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GREAT CANADIAN OIL SANDS EXPERIENCE IN THE COMMERCIAL PROCESSING OF ATHABASCA TAR SANDS

By

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INTRODUCTION

The Great Canadian Oil Sands plant for recovery and upgrading of bitumen from Athabasca Tar Sands officially opened on September 30, 1967. This opening ceremony represented the climax of many years of experimentation, design, construction, and start-up effort that at its peak utilized the services of 2500 people. Some key dates leading up to the event were:

1953 - Great Canadian Oil Sands Limited organized.
1960 - Initial attempt to obtain permission to develop the Athabasca tar sands. Action deferred (1) by the Alberta Oil and Gas Conservation Board until 1962.
1962 - Approval granted (2) by the Oil and Gas Conservation Board for a plant to recover 31,500 barrels per day of synthetic crude.
1963 - Detailed engineering work underway. Construction and operation of an experimental plant 20 miles north of Fort McMurray, Alberta on the west bank of the Athabasca River.
1964 - Approval granted (3) for an increase in the design capacity oι the plant to 45,000 barrels per day. Began construction work on the commercial plant.
1965 - Completed road and bridge connecting the plant site with Fort McMurray. Completed work in the experimental plant.
1966 - Engineering work completed. Construction effort reached its peak.
1967 - Completed construction. Plant start-up. Official opening ceremony.

Some of the design and development problems encountered have been the subject of recent papers. For example:

1. Discussions of problems encountered in the development of the hot water process as utilized by G.C.O.S., including a description of the test plant (4,5).
2. A discussion of the overall commercial plant flow sheet, material and energy balances, and scale-up problems (5).
3. A geological description of the area to be mined by G.C.O.S., with emphasis on the unusual geological exploration problems encountered and how they were overcome (6).
4. A discussion of the properties and processing characteristics of synthetic crude oil produced from Athabasca bitumen in a pilot scale simulation of the G.C.O.S. plant (7).

We wish to build on the background developed in these papers and discuss the commercial plant performance, but first let us take a brief look at the plant itself. Figure 1 shows the plant location in Northeastern Alberta, 20 miles north of Fort McMurray and 270 miles north of Edmonton. Toronto is 1700 miles to the southeast, Salt Lake City 1100 miles directly south, and Seattle 800 miles to the southwest.

Figure 2 is an aerial view of the project. The tar sand mining area is in the background, facilities for extracting bitumen from the tar sand are to the left rear, and process units for upgrading bitumen to synthetic crude are in the center.

Figures 3 through 7 show a series of plant scenes in the same order as the plant operating

