DOE/CE/15446--T5

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DE92 002637

HEAVY OIL RECOVERY PROCESS

CONCEPTUAL ENGINEERING OF A DOWNHOLE METHANATOR AND PRELIMINARY ESTIMATE OF FACILITIES COST FOR APPLICATION TO NORTH SLOPE ALASKA

CONTRACT:DE-FG 01-89 CE 15446

FINAL REPORT

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Dated:10/31/91

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ABSTRACT

The West Sak (Upper Cretaceous) sands,overlaying the Kuparuk field at 2,000-4,500 ft depths,contains heavy oil (12-23 API) at a low GOR. With a range of OOIP estimates of 13 to 25 Billion Barrels,this reservoir,covering 300 sq.miles (777 km2),would rank among the largest known oil fields in the US,but technical difficulties have so far prevented its commercial exploitation.Steam injection is the most successful and the most commonly-used method of heavy oil recovery,but its application to the West Sak presents the following major problems:

1)steam injected from the surface would have to traverse the 2,000 ft (600m)-thick Permafrost layer.Unavoidable heat losses from the steam and production tubings have two detrimental effects:

a) the injected steam quality is greatly reduced,and

b) melting the Permafrost around the cemented casing could cause the well to sink,with potentially disastrous effects;

2)the reservoir consists of 6 major layers,relatively thin and of medium permeability (ca.150 md);

3)the West Sak formation contains some swelling clays,which,when

exposed to steam condensate, further reduce the rock permeability;

4) high well rates, generally required for economic operations in the Arctic, make it imperative to prevent degradation of permeability.

Such difficulties may be overcome by using a novel approach, in which steam is generated downhole in a catalytic Methanator, from Syngas made at the surface from endothermic reactions (Table 1). The Methanator effluent, containing steam and soluble gases resulting from exothermic reactions (Table 1), is cyclically injected into the reservoir by means of a horizontal drainhole while hot produced fluids flow from a second drainhole into a central production tubing. The downhole reactor feed and BFW flow downward in two concentric tubings. The large-diameter casing required to house the downhole reactor assembly is filled above it with Arctic Pack mud, or crude oil, to further reduce heat leaks.

A quantitative analysis of this production scheme for the West Sak required a preliminary engineering of the downhole and surface facilities and a tentative forecast of well production rates. The results, based on published information on the West Sak, have been used to estimate the cost of these facilities, per daily barrel of oil produced. A preliminary economic analysis and conclusions are presented together with an outline of future work. Economic and regulatory conditions which would make this approach viable are discussed.

INTRODUCTION

Published information on the West Sak reservoir, gathered from Ref. 1 to 3, is shown on Table 2. A location map of the field is shown on Fig. 1, where existing roads, sea docks and major flow lines for the Kuparuk and Milne Point fields are also indicated The main characteristics of a 19 API sample of West Sak oil (taken from Ref. 4)

and the results of sand pack displacement runs (Ref.5) of this oil by various fluids,including steam and gas mixtures are summarized in Table 3.Specific relative permeability data are not yet available for the West Sak.From this limited amount of information,only a rough assessment of the technical challenge presented by the exploitation of this very large oil resource can be made:1.The low oil gravity and the low reservoir temperature,together with the medium range of reservoir permeability result in low oil mobility; 2.The relatively small (20-40 ft) thickness of each of the 6 main layers separated by impervious shales is another handicap (see Appendix 1).

ARCO's pilot waterflood,terminated in 1985,has shown that,despite low well productivities (ca.130 B/D/well),in the deeper part of the field,where the oil is lightest,it may be sufficiently mobile to be displaced by water.In most of the reservoir,however,a recovery process resulting in a large increase in oil mobility will be required.

Miscible displacement processes are rarely applicable to heavy oils and the laboratory tests with CO_2-rich injectants presented in Ref.2 and 4 have shown that reservoir plugging by asphaltene precipitation from West Sak oil is likely to offset any benefits of viscosity reduction resulting from mixing with the injected solvent.

A more conventional way of increasing the mobility of heavy oils is through the application of steam heat ("huff and puff"and steamflood). This is a proven technology,responsible for most of the oil production in California.Its implementation in the Arctic,however,presents a very difficult challenge because of the risk of melting the Permafrost wherever heat losses may occur from boilers,steam lines,steam tubings and production tubings carrying hot oil and condensed steam.Preventing any significant melting of the Permafrost around the cemented casings

is a major concern in the design of any Arctic oil well completions. This becomes even more important for thermal wells,where the fluid temperature levels greatly exceed those of naturally flowing wells. Effective removal of the bond between cement and melting Permafrost could result in sinking of the wells,with potential failure of the tubular goods.With a solution GOR of about 250 SCF/B (45 m3/m3) and reservoir pressure exceeding 2,200 psi (15 MPa) in some of the wells, such a failure might cause a potentially catastrophic well blow out.

In addition,any heat loss from the steam prior to its arrival at the well bottom reduces the quality of the injected steam,thus reducing its effectiveness in the reservoir.To eliminate this loss is a major objective of this novel approach,in which the required thermal energy is transported from the surface to the downhole Methanator in chemical form,without significant loss.

BACKGROUND

Laboratory displacement experiments by steam in synthetic porous media (Ref.5) have shown that recoveries of West Sak oil exceeding 75% are achievable and that the addition of an oil-soluble gas (CO_2) further improves the recovery through a combination of various mechanisms: additional viscosity reduction and swelling of the oil in the coldest part of the displacement front,solvent extraction and enhanced steam distillation at the condensing steam front.

The risk of excessive corrosion rates by co-injection of CO_2 with condensing steam at high temperature was,however,demonstrated by the various field tests of combustor-type downhole steam generators (Ref.6),but field experiments with alternate injection of Natural Gas and steam (Ref.7) and laboratory tests (Ref.8) have shown similar

benefits, without the corrosion risk. Similar displacement experiments of West Sak oil by steam and natural gas mixtures (Ref.9) have shown that Methane concentrations of 20 and 25% in the steam/gas mixture improve oil recoveries by 12.5% and 10.4% respectively over that of pure steam at 600 F (315 C). An additional benefit of co-injection of Methane with steam is that produced Methane contributes to lifting the oil and water phases in the production tubing.

Before considering the injection of steam in any specific heavy oil reservoir, it is prudent to check whether the screening criteria are satisfied. A most important parameter is the transmissibility. A minimum transmissibility of 100 md.ft/cp (30 md.m/cp) is required for a steam flood to be successful, according to Taber and Martin's EOR screening guides (Ref.10). Based on the published average characteristics of the West Sak, this threshold is reached for a minimum layer thickness of 23 ft, provided that permeability remains at or close to its original value. This requires that no solids (asphaltenes, coke) be derived from the oil nor any swelling of the reservoir clays be caused by contact with condensed steam. The possible use of conventional steam injection in this heavy oil reservoir is thus made contingent upon maintaining the original transmissibility. The screening guides in the National Petroleum Council EOR study for thermal processes (Ref.11) have later reduced the transmissibilty threshold for steam injection to only 5 md ft/cp, but, considering that reservoir permeability may be lower than the average value (150 md) and that the oil viscosity may be higher than 35 cp in many parts of this very large field, it is still clear that any significant permeability reduction should be prevented for steam injection to be successful in the West Sak.

MAJOR BENEFITS OF NEW APPROACH

To meet the challenge of the West Sak,a novel approach to steam injection has been studied in sufficient detail to estimate its cost and benefits.To eliminate heat losses at the surface and through the tubing,high quality steam is generated downhole.It is mixed with hot gases (Methane also generated downhole,Hydrogen and small amounts of Ammonia gas) prior to injection into the reservoir.

The effects of Methane are expected to be similar to those obtained in the experiments of Ref.9,and analogous to those of CO_2 addition to steam (Ref.5),namely oil viscosity reduction,oil swelling and enhanced effects of steam distillation of the West Sak oil.In Ref.18,R.M.Butler provides the means to evaluate quantitatively the contribution of steam distillation to oil recovery,but this requires information on the reservoir and on distillation curves for West Sak oil containing variable amounts of Methane,which are not available in the published literature.

The effect of Hydrogen is expected to be the prevention of coking or asphaltene precipitation in the reservoir rock,caused by thermal cracking of West Sak oil.Experimental results presented in Ref.12 show that when heavy oils are heated in the presence of rock,steam.water and Hydrogen for periods ranging from 1 to 10 days,significantly less coke is produced and oil recoveries by gravity flow are systematically higher than in the absence of Hydrogen.By analogy with known processes used for refining heavy crudes and residues,these experimental results suggest that,over the long term and in the presence of natural catalysts (clays and other minerals) present in most reservoir rocks, the well-known reactions of Hydrodesulfurization and Hydrocracking become predominent over those of thermal cracking,in the range of

temperature and pressure considered here.The injection of Hydrogen together with steam is intended to prevent plugging of the reservoir rock by oil-derived solids.Specific measurements on representative West Sak core material exposed to stem/hydrogen mixtures for periods of several months are required to confirm the absence of any permeability damage by coke.The feasibility and general results of such long-term experiments,however,was recently demonstrated at LBL on preserved core material from another North Slope Cretaceous heavy oil reservoir,the Ugnu,which overlays a portion of the West Sak.

The effect of NH3,when dissolved in the steam condensate is to form NH4+ cations,which,by exchange with those originally present in the swelling clays of the West Sak formation,will prevent the clays from swelling,thus contributing to maintain the reservoir permeability at or near its original value.This is a well-known phenomenon,but the specific concentration of NH3 required for stabilizing the West Sak clays remains to be determined experimentally on representative core samples of the reservoir.

Other effects of the non-condensible gases in the injectant are to reduce the partial pressure of steam and its condensing temperature. This is expected to increase the rate of expansion of the zone contacted by steam in the reservoir,to a level comparable with that of pure steam injected in shallower reservoirs,thus making the oil displacement process comparable with that of the most successful steam injection projects in California.

A major concern in any thermal project is to be able to deliver heat to the reservoir at a sufficiently high rate.In many cases of low Kh reservoirs,this is accomplished by fracturing the formation.In the present case,it is done by generating high quality steam downhole,as

close to the reservoir as possible and by distributing the injectant through horizontal drainholes extending about 3,000 ft (900 m) into the lower part of each layer.The technical feasibility of drilling and completing horizontal wells in poorly consolidated sands has been demonstrated in Ref.19,where the required drilling mud characteristics are also indicated.

These multiple drainholes,shown on Fig.7,are connected to a single vertical well containing respectively the downhole reactor-steam generator,the concentric tubings feeding the generator and also the single production tubing carrying to the surface the fluids produced from at least one of the drainholes,while the steam/gases mixture is injected into at least one other drainhole.A downhole valve section operated from the surface allows to switch each drainhole from the injection mode to the production mode,in succession,thus alternately operating each drainhole in "huff and puff".Methods and equipment required for the drilling,completion and operation of such a vertical well and multiple drainholes are described in Ref.20.

To provide a high heat rate,not only does the access to the sand face have to be sufficient,but the downhole generator capacity must also be comparable with that of the largest oil field steam generators (> 200 MMBTU/h or > 59 MW).In this respect also,the proposed concept clearly departs from that of all previously tested combustor-type downhole steam generators which had a maximum capacity of only 7.1 MMBTU/h.The total volume of our generator is,of course,much larger.It is essentially a fixed bed catalytic Methanator,cooled by water tubes similar in design to those of a conventional natural circulation boiler.It receives a Syngas feed (H_2,CO,CO_2) and a boiler water feed containing Ammonia,all prepared in conventional facilities at the

surface. It discharges an effluent consisting of steam, Methane, Hydrogen and NH3 into each of the drainholes through the Downhole Valve Section (see Appendix 2).

DESIGN CONCEPT

It is generally recognized that, to offset the high construction and labor costs prevailing on the Alaskan North Slope, the following principles determine the economic viability of petroleum development projects in that area: 1) large reserves, high well productivities and wide well spacings are required,

2) economies of scale in the facilities are the key to achieving acceptable capital and operating costs,

3) design optimization is an essential step to improve the project economics while minimizing operational hazards and risks to the fragile Arctic environment.

4) to the maximum possible extent, proven designs and technologies are preferable. This does not preclude the use of new or emerging technologies, if they have been previously tested elsewhere or if they are fully understood from an engineering point of view. There are many examples of such applications of new technology to the resolution of specific problems on the North Slope. The TAPS pipeline design itself is unique, and the use of horizontal wells at an early date at Prudhoe Bay provides another example of such novelties.

The development of the proposed concept followed those principles: Large catalytic Methanator reactors are used in many industrial plants and the design of water tubes boilers is fully understood. The process chemistry of steam reforming, Methanation, Water Gas shift, etc... is well defined by accurate equilibrium constants in the range of pressures

and temperatures of interest.Suitable materials and catalysts are commercially available which have demonstrated their long-term reliability at the industrial scale in many operating plants.

The high efficiency of transporting thermal energy over long distances by conversion of Natural Gas into Syngas (with endothermic reactions) followed by pipeline transport of the Syngas and final release of the thermal energy by Methanation (with exothermic reactions) has been demonstrated at the industrial scale in Germany in the early 1970's as part of the ADAM/EVA project built by Lurgi and Haldor Topsoe (Ref.13)

The design of the wells and that of the surface units was based on the experience acquired in building similar facilities on the North Slope and elsewhere on the same scale,to get full benefit of economies of scale and to arrive at realistic cost estimates..

Full optimization of such a large project,however,could not be carried out without more detailed information,only available to the West Sak oil owners.The extensive effort of engineering optimization of the most recently developed field on the North Slope,the Endicott field,has shown that large savings could result from such an effort,if it is undertaken in the future.

PROCESS DESCRIPTION

A schematic process flow diagram is shown on Fig.2.The feed to the steam reformer furnace is assumed to be residue gas from the Prudhoe Bay NGL/EOR plant.The produced Syngas provides heat to a waste heat HP boiler and,after partial CO_2 removal,is compressed for injection into a pair of wells equipped with Downhole Methanators/ steam generators.The compressor is driven by a back-pressure steam

turbine.The turbine exhaust provides process steam for the reformer and for BFW preparation and for the CO_2 removal unit.The CO_2 is recovered and compressed for export as a miscible solvent.Auxiliary facilities are shared among four such trains.They include a Liquid N_2 plant,emergency generator and start-up boiler.They are integrated with a small Ammonia synthesis plant (Casale process) deriving its H_2 feed from a slip-stream of compressed Syngas (see Fig.3).Ammonia is dissolved in the BFW stream.

Each Downhole Methanator assembly is hung in a large-diameter cased well and connected to the surface by three concentric tubings.The inner tubing carries produced reservoir fluids to the oil separator unit.The intermediate tubing carries Syngas to feed the Methanator and the outer tubing carries boiler feed water to the steam generator tubes which provide cooling for the fixed bed of catalyst (see Fig.4).

The effluent from the Methanator reactor is mixed with steam and NH_3 from the steam/water separators and distributed to at least one of the drainholes by means of the valve section,while at least one other drainhole is under oil production.Typical temperatures of the fluids flowing in the concentric tubings are listed in Table 4.

These tubings constitute a heat exchanger in which most of the sensible heat of the production stream is captured by the Syngas and BFW streams,effectively preventing this heat from being lost into the Permafrost.This improves the effective heat rate into the reservoir,while reducing the risk of melting the Permafrost.

RESULTS

The entire process (surface and subsurface) was simulated using a Bechtel-proprietary computer program.Material and heat balances are

presented in Table 5 for typical operating conditions.

The net heat rate into the reservoir is 220 MMBTU/h/well (64 MW/well). Composition and enthalpy of each stream are given, including that of the injectant. The heating duty of the steam reformer furnace in each train is equal to that in each of the four process trains of the Syngas/Methanol/Gasoline GTG plant designed by Bechtel and built in New Zealand.

DESIGN OF THE DOWNHOLE METHANATOR ASSEMBLY

The Methanator Assembly consists of a stack of four pre-assembled sections of length compatible with the height (ca.60 ft or 18 m) of the derricks of drilling rigs available on the North Slope. They are connected by threaded joints providing the required flow connections between elements. Each section is made up of two units (Fig.4). Each unit consists of a series of concentric tubes. The outer annulus contains the catalyst and water tubes. The central tubing distributes the BFW to the water tubes and the middle annular spaces are used respectively as steam/water separators and to convey the injectant to the Valve Section, located at the top of the assembly, above the hanger. Dimensions of the various elements are given in Table 6, together with tubings and casing specifications. The numerical justification of the Methanator characteristics and its calculated performance as a steam generator are detailed in Appendix 2, based on the physical properties of the flowing gas mixtures derived in Appendix 2.1

The reactor hanger provides a metal/metal seal with the inner surface of the casing. It divides the cased well into two separate compartments. In the upper part, the annulus between casing and water tubing is filled with a thixotropic completion fluid (Arctic Pack mud

or gelled diesel),so as to prevent any convective heat loss to the casing.The effectiveness of metal/metal seals for thermal well completions has been demonstrated for casing sizes up to 24.5 in.(see Ref.21).In the lower compartment,the annulus between casing and reactor shell is filled with the Syngas feed,which is pre-heated first by heat exchange with the hot produced fluids and finally by contact with the reactor shell,prior to entering the bottom of each catalyst bed. The cross section of Fig.5 shows the reactor hanger and the uppermost unit of the Downhole Methanator assembly while it is run in through the internal upset of the casing,which is the female threaded connection of the hanger.The gas-filled sealed assembly is run-in the water-filled casing.Water in the annular space below the hanger is displaced by pressurized inert gas and a retrievable plug is set at the base of the central water tubing.The drill string used to run the assembly can then be disconnected and pulled out to proceed with the drilling and completion of the horizontal drainholes,the run-in of the Valve Section and its connection to the reactor head and to each of the drainhole liners by means of hydraulically-extended telescopic connector tubes (Fig.6).The latter operations may be conducted with a lighter rig than that used for drilling,completion of the vertical well and installation of the Methanator Assembly in its hanger.An alternative well completion method,using a single rig,is also presented in Appendix 2,together with a more detailed description of the Downhole Methanator assembly and of the Valve Section.

OIL PRODUCTION RATE FORECAST

Production rates of horizontal wells or drainholes may be estimated from their comparison with the production rates of vertical wells

drilled through the same layer,using the curves given in Ref.14.and from the basic equations for steady state radial flow of oil and gas and the fractional flow equation for water/oil displacement (Ref.15). Required reservoir data for this calculation are listed in Table 7. For lack of specific West Sak characteristics,some of them were assumed (including relative permeability data),by comparison with other fields.The result is given in Appendix 3.This calculation only involves the flow characteristics of the reservoir and assumptions regarding the effect of heat on West Sak reservoir fluids within the drainage area of each drainhole.To check the validity of these assumptions,a calculation of the heat loss to the overburden and of the displaced oil volume was made using the simple method of Marx and Langenheim (Ref.22),modified for the geometry of a horizontal drainhole.The results,also shown in Appendix 3,confirm the validity of the previous assumptions.The only merits of such a simple reservoir engineering approach are that it allows to consider separately the fluid flow and heat flow phenomena,with minimum information about the reservoir and without the need for numerical reservoir simulators. At present,only the owners of the West Sak,who have a more detailed reservoir description than that published in Ref.1,are in a position to make use of such numerical simulators,in which both the fluid flow and heat flow are coupled,for each time step.Existing thermal oil reservoir simulators,however,do not take into account the beneficial chemical effects of the injected Hydrogen on the reservoir crude nor the physical effects of the injected Methane.For "huff and puff" steam injection,the reservoir performance predicted by available simulators are often very unreliable (see Ref.23) in the absence of any reservoir production history.This is one more justification for taking a much

simpler approach for estimating the lifetime average oil production of a representative well completed as shown on Fig.7 into four of the 6 reservoir layers (assuming that the other two are too thin to justify their exploitation).The oil production profile vs.time for each drainhole operated in the "huff and puff" mode is shown on Fig.8,which is typical of most "huff and puff" projects.Its characteristics are summarized on Table 13.The oil production decline in time for each drainhole may be offset by sequentially connecting more drainhole pairs to the vertical well until all four layers of sufficient thickness are put into production.It is shown in Appendix 3 that a single drainhole can take the entire steam and gas output of the Downhole Reactor.Conversely,when the vertical cased well is equipped all 8 drainholes,7 of them may be on production,so that the average oil production per drainhole may drop to 1/7 th of its original value while the well production remains constant.In reality,it will be more advantageous to accept a steady decline of production,by bringing into production more of the individual layers at an earlier date than that required for maintaining well production at a constant level,but the latter mode of exploitation leads to a simpler and more conservative forecast of well production.

Considering the total heat rate into the reservoir,the equivalent SOR for conventional surface steam quality of 70% is found to be close to 3/1,which is representative of many similar steam injection projects.The produced WOR is 2/1,which is also reasonable. The oil production rate from a drainhole of 3,000 ft (900 m) reach, completed with a 5 in. ID (12.6 cm) liner is estimated at ca.8,000 B/D,with a probable range of 18,000 to 24,000 B/D for the total fluids produced.The spacing between parallel drainholes in each layer is also

3,000 ft (900 m) and it was assumed that each vertical well would connect in each layer to two opposite drainholes, both alternating in "huff and puff". Assuming that 80 % of the injected Methane is produced and recovered with the gas originally dissolved in the West Sak oil the resulting GLR is 830 SCF/B. This determines the size (9.5/8 in.OD) of the production tubing for a minimum separator pressure of 100 psia.

The steam deliverability through each drainhole was also checked to be sufficient to take the entire output of the downhole generator.

The concentric Syngas tubing and the water tubing were also sized on the basis of the fluids pressure drops and specified as shown in Table 6. The methodology used is described in Appendix 4.

COST ESTIMATES

Drilling and completion times in Table 8 were estimated on the basis of current drilling operations on the North Slope, with the assistance of drilling engineers experienced with conditions and costs of available rigs in that area. A 50 % contingency factor was applied, in view of the fact that horizontal wells have not yet been drilled into the West Sak (although many have been drilled at Prudhoe Bay into the Sadlerochit).

The cost of the downhole reactor assembly was based on vendor quotes established upon the specifications of its various components. Again a 50 % contingency factor was applied to this estimate, presented in Table 9.

The cost of the surface facilities required for the production of the various feeds of the downhole reactors and for all auxiliaries was derived from Bechtel data about the actual cost of the comparable units in the New Zealand GTG plant and from various cost factors

applicable to North Slope facilities designed and built by Bechtel. It was assumed that the plant would consist of heavy (< 2,500 T) modules built on the West Coast, sea-lifted to the North Slope and assembled by the techniques used at Prudhoe Bay. No allowance was made for the cost savings resulting from some of the more recent technical improvements in the mode of construction such as those in the Endicott field, nor from those of barge-mounted plants such as the Prudhoe Bay sea-water treating unit. A conventional 30 % contingency factor was applied to that estimate, presented in Table 10.

A comparison of Tables 8, 9 and 10 shows that the capital cost of a Syngas plant dwarfs those of the wells and subsurface facilities. This is very encouraging because it has been shown that systematic design optimization of such well known process facilities is likely to bring very substantial savings. The development cost of the Endicott field was reduced by nearly 50% through such engineering optimization (Ref. 16).

SIMPLIFIED ECONOMIC ANALYSIS

A pro-forma economic analysis of a basic project consisting of four process trains, and their auxiliaries, feeding eight wells equipped with with downhole reactors and producing at a total rate of 65,000 B/D for 30 years is given in Table 11, including all economic assumptions. This configuration was selected because it corresponds to world-scale facilities providing economies of scale. In view of the magnitude of the West Sak, full development of the field would ultimately require many such basic projects. With a total area of 300 sq. miles (77,670 ha) this is true, even on a 432 ac. (167 ha) spacing for the vertical wells. Each of these wells is connected to opposite pairs of 3,000 ft-reach

horizontal drainholes,each draining an area of 216 ac.(83.6 ha) during the "huff and puff" period of production (Phase 1).Subsequently the mode of operation (in Phase 2) shifts to either in-filled drilling of additional producing wells,also equipped with drainholes,or to a steam flood,as shown in Appendix 5.These production scenarios are compatible with the results of the well production rates,but they should be further analyzed and optimized,based on a more detailed description of the reservoir in the area to be developed.

The area of the field which would could be economically developed by this process,or by any other,is likely to be only a fraction of the total area.From the reservoir description given in Ref.1,it was concluded that application of the proposed process may be limited only to the four upper layers,which present sufficient thicknesses over a major portion of the field.This limitation may prove to be too restrictive.Producing life may also be longer than the assumed 30 years.The assumptions used in calculating the Modified Internal Rate of Return (MIRR) may also be more conservative than those applicable to the West Sak owners.Accordingly,a significant upside potential may exist,which was not considered here,in order to purposely lean toward a conservative evaluation.

It was,however,assumed that only existing roads,major pipelines and sea docks would be used,thus reducing gravel costs to a low level.

The gas used in the plant was assumed to be available at no cost and exempt from royalty and taxes.Conversely,no credit was taken for the CO_2 exported to other fields or to other parts of the West Sak which might be operated under a CO_2 injection recovery process.

Despite present uncertainty about oil prices,a constant 17 $/B at the well head (in constant 1990 dollars) was assumed throughout the 30

years producing life of the basic project.This well head price corresponds to a West-Coast-delivered oil price of about 25 $/B.It was also assumed that Heavy Oil would be exempt from the increase in Severance Tax imposed by the State of Alaska following the Exxon Valdez disaster.

While this price was exceeded briefly at the end of 1990 and during the first Quarter of 1991,it has since dropped,to reach only 24 $/B in the third Quarter.This illustrates the difficulty of making such an economic analysis.In reality,the earliest possible implementation date for a West Sak project of the magnitude considered here is in 1996 or later.For that period,a commonly accepted forecast of world oil prices (Ref.25) reaches values exceeding 25 $/B (in 1990 Dollars) and shows an upward trend.The assumption of a 17 $/B at the well head remaining constant (in 1990 Dollars) over the 30 year life of the project is therefore conservative,based on current information,but this aspect of the economic analysis remains,by far, the largest risk factor.On the hand,the deferment of Heavy Oil production from the known major North Slope reservoirs (West Sak and Ugnu) is not without risk.This is because the produced Heavy Oil,even after its "in-situ" upgrading by the injected steam and gases,remains too viscous to be readily transportable in the existing TAPS facilities,unless it is mixed with a larger volume of lighter crudes produced from other North Slope fields.With the huge Prudhoe Bay field now on decline at a steep rate,the TAPS throughput of lighter crudes is expected to decline sharply and perhaps even to completely vanish during the period 1996-2026 which is considered here for the West Sak Heavy Oil production.In this context,it would be unlikely that oil produced from the West Sak would remain transportable over a 30 years field life

and any further postponement beyond 1996 of the West Sak exploitation would result in a shorter field life,with reduced economic benefits. Fortunately,there are good prospects of finding large new oil fields onshore,if the ANWR is opened to exploration,and offshore,if the large technical problems of safely drilling and producing wells further North in the Beaufort Sea are solved during this timeframe.

The simplified analysis presented in Table 11 shows that,without any benefits of engineering optimization,a 12 % real rate of return (MIRR) is obtained.

This is comparable with that of a marginal North Sea Development (Ref.17) to be implemented during the same period and subject to similar economic conditions.

If,however,further design improvements of the surface facilities lead to a 30 % cost reduction,thus eliminating the assumed contingency factor,Table 12 shows that the rate of return (MIRR) reaches the more attractive value of 19 %.Such a cost reduction appears achievable by considering various alternatives to the duplication of the New Zealand GTG plant and its modularization by the technically proven,but very expensive,Prudhoe Bay construction techniques.With this reduction,the required capital cost per daily barrel of production would fall to 16,185 $/B/D,comparable with that of many current oil development projects in frontier areas.

DISCUSSION

The magnitude of the West Sak and the difficulty of applying steam injection techniques in the Arctic justify a more complete evaluation of the proposed approach than was possible with the limited information available.The present work should be followed by a proper

reservoir engineering study of the oil recovery process by steam and gases.The effects of Methane and steam on West Sak oil need to be fully understood and those of Hydrogen on the reduction of "coking" must be quantified for this particular reservoir,because wide variations of these effects may exist,depending on the oil composition including sulfur content,and on the reservoir mineralogy.These future experimental data should be used in a reservoir simulator to improve our simplified oil production forecast.The extent of the permeability reduction by swelling clays should also be evaluated to arrive at a better estimate of the NH3 requirements.The effectiveness of this additive is well known and the Syngas process lends itself to low-cost synthesis but process optimization is required for this as well as for the Syngas plant.Finally,the downhole Methanator assembly should be field tested in a heavy oil field where such a pilot test can be made at a small fraction of the cost in the Arctic.One such location is presently under consideration by an operating oil company.

Given the long lead time required for the implementation of any new large oil development project on the North Slope,and the necessity to greatly reduce all technical risks before committing the large capital expenditures required in the Arctic,engineering work on the proposed new process and downhole hardware will have to continue at a faster pace in order to resolve all the remaining problems in time for a 1996 start-up.A considerable amount of reservoir engineering work will also have to be done during the same time period by the West Sak owners,who are the only ones,together with the State of Alaska,with access to the detailed information about the reservoir which is required for any worthwhile reservoir simulation.

CONCLUSIONS

1. The possibilty of recovering some of the West Sak heavy oil by a modified steam injection process has been evaluated to the extent possible with the still limited information available on that huge reservoir.

2. The challenge of transporting large amounts of heat through the 2,000 ft (600 m)-thick Permafrost zone is met by providing this energy in chemical form by a Syngas feed stream produced from Natural Gas at the surface.

3. Syngas is converted back to Methane in a downhole catalytic reactor by highly exothermic reactions,which are used to make additional steam.Preliminary engineering of all surface and subsurface facilities show that a heat rate of 220 MMBTU/h (64 MW) per well is achievable with a temperature of injectant (steam and gases) of 620 F (315 C)

4. The injectant (steam and gases) is delivered to the relatively thin and moderately permeable layers of the West Sak through multiple horizontal drainholes operated in "huff and puff" and connected to large-diameter vertical wells,resulting in an estimated average oil production rate of about 8,000 B/D/well.

5. Estimated costs of all the required surface and subsurface facilities were based on preliminary engineering and known cost elements of similar installations.The cost of the surface units,which use conventional technology,dwarfs that of the subsurface facilities. It could be substantially reduced by further engineering optimization.

6. A simplified economic analysis of a basic project consisting of 4 world-scale process units shows that acceptable real rates of return could be obtained,at currently predicted oil prices.

7. Additional work,including a pilot field test in a lower cost area

is being considered.

ACKNOWLEDGEMENT.

This report is an expanded version of SPE Paper 22077,entitled "The Challenge of West Sak Heavy Oil:Analysis of an Innovative Approach",which was co-authored by J.M.Fox of Bechtel Group,Inc.and presented to the SPE International Conference on Arctic Technology,May 1991.

The author wishes to express his thanks to all those in Industry,in the National Laboratories and in Academia who encouraged him to undertake this project.The financial support of the D.O.E.and the technical and financial support of the Bechtel Group Inc. are gratefully acknowledged.Special thanks are given to L.Hansen of LBL for an independent check of the heat transfer parameters of the downhole reactor.

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APPENDICES:

1) Reservoir Description

2) Reactor Design (includes Appendix 2.1 on Physical Properties of Flowing Gas Mixtures)

3) Reservoir Engineering

4) Production Engineering

5) Operating Scenario

TABLE 10
CAPITAL COST SUMMARY
SURFACE FACILITIES SERVING EIGHT DOWNHOLE REACTORS
$MM

	Major Equipment	Bulks	Labor	Total
Reformer Section	165.2	140.8		
Compressor Section	22.0	18.7		
CO2 Removal	17.2	9.0		
Crude Separation	2.8	2.4		
Total On Plot	207.2	170.9	125.0	503.1
Off Plot	70.2	77.7	27.8	175.7
Includes Desalination				
Contractor Indirects				166.2
Catalysts & Chemicals				5.0
Tax, Duties, Freight				8.3
Camp & Catering				16.9
Sealift & Module Movement				86.3
Total Field Cost				961.5
Eng'ng, HO and Fee				75.2
Total Contractor's Cost				1036.7
Owner Costs				99.0
Contingency at 25%				283.9
Total Project - mid 1990				1419.6

TABLE 11
PRO-FORMA ECONOMIC ANALYSIS OF THE BASIC PROJECT

Assumed: 12.5% royalty and 10% sev tax on WH price
48 % of income for State & Federal taxes
17 $/B WH price in constant dollars
355 D/y production at 65,000 B/D constant rate for 30 years
7% p.a.interest rate from year 1 to year 32

Year	Capital Cost($MM)	Disc fact. (aft.I.Tax)	Capex ($MM)	O.& M. Cost($MM)
1	(751)	0.670	(503)	0
2	(752)	0.635	(477)	0

TOTAL (1,503) TOTAL NPC: (980)

	Depr RACRS	O.& M. Cost($MM)	disc fact. (aft.I.T.)	Opex ($MM)	disc Gross (-Roy .S.T.)	Taxable (disc.)	NPV (aft.I.T)
3	(198)	(45)	0.49	(22)	256	38	20
4	(218)	(53)	0.43	(23)	241	0	0
5	(208)	(45)	0.38	(17)	225	0	0
6	(193)	(53)	0.33	(17)	210	0	0
7	(165)	(45)	0.29	(13)	194	16	8 PO
8	0	(61)(W.O.)	0.28	(16)	182	166	86
9	0	(45)	0.23	(10)	170	160	83
10	0	(53)	0.20	(10)	157	147	76
11	0	(45)	0.18	(8)	148	140	73
12	0	(61)(W O)	0.16	(10)	139	129	67
13	0	(45)	0.14	(6)	130	124	64
14	0	(53)	0.12	(6)	120	114	59
15	0	(45)	0.11	(5)	111	106	55
16	0	(61) (W O.)	0.09	(5)	105	100	52
17	0	(45)	0.08	(4)	96	92	48
18	0	(53)	0.07	(4)	90	86	45
19	0	(45)	0.06	(3)	83	80	42
20	0	(53)	0.05	(3)	77	74	38
21	0	(45)	0.04	(2)	74	72	37
22	0	(53)	0.03	(2)	68	66	34
23	0	(45)	0.03	(1)	65	64	33
24	0	(53)	0.02	(1)	59	58	30
25	0	(45)	0.02	(1)	56	55	29
26	0	(53)	0.01	(1)	52	51	27
27	0	(45)	0.01	(1)	50	49	25
28	0	(53)	0.008	(0.3)	48	48	25
29	0	(45)	0.005	(0.2)	47	47	24
30	0	(53)	0.004	(0.2)	46	46	24
31	0	(45)	0.003	(0.1)	45	45	23
32	0	(53)	0.002	(0.1)	44	43	22

TOTAL NPV = 1,148 $MM

REAL RATE OF RETURN = 12% (after tax)

TABLE 12
PRO-FORMA ECONOMIC ANALYSIS AFTER 30% REDUCTION OF SURFACE FACILITIES COST BY ENGINEERING OPTIMIZATION,WITH SAME ASSUMPTIONS AS IN TABLE 13

Year	Rev Capital Cost($MM)	Disc fact (aft I Tax)	Capex ($MM)	O & M Cost($MM)
1	(526)	0.670	(352)	0
2	(526)	0.635	(334)	0

TOTAL (1,052) TOTAL NPC (686)

	Depr. ACRS	O & M. Cost($MM)	disc fact. (aft.I.T.)	Opex ($MM)	disc Gross (-Roy .S.T.)	Taxable (disc.)	NPV (aft.I.T.)
3	(137)	(45)	0.49	(22)	256	97	50
4	(218)	(53)	0.43	(23)	241	0	0
5	(167)	(45)	0.38	(17)	225	41	21
6	(110)	(53)	0.33	(17)	210	83	43 P.O.
7	(54)	(45)	0.29	(13)	194	127	66
8	0	(61) (W.O.)	0.28	(16)	182	166	86
9	0	(45)	0.23	(10)	170	160	83
10	0	(53)	0.20	(10)	157	147	76
11	0	(45)	0.18	(8)	148	140	73
12	0	(61) (W.O.)	0.16	(10)	139	129	67
13	0	(45)	0.14	(6)	130	124	64
14	0	(53)	0.12	(6)	120	114	59
15	0	(45)	0.11	(5)	111	106	55
16	0	(61) (W.O.)	0.09	(5)	105	100	52
17	0	(45)	0.08	(4)	96	92	48
18	0	(53)	0.07	(4)	90	86	45
19	0	(45)	0.06	(3)	83	80	42
20	0	(53)	0.05	(3)	77	74	38
21	0	(45)	0.04	(2)	74	72	37
22	0	(53)	0.03	(2)	68	66	34
23	0	(45)	0.03	(1)	65	64	6433
24	0	(53)	0.02	(1)	59	58	30
25	0	(45)	0.02	(1)	56	55	29
26	0	(53)	0.01	(1)	52	51	27
27	0	(45)	0.01	(1)	50	49	25
28	0	(53)	0.008	(0.3)	48	48	25
29	0	(45)	0.005	(0.2)	47	47	24
30	0	(53)	0.004	(0.2)	46	46	24
31	0	(45)	0.003	(0.1)	45	45	23
32	0	(53)	0.002	(0.1)	44	44	4322

TOTAL NPV = 1,300 $MM (after tax)

REAL RATE OF RETURN = 19%

TABLE 6
REACTOR ASSEMBLY AND WELL CHARACTERISTICS

Catalytic Bed:

Syngas total flow rate >= 100 MMSCF/D
Syngas composition =< 3.5 H2,1 CO
Vertical gas flow through the bed
Total Catalyst Volume = 370 CF
Bed cross section =< 2.0 sq.ft
Total bed length >= 208 ft
Bed porosity = 35 %
Catalyst particle diameter = 0.008 ft
Number of bed sections = 4 in parallel flow
Gas pressure drop per section =< 100 psi
Bed discharge pressure = 2,575 psia
Bed temperature =< 800 F

Water tubes system:

Total cooling water rate = 168,000 #/h
Number of unit bundles = 8 (2 per bed section) in parallel flow
Number of tubes per bundle = 82 in parallel flow
Tube length = 37.5 ft (in helix)
Tube OD = 5/8 inch ;ID 1/2 inch
Bed volume fraction occupied by tubes = 12.3 %
Tube discharge pressure = 2,625 psia
Tube discharge temperature = 675 F
Cooling water temperature at tube inlet >= 400 F

Injectant per well @ 98 % conversion:

Reactor effluent temperature >= 620 F
Reactor effluent pressure: 2,575 psia
Total injected steam:223,000 #/h (@ 100 % quality)
Injectant volume:152 MMSCF/D
Inject.composition:steam 0.744,CH4 0.211,H2 0.039,NH3 0.004,CO 0.002

Characteristics of Tubular Goods Outside Reactor Assembly:

Item	OD (in.)	ID	Drift D.	Coupling OD	Steel	Supplier
Drainh.Liner	5.5	5.0	N.A.	5.5	L-80	Hydril
Prod.Tubing	9.5/8	8.921	N.A.	10.141 (VAM)	L-80	VAM
Gas Tubing	11.3/4	10.772	10.616	11.9375	L-80	Hydril
Water Tubing	13.3/8	12.347	12.191	API	L-80	Any
Casing	24.5	22.5	22.5	API	X-70	Mannesmann

Characteristics of Tubular Internals of Reactor

Item	OD (in.)	ID	Alloy	Supplier
Reactor Shell	21	20.2	Hastelloy C-22	Haynes Int.
Water Tubing	2.3/8	2.0	id.	id.
Gas Tubing	3.5	3.0	id.	id.
Steam/water Sep.	5.5	4.95	Croloy TP 321H	Babcock
Injectant Tubing	7.0	6.4	Hastelloy C-22	Haynes Int.

