DHS}$$

where L = gas pipe length (m)
W = vapour mass flow (t/h)
K = "friction" factor for the pipe corrected for P_{DHS}, ρ_{DHS}
P = pressure (b abs)
ρ = vapour density (kg/m^3)

for 3½" Sched 40, $K = 4.6 \times 10^{-3}$

Thus for a vapour flow equivalent to 6.27 t/h of water and P_{DHS} = 21 bars

$$P_{WHP}^2 = (21)^2 - 4.6 \times 10^{-3} \times 6.27^2 \times 500 \times 21/10.512 = 260.4$$

gives P_{WHP} = 16.14 bars

(For 3" Sched 40, K = 0.0103 then P_{WHP} = 6.0 b. If 7.5 b (abs) is adopted as the minimum required gas pipe well head pressure then the gas pipe must be 3½" size).

(Note that it is acceptable to use the formula for horizontal steam flow because friction is the dominant tern:

$$\text{take } \rho_{ave} = (10.512 + 8.084)/2 = 9.30 \text{ kg/m}^3$$

$$\rho gh = 9.30 \text{ kg/m}^3 \times 9.81 \text{ m/s}^2 \times 500 \text{ m} = 0.456 \times 10^5 \text{ N/m}^2 = 0.46 \text{ bar (gravitational term)}$$

The reader may like to confirm that the momentum change term is smaller than the gravitational term and that the vapour velocity is always well below sonic velocity.)

TURBINE GENERATOR PERFORMANCE

At this stage it is necessary to consider the performance of a turbine generator on geothermal steam. An allowance can be made for the amount of non-condensable gases which will alter turbine performance.

For Ngawha, downhole temperatures are about 225°C. For a condenser condition of 45°C the optimum flash temperature is about 135°C. (Silica saturation temperature is about 130°C). For this analysis a separation pressure of 5 bars (abs) is adopted. Turbine inlet conditions are taken as 4 b (abs) and saturated conditions.

For the water component of the geothermal steam:

h inlet = 2739 kJ/kg (4 b, 143.6°C)

S inlet = 6.897 kJ/(kg.K)
= Sexhaust for N_T = 1
= 0.649 + X1 x 7.500
gives X_1 = 0.833

$\therefore h_1 = 192 + X_1 \times 2392 = 2185$ kJ/kg (0.1 b, 45.8°C)

$\Delta h_1 = 2739 - 2185 = 554$ kJ/kg

The isentropic efficiency of the turbine is assumed to be 74% (for a large single pressure unit).

Δh actual = 0.74 x 554 = 410 kJ/kg

and X actual = 89.3% (which is satisfactory)

For the gas component it is assumed that this is all CO_2, that the CO_2 acts independently of the

CO_2 inlet state is 4 b (abs), 143.6°C ≡ 58 psia 290°F

CO_2 exhaust state is 0.1 b, 45.8°C ≡ 1.5 psia 114°F

From a pressure–enthalpy diagram for CO_2 (7)

$$\left.\begin{array}{l} h\ \text{inlet} = 215\ \text{Btu/lb} \\ h\ \text{outlet} = 177\ \text{Btu/lb} \end{array}\right\}\ \Delta h = 38\ \text{Btu/lb} = 88\ \text{kJ/kg}$$

The turbine output can then be calculated from the known steam composition. The turbine generator unit losses (both mechanical and electrical) are taken as 2% of the gross turbine mechanical work. Thus for Z gas mass fraction the electrical output will be:

Power = 0.98 x Flow (kg/s) x (88 Z + 410 (1 − Z)) kWe

USES FOR THE GAS PIPE STREAM

This economic assessment looks at two uses for the gas rich stream from the gas pipe. The first method is to take the gas streams from the wells to a central small back pressure turbine generator exhausting to atmosphere. An alternative would be to use the gas stream energy to flash the separated water producing more steam for the (main) turbine generator. Consideration will be first directed towards the back pressure turbine option.

POWER OUTPUT FOR DHS CASE

For the average well fitted with a DHS, the fluid flowing in the annulus would be:

Gas: 232/360 x 1.38 = 0.89 t/h
Water: 232/360 x 346.03 = 223.00 t/h

For the separation pressure of 5 bars, the steam fraction is:

X = (930 - 640)/2109 = 0.1375

If all the gas goes to the vapour phase then the geothermal steam is:

Water vapour = 0.1375 x 223.00 = 30.65 t/h
gas = 0.89 t/h

∴ geothermal steam = 31.54 t/h
and gas content Z = 0.89/31.54 = 0.0282 kg/kg

The turbine output would thus be 3443 kWe. However allowance must be made for gas extraction parasitic power − this is estimated (pro-rata) from figures for Ohaaki Power Station (8) to be 80 kWe. Thus the nett output from the annulus flow is 3443 - 80 = 3363 kWe.