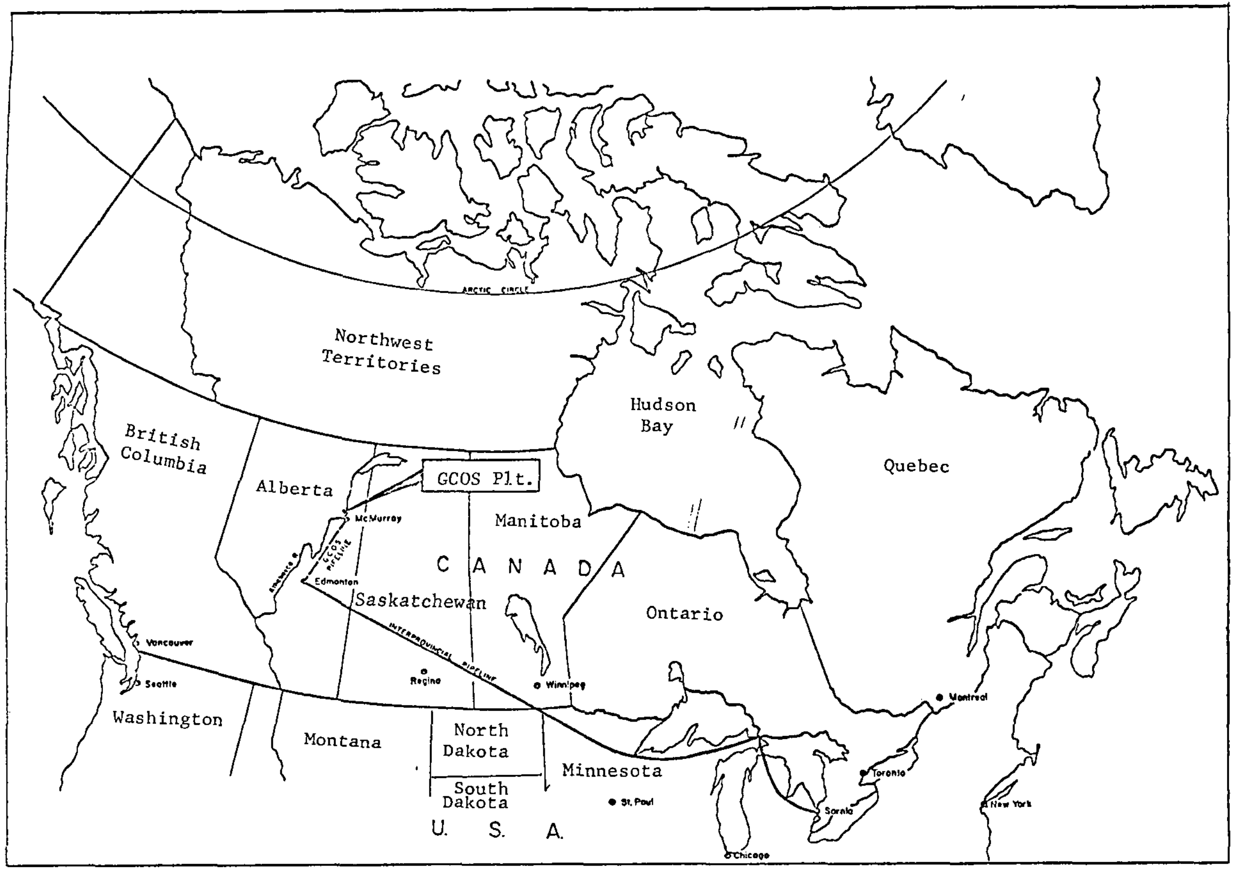

FIGURE 1

MAP SHOWING GEOGRAPHICAL LOCATION OF GREAT CANADIAN OIL SANDS PLANT SITE

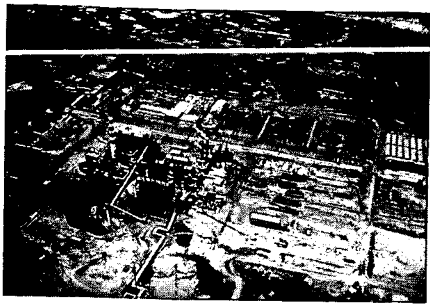

FIGURE 2

AERIAL VIEW OF THE G.C.O.S. PLANT

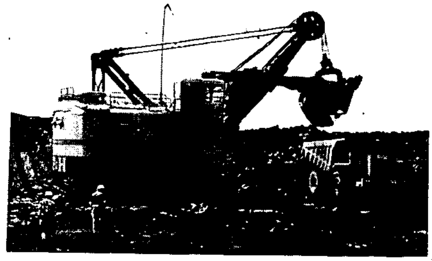

FIGURE 3

OVERBURDEN REMOVAL

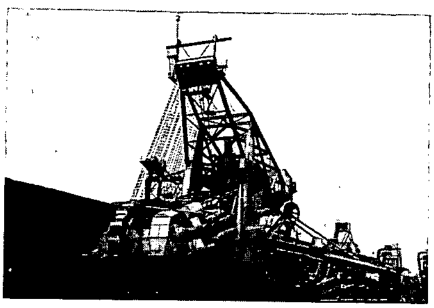

FIGURE 4

BUCKETWHEEL EXCAVATOR

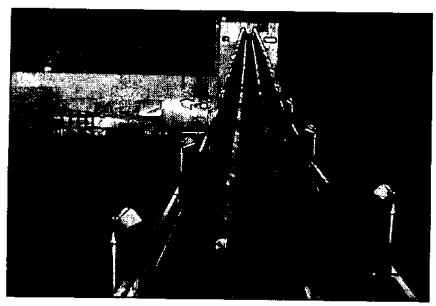

FIGURE 5

FEED CONVEYOR

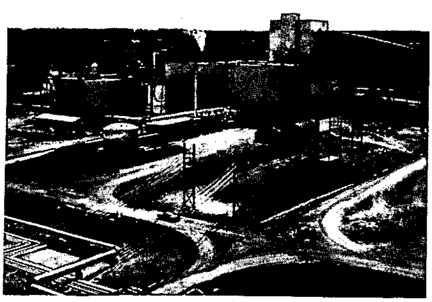

FIGURE 6

EXTRACTION PLANT

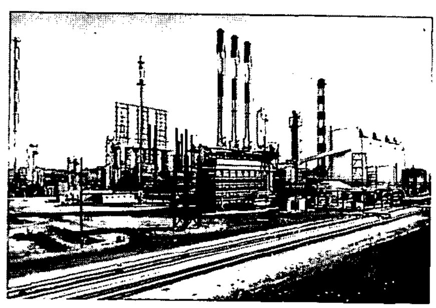

FIGURE 7

BITUMEN PROCESSING AREA AND POWERHOUSE

(Sulfur plant on left, cokers in left center, hydrorefining units in center, hydrogen plant in right center, and powerhouse on the far right) sequence. First the overburden layer is removed (Figure 3) to expose the ore body. Bucket-wheel excavators (Figure 4) mine the tar sand and transfer it to a series of belt conveyors (Figure 5). These conveyors discharge at the extraction plant (Figure 6) where bitumen is separated in a two stage extraction. Bitumen is then upgraded to synthetic crude oil by coking and hydrorefining (Figure 7).

Some essential auxiliary steps in the plant complex are:

1. Disposal of the sand tailings from the extraction plant operation.
2. Hydrogen production.
3. Sulfur recovery.
4. Steam and power generation.
5. Water treating.

Figure 8 (5) shows in more detail how all of these operations are integrated to produce synthetic crude oil from tar sands at the rate of 45,000 barrels per calender day.

Before turning our attention to the performance details of the GCOS plant we would like to mention that this report was prepared in October, 1967 and is based on observations during the startup period (July, August, and September, 1967). Therefore, our data and conclusions should be interpreted as a progress report and not as a final technical evaluation.