TABLE 7
RESERVOIR PARAMETERS USED IN FORECASTING OIL PRODUCTION RATE

WEST SAK UPPER SANDS
Typical Depth: 3,500 ft Pe =< 1,815 psia Te = 75 F
Fracturing Pressure: 2,815 psia

WEST SAK LOWER SANDS
Typical Depth: 4,000 ft Pe =< 2,000 psia Te = 75 F
Fracturing Pressure: 3,215 psia

RESERVOIR FLOW PARAMETERS
Qoh >= 5 Qov for 3,000 ft reach (Ref.14)
Qov = steady state radial flow rate into a vertical well
T injectant = 620 F T av. = 347.5 F
P sep. >= 100 psia
External radius = 1,500 ft
Well radius = 0.229 ft

K (md)	h (ft)	Φ (%)	So	Kro	Krw	Krg	Bo	Bw	μo (cp)	μw	μg(inj.)
150	30	30	>=60	0.9	0.09	0.6	1.4	1.0	1.5	0.14	0.0238

TUBULAR FLOW
K.E.Brown's Pressure Gradient Curves for horizontal and vertical flow

EQUIVALENT STEAM VOLUME FOR ESTIMATE OF EQUIVALENT SOR:
(Total heat rate - 20 % heat loss to overburden) / enthalpy of conventional 70 % quality steam)

TABLE 8
DRILLING AND COMPLETION SEQUENCE AND PRELIMINARY COST ESTIMATE

	OPERATION	COST (1,000 $)
1)	Auger 42 in.hole & install 80 ft of 36 in.conductor pipe	25
2)	Mobilize/demobilize rig (Nabors 18-E) @ 20/day for 12 days	200
3)	Drill 30 in.hole to 3,000 ft (3 days @ 40/day)	120
4)	Run 24 in.casing to 3,000 ft @ 150 $/ft	450
5)	Cost of 4 connector joints in casing string	40
6)	Cement 24 in. casing (5,500 cu.ft)	100
7)	Hang Downhole Reactor Assembly (1 day)	20
8)	Drill 2 horizontal drainholes 8.5 in.,3,000 ft long(12 days)	600
9)	Complete 2 horizontal holes (5.5 in.slotted liners + 2 days)	95
10)	Run 13 3/8 in.tubing, cross-overs, connector/valve assembly	140
11)	Run 11 3/4 in.tubing (36 $/ft + 1 day)	128
12)	Run 9 5/8 in.tubing (31 $/ft + 1 day)	112
13)	Wellhead (w/metal/metal seals)	160
14)	Christmas tree	110
	Total	2,300
	Costs include: mud, drilling tools & bits, cement, tubulars, all freight	
15)	Contingency factor (50 %)	1,150
	Total, incl. contingency	3,450

TABLE 9
COST ESTIMATE OF REACTOR OVERHEAD ASSEMBLY INCLUDING CROSS-OVERS,CONNECTORS AND DOWNHOLE VALVE SECTION

RAW MATERIALS:

Item	Material		Cost (1,000 $)
18 in.OD Outer shell(0.4"wall)	Hastelloy	C-22 @ 9.95 $/#	25
Lower Cross-Over Section	Hastelloy	C-22 @ 10 $/#	10
Twin Connectors Section(7" OD)	Stainless	321 H @	10
Valve Section (90 ft,5.5"OD)	Stainless	321 H @	20
Upper Cross-Over Section	Hastelloy	C-22 @ 10 $/#	10
1/4" Hydraulic Lines (9,000 ft)	Stainless	304 @ 1 $/ft	9
300 tubing clamps	Stainless	304 @ 20 ea	6
	Subtotal raw materials		90
DESIGN,FABRICATION AND TESTING:			500
VALVES,INSTRUMENTATION AND CONTROLS:			
2 Downhole Valves			45
1 Syngas Feed Control Valve			5
4 Temperature Controllers			10
4 Water Level Controllers			20
1 Reactor Control Computer			20
Assembly and Testing			100
	Subtotal Instruments & Controls		200
Warehousing and Freight:			50
CONTINGENCY FACTOR (50%)			420
TOTAL (Incl.Contingency)			1,260

COST ESTIMATE OF DOWNHOLE REACTOR ASSEMBLY

RAW MATERIALS:

Item	Material		Cost(Gulf Coast) (1,000 $)
21 in Reactor shell (250 ft)	Hastelloy	C-22 @ 9.95 $/#	243
2 3/8 Water tubing (250 ft)	Hastelloy	C-22 @ 153.5 $/ft	38
3 1/2 Gas tubing (250 ft)	Hastelloy	C-22 @ 233.5 $/ft	58
5 1/2 Steam-water Sep.(250 ft)	Croloy	TP 321 H @ 64.7 $/f	16
7 in.Injectant tubing (250 ft)	Hastelloy	C-22 @ 595 $/ft	149
5/8 in.Water tubes (25,000 ft)	Croloy	TP 321 H @ 3.25 $/ft	81
Threaded reactor hanger/head	Hastelloy	C-22 @ 10 $/#	41
14 reactor unit head/base	Hastelloy	C-22 @ 10 $/#	388
Reactor base	Hastelloy	C-22 @ 10 $/#	20
Methanation Catalyst			1,000
	Subtotal Materials		2,034
DESIGN,FABRICATION AND TESTING			2,000
Warehousing and Freight			200
	TOTAL DIRECT COST		4,234
CONTINGENCY (50%)			2,117
	TOTAL (incl.contingency)		6,351

TABLE 5

Surface Facilities
Heat and Mass Balance Per Reformer Furnace

	CO2 Compressor				Syngas Compressor							
Stream Number	10	11	12	13	9	9	14	15	16	17	18	19
Stream Name	CO2 Compr Inlet	Stage 1 Outlet	Stage 2 Outlet	CO2 Product	Stage 1 Inlet Gas (CO2 Free)		Stage 1 Outlet	Stage 2 Inlet	Stage 2 Outlet	Stage 3 Inlet	Stage 3 Outlet	Total H2O Produced
Properties												
Composition												
CH4 - MPH	0.0	0.0	0.0	0.0	1570.3	1570.3	1570.3	1570.3	1570.3	1570.3	1570.3	0
C2H6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
C3H8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
C4H10	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
C5H12	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
CO2	1297.9	1297.9	1297.9	1297.9	756.4	756.4	756.4	756.4	756.4	756.4	756.4	0
N2	0.0	0.0	0.0	0.0	76.8	76.8	76.8	76.8	76.8	76.8	76.8	0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
CO	0.0	0.0	0.0	0.0	4741.4	4741.4	4741.4	4741.4	4741.4	4741.4	4741.4	0
H2	0.0	0.0	0.0	0.0	17993.3	17993.3	17993.3	17993.3	17993.3	17993.3	17993.3	0
H2O	0.0	0.0	0.0	0.0	75.3	75.3	75.3	37.6	37.6	18.2	18.2	9707.6
Total - MPH	1297.9	1297.9	1297.9	1297.9	25213.5	25213.5	25213.5	25175.9	25175.9	25156.5	25156.5	9707.6
C atoms	1297.9	1297.9	1297.9	1297.9	7068.1	7068.1	7068.1	7068.1	7068.1	7068.1	7068.1	0.0
H atoms	0.0	0.0	0.0	0.0	42418.5	42418.5	42418.5	42343.2	42343.2	42304.4	42304.4	19415.2
O atoms	2595.8	2595.8	2595.8	2595.8	6329.5	6329.5	6329.5	6291.8	6291.8	6272.4	6272.4	9707.6
N atoms	0.0	0.0	0.0	0.0	153.6	153.6	153.6	153.6	153.6	153.6	153.6	0.0
Total - #/hr	57120	57120	57120	57120	231071	231071	231071	230392	230392	230042	230042	174692
Conditions												
Temp - Deg F	100.0	323.8	323.8	100.0	100.0	231.4	231.4	100.0	231.5	100.0	231.5	100.0
Press - Atm	1.00	7.28	28.57	28.57	21.43	21.43	45.23	43.23	91.21	89.21	188.23	23.14
MM Btu/hr	-214.04	-211.24	-211.24	-214.04	-315.23	-291.54	-291.53	-311.53	-287.86	-309.62	-285.97	-1137.00
Delta H		2.80	2.80		-391.86	23.69	23.70		23.67		23.65	
oxygen/carbon												
steam/carbon												
H2/CO					3.79				3.79		3.79	
(H2-CO2)/(CO+CO2)					3.14				3.14		3.14	
Boudouard Ratio												
Comp Eff		0.77	0.77				0.77		0.77		0.77	
H2O mol fraction					0.0030			0.0015		0.00072		
CO2 mol fraction					0.030			0.030		0.030		
Heat Cap - molar		9.64	9.64			7.15	7.15	7.15		7.15		
k		1.2061	1.2061			1.2779	1.2779	1.2779		1.2779		
(n-1)/n		0.2219	0.2219			0.2824	0.2824		0.2824		0.2824	
Comp. Ratio		4.55	4.55				2.11		2.11		2.11	
Head - ft		35382	35382				78424		78539		78500	
BHP		1326	1326				11866		11869		11868	

TABLE 5

Downhole Facilities
Heat and Mass Balance Per Reformer Furnace
(Per Two Wells)

Stream Number	20	21	22	23
Stream Name	Methanator Feed	Methanator Effluent	Steam Generated	Combined Injectant
Properties				
Composition				
CH4 - MPH	1570.3	7052.4	0.0	7052.4
C2H6	0.0	0.0	0.0	0.0
C3H8	0.0	0.0	0.0	0.0
C4H10	0.0	0.0	0.0	0.0
C5H12	0.0	0.0	0.0	0.0
CO2	756.4	15.5	0.0	15.5
N2	76.8	76.8	0.0	76.8
O2	0.0	0.0	0.0	0.0
CO	4741.4	0.2	0.0	0.2
H2	17993.3	806.1	0.0	806.1
H2O	18.2	6241.2	18650.1	24891.3
Total - MPH	25156.5	14192.3	18650.1	32842.4
C atoms	7068.1	7068.1	0.0	7068.1
H atoms	42304.4	42304.4	37300.2	79604.6
O atoms	6272.4	6272.4	18650.1	24922.5
N atoms	153.6	153.6	0.0	153.6
Total - #/hr	230042	230042	336000	566043
Conditions				
Temp - Deg F	450.0	800	675	736.3
Press - Atm	185	180	180	180
MM Btu/hr	-245.97	-687.39	-1744.50	-2431.89
Delta H	40.00	-441.42		
oxygen/carbon	0.00			
steam/carbon	2.00			
H2/CO	3.79			
(H2-CO2)/(CO+CO2)	3.14			
Boudouard Ratio	0.00458	20089		
O2=0;air=1				
Delta T SMR - Deg F		-20.0		
Eq. Temp - Deg K		710.9		
Delta T WGS - Deg F		-20.0		
Eq. Temp - Deg K		710.9		
Equilibrium Constants				
K - WGS		8.11		
K - SM		0.00047		
K - Boudouard		3849		

TABLE 4
TYPICAL FLUID TEMPERATURES IN CONCENTRIC TUBINGS

Central Production Tubing	Syngas Intermed. Tubing	Water Outer Tubing	Well Casing/ Cement
Surface: 3[illegible]0 F	300 F	100 F	25 F
Bottom: 5[illegible]0 F	400 F	300 F	125 F

TABLE 5

Surface Facilities
Heat and Mass Balance Per Reformer Furnace

	Reformer Furnace								Cooling and Separation		
Stream Number	1	1	2	3	4	5	6	8	7	8	9
Stream Name Properties	Feed Gas Mol %	Feed Gas	CO2 Recycle	Steam	Comb Feed	Effl	Gas Out	Gas Out	Cooled Gas H2O Free	Water	CO2 Free Gas
Composition											
CH4 - lbPH	79.7	6122.8	0.0	0.0	6122.8	1570.3	1570.3	1570.3	1570.3	0.0	1570.3
C2H6	5	384.1	0.0	0.0	384.1		0.0	0.0	0.0	0.0	0.0
C3H8	2	153.6	0.0	0.0	153.6		0.0	0.0	0.0	0.0	0.0
C4H10	0.3	23.0	0.0	0.0	23.0		0.0	0.0	0.0	0.0	0.0
C5H12	0.000	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0
CO2	12	921.9	0.0	0.0	921.9	2054.3	2054.3	2054.3	2054.3	0.0	756.4
N2	1	76.8	0.0	0.0	76.8	76.8	76.8	76.8	76.8	0.0	76.8
O2		0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0
CO		0.0	0.0	0.0	0.0	4741.4	4741.4	4741.4	4741.4	0.0	4741.4
H2		0.0	0.0	0.0	0.0	17993.3	17993.3	17993.3	17993.3	0.0	17993.3
H2O		0.0	0.0	16732.0	16732.0	9725.0	9725.0	9725.0	75.3	9650.5	75.3
Total - MPH	100	7682.3	0.0	16732.0	24414.3	36162.0	36162.0	36162.0	26511.4	9650.5	25213.5
C atoms		8366.0	0.0	0.0	8366.0	8366.0	8366.0	8366.0	8366.0	0.0	7066.1
H atoms		28255.5	0.0	33464.1	61719.6	61719.6	61719.6	61719.6	42418.5	19301.1	42418.5
O atoms		1843.8	0.0	16732.0	18575.8	18575.8	18575.8	18575.8	8925.3	9650.5	8329.8
N atoms		153.6	0.0	0.0	153.6	153.6	153.6	153.6	153.6	0.0	153.6
Total - #/hr		100610	0	301445	402055	402055	402055	402055	268191	173864	231071
Conditions											
Temp - Deg F		200.0	323.8	448.8	1000.0	1616.0	1000.0	318.0	100.0	100.0	100.0
Press - Atm			28.57		28.24	24.84	24.14	23.44	22.74	22.74	21.43
MM Btu/hr		-307.53	0.00	-1597.71	-1743.83	-1032.27	-1222.61	-1419.04	-529.27	-1130.31	-315.23
Delta H					161.62	711.36	-190.34	-196.42		-240.55	
oxygen/carbon					0.00						
steam/carbon					2.00						
H2/CO						3.795	3.79	3.79	3.79		3.79
(H2-CO2)/(CO+CO2)						2.35	2.35	2.35	2.35		3.14
Boudouard Ratio						0.133					
O2=0;air=1											
Delta T SMR - Deg F						15.0	H2O mol fraction		0.0028		
Eq. Temp - Deg K						1144.8	CO2 mol fraction				0.03
Delta T WGS - Deg F						0.0					
Eq. Temp - Deg K						1153.2					
Equilibrium Constants											
K - WGS						0.802					
K - SM						842.55					
K - Boudouard						0.036					

TABLE 3
RESULTS OF SAND PACK DISPLACEMENTS OF WEST SAK OIL WITH VARIOUS FLUIDS

M.W.Hornbrook et al.
19 API West Sak oil in Oklahoma No.1 fluid loss control sand pack:
Displacement by Steam/CO2 mixtures @ 1,690 psi,constant vol.flowrate

Run No.	T (F)	CO2/steam (mol/mol)	Soi (%)	Sgi (%)	Rec.@ 3 PV (%)	WOR @ 3 PV (B/B)
1	660	0/1	89.1	0	67	4.1
2	660	0/1	77 5	11.2	32	13.6
3	660	1/1	88	0	74	1.4
4	660	1/3	89.1	0	80	2.2
5	660	1/4	90.8	0	80	2.1
6	570	1/3	88.1	0	80	2.2
7	572	1/4	89.9	0	73	3

K.Deghani et al.
19 API West Sak oil displacement in sand packs:
Displacement by Steam/CH4 mixtures @ 600 F

CH4 in CH4/steam mixture	Ultimate Recovery Improvement
0 (pure steam)	0 (by definition)
20 %	12.5 %
25 %	10.4 %

A.K.Sharma et al.
19 API West Sak oil in 12 m Slim tubes packed w/ Ottawa sand:
Displacement by NGL or CO2 at various conditions

Run No.	NGL (mol %)	T (F)	P (psia)	Ult.Rec. (%)	Sight Glass	Interpretation (w/23 comp. model)	
1	5.0	80	2,500	39.7	2-phase	Immiscible	Asph.Prec.
2	5.0	80	3,500	63.1	id.	id.	id.
3	17 6	80	1,900	73.8	id.	id.	id.
4	17.6	80	2,350	74.7	id.	id.	id.
5	17 6	80	2,500	73.6	?	id.	id.
6	17.6	80	2,775	75.3	?	id.	id.
7	17.6	250	4,050	54.7	2-phase	id.	id.
8	35.0	80	3,500	78.1	1-phase	MCM	?
Displacement by CO2:							
1		100	1,750	45	2-phase	Immiscible	id.

R.A.De Ruiter et al.
18.5 API(oil A) and 14.0 API(oil B) West Sak oils in 6.1 m Slim tubes
All runs except those with pure C2 show asphaltene precipitation

Pressure (psig)	Average Velocity (ft/day)	Volume of Gas Injected (Fraction PV)	Breakthrough (% PV)	Oil Recovery (% PV)	Result
Ethane					
2000	13.8	1 33	85	97 3	miscible
1500	9 5	1 21	78	96 7	miscible
1200	23 8	1.72	65	94 8	miscible
725	14 3	1.23	70-80	94 3	miscible
433	13 6	1.13	40	70 3	immiscible
*2000	1 3	1 20	86	93 7	miscible
CO2					
2000	16.8	1 08	32	48 8	immiscible
Ethane/Propane/N-Butane (35/34/31 mole %)					
1525	13 5	1.23	86	100.7	miscible
Methane/Ethane/Propane/N-Butane (32 mole % Methane)					
2000	13 1	1 25	75	96 1	miscible
1500	13 5	1 39	83	97 7	miscible
1000	11.6	1 22	73	95 0	miscible
Methane/Ethane/Propane/N-Butane (42.5 mole % Methane)					
2000	8 3	1 22	78	93 5	near-miscible
1500	17 9	1 23	82	87 0	immiscible
Methane/Ethane/Propane/N-Butane (51 mole % Methane)					
2000	17 6	1 20	68-76	84 2	immiscible
1500	14 2	1 26	78-85	88 3	immiscible

All displacements performed with Oil A except displacement marked by *.

TABLE 1
OVERALL REACTIONS AND HEAT BALANCE

SURFACE REFORMER

$CH_4 + H_2O \rightarrow 3H_2 + CO$
$CO + H_2O \leftrightarrow CO_2 + H_2$
AT 800 °C & 25 ATM. OUTLET
ENDOTHERMIC @ 365 MMBTU/HR PER WELL

METHANATION

$3H_2 + CO \rightarrow CH_4 + H_2O$
$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O$
AT ≤ 425 °C & 183 ATM.
EXOTHERMIC @ 229 MMBTU/HR PER WELL

TABLE 2
PUBLISHED INFORMATION ABOUT THE WEST SAK RESERVOIR

Estimated OOIP: 13 to 25 x Billion Barrels
Areal extent: ca.300 sq.miles
Oil Gravity: 10.5 to 23 API
Reservoir depth: 2,500 ft to 4,600 ft
Reservoir temperature: 45 to 100 F
Number of separate layers: 6
Thickness of each layer: 10 to 50 ft
Average porosity: >= 30%
Oil saturation : 60 to 70 %
Typical permeability : 150 md
Lithology: fine-grained quartzitic shaly sandstone, very friable, some swelling clays, Glauconite, Siderite reported,

Characteristics of oil samples:

	Ref.5	**Ref.2 (oil A)**	**(oil B)**
API gravity:	19.2	18.5	14.0
Viscosity:	35.4 cp @ 80 F	256 @ 75 F	5,392 @ 75 F
Solution GOR:	210 SCF/STB	167 @ B.P.	126 @ B.P.
Bubble point:	1,690 psi	1,318 psig @ 60 F	1,234 psig @ 60 F
Oil FVF:	1.069 B/STB	1.053 @ B.P.	1.035 @ B.P.
Gas composition:	98% CH4	N.A.	N.A.
C21+ fraction:	28.82 mol %	69 w%	85 w%
MW:	455 (C21+ frac.)	330 (dead oil)	446 (dead oil)
Sulfur w%	1.82	1.03	1.39
Asphaltene w%	2.8	4.9	4.8

TABLE 13

CHARACTERISTICS OF OIL PRODUCTION PROFILE

1. THE LAG TIME
   INCREASES WITH CYCLE NUMBER

2. THE PEAK OIL RATE
   DECREASES WITH CYCLE NUMBER

3. THE DECLINE RATE
   DECREASES WITH CYCLE NUMBER

4. THE SLOPE OF INCLINE CURVE
   DECREASES WITH CYCLE NUMBER

FIGURE 1 MAP OF EXISTING ROADS AND PIERS

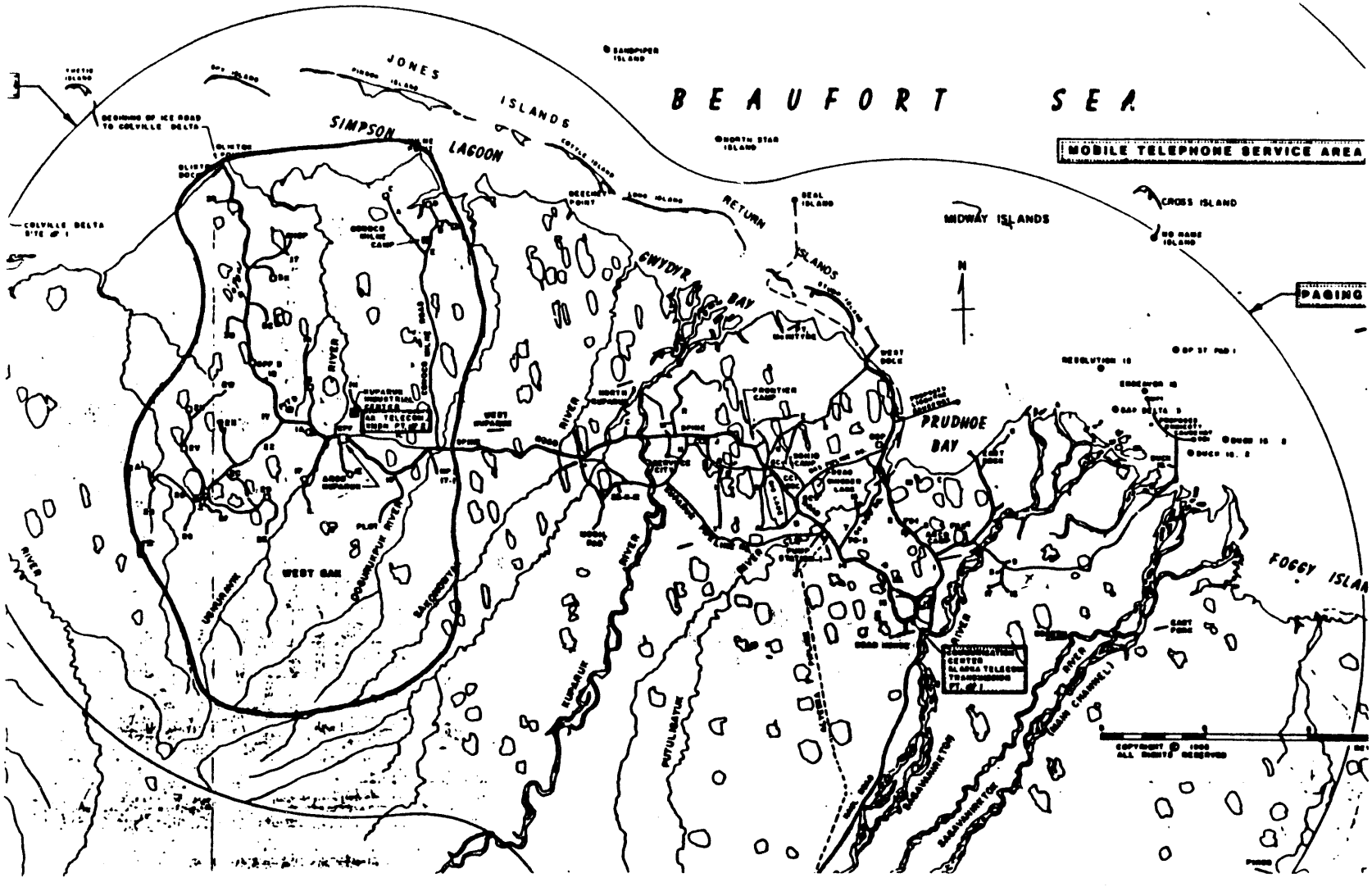
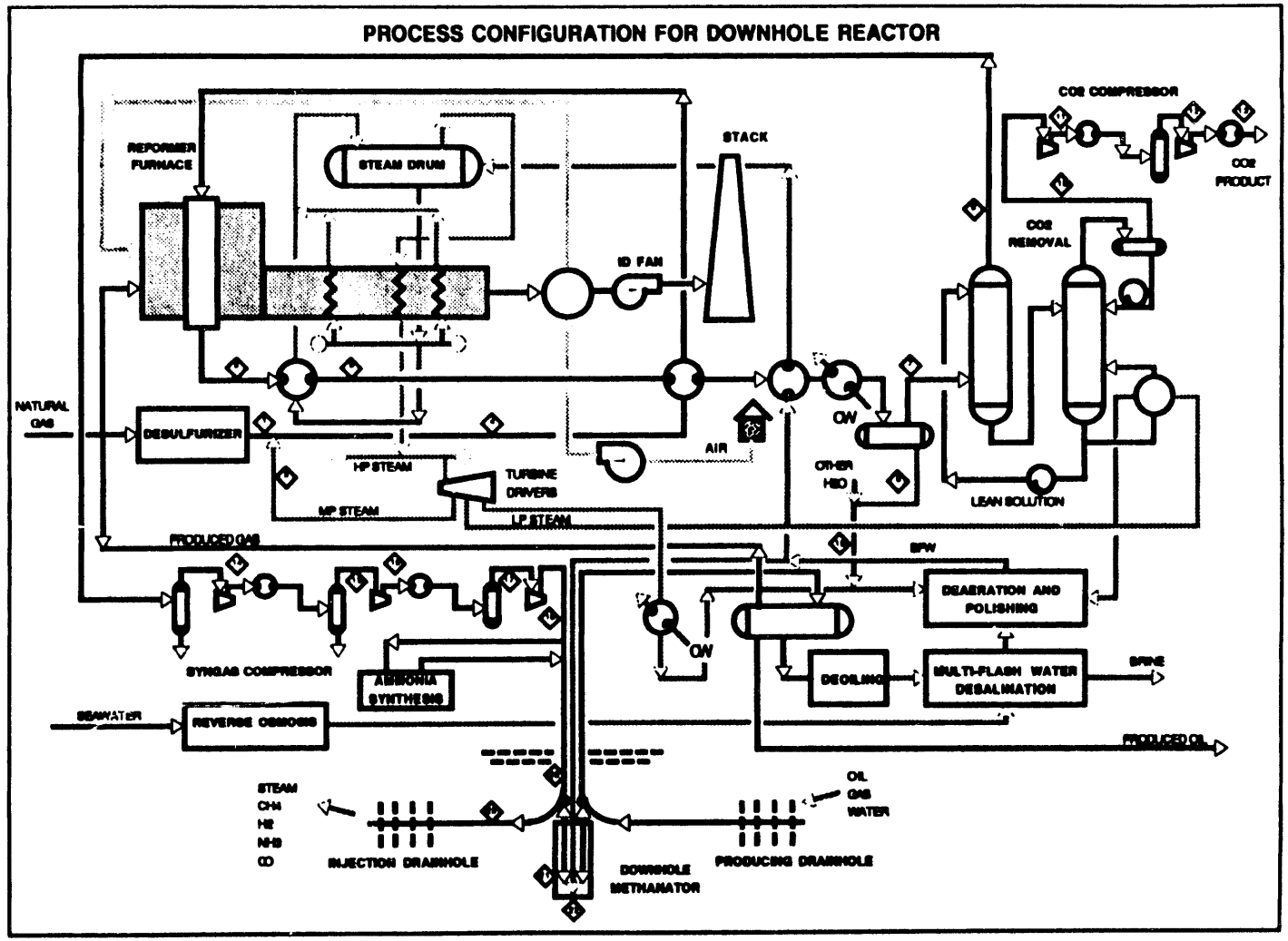


FIGURE 2


PROCESS CONFIGURATION FOR DOWNHOLE REACTOR
REFORMER FURNACE
STEAM DRUM
STACK
ID FAN
CO2 COMPRESSOR
CO2 PRODUCT
CO2 REMOVAL
NATURAL GAS
DESULFURIZER
HP STEAM
MP STEAM
TURBINE DRIVERS
LP STEAM
AIR
CW
OTHER H2O
LEAN SOLUTION
PRODUCED GAS
BFW
DEAERATION AND POLISHING
SYNGAS COMPRESSOR
AMMONIA SYNTHESIS
DEOILING
MULTI-FLASH WATER DESALINATION
BRINE
SEAWATER
REVERSE OSMOSIS
PRODUCED OIL
STEAM
CH4
H2
NH3
CO
INJECTION DRAINHOLE
DOWNHOLE METHANATOR
PRODUCING DRAINHOLE
OIL
GAS
WATER