The fluid flowing in the gas pipe would be:

Gas: 232/360 x 3.30 = 2.124 t/h
Water: 232/360 x 9.29 = 5.979 t/h

It is assumed that the gas pipe discharge from each well would be taken to a single location and that the pipelines would be sized so that the liquid phase (present at the well head pressure) is separated and drained in condensate pots along the pipeline. It is assumed that the pressure drop along the line is 1.5 bars and that the heat loss from the pipeline matches the tendency to superheat as the pressure drops i.e. the gas pipe flow that reaches the turbine is saturated vapour at 1.5 bars below the wellhead pressure. For the average case P_{WHP} = 16.14 bars and the steam fraction is 0.8222, then the vapour at the well head is:

5.979 x 0.8222 + 2.124 = 7.040 t/h

h_{inlet} = 2791 kJ/kg

s_{inlet} = 6.452 kJ/(kg/K)
= s_{outlet} for N_t = 1
= 1.361 + 5.937 X_1
X_1 = 0.858 (for 1.2 b)

∴ h_1 = 439 + 2244 X_1 = 2363 kJ/kg

Δh_{actual} = (2791 − 2363) x 0.72 (isentropic efficiency)
= 308 kJ/kg

For the CO_2 component (the temperature and pressure is determined by the water component which dominates):

h_{inlet} = 235 Btu/lb (14.64 b, 197°C)
h_{outlet} = 200 Btu/lb (1.2 b, 105°C)
Δh_{actual} = 35 Btu/lb = 81.4 kJ/kg

The power output from the gas pipe stream is thus:

0.98 x (2.124 x 81.4 + 5.979 x 0.8222 x 308)/3.6 = 460 kWe

(this is based on the equivalent continuous output).

The combined output of the main turbine generator and the high gas turbine generator (per well) is 3363 + 460 = 3823 kWe

POWER OUTPUT FOR NON-DHS CASE

For the average well without DHS the component flows would be:

Water = 234/360 x 355.3 = 230.95 t/h
Gas = 234/360 x 5.68 = 3.04 t/h

The steam fraction at the separator plant is:

X = (970 - 640)/2109 = 0.1565

and the geothermal steam is:

$$\left.\begin{array}{l} \text{water vapour} = 0.1565 \times 230.95 \\ \qquad = 36.14\ \text{t/h} \\ \text{gas} = 3.04\ \text{t/h} \end{array}\right\}\ \begin{array}{l} 39.18\ \text{t/h} \\ Z = 7.76\% \end{array}$$

The gross turbine power would be 4106 kWe and after deducting parasitic power of 272 kWe, the nett (non-DHS) power is 3834 kWe.

COSTS OF POWER PLANT ITEMS

Before going on to the cost comparison it is necessary to calculate the cost of gas extraction equipment and turbine generators. Data supplied by an equipment manufacturer (9) has been fitted to approximate expressions for the cost of the plant items immediately prior to devaluation. These are:

Exhausters Cost = \$1.207 x 10^6 x √(Dry Gas Flow in t/h)
Turbine generators Cost = \$1.015 x 10^6 x √(MWe)

(These fit limited data in the range up to 40 t/h and 50 MWe)

COST COMPARISON

This economic assessment compares costs on an annual basis assuming a 10% required rate of return and a 25 year economic life. Power plant size is 50 MWe. Maintenance and operation are charged at 5% of the capital cost per annum.

The gas flow is (13.04 x 3.04 =) 39.7 t/h and the cost of gas exhausters would be $7,605,000. The annual cost of gas extraction is thus:

$7,605,000 x (0.11017 + 0.05) = $1,218,000 p.a.
(Note that parasitic power requirements have been included by using the nett non-DHS power output).

For the DHS case:

No. of wells = 14 (50000/3823 = 13.08)

The gas flow would be (13.08 x 0.89 =) 11.63 t/h giving an annual cost for gas exhausters of $659,000 p.a. This is a saving of $559,000 p.a. compared to the non-DHS case.

The size of the main turbine generator can be reduced from 50 MWe to (13.08 x 3363 =) 43989 kWe resulting in an annual saving of:

$$\$1.105 \times 10^6 \times (\sqrt{(50)} - \sqrt{(44)}) \times 0.16017$$
$$= \$71,000 \text{ p.a.}$$

However, a back pressure turbine generator is required to generate (13.08 x 460 =) 6010 kWe. As an estimate of the cost, a MHI Modular 5 back pressure unit would cost about $2.55 x 10^6 (10) (pre-devaluation) giving an annual cost of $409,000 p.a.

The other major cost item which needs to be included is the annual cost of DHS placement and workovers. For 14 wells this gives an annual charge of:

14 x (182,000 x 0.11017 + 220,000)
= $3,361,000 p.a.