RESULTS AND DISCUSSION

Mining

The G.C.O.S. mine is one of the largest open pit operations in North America in terms of daily tonnage. To produce 45,000 barrels per calender day of synthetic crude oil requires, on the average, 97,500 tons of tar sand per calender day. About 0.5 tons of overburden per ton of tar sand must be removed to expose the ore body. This means about 3.3 tons of ore plus overburden must be handled to produce a barrel of synthetic crude oil.

Overburden removal is accomplished by using both scrapers and a power shovel - truck combination. This operation is about 1 year ahead of the mining, so, on a short term basis, it is not critically related to daily production. The total overburden that must be moved averages 12 million cubic yards per year.

Tar sand is mined using two crawler mounted bucketwheel excavators, each with a theoretical capacity of 9,000 tons per hour and nominal average capacity rating of 5,500 tons per hour. This equipment is relatively new to the North American mining scene, although it has been extensively used in Europe, for instance in German brown coal operations.

Because of the non-homogenous nature of the tar sand deposit, G.C.O.S. operating experience with the bucketwheel excavators has varied widely. Operation at up to 8,000 tons per hour per wheel has been demonstrated. On the other hand, occasional rock seams reduce the quantity that can be mined.

Tar sand from the excavator is moved to the plant by a series of belt conveyors. Although belt conveyors have been widely used in other mine operations, there are some new problems peculiar to the handling of tar sand. An earlier paper (4) discussed the problem of tar sand sticking to the conveyor belt as noted during the operation of the G.C.O.S. test plant (1963-65). In inital operation this proved to be at least as bad as expected, if not worse. Not only did the accumulated tar sand cause unbalanced loads resulting in belt training problems; it caused considerable wear on tension pulleys, idlers, and belt scrapers because of the abrasive nature of the material. We made a number of modifications to the system that, while not a complete solution, reduced the problem to a tolerable level. Cleanup of tar sand spills around the conveyor system is an expensive nuisance.

Abnormal loads imposed on the conveyor system by rocks and large lumps of tar sand have caused some mechanical damage. This damage consisted of both broken impact idlers at the transfer points in the conveyor system and holes punched in the belt itself. Replacement of the impact idlers with larger, heavy duty units substantially reduced this problem. Belt repairs still consume maintenance manpower, and when the damage is too extensive to repair during normal shutdowns, contribute to lost production time.

In addition to the mechanical problems associated with mining, the nature of the ore body creates problems in mine planning. On the average, it takes 2.2 tons of tar sand to produce one barrel of synthetic crude oil. However, this can vary from 1.5 to 4.0 tons per barrel depending on the bitumen content of the tar sand. Similarly, the fines content (through 325 mesh) of the ore body varies so that the fresh water requirement in the extraction process can vary between 25 to 300 gallons per ton of tar sand. In turn, this causes a shift in steam demand. Maintaining stable operating conditions in the extraction plant, smooth power plant operation, and a steady flow of bitumen to the process units requires close attention to the nature of the ore body.