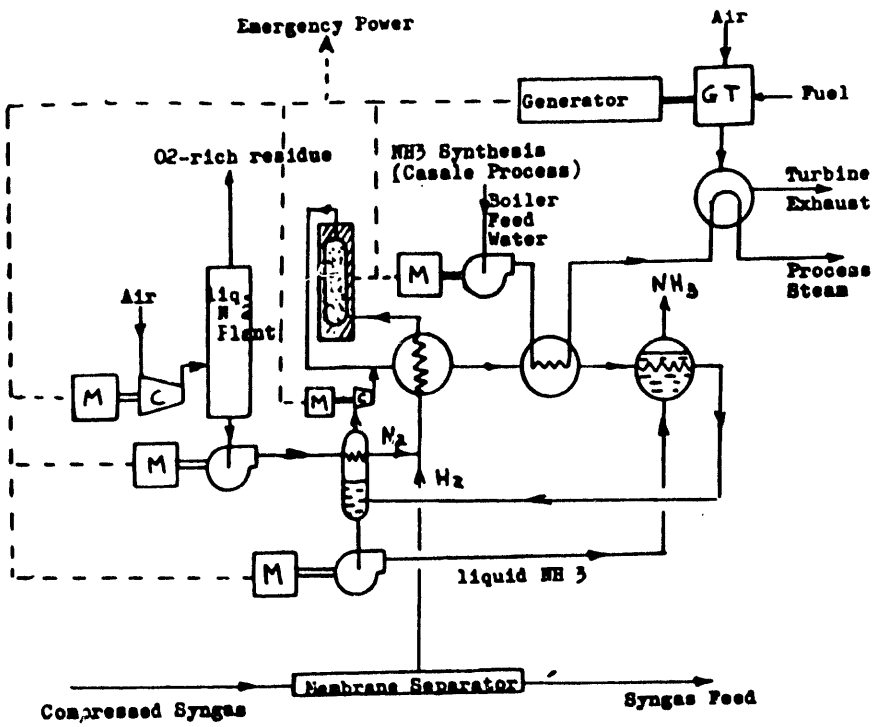


FIGURE 3
MULTI-PURPOSE AUXILIARY FACILITIES/ NH_3 PLANT

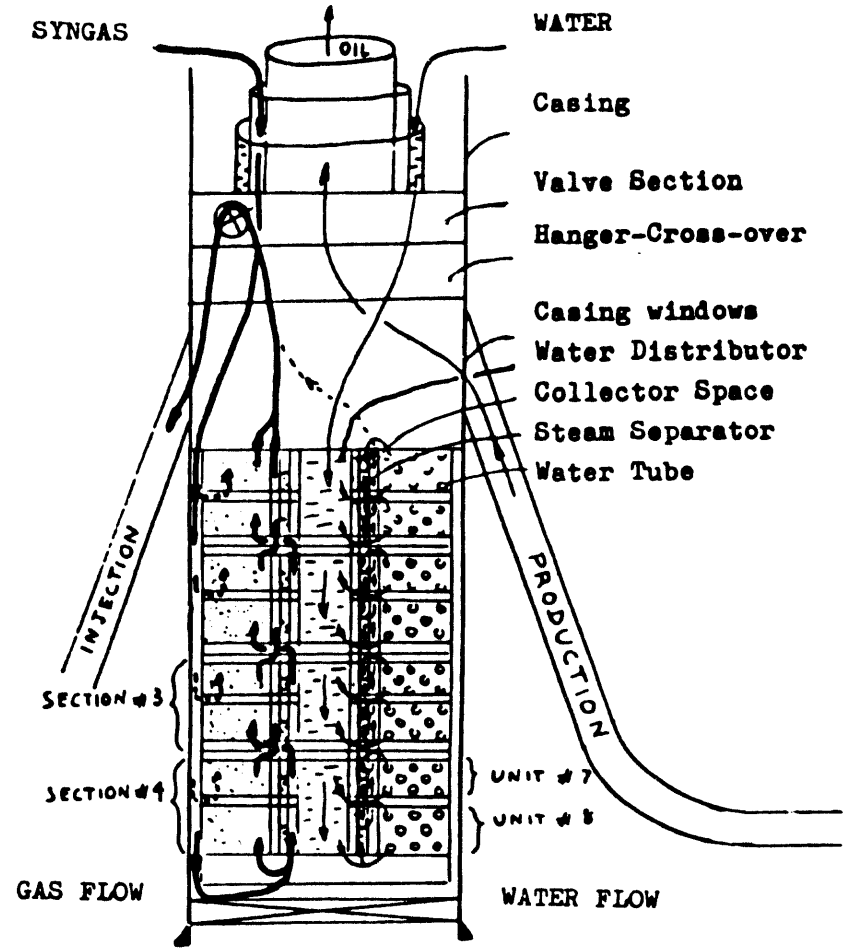


FIGURE 4
VERTICAL FLOW REACTOR - TOP DISCHARGE

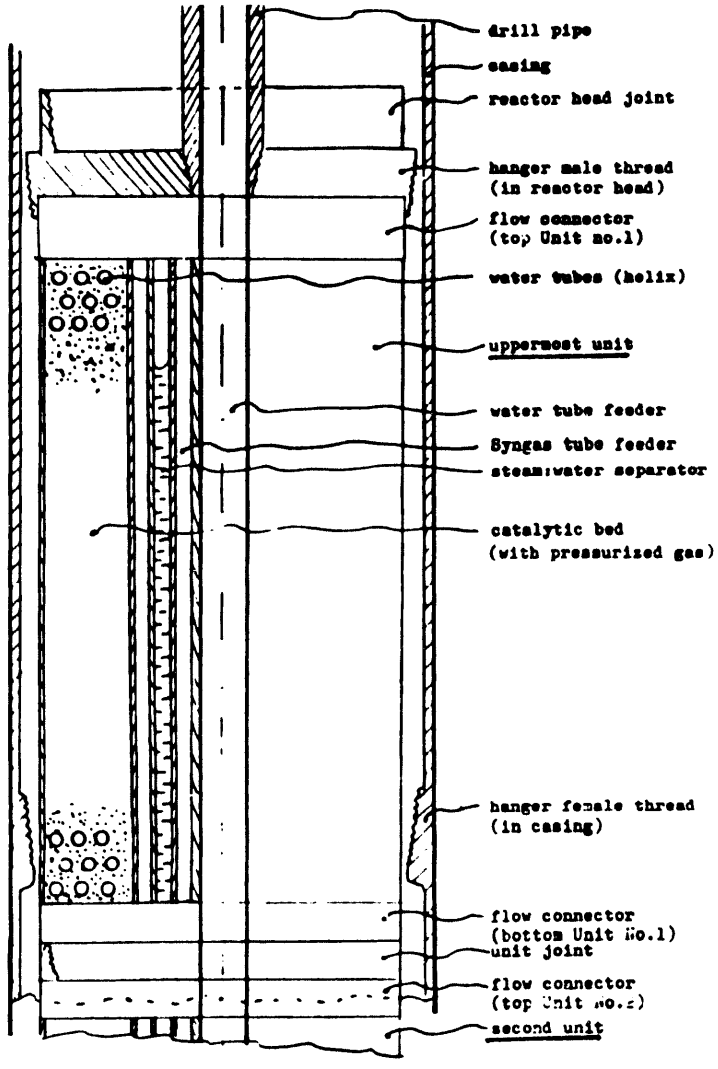
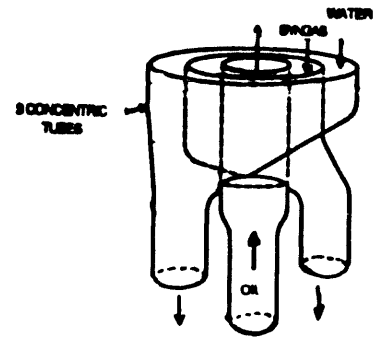


FIGURE 5
HANGER AND UPPERMOST UNIT IN RUN-IN POSITION



DOWNHOLE VALVE SECTION

(DETAIL NOT SHOWN)

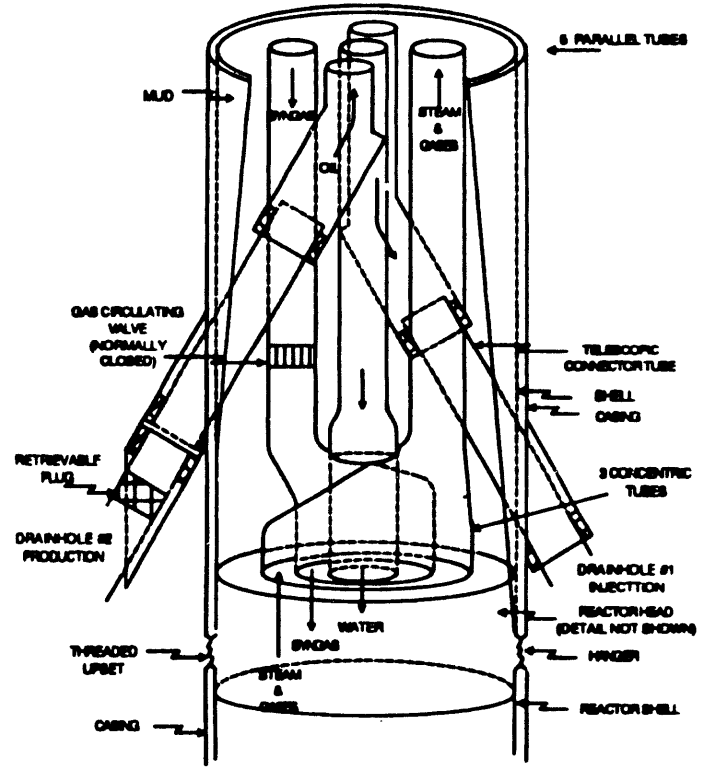


FIGURE 6
COMBINED LOWER CROSS-OVER AND CONNECTOR SECTION

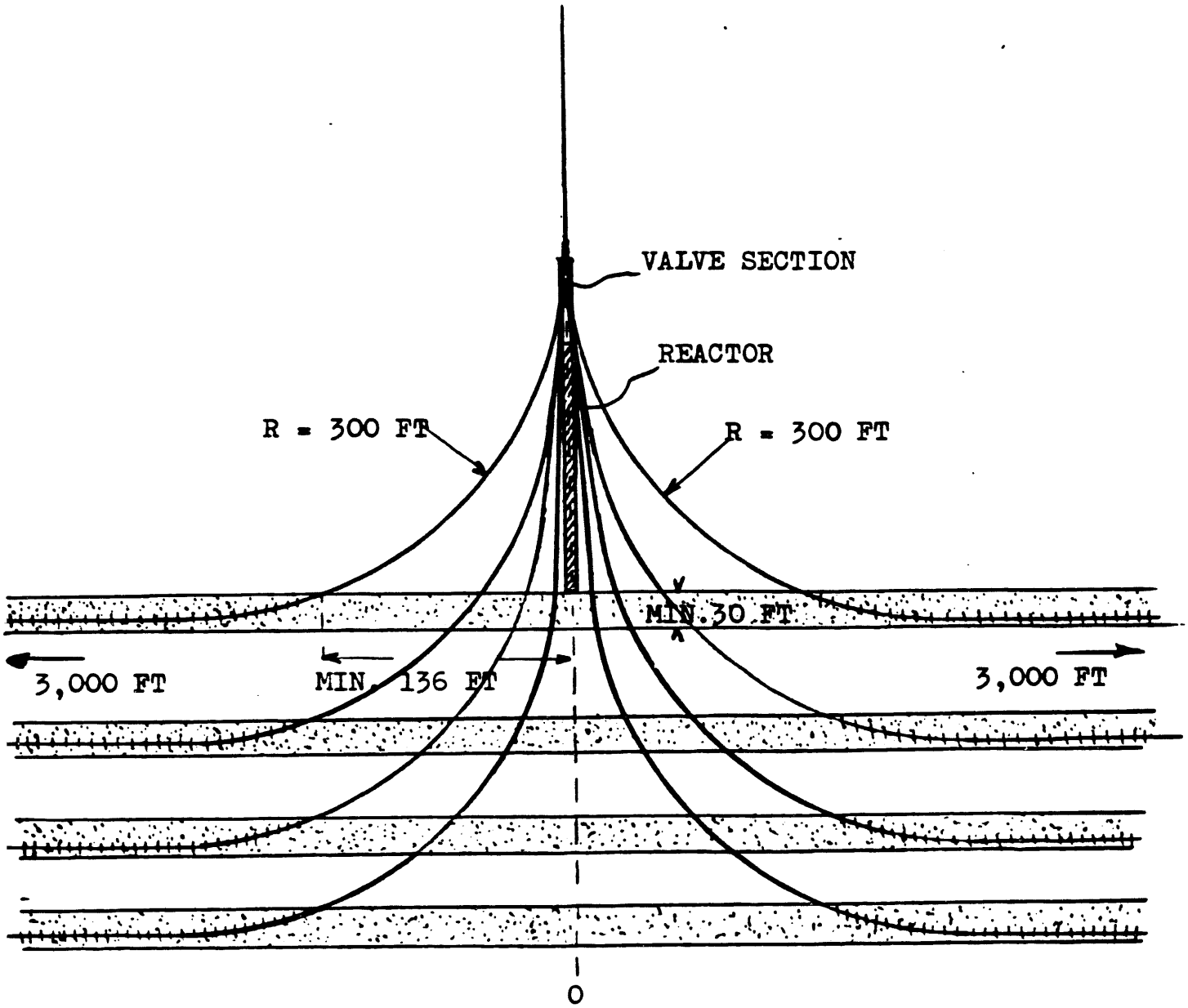
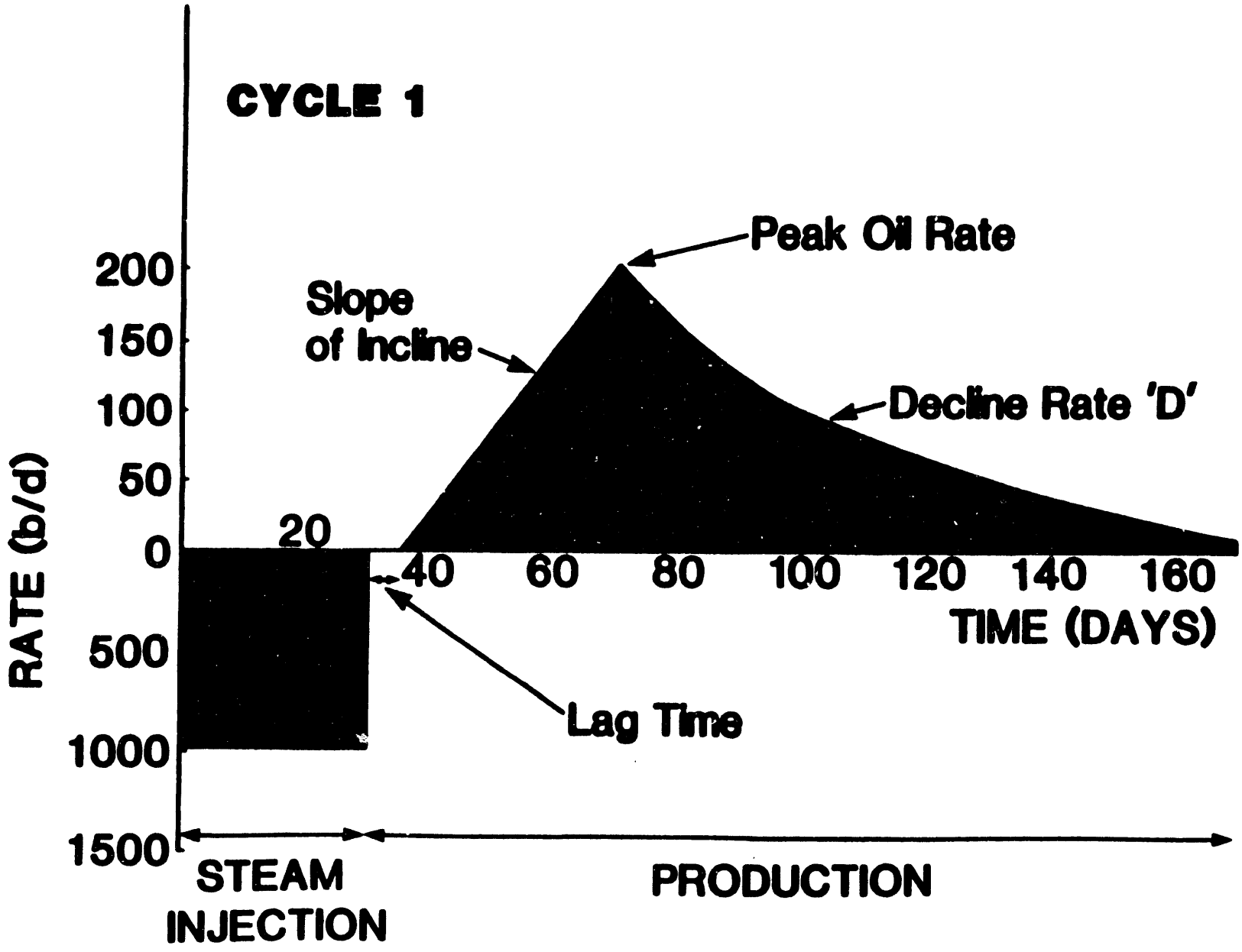


FIG. 7 DRAINHOLE CONFIGURATION

FIG. 8

TYPICAL PROFILE FOR AN INDIVIDUAL WELL DURING A STEAM STIMULATION CYCLE



APPENDIX 1:RESERVOIR DESCRIPTION

1.OBJECTIVES

The main objective of this contract is to evaluate the economic viability of a novel Heavy Oil Recovery Process (US Patent No.4,706,751) for a specific case,namely the West Sak heavy oil reservoir,located on the North Slope of Alaska (see Fig.1a)

1.1.BACKGROUND INFORMATION

This reservoir,of Cretaceous age,is estimated to contain 25 Billion Barrels of heavy oil in place.This heavy oil ,of 10.5 to 22.5 API gravity,lies at depths ranging from 2,500 to 4,600 ft (see Fig.2a). The reservoir is overlain by a 2,000 ft thick Permafrost zone.Its proximity to the reservoir results in low reservoir temperatures,ranging from 45 to 100 degrees F.A complete description of this extensive reservoir is given in Ref.1 (Panda et al.,1989). Six continuous sand layers have been mapped:the Upper Sand No.1 and Upper Sand No.2,separated by a 10 to 40 ft shale interval;the Lower Sand No.1 and Lower Sand No.2,separated by a 30 to 40 ft thick shale,and,finally,the thinner Lower Sands No.3 and No.4,separated from the Lower Sand No.2 by a thick (50 to 100 ft) shale section containing minor sand lenses (see Fig.3a).

The gross pay thickness of each of the two Upper Sands and that of the Lower Sands No.1 and 2 exceeds 25 ft over very large areas (70 to 300 square miles)for each of those four main oil sands.Average porosities are high (>30%) and the oil saturation is 60 to 70%.The

rock characteristics,given in Ref.1 (Panda et al.,1989)and Ref.3 (Kamath et al.1985),show that this reservoir is a fine-grained quartzitic sandstone,very friable,containing Illite clay minerals, Glauconite and Siderite.It exhibits a medium permeability (<150 md).

Under the criteria established by Taber et al.(Ref.10) to serve as technical screening guides for various Enhanced Oil Recovery processes,the conditions required for a successful steam flood are:

- a sand or sandstone of high porosity,

- an oil saturation exceeding 40%,
- a net pay thickness exceeding 20 ft,
- a transmissibility exceeding 100 md.ft/cp.

The viscosity of a 19.2 API sample of West Sak oil,given in Ref.4 (Sharma et al.,1989)is 35.4 cp at a reservoir temperature of 80 degrees F.For a permeability of 150 md,the transmissibility of each of the four top layers,of thickness exceeding 25 ft,is equal to:

150 md x 25 ft / 35.4 cp = 106 md.ft/cp

Accordingly,each of the four layers considered would satisfy Taber and Martin's technical criteria for successful steam flooding.

The high reservoir pressure of those deep sands (1,150 to 2,200 psi) is,however,too high for an economic application of steam injection by conventional techniques.

1.2-TUBING HEAT LOSS ELIMINATION REQUIREMENT

Ref.10 (Taber et al.,1983) also states that "because of excessive heat losses in the wellbore,steam flooded reservoirs should be as shallow as possible".Indeed,at present energy prices,most steam floods become uneconomic at depths below 2,000 ft.

Tubing heat losses for the West Sak must also be reduced to a very low level for important technical reasons,namely that it is essential to prevent melting of the Permafrost.With the well casing necessarily cemented to the Permafrost zone,any significant melting of this 2,000 ft thick zone would destroy the bond between cement and the surrounding frozen rock.This would cause the well casing to sink into the ground and would ultimately lead to the destruction of the well.According to Ref.5 (Hornbrook et al.,1989),the West Sak oil is undersaturated but contains a significant amount of dissolved gas (210 SCF/STB),which would be released in the event of mechanical failure of the well,further endangering the operating personnel and the fragile environment of the Arctic,where a well on fire could have catastrophic consequences.For this reason,the exploitation of the huge oil accumulation in the West Sak reservoir (the largest known oil field in the U.S.) cannot be achieved by conventional steam flooding,despite the fact that the technical criteria established by Taber and Martin (Ref.10)indicate that E.O.R. by steam injection would be successful under normal circumstances.

The heavy oil recovery process described in US Patent No.4,706,751 overcomes this difficulty by providing downhole the large amount of heat required to make the oil flow.This heat is supplied in the form of steam and hot gases by means of a Downhole Methanator.

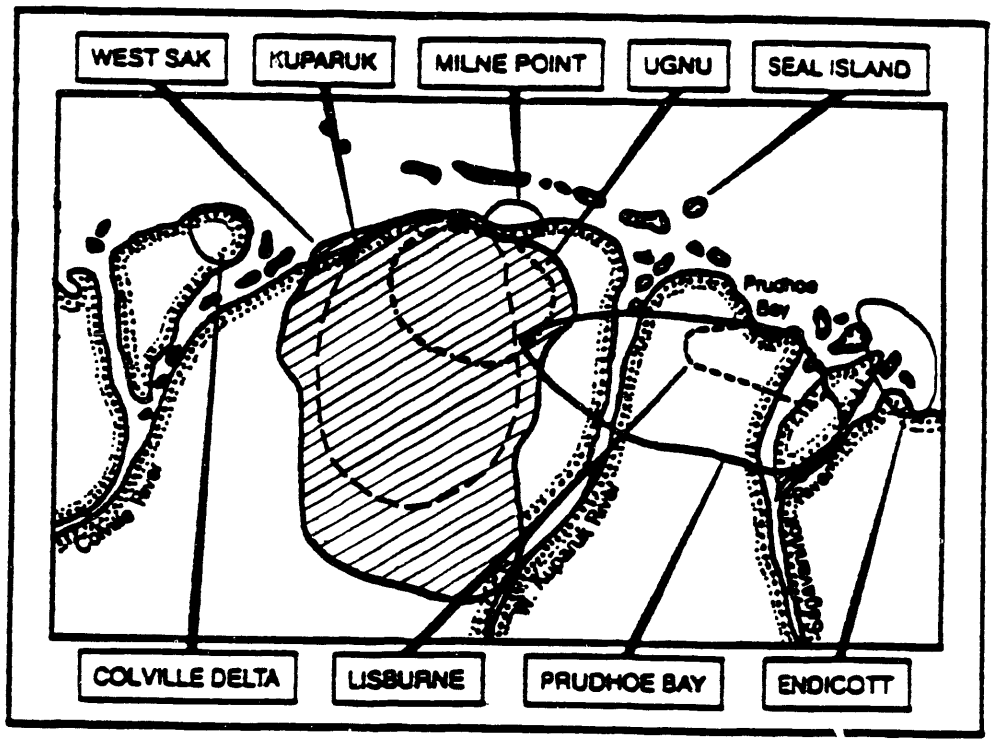


Figure 1: Index map showing major petroleum fields in northern Alaska

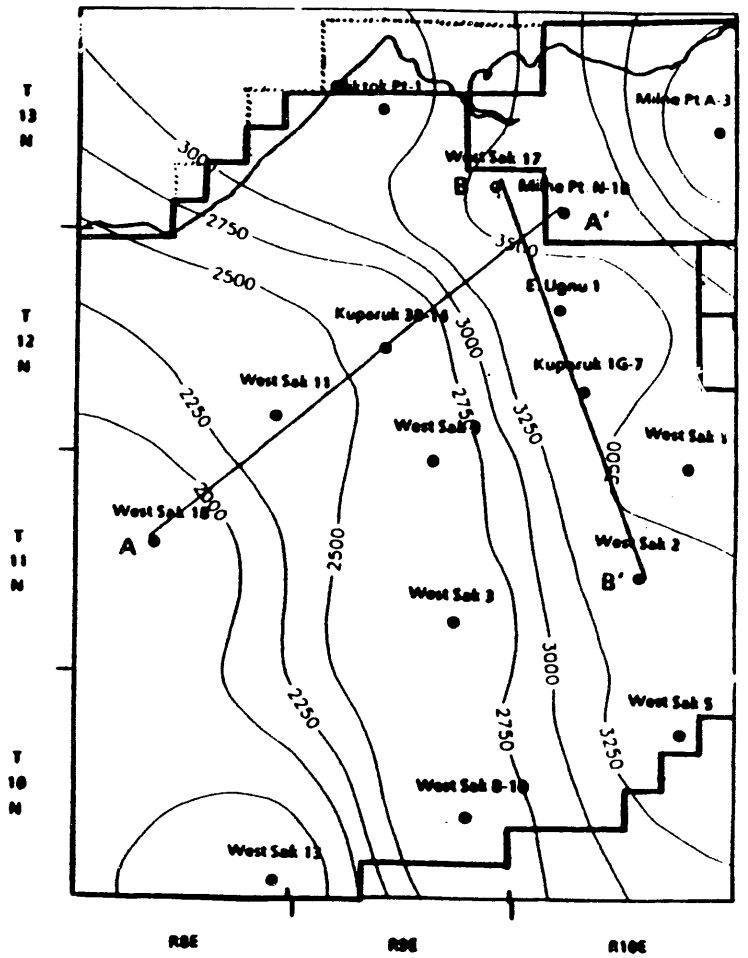


Figure 2

Contour Map Showing Depths to the Top of West Sak Upper Sand 1

(AA': SW-NE Cross-Section, BB': NW-SE Cross-Section)

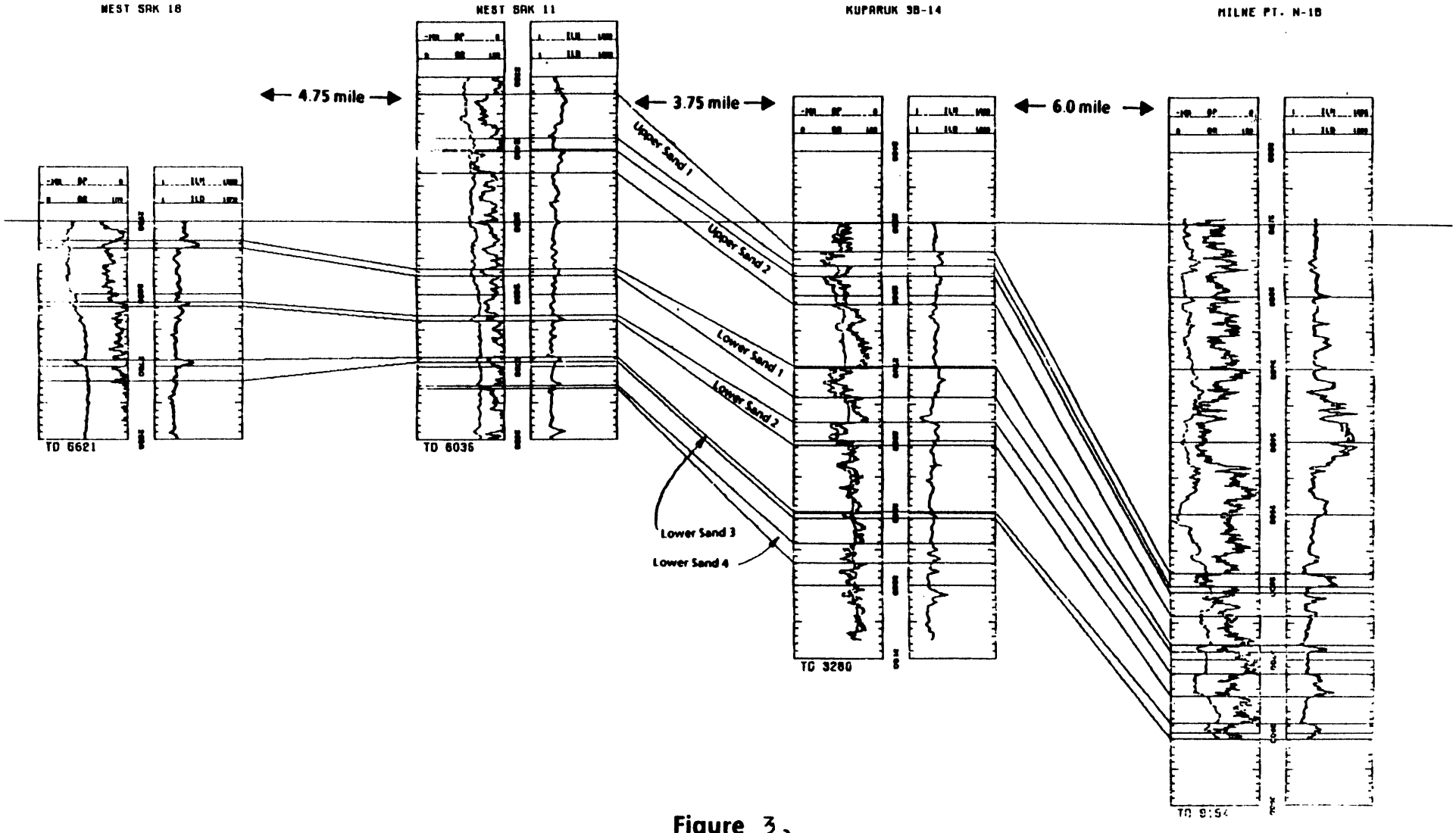


Figure 3a

West Sak Sands Depicted on the SW-NE Stratigraphic Cross-Section AA'

APPENDIX 2:DESIGN OF THE DOWNHOLE REACTOR

Contract DE-FG 01-89 CE 15446

4. TASK No.3:DETERMINATION OF HEAT TRANSFER FROM CATALYST TO FLUIDS IN THE DOWNHOLE METHANATOR

4.1-MAIN CHARACTERISTICS OF THE CATALYTIC REACTOR

In its most conventional form,the catalytic reactor consists of a vertical fixed bed,traversed by the reacting Syngas,and cooled by a set of water tubes immersed in the bed,constituting a once-through boiler.The steam and water effluent from the water tubes is then separated.The hot water,mixed with make-up cooling water is recycled. The steam from the separator is then flashed into the hot gas flowing out of the catalytic bed,to form the injected effluent.This mixture of steam and hot gases is then transferred,by means of appropriate downhole valves,from the Downhole Methanator to one of the horizontal wells connected to the large-diameter vertical well containing the Downhole Methanator.

The Methanation reaction takes place within the internal surface of the catalyst particles,where the heat of reaction is generated,making the catalyst the hottest part of the system.Two heat removal processes contribute to moderate the catalyst temperature:1)the transfer of heat from the external surface of the catalyst particles to the flowing gas

surrounding them;2)the transfer of heat from the catalyst bed to the water tubes immersed in it.

In order to analyze the heat transfer mechanisms involved in the cooling of the catalytic bed by these two mechanisms,it is necessary to establish first some of the dimensional characteristics of the reactor.

4.1.1-VOLUME OF THE CATALYST BED

For the prescribed feed rate of 100 MMSCF/D of Syngas,of composition: 3.5 H2,1 CO,the hourly volume of reactive feed is: (100 MMSCF/D)x(4/4.5) / 24 h/D = 3.70 MMSCF/h.

A typical hourly space velocity for commercial Methanation catalysts operating in the temperature range of interest (550 F to 800 F)with a stoechiometric feed (3 H2,1 CO) is: 10,000 SCF/h per cubic foot of catalyst bed.

Consequently,the volume of catalyst bed required is: (3.70 MMSCF/h) / (10,000 SCF/h / CF) = 370 CF .

4.2 REACTOR CONFIGURATION: A REVIEW OF VARIOUS POSSIBLE CASES

The catalytic bed configuration is always that of a cylindrical annulus of small cross section (1.5 to 2.5 sq.ft) and large total height (148 to 247 ft).

In order to assemble the reactor and to run it into the well, the total height must be divided into several sections of height smaller than that of the derrick (about 85 ft) of the rig used for drilling and completing the vertical well and its associated horizontal drainholes.

Each section may also be divided into several unit beds.Each

unit is cooled by its own set of helical-wound water tubes, all connected in parallel. Each tube in the unit delivers at the top a mixture of steam and hot water to the unit vertical steam-water separator. The water feed at the base of each tube is a mixture of hot water from the separator with a make-up stream of cooling water.

Regarding gas flow, the beds from various units may be connected in parallel, being supplied from a common Syngas feeder tubing, and discharging into a common collector space, into which the wet steam from each unit separator is flashed. The main purpose of the parallel connection is to reduce the pressure drop across each unit bed. This is necessary if gas flow through the beds is vertical (upward or downward). In the case of radial gas flow across the annular section of the bed, however, the pressure drop is small enough to give each unit bed the total length allowable for each section.

For ease of temperature control of each unit bed, these may be grouped in pairs, with both units in the pair connected in series for gas flow and in parallel for water flow.

This temperature control process, developed by A. Foss (Ref. 31, 1987) for autothermal reactors is based on the adjustment of the feed gas temperature for the second unit in the pair by mixing the effluent from the first unit with colder Syngas feed. This provides a faster response to any perturbation in the bed temperature than by adjusting the flow rate of the cooling water make-up in the tubes, but both controls may be used simultaneously in cooled reactors, such as those considered here.

The Methanator effluent may be injected into the horizontal drainholes either through the bottom end or through the top end.

In the first case, the horizontal drainholes are connected to the vertical casing by means of a conventional dual-tubing packer. In the second case, they penetrate into the vertical casing through windows. In either case, a system of surface-controlled retrievable valves, located at the Methanator outlet, allows the effluent to be switched from one drainhole to the other, and conversely allows the fluids produced from the previously injected drainhole to be sent to the production tubing ("huff and puff" mode of steam injection).

To illustrate the various flow options considered, Fig.9, 10, 11 and 12 show schematically the flow distribution of Syngas, water and effluent through the Downhole Methanator assembly, typically composed of four sections, stacked on top of one another.

In Fig.9, each section is shown to consist of a pair of unit beds in series, subject to the type of temperature control described in Ref.31.

Figures 9 and 10 pertain to vertical flow of gas through the beds (upward or downward).

Figures 11 and 12 pertain to radial gas flow through the beds.

The water tubes in each unit are always connected in parallel. Steam effluents from unit to unit and from section to section are also connected in parallel, giving greater uniformity to all units, but requiring that the cooling water tubing extend to the bottom of the Methanator assembly, even in the case of bottom effluent discharge (Fig.9 and 11). This, of course, reduces the cross sectional area of the beds.

The top and discharge configuration makes the valve section directly accessible either from the the mud-filled annular space or through the central production tubing. This flexibility is obtained at the expense of a more complex connection (windowed casing) between the

vertical casing and the horizontal drainholes. For the same reservoir depth, it also requires a 300 ft deeper vertical well, but the unavoidable heat losses through the casing, adjacent to the reactor, are then contributing heat to the oil zone rather than to the overlaying shale.

The bottom end discharge configuration requires the valve section to be located at the base of the Methanator assembly, with access to the valves only through the production tubing.

The radial flow configuration (Fig.11 and 12) provides a very low pressure drop through the relatively thin annular bed, but makes the bed temperature control more difficult, since the reaction rate must be controlled over a few inches of bed thickness, rather than over about 20 to 30 ft.

In all cases, the Syngas feed is pre-heated first by heat exchange (in counter-current) against the hot produced fluids, then by heat transfer from the bed, through the reactor shell, and, finally by heat exchange with the Methanator effluent through the Syngas feeder tubing wall (Fig.9 and 10) or through the inlet screen (Fig.11 and 12).

Conversely, the cooling water is also pre-heated by heat exchange from the warmer fluids from which it is separated by a tubing wall, successively Syngas and produced fluids.

It is clear that pre-heating the cooling water from an external heat source, such as the produced fluids, results in an increase of the total heat rate delivered to the reservoir, a beneficial effect. Most of the heat in the water stream at the tubes inlet, however, is due to the recycled hot water from the steam separators.

This quick comparative look at the various cases shows that their relative merits and costs must be evaluated quantitatively in order to

optimize the configuration of the Methanator assembly.

4.3-DATA REQUIRED FOR HEAT TRANSFER CALCULATIONS

The coefficient of heat transfer from the particles of a fixed bed to the gas flowing through the bed, and the coefficient of heat transfer from the bed to the internal water tubes or to the feed gas tubing or screen in contact with the bed can all be derived from published correlations, based on experimental data.

The variables in these correlations are respectively:

-the Reynolds Number: $Re = v.Dp.\rho f / \mu f$,

-the Thermal Conductivity: Kf of the flowing gas phase,

- the Heat Capacity: Cp' of the catalyst particles,

where: Dp is the catalyst particles mean diameter (ft.),

v is the superficial velocity of the flowing gas phase (ft/s) ($v = Q/S$), where Q is the gas volume flow rate, in ACF/s at the reactor conditions, and S is the total cross-sectional area of the catalytic bed (sq.ft),

μf is the flowing gas viscosity (#/ft.s)

ρf is the flowing gas specific gravity (#/CF) at reactor conditions.

When Kf is expressed in BTU/h.ft.°F

and Cp' is expressed in BTU/#.°F,

the heat transfer coefficients hc will be expressed in BTU/h.sq.ft.°F.

The derivation of μf, Kf, ρf is given in Appendix 2.1 for the following typical conditions:

-Syngas feed rate: 100 MMSCF/D

-Bed cross sectional area: 2.5 sq.ft and 1.5 sq.ft

-Inlet gas conditions: 620°F,2,675 psia

-Outlet gas conditions:680 to 800°F,2,575 psia

The values of Re are also given for the average values of these parameters and for two sizes of catalyst particles:

Dp = 0.004 ft and Dp = 0.008 ft

According to Ref.37 (Zenz and Othmer,1960),the heat transfer coefficient from bed to water tube is maximum when the ratio of particle diameter to tube diameter is: Dp /Dt = 0.15

Correspondingly,two water tube diameters were considered:

Dt = 5/8 inch and 5/16 inch O.D.