The costs and savings can be summarised as follows:

Item		Annual Savings	Annual Costs
(i)	Gas Exhausters	$559,000	-
(ii)	Main Turbine Generator	$71,000	-
(iii)	Gas Turbine Generator	-	$409,000
(iv)	DHS Placement and Workovers	-	$3,361,000

The nett annual cost of using the DHS is $3,140,000 p.a. and this arises primarily as a result of the cost of DHS well workovers.

EFFECT OF OFF-AVERAGE CONDITIONS

The above costs and savings arise for the "average" case. In what situations would it be possible to achieve a lower annual cost and how much lower would it be? To answer this each variable was independently varied to find the value which gave the lowest annual nett cost for operating a DHS. In practice the variables are unlikely to be independent but by this approach a fictitious "Best Case" giving lowest annual cost can be found.

A computer program was written to perform all of the above calculations and allow easy and accurate variation of the main parameters. The variables tested were as follow:

Variable	Units	Range Tested	"Bert Case"	Typical Ngawha Values
Well Output	t/h	250 - 600	550	100 - 540
DHS Pressure	bars	19 - 24	23	
Gas Content	wt%	0.3 - 2.0	1.70	0.91 - 1.70
Well Decline		0.3 - 3.7	0.35	0.3 (expected)
Vapour Separation	-	0.6 - 3.9	0.9	
Liquid Separation	-	0.98 - 3.998	-	
Enthalpy	kJ/kg	960 - 990		900 - 990
DHS Depth	m	400 - 550		-

For the "Best Case" the annual cost of using the

PRODUCTION OF ADDITIONAL STEAM

The other use for the gas pipe stream is flashing of separated water to provide additional steam to the turbine generator. What is the maximum amount of steam which could theoretically be produced from the separated water? To answer this it is necessary to note the gas pipe stream conditions. At the well head the details are:

Water = 5.979	t/h	hgp	= 2450 kJ/kg
Gas = 2.124	t/h	P_{WHP}	= 16.14 b
		T_{WH}	= 202°C

The separated water is available at a temperature of 152°C (5 bars flash pressure). The latent heat of vaporisation is 2109 kJ/kg. The maximum heat transfer would be for the gas pipe stream cooling to 152°C and saturated liquid i.e. hgp = 640 kJ/kg. For the water only the heat transferred is:

$$Q_{max} = 5.979/3.6 \times (2450 - 640) = 3006 \text{ kWth}$$

If however, heat transfer effectiveness is taken as 80% then the actual heat transfer can be calculated:

$$0.80 = (202 - t)/(202 - 152)$$
$$\therefore t = 162°C \text{ and } h_f = 684 \text{ kJ/kg}$$

$$Q_{actual} = 5.979/3.6 \times (2450 - 684) = 2933 \text{ kWth}$$

The extra steam raised would be (2933/2109 x 3.6 =) 5.0 tfh per well. In the main turbine this would generate 560 kWe. (Note that even this is optimistic since pressure and heat losses are not allowed for in this calculation).

The nett output for the DHS ("average") case is thus (3363 + 560 =) 3923 kWe.

COST COMPARISON

It is now necessary to evaluate the costs involved in this option to determine the nett cost of using a MIS. At this stage the cost of heat exchange equipment is neglected.

The number of wells required is 13 (50000/3923 = 12.75). The costs and savings will be:

Item		Annual Savings	Annual Costs
(i)	Gas Extraction	$559,000	-
(ii)	DHS Placement and Workovers	-	$3,121,000
(iii)	Fewer Wells	$110,000	-

The nett annual cost is $2,452,000 p.a.

(For the "Best Case" the nett annual cost was calculated to be $1,420,000 p.a. for a 50 MWe power plant.)

CONCLUSION

This paper has looked at the economics of using the DHS for a power plant at Ngawha. It has been shown (2) that the DHS can achieve good separation of the vapour and liquid phases flowing in a geothermal well and that non-condensable gases reaching the turbine condenser can be markedly reduced. The practicability of placing the DHS has been discussed (3) and it is possible to install the DHS in a geothermal well.

By using the DHS it is possible to achieve a reduction in the cost of purchasing and maintaining gas extraction equipment. However the high cost of well workovers (which would be required annually)

For a 50 MWe power plant at Ngawha the most optimistic situation is for an annual cost for using the DHS of $1,420,000 p.a. The likely cost would be about $2½ million pet annum (based on the average conditions at Ngawha).

It is conceivable that the MIS could be (relatively) more attractive if fluid enthalpy were higher and workovers were required less frequently. However the DHS would have to show a definite advantage over conventional methods of gas extraction before it could be adopted. In this event a more refined economic assessment would be required.

There may well be other applications for the DHS concept. The device was conceived for geothermal developments but this should not detract from its application elsewhere. This is especially so because the separator works very effectively.

ACKNOWLEDGEMENT

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