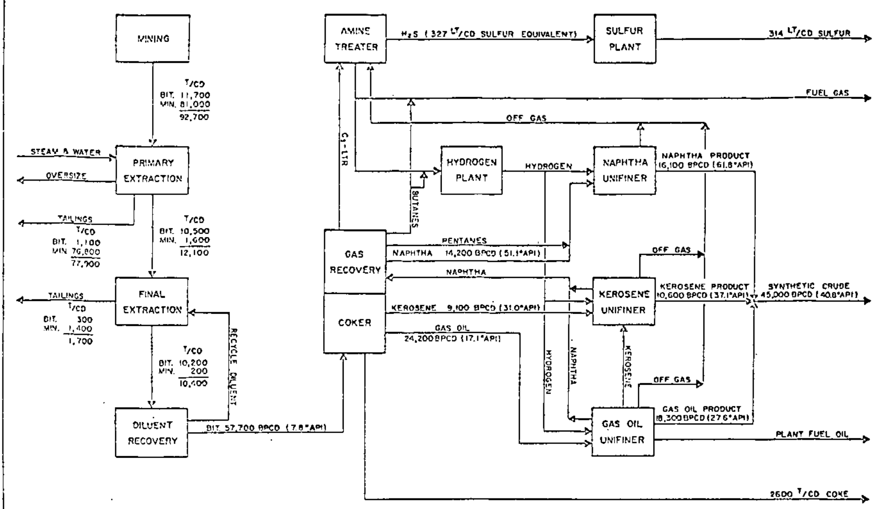

FIGURE 8

GENERALIZED FLOW SHEET SHOWING MAJOR PROCESS STEPS AND MATERIAL FLOWS

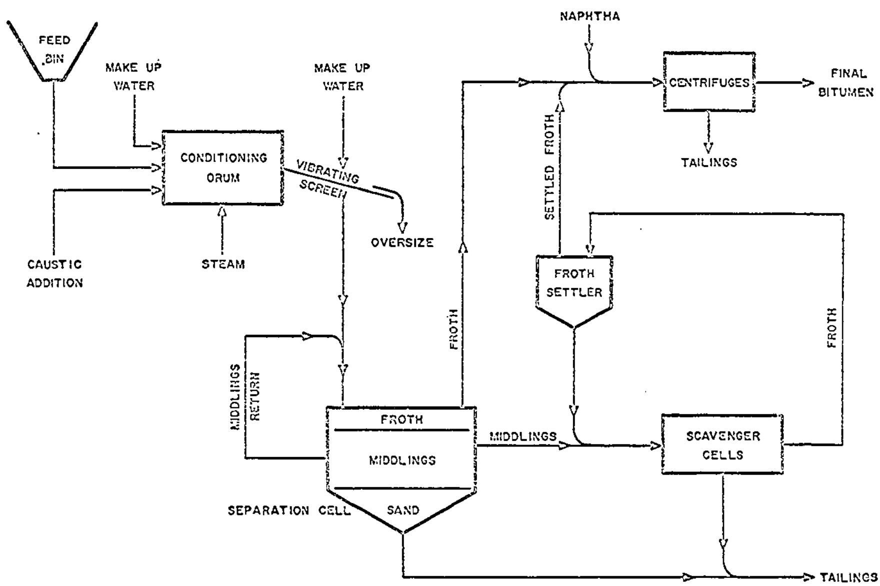

FIGURE 9

EXTRACTION PLANT PROCESS FLOW SHEET

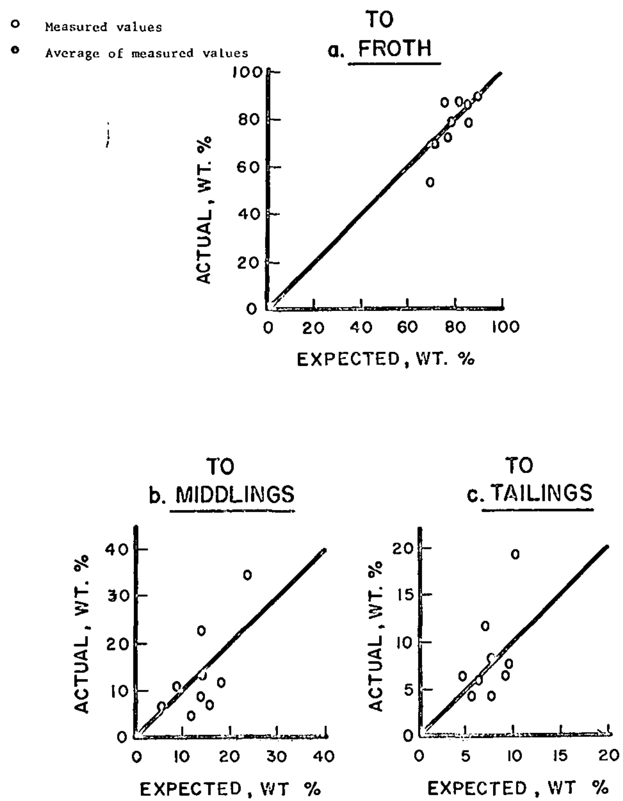

FIGURE 10
SEPARATION CELL BITUMEN DISTRIBUTION

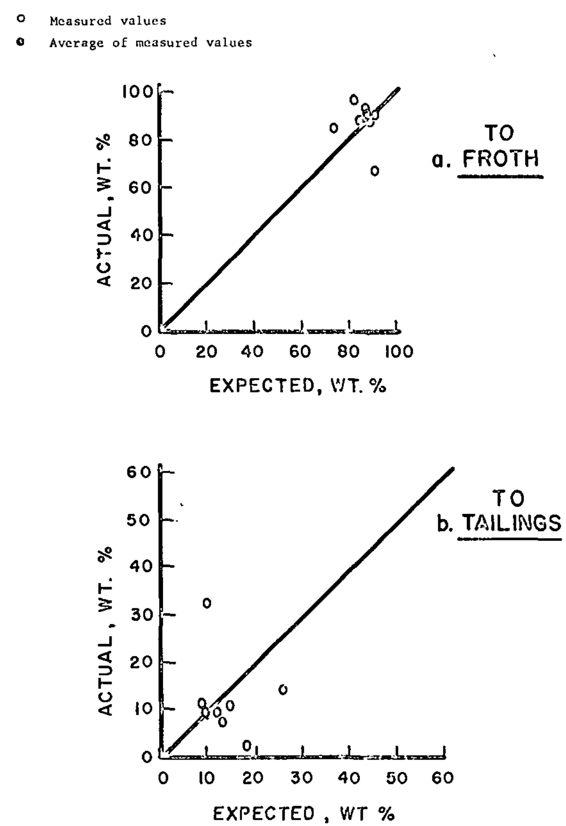

FIGURE 11
SCAVENGER CIRCUIT BITUMEN DISTRIBUTION

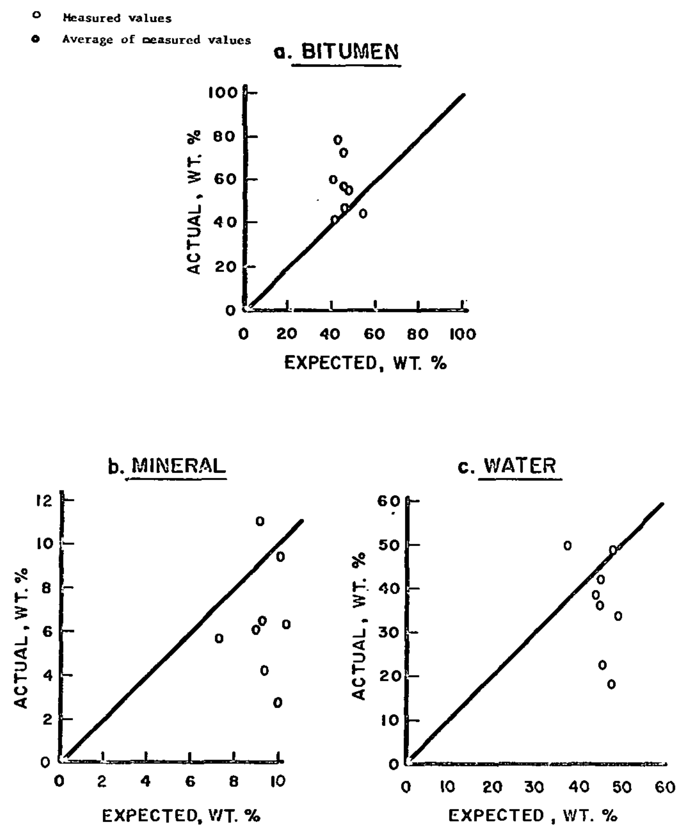

FIGURE 12
SEPARATION CELL FROTH COMPOSITION

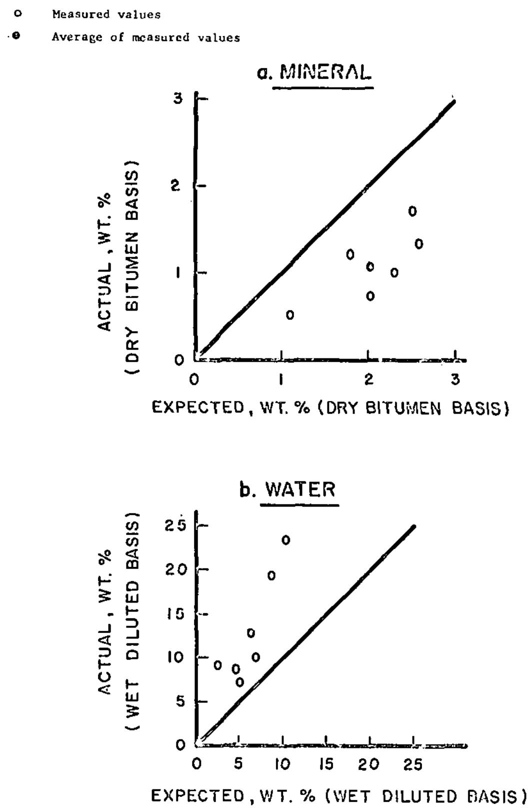

FIGURE 13
MINERAL & WATER IN FINAL EXTRACTION PRODUCT

Bitumen Extraction

Tar sand discharges from the mine conveyor system into feed bins in the primary extraction plant. As the first step in the extraction process, tar sand, caustic, and water are mixed and heated to 170 - 190°F with open steam. This operation is carried out in rotating "conditioning" drums similar in mechanical configuration to rotary kilns or rotary steam tube driers. The conditioned pulp is screened to separate rocks and lumps of unconditioned tar sand. After the addition of enough fresh and/or recycle water to convert the screened pulp to a pumpable slurry, the screen undersize is transferred to separation cells. Here an oil rich emulsion of bitumen, fine mineral, and water rises to surface and overflows into a froth collection system. Sand settles to the bottom and is discharged by a rake mechanism similar to that used in thickeners. Excess water is withdrawn as a "middlings" stream from the center of the separation cells. Part of the middlings is recycled to dilute the screen undersize and the rest is processed through air flotation scavenging cells to recover additional bitumen. Froth from the scavenger