For the pressure and temperature conditions considered,the required internal diameters of the tubes,for commonly available alloy steels are: 3/8 inch and 3/16 inch respectively.

The results are given in Table 1a

The Nusselt Number,Nu, which determines the heat transfer coefficient,hc,from the catalyst to the flowing gas is defined as:

Nu = hc.Dp/Kf

Other dimensionless numbers and dimensional parameters involved in expressing the heat transfer coefficient,hc',from the bed to the external surface of the water tubes are:

ϵ = void fraction in the bed.A representative value for fixed beds is:

ϵ = 0.35

Kf/(Cp.ρf),where Cp is the heat capacity of the flowing gas.A representative value is Cp = 9/(average mol weight of gas),or,in our case: Cp = 9/[(7.77 + 12.75)/2] = 0.876 BTU/#.°F.Consequently

Kf/(Cp.ρf) = 0.0929 / (0.876 x 2.215) = 0.0496 sq.ft/h

Cp/Cs ,where Cs is the heat capacity of the catalyst particles.A representative value is Cs = 0.8 BTU/#.°F.Consequently Cp/Cs = 1.095

$\rho f/\rho p$,where ρp is the specific gravity of the catalyst particles. A representative value is 200 #/CF

4.4-CALCULATION OF HEAT TRANSFER COEFFICIENT FROM CATALYST TO GAS:

Zenz and Othmer,Ref.37,1960 have compiled various experimental data for particle to fluid heat transfer in the form of a set of curves correlating the Nusselt Number versus the Reynolds Number,for various values of the void fraction ϵ(see Fig.13).

For ϵ = 0.35,the range of Nusselt numbers corresponding to the previously determined range of Reynolds Numbers (from Re = 565 for S = 2.5 sq.ft to Re = 1950 for S = 1.5 sq.ft) is:

- Nu = from 130 to 280 for these two cases,respectively.The corresponding values of hc = Nu x Kf /Dp are:
- from hc = 3,136 BTU/h.sq.ft.°F for S = 2.5 sq.ft to hc = 3,251 BTU/h.sq.ft.°F for S = 1.5 sq.ft.

The total external surface area of the catalyst particles in each section of the Methanator is expressed by:

S x L x (1 - ϵ) x (6/Dp) = 487.5 x (total bed volume) for Dp = 0.008

The total bed volume,given in 4.11 above is 370 CF.Consequently the catalyst surface in each section is: 180,375 sq.ft / 4 sect.= 45,094 sq.ft / section for Dp = 0.008 ft.

Correspondingly,the total heat transferred in each of the four sections of the bed to the flowing gas phase ranges from

141 MMBTU/h to 147 MMBTU/h for each °F of temperature difference between the catalyst and the gas phase.Because the total heat to be transferred in pre-heating the Syngas feed is only 25.8 MMBTU/h for the entire reactor,or 6.45 MMBTU/h per section,it is clear that all of this heat can be transferred from the catalyst,but most of it

will in fact be transferred from the reactor shell and feed gas tubing, leaving most of the reaction heat to be transferred to the water tubes. This will be analyzed in Tasks No.4 and No.5.

4.5-HEAT TRANSFER FROM THE CATALYTIC BED TO THE WATER TUBES FOR VERTICAL GAS FLOW:

Zenz and Othmer, in Ref.37, 1960 also provide a correlation for heat transfer from the catalyst bed to the water tubes immersed in it. The correlation, developed by Leva et al.(1952), is expressed as:

$$hc' = 3.50\ (Kf/Dt) \times (\exp.-4.6\ Dp/Dt) \times [Re]^{0.7}$$

for fixed beds cooled by water tubes or jackets. To maximize the heat transfer, it is shown that Dp/Dt should be about 0.15, according to Ref.37.

Dp/Dt = 0.154 for all cases considered, which simplifies the expression of hc' to:

$$hc' = 1.72\ (Kf/Dt) \times [Re]^{0.7}$$

TABLE 1a

T	S	Dp	Dt	Re	Kf	hc'(BTU/h sq.ft °F)
680	2.5	0.004	0.026	585	0.0929	531
800	2.5	0.004	0.026	565	0.0965	539
680	2.5	0.008	0.052	1170	0.0929	432
800	2.5	0.008	0.052	1130	0.0965	438
680	1.5	0.004	0.026	975	0.0929	760
800	1.5	0.004	0.026	942	0.0965	771
680	1.5	0.008	0.052	1950	0.0929	617
800	1.5	0.008	0.052	1884	0.0965	629

The temperature gradient at the contact between the catalyst bed and the water tubes may be as much as 200 °F, so that the outer surface area of the tubes required for transfer of 122.2/4 = 30.6 MMBTU/h from each section of the reactor may be as little as:

(30.6 MMBTU/h)/(760 BTU/h sq.ft.°F x 200°F) = 201.3 sq.ft/section

For this case, the outer diameter of the water tubes is 5/16 inch, or 0.0817 ft in circumference, the total length of water tubes per section is 2,464 ft. The length of the section in that case is at least 61.75 ft, and the length of each helical tube may be about 120 ft. This would require less than 21 tubes in parallel to provide the needed surface of heat transfer. The total water tube volume per section remains a very small fraction of the catalyst bed volume:

$[21 \times \pi \times (0.013\ ft)^2 \times 120\ ft] / (370\ CF / 4) = 1.5\ \%$

Even for the larger cross section of bed (1.5 sq,ft,hc' = 509), this conclusion applies to the design concept of vertical flow reactors. The required surface area of 5/16 in.OD tubes becomes 283.8 sq.ft. For a bed length of 37 ft, the length of each water tube in helix is only about 75 ft, so that a larger number of tubes in parallel is required: 47 tubes, of 5/16 in.O.D., whose volume represents about 2 percent of the catalytic bed volume.

This confirms the feasibility of this rather conventional reactor design concept for vertical flow. Further optimization of this design depends on the option selected: discharge at the bottom end or at the top end of the Methanator assembly, diameters of the respective tubings, which largely determine the heat transfer to the cooling water, and diameters of the casing, reactor shell and collector wall, which control the pre-heating of the Syngas feed. This will be

part of Task No.5,which will analyze the water tube system as that of a boiler,and determine additional design parameters.

4.6-GAS PRESSURE DROP IN EACH SECTION FOR VERTICAL FLOW:

The correlation of Rose and Rizk,also given in Ref.37,gives the friction factor $f = (g.Dp.\Delta P) / (L.\rho f.v^2)$ as a function of the Reynolds Number for $\epsilon = 0.4$ and provides a correction factor for other values of ϵ(see Fig.14 and 15).Wall effects may also be taken into account. For the values of Re = 565 to 1950 previously considered,the respective values of f are 21 and 18,with a correction factor of 1.5 for $\epsilon = 0.35$.The corresponding pressure drops in each section are:

ΔP = 187 psi for S = 2.5 sq.ft, Dp = 0.004 ft and

ΔP = 374 psi for S = 1.5 sq.ft, Dp = 0.008 ft

The assumed Syngas feed pressure of 2,675 psia,corresponding to a 100 psi pressure drop is a little too low for these extreme conditions.This could be remedied by either increasing the bed inlet pressure,at the cost of a bigger compressor for the Syngas feed,or by increasing the number of sections from 4 to 6,at the cost of greater complexity in the Methanator assembly.

A simpler option is to consider the intermediate case for coarser particles and larger tubes:

Re = 1170 to 1130 for Dp = 0.008 ft and S = 2.5 sq.ft

The friction factor is then f = 19 x 1.5 and ΔP = 85 psi,well within the assumed 100 psi pressure drop in each of the four sections. This,however,reduces the heat transfer coefficient hc'from 760 to 438 and correspondingly increases the surface of water tubes required from 201.3 sq.ft to 349.3 sq.ft.The corresponding number of 75 ft long

tubes of 5/8 inch OD is: 29 tubes.

A more detailed analysis of this case will be presented as part of Task 5.

4.7-HEAT TRANSFER FROM CATALYST TO GAS IN RADIAL FLOW:

For radial flow, a coarse particle size may be preferable. A value of Dp = 0.008 ft is assumed.

The average superficial velocity, corresponding to the average gas composition, with an average shrinkage of (1 + 0.5644) x 0.5 = 0.7822 is Q/S for a flow rate at 2,575 psia, 680 F in each section of: Q = 100 MMSCF/D x 0.7822 x (1,140 R/492 R) x (14.7 psia/2575 psia) x 1/4 = 0.258 MMACF/D = 2.9 ACF/s.

Assuming that the mid-section S of the annular bed has a radius of 9", the flow velocity in a 37 ft tall section is only:

v = Q/S = 2.9 ACF/s / (π x 1.5 ft x 37 ft) = 0.016 ft/s, about 1/60th of that in a vertical flow reactor of same dimensions.

The corresponding Reynolds Number at 680 F is:

Re = v.Dp.ρf / μf = 0.016 ft/s x 0.008 ft x 2.291 #/CF / 1.619 (10.-5)#/ft.s = 19

The Nusselt Number representing the heat transfer from catalyst to gas is now: Nu = 13

The heat transfer coefficient is:

hc = 13 x 0.0929 BTU/h.ft.°F / 0.008 ft = 151 BTU/h.sq.ft.°F, but the surface of the catalyst is now:

370 CF x (1 - 0.35) / (4 x Dp/6) = 45,094 sq.ft. per section

The heat transferred from catalyst to gas in each section is then:

45,094 sq.ft x 151 BTU/h.sq.ft.°F = 6.8 MMBTU/h.°F, which is 105 % of the Syngas preheating load if the temperature gradient between

catalyst and gas is only one degree F.Accordingly,most of the catalyst heat will again be transferred to the water tubes.

4.8-HEAT TRANSFER FROM THE CATALYST BED TO THE WATER TUBES WITH RADIAL GAS FLOW:

Leva's correlation,obtained for a range of 250 < Re < 3,000,is not applicable in this case,but the experiments of Plantz,Felix and Yagi,extending down to a Reynolds number of 20,provide a valid correlation,also given in Ref.37.

For a Reynolds Number of 19,and a ratio Dp/Dt = 0.154 (5/8 inch OD tubes),the transfer of heat from the bed to the water tubes is read directly from this correlation.

The heat transfer coefficient,hc',is:

hc' = 24 BTU/h.sq.ft.°F

For the same heat load (30.6 MMBTU/h/section) and the same assumed LMTD (200°F),the required surface area of the tubes is now:

30.6 MMBTU/h / (24 BTU/h.sq.ft.°F x 200 °F) = 6,375 sq.ft/section.

The required surface area is nearly 22 times that required for vertical gas flow reactors,making it unlikely that water tubes can still be used for cooling the bed in radial flow.

The cross section area of an annular bed of diameters 12 and 24 inches (6 inch bed of average diameter 18 inches) is 2.356 sq.ft and the total height of the bed is:370 CF / 2.356 sq.ft = 157 ft. Under the present assumption of 4 sections,each section will be 40 ft tall and each helical water tube may be 100 ft long.With an external surface of 0.1634 sq.ft/ft,the required surface may be provided by 390 tubes of 5/8 in. O.D.,which is excessive.The tube volume represents 89 % of the catalyst volume,or 47 % of the total bed volume.

In view of the low gas pressure drop,it may be possible to reduce the number of sections from 4 to 3 ,thus reducing the complexity of the Methanator assembly.Contrary to the case of vertical gas flow reactors,this does not change the Reynolds number,nor the heat transfer coefficient,nor the pressure drop.The tube surface required in each of the three 53 ft-tall sections would,however, increase to 8,500 sq.ft,again occupying 47 % of the total bed volume.

The number of 5/8 inch O.D. water tubes remains the same.Another water cooling process,by direct flashing of water into the hot flowing gas phase would be preferable to reduce the reactor volume.

4.9-PRESSURE DROP IN REACTORS WITH RADIAL FLOW:

The same correlation,from Rose and Rizk,gives a friction factor which,for the value of the Reynolds Number of Re = 19,is equal to:

$f = 90 \times 1.5 = 135$

The corresponding pressure drop,for a typical bed thickness of 6" and 40% void fraction is:

$\Delta P = 135 \times 0.5$ ft $\times$ 2.215 #/CF $\times$ $(0.016 \text{ ft/s})^2$ /(32.2 ft/s.s $\times$ 0.008 ft $\times$ 144 sq.in/sq.ft) = 0.002 psi .The inlet pressure is practically the same as the discharge pressure of 2,575 psia.The assumption of a flowing gas phase at 2,575 psia in the calculation of the volume flow rate is therefore justified.

4.10-CONCLUSIONS OF TASK No.3

1)-Within the geometry of a large-diameter cased well,the preceding analysis shows that it is feasible to quickly bring the feed gas to reaction temperature by heat transfer from the catalyst particles.

2)-Most of the heat of reaction will be transferred to the cooling

water.

3)-With a reactor in vertical gas flow consisting of 4 sections connected in parallel,the gas pressure drop in each 37 ft section is less than 100 psi,if the bed cross section is 2.5 sq.ft and the catalyst particle size is 0.008 ft.

4)-In such a vertical flow reactor,the entire heat load required to vaporize the cooling water may be transferred to a relatively small number of 5/8 inch O.D. water tubes,occupying a small fraction of the reactor volume.

5)-With a reactor in radial gas flow,the gas pressure drop is negligible,which may lead to a reduction in the number of parallel sections to three instead of four,thus reducing the complexity of the reactor assembly.

6)-Because of the much lower heat transfer coefficient in the radial gas flow reactor,the number of 5/8 inch O.D. tubes required to transfer the full heat load to the cooling water becomes excessive, since they occupy a large fraction (47 %) of the total bed volume.

7)-Additional design parameters will be determined from an analysis of the water tube system as a boiler.This will be done in Task No.5.

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5. TASK No.4-CALCULATION OF OIL PRODUCTION RATE UNDER VARIOUS ASSUMPTIONS :(see Appendix 3)

5.1.-EFFECT OF OIL PRODUCTION RATE ON THE REACTOR DESIGN:

A rough estimate of the production rate to be expected from a horizontal well extending 2,000 to 3,000 ft into one of the West Sak sands of thickness exceeding 25 ft is necessary to proceed with the

conceptual design of the Downhole Methanator.This is required for the following reasons:

1) in the cases where the injected effluent is discharged at the bottom of the Methanator assembly,the production tubing occupies the central part of the assembly,thus limiting the space available for the reactor proper.

2) the fluids produced from the well,after a period of injection with steam and hot gases,are very hot.In their flow to the surface they exchange some of their heat with the Syngas feed and cooling water streams,thus determining the temperatures of these streams at their entry into the reactor.

3) the oil production rate is an essential part of the economic evaluation of the entire process,because the resulting cash flow is the only income offsetting the capital and operating costs of the process.

The total liquid rate is estimated at 18,000 to 24,000B/D,of which about 1/3 may be oil and 2/3 water.Determining the size of the production tubing for this large flow rate will require a careful analysis,which will be conducted as part of Task No.6. (see Appendix 4) The GOR,corresponding to 80% recovery of the injected Methane is 2,430 SCF/STB.The corresponding GLR is then:2,430 / 3 = 810 SCF/B.With proper tubing size,this appears to be sufficient to lift the liquid stream to the surface.

The productivity of a horizontal well is largely dependent upon the number of fractures intersected by the well.In the absence of fractures,it increases with the horizontal reach of the well and with the ratio of vertical to horizontal permeabilities.No information is available on either subject and no horizontal well has yet been

drilled in the West Sak reservoir. In the event that a single producer of maximum reach is insufficient, several horizontal drainholes connected together may be used instead. High pressure steam is often used to create fractures or to open existing ones. The known presence of many fractures and faults in the underlying Kuparuk reservoir makes it very likely that horizontal wells in the West Sak would be capable of producing low viscosity fluids (hot hydrogenated oil and water) at high rates. The cost of additional horizontal drainholes is relatively low, compared to that of the large-diameter vertical hole, casing, Methanator and vertical tubings. Consequently, the uncertainty regarding the productivity index of each horizontal drainhole is not of major consequence for the economic viability of the process in the West Sak.

A complete reservoir engineering analysis, which is beyond the scope of this contract and which would require more specific data on the area of the West Sak field subjected to a pilot test of this novel process, would be required to arrive at a better estimate of the oil production.

5.4-HEAT RECOVERED FROM THE PRODUCTION STREAM:

The temperature of this oil/water/gas mixture flowing from the horizontal well after injection of steam at 620 F may be 450 to 550 F These hot fluids, when brought to the surface, may exchange heat through the wall of the production tubing. Assuming that the well-head temperature of the production stream is 320 F, corresponding to an average temperature drop of 180 F, the heat content of 21,000 B/D of oil and water which could be transferred to the Syngas feed by convective heat exchange, in counter-current, is about:

Rate ρ_{av} Cp $(T_1 - T_0)$ ΔH ΔH

	(B/D)	(#/B)	(BTU/#.F)	(F)	(MMBTU/D)	(MMBTU/h)
Oil	>= 7,647	270	0.50	180	186	7.7
Water	>= 13,353	286	1.12	180	770	32.1
total	>= 21,000				total:	39.8

The enthalpy of the produced gas volume (0.8 x 21.8 MMSCF/D) has been omitted because most of it is ultimately spent in isentropic expansion,to produce the work of lifting the liquids to the surface. This additional heat,recovered from the heated reservoir,is used to pre-heat the Syngas feed and cooling water.It is essential that the Permafrost be protected from this important heat source,and there is no better way than by capturing the heat,by counter-current exchange with the fluids sent to the downhole Methanator.

The feasibility of transferring this amount of heat (39.8 MMBTU/h) through the tubing walls can be verified quickly:
Typical tubing diameter for the production tubing is 9.5/8 in.for a 3,000 ft depth.The over-all heat transfer coefficient is about 80 BTU/h sq.ft.F.Correspondingly,the LMTD required for this heat exchange is:
39.8 MMBTU/h / [80 BTU/h sq.ft.F x π x(9.625"/12"/ft) x 3,000 ft] = 65.8 F.In reality ,due to the large temperature differences between the various streams,the LMTD will be larger than that required.This shows that the concentric well tubings constitute a very good heat exchanger system for the recovery of this heat,representing 23% of the previously injected heat.

5.5-HEAT BALANCE IN THE METHANATOR:

The recovery of this heat,otherwise wasted,is included in the

over-all heat balance:

Methanation reaction heat @ 620 F:	223.6 MMBTU/h
Heat recovered from produced fluids:	33.3
Pre-heating Feed gas (300-620 F):	(25.3)*(to 620 F
Enthalpy of non-condensible gases:	(26.1)*
Enthalpy of reaction steam:	(49.5)*
Heat added to cooling water:	162.0
Enthalpy of injected steam:	188.1
Enthalpy of injected effluent @ 620 F	214.2
Cooling water rate: (ΔH = 1,085 BTU/#)	149.300 #/h

(N.B.After optimization of surface facilities which increases by 15 % the feed rate,but leaves 6% of CH_4 and a lower H_2/CO ratio in the feed,the Bechtel simulator shows that the total heat available for injection may be increased by about 3 % to 220.6 MMBTU/h/well,per Table 5). When all the minor differences in the assumptions on which both independent estimates were made are considered,this shows an excellent agreement.

The effluent,steam and water, from the water tubes must be at a pressure higher than that of the bed effluent (2,575 psia). Assuming that there is a 50 psi pressure drop in the throttling steam valves,the tube outlet pressure is 2.625 psia,corresponding to a vaporization temperature of 675 F.

The temperature of the [illegible] must [illegible] than 675 F in [illegible] thermal resistances from bed to steam [illegible] heat transfer from bed to tube,metal heat conduction resistance, boiling heat transfer and forced convection within the tube.

The enthalpy of steam at 675 F is 1,080 BTU/#,but that of steam at

620 F.the equilibrium temperature of steam in the injected effluent,is 1,153 BTU/#,higher than that of the steam flashed from the steam separators.This means that some of the steam is condensed during the flash.The droplets of condensed steam may,however,re-evaporate if additional heat is available to offset the deficit (73 BTU/# x cooling water rate #/h >= 8.2 MMBTU/h).Two sources of heat are present:

1) the effluent from the beds is at a temperature T higher than 675 F: it will give off the corresponding sensible heat.Assuming that T =< 800 F,this sensible heat is =< 13.8 MMBTU/h ,which is more than enough to re-vaporize the flash condensate in the reactor effluent.

2)the wall of the bed is also at a temperature T higher than 675 F and transfers heat to the effluent flowing against it in the collector space.

6-TASK No.5-REACTOR CONFIGURATION AND DIMENSIONS:

6.1-THE WATER TUBES SYSTEM AS A HIGH PRESSURE BOILER:

A major concern in the design of high pressure boilers is to prevent tube burnout due to excessive tube wall temperature.For safe operation,it is necessary to avoid the Departure from Nucleate Boiling (DNB) regime,which,at high pressure,may occur at relatively low steam quality values.This unstable regime,corresponding to the phenomenon of film boiling,is characterized by a large increase in the tube wall temperature,relative to that of steam,as shown in Ref.40 (Babcock and Wilcox,1972).The same reference indicates that,at a pressure of 2,625

psia and for heat flux of 40,000 BTU/h.sq.ft of internal surface,the maximum steam quality obtainable safely is only about 33% (see Fig 17).This calls for recycling 67% of the water and increases nearly threefold the water flow rate through the tube system.

The total cooling water rate being 149,300 #/h,the total water rate through the tube system would be,for 3 passes:
3 x 149,300 = 447,900 #/h.(The minimum value for safe operations would be:149,300 / 0.33 = 452,424 #/h,consequently this includes a 1% margin of safety).

Ref.39 also provides the limit between safe and unsafe operation as a function of the water tube specific heat flux,in BTU/h.sq.ft of inner surface,at an even higher pressure of 2,700 psia.

These data were used in a proprietary boiler design program to determine the number of tubes required in each bundle to bring the water mixture (BFW and recycle) to the vaporization temperature of 675 F and to vaporize it under nucleate boiling conditions.It is essential to eliminate the risk of getting into the unstable regime of film boiling in order to prevent tube burn-out.

In reality,the total heat load is made up of the vaporizing load in the upper part of the tube and of a water pre-heat convective heat transfer load,in the bottom part,which must be analyzed separately.

It also shows the interdependence of the heat transfer from bed to outer tube surface with that from inner tube surface to water in nucleate boiling and with the water pre-heat.A numerical solution including all these factors,as well as the thermal resistance of the metal tube wall was used for this purpose.

6.2-HEAT TRANSFER BY CONVECTION IN THE PRE-HEAT PORTION OF

THE TUBE BUNDLES:

Heat transfer,by convection,to water flowing through the water tubes is determined by the Reynolds Number for the tube:

$V.Dti.\rho w/\mu w$,where:

- V is the water velocity in the tube (ft/s)
- Dti is the tube internal diameter,first assumed to be 3/8 th of an inch, or 0.031 ft
- ρw is the specific gravity of the flowing water (#/CF)
- μw is the viscosity of water in the tube(#/ft.s).

It also depends upon the Prandtl Number,$Cw.\mu w/Kw$,where:

- Cw is the specific heat of water(BTU/#.°F),
- Kw is the thermal conductivity of water (BTU/h.ft.°F)

For the assumed average conditions in the tube:600 °F,3,000 psia,these parameters have the following values:

ρw = 42.4 #/CF

μw = 0.058 x (10.-3)#/ft.s

Kw = 0.292 BTU/h.ft.°F

Cw = 1.51 BTU/#.°F

The corresponding Prandtl Number is:

Pr = 1.51 BTU/#.°F x 0.058 x (10.-3)#/ft.s x 3600 s/h / 0.292 BTU/h°F

Pr = 1.080

Assuming that the tube bundles are all connected in parallel,the water flow in each bundle is:447,300 #/h / (3600 s/h x 8 units) = 15.55 #/s per unit,and the volumetric water flow rate is:

15.55 #/s / 42.4 #/CF = 0.367 CF/s per unit bundle.

For comparison,it is assumed that the number of tubes in each bundle is 100,for a maximum heat flux of 40,000 BTU/h.sq.ft,

V = 0.367 CF/s / [100 x 0.00076 sq.ft] = 4.8 ft/s

The Reynolds Number can then be calculated, for water at 675 F.
W.M.Mc Adams, in Ref.34, 1964 provides a correlation:

$$hc'' = 0.023 \times (Kw/Dti) \times Pr^{0.33} \times Re^{0.8}$$

from which the convective heat transfer coefficient can be calculated (in BTU/h.sq.ft°F) from the inner tube surface to the water stream.
The over-all heat transfer coefficient U, from bed to water is:

$$1/U = 1/hc' + (So/Si)*(1/hc'') + Rm + Rfo + (So/Si)*Rfi$$

where So and Si are respectively the outer and inner surfaces of the tubes and Rm, Rfo and Rfi are respectively the thermal resistances of the metal tube and its fouling resistances outside and inside.

So/Si = 0.1634 / 0.098 = 1.667

Rfo = 0.001

Rfi = 0.001

Rm = 0.0006

hc' = 438 BTU/h sq.ft F for T = 800 F (see section 4.5 above)

6.3- HEAT TRANSFER BY NUCLEATE BOILING IN THE UPPER PART OF THE TUBE

Thom's correlation (Ref.41) is used for the part of the tube in the forced nucleate boiling regime.
It relates the heat rate per unit area of the inner tube surface to the square of the difference between the inner surface temperature and the boiling temperature.

6.4-NUMERICAL SOLUTION OF THE VARIOUS HEAT TRANSFER FUNCTIONS OF THE TUBE SYSTEM:

Nomenclature:

n = number of tubes per unit

Q1 = heat rate (BTU/h) from bed to tube in each unit

q1 = heat flux (BTU/h sq.ft of inner surface) in vaporizing regime

Lv = % length of tube in vaporizing regime

Lc = % length of tube in convection regime = 100% - Lv

Ktm = thermal conductivity of tube wall metal

Rfo = fouling resistance outside = 0.001 for gas

Rfi = fouling resistance inside = 0.0005 for steam

= 0.001 for water

α = maximum steam quality

Tw = temperature of water (recycled + cooling water)at tube inlet

Tcw = cooling water temperature

U = over-all heat transfer coefficient in convection regime

hcb = boiling heat transfer coefficient,derived from Thom's correlation for fully developed nucleate boiling (Ref.41,Applied heat transfer. V.Ganapathy,1982).

Q2 = heat rate(BTU/h) in convection regime

q2 = heat flux in convection regime (BTU/h sq.ft of inner surface)

Assumptions:

8 units in parallel

tube characteristics Dto = 0.052 ft

Dti = 0.031 ft

length = 75 ft / 2 = 37.5 ft in each unit

So = 0.1634 sq.ft/ft temperature = To

Si = 0.098 sq.ft/ft temperature = Ti

Metal:Croloy 2.1/4 % Cr,1% Mo

Heat transfer from bed to tube outer surface (see 4.5 above):

T (F) hc'(BTU/h sq.ft.F)

680	432
800	438

Average value of hc' = 435 BTU/h sq.ft.F

The program solves the simultaneous equations expressing:

1) The heat rate from the bed into the part of the unit bundle in vaporizing regime,

2)Heat conduction through tube wall in vaporizing part of unit,

3)Nucleate Boiling heat transfer inside tube in vaporizing part,

4)Thom's correlation for forced convective boiling in the upper part of the tube,

5)The inlet temperature of the mixture of BFW and recycled water.

6)As shown on Fig.17, α is directly related to Q1,since the pressure (2,625 psia) is fixed.

The bed temperature T,Lv and the number of tubes n may be considered as the main design parameters.

6.5-JUSTIFICATION OF THE TUBE DIMENSIONS AND ALLOY COMPOSITION:

The maximum allowable design stress for Croloy 2 1/4 is,per the ASME specification SA 213,T22,in excess of 13,100 psi at a temperature of 900 F (100 F above the maximum operating temperature of the bed,to allow for the risk of hot spots in the catalytic bed).

The stress resulting from the maximum internal pressure in the tube: 2,625 psia,for the dimensions considered (5/8 " OD,3/8 " ID) is: (2,625 psia x 5/8 ")/[2 x (1/8 ")] = 6,562 psi,or about 50% of the allowable stress,thus providing a two to one safety margin.The temperature at which this maximum stress,corresponding to zero Syngas pressure and 2,625 psia steam pressure,would exceed the allowable stress for this alloy is about 1,030 F.This covers the risk of

a hot spot in the catalytic bed,230 F above the upper design operating temperature.A less conservative design may be preferred provided that the tests with commercially available catalysts scheduled in Phase 2 show that it is possible to safely reduce the tube wall thickness.

6.6-EFFECT OF A REDUCTION OF THE TUBING WALL THICKNESS:

If the risk of hot spots developing in the catalytic bed is shown to be minimal,the tube wall thickness may be 1/16 inch,instead of 1/8 inch.This will make: Dti = 0.042 ft instead of 0.031 ft

Si = 0.1319 sq.ft/ft instead of 0.098 sq.ft/ft

Tube inner section = 0.00138 sq.ft instead of 0.00076 sq.ft

hc' remains unchanged,but the numerical equations are changed.
The solution shows that a 15 % reduction of the number of tubes is possible by the use of 1/16" thick tubes. The adoption of this tube ID,coupled with the use of an alloy providing a higher allowable stress at high temperature are fully justified.

Such alloys are available,for instance Croloy 321 H,for which the allowable design stress at 1,050 F is still 10,100 psi,thus providing a safe operation at this high temperature,even if the differential pressure between steam and Syngas reaches a value of 2,020 psi. This is an unlikely occurrence,with suitable controls and check valves on the Syngas compressor at the surface.

The bed volume fraction occupied by water tubes is only 12.3 % of the total bed volume.To make up for the tubes volume,the height of each bed section is increased to 42 ft.When the heights of the connectors between units and between sections and those of the valve section,reactor hanger,cross-over section,etc...are added,the over-all height of the reactor assembly exceeds 200 ft,for a total volume

greater than 600 CF.The weight of the reactor assembly and that of the outer tubing,for the West Sak depth,however,remain well within the capacity of the draw-works available on the heavy drilling rigs presently in use on the North Slope of Alaska.

6.7-CONCLUSIONS OF TASK No.5:

1) The water tubes system is essentially a high-pressure boiler in which each tube is operating successively in two different regimes over two adjacent portions of its total length (37.5 ft):

A) a forced convection regime to pre-heat the water up to the boiling temperature of 675 F,and

B) a fully developed nucleate boiling regime to vaporize the water into a low quality steam/water mixture.

2) The steam quality is limited by the need to prevent tube burn-out in the vaporizing regime.It decreases with increasing heat flux in that portion of the tube and with increasing boiling pressure (here: 2,625 psia).

3) To minimize the number of tubes,and their volume share of the reactor,it is advantageous to :

A) maximize the steam quality $\dot{\alpha}$ by operating the vaporizing part of the tube system at a relatively low heat flux,

B) pre-heat the cooling water prior to its entry into the tubes system.This shows the interdependence between the design of the tubes system and the heat exchange characteristics of the concentric tubings.

4) With 8 unit bundles in parallel and a relatively small number of tubes,the friction pressure drop through the water pre-heat part of

the 0.5 inch ID tubes is small.

5) Tube dimensions of 5/8 inch OD,1/2 inch ID result in significantly fewer tubes per bundle and safe operations can still be maintained by selecting a higher-strength alloy,such as Croloy 321 H.

Until further optimisation,this will be the preferred design.A recapitulation of the reactor characteristics is given in Table 6.

7.DESIGN OF THE DOWNHOLE VALVE SECTION

The valve section,located above the reactor hanger,includes 3 major elements:1) a set of 5 parallel tubings,of which two are terminated by telescopic tubes providing a retrievable connection to the various drainholes,while the other three respectively convey the Syngas and water streams to the reactor below and the reactor effluent to the valves.One of these 3 tubings may be the outer shell of the section.

2) a set of wireline retrievable downhole three-way valves located within the first two parallel tubings.

3) upper and lower cross-over flow connectors transforming the parallel tubings arrangement into the concentric tubing configuration required to mate with the respective ends of the 3 tubings above and with the reactor head below (see Fig.6).

The function of the three-way downhole valves is to alternately connect the drainholes to the central production tubing above or to the reactor effluent outlet below.These valves are operated from the surface by hydraulic lines strapped on the outside of the water tubing.The valve actuators are similar to those of conventional subsurface safety valves.The valves provide a full opening path from each drainhole to the surface through the production tubing,so that logging,wireline tools and coiled tubing can be run through,for

surveillance and maintenance of production from each drainhole over the field's life.

The minimum number of downhole valves is 2,but,for added flexibilty in the well operation,it may ultimately reach 10,if 8 drainholes are to be operated independently of each other in three modes:injection, steam soak and production.Each valve is hung and locked into a suitable nipple from which it can also be retrived and brought to the surface by conventional wireline techniques.Some of the nipples may also include a sliding sleeve.Typical downhole valve configurations are shown on Fig.18.

The movable part of each valve (ball,flapper or sliding sleeve) provides metal/metal seals against elements of the valve body,so that only one type of stream (produced fluids or reactor effluent) can flow through at any time,while the other stream is shut-off.The surface-controlled position of the valve's movable part determines which type of stream successively flows through the valve (see Fig.19 and 20).

7.1-SEQUENCE OF OPERATIONS FOR DRILLING AND COMPLETION OF THE DRAINHOLES AND FOR THEIR CONNECTION TO THE VALVE SECTION

After drilling the large-diameter vertical hole,the interval over which the multiple drainholes will be kicked-off is underreamed to provide space for all the drainhole liner stubs in their extended position.The 24.5 in. casing string is then made-up and run-in. It includes a special joint 300 ft above the shoe,incorporating the reactor hanger and a joint equipped with telescopic liner stubs through which the respective drainholes will be kicked-off,drilled and completed.Each stub is closed by a drillable plug and can slide

through an elliptical window machined into the casing at the kick-off angle. Drillable aluminum guides within the casing provide mechanical support to each stub in its retracted position, while the casing is run into the hole, and during its telecopic extension. With the casing hung at the surface, a work string is run-in and stabbed into the casing shoe to proceed with the stab-in cementing operation. When the cement slurry has been displaced behind the casing using conventional procedures, the mud pressure in the annulus between the work string and the casing is increased, so as to drive the telescopic stubs into the slurry. The tail end of each stub is designed to provide a metal/metal seal between the stub and the casing window. After the cement has set and the work string has been pulled out, the aluminum stub guides are drilled out and any protrusion into the casing is removed. The drainhole drilling string, including its downhole motor, bent sub and guiding system is run-in and inserted into one of the extended and cemented liner stubs, to proceed with drilling the stub plug, cement and formation at the kick-off angle, and from there, to build-up the angle until the horizontal part of the drainhole is reached.

The drainhole is then completed using conventional techniques (perforated liners and screens, gravel packing if necessary) and its tail end is cemented using cement-inflated external packers. The end of the liner is connected to the extended end of the stub by means of a hanger [illegible] within the cemented stub, and terminated by a polished bore receptacle [illegible] an equivalent female metal seal.

When all the planned drainholes have been drilled and completed, some of the cemented stubs may remain [illegible] and completion of additional drainholes [illegible] layers

The drilling mud is displaced by water from each completed drainhole and from the vertical casing,which is thoroughly cleaned of any mud solids.

The reactor assembly is then made-up,leak tested and run-in the water-filled casing.The reactor is threaded through its hanger and the metal/metal seal is leak tested.Each of the 8 units composing the reactor was factory assembled,loaded with activated catalyst,filled with pressurized inert gas,leak tested and sealed prior to being sent to the rig.The integrity of the seals between adjacent units is also checked during make-up on the rig floor,prior to run-in.All flow connections between the reactor head and the valve section are protected from water entry from the well by spring-loaded check valves.