cells is an unstable water-rich emulsion of bitumen, fine mineral, and water that, upon settling, will break down into a bitumen-rich emulsion similar to that produced in the separation cells and a bottoms layer of about the same composition as middlings. Settler tailings are recycled to extinction and the settled scavenger froth is combined with the separation cell froth. This combined froth, still containing substantial quantities of fine mineral matter and water, is the product from the primary extraction step. Primary extraction tailings are diluted with cold water and pumped to the tailings disposal area.

To upgrade the froth from the primary extraction step so that it will be suitable for processing into synthetic crude, it is heated to 190 - 200°F with open steam, diluted with naphtha to reduce the viscosity and density of the hydrocarbon phase, and centrifuged.

Figure 9 (5) shows these steps schematically.

In the commercial plant, there are:

1. A four-compartment feed bin.
2. Four conditioning drums.
3. Four vibrating screens.
4. Four separation cells.
5. Thirty-six air flotation cells.
6. Two scavenger froth settlers.
7. Forty centrifuges.

To compensate for feed variations, these units are sized so that three-fourths of the installed equipment can handle the nominal capacity.

Many aspects of this approach to the recovery of bitumen from tar sand have been the subject of considerable research and small-scale plant operation over the past 30 years (8, 9, 10, 11). However, this early work did not supply sufficient information on the basic mechanism of the separation process to permit design of the G. C. O. S. plant without additional laboratory and pilot plant study (5).

The commercial plant equipment design correlations were derived by first establishing a theoretically reasonable process model and then empirically fitting the experimental data to define the constants and coefficients that would give a statistically sound representation of the relationship between equipment size and design, feed stock characterization, and process performance. Our interest at this time is in how well these design correlations represent commercial plant performance:

1. Bitumen recovery - There are four streams leaving the extraction plant that contribute to bitumen losses. These are: screen oversize, separation cell tailings, scavenger cell tailings, and centrifuge tailings. Table 1 compares plant measured losses with expected losses calculated from the design correlations.

Overall, the commercial plant gives about the recovery we expected from pilot plant performance. The only difference of any substance is in the screen oversize loss, and note that we did not have a very precise prediction for this value. Reference (5) discusses this point in considerable detail. Briefly, this is the story. Screen oversize results from tar sand lumps not broken down in the conditioning drum. The two most important factors affecting this breakdown of lumps are the rate of heat transfer in the conditioning drum and the feed lump size distribution. For our predictions, we were far from certain what the feed lump size would be. Furthermore, during winter operation some of the tar sand lumps will be frozen and this will change the heat transfer characteristics of the system. Consequently, we still do not have a complete picture of what the average year-round oversize loss will be.

Table 1 is based on averaged values for a number of steady-state operating cases. Figures 10 and 11 show in more detail operating data on the bitumen distribution in the separation cells and scavenger cells for these cases. Note the variations in the expected values. The principal reasons for this are variations in feed bitumen content, feed fines content, and feed fines to fresh water ratio.

TABLE 1

EXTRACTION PLANT BITUMEN DISTRIBUTION (7,8/67)

Basis:
a. Average bitumen in tar sand = 11.8 wt.%
b. Average < 325 mesh fines in the tar sand mineral fraction = 20.1 wt.%
c. Average water input = 0.58 tons per ton of tar sand
d. Average operating temperature = 185°F

	WT.% OF BITUMEN IN TAR SAND	
	Measured	Expected
LOSSES		
Screen oversize	< 0.1	0.5 to 3.0
Separation cell tailings	8.3	7.9
Scavenger cell tailings	1.6	2.0
Final extraction tailings	2.2	1.8
PRODUCT	87.9	87.8 to 85.3

As feed bitumen content varies the loss to separation cell tailings, expressed as per cent of the feed bitumen, varies inversely because this loss is (approximately) constant per ton of tar sand. As feed fines vary at a fixed fines to water ratio, the volume of middlings changes in direct proportion without much change in middlings composition. Thus more or less bitumen leaves the separation cell in the middlings stream. In a similar manner, a change in the fines to water ratio affects the bitumen leaving the separation cell in the middlings, but the relationship is more complex because this also changes middlings composition. Ideally, the process should be operated at a maximum fines to water ratio as discussed in (5) but during startup this wasn't always possible.

Our predictions of bitumen distribution in the scavenger cells are based on residence time (5). Thus any changes in the separation cell that affect middlings rate would have an inverse effect on scavenger froth yield. We've had some indication that bitumen concentration in the middlings stream should be a parameter, but this relationship hasn't been fully developed.

Within practical operating limits, we expected the final extraction plant bitumen losses to be a fixed percentage of the final extraction plant feed, so the only variation in relating this loss back to bitumen in tar sand is a slight correction for changes in primary extraction recovery. We don't consider the small difference between the measured and expected values significant. Our test plant data, used as a basis for the expected loss had a standard deviation of $\pm$ 1% and the standard deviation of the individual values that make up the average measured plant loss is $\pm$ 1.6%.

2. Primary Extraction Plant Froth Composition: On the average, the primary extraction plant froth was richer in bitumen than expected:

TABLE 2

FROTH COMPOSITION (7,8/67)

Basis: Same as Table 1.

	Actual (Wt.%)	Expected (Wt.%)
Bitumen	57.3	45.9
Mineral	6.5	9.3
Water	36.2	44.8

We do not know why the commercial plant produces this better quality froth. Figure 12 shows the individual data points that make up the average value, from the same operation as the bitumen distribution values already discussed. Note that we did not expect a great deal of variation in the froth composition. What small differences we did expect were related to the bitumen and fines content of the tar sand. In addition to a better average composition in the commercial plant product, there is a much wider variation.

3. Final Extraction Plant Product Composition - Even after the two stage centrifuge operation, the product contains some fine mineral matter and water. It is desirable to minimize these. Residual mineral is concentrated in coke produced in subsequent processing. This coke is the primary source of plant fuel, and excessive mineral makes it more difficult to pulverize and increases furnace ash. Residual water must be vaporized in subsequent processing so minimizing the water content of the product saves heat and reduces foaming problems. Figure 13 shows a comparison of plant product mineral and water contents versus what we expected at these operating conditions from our pilot plant work. The actual mineral content is lower than we expected by a factor of about 2. On the other hand, the product water content is high by about the same factor. We don't know why the mineral content is better, but subsequent to this study we did find the product water content could be improved by modifying the second stage centrifuge internals to get more even feed distribution. After this modification, we observed water contents in the 5 to 10 per cent range rather than the 10 to 15 per cent shown here. We would like to get to 5 per cent maximum water in the centrifuged product.

Tailings Disposal - Tailings from the primary and final extraction plants are pumped as a water slurry to a 700 acre diked tailings pond. Water is recovered from the pond and recirculated through the primary extraction plant. Operation has not progressed far enough to permit evaluation.