With the reactor assembly hung into the well,the oriented valve section is run-in with the water tubing and stabbed into the reactor head.Hydraulic control lines are strapped to the outside of that tubing are used to hydraulically thrust into each PBR (or equivalent) a telescopic connector tube,as shown on Fig.6 to establish leak-proof metal/metal connection with the liner of each completed drainhole.

The Syngas tubing is then made-up and run-in.It is used to fill the annulus between water and Syngas tubings with inert gas prior to connecting the Syngas tubing to the upper cross-over connector of the valve section.

Finally,the central production tubing,filled with inert gas,is run-in and connected to the upper cross-over connector.

Thermal expansion joints and other common features of thermal well completion and thermal well head are included,but are not specified in detail.Reactor start-up and emergency shut-down procedures have also

been developed.

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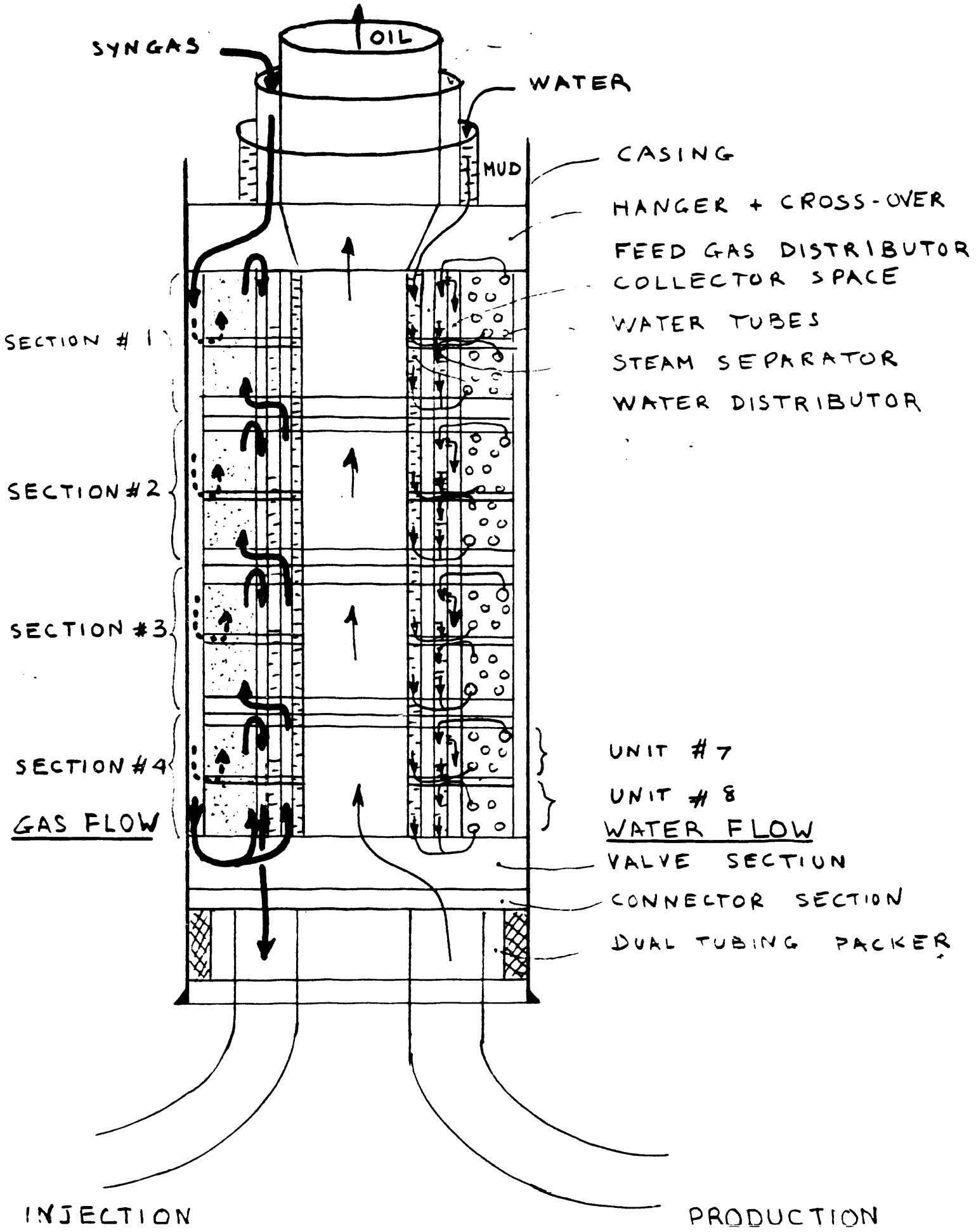
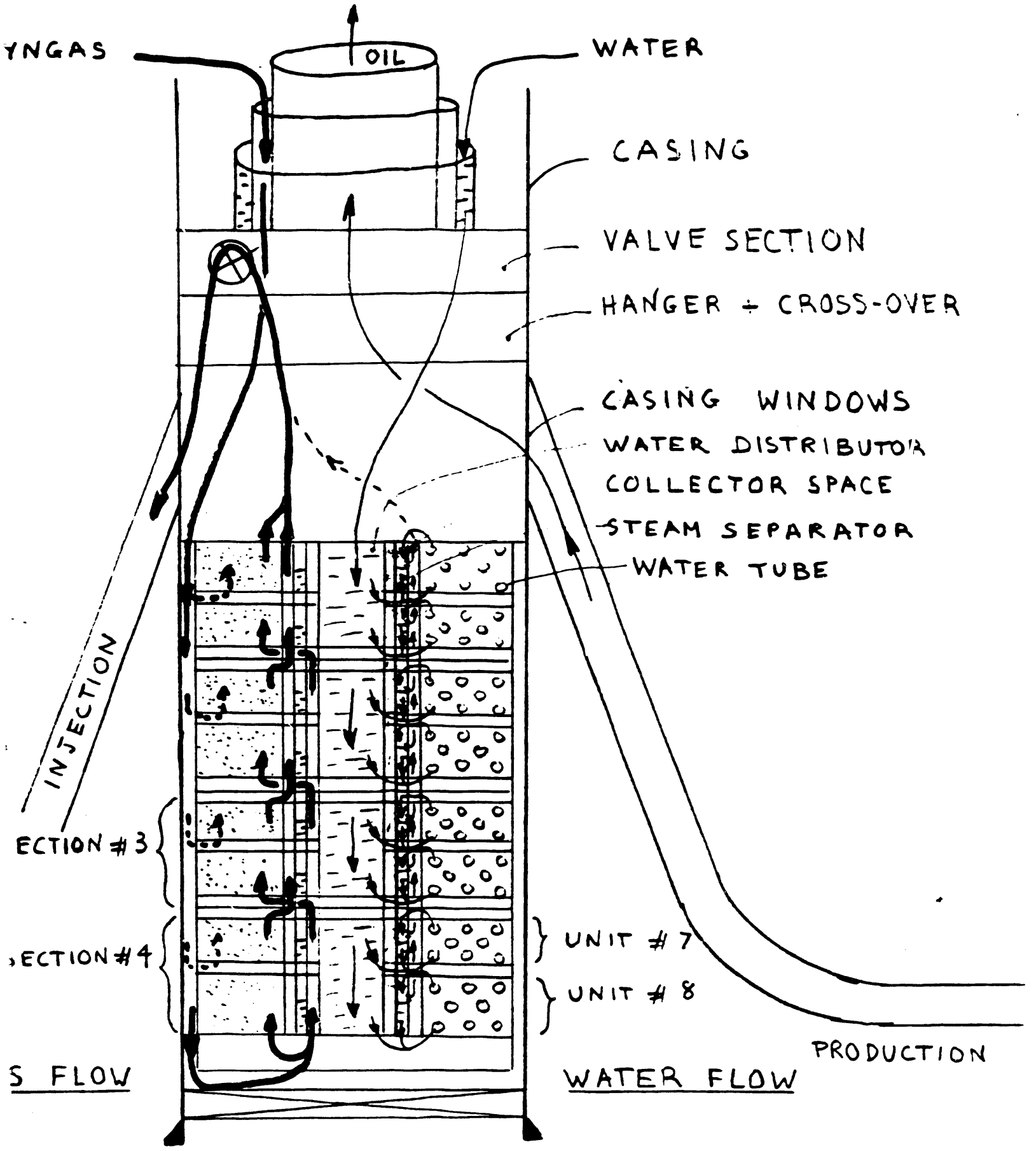


FIG. 9 - VERTICAL FLOW REACTOR - BOTTOM DISCHARGE



G. 10. VERTICAL FLOW REACTOR - TOP DISCHARGE

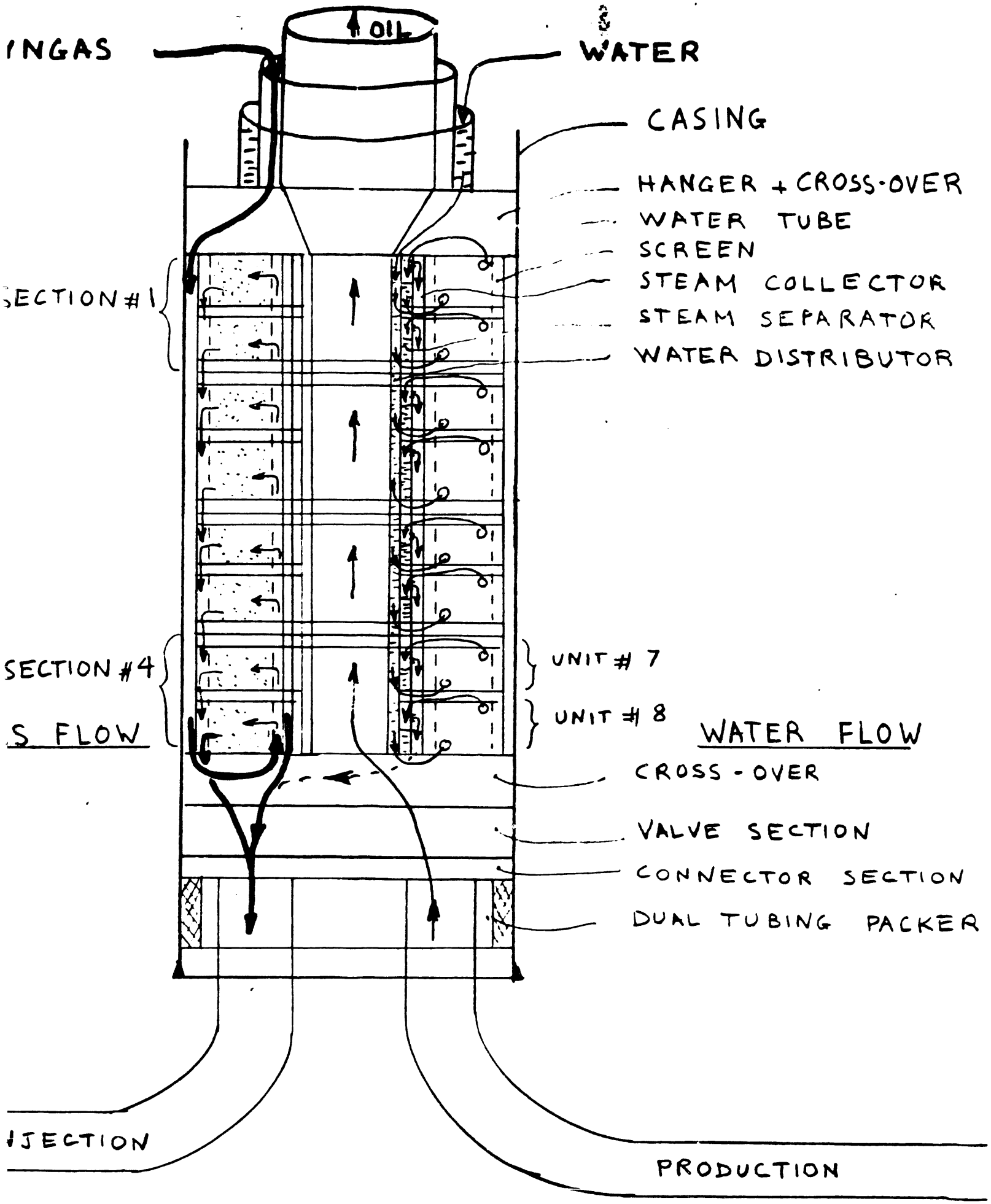


FIG. 11 - RADIAL FLOW REACTOR - BOTTOM DISCHARGE

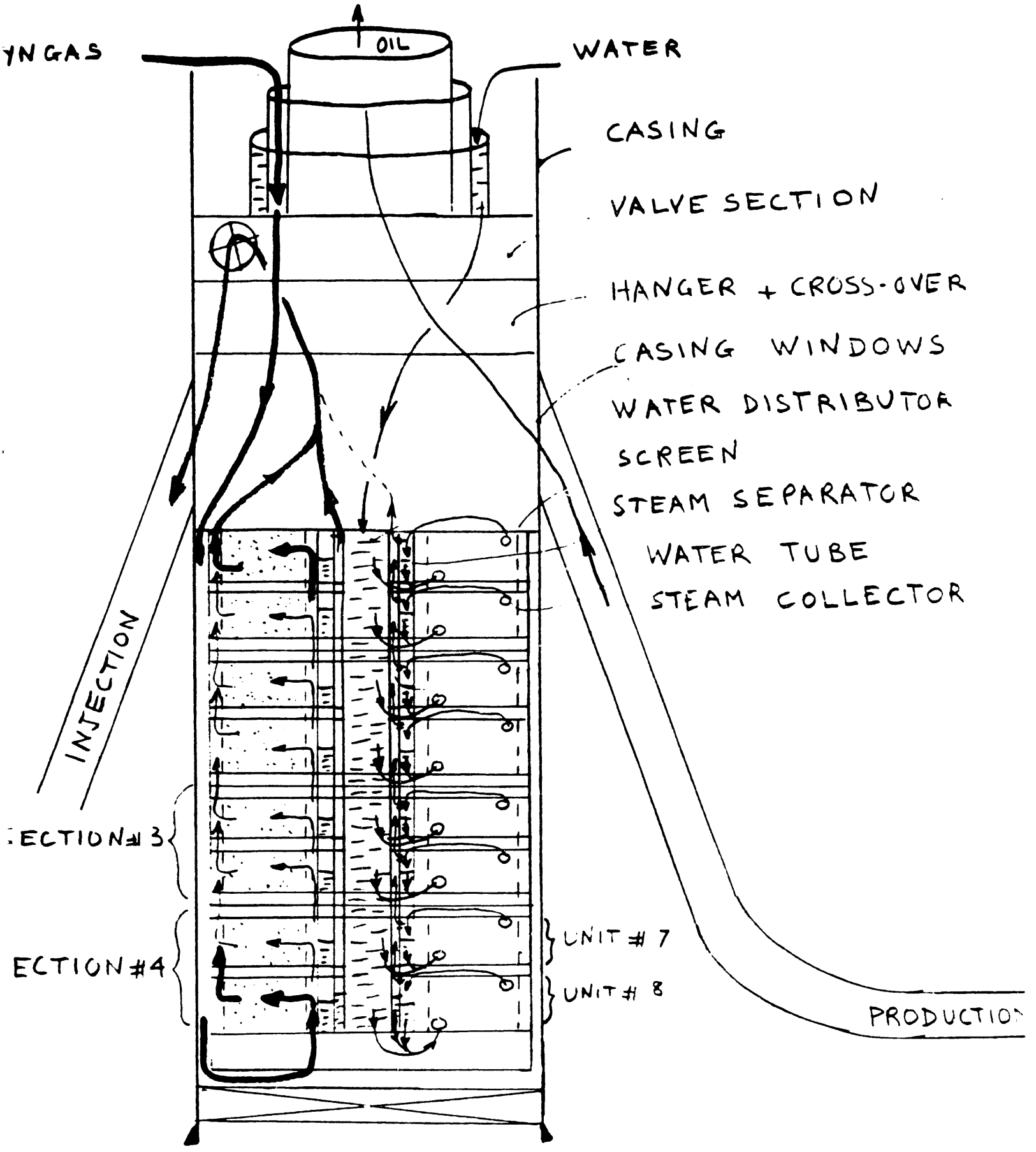


FIG. 12 - RADIAL FLOW REACTOR - TOP DISCHARGE

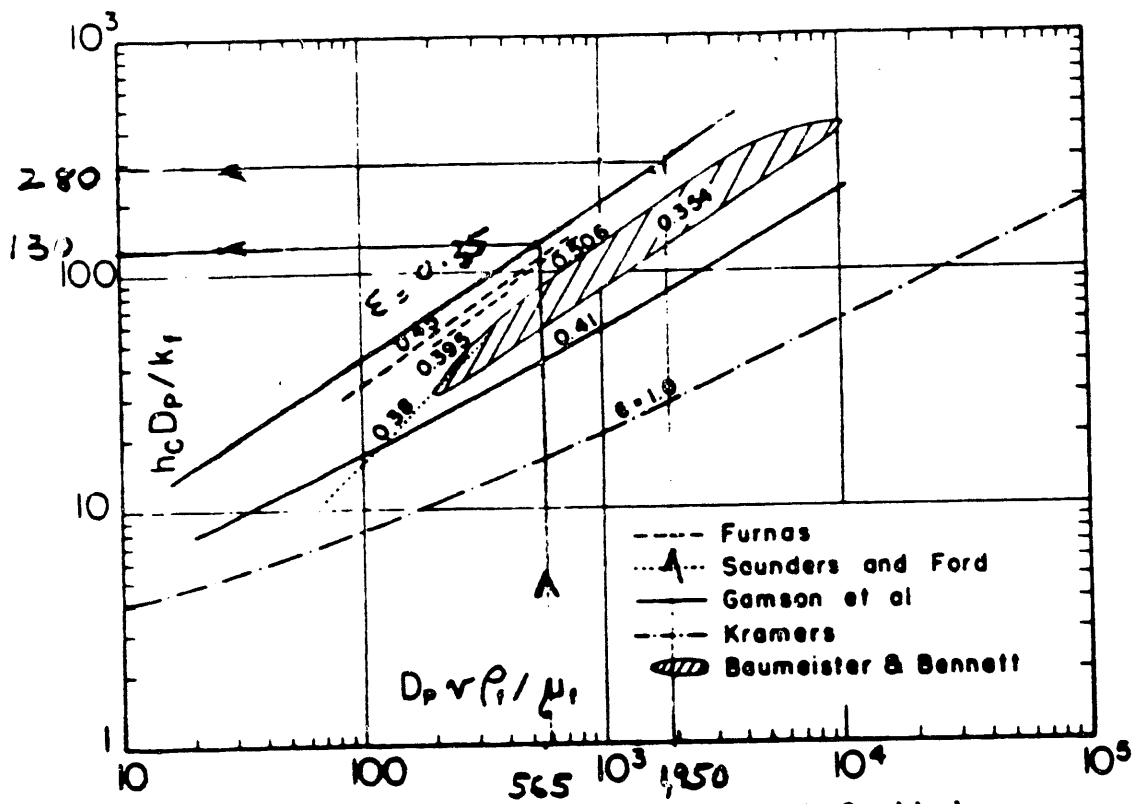


Figure 13. Particle-to-fluid heat transfer in fixed beds

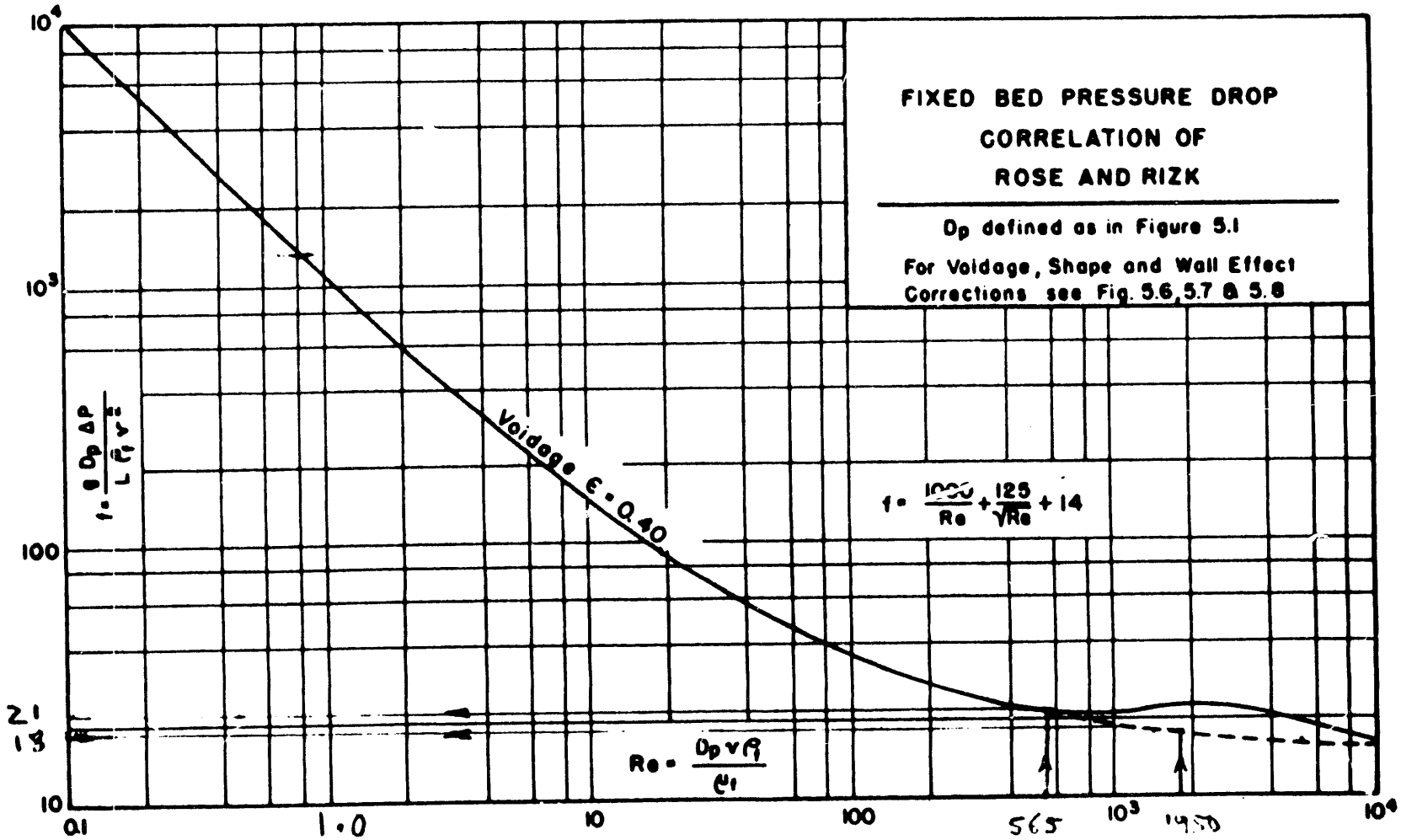

FIXED BED PRESSURE DROP
CORRELATION OF
ROSE AND RIZK
Dp defined as in Figure 5.1
For Voidage, Shape and Wall Effect
Corrections see Fig. 5.6, 5.7 & 5.8
Voidage ε = 0.40
f = 1000/Re + 125/√Re + 14
Re = Dp v ρ / μ
f = g Dp ΔP / L ρ v²
10⁴
10³
100
21
13
10
0.1
1.0
10
100
565
10³
1450
10⁴

Figure 14

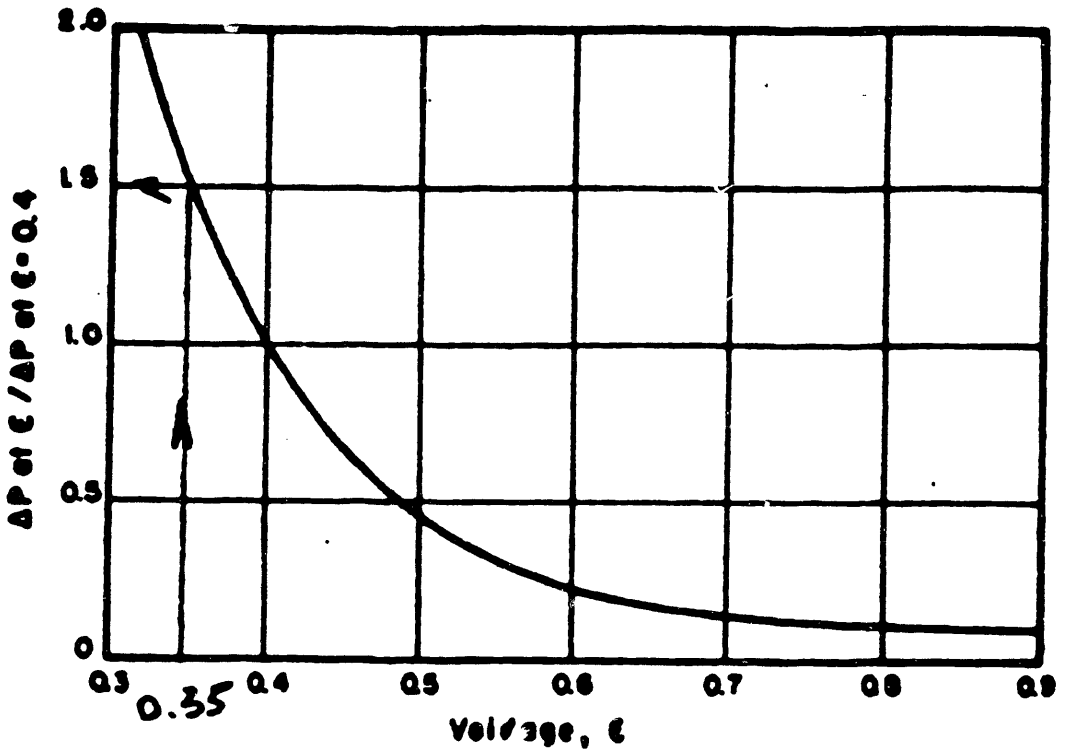


Figure 15 Rose and Rizk's fixed-bed voidage function

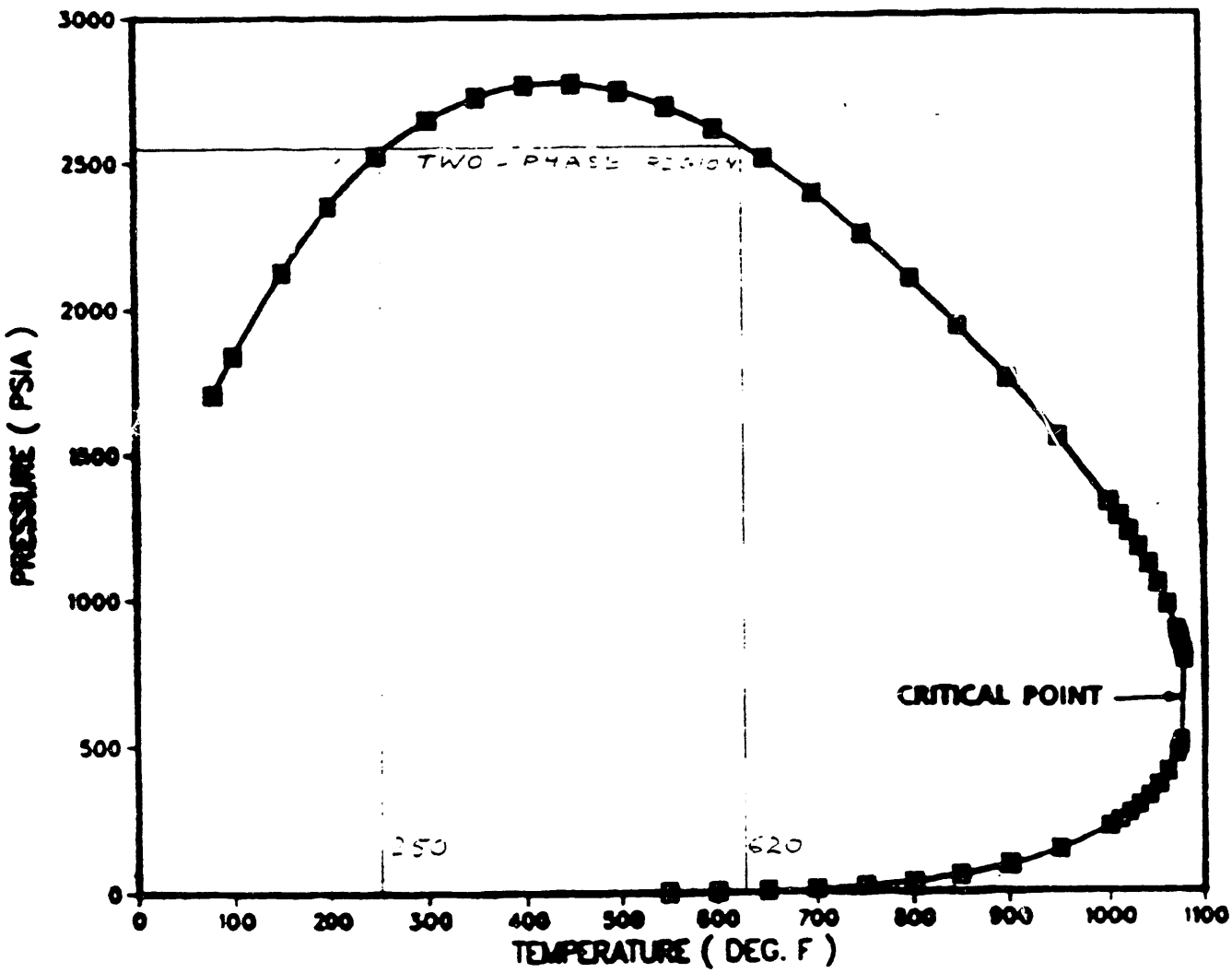


Figure 16 EOS Predicted P-T Envelope for West Sak Crude

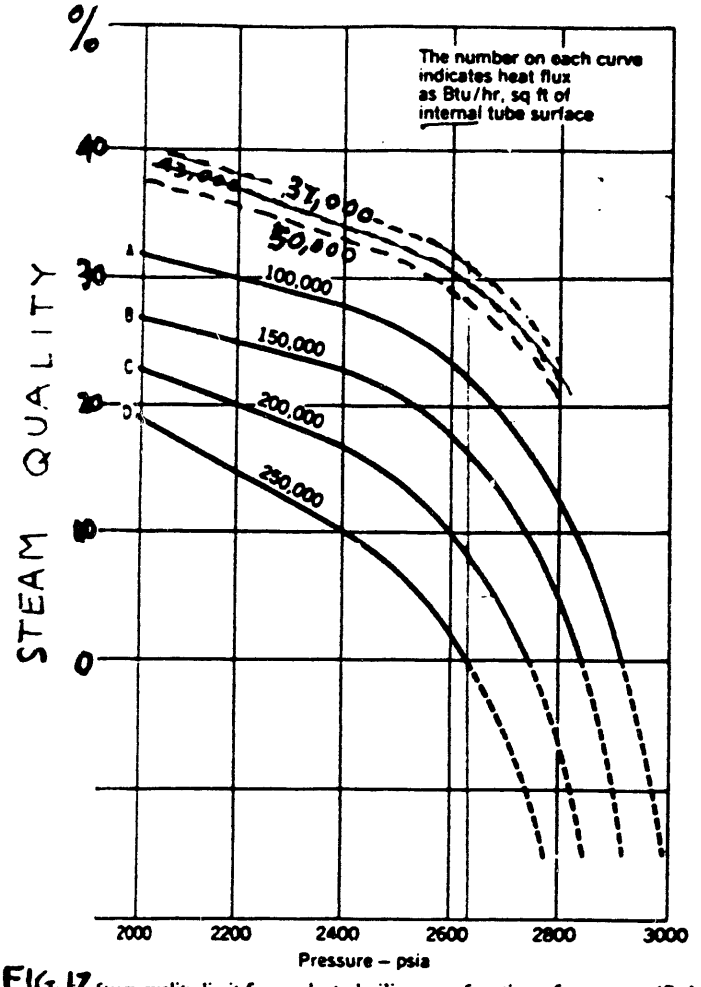


FIG. 17. Steam quality limit for nucleate boiling as a function of pressure. (Babcock & Wilcox.)