Mechanical problems - One of the major mechanical problems so far has been the tendency of the rich tar sand (>13% bitumen) to cause bridging in the feed bin. To a lesser degree, plant conveyors have encountered many of the same problems as the mine conveyors. Erosion is always a problem handling this type of material; consequently, metal wear plates or elastomer liners in critical areas were part of the original design. In some cases, it has been necessary to build up the original wearing surfaces or change to alternate materials.

At this point in the process, we have as a product a heavy (8°API) viscous (500 SSU at 210°F) oil diluted with light hydrocarbon so it can be pumped and stored in tanks. The work discussed so far, starting with mining and continuing through both extraction stages, is an adaptation of mine - ore mill technology. To upgrade this product to a suitable substitute for conventional crude oil requires further processing based on petroleum refining technology.

Upgrading Bitumen to Synthetic Crude Oil

The diluted bitumen product from the final extraction step is upgraded to synthetic crude oil by first separating the light naphtha diluent, then coking the bitumen, and finally hydrorefining the coker distillate fractions. In addition to the finished synthetic crude, these operations also yield coke for power plant fuel, gas for hydrogen plant feed and furnace fuel, supplementary fuel oil as required, make-up diluent for the final extraction plant, and hydrogen sulfide which is further processed to recover elemental sulfur.

From intermediate storage, the diluted bitumen is charged to the diluent recovery system. This unit closely resembles a conventional crude oil distillation system. Diluent is recovered as an overhead product and returned to storage for recycle through the final extraction plant. Bitumen is recovered as the bottoms product. The capacity of the system is 135,000 barrels of combined diluent and bitumen per stream day.

Bottoms from diluent recovery are fed to the coker furnaces at approximately 500°F. In passing through these furnaces, the temperature is raised to 900 - 910°F to initiate thermal cracking. From the furnaces, the bitumen discharges into delayed cokers. Coke accumulates until a drum is filled. Then that drum is cooled down, steamed to strip out residual hydrocarbon vapors, the coke dumped, and the drum preheated to prepare it for the next cycle. During the filling cycle, the hydrocarbon vapor products are continuously withdrawn. To provide a continuous flow of coker vapor product, three furnaces and three pairs of 26 foot diameter coke drums are used. At any given time, three drums (one for each furnace) are taking feed, one is cooling down, one is discharging coke, and one is being preheated.

Vapor from the coker is processed through a fractionating tower to separate it into wet gas, unstabilized naphtha (375°F end point), kerosene (375°F - 500°F), gas oil (500°F to 850°F), and bottoms (850°F+). The wet gas and naphtha fractions are further processed in a gas plant to give a C_3 and lighter gas for plant fuel and hydrogen production, a C_4 fraction which can be used as supplementary fuel or blended into synthetic crude to control its vapor pressure, and a stabilized C_5 - 375°F naphtha fraction for hydrorefining to a synthetic crude component. The kerosine and gas oil fractions from the coker fractionator are fed directly to hydrorefining units for upgrading to synthetic crude components. The heavy bottoms fraction can either be used for furnace fuel as needed or recycled through the cokers.

The following outline is a brief summary of our experience with this portion of the G.C.O.S. plant.

1. Diluent recovery - In our discussion of the final extraction plant, we briefly mentioned the problem with foaming resulting from excess water in the diluted bitumen. This foaming problem was exaggerated in our initial operation because of the tendency of the high water content product to stratify in storage. In addition to the work in final extraction to reduce the total quantity of water, tank mixers were installed to minimize stratification.

2. Coking - Initial operation of this unit has been most satisfactory, because several potential problems that we anticipated have not so far materialized. We thought residual clay might create tube fouling problems in the coker furnaces. But after three months of operation, we inspected the tubes and found no significant buildup. Since then, there has been no measurable increase in furnace tube pressure drop. We were also concerned over the potential mechanical problems with the high pressure water jet system used for cutting coke because it had never been used in coke drums this large. However, except for one case of a water pump jammed by dirt from the water (we now use clarified water), this has not been a problem. The coke has contained more water than anticipated, thus increasing the draining and air drying time before it can be transferred to the power house bunkers, but this is not a serious problem.

The only significant operating problem related to the design of the coker system is overquenching in the bottom section of the coker fractionator. This has reduced the end point of the gas oil fraction and resulted in a loss of gas oil yield. Mechanical modifications are planned on the first opportunity and should correct this problem.

The following table compares expected and actual coker product distributions:

TABLE 3

COKER PRODUCT DISTRIBUTION (9/67)

Component	Measured wt.%	Expected wt.%
Gas (C_4 & lighter)	7.9	8.3
Naphtha	12.7	12.1
Kerosine	15.0	10.0
Gas Oil	36.2	41.4
Fuel Oil	6.0	4.2
Coke	22.2	22.7

The discrepency in gas oil yield is a reflection of the fractionator problem we just discussed. The excessive recycle that results contributes to increased kerosine yield by overcracking. The remaining kerosine comes from high end point naphtha that was used as the initial diluent inventory. Part of this material was not recovered in the diluent recovery system, thus showing up in the coker product.

3. Hydrorefining - Processing problems were minimal at startup. However, we did have a number of mechanical problems. Perfecting the seal oil system for the hydrogen recycle and booster compressors was a major one that has since been overcome.