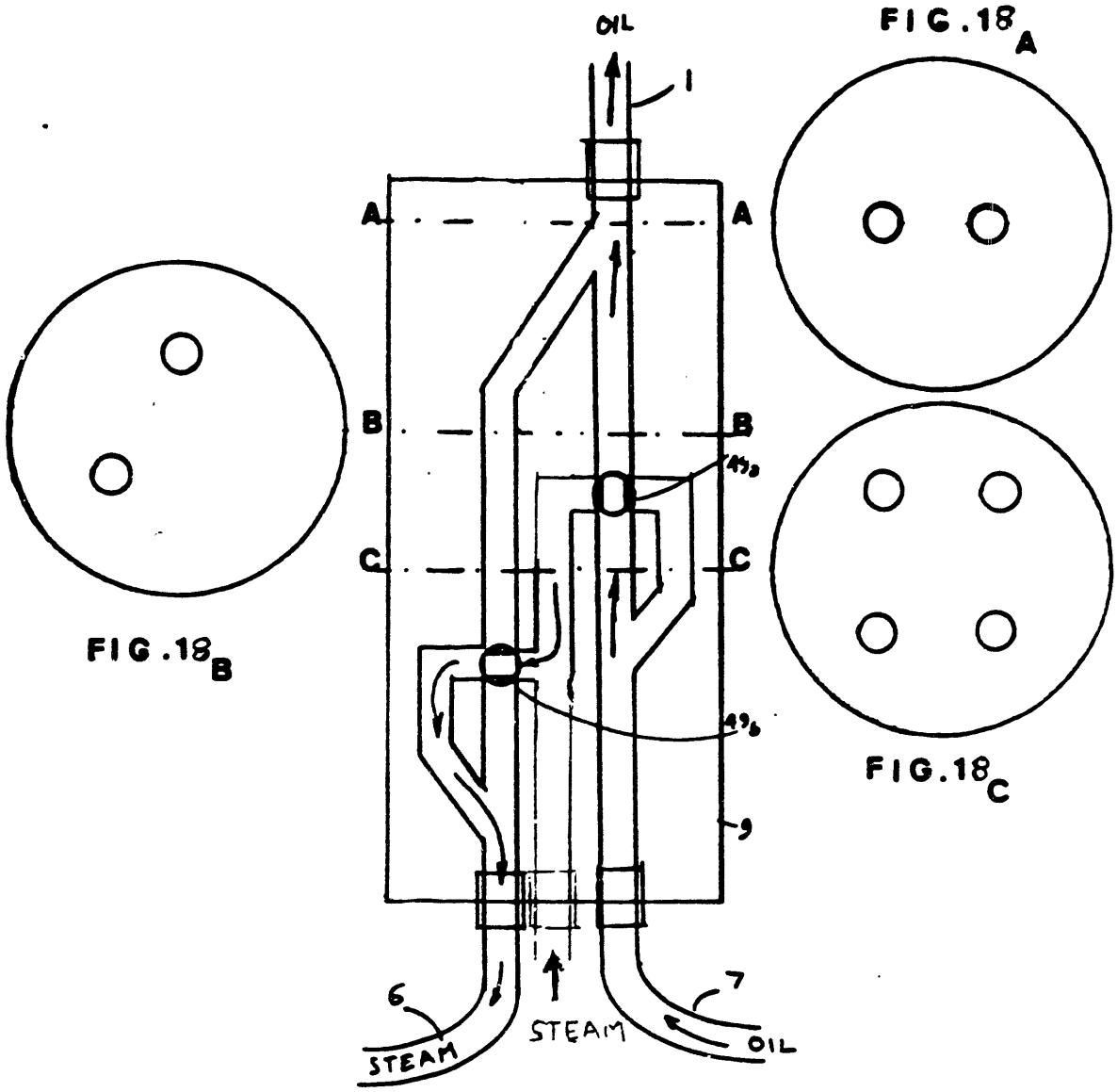


FIG.18A

FIG.18B

FIG.18C

FIG.18

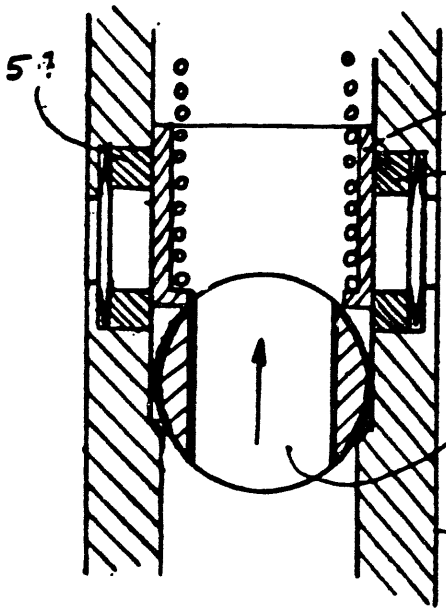


FIG. 19A

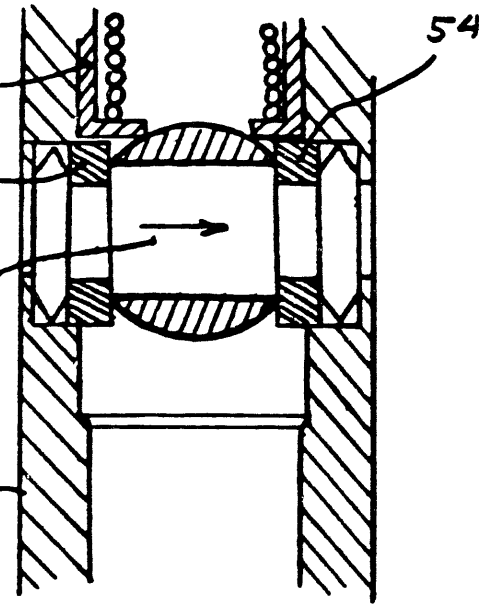


FIG. 19B

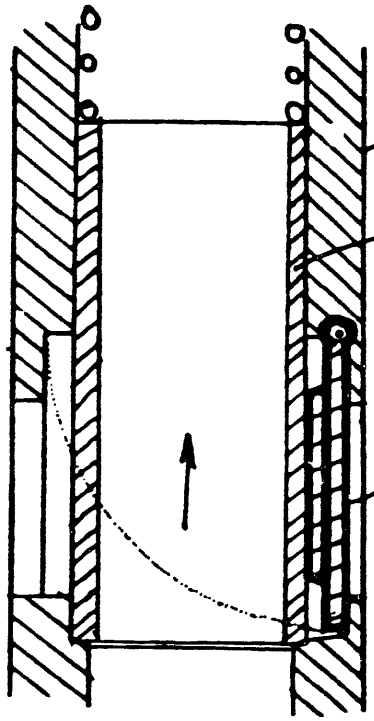


FIG. 19C

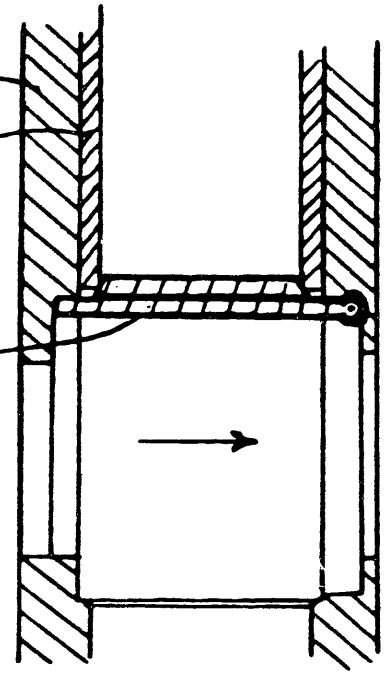


FIG. 19D

APPENDIX 3:PRELIMINARY RESERVOIR ENGINEERING

5.WEST SAK OIL PHASE BEHAVIOR

The Pressure,Temperature phase envelope predicted by the Equation of State method (EOS) for the recombined West Sak crude studied by A.K.Sharma et al.(1989)is reproduced on Fig.16,taken from Ref.4.It shows that,at the injection temperature of 620 F,the bubble point of West Sak oil is about 2,500 psia.It also shows that,at that pressure,two-phase mixtures exist between 620 F and about 250 F.This indicates that injected Methane will dissolve preferentially in the colder region,ahead of the condensing steam front.In the part of the reservoir at a temperature exceeding 250 F,the injected Methane will provide full pressure support,as a gas phase.

5.2-ELEMENTS OF A FORECAST OF WELL PRODUCTION:

An accurate forecast of the well production rate and fluid composition could only be obtained by using a numerical simulator. In such simulators,the reservoir around the horizontal well is divided into representative grid blocks.The simulator performs a material balance in each block during each time step.It also keeps track of the

in-flow and out-flow of the various phases,under the prevailing pressure gradients for each block and each time step in order to determine the total flow of each phase into the well as a function of time.Among the many steam injection simulators available,the MULKOM numerical simulator,developed under D.O.E. funding by K.Pruess et al (1983) at the Lawrence Berkeley Laboratory (Ref.39)appears to be most suitable because it is also capable of considering the changes in composition and transport properties of the oil and gas phases resulting from "in situ"hydrogenation and steam distillation of oil within the reservoir.

The present budget precludes the use of any such simulator. The simple approach taken here consists in considering the over-all material balance of injected and produced fluids and in making assumptions based on results of conventional steam injection processes.To the extent possible,the validity of these assumptions is verified against experience in other fields than the West Sak.

Ultimately,some numerical simulation will be required for confirmation of the rough estimates obtained below.

5.21 -RISK OF PERMEABILITY REDUCTION:

A major uncertainty regarding the production rate is the possibility that fresh water from condensed steam may cause the clay minerals in the reservoir rock to swell,thereby drastically reducing the rock permeability in the steam-injected zone.The only published results on West Sak oil recovery from experimental steam injection runs in the laboratory,given in Ref.5,were obtained with sand packs containing no swelling clays,but the presence of this type of clays was reported in some of the descriptions of the West Sak

reservoir rock. There is no published information regarding the areal distribution of such swelling clays within the reservoir and it is not possible at this time to assess the importance of this potentially serious problem. Dilution of the connate water by condensed steam may reduce the ionic strength of the solution in direct contact with the clay minerals to such an extent that the negative charges of the clay lattice must now be balanced largely by H+ ions resulting from the ionization of water. Because of their larger hydration number, H+ ions carry into the bound water phase, adjacent to the clay mineral, much more water than the original counter-ions (Na+, Ca++). This water causes the clay to swell, an effect which can only be prevented by replacing the H+ ions with cations exhibiting a lower hydration number.

Fortunately, the present process offers a remedial action against clay swelling. This can be prevented by adding to the Syngas feed or to the cooling water feed a small percentage of Ammonia gas, NH3, which will be injected into the reservoir, together with the steam, CH4, H2 mixture, increasing the Ph and the ionic strength of the steam condensate.

The hydration number of NH4+ ions is less than one third of that of H+ ions. Consequently, they carry much less water with them than do H+ ions.

In the non mobile water phase, also called "bound water" of the clay, the collapse of the electrical double layer adjacent to the clay platelets by NH4+ ions will prevent clay swelling, thus eliminating the risk of blocking the flow paths within the reservoir pore space.

In the mobile water phase, however, the addition of NH4+ ions to the mixture of connate water and condensed steam results in the formation of NH4 Cl, a highly soluble salt.

The amount of NH_4 ions required to stabilize the clays must be determined experimentally on reservoir core samples. Assuming that the original counter ions in the clays represent a Na^+ ion concentration of 5,000 PPM in the formation water, the NH_3 requirements for a total injected steam rate of 155,565 #/h is only 575 #/h or 7 ton/D.

The produced NH_3 stream obtained by the process shown on Fig.2 may then be scrubbed with water and mixed with the cooling water feed or re-vaporized and mixed with the Syngas stream upstream of the booster compressor. It remains unaffected by heat in the Downhole Methanator and becomes part of the Injectant mixture. It will, however, dissolve in the formation water and condensed steam mixture flowing through the reservoir. The NH_4^+ ions thus produced are strongly attracted by the electrical charges in the lattice of the clay minerals, which become stabilized, despite the addition of large volumes of steam condensate.

Despite its very small size and the elimination of the costly feed preparation step, the Ammonia plant still represents a sizable investment, which can only be justified if experimental steam injection runs into representative core samples of the West Sak reservoir conclusively show that clay swelling is excessive and that remedial measures must be taken to preserve the rock permeability, which is only moderate (150 md) for the West Sak reservoir. Such laboratory experiments should be given a high priority in future research.

5.20 - FLUID DRIVE MECHANISMS:

Assuming that the average pressure in the hot part (800 F to 680 F) of the reservoir is 1,000 psia and that half of the injected Methane remaining in the reservoir stays in that zone, the volume of liquids which this gas phase can displace from the reservoir is:

0.5 x 0.2 x21.8 MMSCF/D x (14.7 psia / 2,300 psia) x (890 R / 492 R) = 7,675 CF/D or 1,367 B/D.

Additional drive mechanisms,at constant reservoir pressure,include:

-swelling of oil by the rest of unrecovered Methane injected in the colder part of the reservoir,with an estimated incremental volume of 2.18 MMSCF/D,

-thermal expansion of fluids in the heated zone,

-gravity drainage of oil from the steam injected zone.

In addition,with a pressure draw-down applied from the well,three more drive mechanisms come into play:

-injected steam (15,370 B/D/well) and gases provide another source of pressure support,as shown in Ref.38 (P.J.Briggs et al.,1988),

-edge water influx from the large Cretaceous aquifer,

-compaction of the friable West Sak reservoir rock.

The effect of oil swelling by unrecovered Methane in the colder zone can be evaluated quantitatively from the apparent density of natural gas dissolved in crude oils (Ref.33,D.L.Katz et al.1959) For a 20 API crude at 2,300 psia and an average temperature of 175 F,the apparent density of Methane dissolved in the colder reservoir crude is: 20.5 #/CF.The incremental volume due to that effect is:2.18 MMSCF/D x 16 #/#.mol /(380 SCF/#.mol x 20.5 #/CF) = 4,178 CF/D or 798 B/D.

Thermal expansion of fluids in the heated zone may be calculated from the incremental volume of that zone.The daily heat input into the reservoir,assuming that 20% of the total heat is lost to the overburden is,215 MMBTU/h x 0.8 x 24h/D = 4,128 MMBTU/D.

The average temperature in the heated zone is:

(620 + 75) / 2 = 347.5 F,which corresponds to an increase of 272.5 F

over the original reservoir temperature (75 F).The incremental volume of the heated zone is: 4,128 MMBTU/D / (Cr x 272.5 F) ,where Cr is the specific heat per cubic foot of fluid-saturated reservoir.A typical value of Cr is 30 BTU/CF.The incremental volume per day of the heated zone is therefore:

504,354 CF/D.With 30% porosity,the fluid volume is:151.486 CF/D

The thermal expansion of the oil and water is about 10% or 15,149 CF/D or 2,698 B/D.

Adding up these three volumes to that of the injected steam gives a minimum value of the daily total fluid production:

15,370 B/D + 2,698 B/D + 798 B/D + 1,367 B/D = 20,233 B/D.,rounded to 20,000 B/D.

Gravity drainage,pressure support from steam and gases,water influx and reservoir compaction effects cannot be predicted without more specific information about the reservoir.

A conservative estimate of these various increments would be 2,000 to 4,000 B/D,bringing the total fluid production to the range of up to 24,000 B/D.

5.9-FORECAST OF OIL RATE FROM RESERVOIR VOIDAGE:

The produced fluids consist largely of water,a mixture of steam condensate with formation water,and oil,some from the region affected by "in situ" hydrogenation and some from the colder regions of the reservoir.The volume of steam condensate alone is 15,370 B/D.The WOR depends upon the relative permeabilities to oil and water of the West Sak reservoir rock,for which no data have yet been published.

A more commonly known yardstick for conventional steam injection projects is the Oil/Steam Ratio (OSR).For a successful project using

horizontal wells in a relatively thin reservoir layer, it is not unreasonable to expect the OSR to be 0.5. This would bring the oil rate to a minimum value of 7,885 B/D, discounting any benefits from the hot injected gases and from 'in situ' hydrogenation. If, however, only the oil subjected to hydrogenation is recovered, the oil rate will be 7,847 B/D. Consequently an oil rate forecast, based on reservoir voidage, of 8,000 B/D is conservative.

5.4-FORECAST OF OIL RATE FROM THE FLOW POTENTIAL OF A DRAINHOLE:

The production of horizontal drainholes, Qoh, was compared to that of vertical wells, Qov, drilled in the same formation by S.Joshi (Ref.14).

For the reservoir layer characteristics considered here, Qoh >= 5 Qov

$Qov = 7.08 \times K \times Kro \times h \times (Pe-Pwf) / Bo \times \mu o \ (Ln\ re/rw - 0.75)$

We assume the following values in the drainage area of the drainhole:

(Pe-Pwf) = 850 psi

K = 0.15 darcy

Kro = 0.9

h = 30 ft

Bo = 1.4

μo = 1.5 cp

re = 1,500 ft

rw = 0.229 ft

Consequently, Qov = 1,649 STB/D

Qoh = 5 x 1,649 = 8,246 B/D, rounded to 8,000 B/D

5.5-FORECAST OF OIL RATE FROM THE HEAT FRONT DISPLACEMENT:

The method used is that of Marx and Langenheim (REF 22), modified for the geometry of a linear frontal displacement. The thermal

properties of the reservoir and overburden formations were derived from the experimental data published by W.Somerton (Ref.42). This method only considers the displacement of the heat front. After 15 years of continuous steam injection, the thermal front area predicted is 19 MMsq.ft, corresponding approximately to the drainage area of two drainholes during Phase 1 ("huff and puff" mode). Conversely, the oil displacement rate calculated is 7,820 STB/D (rounded to 8,000 B/D).

At the end of Phase 1, it was assumed that steam injection would be continued as a steamflood, displacing oil towards newly drilled production drainholes, parallel to the injection drainholes and with a separation of about 3,000 ft. The original injection wells are assumed to be arranged on a square grid and all the new producer wells are expected to be drilled at year 16. The steamflood sweep efficiency was estimated at 40 %. Under those conditions, the predicted results of Phase 2 are:

- Steamflood duration: 14.0 to 18.7 years
- Average Oil Rate: 8,000 B/D
- Average Steam/Oil Ratio = 3

5.7-CONCLUSIONS OF TASK No.4:

1) Total liquid production from the well is estimated at 18,000 to 24,000 B/D, with a minimum GLR of 800 SCF/B.

2) The liquid stream is approximately 1/3 oil and 2/3 water.

3) More accurate estimates will require additional information about the West Bak reservoir and the use of numerical reservoir simulators.

4) The predicted production lifetime equals or exceeds 30 years.

5) These forecasts are internally consistent when widely different methods are used.

APPENDIX 2.1

PHYSICAL PROPERTIES OF GAS MIXTURES FLOWING IN THE DOWNHOLE METHANATOR

A.1. VISCOSITY OF H_2,CO MIXTURES

The viscosity of gas mixtures can be derived from the values for pure components in the mixture,provided that the following conditions are met:

1) a suitable mixing rule is known,

2) the effect of temperature is taken into account,

3) the effect of pressure is accounted for.

For mixtures containing a large fraction of hydrogen,with its low viscosity,compared to that of other gases, it is prudent to verify the

validity of the calculations by comparing results with experimental data on similar mixtures. Such data are available from work published by N.V.Tsederberg (Ref.32,1965) for a mixture of 0.795 H2, 0.205 CO, at 14.7 psia and 32°F.

A suitable mixing rule for viscosity calculations is that of Carr (see Ref.33, D.L.Katz, 1959):

$\mu f(mix) = \Sigma\, x_i . \sqrt{M_i} . \mu_i \;/\; \Sigma\, x_i . \sqrt{M_i}$

where: x_i are the mol fractions of the pure components in the mixture,

M_i are their respective molecular weights,

μ_i are their respective viscosities.

This rule will be first verified for the Tsederberg mixture, based on data for pure components, given in Ref.34 (F.Kreith, 1964) at various temperatures.

components:	CO	H2	Tsederberg Mixture
mol.fraction	0.205	0.795	1.000
mol weight	28	2	
μ_i @ 14.7 psia			
and @ 32°F :	1.065	0.540	0.800
and @ 62°F :	1.932	0.950	
$\times 10^{-5}$ #/ft.s			

The experimental value given by Tsederberg is 0.876 (less than 10% above that calculated by Carr's mixing rule).

2. EFFECT OF A DIFFERENT COMPOSITION:

The composition of the feed Syngas is slightly different from that of the Tsederberg mixture:

components:	CO	H2	Syngas Mixture
mol.fraction	0.222	0.778	1.000
μi @ 14.7 psia and at 32°F :	1.065	0.540	0.811

By comparison with Tsederberg's experimental data, the true value of viscosity for the Syngas mixture should be:

0.876 x (0.811 / 0.800) = 0.888 (10.-5)#/ft.s

A.3-EFFECT OF TEMPERATURE:

The calculated viscosity of the Syngas feed @ 620 °F and 14.7 psia is:

	CO	H2	Syngas Mixture
mol.fraction	0.222	0.778	1.000
μi @ 14.7 psia and @ 620 °F :	1.932	0.950	1.457

True value of viscosity: 1.457 x (0.876 / 0.800) = 1.595 (10.-5)#/ft.s

A.4-EFFECT OF PRESSURE:

The effect of pressure on gas viscosity is relatively small in the region far from the critical point. Comings and Egly (Ref.35,1941) have provided correlation charts in terms of reduced pressure and reduced temperature.

The critical parameters for the Syngas mixture are as in the First Quarter Report (p.12). They are :

Pc = 259 psia and Tc = 170 °R

In order to offset the expected pressure drop across the bed, the

Syngas feed [illegible] delivered at a pressure somewhat higher than that of the effluent, for instance at [illegible]675 psia, allowing a 100 psi pressure drop. The temperature of the feed gas, after pre-heating [illegible] with the catalyst is, however, very close to 620 °F. Corresponding [illegible] reduced pressure and reduced temperature [illegible] are respectively:

Pr = 2,675 / [illegible]

Tr = 1080 [illegible]

The Cummings [illegible] given in Ref.17 do not extend to that [illegible] shows that the correction [illegible] effects does not exceed 1.07.

[illegible] OF SYNGAS AT INLET CONDITIONS:

[illegible] size of gas viscosity at the inlet of the bed is:

μf = 1.07 × 1.595 × (10.-5) #/ft.s = 1.707 × (10.-5) #/ft.s

[illegible]-VISCOSITY OF H2, CH4, STEAM MIXTUTRES:

The viscosities of most gases, other than Hydrogen are relatively close. Because the mol fraction of H2 in the effluent from the beds [illegible] the [illegible] effluent are much lower than that in the Syngas [illegible] it is possible to apply Carr's mixing rule without further correction. The composition of these two mixtures are given below.

[illegible] EFFECT OF TEMPERATURE:

[illegible]	H2O	CH4	H2	CO
mol.weight	13	[illegible]	[illegible]	29
viscosity				viscosity

(10.-5)#/ft.s (10.-5)#/ft.s)

@						
620 F,14.7psia	1.447	1.243	0.950	1.932	of mixture	
675 F,14.7 psia	1.521	1.310	0.981	1.994	@14.7 psia	
680 F,14.7 psia	1.528	1.316	0.984	2.000	and T(F)	
700 F,14.7 psia	1.553	1.320	0.994	2.022		
800 F,14.7 psia	1.700	1.415	1.050	2.134		
Injected effluent						
mol.fraction	0.694	0.192	0.110	0.004	1.390	620
Bed effluent						
mol.fraction	0.385	0.385	0.222	0.008	1.320	620
					1.386	675
					1.393	680
					1.408	700
					1.523	800

A.7-EFFECT OF PRESSURE:

components	H2O	CH4	H2	CO	Effluent: Injected	Bed
Pc (psia)	3,199	666	188	507	2,761	1,574
Tc (R)	1,165	344	[illegible]	440	[illegible]	[illegible]
Pr @ 2.575 psia					[illegible]	[illegible]
Tr @ 620 F (1,080 R)					[illegible]	[illegible]
Tr @ 675 F (1,135 R)						[illegible]
Tr @ 680 F (1,140 R)						[illegible]
Tr @ 700 F (1,160 R)						[illegible]
Tr @ 800 F (1,260 R)						[illegible]

For these steam-rich mixtures, the pressure correction factors are more

significant than that for the Syngas feed: 1.15 1.11

t = 1.18

A.8-VISCOSITY OF EFFLUENT FROM THE BED:

Combining the effects of temperature and pressure for the composition of this effluent, the viscosity is at 2,575 psia, at T°F:

1.10 x 1.393 x (10.-5)#/ft.s	= 1.532 x (10.-5)#/ft.s	@ 680
1.09 x 1.408	= 1.535	@ 700
1.08 x 1.523	= 1.645	@ 800

A.9-VISCOSITY OF INJECTED EFFLUENT:

The viscosity of the injected effluent, at 2,575 psia, 620 F is:

1.15 x 1.390 x (10.-5)#/ft.s = 1.598 x (10.-5)#/ft.s

A.10-THERMAL CONDUCTIVITY OF GAS MIXTURES CONTAINING HYDROGEN:

The thermal conductivity of gas mixtures can again be derived from those of the pure components, provided that a suitable mixing rule is available.

For thermal conductivities, the mixing rule recommanded by Perry's Handbook of Chemical Engineering (Ref.36, 1987) is:

$$(K_{mix}) = \sum^{3} x_i . \sqrt{M_i} . K_i \Big/ \sum^{3} x_i . \sqrt{M_i}$$

This is similar to Perry's rule, except that the weighting factors are now the cubic roots of the molecular weights, instead of their square roots. Again, for Hydrogen-rich mixtures, it is desirable to verify the mixing rule against the experimental data provided by Tsederberg for his mixture (0.205 CO, 0.795 H2).

The pure component thermal conductivities are again taken from the

compilation of F.Kreith (Ref.34),at the same conditions (14.7 psia, 32°F)

components:	CO	H2	Tsederberg mixture
mol fraction	0.205	0.795	1.000
mol weight	28	2	
Kf @32F,14.7psia	55	411	275

(x 10.-6)cal/cm.s.°K

This is in excellent agreement with the experimental value given by Tsederberg: Kf(mix) = 270 x (10.-6)cal/cm.s.°K

The discrepancy being less than 2%,no correction need to be applied.

A.11-EFFECT OF TEMPERATURE:

Perry's mixing rule will be applied to the experimental data compiled by Kreith (Ref.34) for pure components at 620°F,14.7 psia, for the Syngas composition:

components:	CO	H2	Syngas feed
mol fraction	0.222	0.778	1.000
mol weight	28	2	7.772
Kf @620F,14.7psia	0.0259	0.171	0.112

(BTU/h.ft.°F)

A.12-EFFECT OF PRESSURE:

This effect is again quite small for gases which are far from their critical point (Pr = 10.3 , Tr = 10.8).

By extrapolation of the charts given in Ref.17,the pressure correction factor is about 1.05.Consequently,the thermal conductivity of the Syngas feed is,at 2,675 psia,620 F:

1.05 x 0.112 BTU/h.ft.°F = 0.118 BTU/h.ft.°F

A.13-THERMAL CONDUCTIVITIES OF EFFLUENT MIXTURES:

The same method will be applied to the calculation of the thermal conductivities of the effluent from the beds and of the injected effluent,respectively.The data compiled by Kreith will be used.

A.14-EFFECT OF TEMPERATURE:

components:	H_2O	CH_4	H_2	CO	
mol weight	18	16	2	28	Effluent's
Kf @14.7psia (BTU/h.ft°F)					Thermal
and T (°F)					Conductivity
620	0.0264	0.049	0.171	0.0249	(BTU/h.ft°F)
675	0.0280	0.052	0.177	0.0260	
680	0.0282	0.053	0.178	0.0261	@14.7 psia
700	0.0288	0.055	0.179	0.0265	
800	0.0321	0.062	0.189	0.0285	
Injected effluent					Kf @ T(F)
mol fraction	0.694	0.192	0.110	0.004	0.0391 620
Bed effluent					Kf @ T(F)
mol fraction	0.385	0.385	0.222	0.008	0.0537 620
					0.0560 675
					0.0570 680
					0.0582 700
					0.0641 800

A.15-EFFECT OF PRESSURE:

The pressure correction factor is again dependent upon the reduced pressure and reduced temperature of each effluent mixture.

Effluent:	Pr	Tr	Correction Factor
Injected	1.09	1.22 @620 F	1.3
Bed	1.68	1.81 @620 F	1.2
		1.90 @675 F	1.19
		1.91 @680 F	1.19
		1.94 @700 F	1.18
		2.11 @800 F	1.17

A.16-THERMAL CONDUCTIVITY OF EFFLUENT FROM THE BEDS @ 2575 psia and T:

Kf = 1.19 x 0.0570 BTU/h.ft.°F = 0.0678 BTU/h.ft.°F @ 680 F

1.18 x 0.0582 = 0.0687 @ 700 F

1.17 x 0.0641 = 0.0750 @ 800 F

A.17-THERMAL CONDUCTIVITY OF INJECTED EFFLUENT:

Kf = 1.3 x 0.0391 BTU/h.ft.°F = 0.0508 BTU/h.ft.°F @ 2,575 psia, and 620 F

A.18-SPECIFIC GRAVITY OF SYNGAS FEED:

The molecular weight of this mixture was calculated in A.331 above. It is 7.772 #/#mol.

At the bed inlet conditions (2,675 psia,620 F),the specific gravity of Syngas is:

(7.772#/#mol/380scf/#mol)x(492°R/1,080°R)x(2,675psia/14.7psia) = 1.675 #/cf

A.19-SPECIFIC GRAVITIES OF THE EFFLUENTS:

The molecular weights of the two effluent mixtures are respectively:

Injected Effluent Bed Effluent

mol.weight	15.9	13.75
specific gravity		
@ 2575 psia,620 F	3.339 #/ACF	2.887 #/ACF
@ 2575 psia 800 F		2.474 #/ACF

A.20-SUPERFICIAL VELOCITIES OF GAS FLOWING THROUGH THE CATALYST BED:

The superficial velocity depends on the reactor configuration and on the cross-sectional area of the bed.

For the configurations of Fig.9 to 12 with four parallel sections,each one made up of two unit beds of cross-section 1,5 to 2.5 sq.ft,in series,each of the beds receives 1/4 of the total gas flow rate,in vertical flow.

Due to the large shrinkage of gas in the Methanation reaction,it is necessary to evaluate separately the superficial velocity at the inlet and at the outlet of the bed.The shrinkage factor is:

(4/4.5) + [(0.98 x 0.5)+(0.02 x 1)] + (0.5/4.5) = 0.5644

A.21-INLET SUPERFICIAL VELOCITY FOR VERTICAL FLOW:

The gas bed conditions at the inlet of the bed are:3.675 psia and 620°F (1 [illegible]).The Syngas mixture is essentially an ideal gas and the actual volume flow rate into each section is: Q =

100 MMSCF/D / (86,400 s/D x 4 sections) x [illegible] x 14.7/3.675 x

3.49 ACF/s.

For a bed cross section of 1.5 to 2.5 sq.ft this corresponds to a range of inlet superficial velocities of:

3.49 ACF/s / 2.5 sq.ft = 1.396 ft/s [illegible]

3.49 ACF/s / 1.5 sq.ft = 2.327 ft/s

A.22-OUTLET SUPERFICIAL VELOCITY FOR VERTICAL FLOW:

The bed effluent total flow rate is Qt = 100 MMSCF/D x (shrinkage factor), or: 100 MMSCF/D x 0.5644 = 56.44 MMSCF/D and, for each section Q = 56.44 MMSCF/D / 4 = 14.11 MMSCF/D

The hottest conditions at the outlet of the bed, corresponding to the highest velocity, are:

2,575 psia, 800°F.

The compressibility factor for such a mixture, of mol.weight 13.75 is approximately 1.02. Consequently, the actual volume flow rate at the outlet of each section of the bed is: Q =

(14.11 MMSCF/D / 86,400 s/D) x (1,260/492) x (1/1.02) x (14.7/2,575) = 2.3 ACF/s.

The corresponding superficial flow velocity at the outlet ranges from:

v = Q/S = 2.3 ACF/s / 2.5 sq.ft = 0.92 ft/s to:

v = Q/S = 2.3 ACF/s / 1.5 sq.ft = 1.55 ft/s

A.23-AVERAGE CONDITIONS FOR BEDS IN VERTICAL FLOW:

The representative conditions for each of the 4 bed sections are the average of those at the inlet and outlet. These are compiled below:

Parameter	Inlet	Outlet	Average	Case
v (ft/s)	1.396	0.92	1.158	S = 2.5 sq.ft
	2.327	1.55	1.938	S = 1.5 sq.ft
ρv (#/ACF)	1.695	2.775	2.215	
μf (#/ft.s)		T °F		
x(10,-5)	1.707	1.531 [illegible]	1.619	
		1.575 700	1.601	

		1.645	800	1.676
Kf(BTU/h.ft.°F)	0.118	0.0678	680	0.0929
		0.0687	700	0.0933
		0.0750	800	0.0965

A.24-CALCULATION OF DIMENSIONLESS PARAMETERS FOR BEDS IN VERTICAL FLOW

For vertical beds,it is advantageous to use a catalyst made of coarse particles,so as to reduce the pressure drop in the bed and to enhance the heat transfer to the water tubes.For these reasons,it will be assumed that the Methanation catalyst average particle diameter is Dp = 0.004 ft or Dp = 0.008 ft

The total mass flow rate / h is: G = 100 MMSCF/D x 7.778 #mol / (24 h/D x 380 SCF/#mol) = 85,285 #/h.It is divided among all the n reactor sections (4 minimum).Consequently V_f = G / (n x S x 3,600 s/h) Conversely,Re = G x Dp / (μf x n x S x 3,600)

The Reynolds Numbers for the two values of the bed cross section and the two particle sizes are at various temperatures:

Re	T (°F)	S (sq.ft)	Dp (ft)
587	680	2.5	0.004
1170	680	2.5	0.008
975	680	1.5	0.004
1950	680	1.5	0.008
584	700	2.5	0.004
1168	700	2.5	0.008
974	700	1.5	0.004
1948	700	1.5	0.008
565	800	2.5	0.004

1130	800	2.5	0.008
942	800	1.5	0.004
1884	800	1.5	0.008

TABLE 1a

SELECTED DESIGN PARAMETERS

Assumptions:

G (#/h) = 85,283 for entire reactor

Dp (ft)	0.004	0.008
Dp / Dt	0.154	0.154
Dto (ft) OD	0.026	0.052
Dti (ft) ID	0.0156	0.031
Tube Outer Surface So (sq.ft/ft)	0.0817	0.1634
Tube Inner Surface Si (sq.ft/ft)	0.049	0.098
Tube Inner Section (sq.ft)	0.00019	0.00076
Tube Outer Section (sq.ft)	0.00053	0.00212

Intermediate Results for vertical gas flow:

	S = 2.5 sq.ft	S = 1.5 sq.ft
V (ft/s) @ 800 F	1.158	1.938
μf (#/ft.s)×(10.-5) for 680 F < T effluent < 800 F		1.617 to 1.676
Kf (BTU/h ft °F)		0.0929 to 0.0965

Re	S (sq.ft)	Dp (ft)
585 to 565	2.5	0.004
1170 to 1130	2.5	0.008

975 to 942	1.5	0.004

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APPENDIX 4

PRODUCTION ENGINEERING OF THE WELL AND DRAINHOLES

7. TASK No.6:CALCULATION OF PRESSURE DROPS IN VARIOUS DOWNHOLE TUBES

It was shown in Appendix 3 that the average total fluids production expected from the horizontal drainholes associated with the Downhole Methanator was of the order of 18,000 to 24,000 B/D,with a GLR of 830 SCF/B and a WOR of about 2/1.The objective of Task No.6 is to determine the size of the liners in the horizontal drainholes and the size of the production tubing required to convey the produced fluids to the surface.The respective sizes of the Syngas feed tubing concentric with the production tubing and of the water tubing,which is concentric with the Syngas tubing,can then be determined,for the conditions specified.These are:

For the WEST SAK UPPER SANDS:

Typical Depth:3,500 ft

Fracturing pressure gradient >= 0.8 psi/ft or Pmax >= 2,815 psia

Reservoir fluid pressure:Pe = 1,800 psig = 1,815 psia

Net thickness of each sand:h = 30 ft

Average Permeability:K = 150 md.In the absence of any information on directional permeabilities,it is assumed that Kv/Kh = 0.25.

For the WEST SAK LOWER SANDS:

Typical Depth:4,000 ft

Fracturing pressure: 3,215 psia

Reservoir fluid pressure: Pe = 2,000 psia

Net thickness of sands of interest: h = 30 ft

Average Permeability: 150 md.

Assumptions:

In both cases, the average temperature in the drainage area is:

(620 + 75)/2 = 347.5 F

Water viscosity at that temperature is μw = 0.14 cp

Steam quality at the water tubes exit: 31 %

The produced oil viscosity at that temperature is estimated at 1.5 cp

The oil volume factor Bo = 1.4

The total reach of the horizontal drainhole is L = 3,000 ft

The radius of the drainage area is re = 1,500 ft

The drainholes in Phase 1 are operated in "huff and puff" and are in the production mode 50% of the time.

During Phase 1, the drainage area of each drainhole is 9 MM sq.ft = 206.6 ac.

Average flowing pressure into the drainhole liner: Pwf = 860 to 1,060 psia

7.2- PRESSURE DROP IN THE DRAINHOLE LINER

The pressure drop for a total flow rate of 20,000 B/D with a GLR of 330 is obtained from the corresponding graph (Fig.20) taken from Ref.44 (K.E.Brown), for a 5 in.ID liner, with an effective length of 3,000 / 2 = 1,500ft. The Bottom flowing pressure into the vertical production tubing, pwb, is then 820 psia for the case of the Upper Sands and 820 psia for the Lower Sands.

7.3- PRESSURE DROP IN THE PRODUCTION TUBING

A tubing size of 9.5/8 in.OD (8.921 in.ID) was selected.The pressure drop for the same flow rate (20,000 B/D,930 GLR) in vertical three phase flow is again determined from Ref.44 using the graph of Fig.21. The corresponding well head flowing pressure is:

160 psia for the Upper Sands (Vertical flow depth = 3,500 ft),and

180 psia for the Lower Sands (Vertical flow depth = 4,000 ft)

In either case,the well head pressure is sufficient to discharge the produced fluids into a separator at 100 psia operating pressure.

To minimize the outside diameter of the couplings,a VAM "special clearance" connection was selected,with an OD = 10.141 in.

7.4- VERIFICATION OF STEAM DELIVERABILITY

The worst case is that of the Lower Sands (Pe = 2,000 psia).

The injection pressure is 2,575 psia,for the mixture of steam and gases discharged from the Downhole Methanator.This is 240 psi below the fracturing pressure of the Upper Sands.

The relative permeability to gas is $K_{rg} \geq 0.6$

The average injectant viscosity is $\mu g = 1.598 \times 10.^{-5}$ #/ft.s = $2.38 \times 10.^{-2}$ cp.

The average temperature in the drainage area is 347.5 F = 807.5 R

The maximum flow rate into the reservoir for this pressure gradient is:

$Q_g = 5 \times 703 \times 0.150 \times 0.6 \times 30 \times [(2,575)^2 - (2,000)^2] / \{2.38.10^{-2} \times 807.5 [\ln (1,500 / 0.229) - 0.75]\} = 162$ MMSCF/D

The total volume to be injected is only 149.8 MMSCF/D/well,providing a 7.5 % margin.

The 5 in.ID liner of the drainholes is therefore acceptable even in the worst case.