Table 4 shows inspection data on hydrorefined synthetic crude components. The composite synthetic crude from a blend of these components is shown in Table 5, along with comparable data on pilot plant product (7). The principal differences are the low end point and the greater quantity of material in the 375 - 500°F boiling range that are results of the distillation problem we discussed earlier. When this is resolved, we think the products will be comparable.

TABLE 4

HYDROREFINED SYNTHETIC CRUDE COMPONENTS (9/67)

	NAPHTHA		KEROSINE		GAS OIL	
	Commercial Plant	Pilot Plant(7)	Commercial Plant	Pilot Plant(7)	Commercial Plant	Pilot Plant(7)
API Gravity @ 60°F	55.3	50.9	38.6	39.7	27.5	28.7
Distillation (D-86)						
IBP	162	174	358	388	498	499
5	194	260	385	398	526	512
10	206	274	398	402	540	522
30	238	282	418	411	568	561
50	278	296	438	415	588	611
70	316	310	460	423	615	655
90	369	334	496	433	675	740
95	396	344	513	448	706	785
EP	462	366	533	468	715	869
Aromatics, vol.%	-	18	12.7	13.8	25.3	29.8
Sulfur, ppm	15	50	50	50	410	800
Yield, vol.% of synthetic crude	30.8	30.6	27.2	19.0	42.0	50.4

TABLE 5

G.C.O.S. SYNTHETIC CRUDE (9/67)

	Initial Commercial Production	Pilot Plant Product
Gravity, °API	38.3	37.6
Distillation (D-86)		
IBP	162	210*
5%	221	277
10%	254	300
30%	408	379
50%	507	386
70%	588	574
90%	615	688
95%	675	738
EP	715	833
Sulfur, wt.%	.022	.030

* Pilot plant product was stabilized to C_6+, but commercial plant product is C_5+. In preparing the pilot plant sample approximately 5 vol.% of a 174°F to 210°F fraction was stripped off. There isn't any real difference in the front end.

Hydrogen Production

Amine treated coker dry gas is used as raw material for production of 63.5 mm CFD of 95.0% H_2. Residual sulfur is removed by caustic washing, then the coker gas passes over a cobalt molybdenum catalyst in order to saturate olefins and convert mercaptans to hydrogen sulfide. This H_2S is then removed by passage through a zinc oxide bed at 700°F. The gas is then mixed with steam and passed over the reforming catalyst at furnace outlet conditions of 1525°F and 300 PSIG. The gas is quenched and then passed through high and low temperature shift converters. Carbon dioxide is removed by absorption in a promoted potassium carbonate solution.

The hydrogen plant was started up on natural gas. There were some problems with leaks in the system, but once these were corrected the unit performed well. Recently, the hydrogen plant operation was changed to plant gas, but we do not have any performance data yet.

Sulfur Production

Off gases from the coker and hydrorefiners are treated with monoethanolamine to remove hydrogen sulfide. The recovered H_2S is converted to sulfur in a conventional two stage oxidation plant. Little contamination of amine solution by carbonyl sulfide has been observed. Filtering and reclaiming facilities are used to maintain a clean amine system.

Utilities

The powerhouse provides all utilities for the project. Three coke fired boilers supply 2,250,000 pounds per hour of steam at 800 PSIG and 750°F. This steam is exhausted through two turbo-generators with a combined capacity of 76,500 KVA at 13.8 KV. Steam is extracted at 425 PSIG and exhausted at 50 PSIG. The higher pressure steam is used to power turbine drives in the powerhouse and refinery. The 50 PSIG steam is used primarily in the extraction plant.

The major initial operating problems were excessive superheating of the steam, creating metallurgical problems with steam piping, and fouling of boiler tubes on the fire side. Removal of some superheat tube surface eliminated the first problems. Installation of soot blowers and reduction of the ball load in the coke pulverizers to avoid over grinding coke reduced tube fouling. Control of the turbo-generators is a continuing problem, because many of the plant electrical loads are intermittent. For example, when a bucketwheel excavator completes a cut and starts repositioning itself power demand drops substantially and, the powerhouse must adjust to handle the reduced load. In as little as five or ten minutes, when the bucketwheel is repositioned, it must shift again to pick up the load.

Water treatment facilities for boiler feed water consists of a clarifier, hot lime reactors, filters and ion exchange. It is the largest boiler feed water plant in Canada and it can maintain total make-up to the boilers if required. The rather elaborate water treating facilities are necessary because at times the Athabasca River contains up to 2,500 ppm turbidity and substantial quantities of silica that the high pressure boilers cannot accommodate. This unit has operated satisfactorily.

Overall Process Evaluation

As we noted in our earlier remarks, it is too early for a complete technical evaluation of the G.C.O.S. plant. However, we do feel that this preliminary look shows that the basic process concepts are sound. The hot water extraction process had been demonstrated only at a hundred-fold smaller scale. The scaled up version performs about as expected. Coking and hydrorefining had never been applied to bitumen and products derived from bitumen except in pilot plant equipment. These also perform about as expected. What we cannot adequately evaluate is the long term mechanical reliability of these processing schemes. We've noted the major mechanical problems encountered so far, some apparently solved and others we still have. It will probably take many months, perhaps even a year or two, for a definitive mechanical evaluation to take shape.

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