7.5-SELECTION OF THE SYNGAS FEED TUBING

The Syngas feed tubing is concentric with the production tubing.Its drift diameter must exceed the 10.142 in.diameter of the VAM "special clearance" couplings of the 9.5/8 in.production tubing.

A 11.3/4 in.Hydril Super FJP tubing,with an ID of 10.772 in. and a drift diameter of 10.616 in.was selected.

Its annular cross section is 18.37 sq.in,equivalent to that of a tubing of 4.836 in.ID.The OD of the couplings of the 9.5/8 tubing is 10.141 in.for "special clearance" VAM joints.This leaves a clearance of 10.616 - 10.141 = 0.475 in. between the concentric pipes.Besides a reduced OD,another advantage of the Hydril Super FJP is to provide 3 metal/metal seals to prevent leakage of the Hydrogen-rich Syngas.
The catalytic bed pressure is about 2,700 psia at the inlet of the bed.Considering the pressure drop through the control valve and the pressure drop through the gas tubing,the injection pressure at the well-head may be about 3,000 psia.

A simple verification of the pressure drop is obtained with the chart of Fig.22,taken from Ref.45 (API),which gives a value of $P\Delta P/T$ per 100 ft of about 15. The minimum temperature is 300 F (760 R).
The general arrangement of the tubings and drainholes is that of Fig.10,described in Appendix 3.The Downhole Methanator is located in the large-diameter casing below the windows through which the drainhole liners are connected to the casing.This configuration is preferred when the diameter of the production tubing is large,because it leaves a greater part of the reactor cross section available for

the catalytic bed.The length of the reactor and valve section is estimated at about 250 ft.

The tubing length is,for the Lower Sands: 4,000 - 250 = 3,750 ft

Consequently ΔP = 15 x 37.5 x760 / 3,000 = 142.5 psi.

The grade of steel selected is C-75,which provides a collapse pressure of 3,070 psig and a burst pressure of 5,460 psig (well above the 3,000 psia gas pressure at the well head).The limiting design parameter is the collapse pressure.In the event of a failure of the Syngas feed compressor and of partial venting of the gas in the tubing,the full pressure of water in the annular space between the water tubing and the Syngas tubing might become applicable,thus requiring a collapse pressure in excess of about 2,800 psig for the gas tubing.

7.6- SELECTION OF THE WATER TUBING

The water tubing is concentric with the Syngas tubing.Accordingly, its internal diameter and drift diameter must both be greater than that of the Hydril Super FJP couplings of the 11.75 in.gas tubing,which is 11.9375in.

Its collapse pressure must be sufficient to withstand the full hydrostatic pressure of the Arctic Pack mud filling the annulus between this tubing and the large-diameter casing,in the event that the cooling water pump fails and that the water tubing empties while the Downhole Methanator shuts down.For the Lower Sands,the height of the mud column is 3,750 ft.With a maximum mud density of 10.0 #/gal,the minimum allowable collapse pressure is 1,946 psig.

A 13.3/8 in.tubing was selected,with a wall thickness of 0.514 in. and a drift diameter of 12.191 in.This leaves a clearance of .2555 in.

for running the Hydril gas tubing within the concentric water tubing. In order to meet the required collapse pressure with a safety factor of 1.4, corresponding to 2,725 psig, a high grade steel, P-110, was selected. The collapse pressure, for a wall thickness of 0.514 in. is 2,890 psig. The tubing ID is 12.347 in. and the drift diameter is 12.191 in. Regular API couplings are preferred in this case where the couplings OD is not critical in view of the large clearance available between the water tubing and the large-diameter casing.

The flow cross section is 11.3 sq.in., equivalent to a tubing of 3.793 in. ID. The volume of cooling water to be conveyed to the reactor is 149,300 #/h / 61.2 #/CF = 2,440 CF/h or 10,428 B/D, assuming that the average water temperature is 150 F. The corresponding water velocity is.

2,440 CF/h / (3,600 s/h x 11.3 sq.in / 144 sq.in/sq.ft) = 8.6 ft/s.

The pressure drop due to friction is determined from the chart of Fig.23, also taken from ref.45. For a flow rate of 10,428 B/D, the friction pressure drop is 3 psi/100 ft.

The total friction pressure drop is: 3 / 100 x 3,750 = 112 psi

The discharge pressure of the water tubes in the reactor is 2,625 psia. Assuming that the pressure drop on the water circulation system and water control valves is 250 psi, the maximum pressure applied by the cooling water on the external surface of the gas tubing is 2,625 + 250 + 112 = 2,987 psia or 3,002 psig, almost equal to the nominal limit of 3,000 psig previously considered for the gas tubing collapse pressure. The system of concentric tubings selected meets all conditions for safe operation.

7.7-CRITERIA FOR THE SELECTION OF CASING SIZE

A 24 in. casing was selected as the best compromise between conflicting requirements:1) a cross section sufficient to provide about 2.5 sq.ft for the catalytic bed cross section

2) a wall thickness sufficient to prevent collapse under the reservoir fluid pressure (2,000 psia).The collapse pressure is proportional to the cube of the ratio of wall thickness to diameter,thus requiring a larger wall thickness for a larger diameter pipe.The proportionality factor is directly related to the Young's modulus of the steel,so a high strength steel is preferable.

3) a total weight within the capacity of the heavy drilling rigs available on the North Slope of Alaska.

7.8-CEMENTING OF THE LARGE CASING

The largest high pressure API casing supplied by Mannesmann Rohrenwerke A.G.has a diameter of 24.5 in.,which would be suitable for this application.The upper part of the casing,above the reactor,will always be filled with drilling mud while it is lowered into the hole and will remain liquid-filled during cementing of the casing.This operation will be done by displacement of the cement slurry through an inner stinger and/or drill-pipe.This well-known method results in a considerable reduction of the weight applied by the cement slurry during its displacement.Other advantages of this technique include:

a) shorter pumping time

b) reduced contamination of the cement slurry before its flow behind the casing

c) the inside surface of the casing remains in contact with mud and stays clean.

Fig.24 shows the special stab-in cementing shoe and stinger (also

called stab-in unit) which are used for this operation.

7.9-PROTECTION AGAINST COLLAPSE OF THE LARGE CASING

The upper part of the casing,above the reactor is filled with Arctic Pack mud in the annulus around the 13.3/8 water casing. Consequently,this portion of the casing is not subject to collapse.

The bottom part of the casing,opposite the reactor,is normally pressurized by the Syngas feed flowing through the annulus between the casing and the reactor shell.The risk of casing collapse only occurs if the Syngas pressure is released.This may be prevented by installing a set of two check valves,one on the reactor inlet and the other on the outlet from the catalytic bed.Each check-valve will be pre-set to close respectively if the upstream gas pressure falls below the valve preset pressure value.In that event,the reactor will shut down and will gradually cool off under continued water injection.The gas pressure in the annulus between casing and reactor shell will drop as the gas cools down,but remains sufficient to prevent the casing from collapsing.

Should the reactor have to be disconnected from its hanger and brought to the surface for repairs,the Arctic Pack mud would first be displaced by brine of suitable density,which would be circulated out and replaced by fresh water before breaking the connection. All drainholes would also have been temporarily plugged with wireline retrievable plugs,prior to breaking their connection to the reactor.

Gas trapped in the casing below the reactor hanger would then be replaced by fresh water after this seal has been broken and while the reactor and water tubing are pulled out.Following reactor repairs,the reactor assembly,including the Valve Section,Connector and Cross-over

Sections,hung on the water tubing is lowered into the water-filled casing and threaded into the hanger.Water in the annulus outside the reactor shell below the reactor hanger is displaced by pressurized inert gas.Conversely, water above the reactor hanger in the casing annulus outside the water tubing is displaced by Arctic Pack mud.After reassembling the Syngas tubing and the production tubing,water is displaced out from the annulus between the Syngas tubing and the production tubing and replaced by pressurized inert gas.The temporary plugs in the drainholes are then removed and the reactor is now again ready for re-start.At all times during these operations the internal surface of the casing is pressurized and the risk of casing collapse while the reactor is in operation or being serviced is thus eliminated.

Another risk of casing collapse exists during the displacement of the cement slurry behind the casing.The weight of the slurry is about 15.4 #/gal.With a 10 #/gal mud inside the casing,the net external pressure applied by the cement slurry,in the case of the Lower Sands is: (0.800 psi/ft - 0.519 psi/ft) x 4,000ft = 1,124 psig
With a safety factor of 1.5,the required collapse pressure is 1,686 psig.To meet this requirement with a 24.5 in OD pipe,a high streength steel is required (C-75 for seamless pipe or X-70 for ERW pipe) and a wall thickness of about 1.00 in.Mannesmann Rohrenwerke appears to be the only supplier for seamless pipe of this size,but several suppliers are available for ERW pipe of that size and larger,most of them are in Japan and in Europe.

An alternative which would be worth considering in the future is the possibility of using a stage-cementing operation to reduce the height of the column of cement slurry and thus reduce its hydrostatic

pressure. Such an operation would require the inclusion in the casing string of several stab-in stage cementing collars. These are readily available.

The possibility of using a lower density cement in the upper part of the well could also be investigated when more information becomes available regarding the fracturing pressure gradient in the uphole section.

The use of either of these two cementing alternatives would result in significant reductions in the collapse pressure specification and consequently in the casing thickness and weight.

7.10-DESIGN OF THE REACTOR HANGER

In view of the large weight of the reactor assembly and of the necessity of establishing a gas-tight seal between the gas-filled annulus opposite the reactor shell and the annulus filled with Arctic Pack mud above the reactor, a threaded connection providing a reliable metal/metal seal is preferred to the use of a conventional-type liner hanger designed for the large-diameter casing. The location of the hanger and that of the connections of the drainhole liners with the casing are known in advance and the use of the drill pipe for cementing the casing makes it possible to insert between two casing sections a short joint containing a buttress thread on an internal upset of sufficient wall thickness and length to withstand the vertical load. The reactor head can then be fitted with a threaded male connector which, when screwed into the threaded upset joint, provides both the required seal and the structural support for the reactor assembly. The male connector thread is located on an external upset with respect to the outside diameter of the reactor shell.

Fig. 5 shows in concept the reactor being run-in through the casing prior to mating the threaded hanger into the casing internal upset. The internal upset thickness in the casing joint is typically 0.5 in.

The male connector OD is equal to the casing ID, minus a 0.25 in. clearance. This leaves 0.375 in. for the maximum thread height. Conversely, the reactor shell OD is equal to the drift ID of the casing, opposite the buttress thread, minus a 0.25 in. clearance.

7.12-DETERMINATION OF THE SIZE OF THE REACTOR SHELL

From the previous analysis, the outside diameter of the reactor shell is then: 24.5 - 2 x (1 + 0.5) - 0.25 = 21.25 in. When the reactor is in operation, the gas pressures inside and outside the reactor shell only differ by the friction pressure drops due to the gas flow, of which a major component is the pressure drop through the catalytic beds. This pressure drop is only a few 100 psi, which produces a very low stress in the reactor shell. When the reactor is lowered into the casing or pulled out for repairs, the casing is filled with fresh water and the pore space of the catalytic bed is filled with pressurized inert gas, at a typical pressure of 1,700 psia. The shell thickness is thus determined by this internal gas pressure acting only when the reactor is cold and under external pressure of 14.7 psia, or under an external water pressure of 2,000 psia, resulting in a net external pressure of 300 psi. The design conditions are then: 1,715 psig burst pressure and [illegible] psi collapse pressure.

The alloy steel selected for the shell [illegible] for [illegible] at high temperature in the presence of [illegible] Hydrogen gas. A suitable alloy is Hastelloy C, with a yield strength of 100,000 psi and a Young's modulus, at 800 F, of 25 million

psi/in. A shell thickness of 0.5 in.is sufficient to meet those two conditions,neglecting the internal support provided by the catalyst particles tightly packed inside the shell,which would increase the resistance to collapse by buckling of the shell.Consequently the ID of the reactor shell is 20.25 in.This shows that the design assumption of a bed cross section of 2.5 sq.ft is not achievable,since the internal cross section of the reactor shell is only 2.225 sq.ft and includes not only the annular bed but also the system of concentric pipes required for the reactor internals.If these internals can be limited to a maximum diameter of 3 in.,the bed cross section will effectively be limited to 1.89 sq.ft.This will correspondingly increase the height of the bed to: 370 CF / 1.89 sq.ft = 196 ft.

When the additional heights required for the connectors between units and sections and for the reactor head and valve section are considered,the previously assumed total height of 250 ft still appears conservative.The reduced bed cross section and increased length will of course,increase the gas pressure drop through the bed,to about 250 psi for two unit beds in series.This confirms that the stress in the reactor shell is quite low when the reactor is in operation.

7.13-DESIGN OF REACTOR INTERNALS

The reactor concept is that of Fig.10,in which the cooling water feed tubing occupies the central part.With a maximum flow of 2,440 CF/h opposite the top section of the reactor,decreasing to 810 CF/h opposite the bottom section,a small-diameter tubing can be used,corresponding to the average flow rate of 1,625 CF/h.The average water temperature is (100 + 216) / 2 = 158 F,for which the water viscosity is only 0.29 10-3 #/ft.s and its specific gravity is

61 #/CF.A 2 in.ID (2.1/4 in.OD) pipe size is selected to permit the use of 1.75 in.OD wireline tools.The corresponding flow velocity is:

1,525 CF/h / (3,600 s/h x 3.1416 sq.in./ 144 sq.in./sq.ft) = 19.4 ft/s

The Reynolds number is: 19.4 x 0.166 ft x 61 #/CF / 0.29 10-3 #/ft.s = 677,000 .The corresponding Fanning friction factor is 0.0045 and the pressure drop by Fanning's equation is:

$\Delta P = 0.323 \times 0.0045 \times (61/62.4) \times (16.4)^2 \times 250 / 2 = 47.8$ psi,which is acceptable.

Concentric with this water pipe is the feed gas pipe.A 3 in ID pipe produces a pressure drop $P\Delta P$ / T(R) =< 50 psi / 100 ft ,or for 250 ft length and P = 3,000 psi.T = 1,085 R the pressure drop is =< 65 psi. The OD of this pipe,made of Hastelloy C,is only 3.5 in.

Concentric with this Syngas pipe are the eight steam/water separators.For a water/steam contact at 90 % of the separator height the water flow through each separator is:

2,440 CF/h / (0.31 x 8) = 983.8 CF/h or 122.6 gpm and the length of the flow path in each separator is about 0.9 x (196 ft / 8) = 22 ft

It is necessary to maintain the natural circulation of water in the water tubes,in which the sub-cooled water level is:

0.65 x (196 ft /8) = 15.9 ft,according to section 6.6.

The head loss due to friction in each of the 82 water tubes is less than 1 ft of water.Consequently the required head in the separator tubes must exceed 15.9 + 1 = 16.9 ft of water.It follows that the head loss due to water friction in the separators must be limited to about 5.1 ft of water over the 49 ft height of the separator (a pressure gradient of 4.5 psi/100ft).This corresponds to a flow cross-section equivalent to that of a 2.75 in.ID pipe.

The annulus between the 3.5 OD gas pipe and a pipe of 4.5 in.ID

more than satisfies this condition.This third concentric pipe is made of alloy steel compatible with high temperature steam.such as Croloy TP 321 H,with a high yield strength.In view of the low stress to which it is subjected,the pipe OD is only 5.0 in.

Finally.a fourth pipe,adjacent to the catalytic bed and concentric with the steam/water separator pipe,provides an annulus for the flow of the injectant mixture.With a reactor cooling water rate of 149,400 #/h,the total mass flow rate of injectant is 566,043 #/h.
Its pressure is 2,575 psia and its temperature 620 F (1,080 R).
It was shown in section 7.4 that an injection pressure of 2,520 psia is sufficient to provide the required deliverability from the drainhole.A ΔP of 55 psi in the injectant collector space over the total length of the bed (196 ft),or 28.1 psi/100 ft,is acceptable.
The friction pressure gradient for this steam-rich mixture may be obtained from Fritsche's formula.

The graph of Fig.25 (taken from Ref.17) shows that a cross-section equivalent to that of a 5.25 in.ID pipe is sufficient.The annulus between a 7.65 in.ID (8 in.OD) pipe and the 5.0 in.OD pipe is sufficient to provide such a cross section.This confirms the possibility of including all the reactor internals within a 8 in. pipe,as assumed in section 7.12.The alloy steel selected for this pipe,in direct contact with the hot catalytic bed and with its Hydrogen-rich feed gas is again Hastelloy C.

7.14- POTENTIAL FOR FURTHER OPTIMIZATION

The gas pressure drop through the catalytic beds could be further reduced by selecting slightly larger catalyst particles (Dp > 0.008 ft).It could also be reduced by increasing the bed cross section.This

could be achieved by tapering the various concentric pipes providing the reactor internals.This tapering would reflect the gradual consumption of cooling water from unit to unit starting from the reactor top.It could also reflect the gradual consumption of Syngas from the bottom of the reactor to the top.

By interchanging the respective uses of the central injectant collector space and of the casing/reactor shell annulus,the stress in the reactor shell may be changed from compression (collapse) to tension (burst).This might make it possible to reduce the wall thickness of the reactor shell,thus saving on Hastelloy costs.

Another option,which would maximize the bed cross section,consists in using the annulus between casing and reactor shell as a single separator for the steam and water effluent from all units.The steam outlet and the cooling water inlet would be located in the reactor head,thus eliminating the need for the central water distribution tube within the reactor.The reactor internals would then consist of only two concentric tubings,respectively distributing the Syngas feed and collecting the effluents from all the beds.Their outer diameter could be reduced to less than 7 in.

The possibility of designing the water tubes system in the catalyst bed as a set of forced circulation,once-through,high pressure steam generator instead of a natural circulation boiler should also be investigated as a way of further simplifying all the reactor internals.Such an approach would result in a small reduction of the steam quality (to about 70%) in the injectant mixture,but it might simplify the operating controls of the reactor,and also reduce slightly the volume of the reactor internals.

Only a limited amount of design optimization was done.The results

are reflected in the dimensions given in Table 6,which are slightly different from those indicated in Section 7.13,and which were used for cost estimating purposes.

Too little information is presently available on the strength of the rocks in the West Sak reservoir to evaluate whether the horizontal drainholes will have to be gravel-packed in order to withstand the large flowing pressure gradient without excessive sand production,but this important question should also be addressed in future studies.

7.15- CONCLUSIONS OF TASK No.6

1) The calculation of pressure drops in the flow of fluids entering or leaving the reactor and the selection of the various tubular goods required for these flow systems further demonstrates the technical feasibility of the application of this process to the West Sak.

2) The production rate of 18,000 to 24,000 B/D with a GLR of 830 and a WOR of about 2 may be achieved with 3,000 ft long horizontal drainholes lined with a 5 in. ID perforated liner and with a 9.5/8 in. vertical tubing.

3) This work also provides a firm technical basis for estimating the cost of all the downhole facilities.

4) These facilities use available alloy steels,conventional pipe sizes and known technologies to reduce technical risk.This greatly enhances the reliability of the cost estimates.

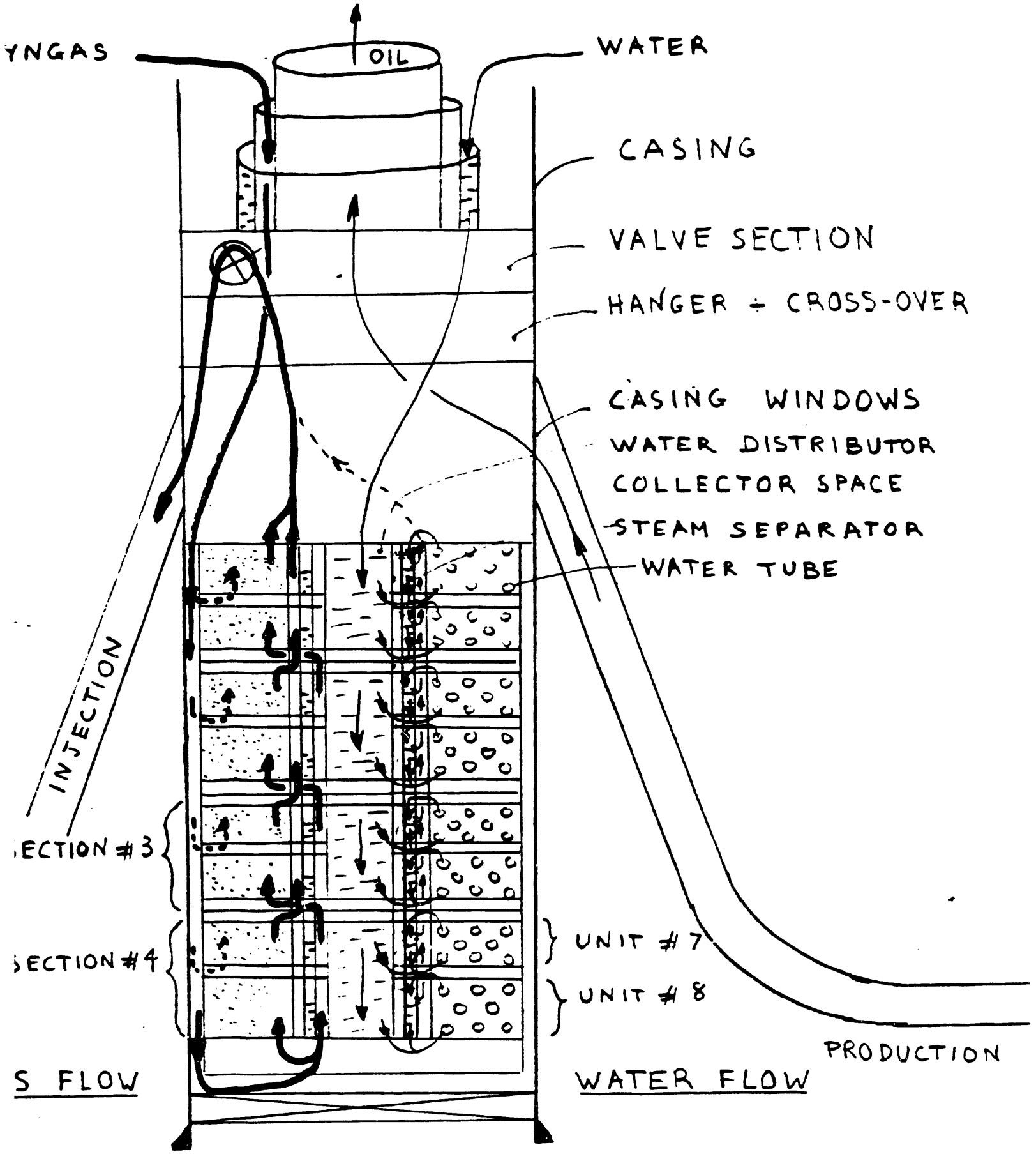
REFERENCES:

42) Craft,B.C. and Hawkins,M.F.,"Applied Petroleum Reservoir Engineering",Prtentice Hall,1959

43) GPSA Engineering Data Book,1987

44) K.E.Brown,"The technology of artificial lift methods,Vol.3a Pressure Gradient Curves,Pennwell Pub.Co.,Tulsa,OK,1980

45) API-Recommended Practice for Design and Installation of Offshore Production Platforms Piping Systems,1984



IG. 10. VERTICAL FLOW REACTOR - TOP DISCHARGE

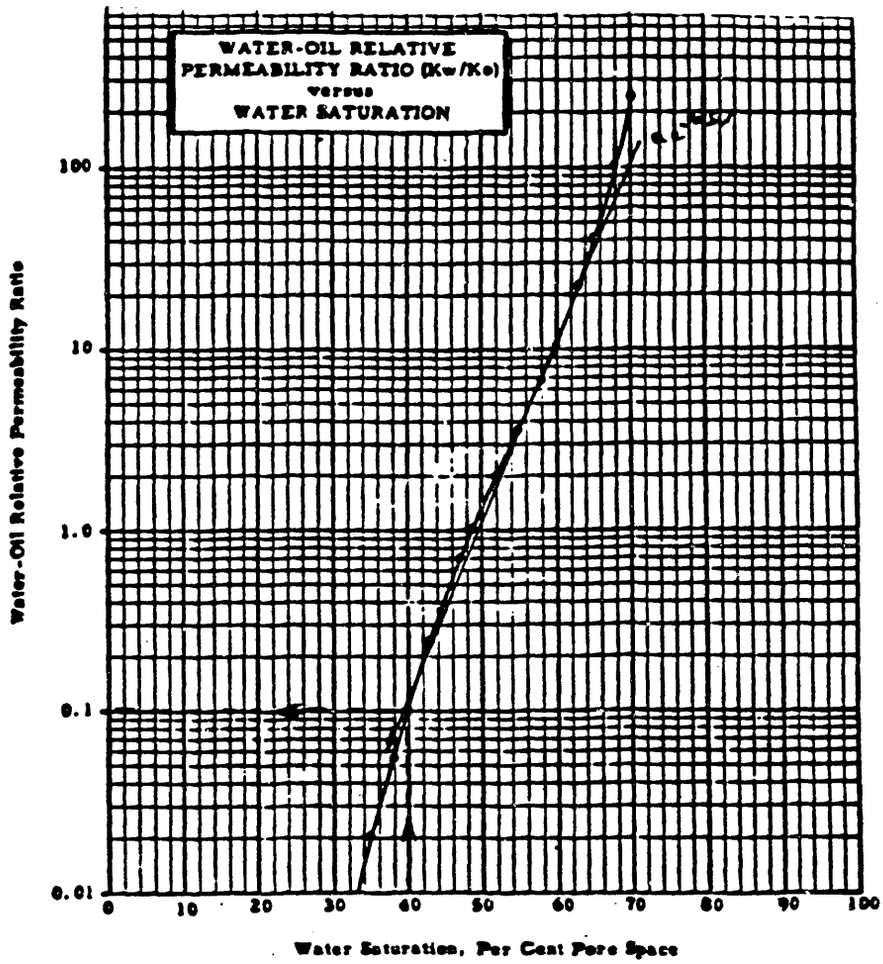


FIG. 19a

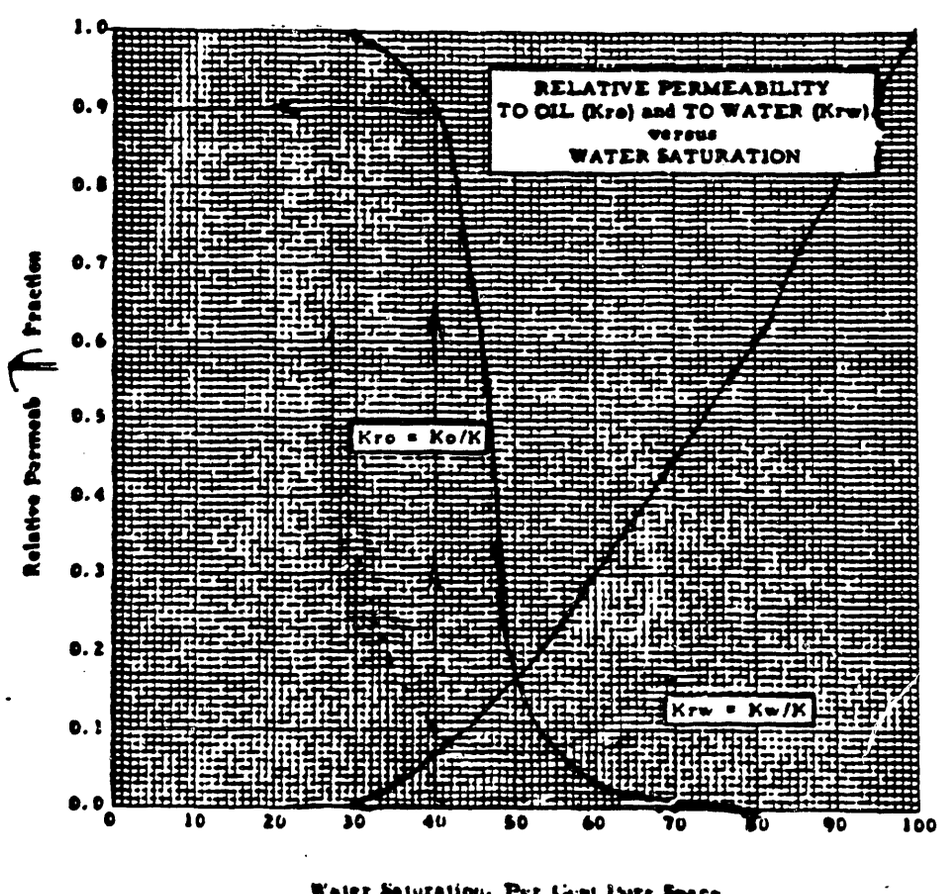


Fig. 19b Typical Oil and Water Relative Permeability Curves.

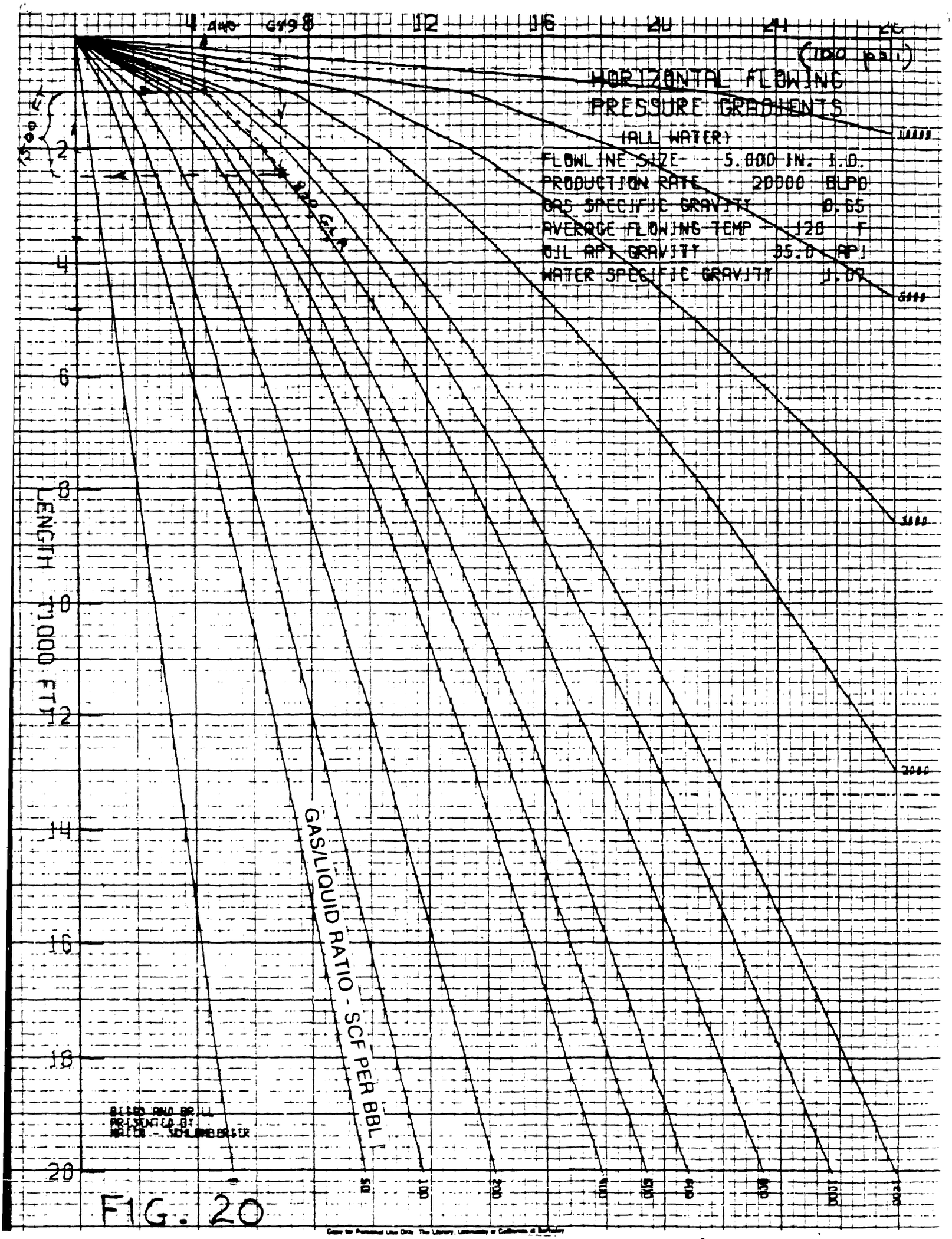

HORIZONTAL FLOWING
PRESSURE GRADIENTS
(ALL WATER)
FLOWLINE SIZE 5.000 IN. I.D.
PRODUCTION RATE 20000 BLPD
GAS SPECIFIC GRAVITY 0.65
AVERAGE FLOWING TEMP 120 F
OIL API GRAVITY 35.0 API
WATER SPECIFIC GRAVITY 1.07
(100 psi)
LENGTH 1000 FT
GAS/LIQUID RATIO - SCF PER BBL
4
8
12
16
20
24
2
4
6
8
10
12
14
16
18
20
50
100
200
400
600
800
1000
1200
5000
3000
2000

FIG. 20

FIG. A-662

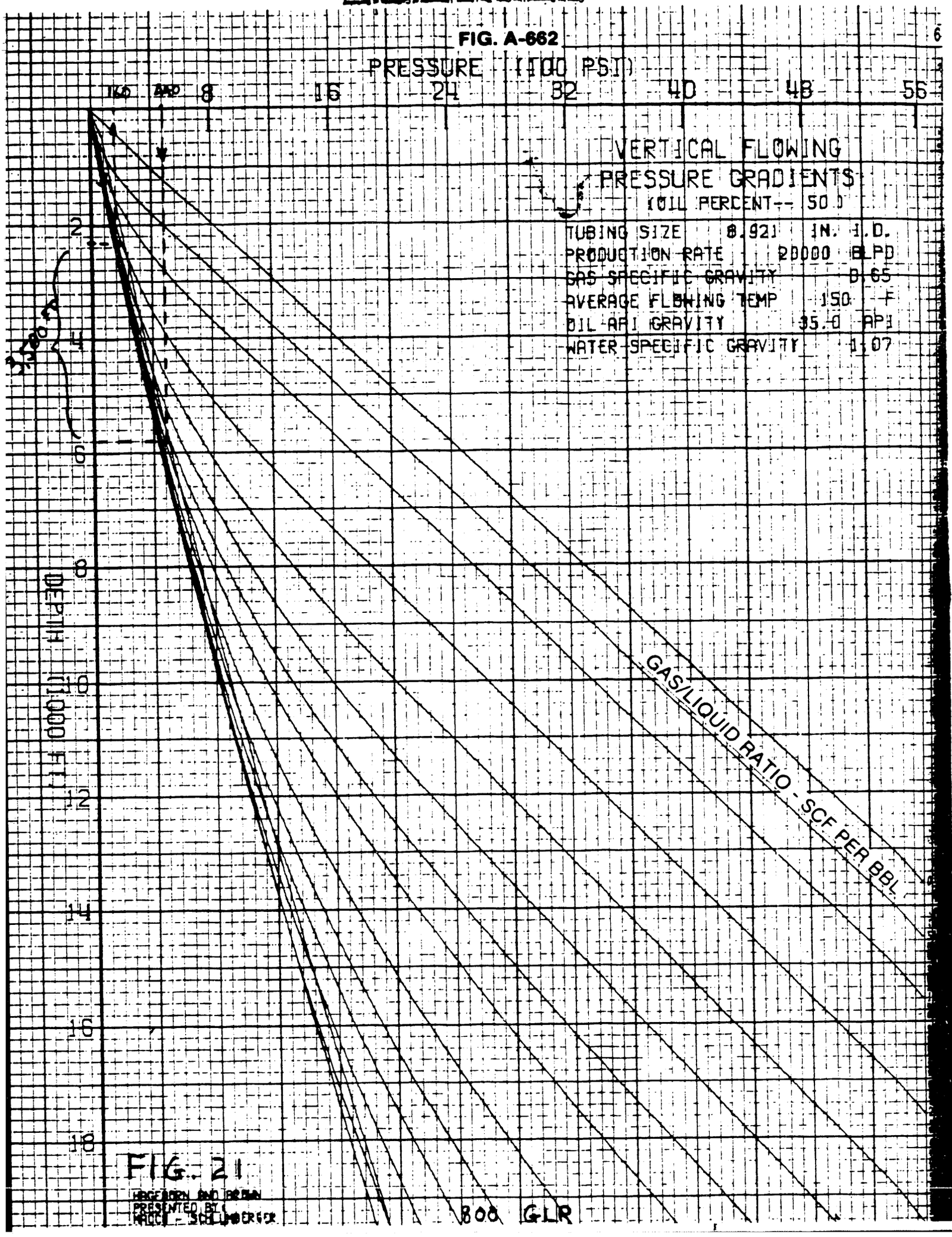


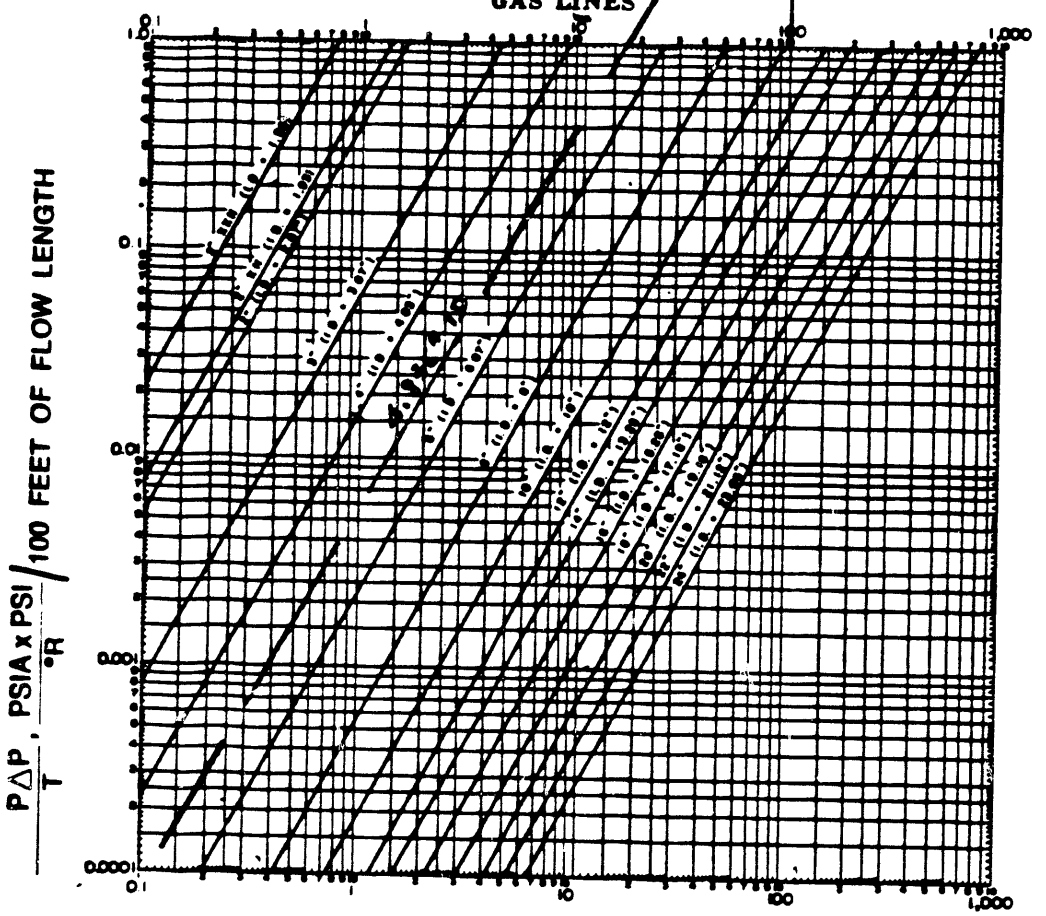
FIG. 21

HAGEDORN AND BROWN
PRESENTED BY:
HALCO - SCHLUMBERGER

FIG. 22

≈15 psi/100 ft for Mw = 7.8 #/lb mi

FIGURE 2.4
PRESSURE DROP IN SINGLE-PHASE GAS LINES



NOTES:

1. To find pressure drop, multiply ordinate by $\frac{T\ °R}{P_{PSIA}}$ and by mol. wt. correction factor. Mol. wt. correction factors not tabulated may be interpolated.
2. Length of flow includes the equivalent pipe length due to friction in fittings.
3. Do not use these curves when total pressure drop exceeds 10% of line pressure.

MOLECULAR WEIGHT CORRECTION FACTORS

1000-100,000 MCF/DAY

	4"-24"		Approximation		
M.W.	20	1.05	- 5%	to	+ 5%
M.W.	40	1.97	-10%	to	+10%
M.W.	60	2.89	-15%	to	+15%
M.W.	80	3.66	-15%	to	+15%

100-1000 MCF/DAY

	2"-4"				
M.W.	20	1.05	- 5%	to	+ 5%
M.W.	40	1.99	-15%	to	+15%
M.W.	60	2.47	-20%	to	+20%
M.W.	80	3.10	-25%	to	+25%

4. Pressure drop equation (Fanning) used is based on friction factor values given by the NGPSA Data Book (1948-1957 Editions).

$$\Delta p = \frac{S_g\, L\, f\, T\, Q_g^2}{14.9\, P\, d_i^5} \cdot 1000$$

Q_g = gas flow rate, MMCF/D @ 14.7 psia & 60°F.
S_g = gas specific gravity (air = 1). $\left(\frac{\text{MOL. WT.}}{28.9}\right)$
T = flowing temperature, °R (460 + °F).
L = equivalent length of line, feet.
P = operating pressure, psia.
d_i = pipe inside diam., inches.
Δp = pressure drop, psi.
f = friction factor from NGPSA graph, or

$$f = \frac{0.021}{\left(\frac{d_i V_g \rho_g}{\mu_g}\right)^{.38}} + 0.0025$$

V_g = average gas velocity, ft/sec.
μ_g = gas viscosity at flowing pressure and temperature, centipoise.
ρ_g = gas density at flowing pressure and temperature, lbs/ft.^{3}

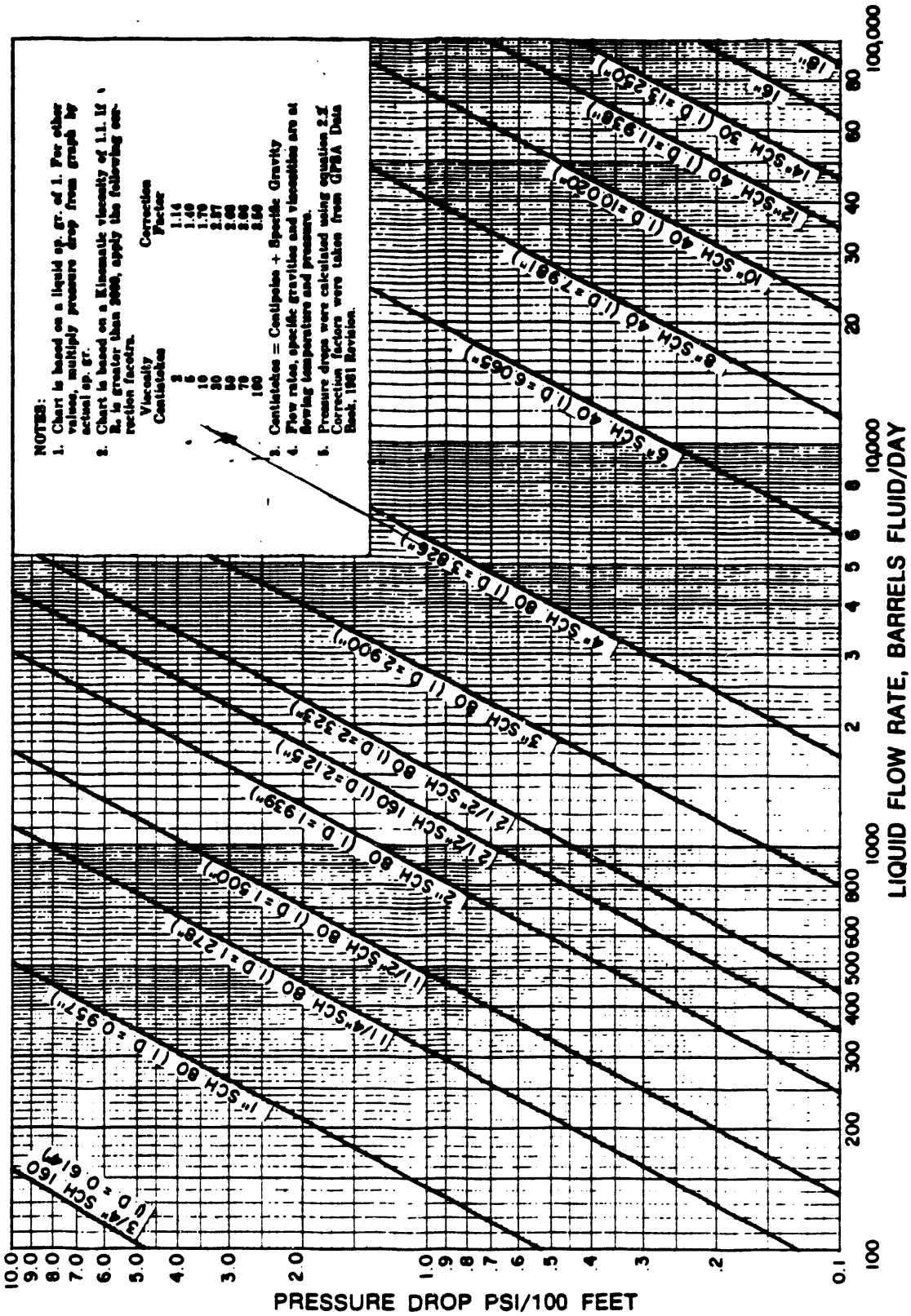


FIGURE 23
PRESSURE DROP IN LIQUID LINES

FIG. 24

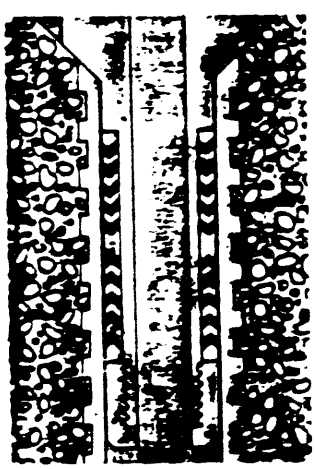
STAB-IN CEMENTING EQUIPMENT

BAKERLINE®
A DIVISION OF BAKER INTERNATIONAL CORPORATION

Bakerline stab-in equipment is used for efficient and safe cementing for large diameter casing.

Advantages of Stab-In

- Cement displaced by drill pipe
- Cement mixing can be terminated at any time
- Casing protected: cementing pressures confined to drill pipe (like a squeeze job)
- No plug(s) required to get even less contamination and greater displacement accuracy
- Minimal drill-out required
- Stab-In allows tighter control of cementing operation
- Stab-in takes the guesswork out of figuring cement volumes
- Stab-in cuts down the possibility of flash setting in casing due to pumping a large displacement volume
- When cement return appears, only the volume remaining in the drill pipe has to be displaced
- Stab-In reduces pump displacement time by cutting the cement volume needed (Conventional 20 in. casing volume at 2,000 ft is 710 bbls of cement. Stab-In cementing through 4-1/2 in. drill pipe requires only 28 bbls of cement in the pipe at 2,000 ft)



STAB-IN SEAL SUB IN STAB-IN FLOAT SHOE

CHEVRON SEALS MAKE THE DIFFERENCE

The chevrons seal more dependably than conventional O-ring seals. The chevron configuration eliminates the risks of swab-off, and eliminates the chances of damage caused by seals

BAKERLINE STAB-IN CEMENT FLOAT SHOE
Product No. 254-03

Recommended for shallow to medium depth conductor and surface casing cementing jobs where displacement volumes can be accurately controlled.

BAKERLINE STAB-IN CEMENT FLOAT COLLAR
Product No. 255-02

BAKERLINE GUIDE SHOE
Product No. 102-01

Recommended for large diameter casing cementing from shallow to deep applications where a "shoe joint" is needed to prevent over-displacement of cement. A float collar is desired to remove the back-pressure device from the bottom of the string.

BAKERLINE STAB-IN STAGE COLLAR
Product No. 200-40 (not shown)

This product provides the advantages of the Bakerline Model J Stage Cementing Collar for large diameter

BAKERLINE STAB-IN CEMENT COLLAR
Product No. 255-03

BAKERLINE FLOAT SHOE
Product No. 100-01

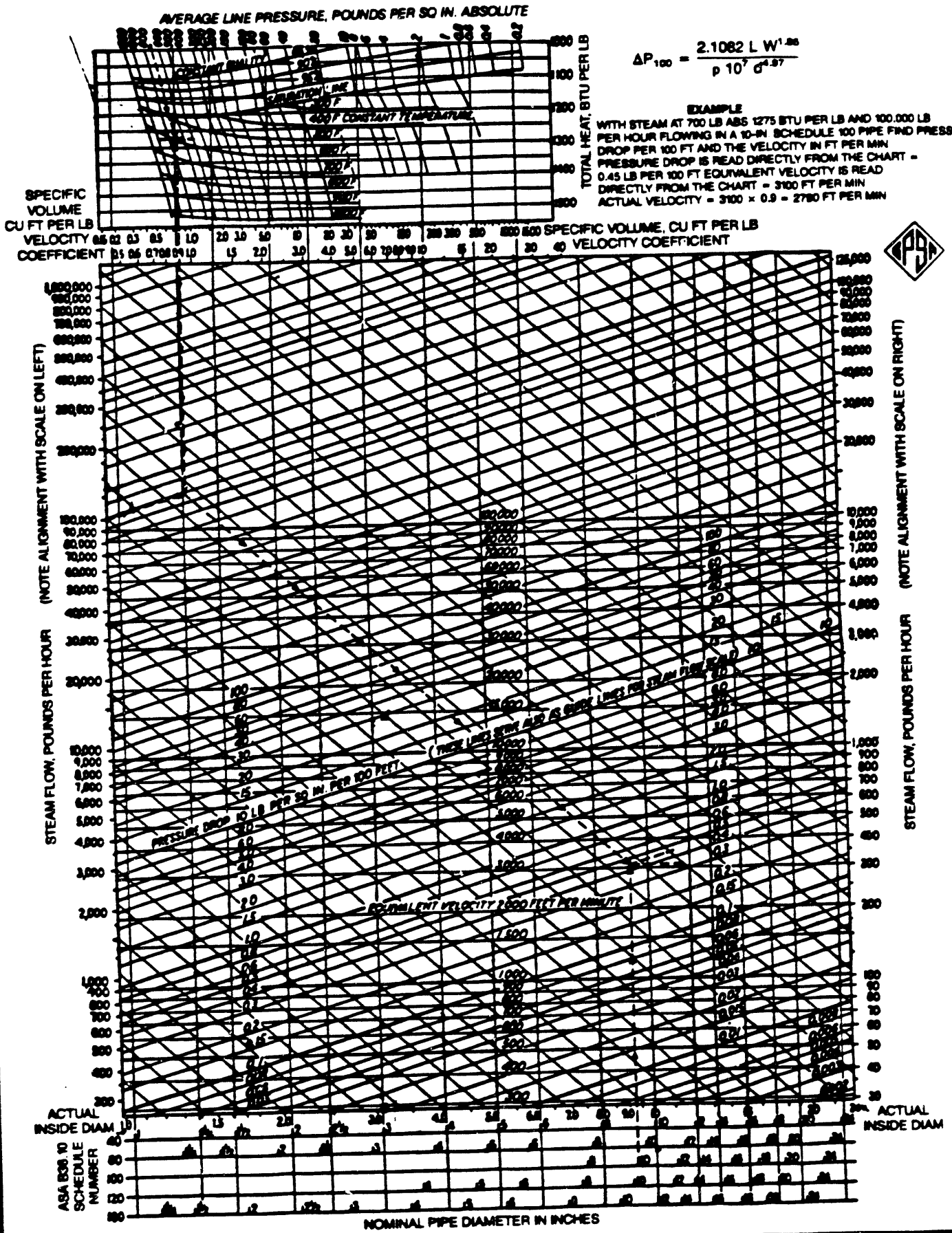
Recommended for large diameter casing cementing from shallow to deep applications where a "shoe joint" is required and a float shoe is needed to enhance cementing set-up around the "shoe joint".

BAKERLINE STAB-IN NON-CIRCULATING FLEXIFLOW FILL-UP COLLAR
Product No. 255-05

BAKERLINE GUIDE SHOE
Product No. 102-01

Recommended for applications where it is desired to provide an automatic fill feature as casing is run in the hole. Flexiflow Collar is converted to back-pressure device by applying pressure to drill pipe after stabbing seals into collar seal bore.

Pressure Drop in Steam Pipes by Fritzsche's Formula[11]



$$\Delta P_{100} = \frac{2.1082\ L\ W^{1.86}}{\rho\ 10^7\ d^{4.97}}$$

EXAMPLE

WITH STEAM AT 700 LB ABS 1275 BTU PER LB AND 100,000 LB PER HOUR FLOWING IN A 10-IN SCHEDULE 100 PIPE FIND PRESSURE DROP PER 100 FT AND THE VELOCITY IN FT PER MIN PRESSURE DROP IS READ DIRECTLY FROM THE CHART = 0.45 LB PER 100 FT EQUIVALENT VELOCITY IS READ DIRECTLY FROM THE CHART = 3100 FT PER MIN ACTUAL VELOCITY = 3100 × 0.9 = 2790 FT PER MIN

FIG. 25

Contract:DE-FG 01-89 CE 15446

APPENDIX 5

OPERATING SCENARIO OF THE WEST SAK PROJECT AREA UNDER INJECTION OF STEAM AND GASES

8.-THREE PRODUCTION PHASES

Over a lifetime of about 30 years,most of the large oil fields are subject to several different modes of exploitation,taking place sequentially,in order to maximize oil recovery and cumulative cash flow.

For the West Sak project,three main phases have been assumed:

1) "Huff and puff" operation of multiple drainholes connected to vertical wells,each one equipped with a Downhole Methanator/Steam Generator.Well drainage area:ca.14,000 ft x 3,000 ft (see Fig.26) over a minimum of 4 separate layers.

2) In-fill drilling of production wells and drainholes at mid distance between the original wells,for conversion to a steamflood.Drainage area of each injector:ca.14,000 ft x 6,000 ft.(see Fig.27)

3) Heat scavenging by water injection in the original (Phase 1) drainage area and re-use of the Downhole Methanators/Steam Generators in new wells,drilled adjacent to the Phase 2 drainage area.

An alternative Phase 2 operating mode is to transfer the Downhole Reactors from the Phase 1 area to newly drilled large-diameter wells in between the original wells,which are converted to heat scavenging.

8.1- PROJECT ARRAY

With a Syngas plant consisting of four trains,each one with a sufficient capacity to serve two Downhole Reactors,and with the Phase 1 wells drilled on a nearly square grid,the project area is a rectangle,or parallelogram,of sides ca.14,000ft x 28,000 ft.

In Phase 1,only 8 large diameter wells are drilled,each of them equipped with a Downhole Reactor and connected to at least 4 pairs of horizontal drainholes.The horizontal distance between wells is ca.6,300 ft,when the drainholes coincide with the inter-well line,to ca.7,000 ft,when the wells are staggered (see Fig.28)

In Phase 2,in-fill drilling adds 10 production wells in the case of conversion to a steamflood.These are normal size vertical wells,each one connected to at least 4 pairs of drainholes.In the case of the transfer of "huff and puff" operations to the median areas,the in-fill wells are the more costly,large-diameter wells required for being fitted with the transferred Downhole Reactors.

8.2-REASONS FOR CHANGING THE OPERATIONAL MODE

Major problems with steam injection methods of oil recovery are the tendencies towards steam fingering and steam override,which may result in by-passing some of the oil in place.The use of horizontal drainholes drilled in the lower part of the reservoir results in the formation of a steadily increasing steam chamber,along the sides of which the heated oil flows downward by gravity,towards the drainhole during its production ("puff") part of the cycle.The advantages of this type of counter-current flow of steam and oil are well described by R.M.Butler in Ref.46.

A very important feature of the Phase 1 "huff and puff" mode

of operation is that the heat loss to the overburden is reduced as a result of the fact that the heat front in the overburden is alternatively advancing and receding.This maximizes the effectiveness of the injected steam to mobilize the oil in the zones closest to the drainholes.This is an effective way of reducing the pressure drop around the drainhole and of maintaining the well productivity.

8.3-PROPOSED PHASE 2 MODE OF OPERATION

Ultimately,the upper part of the steam chamber extends over the entire drainage area (>= 216 ac.) of the drainhole and heat losses to the overburden become comparable with those experienced in a steam flood,and the production part of the "huff and puff" cycle peaks at a much lower rate and extends over very long periods.This is the main justification for conversion to a steam flood in Phase 2.

Phase 2 is predicted to begin after about 15 years of cyclic steam injection.By then,the oil saturation around the original drainholes has been significantly reduced and the viscosities of oil and water in that zone have also been drastically reduced.This situation makes it easy to inject steam at a high rate,a primary requirement for a successful steam flood.The steam flood fronts in the pattern shown on Fig.27 effectively starts at a distance of about 1,500 ft from the injection drainholes and has another 1,500 ft to travel before reaching the producing drainholes.Over such long distances,the reservoir hetereogeneities,sicous fingering and gravity segregation are expected to make the frontal advance very irregular.The over-all sweep efficiency of the steam and gases is assumed to drop from 70 % in Phase 1 to only 40 % in Phases 2 and 3.On that basis,material balance predicts that the total duration of these two last phases will

also be about 15 years. It is also expected that during Phase 2 the OSR (Oil/Steam Ratio, a determining economic factor) will steadily decline.

8.4-PHASE 3 OPERATION AND OVER-ALL RESULTS FORECAST

Phase 3 will begin when the irregularity of the steam front causes early break-through of the injectant at the producing drainholes. To offset the declining OSR, water will be substituted for steam in the injector drainholes responsible for the break-throughs. Heat stored in the reservoir rock will vaporize some of the injected water, thus increasing the total volume of steam behind the front. Heat scavenged from the rock will also provide a hot water front traveling behind the steam front. Although this hot water is less effective than steam to displace heavy oil, its contribution to oil recovery is not insignificant and its cost is low.

The Downhole Reactors, pulled out of the wells converted to water injection, are then transferred to virgin zones outside the original pattern, as illustrated on Fig.27, for a new "huff and puff" project. The lifetime of the reactors was estimated at 30 years, with periodic replacement of the catalyst and routine repairs of the internals.

With this scenario, the over-all West Sak recovery after 30 years is calculated to reach about 44 %, a value consistant with steam injection results in several mature Heavy Oil fields of California.

8.5-DISCUSSION

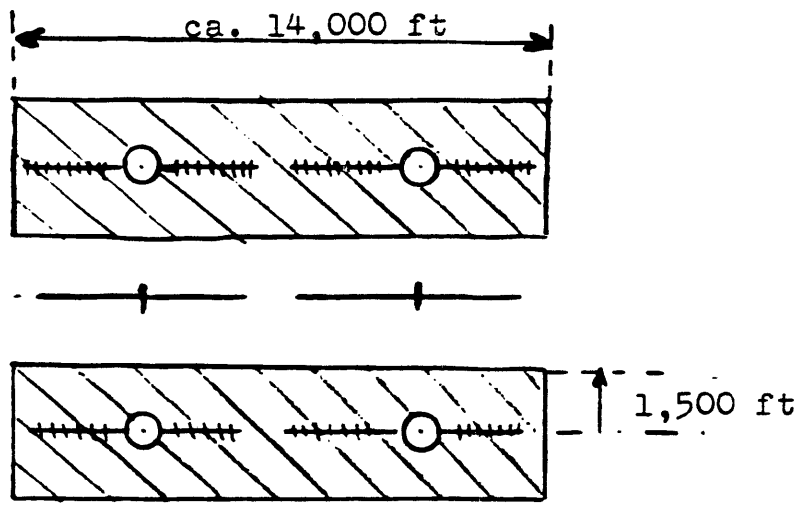
For lack of detailed information about the West Sak reservoir, no specific area was considered for this scenario, but, in a field covering 300 sq.miles, it is unlikely that a suitable areal portion of only about 14 sq.miles could not be found, which would correspond to the

general conditions assumed here. With a potential OOIP of up to 25 Billion Barrels, it is highly probable that several parts of the field covering more than 5% of the total area, could be exploited under the proposed scenarios.

REFERENCE:

R.M. Butler, "A New Approach to the Modeling of Steam-Assisted Gravity Drainage", J.Can.Pet.Tech. (May-June 1985, P.42)

PHASE 1 DRAINAGE (MARX-LANGENHEIM)



PHASE 2 (ALTERNATIVE 1) :

IN-FILL DRILLING AND REACTOR TRANSFER

PHASE 2 (ALTERNATIVE 2) :

CONVERSION TO STEAM FLOOD

PHASE 3 (IN BOTH CASES) :

HEAT SCAVENGING BY SEA WATER INJECTION

FIG. 26 . OPERATING SCENARIOS

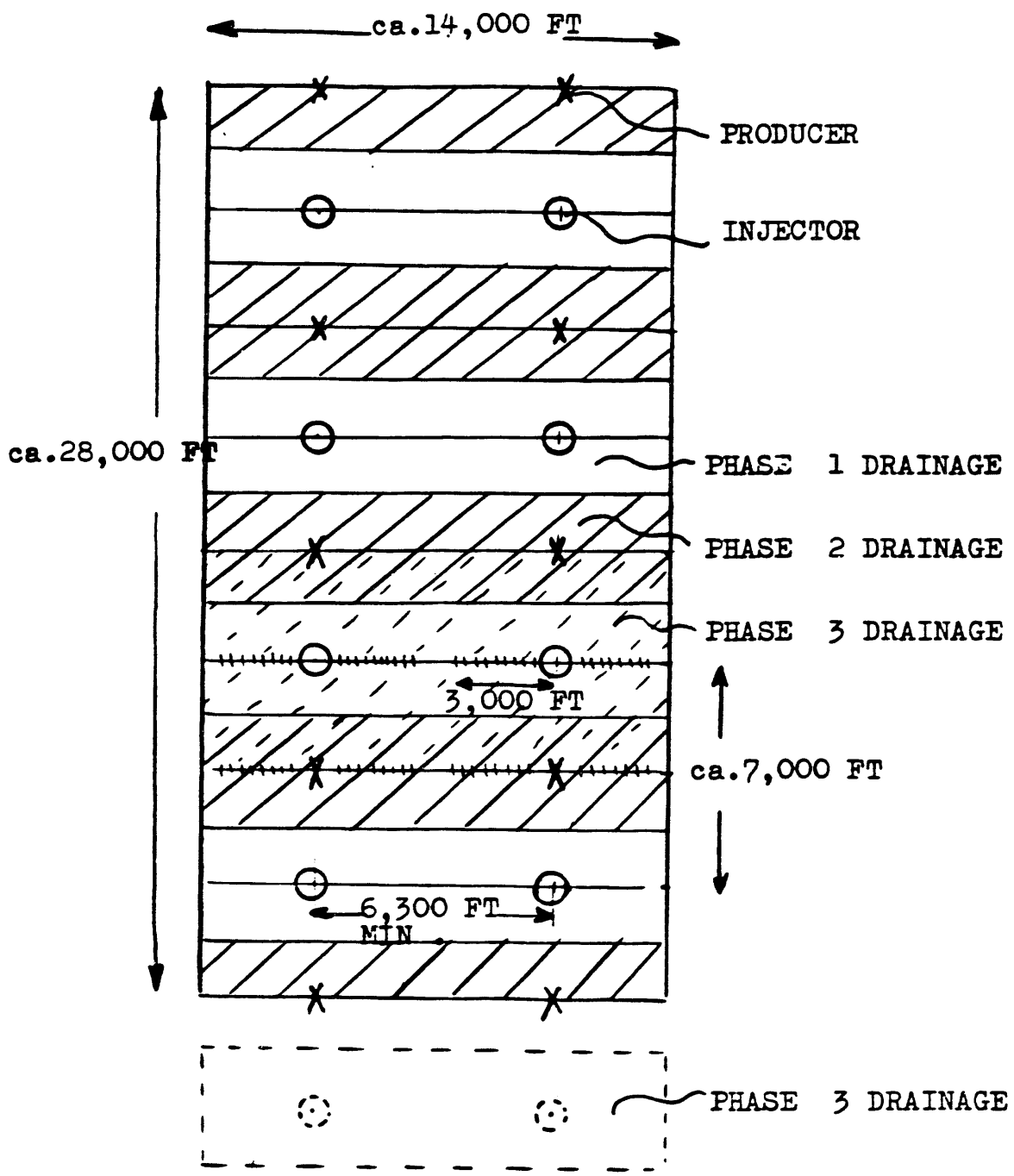
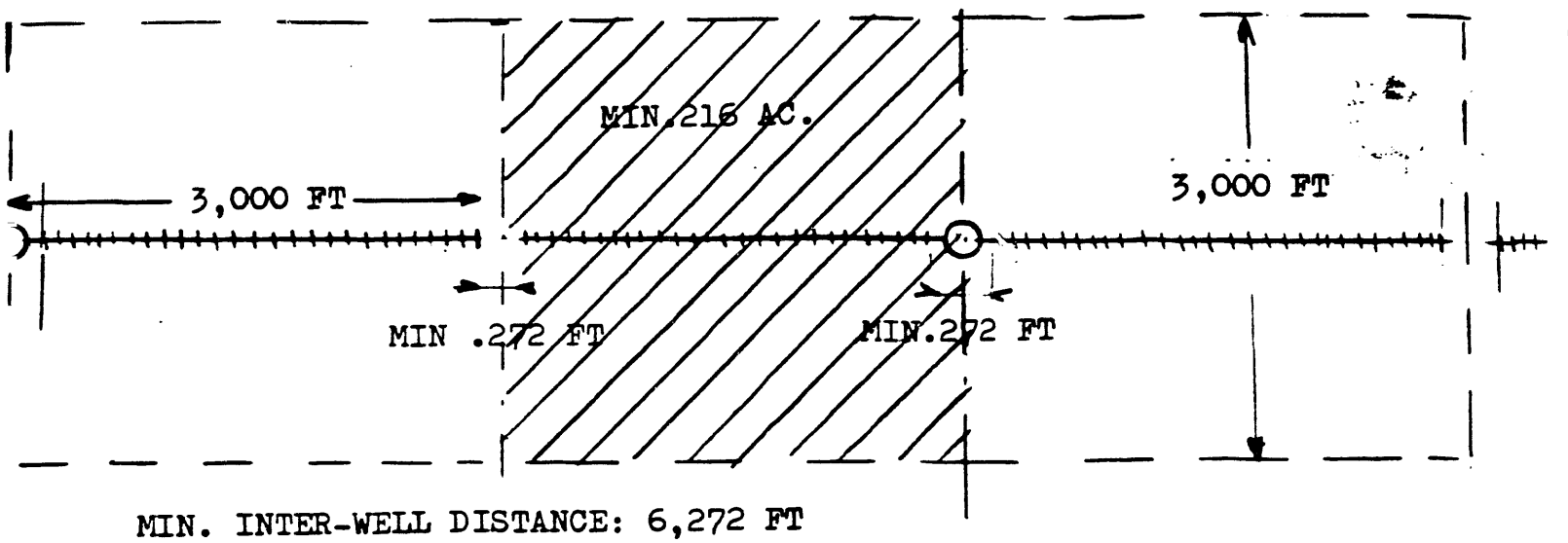


FIG.27 EXAMPLE OF PROJECT ARRAY (4 ROWS OF 2 INJECTORS)

N.B. WITH STAGGERED WELLS, THE ARRAY IS A PARALLELOGRAM



DRAINAGE AREA OF DRAINHOLES IN PHASE 1

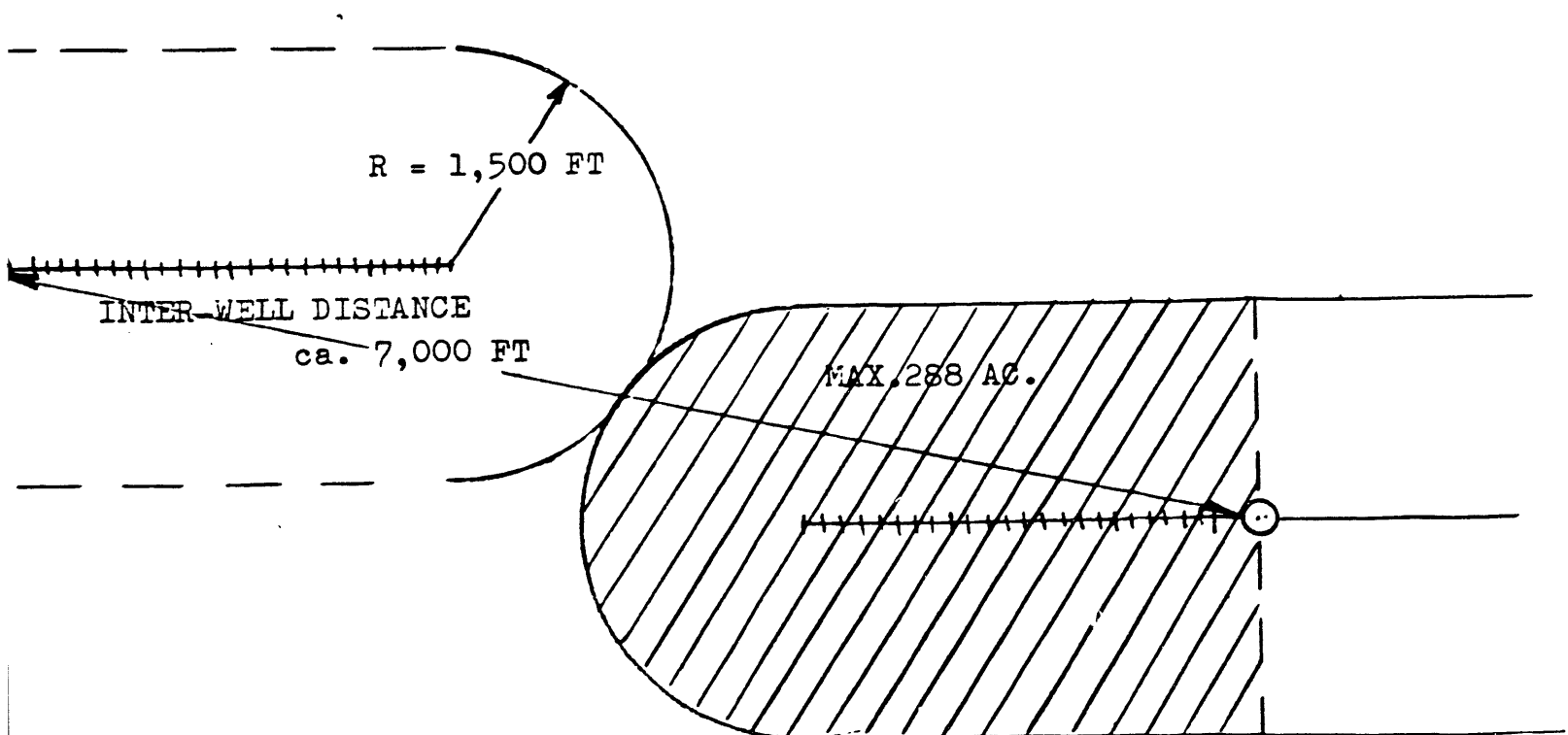


FIG. 28 - PHASE 1 DRAINAGE AREAS

END

DATE FILMED

01/06/92